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EXPERIMENTAL STUDIES OF GAS-LIQUID DISPERSION
IN A SPLIT-CYLINDER AIRLIFT FERMENTOR

by

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INTRODUCTION

Gas-liquid dispersions are important in many chemical engineering operations, involving heat and mass transfer. Dispersions are produced by injecting gas into liquids by means of a sparger. Surface tension forces cause the gas to break into bubbles, which may further break up due to turbulent or shear forces. Agitation is created in airlift columns by sparging gas at a high rate.

Types of airlift fermentors include the split-cylinder airlift (Orazem et al. 1979; Belfield 1976), the concentric cylinder airlift (Hatch 1973, Blenke 1979), the external return flow airlift (Kanazawa 1975), airlift fermentors with motionless mixers (Viesturs et al. 1980; Voigt et al. 1979), and multistage airlift columns (Orazem et al. 1979).

In the split-cylinder airlift column used in this work, density differential is created by sparging air into one of the two regions. This provides the driving force for liquid circulation. The liquid circulation depends on the difference between the gas holdups in the two regions. The interdependence of these variables complicates the design and modeling of an airlift.

Gas holdup is closely related to the bubble size distribution in the airlift. Smaller bubbles give rise to larger interfacial areas per unit volume, and high overall oxygen transfer coefficients.

The factors which affect bubble size are fluid properties, such as surface tension and viscosity, and the presence of surfactants and electrolytes in the fermentation medium, as well as the operational and geometric factors. Fluid properties affect the rate of coalescence of bubbles. It is desirable to have low coalescence rates, so that small bubble size can be maintained.

In this work, the effects of electrolytes (sodium chloride) and surfactants (sodium lauryl sulfate) on gas holdup are examined by measuring the gas

holdups at various positions in the fermentor, using single tube manometers. The effect of viscosity was studied by performing similar experiments on CMC solutions of viscosities up to 4 poise. The combined effect of viscosity and surfactant was examined by adding 10 ppm sodium lauryl sulfate to CMC solutions of various concentrations. Experiments were carried out for different liquid heights.

The following section covers a brief review of related work that has been reported in literature. The description of experimental methods and a discussion of the effects of viscosity, surfactants and electrolytes on the circulation rates are included in the next two sections. Conclusions and recommendations are listed in the final section.

LITERATURE REVIEW

Previous works on airlift fermentors and related systems are reviewed. The behavior of gas-liquid dispersions in airlift fermentors depends on a number of factors. One of the important factors, which governs bubble size, is turbulence in the liquid, as it causes bubble break-up. This, along with the opposing factor, the rate of coalescence of bubbles, determine the dynamics of bubble size distribution. Both of them depend on physical properties of the liquid. The background of the problem and related work reported elsewhere is described in the following pages.

Turbulence:

In aerobic fermentations, turbulence in the liquid is important in order to bring about bubble break-up. In mechanically agitated tanks, break-up is induced by the turbulence which is created by means of an impeller. In an airlift fermentor it depends on the turbulence of the fluid flow field, which depends on the power per unit volume associated with the airflow.

Turbulent flow produces primary eddies which have a scale of similar magnitude to the dimensions of the main flow. The large eddies break down into smaller eddies due to vortex stretching. The motion of small eddies is mainly governed by viscous forces, which then dissipate the very small eddies into heat (Tennekes and Lumley 1978). Kolmogoroff has postulated that most of the energy dissipation occurs at the smallest scales. He has concluded that local isotropic turbulence exists even though the turbulent motion of the larger eddies may not be isotropic.

The approximate scale of the largest eddies is given by

$$l \sim d \quad (1)$$

where d is the characteristic dimension of the vessel.

The scale of energy dissipating eddies is given approximately by

$$\eta = \frac{\mu_L^{3/4}}{\rho_L^{1/2}} \left[\frac{P}{V} \right]^{-1/4} \quad (2)$$

where η is the Kolmogoroff microscale, and $\frac{P}{V}$ is power per unit volume of the upflow section, which is given by (Bhavaraju 1978)

$$\frac{P}{V} = \frac{4Q_M \rho_L g}{\pi d^2} \quad (3)$$

Turbulence is present if the scale of the energy containing eddies is much larger than the scale of the energy dissipating eddies. The ratio of the two scales is defined as

$$\frac{\ell}{\eta} = \frac{d \rho_L^{1/2}}{\mu_L^{3/4}} \left[\frac{P}{V} \right]^{1/4} \quad (4)$$

If turbulence is present, the mean bubble diameter in the liquid is given by (Calderbank 1967)

$$d_{BM} = C_1 \frac{\sigma^{0.6}}{\left(\frac{P}{V} \right)^{0.4} (\rho_L)^{0.2}} \left[\frac{\mu_L}{\mu_G} \right]^{C_2} \quad (5)$$

where C_1 and C_2 are empirical constants. This correlation is valid when

$$\ell \gg d_{BM} \gg \eta .$$

Bhavaraju et al. (1976), from their experiments on 0.1% carbopol solution (viscosity ~ 0.3 p), have suggested that $\frac{\ell}{\eta}$ should exceed 200 for turbulent conditions.

When turbulence is absent, at high viscosity, bubble size is governed by the conditions at the orifice, and the bubble diameter is given by (Bhavaraju et al. 1978)

$$d_B = 3.23 \frac{d_0^{0.05} \mu_L^{0.1} Q^{0.31}}{\rho_L^{0.1}} \quad (6)$$

This indicates that the bubble size is governed mainly by the gas flow rate per orifice. For highly viscous liquids, therefore, sparger design may be important.

Bhavaraju et al. (1978) have further shown that for both high and low viscosity liquids, bubble breakup is related to liquid turbulence in the vessel rather than gas turbulence in the orifice.

The power input per unit volume in sparged vessels is insensitive to changes in viscosity [see eqn. (3)]. The flow behavior of the two phase mixture depends on the physical properties of the fluids and the power input per unit volume. For turbulent pipe flow, Taitel et al. (1980) have expressed the rate of energy dissipation per unit mass as

$$\varepsilon = \left| \frac{dp}{dz} \right| \frac{V_M}{\rho_M} \quad (7)$$

where

$$\frac{dp}{dz} = \frac{2f}{D} \rho_M V_M^2 \quad (8)$$

and

$$f = 0.046 \left(\frac{V_M^D}{v_L} \right)^{-0.2} \quad (9)$$

For turbulent wake behind a bubble in a power-law liquid, viscous dissipation is important. Lumley (1964) has discussed the mechanism of turbulent energy dissipation in a non-Newtonian liquid.

Bubble breakup:

Fundamental work on bubble breakup was done by Hinze (1955). When a gas is injected into a liquid, the potential energy of the gas phase is converted into kinetic energy, and simultaneously the two phases interact. It generates strains in the dispersed phase, which may cause it to break into smaller fragments. In the case of turbulence it is the kinetic energy of the turbulent motion in the continuous phase that brings about the breakup.

The disintegration takes place in three stages. First, the large bubbles are deformed. They break into medium size bubbles and later into smaller bubbles until equilibrium is reached. The type of deformation depends on dynamic pressures, viscous stresses, and flow pattern around the globule. If the flow is viscous, the deformation is lenticular, the tangential viscous forces on the interface cause a toroidal internal circulation. This produces normal forces at the interface. These along with the external normal component of the viscous stresses, deforms the globule. A preference for a given type of deformation and breakup depends on the local flow pattern, and physical properties of the two phases, such as densities, viscosities, and interfacial tension.

The final breakup may occur in a turbulent flow field by local viscous shearing action, or by dynamic pressure fluctuations over distances equal to the bubble diameter. The deformation gives rise to internal flows which cause viscous stresses $(\frac{\mu_G}{d_B} \sqrt{\frac{\tau}{\rho_G}})$ as well as dynamic pressures (τ) . The inter-

facial tension σ gives rise to a surface force that counteracts the deformation. The surface tension force is of the order σ/d_B .

The surface tension force around the circumference of the bubble is $\pi d_B \sigma$. This acts on an area of $\pi d_B^2/4$. Thus, the pressure holding the bubble together is $4\pi d_B \sigma / \pi d_B^2 = 4\sigma/d_B$. The ratio of the dynamic pressure to the surface tension is defined as the Weber number.

$$We = \frac{\tau d_{Be}}{\sigma} \quad (10)$$

This determines the equivalent diameter d_{Be} of the largest bubbles. It is the maximum diameter which exists at dynamic equilibrium in a turbulent flow field. A higher Weber number leads to greater deformation. The critical Weber number is defined as the point where splitting occurs. It is a function of the viscosity group as shown below.

$$N_{vi} = \frac{\mu_G}{\sqrt{\rho_G \sigma d_B}} \quad (11)$$

In general, the critical Weber number is characteristic of the dispersion equipment and physical properties of the fluids.

Grace et al. (1978) developed an empirical model, based on the 2-dimensional linearized Rayleigh-Taylor stability theory to describe the breakup of bubbles and drops in stagnant fluids.

Bubble breakup may occur in an airlift fermentor, either by viscous shearing or due to turbulent conditions in the liquid. While bubble breakup decreases the mean bubble size, coalescence of bubbles increases it. The final bubble size is determined by a balance between the two processes.

Coalescence of bubbles:

In gas-liquid contacting equipment, bubble coalescence is an important factor. The coalescence rate is a significant factor in determining the bubble size, interfacial area, gas residence time, and the mass transfer efficiency of the equipment. A low coalescence rate leads to smaller bubbles and higher interfacial area per unit volume.

Coalescence is a two-step process. First, the bubbles approach to within 10^{-4} - 10^{-3} cm. This is followed by thinning and rupture of the liquid layer between the bubbles (McBride, 1979). If time available for coalescence is greater than film-thinning time, coalescence takes place. McBride et al. (1979) have shown that in laminar flow, film-thinning time increases with increasing apparent viscosity. In laminar flow, the force due to excess pressure causes coalescence, whereas in turbulent flow, inertial force of the liquid moving the bubbles together, causes coalescence.

In stagnant fluids, the wake capture mechanism prevails. Calderbank et al. (1965) have concluded, from their experiments, that for liquids with viscosities greater than 70 cp, coalescence is very rapid. This is due to a longer wake behind the leading bubble, when the liquid is more viscous (Acharya and Ulbrecht, 1978).

Several models have been proposed to describe the coalescence process. (Happel and Brenner, 1965; Clift et al., 1976; Crabtree and Bridgewater, 1971; DeNevers and Wu, 1971). Most models are based on interaction between two solid spheres. Narayanan et al. (1974) experimentally studied two bubbles of equal size. The mode of coalescence was found to depend on the wake structure.

The interaction is more complex when bubbles are deformed, as in the case of highly viscous and non-Newtonian liquids. The model for bubble coalescence in viscous liquids, which has been proposed by DeNevers and Wu (1971), assumes that the only forces acting on the trailing bubble are the buoyancy force and the drag. The drag decreases as the bubble enters the wake of the leading bubble. The trailing bubble gets accelerated and then gets caught in the wake.

Bhaga (1979) performed experiments on bubbles in highly viscous liquids. He studied the interaction and coalescence of 2 vertically aligned bubbles, photographically. He observed that if the separation between the bubbles was too large or if the trailing bubble was too small compared to the leading bubble, the height of the column was insufficient for coalescence to occur. The trailing bubble has relatively little influence on the velocity and the shape of the leading bubble, but the wake of the leading bubble has a significant effect on the velocity and shape of the trailing bubble. The behavior of the bubble formed by coalescence depends on the viscosity of the liquid as well as the sizes of the interacting bubbles.

Otake et al. (1977) studied coalescence and breakup of bubbles in liquids, using a photographic technique. They observed that there was a critical distance at which the leading bubble began to exert a noticeable influence on the following one. The condition for the occurrence of coalescence was determined in terms of overlap between the projected areas of the two bubbles.

Hatch (1973) has shown that for an airlift system, at moderate superficial gas velocities (between 30 and 600 cm/min), coalescence begins to influence the bubble size distribution more than the orifice diameter. At higher gas flow rates the gas holdup is increased and consequently coalescence is also increased.

Coalescence within a fermentor leads to variation of bubble size and holdup, with position. Larger bubbles are formed as a result of coalescence. Sufficiently large bubbles move upward in the downflow region, instead of being carried downward by the circulating liquid. This leads to further coalescence, which results in an accumulation of bubbles at the top, and fewer bubbles at the bottom. On the other hand, hindrance of coalescence would lead to a relatively uniform distribution of bubbles throughout the downflow region. Coalescence in the upflow region may form bubbles that are too large to be carried into the downflow region. They may escape into atmosphere.

Effect of electrolytes and surfactants on bubble size and rates of coalescence:

In fermentation, both inorganic and organic compounds are frequently present in the broth. The effects of these compounds on the gas-liquid dispersion in airlift fermentors has not been systematically investigated. Electrolytes and surfactants are two groups of substances which are known to affect the gas-liquid dispersion in airlift fermentors. In order to understand their behavior in airlift fermentors, the literature on electrolytes and surfactants in simple gas-liquid dispersions is examined.

The addition of small quantities of some organic and inorganic compounds to water decreases the bubble size even when other physical properties of the liquid, such as viscosity and surface tension, remain unchanged. (McTaggart 1914; Marrucci et al. 1967).

Effect of electrolytes:

A systematic investigation of bubble size and coalescence was made in the 1920's (Foult 1929; Foult et al. 1929; Schnurman 1929), when a theory of liquid film formation was developed, while investigating foaming and priming of boiling water. It has been suggested that solutes tend to concentrate on or recede from liquid surface, whenever there is a difference in concentrations

of the surface layer and bulk liquid. In either case the movement of matter requires work, hence there is a tendency to resist mixing. In subsequent works on porous plates, workers mentioned the effect of electrolytes on bubble size (e.g., Siemes et al., 1960 ; Yoshida et al. 1965 ; Burckhart et al. 1974), but no effort was made to determine the mechanism of the phenomenon, until 1967, when Marrucci and Nicodemo (1967), reported their results on coalescence of bubbles near a porous distributor plate. They experimented with various electrolytes of different valence and concentrations. Their observations indicated that the electrical repulsive forces hinder coalescence between bubbles, and that the coalescence process is influenced by the gas flow rate and the concentration and valence of the solute.

Their experiments on nonelectrolytes (ethyl alcohol solution of different concentrations) also indicated a decrease in bubble size with increasing solute concentration. Later studies (Zieminski and Whittemore, 1971 ; Lessard and Zieminski, 1971) on the effects of salts on interfacial area of dispersion and oxygen transfer coefficient indicate a strong dependence of these quantities on valence of the ionic species and salt concentration. The results of the observations were utilized to suggest a mechanism for coalescence-prevention due to the salts, in terms of ion-water interactions.

Zieminski and Whittemore (1971) allowed bubbles from a disperser to rise freely in a plexiglass column, containing solutions of various electrolytes. The bubble size was determined by photographic methods. The rate of oxygen absorption was also measured. They concluded that the mechanism of bubble coalescence in electrolyte solutions is not a unique function of electric forces produced by surface potentials, but is a more complex phenomenon depending on the effects of ion-water interactions.

An ion introduced into water interferes with its structure. The effects may be steric in nature, and hence depend upon size of the ions. The ions affect neighboring water molecules, loosening or tightening the structure. This may decrease or increase viscosity. Zieminski and Whittemore (1971) found that small and highly charged ions are the most effective in hindering coalescence. Since coalescence rates depend on the draining of film between bubbles, ions which are structure makers (and hence increase viscosity) will increase the rigidity of surface film and decrease coalescence.

Lessard and Zieminski (1971) studied bubble coalescence and gas transfer in aqueous electrolytic solutions. They contacted pairs of gas bubbles, in order to evaluate the coalescence percentage as a function of solute concentration. They found sharp transition concentrations, at which, the coalescence percentage decreased rapidly.

They found a very satisfactory correlation between the degree of coalescence and ionic entropy of hydration. Increase in entropy can be related to a decrease in viscosity (Gurney, 1953) and increase in coalescence. This appears to support the mechanism of coalescence based on ion-water interactions.

The effect of electrolyte concentration on the mass transfer coefficient is quite small (Lessard and Zieminski 1971). At higher concentrations, the solutions show a much higher oxygen transfer coefficient ($K_L a$) in stirred tanks (Zlokarnik, 1978). This may be due to an increase in interfacial area.

Effect of surfactants:

A surface active agent or surfactant is a substance that has both hydrophilic and hydrophobic molecular sections, and is attracted to the interface between two unlike phases. The presence of these molecules at an interface reduces interfacial tension. This property is important in an operation

involving dispersion of phases, as it could result in an increase in the interfacial area.

The effect of surfactants on coalescence rates and oxygen transfer coefficient has been studied (e.g., Mancy and Okun, 1963; McKeown and Okun 1963; Eckenfelder and Barnhart, 1961; Aiba and Toda, 1963; Carver, 1955). It was found that the addition of sodium lauryl sulfate to water (Eckenfelder et al., 1959) resulted in a sharp decrease in $K_L a$ at low concentrations, and a gradual increase at higher concentrations. The increase in $K_L a$ was attributed to the decrease in bubble size, and the resultant increase in interfacial area. The effect of various organic substances was studied by Eckenfelder (1961), Zieminski et al. (1960), and Aiba et al. (1961). It was discovered that the increase in the interfacial area was much greater than the decrease in K_L .

The hydrodynamic behavior of a 2-phase system can be changed by the addition of surfactants (Levich, 1962). The adsorption of a surface active material on the surface of a drop (or bubble) has a noticeable effect on its motion. Theoretical studies on the retardation effects of surfactants on the velocity of a drop were done by Levich (1962), Newman (1967) and Gal-or and Waslo (1968). Gal-or and Waslo (1968) developed a free surface model, which was a generalization of the earlier models. It takes into account, the interaction between bubbles. The effect of gas holdup and surfactants on the velocity of ensemble is quantitatively expressed. A detailed account of theoretical studies is given by Resnick and Gal-or (1968).

The presence of surfactant tends to stop the circulation in the bubble, due to the formation of a rigid film. Hammerton and Garner (1954) showed, from their experiments on air bubbles in various liquids, that the presence

of surface active material can change a mobile surface of a bubble to a rigid one. This leads to a decrease in mass transfer coefficient.

The effect of surfactants on the mechanism of coalescence was reported by Lindblad (1964), who compared the coalescence rates of air bubbles in fresh distilled water to those in water stored for 30 hours. Voigt and Schügerl (1979) pointed out that electrolytes reduce coalescence immediately following bubble formation while surface active agents reduce coalescence only after the bubble has been surrounded by surfactants.

Previous work on bubble coalescence in the presence of electrolytes and surfactants indicates that higher interfacial areas and gas holdups are possible in fermentation broths. Related work has been done for bubble columns (Burckhart and Deckwer, 1974 ; Voigt and Schügerl, 1979 ; Mashelkar and Sharma, 1970 ; Abdel-Aal et al., 1966) . The upflow region of an airlift fermentor is similar to a bubble column. There appears to be a need to explore the downflow region, with respect to the effects of electrolytes and surfactants on gas holdup.

Effect of viscosity and non-Newtonian behavior:

In recent years, industrial fermentations involving highly viscous, non-Newtonian fermentation broths have received considerable attention. Work on antibiotic fermentations involving mycelial broths has been done, and the effect of liquid physical properties on oxygen transfer has been reported. (Tuffile and Pinho, 1970 ; Diendoerfer and Gaden, 1955 ; Maxon, 1959 ; Steel and Maxon, 1961 ; Wang and Fewkes, 1977) . Some work has been reported on highly viscous, polysaccharide fermentations. (Swallow et al., 1978 ; Leduy et al., 1974) . The high power requirements appear to pose a problem in the design of the fermentor. Charles (1978), in a review on technical aspects

of the rheological properties of microbial cultures, has noted that the turbine impeller may not be the optimum design for non-Newtonian cultures.

The problems in the scale up of such media were considered by Bartholomew (1960). According to him, the success of the scale up is based on equal mass transfer coefficients. This condition is not satisfied by some broths, such as the penicillin fermentation broth. The increase in viscosity leads to a decrease in gas-liquid interfacial area and the volumetric mass transfer coefficient. The mechanically agitated system has the disadvantages of high energy requirements and the adverse effect on the growth patterns of the organism. Airlift fermentors are, therefore being considered as low-energy alternatives to stirred tanks, for fermentations involving highly viscous broths (Snedecor and Wang, 1976; Malfait et al., 1980).

The nature of fermentation broths has been reviewed (Taguchi, 1971; Blanch et al., 1976). Many fermentation broths display pseudoplastic behavior, and a few display Bingham plastic behavior.

Several rheological models have been developed to describe pseudoplastic liquids (Wilkinson, 1960). The simplest and the most widely used one is the power law model, which may be stated as

Shear stress $\propto$ (shear rate)ⁿ⁻¹ for a one-dimensional case

$$\eta_{app} = K \left| \frac{dv}{dr} \right|^{n-1} \quad (12)$$

where η_{app} is the apparent viscosity, K is the consistency index (in pseudo poise), n is the power-law index, and $\frac{dv}{dr}$ is the shear rate. K and n are constant over limited ranges of shear rate. Hence, it is necessary to specify shear rate, in reporting rheological constants for a liquid.

Bingham plastics are fluids which exhibit Newtonian behavior once flow is initiated; that is, once the yield stress is exceeded. Mathematically, the behavior is expressed as

$$\tau = \tau_y + \eta_p \left| \frac{dV}{dr} \right| \quad (13)$$

where τ_y is the yield stress, and η_p is the plastic viscosity.

Roels et al. (1974) have developed a model which is based on the Casson equation. It incorporates cell morphology and cell concentration. Charles (1978), in a review article, has listed rheological properties of many mycelial culture fluids, which have been reported in literature.

The difficulties in the design of gas sparged contactors involving highly viscous media have been considered by Bhavaraju et al. (1976), and a liquid circulation model has been proposed. The hydrodynamics of an aerated column depends on the manner in which the rising bubbles influence the liquid motion, the interfacial behavior of the bubbles and liquid phase turbulence (Bhavaraju, 1978). All these factors are important in the design of an airlift fermentor. In highly viscous fermentation media, both rise velocity and liquid phase turbulence are affected.

Bubble rise velocities are largely governed by density differences, bubble size and liquid viscosity, and to a lesser extent, by the surface tension of the liquid in which they rise. Literature on bubble rise velocities in low viscosity liquids is abundant. (e.g., Haberman and Morton, 1956; Garner, 1954; Van Krevelen and Hoftijzer, 1950; Rosenberg, 1955). Peebles and Garber (1953) suggested different correlations for bubble rise velocities in different ranges of Reynolds numbers from their experiments on air bubbles in water. For highly viscous liquids, however, experimental data are relatively scarce. Kojima et al. (1968) made an extensive study of bubble rise velocities as a function of bubble diameter in liquids with viscosities from 1.57 to 103.8 poise. Their data show that the rise velocity of a bubble decreases with liquid viscosity. Redfield and Houghton (1965) reported the rise

velocities of carbon dioxide bubbles in dextrose solution. Angelino (1966) studied the velocities of air bubbles in glycerol solutions. Recently, Grace et al. (1976) presented a generalized correlation for terminal velocities of liquid drops and bubbles. Acharya et al. (1977) recently reported experimental results on drag coefficients for isolated bubbles, bubble shapes, and velocity transition (beyond Stokes region) in non-Newtonian media.

One of the earliest works on motion of gas bubbles in non-Newtonian liquids was that by Astarita and Apuzzo (1965). They suggested the possibility of extending the available theoretical knowledge of the motion of gas bubbles in Newtonian liquids to non-Newtonian liquids, and interpreted the results of their experiments on carbopol solution on the basis of theoretical knowledge. They defined a modified Reynolds number

$$Re_n = \frac{(d_B)^n \rho_L V_r^{2-n}}{K} \quad (14)$$

where n and K are the power-law parameters, and V_r is the bubble rise velocity.

For $n = 1$, this is reduced to

$$Re = \frac{d_B \rho_L V_r}{\mu_L} \quad (15)$$

where $K = \mu_L$ when $n = 1$.

At very low Reynolds numbers, creeping flow takes place. By equating the drag to buoyancy, Astarita and Apuzzo obtained the following equation for the rise velocity of a bubble in a non-Newtonian liquid:

$$V_r = \left[\frac{\rho_L g}{K} \frac{2^{1+n}}{X_n} \left(\frac{4\pi}{3} \right)^{\frac{2-n}{3}} \right]^{1/n} v^{(1+n)/3n} \quad (16)$$

where v is the bubble volume, and X_n is a function of the flow index n .

Wasserman and Slattey (1964) have reported upper and lower bounds for X_n .

For pseudoplastic fluids, V_r increases with increasing bubble diameter more rapidly than for Newtonian fluids.

For $n = 1$, under the condition of Stokes flow,

$$V_r = \frac{g\Delta\rho d_B^2}{18\mu_L} \quad (17)$$

and

$$C_D = \frac{24}{Re} \quad (18)$$

In this flow regime, the bubbles are spherical and rigid, and internal circulation is absent. At $Re \gtrsim 2$, the bubble surface becomes free, and the rise velocity is given by

$$V_r = \frac{g\Delta\rho d_B^2}{12\mu_L} \quad (19)$$

and

$$C_D = \frac{16}{Re} \quad (20)$$

This is known as the Hadamard-Rybczynski equation.

Bubble rise velocity and bubble size are interdependent. Large bubbles rise rapidly and provide little interfacial area. Small bubbles increase interfacial area, but they rise slowly. As a result, a small bubble may be depleted of oxygen, and remain in the liquid without serving any purpose. It is, therefore, necessary to have an optimum bubble size in the airlift.

Bubble rise velocity is important in the calculation of the gas holdup in the fermentor. For low gas holdups (DeNevers 1962) in upflow,

$$\phi_u = \frac{V_s}{V_s + V_r} \quad (21)$$

It is related to the gas holdup in the downflow by the following equation:

$$\phi_d = \phi_u - (\Delta P_{\text{Total}}) g_c / \rho_L g H \quad (22)$$

Interfacial area is related to gas holdup by the equation

$$a = 6\phi/d_{BM} \quad (23)$$

Many workers have reported a jump in the bubble rise velocity in the Hadamard-Rybczynski region. Circulation inside gas bubbles was investigated by visual and indirect means by Garner and Hammerton (1954). Leal et al. (1971) reported a jump in rise velocity of a bubble in viscoelastic-liquids.

Surfactants are known to hinder internal circulation, and make the surface rigid. Calderbank (1967) has shown that the oxygen transfer coefficient is higher for mobile interface, than for a rigid interface.

At low bubble diameters, the bubble interface in CMC solution, is rigid. The stresses set up due to the difference in surface tension between the front and the rear part of the bubble have a retarding influence on bubble motion. With increasing bubble size, these stresses diminish with respect to the viscous stresses and the bubble surface becomes free (Acharya et al. 1977).

Surface effects are known to influence the transition radius, for a bubble to change over from rigid to mobile interface. The mechanism of the jump in the rise velocity is not yet clear.

The shapes of bubbles are important in determining bubble behavior. Large bubbles in highly viscous liquids change from spherical to ellipsoidal to a spherical cap shape. Correlations for bubble shapes and deformations have been published by Acharya et al. (1977) and Bhaga (1978).

In practice, swarms of bubbles, instead of single bubbles, are encountered. The effect of gas holdup on the motion of a swarm of bubbles in Newtonian liquids was examined by Gal-or et al. (1968). Bhavaraju et al. (1978) developed equations for rise velocities and mass transfer for swarms of bubbles, in terms of gas holdup and pseudoplastic parameters.

Their results indicate that the velocity of a bubble swarm is less than that of a single bubble, and the ratio of velocity of a bubble swarm to that of a single bubble decreases with increasing gas holdup. Further, as the liquid approaches Newtonian behavior, this ratio deviates from 1. This implies that for a constant gas holdup, the retarding influence of neighboring gas bubbles on the swarm rise velocity is more in the case of a Newtonian fluid than in the case of a pseudoplastic fluid.

Bubble swarm behavior and oxygen transfer rates in sparged columns for non-Newtonian fluids were investigated by Buchholz et al. (1978). The bubble size distributions were recorded photographically. CMC solutions of various concentrations were used, and their rheological parameters and apparent viscosities were reported, in terms of concentrations. The distribution of bubble diameters was found to be log-normal. The mean gas holdups at superficial gas velocities varying between 1 and 6 cm/s were reported. No effort was made to determine these parameters at various points in the column.

It is the purpose of this work to examine the effect of non-Newtonian media on bubble behavior in both upflow and downflow regions of an airlift.

Liquid circulation and gas holdup

One of the important characteristics that distinguish an airlift from conventional bubble columns is the presence of a directional circulation of the liquid. The driving force for the circulation results from a pressure

difference between the two regions of the fermentor. The hydrodynamics in the upflow region is similar to that reported for bubble columns.

A fraction of the bubbles that is entrained in the downflow flows downward with the circulating liquid. Large bubbles, formed due to coalescence, rise through the liquid, thus increasing drag. The difference in gas holdups in the two regions partly depends on the liquid velocity. The interdependence of velocity and holdup complicates modelling of an airlift column.

Liquid circulations are known to exist in gas-sparged vessels. DeNevers (1968) examined fluid circulations in systems with or without a baffle. He showed the mechanism of circulation to be quite similar to natural convection, but with much larger driving force. The liquid with lower density, which contains larger fraction of entrained bubbles, moves upward. The work to drive the circulation is supplied by buoyant work of the rising bubbles. It is dissipated as friction with the walls and with other bubbles or liquids. Nicklin (1962) calculated energy loss due to wall friction and slip. Chakravarty et al. (1974) developed a model for liquid velocity in a gas-lift column, by making an energy balance. They assumed a non-coalescing system. Hatch (1973) measured liquid circulation velocity in an airlift fermentor, using hot water tracers. He used a homogeneous flow model, proposed by Wallis (1969) to calculate pressure gradients in different regions of the fermentor.

A model for prediction of liquid circulation rate was developed by Bhavaraju et al. (1978). They assumed that an imaginary tube separated the upflow and downflow zones. Liquid velocity in the imaginary tube was calculated for laminar and turbulent conditions, using steady state macroscopic energy balances for the gas and liquid phases. Equations were also developed for non-Newtonian liquids.

Hsu and Dudukovic (1980) developed a model for predicting the liquid circulation velocity in a loop reactor. Their model is based on an overall momentum balance of a gas-lift loop. Experimental correlations for gas holdups and two-phase pressure drops were used. Within the ranges of viscosity, density and surface tension, these properties did not affect gas holdups. Recently, Merchuk and Stein (1980) have proposed a model for liquid circulation, from their studies of local variation of gas holdups in the upflow region of an airlift system. No such studies on the downflow section have been reported.

Steady state momentum balance for the loop is written as (Hsu and Dudukovic, 1980)

$$\begin{aligned}
 & g[\phi_d \rho_G + (1 - \phi_d) \rho_L] H_d \\
 & - g[\phi_u \rho_G + (1 - \phi_u) \rho_L] H_u \\
 & = \left[\frac{4f_{TP}}{D} H \left(\frac{\rho_M V_M^2}{2} \right) \right]_u + \rho_L V_L \left(\frac{V_L}{1 - \phi_u} - V_L \right) \\
 & + \left[\frac{4f_{TP}}{D} H \left(\frac{\rho_M V_M^2}{2} \right) \right]_d + \Delta P_t
 \end{aligned} \tag{24}$$

In equation (24) left hand side represents the hydrostatic heads of the downflow and upflow sections. First and third terms on right hand side represent frictional losses for the two regions. Second term represents the accelerational loss across the sparger. ΔP_t includes exit and entry losses, and frictional pressure drop due to turnaround.

The mixture density and superficial velocity of the mixture are calculated from

$$\rho_M = \rho_L (1 - \phi) + \rho_G \phi \quad (25)$$

and

$$V_M = V_s + V_l \quad (26)$$

Gas holdup is defined as

$$\phi = \frac{\text{Volume of gas}}{\text{Total volume of mixture in the fermentor}}$$

The accelerational pressure drop is known to be less than 1% of the total pressure drop (Hatch, 1973), and can be neglected. Hence, for steady state, the pressure balance for flow in the airlift can be written as

$$\Delta P_{H_d} - \Delta P_{H_u} = \Delta P_{F_d} + \Delta P_{F_u} + \Delta P_t \quad (27)$$

where ΔP_t is the pressure drop due to the turnaround .

The left hand side, which is the difference of hydrostatic heads, is equal to $(\phi_u - \phi_d)g\rho_L H/g_c$.

The frictional contributions to pressure drops can also be calculated, assuming homogeneous flow (Wallis, 1969), and using standard equations. Hence, the frictional pressure drops in the two regions may be written as

$$\Delta P_{F_d} = \frac{2C_{fd} \rho_{Md} H V_L^2}{g_c D} \quad (28)$$

$$\Delta P_{F_u} = \frac{2C_{fu} \rho_{Mu} H V_{Lu}^2}{g_c D} \quad (29)$$

where

$$V_{Lu} = V_L \left(\frac{1 - \phi_d}{1 - \phi_u} \right) \quad (\text{Hatch, 1973}) \quad (30)$$

For the turnaround pressure drop

$$\Delta P_t = \frac{2C_f \rho_L L_e V_L}{D_{h_B}} \quad (31)$$

where L_e is an equivalent length of straight tube, which is determined from a pressure balance (Hatch, 1973).

If gas holdups are low (< 20%), the friction factors C_{fd} , C_{fu} and C_f can be determined from standard correlations for laminar and turbulent conditions, in terms of the liquid Reynolds number

$$Re_L = \frac{D V_L \rho_L}{\mu_L} \quad (32)$$

$$C_f = \frac{16}{Re_L} \quad (33)$$

for laminar, and

$$C_f = \frac{0.0791}{Re_L^{0.25}} \quad (34)$$

for turbulent conditions. Hsu and Dudukovic (1980) used the Reynolds number of the mixture

$$Re_M = \frac{D V_M \rho_M}{\mu_L} \quad (35)$$

and pointed out that the two phase friction factors are in most cases higher than those predicted by Equations (33) and (34) based on Re_m . The transition from laminar to turbulent flow appeared to occur at a mixture Reynolds number of about 1000.

Pressburg 1961, Dukler et al. 1964, Nicklin et al. 1974, Brown et al. 1960, Oshinowo and Charles 1974), and several empirical correlations for two-phase gas holdups have been proposed. One of the widely used correlations was developed by Lockhart and Martinelli (1949), for flow in horizontal pipes. Hughmark and Pressburgh (1961) suggested that the two-phase frictional pressure drop is due to two mechanisms, namely friction between the fluid and walls, and turbulent interaction between the two phases. Both of them cause energy losses. Energy loss due to turbulence is expected to be a function of slip velocity. Bhaga and Weber (1972) presented a theoretical analysis of 2- and 3-phase vertical flows. Recently, Hsu and Dudukovic (1980) used their liquid circulation data to develop an empirical correlation for gas holdup in air-water system.

On reviewing the literature, there appears to be a need to investigate these parameters in terms of physical properties of the dispersions, in airlift columns, and their variation in different parts of the columns.

The objectives of this work were

- 1) To examine the effects of different liquid properties, such as non-Newtonian viscosity and presence of ions and surfactants in the medium, on bubble behavior;
- 2) to determine gas holdups in the two regions and liquid circulation velocities in the downflow section with an aim to correlate them;
- 3) to study the variation of gas holdup with vertical distance in the fermentor. This is particularly important when the medium is highly viscous, and
- 4) to devise ways to improve the interfacial area and bubble size distribution with an ultimate aim to improve the overall oxygen transfer.

EXPERIMENTAL PROCEDURE

The 15.24 cm diameter column used in this work was the single stage split cylinder airlift tower used earlier by Orazem and Erickson (1979). Additional pressure taps were installed such that the pressure could be measured using open end manometer containing water at several locations along the downflow section. Pressure taps were located at $z = 0, 24.5, 51.5, 80.5$ and 105.5 cm where z is the distance below the top of the 107.1 cm baffle. Gas holdups in various parts of the downflow section were determined from the differences between two manometer readings, using the equation

$$\phi = \frac{\Delta H}{\Delta z} \quad (36)$$

where ΔH is the difference between two manometer readings, and Δz is the distance between the two pressure taps which are connected to the manometers.

Experiments were conducted, for several ungassed liquid heights ranging from 10 cm below to 20 cm above the top of the baffle. The liquids used included distilled water, distilled water with various concentrations of electrolyte, distilled water with various concentrations of CMC (Carboxy methyl cellulose), distilled water with various concentrations of sodium lauryl sulfate (surfactant), and solutions with surfactant and CMC.

Gas holdups were determined in various parts of the downflow section, the total upflow section, and the total downflow section using manometers. The total gas holdup in the fermentor was determined from the rise in liquid level after air was sparged into the fermentor. Liquid velocities were estimated by measuring the time of travel of 0.6 cm diameter particiales of coated paper which were neutrally buoyant. The time of travel was taken to be the average of several readings. In the case of distilled water and surfactant solutions, the velocity fluctuations were large due to

turbulence in the liquid. In highly viscous solutions, the flow appeared to be laminar near the walls, and the liquid velocity was high at the center, and nearly zero close to the walls. The maximum velocity was recorded in order to estimate the shear rate, which was used in determining the apparent viscosity for each concentration of CMC (Buchholz et al. 1978).

In determining gas holdups, the levels of manometers were read at 30 second intervals, over a period of 2 minutes, in order to average the fluctuations in levels. Similarly, an average liquid height in the column was recorded in the calculation of total gas holdup. Precautions were taken to remove air bubble from all the manometer lines.

DISCUSSION OF RESULTS

The results of the experiments show the influence of liquid properties examined here on gas-liquid dispersion. The quantities of interest are flow behavior, bubble sizes, and local and total gas holdups in the fermentor. The results are shown in Figures 1-22.

Flow behavior

For the range of gas and liquid velocities examined in this work, bubble flow regime was present, except for CMC solutions, when the spherical cap bubble rising through the liquid, caused slug-flow like conditions at the top. In water, and NaCl and Na LSO_4 solutions, the bubble size was small, varying from less than 1 mm to about 0.5 cm in the downflow section.

The liquid flow pattern appears to be turbulent, from movement of the solid tracer, for low viscosity liquids, such as distilled water and solutions of electrolytes and surfactants. The presence of these substances does not significantly alter the viscosity of the solution (Hatschek, 1928). The movement of the solid tracer for measuring liquid flowrate indicates random behavior of the circulating liquid. The coalescence of bubbles also contributes to the turbulence. Visual observation indicates a marked decrease in coalescence rates when either electrolyte or surfactant is added to distilled water. Large bubbles, which are formed due to coalescence, reverse the direction of flow, and increase the drag on the circulating liquid in the downflow region. The bubbles in downflow are much smaller, and move slower. In CMC solutions, the flow pattern changes from turbulent at the very top to a greater fraction of laminar flow throughout the downflow region. The larger bubbles at the center have diameter of about 4-5 cm and a spherical cap shape. Smaller bubbles either remain almost stationary near the

walls or are captured in the wake of a large bubble which is rising. As a result, there is an accumulation of larger bubbles near the top, and the conditions are more turbulent. In the upflow region the flow pattern appears to be turbulent, and the bubbles are larger. For highly viscous liquids (CMC solutions), for small bubbles, conditions were close to creeping flow, as indicated by the approximate calculation of bubble Reynolds number, for different bubble diameters (Table I).

The rise velocities required for these calculations were calculated from the appropriate expression based on the bubble Reynolds number [See equations (17) and (19)]. The results of these calculations have been reported (Deshpande, 1979) (see Appendix). They indicate that the effect of viscosity on bubble rise velocities becomes weaker beyond Stokes region, and there is a jump in the rise velocity as the bubble interface becomes free. The drag coefficients are calculated from equations (18) and (20) for Newtonian liquids, and from the following equations for power-law liquids (Acharya et al. 1977);

$$C_D = \frac{16 F_1(n)}{Re_n} \quad (37)$$

$$Re_n = \frac{d_B^n V_r^{2-n} \rho_L}{K} \quad (38)$$

In these calculations, the power-law index n was assumed to be nearly equal to 1. $F_1(n)$ has been defined as a function of n , by Hirose and Moo-Young (1969) and also by Bhavaraju et al. (1978), and is equal to 1 when n is nearly equal to 1.

The drag coefficients in Table I indicate that drag decreases with increasing bubble volume, but increases rapidly with increase in viscosity.

The drag on a bubble in a power-law fluid is higher than that for a Newtonian fluid with same apparent viscosity.

The energy dissipation in turbulent flow was calculated, using Taitel's equation (1980). For a liquid viscosity of 1 poise, it was found to be roughly 2.5 times that for a viscosity of 0.01 poise. Thus, where turbulent conditions are generated by the rising gas phase, the energy is rapidly dissipated for CMC solutions. Visual observations indicate that laminar conditions exist close to the walls. Because of the high viscosity, near the walls small bubbles and the liquid moved very slowly compared to the velocities farther from the walls.

Liquid circulation

The experimentally measured liquid circulation velocities are plotted as functions of liquid height, for distilled water and three different solutions in Fig. 18. The liquid height has considerable effect on V_L for CMC and surfactant solutions. These data reasonably agree with the plots of difference in gas holdups in the upflow and downflow regions vs. h_L (Figs. 19-22). At a very low liquid height ($h_L \approx -10$ cm), the difference in holdups is the lowest for all the liquids. This explains a low liquid circulation rate for corresponding height in Fig. 18. The high liquid circulation rates also correspond to larger differences in gas holdups. The latter, however, are rather poor in accuracy, due to the limits of the equipment, especially for 1% CMC solution. The liquid circulation rates as measured, are average values. For example, the variance and standard deviation for 0.2 M NaCl solution in the measurement of liquid velocities were 3.5 (cm/sec)^2 and 1.87 (cm/sec) respectively.

The liquid Reynolds number is higher for a power-law fluid with $n < 1$ than for a Newtonian one with the same apparent viscosity. (See Table IV) It is given by (Skelland, 1967)

$$Re_{LN} = \frac{D V_L^{2-n} \rho_L}{8^{n-1} K \left(\frac{3n+1}{4n} \right)^n} \quad (39)$$

Shear rates were calculated using experimentally determined liquid velocities, and apparent viscosities were computed for CMC solutions using the power-law parameters for CMC solutions listed by Buchholz et al. (1978).

The liquid velocity appears to be nearly independent of concentrations of solutions and only weakly dependent on liquid properties. The liquid in the upflow region experiences more velocity fluctuations based on the path as indicated by the motion of the solid tracer. The liquid Reynolds number for CMC solutions are always in the laminar region. For water and Na_2SO_4 solution, the liquid circulation is clearly turbulent (See Table IV).

Gas holdup

The effect of ungassed liquid height above the baffle was investigated for each of the different aqueous solutions. The effect of ungassed liquid height on holdup in the downflow section is presented graphically in Figures 8, 10, 11, and 16. Figures 19-22 show the variation of holdup in both the upflow and downflow sections with h_L . The effect of h_L on the variation of holdup in different parts of the downflow section is shown in Figures 2-6 and 14. Figures 8, 10, 11 and 16 show that the largest holdup in the downflow section is found for the smallest values of ungassed liquid height above the baffle, h_L . These figures also show that the smallest holdup in the downflow section for each solution is usually for the largest value of h_L . Figures 2-6 and 14 show that the greatest variations of holdup with distance below the baffle, z , are associated with the smallest values of h_L . The highest holdup in the

top of the downflow section and the lowest holdup in the bottom of the downflow section are usually for the smallest value of h_L . Examination of Figure 18 shows that the slowest liquid velocity is also associated with the smallest value of h_L . Figures 19-22 show that the difference in the gas holdup between the upflow and downflow sections, which is the driving force for liquid circulation is frequently smallest for the smallest value of h_L . Thus, coalescence and escape of gas bubbles in the downflow section appears to be largest for small values of h_L because of the reduced rate of liquid circulation. The reduced rate of liquid circulation is related to the reduced driving force for flow over the baffle as h_L is decreased.

These results suggest that an optimum liquid level exists for a given baffle height. It appears that the ungassed liquid height should be less than or equal to the baffle height. This should give relatively high gas holdup in all parts of the downflow section. The results indicate that the ungassed liquid height should be lower when the fluid contains surface active substances which increase the gas holdup. For example, Figure 14 shows that the condition of $h_L = -10$ cm has resulted in the highest holdups in all parts of the downflow section for an aqueous solution of 10 ppm NaLSO_4 .

The effect of h_L on differences in gas holdups in the two regions is displayed in Figs. 19-22, for various solutions. In all cases, the difference is nearly constant, for ungassed liquid heights above the baffle greater than 0 cm and slight lower for very low values of h_L . The quantity of gas in the upflow section that is entrained in the downflow section shows a stronger dependence on the liquid properties than on the liquid height. The decrease in the holdups from the upflow to downflow sections varies from 10.3% (lowest) to 17.9% (highest) in the CMC + surfactant case. In CMC solution these values are 19.1% and 52.6% (see Table II). Table II shows that larger quantities of air escape into the atmosphere with CMC solution and distilled water than with

solutions containing a surfactant or electrolyte. These substances help maintain smaller bubble size, and therefore increase the number of bubbles that can be carried downward.

The variation in the gas holdup in the downflow section with distance from the top of the baffle is depicted in Figures 2-6, and 12, 14, 15 and 17. Figures 3, 4 and 5 for water (distilled and tap) and sodium chloride solution show a gradual decrease in the holdups as the liquid flow downward. At a very low liquid height, h_L , the gas holdup is very high in the top of the downflow section, but it drops rapidly with distance, z . Figure 12 shows variation of ϕ with distance, z , at $h_L=0$, for four different liquids. The holdup decreases rapidly for 1% CMC, but is almost constant for solutions containing surfactant. For distilled water the decrease is slower than in 1% CMC solution. The coalescence rates are high in both distilled water and 1% CMC solution. As a result, larger bubbles are formed, which flow upward near the top. CMC solutions of various concentrations display similar trends (Fig. 15) in the variation of gas holdup with distance. Comparison of Figs. 15 and 17 shows that for 1% CMC solution, at $h_L = 0$ cm the holdup in the lower part is 25% of the average holdup in the downflow, and only 10% of the holdup in the top part. For 10 ppm Na_2SO_4 solution these values are 95.7% and 84.6%, respectively. For a 1% CMC solution containing 10 ppm Na_2SO_4 , these values are 95.7% and 68.8%, respectively. This suggests that a small amount of surfactant greatly reduces coalescence rates in CMC solution, and helps achieve a smaller average bubble size. A comparison of holdups in the top and bottom of the downflow section is shown in Fig. 6. It appears that the difference between the two is higher at low liquid heights. The difference is lowest for surfactant solution.

Table III lists holdups in the upflow and downflow regions, and the total holdups for different solutions. The measured value of the total holdup (from the gassed and ungassed liquid heights) is larger than the average holdup in the upflow and downflow sections. The difference is partly due to holdup in the head region, oscillation of the liquid level, and foaming which makes it difficult to identify the gassed liquid level. This is more pronounced in surfactant and electrolyte solutions. In CMC solutions, large bubbles which are carried to the top often burst, and make it difficult to read the liquid level. At lower liquid heights the agreement between the average and measured values of the total gas holdup is somewhat better. This is because the holdup in the head region was neglected in finding the average value. In an airlift fermentor, the highest holdup is usually in the head region.

Effect of surfactant concentration

Figures 8 and 9 show the effect of sodium lauryl sulfate concentration on the gas holdup in the downflow section of the airlift fermentor. The gas holdup increases as surfactant concentration increases. The addition of surfactant reduces coalescence and greatly increases the holdup in the lower part of the downflow region as shown in Fig. 9. While surfactants increase the gas holdup, they also increase foaming problems. For 25 ppm NaLSO_4 , superficial gas velocities of 400 cm/min and an ungassed liquid height of 10 cm above the baffle, the total volume of the equipment was used because of the significantly higher gas holdup and the foam at the top of the head region.

The presence of surfactant reduces internal circulation, thereby increasing drag. The drag coefficient for bubbles in contaminated water is about twice that obtained from the standard drag curve (Clift et al. 1978).

At high viscosity the viscous resistance of the fluid limits internal motion. The liquid motion carries the surfactant particles to the rear of the bubble. With an increasing surfactant concentration, there is an accumulation of the particles on one side of the bubble. Although this usually results in a decreased oxygen transfer coefficient, the increase in the interfacial area and a more uniform distribution of bubbles throughout the fermentor may outweigh that factor.

Effect of electrolyte concentration

Figure 7 shows the variation of gas holdup in the downflow section with concentration of sodium chloride solution. Addition of a small amount of NaCl to distilled water increases the gas holdup by about 33%. A further increase in the concentration causes a relatively smaller increase in the holdup. The presence of a maximum in the curve, at 0.2 M NaCl agrees with the results of Lessard and Zieminski (1971), which predicted a sudden decrease in the coalescence rate at nearly the same concentration of NaCl. In fermentation, electrolytes are almost always present; however, their concentrations are frequently fixed according to the nutrient needs of the culture rather than mass transfer considerations. Both factors should be considered.

Effect of viscosity

Figure 13 shows the effect of concentration in terms of wt % CMC on the total holdup in the downflow section. As CMC concentration increases, the viscosity increases and the rise velocity of small bubbles decreases. Although highly viscous solutions reduce coalescence rates because film thinning times are increased, under laminar conditions the time available for coalescence is often greater than film thinning time (McBride, 1979). Holdup is relatively high in the upper part of the downflow section, but it decreases rapidly with vertical position due to coalescence (Fig. 15).

Visual observation of the highly viscous solutions in the airlift column reveals the existence of very large bubbles in the upflow region for solutions with and without surfactant. Because of the high viscosity, flow is laminar and very slow near the column walls and baffle in both the upflow and downflow regions. In the downflow region, bubbles rise near the vessel wall while similar bubbles at the center are carried downward due to high liquid velocity near the center. Relatively large bubbles which rise slowly near the wall in the viscous liquid frequently breakup if they are carried away from the wall into the center of the flow field.

A large number of small bubbles were present in the gas-liquid dispersion when surfactant and CMC were used together. The mechanism by which these small bubbles are generated in viscous fluids is not fully understood. Since the sparger diameter is 0.95 cm with only one opening, small bubbles are not expected to be generated in large numbers because of sparger design (Bhavaraju et al. 1978). Some of the small bubbles may be the result of local turbulence with bubble breakup due to dynamic pressure fluctuations. However, these small bubbles are generated by bubble breakup under conditions where the liquid Reynolds number indicates laminar flow conditions. Resonance and dynamic pressure fluctuations may lead to bubble breakup under these conditions. The wake of rising bubbles influences the path of trailing bubbles such that bubble deformation occurs and large bubbles coalesce with each other. Vortex shedding may lead to relative gas-liquid velocities sufficient to cause breakup. Bubbles may also be formed at the gas-liquid interface as large bubbles pass from the liquid continuous phase to the gas phase above the liquid.

The variation of the gas holdup in the downflow section with liquid height above baffle, is shown in Fig. 16. For 1% CMC, the holdup decreases from -10 to

10 cm, and remains constant for higher liquid heights. On addition of 10 ppm surfactant to the 1% CMC solution, the holdup is nearly constant for all liquid heights. The total holdup in the downflow section increases by 62.5% to 333% due to addition of the surfactant. For 2% CMC containing 10 ppm NaLSO_4 , the holdup in the downflow section is higher than that for 1% CMC, but the values are considerably smaller than those for 1% CMC with 10 ppm NaLSO_4 .

Figure 17 shows the variation of the same solutions along the vertical distance below the baffle, z , in the fermentor. For 1% CMC, the holdup decreases from the top to the bottom of the column. Very few bubbles are carried to the lower part because coalescence is rapid and large bubbles rise and leave at the top. These rising bubbles accelerate the coalescence process. When surfactant is present, more bubbles are carried downward.

In case of 1% CMC with 10 ppm NaLSO_4 the gas holdup falls by only 25% in going from top to bottom. Figure 17 shows that the 2% CMC solution containing NaLSO_4 shows considerably greater variation (about 70%) but the variation is not so large as that for 1% CMC solution without surfactant.

An increase in the superficial gas velocity increases the holdup in the downflow section. This effect is more pronounced in a CMC solution than in a NaLSO_4 solution (Fig. 10). The liquid height has little effect on the holdup in the lowest part of the downflow section, but that in the top stage is much higher at low liquid heights (Fig. 2). The liquid circulation is considerably slower in this case, and very few bubbles are carried beyond manometer 6 (Fig. 1).

The manometer readings were almost steady in CMC solutions. The gas holdups, however, were much smaller, especially in the lower half. This increases the possibility of error. In some cases, the differences in readings between consecutive monometers were very small with CMC solutions because of the low holdup in the lower part of the downflow section.

The results of pressure drop measurements are listed in Table V.

The total pressure drop is maximum (4790 dynes/cm²) for water, and decreases with increase in viscosity. The comparison of results for distilled water and NaLSO₄ solution, and also for 1% CMC and 1% CMC containing NaLSO₄ indicates that the presence of surfactant lowers the total pressure drop in the fermentor.

The equivalent length of straight pipe was calculated from the following equation (Hatch, 1973):

$$H\rho_L(\phi_d - \phi_a) \frac{g}{g_c} = \frac{2C_f \rho_L Q_L (Q_L + Q_{gd}) Le}{D_{hb} A_b^2 g_c} + \frac{2C_f \rho_L Q_L H}{g_c} \left[\frac{Q_L + Q_{gd}}{D A_d^2} + \frac{Q_L + Q_{gu}}{D A_u^2} \right] \quad (40)$$

Q_{gu} was assumed to be equal to the volumetric air flow rate. Q_{gd} was assumed to be equal to Q_{gu} times ϕ_d .

The equivalent lengths reported by Hatch (1500 cm) and Merchuk and Stein (9200 cm) are comparable with the value of 1500 cm reported for distilled water, in this work. For the CMC solutions, the equivalent length is small. This may be because the measured liquid velocities are larger than the average liquid velocity due to the nature of the velocity profile under laminar flow conditions.

The friction factors required for these calculations were estimated by means of equations (33) and (34). They are listed in Table IV. The Reynolds numbers for power-law liquids are higher, and the drag coefficients are lower.

The errors in liquid velocities arise due to the fluctuations in the path of the tracer. Several readings were taken to get the average liquid velocity. The mean and variance for 0.2 M NaCl are reported earlier. In CMC solutions, the average measured liquid velocity is close to the maximum value.

Errors in the measurement of gas holdup are due to fluctuations of manometers. The means and variances which are listed in Table VI indicate fewer fluctuations in the CMC data. However, the differences in holdups in highly viscous liquids are very small, especially in lower parts of the fermentor. Sometimes it was difficult to detect a density difference in the lower part when bubbles were concentrated near the top . This limits the accuracy of the results.

CONCLUSIONS AND RECOMMENDATIONS

For both distilled water and CMC solutions, the addition of a surfactant greatly increased gas holdup. The holdup in the downflow region decreased rapidly with distance below the top of the baffle in the absence of surfactant. The surfactant helps achieve a more uniform distribution of bubbles in the downflow region. The values of gas holdup for solutions of 1% CMC and 10 mg/liter surfactant were frequently higher than the values for similar solutions without CMC. It may, therefore, be desirable to have surfactants in the fermentation medium, provided that foaming can be controlled.

The presence of electrolytes yields similar results for gas holdup. Electrolytes are always present in fermentation broths as salts in various amounts.

The effect of electrolytes on bubble behavior needs to be investigated when actual fermentation broth is used. Further, oxygen transfer rates for solutions containing electrolytes and surfactants need to be investigated to find the optimum mass transfer conditions.

The experiments on CMC solutions indicate that although the liquid circulation rates are satisfactory in highly viscous media, the bubbles tend to coalesce and accumulate in the top portion of the fermentor. One way to resolve this problem is to add a surfactant to the liquid.

The addition of surfactant to CMC solutions resulted in a large number of small bubbles. The mechanisms by which these small bubbles are generated needs to be investigated further.

The results of this investigation have shown that the ungassed liquid height relative to the baffle height influences gas holdup in the airlift fermentor. The variation of holdup with distance below the baffle in the downflow section is largest for the lowest ungassed liquid height. The liquid circulation

velocity is also smallest for the lowest ungassed liquid height. The hold-up in the downflow region is largest for the smallest ungassed liquid height and smallest for the largest ungassed liquid height. The optimum ungassed liquid height depends on fluid properties. Ungassed liquid heights less than or equal to the baffle height are recommended.

The accuracy of the results is largely determined by the sensitivity of the manometer levels to changes in gas holdups. This is more important in highly viscous media since the fraction of gas in the downflow section is very small. In some cases the holdup is too small to accurately determine changes in the holdup with position by the existing equipment. Better ways to measure the gas holdup and the liquid circulation rates are desirable to improve existing designs.

**THIS BOOK
CONTAINS
NUMEROUS PAGES
WITH DIAGRAMS
THAT ARE CROOKED
COMPARED TO THE
REST OF THE
INFORMATION ON
THE PAGE.**

**THIS IS AS
RECEIVED FROM
CUSTOMER.**

Table I. Drag on bubbles with different diameters
for $\mu_L = 0.1, 1.6$ and 4.0 poises.

μ_L (poise)	D_B (cm)	Re_B	C_D
0.01	0.2	528	0.0303
0.01	0.5	1101	0.0145
0.01	1.0	2471	0.0065
1.6	0.2	0.17	141.2
1.6	0.5	3.99	4.01
1.6	1.0	15.44	1.04
4.0	0.2	0.027	888.9
4.0	0.5	0.426	56.3
4.0	1.0	5.11	3.1

Table II. Percent decrease in gas holdup, ϕ , as the liquid flows into the downflow section.

Solution	$\phi_u - \phi_d / \phi_u$	
	Lowest	Highest
1% CMC + NaLSO ₄	10.3%	17.9%
1% CMC	19.1%	52.6%
10 ppm NaLSO ₄	-5.4%	25%
Distilled water	24%	57.5%

Table III. Gas holdups in airlift fermentor for different solutions at a superficial gas velocity of 400 cm/min.

Solution	h_L (cm)	ϕ_u	ϕ_d	ϕ_T	$\frac{\phi_u + \phi_d}{2}$
1% CMC	20	0.0550	0.0288	0.07353	0.0419
	-10	0.1150	0.0899	0.1351	0.1025
1.5% CMC	20	0.024	0.0120	0.0667	0.018
	-10	0.1162	0.0958	0.1312	0.106
Distilled water	20	0.0959	0.0599	0.0935	0.0779
	-10	0.1126	0.0899	0.2066	0.1013
10 ppm NaLSO ₄	10	0.1220	0.0947	0.1533	0.1084
	-10	0.1330	0.1420	0.1587	0.1375
0.6 M NaCl solution	20	0.1282	.0719	0.1127	0.1000
	-10	0.1198	0.0947	0.1351	0.1072

Table V. Experimentally determined pressure drops and equivalent lengths.

Solution	η_{app} (poise)	V_L (cm/s)	Re_M	ΔP_{Total} $\frac{\text{dynes}}{\text{cm}^2}$	Equivalent length (cm)
Distilled	0.01	35.4	7.15×10^5	4790	1500
10 ppm NaLSO_4	0.01	44.4	7.02×10^5	3670	838
1% CMC	0.82	41.8	1130	3580	43.8
1% CMC + NaLSO_4	0.82	41.8	985	2650	38.8
1.3% CMC	1.18	45.9	651	3300	9.7
1.5% CMC	1.66	41.8	514	3200	3.84

Table VI. Variances and standard deviations of manometer readings
($h_L = 20$ cm, $V_s = 400$ cm³/min).

Liquid	Manometer							
		0	2	3	4	6	7	8
10 ppm Na ₂ SO ₄	Mean	38.9875	34	34.5	35.35	36.5	37.3625	38.3
	σ^2 (cm ²)	0.0373	0.00667	0	0.0033	0	0.0123	0.005
	σ (cm)	0.193	0.0816	0	.0577	0	0.1109	0.0707
1% CMC	Mean	35.7	33.4	33.8	33.85	34.1	34.1	35
	σ^2 (cm ²)	0.0167	0	0	0.00167	0	0	0
	σ (cm)	0.129	0	0	0.041	0	0	0
0.2 M NaCl	Mean	39.91	34.4	34.75	35.55	36.81	37.84	39.01
	σ^2 (cm ²)	0.078	0	0	0.0125	0.01175	0.083	0.0905
	σ (cm)	0.279	0	0	0.1118	0.1084	0.288	0.301

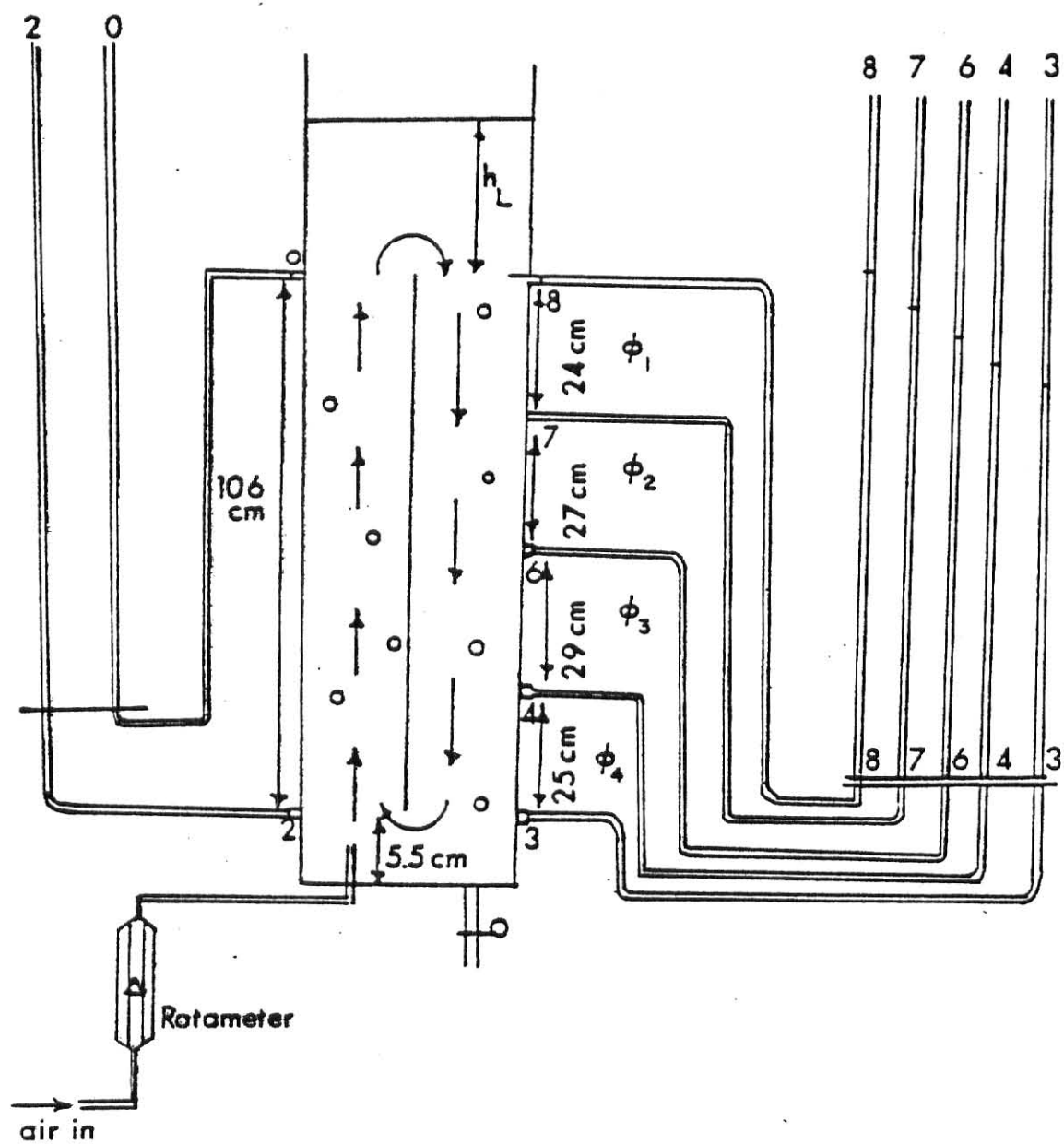


Fig. 1. Airlift column and manometers.

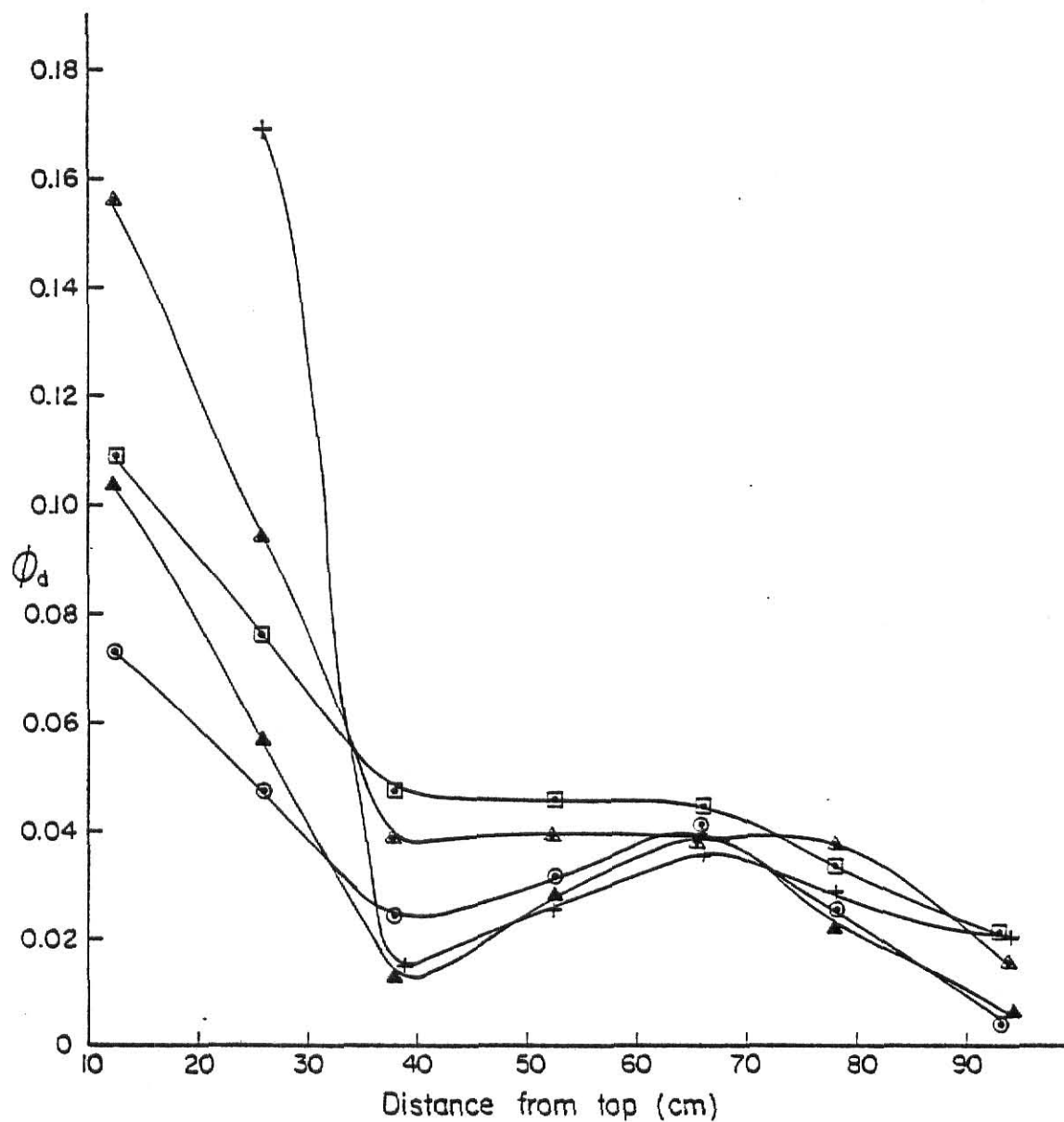


Fig. 2. Variation of gas holdup in the downflow section of an air-lift fermentor with distance from the top of the baffle, z , for 0.74% CMC solution at a superficial gas velocity of 400 cm/min and ungassed liquid height above baffle, h_L as parameter. $h_L = \circ$ -20 cm, $\blacktriangle$ -10 cm, $\square$ -0 cm, $\triangle$ --5 cm, $+$ --10 cm.

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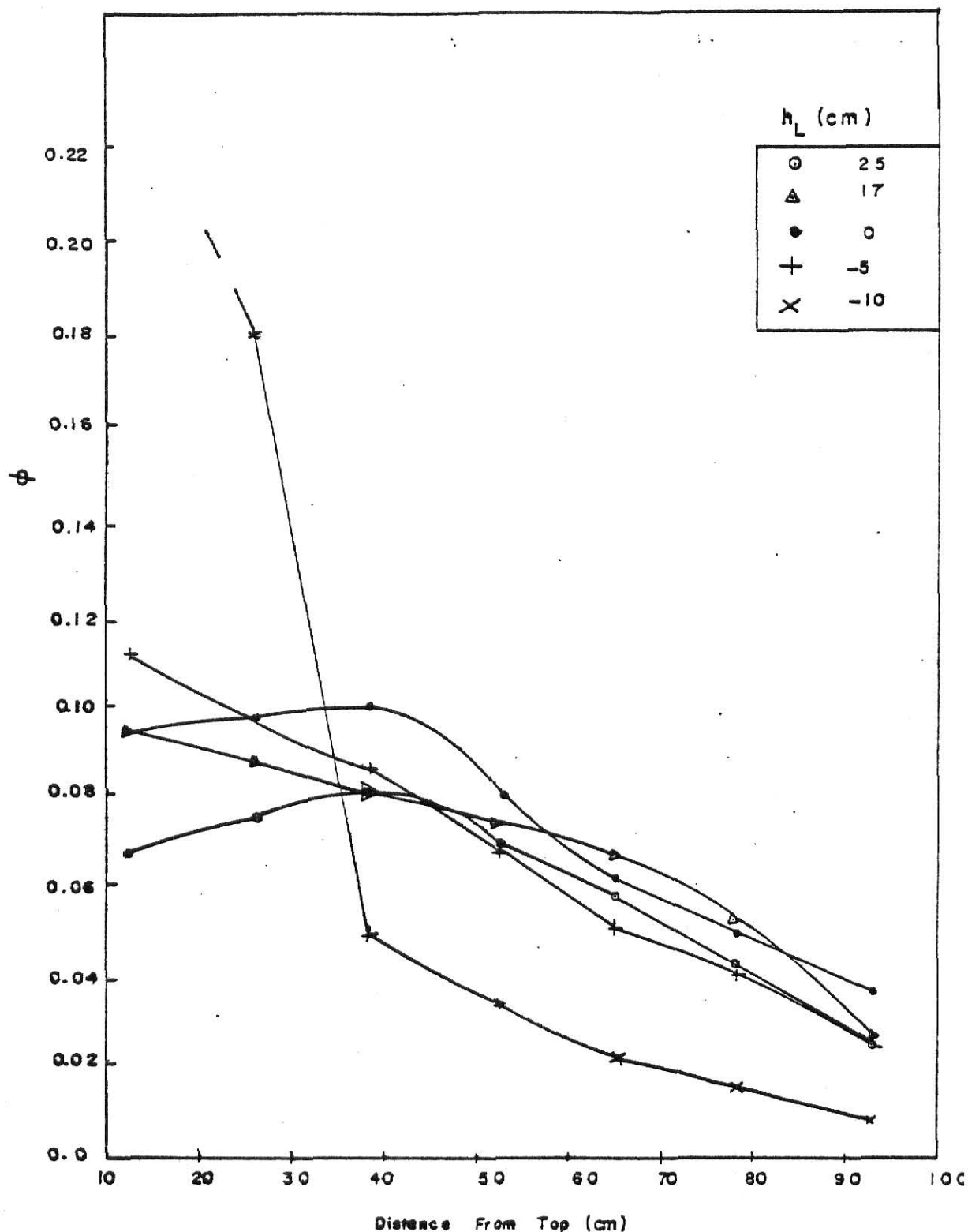


Fig. 3. Variation of gas holdup in the downflow section of an air-lift fermentor with distance from the top of the baffle, z , for distilled water at a superficial gas velocity of 400 cm/min. and ungasged liquid height above baffle h_L , as parameter.

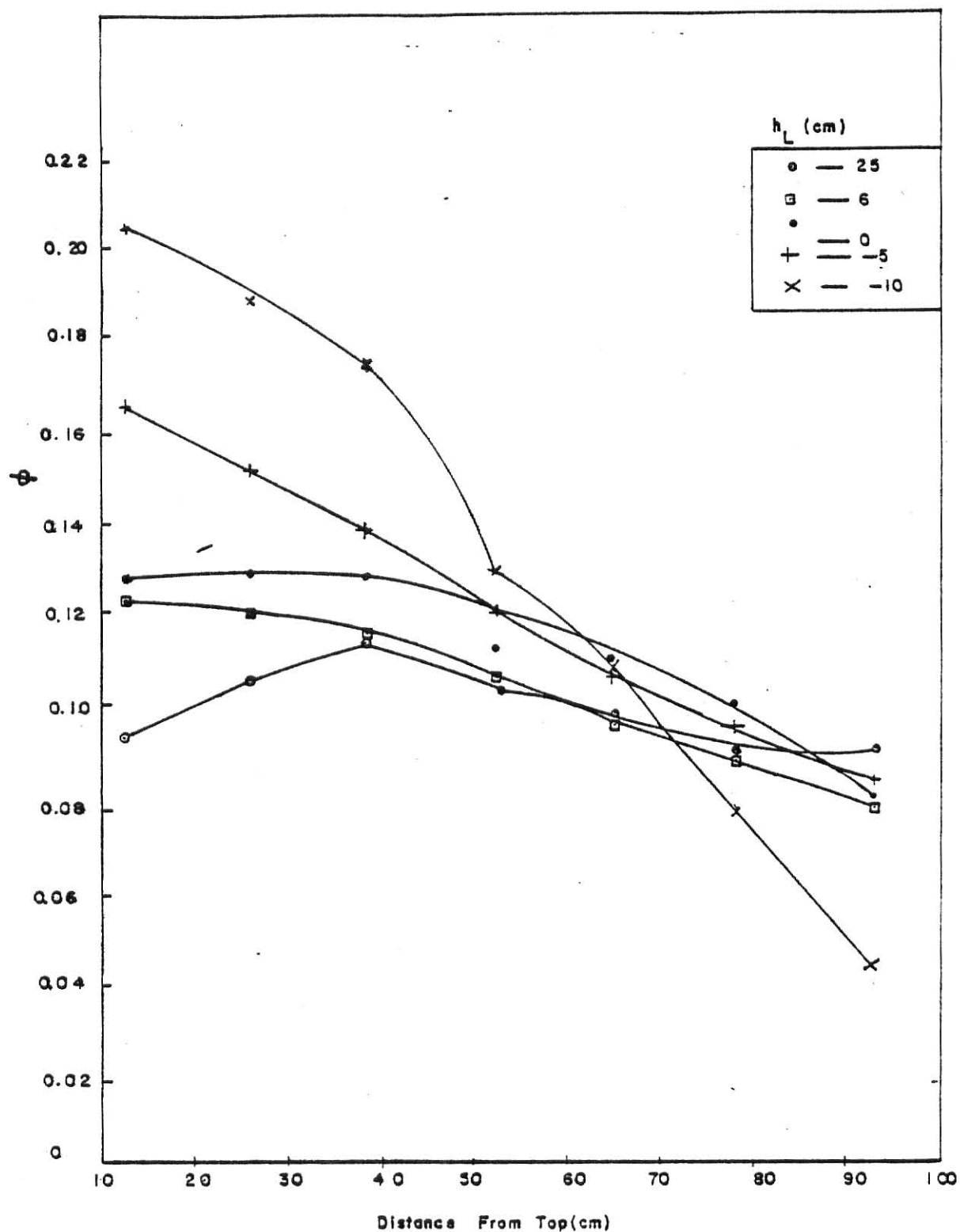


Fig. 4. Variation of gas holdup in the downflow section of an air-lift fermentor with distance from the top of the baffle, z , for tap water at a superficial gas velocity of 400 cm/min. and ungassed liquid height above baffle, h_L , as parameter.

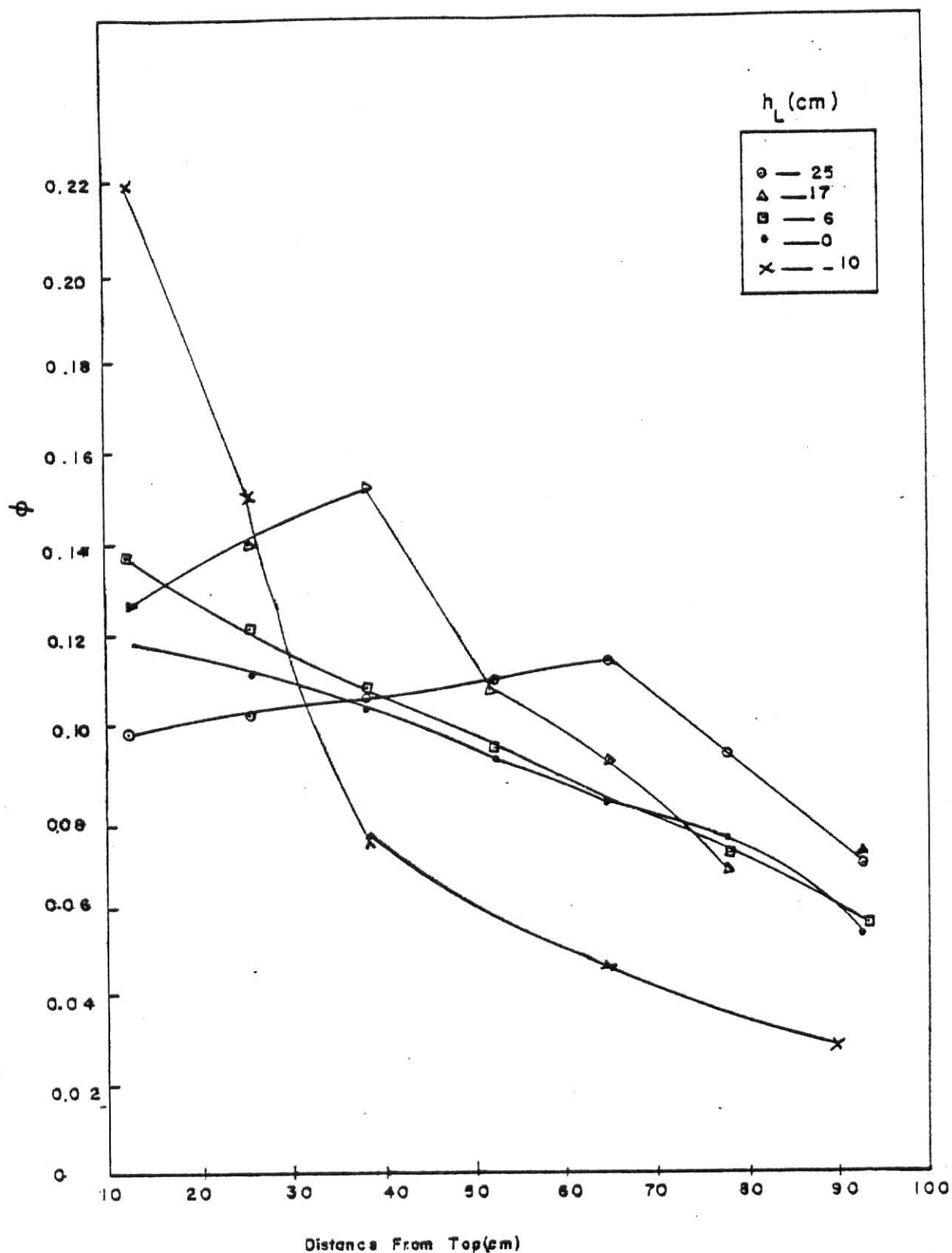


Fig. 5. Variation of gas holdup in the downflow section of an air-lift fermentor with distance from the top of the baffle, z , for 0.6 M NaCl solution at a superficial gas velocity of 400 cm/min. and ungassed liquid height above baffle, h_L , as parameter.

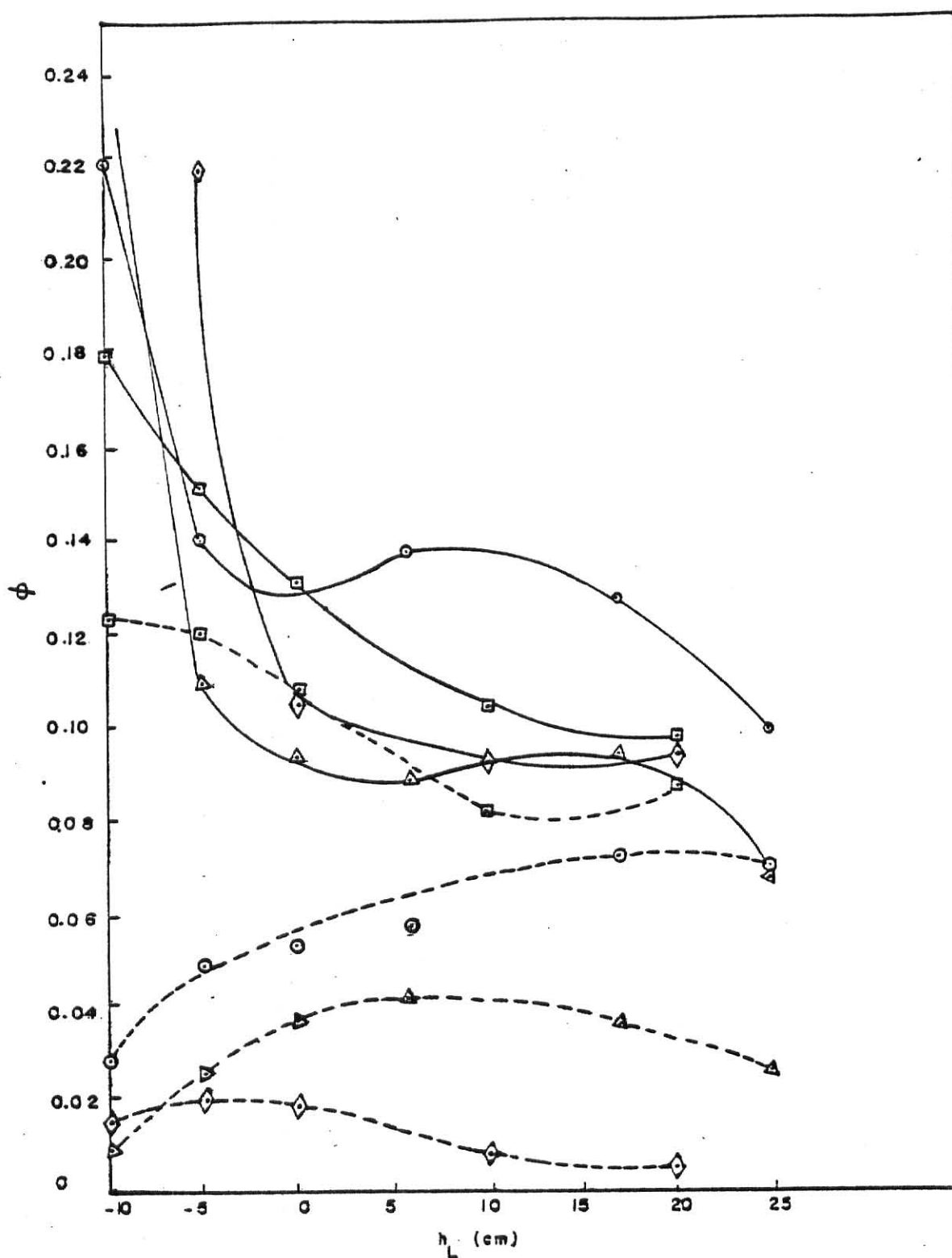


Fig. 6. Variation of gas holdup in the top and bottom of the downflow section of an airlift fermentor with ungassed liquid height above baffle, h_L , for Δ - distilled water, $\odot$ - 0.6M NaCl, $\square$ - 10 ppm Na_2SO_4 , and $\diamond$ - 1% CMC. Solid lines refer to top part and dashed lines to bottom part of the downflow section.

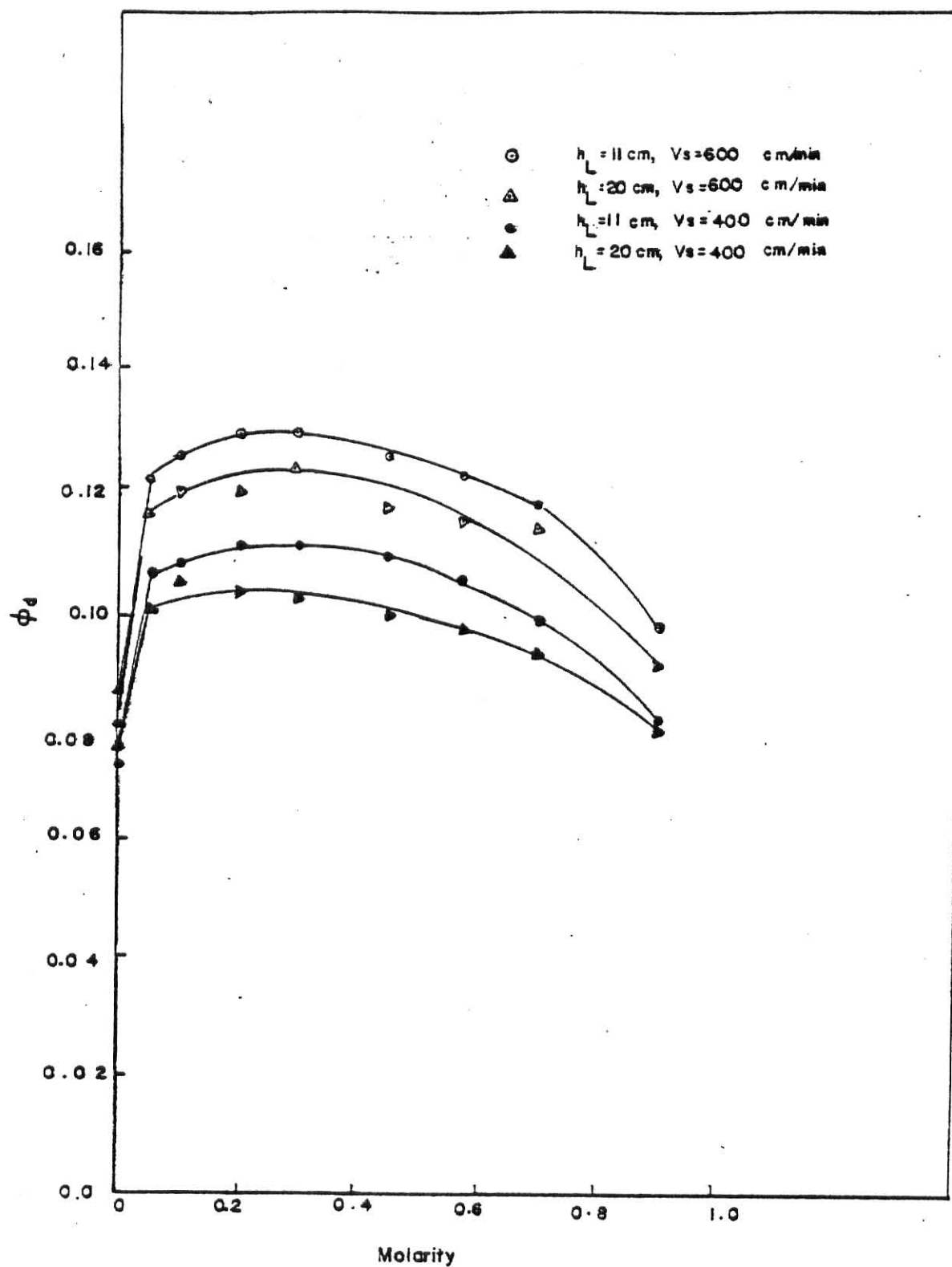


Fig. 7. Variation of gas holdup in the downflow section of an air-lift with concentration of NaCl for two superficial gas velocities, V_s , and two ungassed liquid heights above the baffle, h_L .

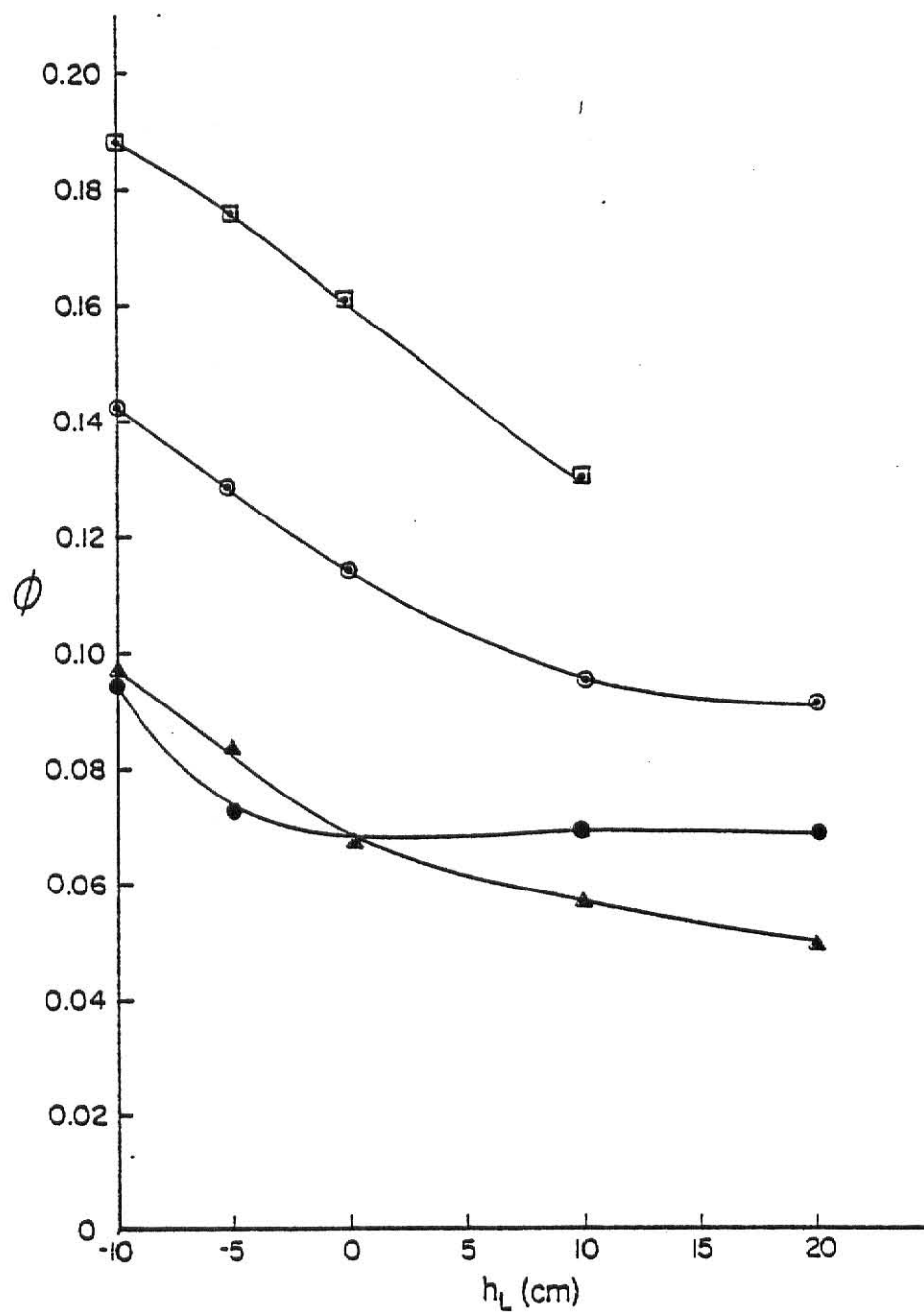


Fig. 8. Variation of gas holdup in the downflow section of an air-lift fermentor with ungassed liquid height above the baffle, h_L , at a superficial gas velocity of 400 cm/min. For ● - distilled water, Δ -2 ppm NaLSO₄, ⊙ -10 ppm NaLSO₄, and ◻ -25 ppm NaLSO₄.

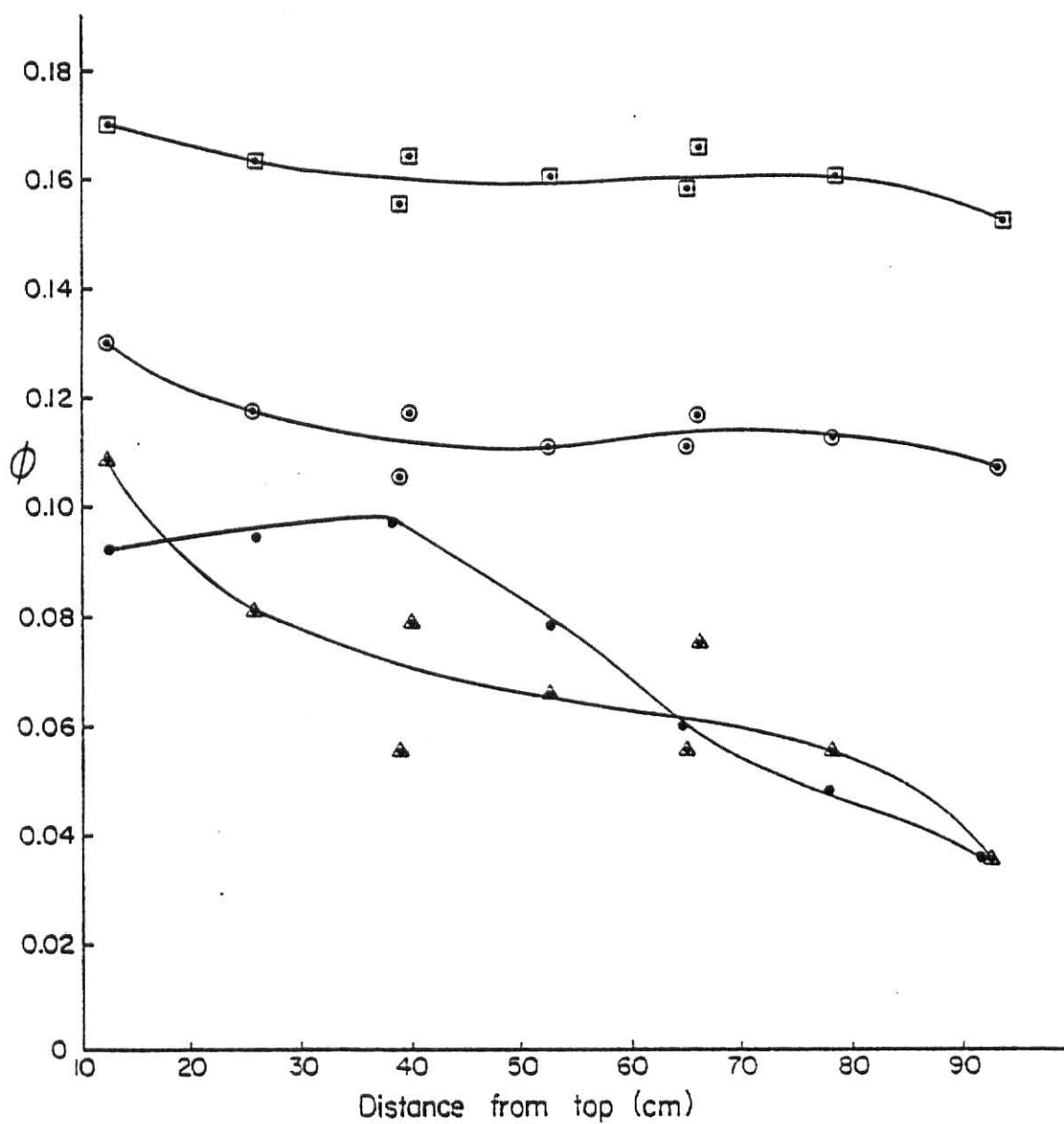


Fig. 9. Variation of gas holdup in the downflow section of an air-lift fermentor with distance below the top of the baffle, z , at a superficial gas velocity of 400 cm/min for $h_L = 0$.
 ● - distilled water, Δ -2 ppm NaLSO₄, ○ -10 ppm NaLSO₄, and ◻ -25 ppm NaLSO₄.

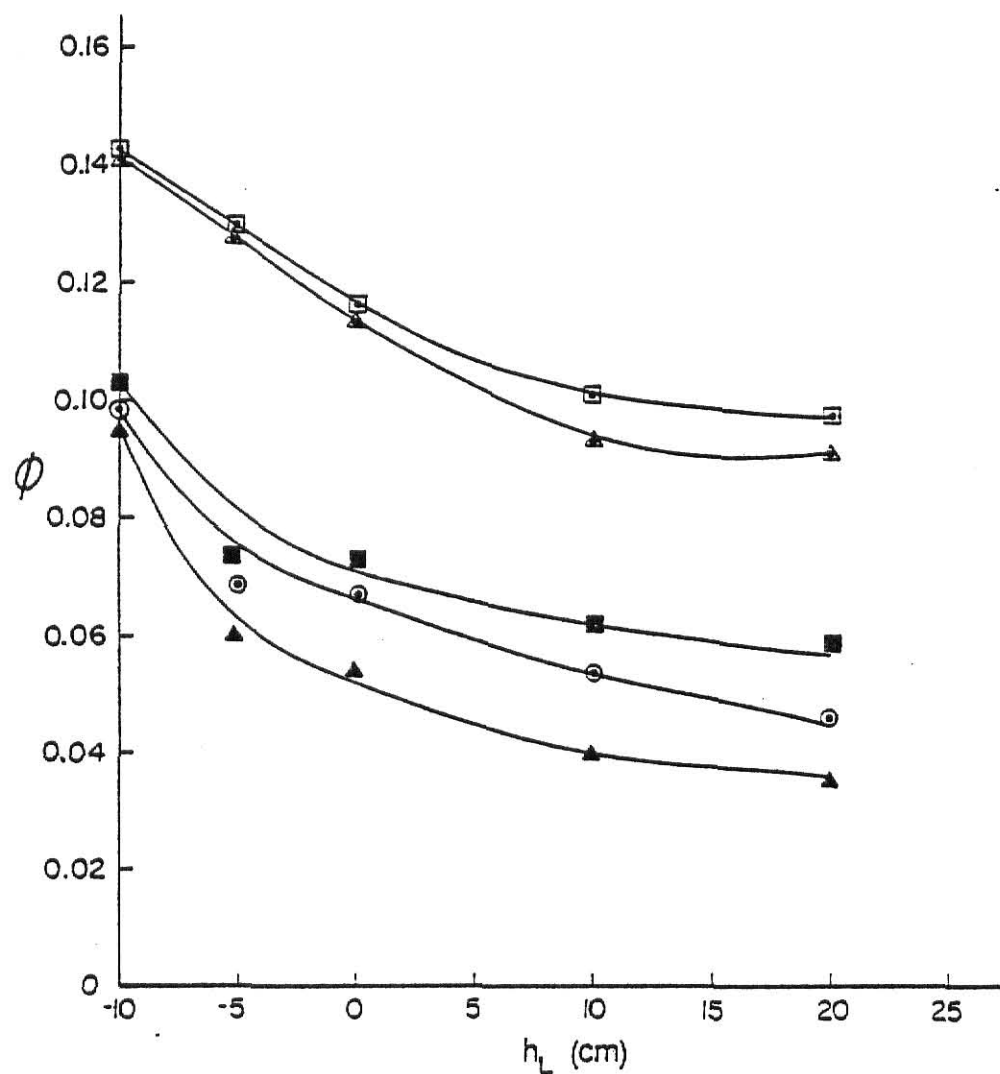


Fig. 10. Variation of gas holdup in the downflow section of an air-lift fermentor with ungassed liquid height above the baffle, h_L , for 10 ppm Na_2SO_4 with superficial gas velocities of $\square$ -400 cm/min and $\triangle$ -600 cm/min, and for 0.74% CMC with superficial gas velocities of $\blacktriangle$ 400 cm/min, $\bullet$ -600 cm/min and $\blacksquare$ -760 cm/min.

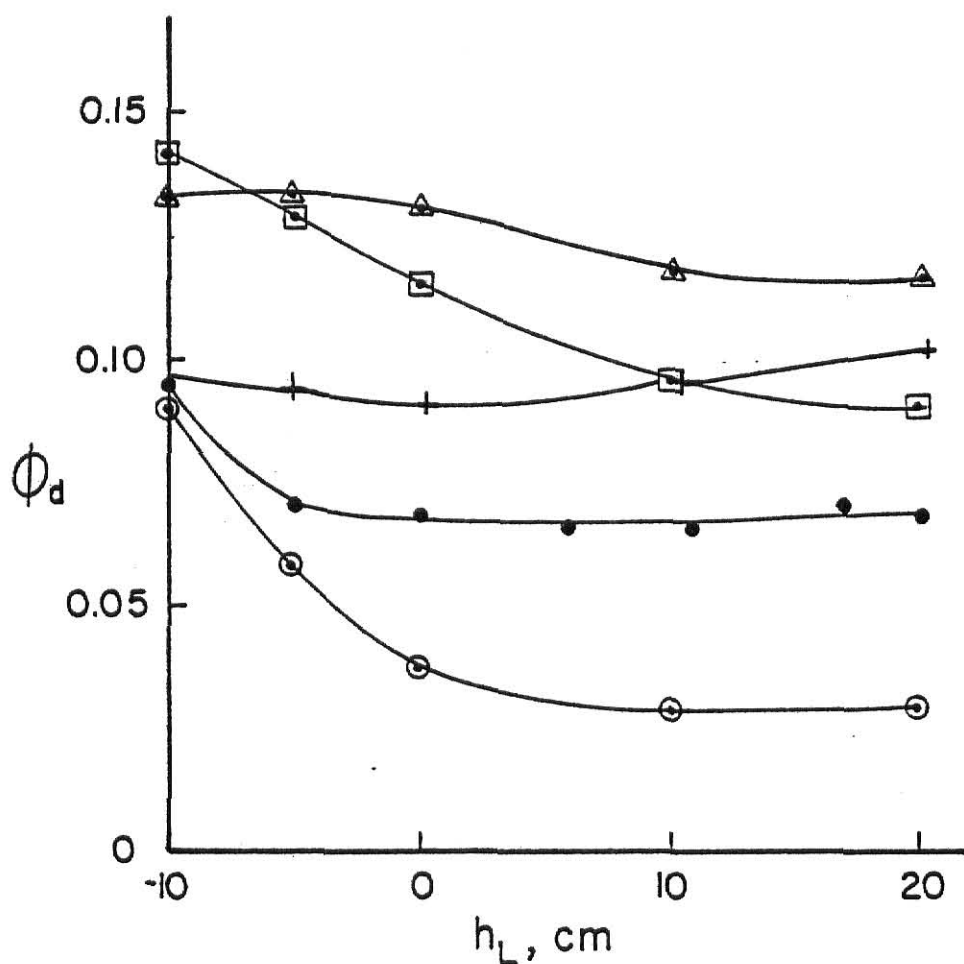


Fig. 11. Variation of gas holdup in downflow section of an airlift fermentor, with ungassed liquid height above baffle, for a superficial gas velocity of 400 cm/min. $\odot$ -1% CMC, $\square$ -10 ppm NaLSO₄, $\triangle$ -1% CMC + 10 ppm NaLSO₄, $\bullet$ -distilled water, and $+$ -0.6 M NaCl.

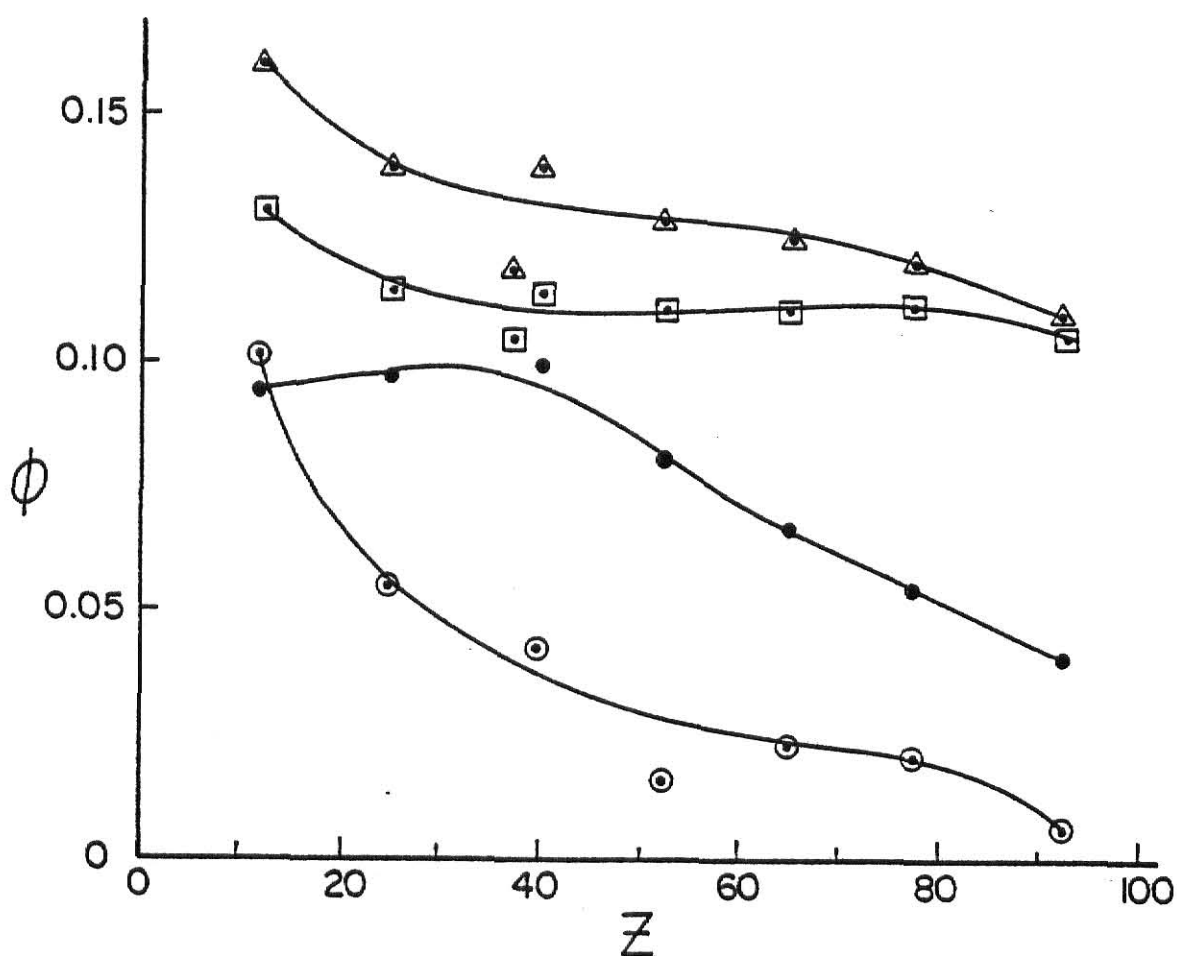


Fig. 12. Variation of gas holdup in downflow section of an airlift fermentor, with distance from top of the baffle, z , for an ungasged liquid height equal to baffle height and a superficial gas velocity of 400 cm/min. $\circ$ -1% CMC, $\triangle$ -1% CMC + 10 ppm NaLSO₄, $\square$ -10 ppm NaLSO₄, and $\bullet$ -distilled water.

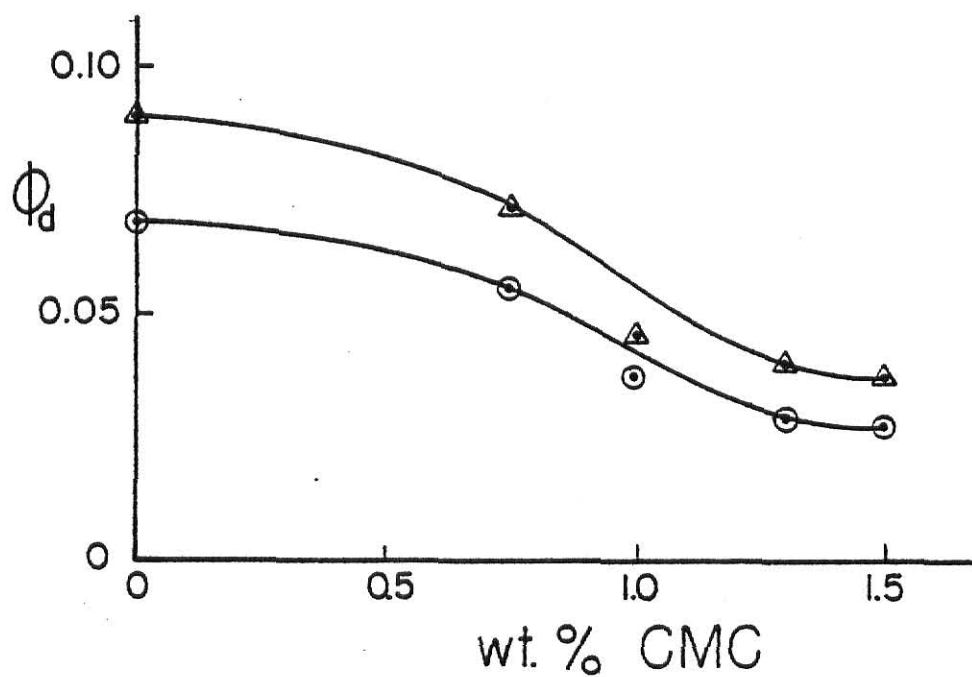


Fig. 13. Effect of concentration of CMC on gas holdup in downflow section of an airlift fermentor at superficial gas velocities of ○ = 400 cm/min and △ = 760 cm/min for $h_L = 0$.

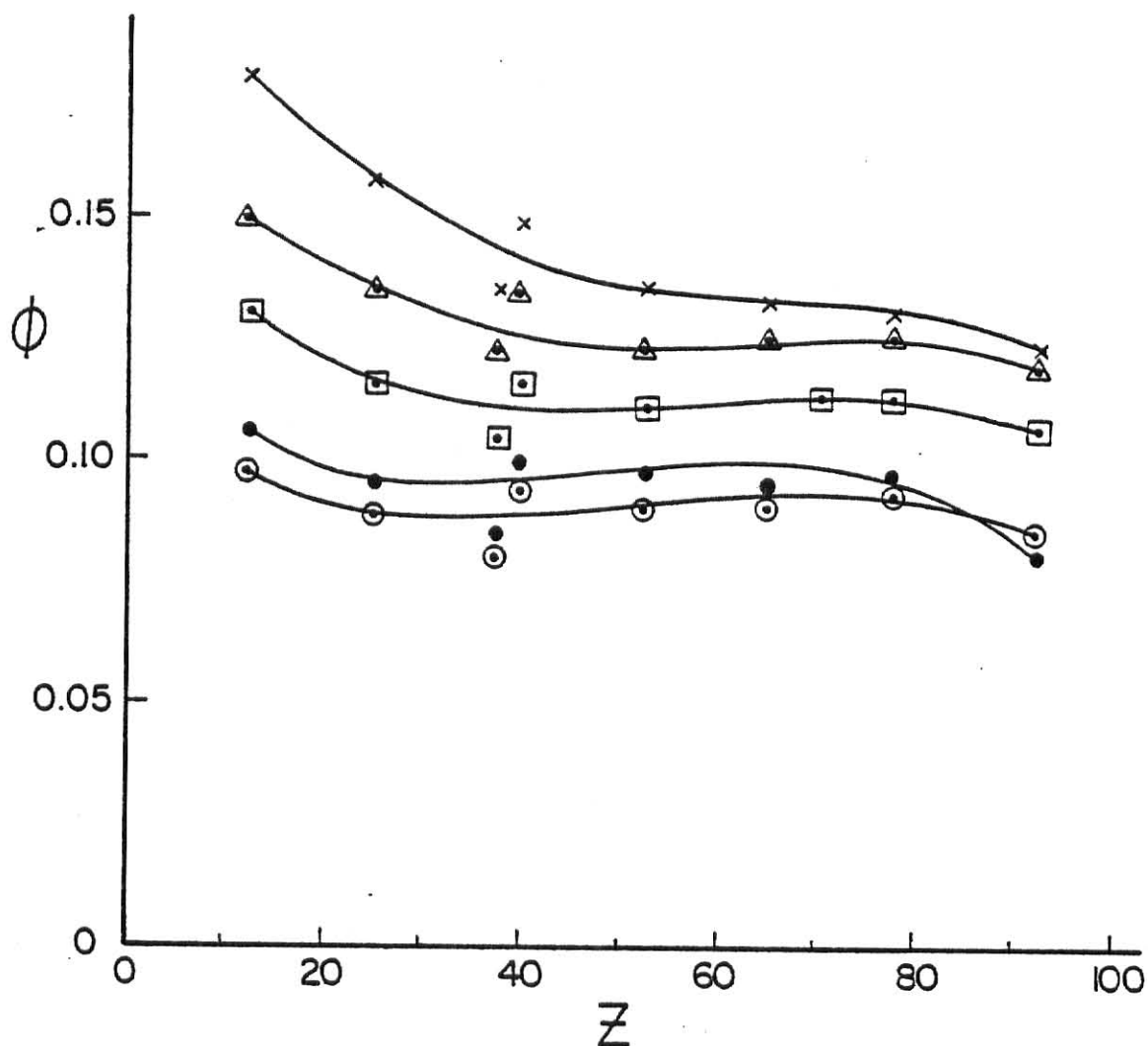


Fig. 14. Variation of gas holdup in downflow section of an airlift fermentor with distance from top of the baffle, z , for 10 ppm sodium lauryl sulfate at a superficial gas velocity of 400 cm/min and ungassed liquid height above baffle as parameter. Values of h_L are x - -10 cm, Δ - -5 cm, $\square$ - 0, $\bullet$ - 10 cm, and $\circ$ - 20 cm.

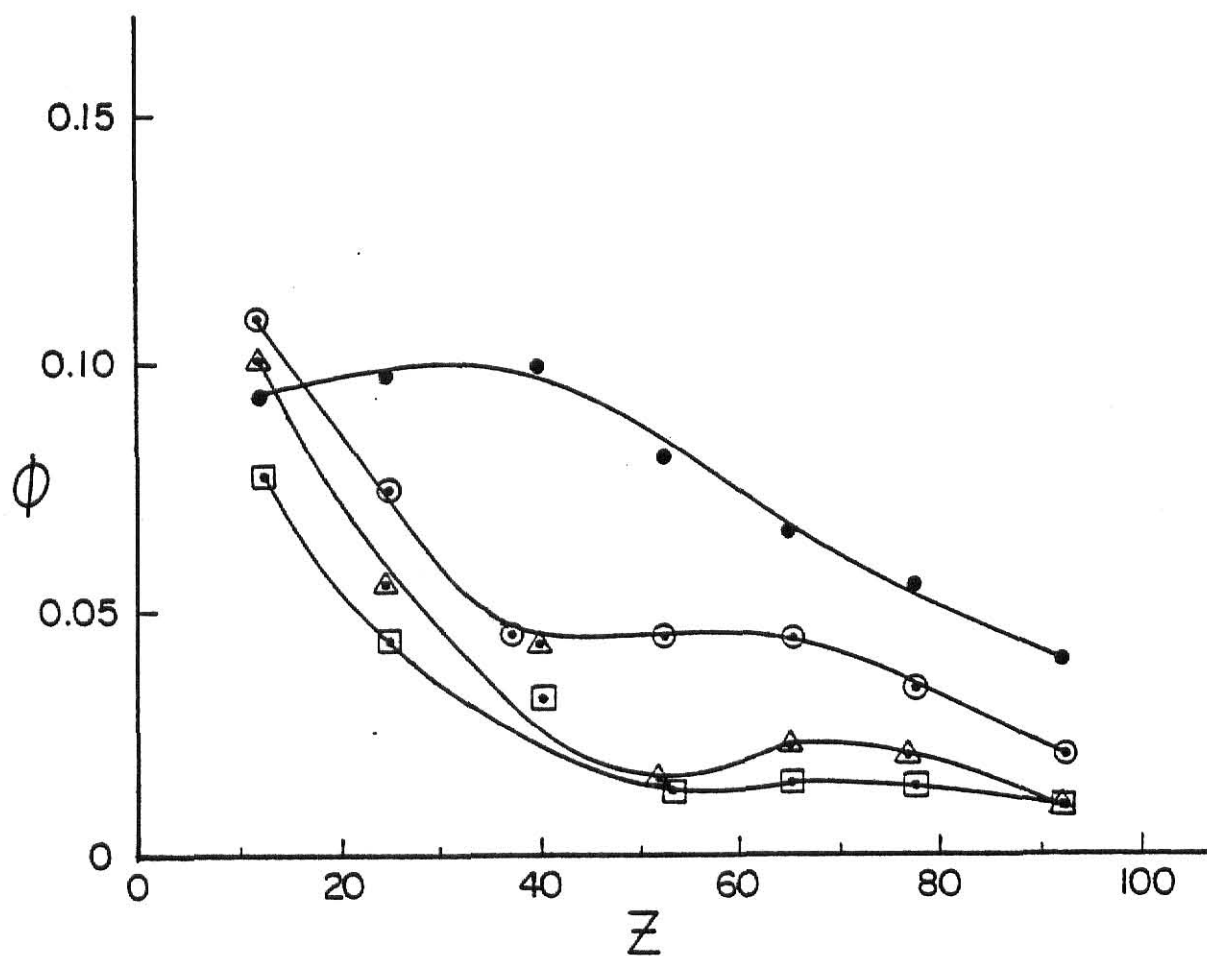


Fig. 15. Variation of gas holdup in downflow section of an airlift fermentor with distance from top of the baffle, z , at a superficial gas velocity of 400 cm/min and ungassed liquid height equal to baffle height. Concentrations of CMC in weight percent are $\bullet$ - 0%, $\circ$ - 0.74%, Δ - 1.0% and $\square$ - 1.5%.

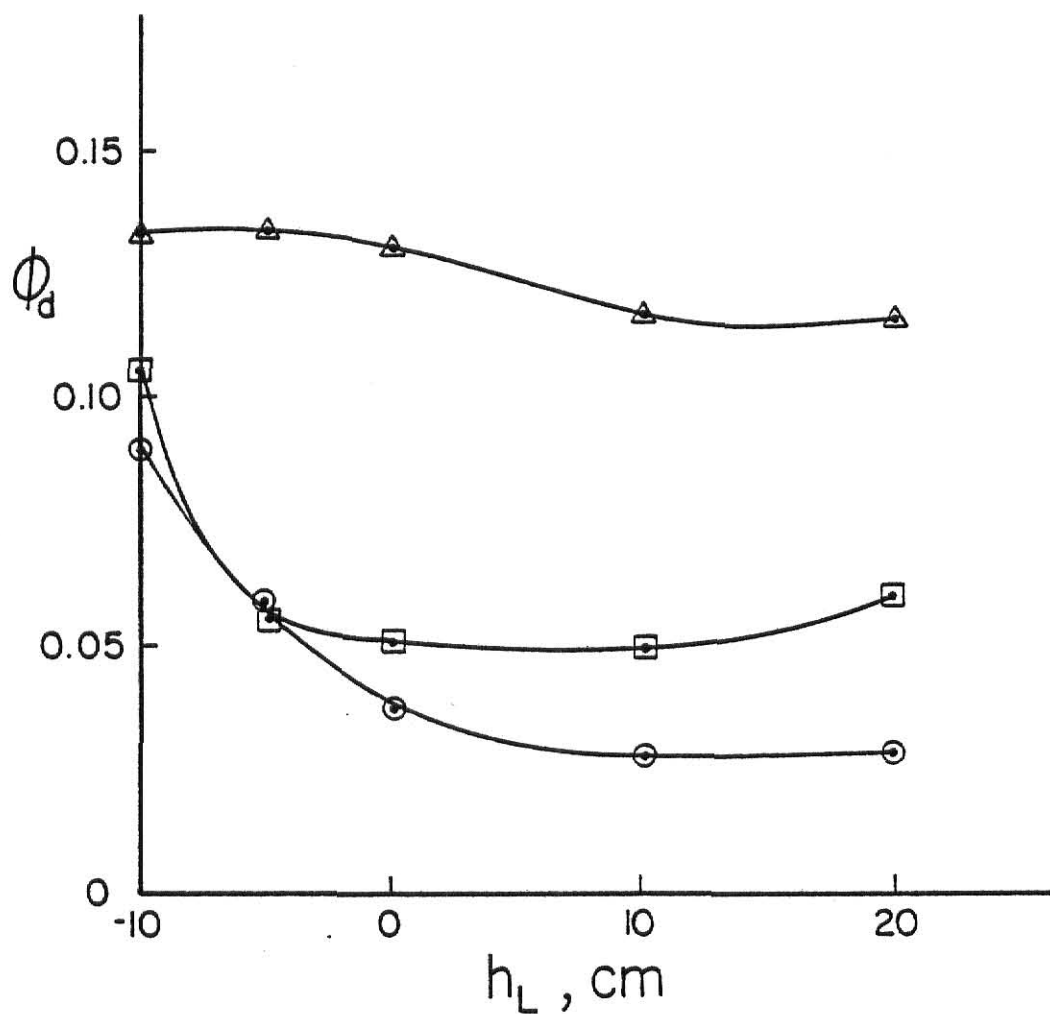


Fig. 16. Variation of gas holdup in downflow section of an airlift fermentor with ungassed liquid height above baffle for a superficial gas velocity of 400 cm/min. $\odot$ -1% CMC, Δ -1% CMC + 10 ppm NaLSO₄, and $\square$ -2% CMC + 10 ppm NaLSO₄.

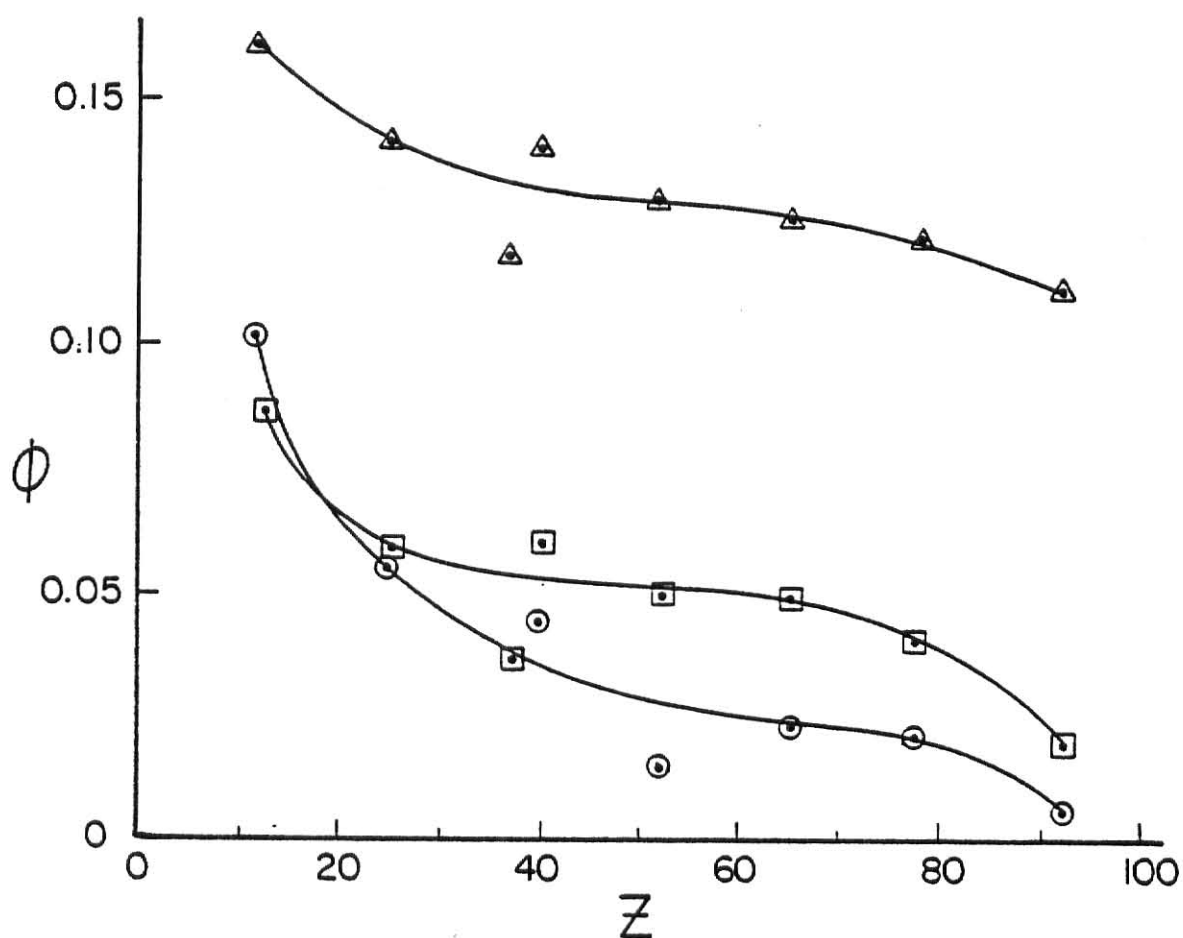


Fig. 17. Variation of gas holdup in downflow section of an airlift fermentor with distance from top of the baffle, z , for an ungassed liquid height equal to the baffle height and a superficial gas velocity of 400 cm/min. $\circ$ - 1% CMC, Δ - 1% CMC + 10 ppm NaLSO₄, and $\square$ - 2% CMC + 10 ppm NaLSO₄.

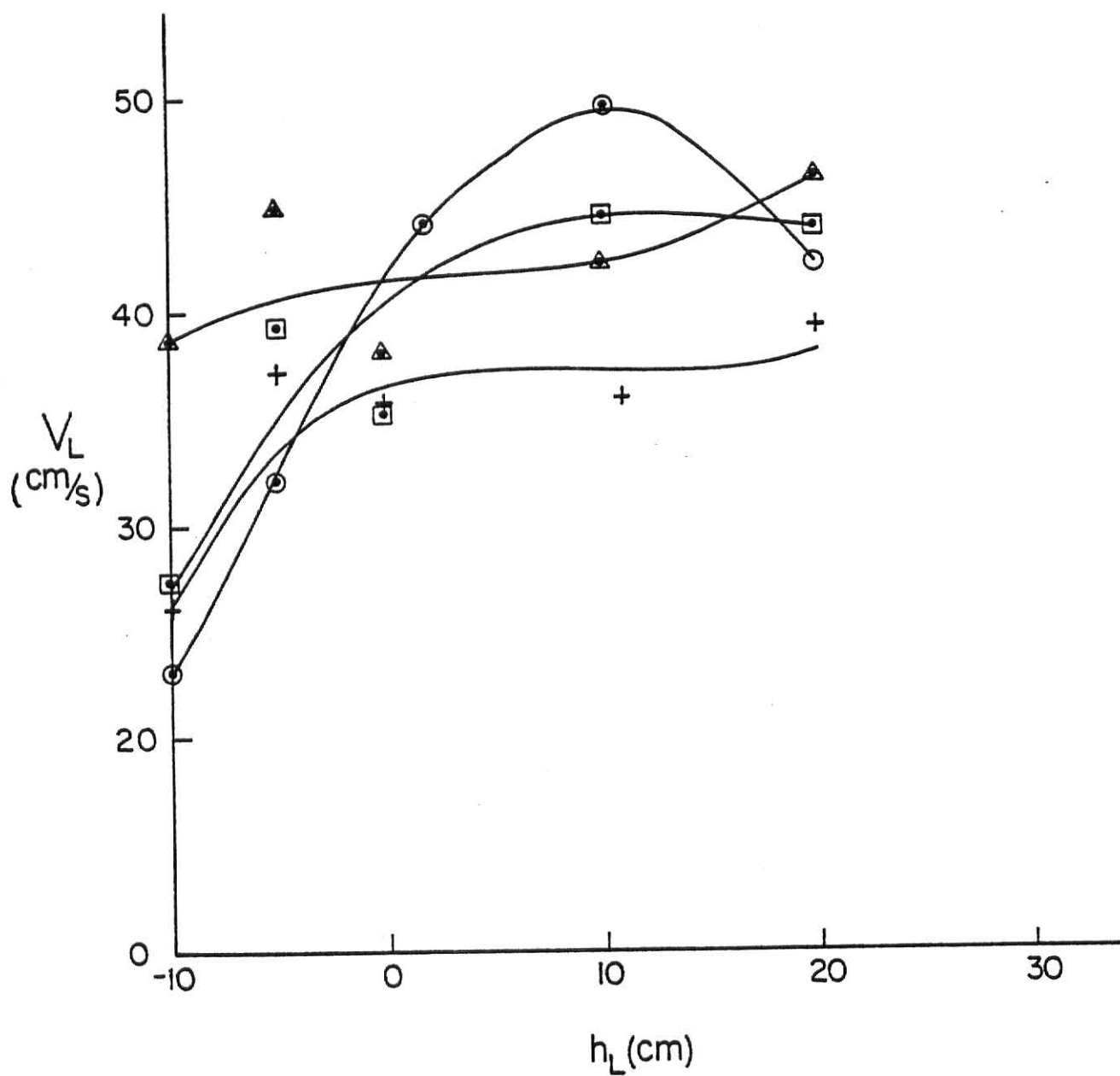


Fig. 18. Variation of liquid circulation velocity in an airlift fermentor with ungassed liquid height above baffle, h_L , at a superficial gas velocity of 400 cm/min for ○ -1% CMC, ◻ -10 ppm NaLSO₄, Δ -1% CMC with 10 ppm NaLSO₄, and + - distilled water.

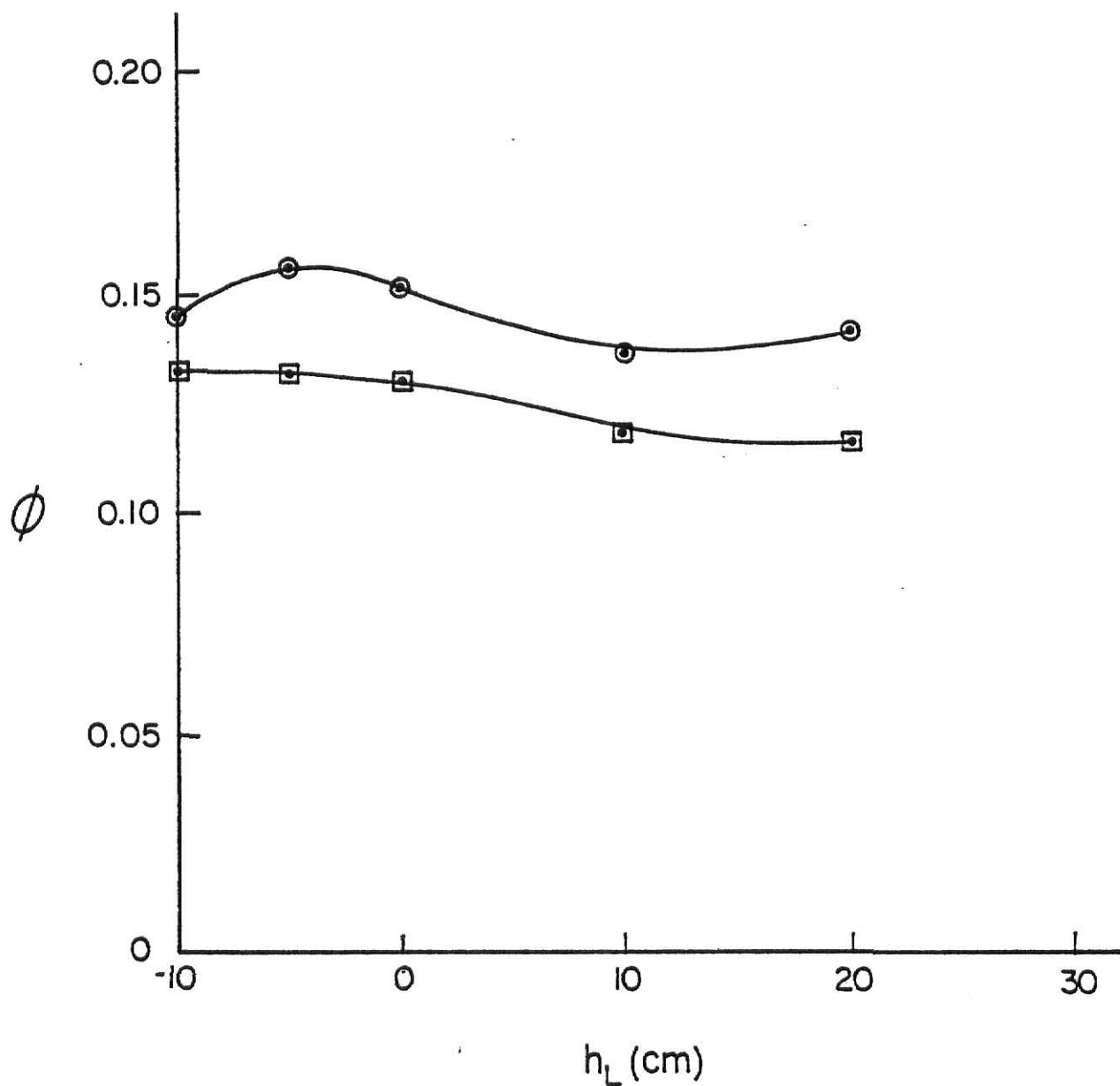


Fig. 19. Variation of gas holdup with ungassed liquid height above the baffle, h_L , at a superficial gas velocity of 400 cm/min in the $\odot$ - upflow and $\square$ - downflow sections of an airlift fermentor containing 1% CMC with 10 ppm Na_2SO_4 .

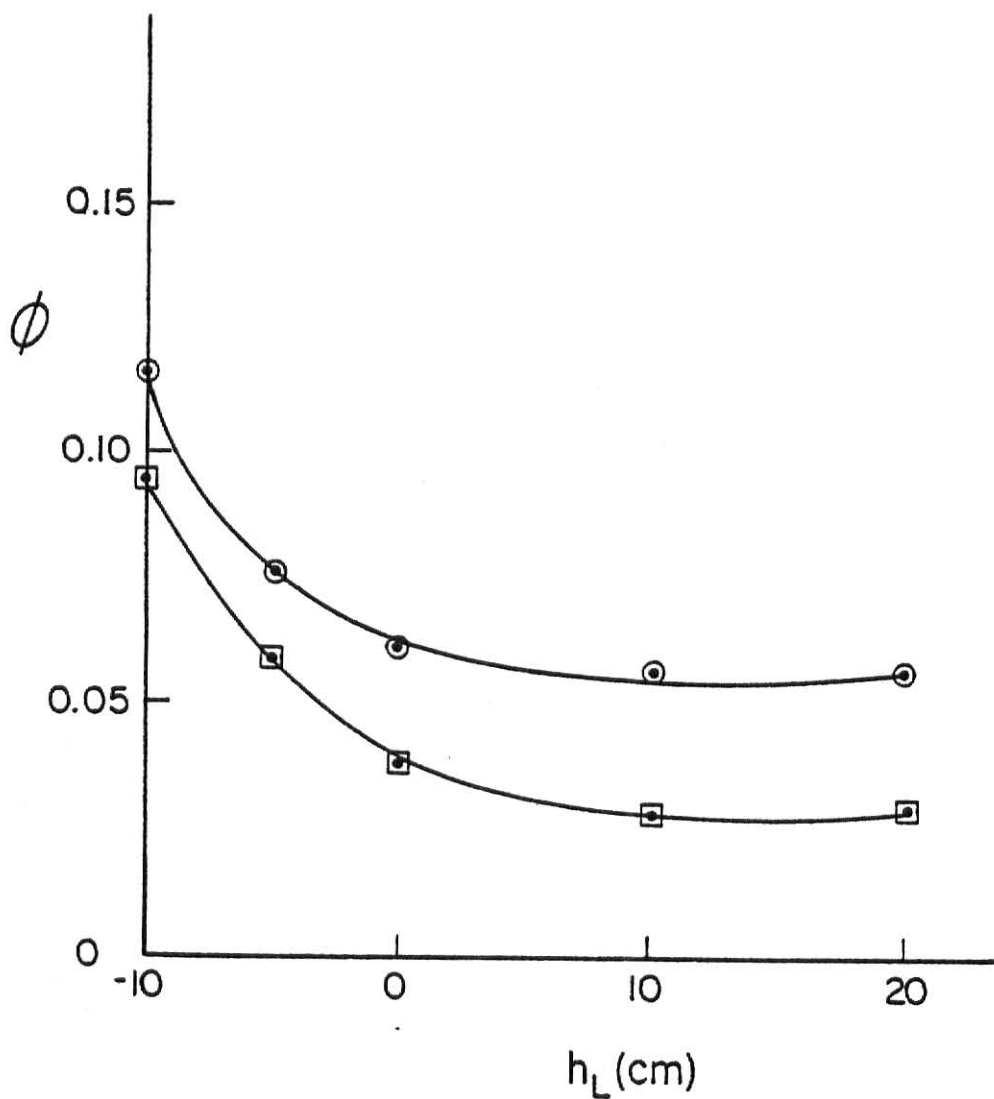


Fig. 20. Variation of gas holdup with ungassed liquid height above the baffle, h_L , at a superficial gas velocity of 400 cm/min in the $\circ$ - upflow and $\square$ - downflow sections of an airlift fermentor containing 1% CMC.

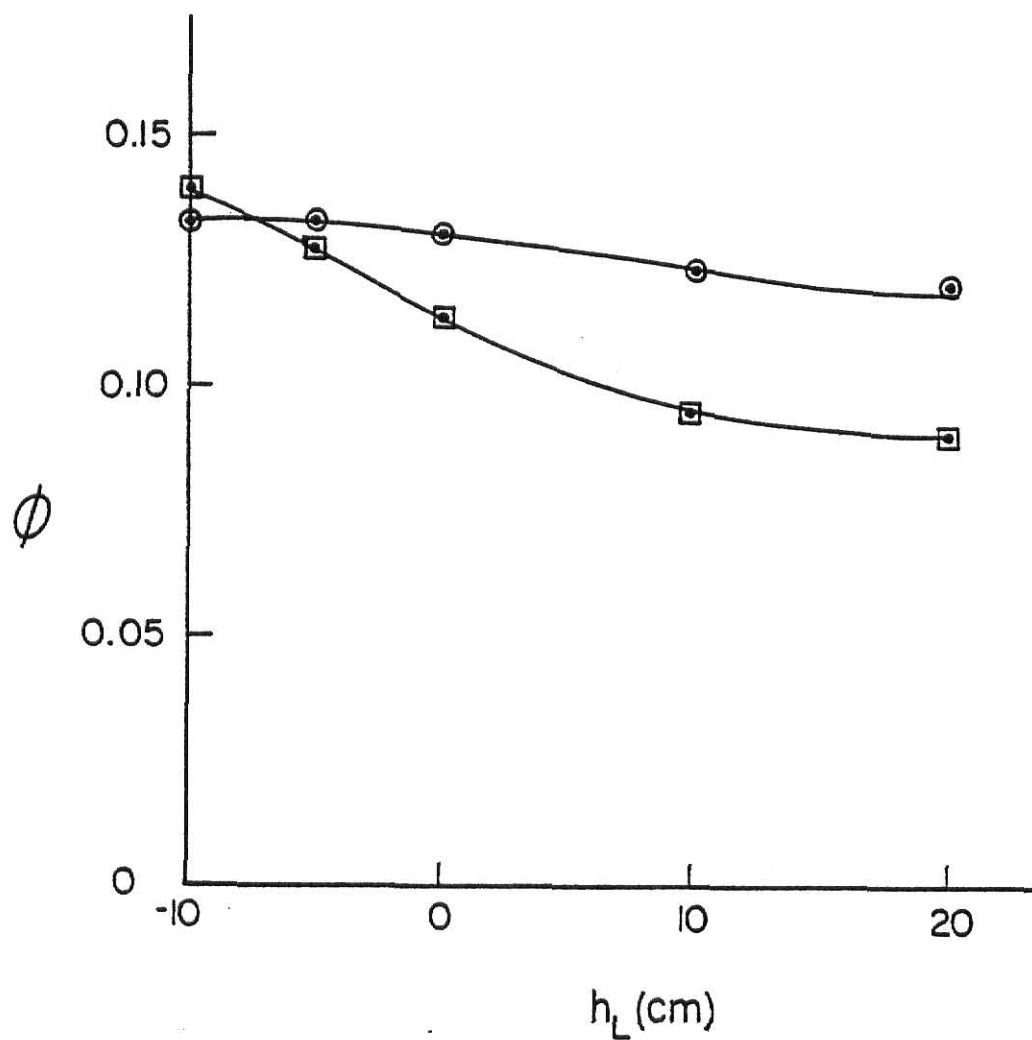


Fig. 21. Variation of gas holdup with ungassed liquid height above the baffle, h_L , at a superficial gas velocity of 400 cm/min in the $\odot$ - upflow and $\square$ - downflow sections of an airlift fermentor containing 10 ppm NaLSO_4 .

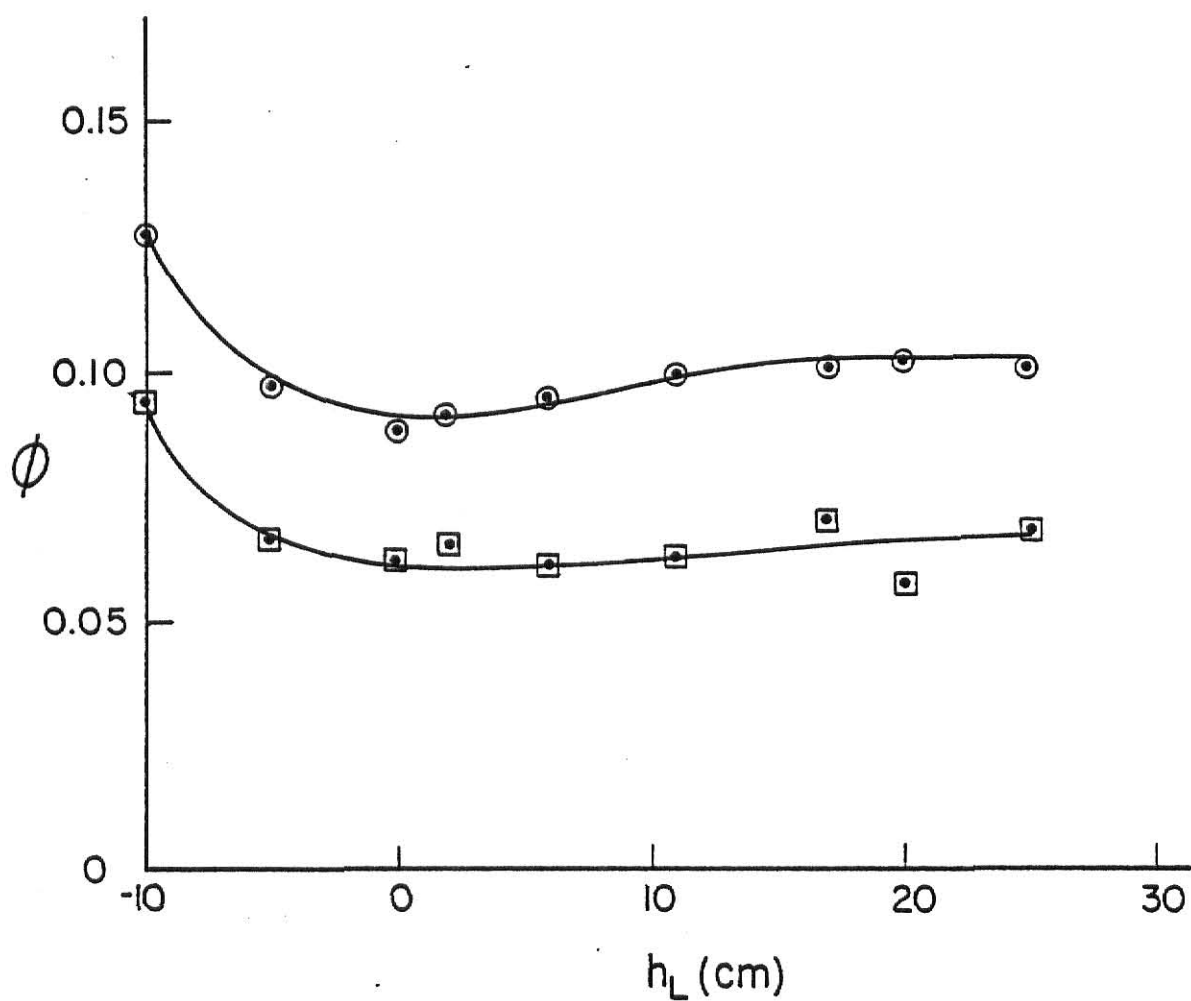


Fig. 22. Variation of gas holdup with ungassed liquid height above the baffle, h_L , at a superficial gas velocity of 400 cm/min in the $\odot$ - upflow and $\square$ - downflow sections of an airlift fermentor containing distilled water.

NOMENCLATURE

a	interfacial area per unit volume
A_d	cross-sectional area of the downflow section
A_u	cross-sectional area of the upflow section
C_1, C_2	empirical constants
C_D	drag coefficient for a bubble
C_f	friction factor
c_{fN}	friction factor for a non-Newtonian liquid
d	diameter of the airlift column
d_B	bubble diameter
d_{Be}	equivalent diameter of a bubble
d_{BM}	mean bubble diameter
d_o	orifice diameter
dp/dz	pressure gradient
D	equivalent diameter of the tube ($=4R_h$)
D_{hb}	hydraulic diameter of base
f	friction factor
f_{TP}	two-phase friction factor
$F_1(n)$	correction factor for non-Newtonian behavior [Hirose and Moo-Young (1969), eqn. (30)]
g	acceleration due to gravity
g_c	conversion factor
h_L	liquid height above the baffle
H	liquid height in the fermentor
K	consistency index
ℓ	scale of largest eddies
L_e	equivalent length
n	power-law index

P	pressure
ΔP_F	frictional pressure drop
ΔP_H	pressure drop due to hydrostatic head
ΔP_t	turnaround pressure drop
ΔP_{Total}	total pressure drop
P/V	power per unit volume
Q	volumetric gas flow rate
Q_{gd}	volumetric gas flow rate in the downflow section
Q_{gu}	volumetric gas flow rate in the upflow section
Q_L	volumetric liquid flow rate
Q_M	volumetric mean gas flow rate
Re	Reynolds number for a bubble ($=d_B V_r \rho_L / \mu_L$)
Re_L	liquid Reynolds number ($=DV_L \rho_L / \mu_L$)
Re_{LN}	liquid Reynolds number for a power-law liquid (Eqn.(39))
Re_n	bubble Reynolds number in a power-law liquid (Eqn.(38))
Re_M	mixture Reynolds number ($=DV_M \rho_M / \mu_L$)
R_h	hydraulic radius
v	bubble volume
V_ℓ	superficial liquid velocity
V_L	liquid circulation velocity
V_M	superficial velocity of a mixture
V_r	rise velocity of a bubble
V_s	superficial gas velocity
We	Weber number
X_n	drag coefficient for a power-law liquid

Greek Letters:

ϵ	rate of energy dissipation per unit mass
η	Kolmogoroff microscale
η_{app}	apparent viscosity
η_p	plastic viscosity
σ	surface tension or interfacial tension
μ_d	viscosity of the dispersed phase
μ_G	gas viscosity
μ_L	liquid viscosity
ρ_d	density of the dispersed phase
ρ_g	gas density
ρ_L	liquid density
ρ_M	mixture density
$\Delta\rho$	difference in densities between two phases
ν_L	kinematic viscosity of liquid
τ	dynamic pressure
τ_y	yield stress
ϕ	gas holdup
ϕ_T	total gas holdup

Subscripts:

u	upflow
d	downflow
L	liquid
g,G	gas

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APPENDIX

METHODS OF CALCULATION

The region of bubble rise velocity varies considerably with bubble Reynolds number. The following equations which are available in literature, were examined. Region I: Very small bubbles follow Stokes' eqn.

$$V_r = \frac{g \Delta \rho d_B^2}{18 \mu_L} \quad (1)$$

For low viscosity fluids, this equation is valid up to about $Re_b = 2$. (Peebles and Garber, 1953). The bubbles are spherical and rigid, and viscous forces dominate inertia and surface tension forces. Medium-size bubbles in viscous liquids are also likely to follow this equation at low bubble Reynolds numbers. Acharya et al. (1977) have shown that the bubble radius at which transition occurs may be predicted for viscous liquids. Region II: At higher Reynolds numbers, the bubble interface is freed and there is a transition to the Hadamard-Rybczynski region. Many workers have reported a jump in the bubble rise velocity in this region. The Hadamard equation is

$$V_r = \frac{g \Delta \rho d_B^2}{12 \mu_L} \quad (2a)$$

Peebles and Garber defined this region by

$$2 \leq Re \leq 4.02 \left[\frac{g \mu_L^4}{\rho_L \sigma^3} \right]^{-0.214} \quad (2b)$$

and suggested an empirical correlation for rise velocity, from their experiments on several liquids.

$$V_r = 0.33 g^{0.76} \left(\frac{\rho_L}{\mu_L} \right)^{0.52} \left(\frac{d_B}{2} \right)^{1.28} \quad (2c)$$

This indicates weaker influence of viscosity. It does not, however, account for the velocity jump. In the absence of more experimental data in the transition region, it appears more appropriate to use eqn. (2a) than eqn. (2c).

Region III: At higher Reynolds numbers, bubbles are deformed, and the rise velocity is independent of viscosity. Mendelson (1967) used the wave analogy to derive the following eqn.

$$V_r = \left(\frac{2\sigma}{\rho d_B} + g \frac{d_B}{2} \right)^{1/2} \quad (3)$$

For small bubble diameters, the second term is negligible in comparison with the first term, and bubble rise velocity decreases with increasing bubble diameter; however for large bubbles, the second term is more important (Acharya et al. 1977).

In the present work, eqn. (3) has been used throughout region III. It is not possible to define the boundary between regions II and III, as the inequality (2b) is no longer valid for highly viscous liquids. The equation giving lower rise velocity was chosen beyond Stokes region.

The results of the calculations are shown in Fig. 23.

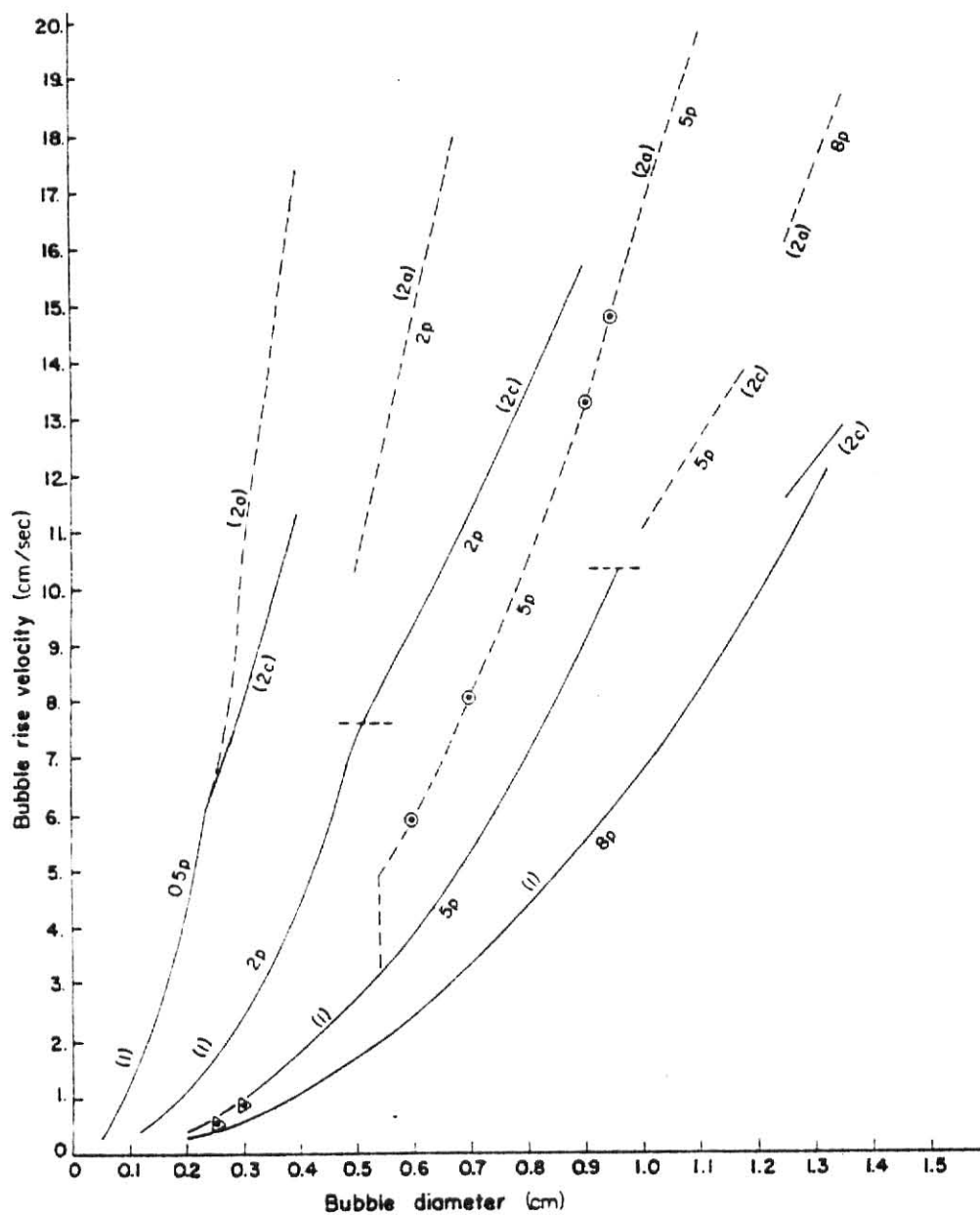


Fig. 23. Bubble rise velocities for different bubble diameters, with viscosity as parameter.

EXPERIMENTAL STUDIES OF GAS-LIQUID DISPERSION
IN A SPLIT-CYLINDER AIRLIFT FERMENTOR

by

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Studies were carried out on the effects of physical properties, e.g., viscosity, and presence of surfactants and electrolytes on the behavior of gas-liquid dispersions in a split-cylinder airlift fermentor. The axial variation of gas holdup was experimentally determined using open end manometers. The effect of viscosity on gas holdup in the downflow section of the column was studied using 0 to 1.5% solutions of carboxy methyl cellulose with apparent viscosities of up to about 2 poise. The gas holdup was much lower at higher viscosities. The axial variation of holdup in the downflow section was largest for the CMC solutions. The holdup in the bottom of the downflow section was only 10% of the holdup at top, for highly viscous solutions (1% CMC). For CMC solution containing a surfactant it was 68.8% of the holdup at top.

The addition of a surfactant (sodium lauryl sulfate) to distilled water and CMC solution increased holdup because of decreased coalescence. The variation of gas holdup with axial position was much smaller than for rapidly coalescing solutions of CMC and for distilled water. The addition of an electrolyte (sodium chloride) to distilled water also decreased coalescence and resulted in higher holdups. The variation of gas holdup with distance was, however, greater than in the case of surfactant. In both cases smaller bubble size was maintained.

The height of the ungassed liquid relative to the baffle height was investigated in order to examine its effect on holdup in the downflow section. It was found to be small. The greatest effects on holdup appeared to occur when the ungassed liquid height was below the baffle.