HYDRODYNAMIC CHARACTERISTICS OF THREE-PHASE FLUIDIZED BEDS

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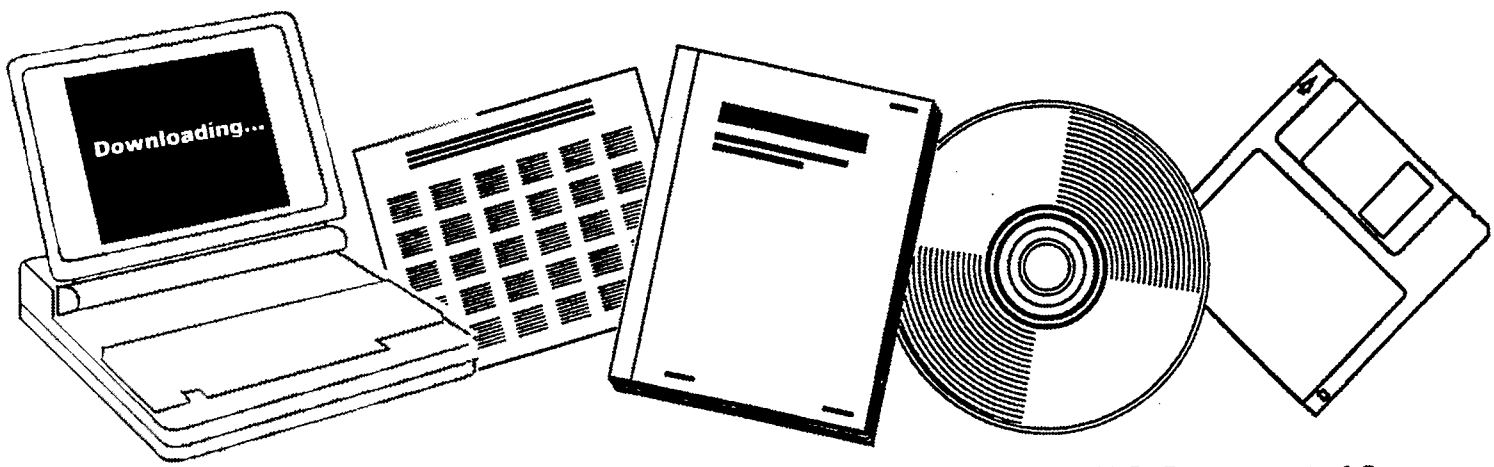

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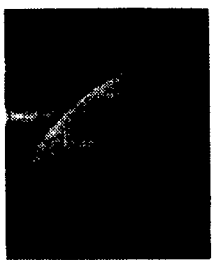

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HYDRODYNAMIC CHARACTERISTICS OF THREE-PHASE FLUIDIZED BEDS*

J. M. Begovich
J. S. Watson

Oak Ridge National Laboratory
Post Office Box X
Oak Ridge, Tennessee 37830

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HYDRODYNAMIC CHARACTERISTICS OF THREE-PHASE FLUIDIZED BEDS

By John M. Begovich and J. S. Watson

Chemical Technology Division

Oak Ridge National Laboratory

Oak Ridge, Tennessee 37830

SYNOPSIS

The hydrodynamics of three-phase (gas-liquid-solid) fluidized beds has been studied in two columns of 7.62 and 15.2 cm in diameter. The minimum gas and liquid velocities necessary to fluidize a bed were determined as a function of the particle size and density and the liquid viscosity; no effect of the initial bed height or column diameter was found. An electroconductivity technique was used to measure local concentrations of solid, liquid, and gas in the fluidized bed; these profiles were fitted using the error function. Overall solid holdup data were combined with over 1350 points from the literature to yield a dimensional correlation involving the physical parameters of the systems studied.

NOMENCLATURE

A	cross-sectional bed area	[cm^2]
Ar	Archimedes number = $d_p^3 \rho_L(\rho_S - \rho_L)\, g/\mu_L^2$	[-]
d_p	particle diameter	[cm]
erf(x)	error function = $\frac{2}{\sqrt{\pi}} \int_0^x e^{-z^2}\, dz$	[-]
Fr	Froude number = U_G^2/gd_p	[-]
g	gravitational acceleration	[cm/sec^2]
h	axial column position	[cm]
H	bed height	[cm]
I	inflection point in local holdup vs height curve	[cm]
M	mass	[g]
ΔP	pressure drop across bed	[dyn/cm^2]
P	curve-fitting parameter, defined in Eqs. (7)-(10)	[-]
Re	Reynolds number = $U_L d_p \rho_L/\mu_L$	[-]
U	superficial velocity	[cm/sec]

Greek Letters

ε	local holdup (volume fraction)	[-]
$\bar{\varepsilon}$	average holdup	[-]
μ	viscosity	[g/cm·sec]
ρ	density	[g/cm^3]
σ	standard deviation	[cm]
$\Phi(x)$	probability integral = $\frac{1}{\sqrt{2\pi}} \int_0^x e^{-w^2/2}\, dw$	[-]

Subscripts

G	gas phase
L	liquid phase
mf	minimum fluidization
S	solid phase

Superscripts

''	two-phase
'''	three-phase

INTRODUCTION

Three-phase fluidized beds, containing solid particles fluidized by the upward cocurrent flow of liquid and gas which form continuous and discontinuous phases, respectively, have important present and future applications in hydrocarbon and coal processing and in some biological reactors. However, accurate design of such reactors is complicated by such factors as (1) lack of knowledge of the minimum fluid velocities required to achieve fluidization, and (2) axial variations in reactor properties, particularly distribution of the solid phase. No published data or equations exist for predicting reactor performance under high fluid flow rates where axial variations are important, and only a limited amount of data exists (Vail et al., 1970; Burck et al., 1975; Bloxom et al., 1975) for predicting

minimum fluidization (MF) velocities in three-phase fluidized beds.

The minimum fluid flow rates required to achieve fluidization are determined by a plot of the pressure drop across the bed vs the superficial liquid velocity at constant gas flow rate. When fluidized, the pressure drop across the bed will no longer change with increasing liquid flow rate. Thus the flow rates at which a break in the curve occurs correspond to the MF velocities.

The following equations have typically been used to determine the volume fraction (holdup) of each phase in a three-phase fluidized bed:

$$\varepsilon_L + \varepsilon_G + \varepsilon_S = 1 \qquad (1)$$

$$\Delta P = gH(\rho_L\varepsilon_L + \rho_G\varepsilon_G + \rho_S\varepsilon_S) \qquad (2)$$

$$\varepsilon_S = M_S/\rho_S AH \ , \qquad (3)$$

where the bed height in Eqs. (2) and (3) is obtained either visually or from the measured pressure gradient (Kim et al., 1975; Bhatia and Epstein, 1974). At high flow rates, neither method is satisfactory because the indistinct bed height makes visual measurements extremely subjective, while the measured pressure gradient yields a bed height based on an unrealistic homogeneous bed.

Experimental

To provide an alternative equation to Eq. (3), the electroconductivity method of Achwal and Stepanek (1975, 1976) was modified for application to a three-phase fluidized bed. The apparatus used in this study, as well as details of the procedure to obtain the holdup for each phase as a function of axial position within the column, have been described elsewhere (Begovich and Watson). Briefly, however, the experiment involves the use of various solids fluidized by air and water in either a 7.62- or a 15.2-cm-ID column. A series of liquid manometers located at regular intervals along the column walls provided the pressure gradient in the bed. Two small pieces of platinum sheet that were attached to opposite sides of a movable plexiglass ring and connected to a conductivity meter permitted measurement of the electrical conductivity of the bed at any axial position in the column. Using the ratio of conductivities in the bed to those in the liquid alone as the liquid holdup, together with the measured pressure gradient and Eq. (1), all three phase holdups could be determined as a function of height in the column.

The MF velocities required to achieve fluidization were determined from the intersection of the static and fluidized bed pressure drop curves on the plot of bed pressure drop vs superficial liquid velocity at a constant gas flow rate. The range of experimental conditions used in this study are detailed in Table 1.

Table 1. Range of experimental conditions used in three-phase fluidization studies

U_G, cm/sec	:	0 - 17.3
U_L, cm/sec	:	0 - 12.0
Column diameter, cm	:	7.62 and 15.2
Initial bed height, cm	:	22 - 45

RESULTS

Minimum Fluidization

The effects of column diameter and static bed height (or bed mass) on the minimum fluidization velocities for the air-water-4.6-mm-glass beads system are shown in Fig. 1. The minimum liquid velocity required to fluidize the bed with no gas phase present is indicated by the arrow on the ordinate of the plot as calculated from the two-phase correlation of Wen and Yu (1966). Excellent agreement between the calculated and experimental point of this system can be observed, as was the case for each system studied.

Neither the column diameter nor the mass of solids present in the column appeared to have any significant effect upon the MF velocities. Since fluidization of a bed is achieved when the upward inertial and drag forces exerted on the particles by the fluids equals the buoyant weight of the bed, an effect of static bed height on the MF velocities would only be expected if end effects were present in the bed. Likewise, one would not expect the MF velocities to be a function of column diameter unless the size of the gas bubbles approached that of the column diameter or unless channeling occurred.

MF velocities are shown in Fig. 2 for each of the systems studied. As the gas velocity was increased, the minimum liquid velocity required to achieve fluidization in each of the systems decreased. The magnitude of this decrease is considerably different for the plexiglass beads with their small solid/liquid density difference. In their two-phase correlation, Wen and Yu (1966) noted that the MF velocity increases with increasing particle diameter and increasing solid/liquid density difference but decreases with increasing fluid viscosity. The plexiglass beads have the same diameter as the alumina and one of the glass beads and also have a much smaller solid/liquid density difference. Thus they fluidize at lower velocities. The alumina and alumino-silicate beads have approximately the same density, but the smaller diameter of the latter particles causes them to fluidize at lower velocities.

Likewise, the 4.6-mm-diam glass beads fluidize at lower velocities than do the 6.2-mm-diam glass beads. It is of interest to note that although the curves of the alumina and 6.2-mm-diam glass beads start at essentially the same point for zero gas velocity, for increasing gas velocities they rapidly diverge until the gas velocities reach in excess of 8 cm/sec; at this point, the curve for the alumina beads merges with the curve of the 4.6-mm-diam glass beads.

The effect of liquid viscosity on the MF velocities for the 4.6-mm-diam glass beads is shown in Fig. 3. A single smooth curve in Fig. 1 fits all data for the same glass beads in both the 7.62- and 15.2-cm-ID columns. The arrows shown for zero gas velocity in Fig. 3 are calculated values and are shown only to indicate the expected influence of liquid viscosity. For a given gas velocity, the minimum liquid fluidization velocity decreased as the liquid viscosity was increased. However, the influence of the liquid viscosity appeared to decrease for the higher viscosities, particularly at the higher gas velocities. The gas velocity itself did not appreciably affect the minimum liquid fluidization velocity for the more viscous aqueous glycerol solutions studied.

Axial Variation in Holdups

As mentioned previously, bed heights are indistinct at high fluid flow rates, and the holdups calculated using Eqs. (1)-(3) represent those assuming an unrealistic homogeneous bed. Using the electroconductivity of the bed and the measured pressure gradient allows the holdups to be determined. Liquid and solid holdup values are plotted as a function of height, as shown in Fig. 4. The solid holdup, typically, is fairly uniform in the lower section of the bed; however, near the top of the bed, a transition region of slowly decreasing solids concentration connects with a two-phase gas-liquid region above the bed. The liquid holdup remains fairly constant in the bed and then increases to a constant value in the gas-liquid region. Figure 4 also shows that the same relationship between the holdups (solid and liquid) and height in the column was obtained in both the 7.62- and the 15.2-cm-ID columns under identical conditions (i.e., identical gas velocities, liquid velocities, particle type, and static bed height).

The effect of liquid velocity on the axial variation in the 4.6-mm-diam glass bead holdup is shown in Fig. 5 under conditions of constant gas velocity in the 7.62-cm-ID column. As the liquid velocity was increased, the bed expanded, and thus the solid holdup decreased. The calculated bed height, as found from the intersection of the measured pressure gradients in and above the bed, is indicated on the curves for each flow rate. As noted previously by Bhatia and Epstein (1974), this bed height corresponds to the height the same bed would have if the solids concentration in the column were uniformly distributed. The highest position where solids were detected was higher than this calculated bed height, however, since the bed contains a rather wide transition region over which flow changes from a three-phase to a two-phase column. The width of this transition region appeared to remain fairly constant with changing liquid velocity; that is, the solid holdup decreased from the approximately constant value in the bed to zero over about 20 cm of column length.

When the liquid velocity was held constant and the gas velocity was increased, the width of the transition region increased substantially, as illustrated in Fig. 6 which is typical of all of the data. The solid holdup in the lower portion of the bed was decreased slightly by the increase in gas velocity; however, the transition region increased from 20 cm in width to approximately 35 cm. As expected, the calculated bed height for the higher gas velocity indicated a much lower bed height than that observed visually (highest position with solids).

These results demonstrate the shortcomings of assuming a distinct bed height and a uniform bed. The transition region is a significant fraction of the total bed height and must be considered in realistic designs of three-phase systems. In commercial units operating with taller beds, the transition region could be less significant; however, the higher gas rates often employed by such units could cause the transition region to remain a significant fraction of the total bed height.

Overall-Phase Holdups

Figures 5 and 6, in addition to indicating the axial variation of the solid phase holdups, also demonstrate that Eqs. (1)-(3) are approximately correct for calculating the bed height via the pressure gradients (Epstein, 1977). The equivalent homogeneous bed may be sufficient as a model for the actual bed in some applications.

Since all the data necessary to use Eqs. (1)-(3) for predicting equivalent homogeneous bed heights were obtained simultaneously with the measurements for the local holdups, it was a simple matter to calculate overall, or average, phase holdups based on an equivalent homogeneous bed. For the systems studied, increasing the liquid velocity and holding all other conditions constant resulted in: (1) a decrease in the overall solid holdup, and (2) a slight decrease in the overall gas holdup. Since the holdups must sum to unity, the increased liquid velocity thus increased the overall liquid holdup. At constant liquid velocity, increasing the gas velocity caused the overall gas holdup to increase but only slightly decreased the overall solid holdup. Using aqueous glycerol solutions to evaluate the effects of higher liquid

viscosities on this same system, Bloxom et al. (1975) showed that the overall solid holdup decreased and the overall liquid holdup increased with increasing viscosity, while the overall gas holdup was unaffected. These results are in good agreement with previous investigators (Burck et al., 1975; Kim et al., 1975; Bhatia and Epstein, 1974; Michelsen and Ostergaard, 1970; Bruce and Revel-Chion, 1974; Dakshinamurty et al., 1971; Efremov and Vakhrushev, 1970; Mukherjee et al., 1974; Ostergaard and Michelsen, 1968; Ostergaard and Thiesen, 1966; Rigby and Capes, 1970; Ostergaard, 1965). All liquid-solid fluidization systems studied in these experiments expanded upon introducing gas into the bed, in agreement with the results predicted by the criterion developed by Bhatia and Epstein (1974) and Epstein (1976).

CORRELATION OF RESULTS

Minimum Fluidization

The minimum liquid fluidization velocities shown in Figs. 2 and 3 were correlated with the system parameters and resulted in the following dimensionless correlation:

$$Re_{mf} = a\, Ar^{b}\, Fr^{c} , \qquad (4)$$

where the constants and their 99% confidence limits are:

$$a = 5.121 \times 10^{-3} \quad \pm 0.004$$
$$b = 0.662 \quad \pm 0.062$$
$$c = -0.118 \quad \pm 0.048 .$$

Eq. (4) had a correlation coefficient of 0.94 and an F-value of 440 using a total of 112 points.

Unfortunately, Eq. (4) is not valid for zero gas flow rate. To produce a three-phase correlation that degenerates to an acceptable two-phase correlation as the gas flow rate goes to zero, the MF velocity predicted by the two-phase correlation of Wen and Yu (1966) was used and resulted in the following:

$$U_{L,mf}/U_{L,mf}'' = 1-U_G^{a}\, \mu_L^{b}\, d_p^{c}\, (\rho_S-\rho_L)^{d} , \qquad (5)$$

where the exponents and their 99% confidence limits are:

$$a = 0.436 \pm 0.088 \qquad c = 0.598 \pm 0.289$$
$$b = 0.227 \pm 0.058 \qquad d = -0.305 \pm 0.146 .$$

Eq. (5) had a correlation coefficient of 0.93 and an F-value of 179 using a total of 125 data points. The dimensional correlation of Eq. (5) is somewhat less satisfactory statistically, but it behaves correctly as gas velocity approaches zero.

Local Holdups

Figures 4-6 clearly indicate that each of the holdups is fairly constant in two regions: (1) the lower portion of the bed, and (2) the gas-liquid region above the bed. The transition region between these two extremes was seen to depend on the gas velocity and the physical characteristics of the solid particles. An inflection point was observed on each curve with a spread about that point proportional to the width of the transition region. If each curve were differentiated, these two parameters would correspond to the mean and standard deviation of the normalized Gaussian curves. The error function was used to fit the gas and solid holdup curves, and the liquid holdup curve was determined as the residual of Eq. (1). Use of the error function is essentially equivalent to use of the probability integral, since the two are related by the following:

$$\mathrm{erf}(x) = 2\,\Phi\,(\sqrt{2}\,x) . \qquad (6)$$

Thus the gas holdup curves could be fitted by the following:

$$\varepsilon_G = [(P_G-1)/-2]\varepsilon_G''' + [(P_G+1)/2]\varepsilon_G'' , \qquad (7)$$

where

$$P_G = \mathrm{erf}[(h - I_G)/\sigma_G] . \qquad (8)$$

The solid holdup was fitted in a similar manner using the error function and the knowledge that the solid holdup in the gas-liquid region of the column is zero:

$$\varepsilon_S = [(P_S + 1)/2]\varepsilon_S''' , \qquad (9)$$

where

$$P_S = -\mathrm{erf}[(h - I_S)/\sigma_S] . \qquad (10)$$

The liquid holdup at each point was obtained from the residual of Eq. (1).

Thus knowledge of six parameters — ε_G''', ε_G'', ε_S''', σ_G, σ_S, and I_G — allows one to construct each of the phase holdups vs axial column position curves. Treatment of the experimental data in this way and correlation of the six parameters with fluid and solid properties and experimental conditions is in progress. The fit of the error functions to experimental data is illustrated in Fig. 7. For the system shown, the six parameters are:

$$\varepsilon_G''' = 0.072 \qquad \sigma_G = 2.64 \text{ cm}$$
$$\varepsilon_G'' = 0.129 \qquad \sigma_S = 2.83 \text{ cm}$$
$$\varepsilon_S''' = 0.511 \qquad I_G = 45.7 \text{ cm} .$$

Overall Phase Holdups

The overall solid holdups from these studies were combined with 1355 points from the literature (Kim et al., 1975; Bhatia and Epstein, 1974; Michelsen and Ostergaard, 1970; Bruce and Revel-Chion, 1974; Dakshinamurty et al., 1971; Efremov and Vakhrushev, 1970; Ostergaard and Michelsen, 1968; Ostergaard and Thiesen, 1966; Rigby and Capes, 1970; Ostergaard, 1965) to yield the following dimensional correlation:

$$1 - \varepsilon_S = a\, U_L^{\,b}\, U_G^{\,c}\, (\rho_S - \rho_L)^d\, d_p^{\,e}\, \mu_L^{\,f}\, D_c^{\,g}\,, \quad (11)$$

where

a = 0.371 ± 0.017	e = −0.268 ± 0.010
b = 0.271 ± 0.011	f = 0.055 ± 0.008
c = 0.041 ± 0.005	g = −0.033 ± 0.013 .
d = −0.316 ± 0.011	

Eq. (11) had a correlation coefficient of 0.87, an F-value of 1178, and was based on a total of 2381 points.

Combining the gas holdup with 169 points available from the literature (Kim et al., 1975; Bhatia and Epstein, 1974; Michelsen and Ostergaard, 1970; Efremov and Vakhrushev, 1970; Ostergaard and Michelsen, 1968) resulted in the following correlation:

$$\varepsilon_G = (0.048 \pm 0.010)\, U_G^{\,0.720\pm0.028}\, d_p^{\,0.168\pm0.061}\, D_c^{\,-0.125\pm0.088}\,. \quad (12)$$

Eq. (12), based on a total of 913 points, had a correlation coefficient of 0.93 and an F-value of 1793. Note that Eq. (11) does not hold for zero gas velocity, for which it would predict a solid holdup of unity. In fact, it is recommended that the equations presented herein not be used for conditions far removed from those tested.

CONCLUSIONS

The minimum gas and liquid velocities required to fluidize various types of solids have been determined as a function of particle size, particle density, and liquid viscosity; no effect of the initial bed height or column diameter was found.

An electroconductivity technique was adapted for use in three-phase fluidized beds and allowed each of the three-phase holdups to be determined as a function of height in the columns. The transition region where the solids concentration drops to zero was found to increase in width with increasing gas velocity, while remaining fairly constant in width with changing liquid velocity.

Using six parameters determined from the local gas and solid holdup profiles, it was possible to fit each of the holdup vs height in the column curves. Correlation of these parameters should give a reactor designer more information concerning vitally important phase distributions than that available using the homogeneous bed model.

ACKNOWLEDGMENT

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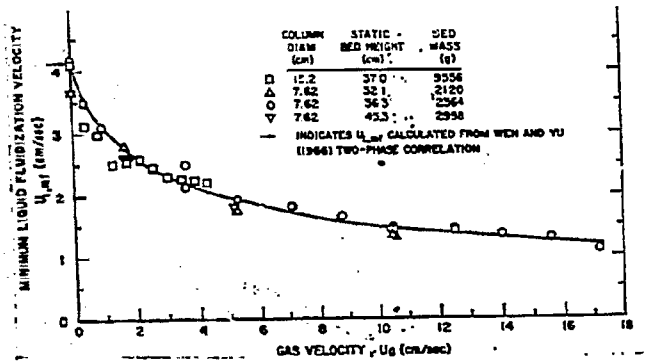

Fig. 1. MF velocities for the air-water-glass beads system.

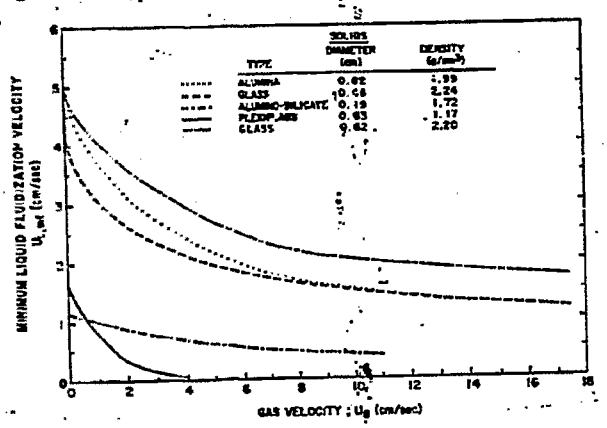

Fig. 2. Effect of particle size and density on the MF velocities.

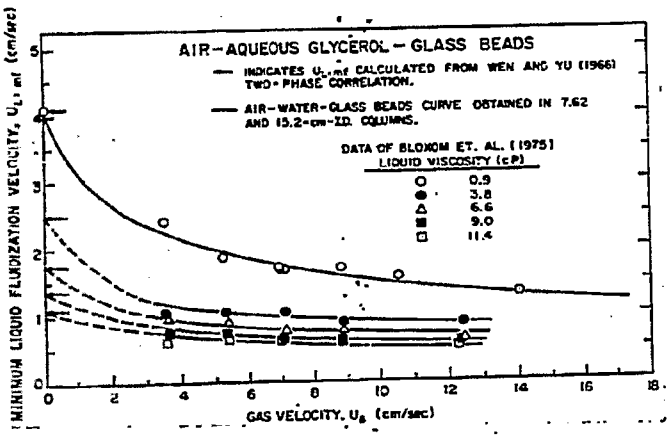

Fig. 3. Effect of liquid viscosity on the MF velocities.

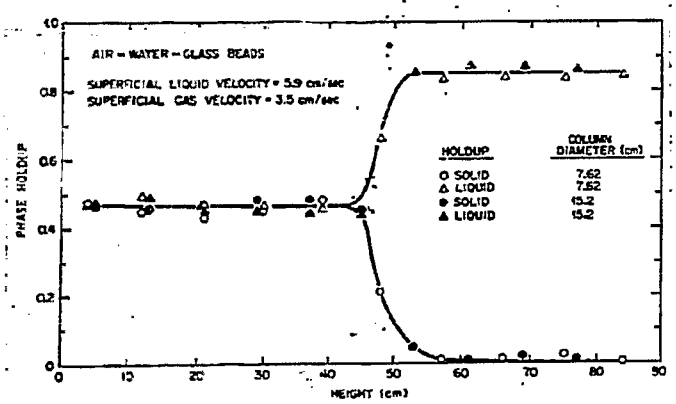

Fig. 4. Effect of column diameter upon the axial variation in holdups.

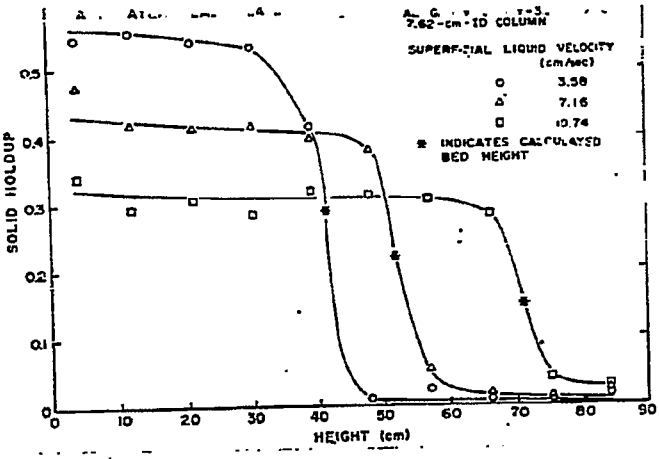

Fig. 5. Effect of liquid velocity upon the axial variation in the solid holdup.

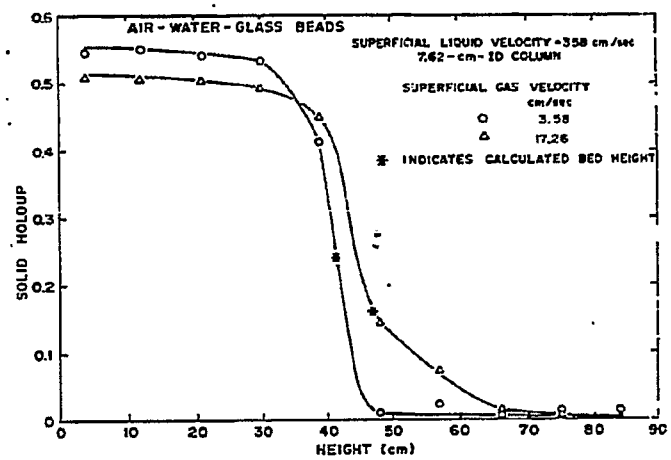

Fig. 6. Effect of gas velocity upon the axial variation in the solid holdup.

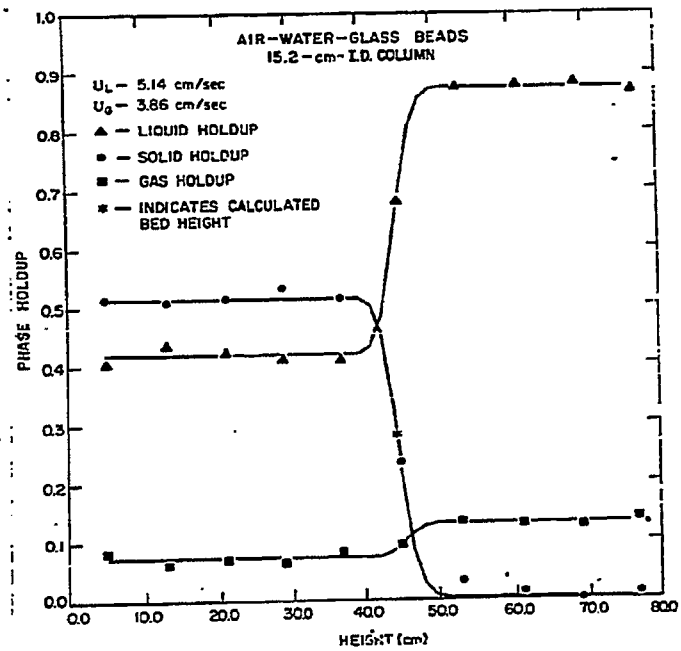

Fig. 7. Fit of local holdup profiles for the air-water-glass beads system.

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