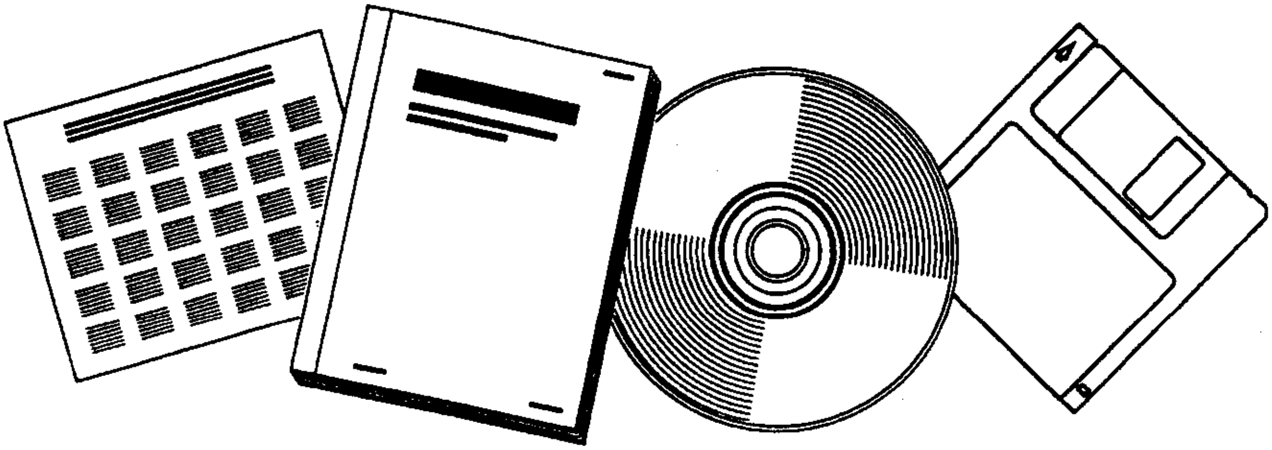
DE82007255

SOLVENT-REFINED-COAL (SRC) PROCESS. HYDRODYNAMICS AND AXIAL MIXING IN A THREE-PHASE BUBBLE COLUMN

PITTSBURG AND MIDWAY COAL MINING CO.
ENGLEWOOD, CO

JAN 1982



U.S. DEPARTMENT OF COMMERCE
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082094

DOE/ET/10104-28
(DE82007255)

FOSSIL ENERGY

HYDRODYNAMICS AND AXIAL MIXING IN A THREE-PHASE BUBBLE COLUMN

Solvent Refined Coal (SRC) Process

By
S. Kara
B. G. Kelkar
Y. T. Shah

January 1982

Work Performed Under Contract No. AC05-76ET10104

The Pittsburg & Midway Coal Mining Company
Englewood, Colorado

U. S. DEPARTMENT OF ENERGY

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DOE/ET/10104-28
(DE82007255)
Distribution Category UC-90d

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Vol. V: Research & Development Technical Support Programs
Part 3: Modeling and Simulation

Research and Development Report No. 53
Interim Report No. 51

January 1982

The Pittsburg & Midway Coal Mining Co.
Post Office Box 3396
Englewood, Colorado 80155

Prepared for the
U. S. Department of Energy
Division of Coal Conversion and Utilization
Under Contract No. DEAC-0579-ET10104

HYDRODYNAMICS AND AXIAL MIXING IN A THREE-PHASE BUBBLE COLUMN

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March 1981

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ABSTRACT

The hydrodynamics and the mixing properties in a 15.2 cm diameter x 335.3 cm high stainless steel column with three phases (air, water and coal) in cocurrent upflow were studied. The effect of solids particles size, solids concentration, slurry velocity and gas velocity on gas holdup and heat dispersion coefficients were determined. All the experimental data were correlated empirically and the results were compared with other published literature. Theoretical explanations of some results were provided.

INTRODUCTION

Three phase bubble columns are widely used in the chemical process industries as absorbers or reactors because of their low maintenance costs, simplicity of construction and absence of moving parts. They are typically applied to hydrogenation, chlorination, and sulfonation reactions. In recent years, they are widely used in coal utilization and conversion processes, such as direct coal liquefaction, oxydesulfurization of coal etc. Although, easy to construct, the proper design and scaleup of a bubble column requires a thorough understanding of the prevailing hydrodynamic and mixing properties associated with it.

In recent years the holdups of gas, liquid and solid phases in three-phase slurry operations have been examined using a number of techniques, such as pressure drop measurement, mean residence time distribution, x-ray transmission, gamma radiation, and electroconductivity. Recently, Shah (1979) has reviewed these techniques, quite extensively. The published literature (Imafuku et al., 1968; Kato et al., 1972; Viswanathan et al., 1965; Ostergaard and Michelsen, 1968), indicated that for particles less than 1 mm size, gas holdup was significantly reduced by the presence of solid particles. This was attributed to the fact that small particles promote bubble coalescence which results in higher rising velocities. The effect of larger particles was found to be less significant, since these particles instead tend to cause breakup of bubbles. The effect of solids concentration on gas holdup were investigated by Kato et al. (1972) and Ying et al. (1980). The work by Kato et al. (1972) showed that the increase in solid concentrations generally decreased gas holdup, but the effect becomes insignificant at high gas velocities (greater than 10 to

20 cm/sec). However, at low gas velocities (< .1 ft/sec) existence of solids did not change the holdup according to Ying et al. (1980). In general, an increase in gas velocity increased the gas holdup (Metzner and Brown, 1956; Viswanathan et al., 1965; Ostergaard, 1964, Kim et al., 1972, 1975; Kato et al., 1972). The effect of liquid velocity on gas holdup appeared to depend upon the free settling velocity of the particulate phase. In beds of large and heavy particles, the gas holdup decreased with liquid velocity; whereas in the beds of smaller particles, gas holdup was found to be independent of liquid velocity (Sherrard, 1966; Ostergaard and Michelsen, 1968; Kim et al., 1972, 1975; Ying et al., 1980). Some studies also reported an increase in gas holdup with decrease in liquid velocity (Michelsen and Ostergaard, 1970: Chen, 1975, 1976). By examining a wide range of liquid velocity (0-2.5 m/s), Hills (1976) reported similar results for an air-water system.

The extent of backmixing in the individual phases of multiphase bubble columns has been largely evaluated with the help of one dimensional axial dispersion model. Other models, such as those by Deckwer (1973), Baird and Rice (1975) and Joshi (1980) have also been proposed to explain the experimental data. The experimental data for backmixing in three phase fluidized bed were reported by Kato et al., (1972); Ostergaard, (1978); Michelsen and Ostergaard, (1970); and Vail et al., (1970). These data were obtained for large (> 1 mm) and heavy particles where the terminal velocity of the particles was significant. The data showed that the axial dispersion coefficient increased with gas velocity but remained essentially independent of liquid velocity. The information on bubble size distribution, in the presence of solids was not available in the published literature. Recently, Ying et al. (1980) have shown that the presence of solids decreased the axial dispersion coefficient in five and twelve-inch diameter columns. In their work solid particle sizes, less than 100 mesh, were used. Solid concentration was ranging from zero to 12 lb/ft^3.

The purpose of this paper is to illustrate the behavior of phase holdups and the axial heat dispersion coefficient in gas-liquid-solids fluidized bed system. The independent system parameters examined are gas velocity, slurry velocity, solid particles size and concentration. Since one of the major applications of the present work is for coal conversion processes, only different sizes of coal particles have been investigated. Unlike other studies, this work was concentrated on relatively small particles (3-300μ) and for a wider range of slurry velocity (0 to 10 cm/sec). The heat dispersion coefficients obtained from this work have been compared with both the mass dispersion coefficients (Deckwer, 1972; Towell and Ackerman, 1972; Kato et al., 1972) and the theoretical correlations obtained by Joshi (1980) and Baird and Rice (1975).

Experimental Setup and Procedure

The schematic diagram of the experimental setup is shown in Figure 1. The experimental study was carried out in an eleven-feet-high and six-inch stainless steel column having three major sections:

1) conical bottom section for gas inlet and distribution,
2) one foot high "calming section" filled with 1"x1/2" copper, rings for mixing of gas and slurry,
3) main test section below the heater, and
4) heat source at the top.

Two distributor plates, one between the test section and calming section, had sixty 10 mm diameter holes; the other between the calming section and the conical bottom had 239 holes of 1 mm size. These were used to ensure uniform flow at the bottom of the test section.

Although six pressure taps at intervals of 2 feed were provided, only the two bottom most and the two top most pressure taps, connected to two different manometers were normally used. To overcome the difficulty of plugging these manometer lines, a water backflushing system was used.

In addition to the pressure taps, six sampling tubes of 1/4" in diameter were inserted to collect solid-liquid samples. Each of these "L" shaped sampling tubes had five 1/8" holes drilled on one leg and was installed horizontally. This arrangement provided radially averaged phase concentration at each location. In case of plugging, these tubes could be backflushed. The solids concentration in the sampled slurry was measured gravimetrically.

For heat tracer experiments, eleven iron constantan thermocouples, located one foot apart from each other along the length of the column were used. All thermocouples were connected to a single digital readout device (with ± 0.1°C accuracy). A shell and tube heat exchanger was located in the outlet line of the column to cool the heated slurry to the desired inlet temperature of slurry. This was employed during the steady state heat tracer experiments. A steam-heated coil was used as a heat tracer. Steam was introduced at the top through copper coils, at a temperature of 120°C. Gas and slurry flow rates were measured by rotameter and ultrasonic device (Hersey Products, Inc.), respectively. All the experiments were carried out at atmospheric pressure, with cocurrent upflow of gas and slurry, in a continuous steady manner.

The ranges of system parameters studied are given in Table 1. The settling properties of the particles as a function of the particle size, and the slurry properties, such as density, viscosity etc. are summarized in Tables 2 and 3, respectively. The viscosities of the coal-water slurry were calculated using the relationship given by Larosa (1979).

Analysis of Raw Data

A) Holdups

The manometer reading measurements were converted to absolute pressures by a simple hydrostatic head technqiue. It was required that the holdup of gas, liquid and solids, expressed as the fractional volumes, should add up to unity. A second equation was obtained from the pressure drop measurements. Overall pressure drop, measured between the two points, consists of two terms: the hydrostatic head and the frictional pressure drop term. It was found that frictional pressure drop was negligible and pressures measured were linear with respect to the axial distance. This proves that the most important term which contributed to the overall pressure drop was the hydrostatic head term that includes the holdups and densities of the phases. The final form that was used is given by the following equation for the gas holdup:

$$\varepsilon_G = \frac{HH_{VG=0} - HH}{HH_{VG=0}} \qquad (1)$$

To determine solid and liquid holdups, solid-liquid samples were collected at several locations along the length of the column at various gas and liquid velocities. By measuring the volume and weight of each sample, the slurry density was obtained. After filtering and drying the samples, it was possible to calculate the fractional volume and weight of solid and/or liquid within the slurry itself. From this information, individual liquid and solid holdups were calculated by using the equations 3 and 4.

$$\varepsilon_{SL} = 1 - \varepsilon_G \qquad (2)$$

Sampling data provided:

$$\frac{\varepsilon_S}{\varepsilon_{SL}} = \frac{\varepsilon_S}{\varepsilon_L + \varepsilon_S} = \frac{\text{Volume of Coal}}{\text{Volume of Coal + Volume of Water}}$$

and

$$\varepsilon_S = \varepsilon_{SL} \left(\frac{\varepsilon_S}{\varepsilon_{SL}}\right) \qquad (3)$$

Therefore, $$\varepsilon_L = 1 - \varepsilon_G - \varepsilon_S \qquad (4)$$

B) Axial Mixing and Dispersion Coefficients

The measured temperature distributions were correlated by the axial dispersion model based on the following assumptions:

1) the contribution due to the latent heat of vaporization was negligible. (This assumption was justified by calculation under experimental conditions.)

2) the thermal conductivity and the density of the gas phase were small compared to those of slurry and was therefore neglected.

3) the molecular thermal conductivity of slurry was negligible compared to the axial dispersion coefficient of heat.

4) there was no radial temperature gradient over the entire length of the column. (This assumption was checked by measuring the temperatures at different radial positions for different sets of parameters.)

5) wall heat losses were negligible. (Column was insulated with one-inch thick fiberglass insulation.)

6) steady state flow.

7) no chemical reaction is occurring.

8) solid-liquid mixture behaves as homogeneous slurry.

9) gas phase backmixing was relatively small.

Based on these assumptions, the following Equation (5) was obtained:

$$- A \frac{d^2T}{dZ^2} + B \frac{dT}{dZ} = 0 \tag{5}$$

where $A = D_{SL} \rho_{SL} Cp_{SL} \varepsilon_{SL}$

$B = \rho_{SL} V_{SL} Cp_{SL}$

Solution of this differential equation with the general boundary conditions of the form,

$$\begin{aligned} @\ Z &= Z_C \quad T = T_C \\ @\ Z &= Z_H \quad T = T_H \end{aligned} \tag{6}$$

resulted in the following relationship:

$$\frac{T-T_C}{T_H-T_C} = \frac{e^{\frac{V_{SL}}{D_{SL}\varepsilon_{SL}}(Z-Z_C)} - 1}{e^{\frac{V_{SL}}{D_{SL}\varepsilon_{SL}}(Z_H-Z_C)} - 1} \tag{7}$$

To eliminate the end effects due to the existence of the steam coil at the top and the distributor plate at the bottom of the test section, the temperatures at a distance of 23.5 and 267.3 cm from the distributor plate were taken as the boundary conditions (Equation (6)).

Some typical data for the measured and calculated temperatures as a function of the axial distance from the distributor plate are illustrated in Figure 2.

The calculated values of dispersion coefficients were found to be independent of the amount of heat input to the system.

Results and Discussion

A. Flow Regimes

Determination of the flow regime in a multiphase mixture is one of the most important steps in evaluating the characteristics of the flow, because each flow regime provides different phase behavior and mixing characteristics. In the present study, a flow regime chart was prepared similar to the one given by Darton and Harrsion (1975), using the results of the measured gas holdup and the drift flux theory. The chart is based on the slip velocity which is defined as,

$$U_S = \frac{V_G}{\varepsilon_G} - \frac{V_L}{\varepsilon_L} \qquad (8)$$

and the drift flux of gas, ν_{CD}, (Wallis, 1962, 1969) which is defined as the volumetric flux of gas relative to a surface moving at the average velocity, i.e.

$$\nu_{CD} = U_S \varepsilon_g (1 - \varepsilon_g) \qquad (9)$$

The drift flux plotted as a function of gas holdup appears to fall into two regions: one in which the gas holdup increases rapidly with the drift flux, and the other, in which its increase is markedly slower. These correspond to the "uniform bubbling" and "churn turbulent" regimes, respectively.

It can be seen from Figures 5-7 that most of our data fall in the transition region. For the data that fall in the uniform bubbling regime, rise velocity of bubble swarms was equal to the rise velocity of single bubble in an infinite medium, which was 180 mm/sec. In this region there was no effect of liquid flow rate, solids concentration or particle size, on either drift flux velocities or gas holdup. However, deviation from the uniform bubbling regime occurs at smaller gas holdup values (or lower superficial gas velocities) as either the particle size, solid concentration or liquid velocity increases.

The point of transition from one regime to the other was hypothesized to depend chiefly upon the physical properties of the gas and liquid, the particle size and the fluid flow-rates as was observed in the present work, since these factors principally determine the onset of coalescence.

A relationship between the drift flux and the gas holdup of the following general form has been suggested for both bubble and slug-flow regimes

$$\nu_{CD} = \varepsilon_g U_{BO} (1 - \varepsilon_g)^m \qquad (10)$$

Different values of m ($0 \leq m \leq 2$) have been reported by different investigators (Nicklin et al., 1962; Turner, 1966; Richardson and Zaki, 1954) for two-phase systems. The value of m was considered to be an indication of how the rise of a particular bubble is affected by its neighbors as well as surrounding fluid properties. Darton and Harrison (1975) indicated that the knowledge of how the drift flux-gas holdup plot changes with solids concentration, particle size and slurry velocity was needed.

In the present study m was correlated to slurry and particle Reynolds numbers, and liquid and solid holdup according to the relation

$$m = - (1.95 \times 10^{-4}\ Re_{SL} + 16.13\ \frac{\varepsilon_S}{\varepsilon_S + \varepsilon_L} + 0.97\ Re_p + 3.34) \tag{11}$$

As shown in Figures 5 to 7, the predictions of Equations (10) and (11) agreed quite well with the present experimental data.

B. Holdups

1. Gas Holdup

Within the range of system variables examined in this work, gas holdups were found to depend on gas and liquid velocities, solid concentration and solid particle size. Some typical results illustrating the effect of superficial gas and liquid velocities on the gas holdup is shown in Figure 6. At low gas velocities (5-10 cm/sec), the bubbles rise separately in the liquid, i.e. the homogeneous bubble flow regime prevails, and the effect of gas velocity on the gas holdup is quite pronounced. Similar results have been reported for two-phase (i.e. gas-liquid) systems (Lockett and Kirkpatrick, 1975). At high gas velocities, significant coalescence of bubbles occurs, and a large fraction of the column was occupied by large bubbles. As reported previously (Nicklin, 1962) under this situation, an increase in gas velocity does not significantly increase the gas holdup. The present data for both low and high gas velocities were correlated by an empirical correlation

$$\varepsilon_g = \frac{Re_G}{A + BRe_G + CRe_{SL} + D\ (\frac{\varepsilon_S}{\varepsilon_S + \varepsilon_L})} \tag{12}$$

where the values of the four constants are given in Table 4. The typical predictions of ϵ_g from Equation (12) are shown by the solid curves in Figures 6-10. The parity plot, Figure 11, presents the comparison between all the predicted and observed values of gas holdups obtained in this work.

In the bubble flow regime, a linear relationship between gas holdup and superficial gas velocity exists. Although it is generally claimed that in a bubble column there is little effect of liquid (or slurry) velocity on the gas holdup, the present work shows a decrease in gas holdup with an increase in liquid velocity. This result is in agreement with the observations of Michelsen and Ostergaard (1970) and Kim et al. (1972) for three phase systems; and of Hills (1976) for two phase systems.

It has been found previously that liquid viscosity plays an important role in determining the behavior of fluidized beds (bubble coalescing or bubble disintegrating). In general, low viscosity solutions exhibit bubble disintegrating behavior. However, a trend towards bubble coalescing behavior is observed by increasing the viscosity of the liquid fluidizing medium. This finding is supported by the work of Calderbank et al. (1967) and Vasalos et al. (1980) who observed that bubble coalescence was very much more pronounced in beds containing liquids of higher viscosity. This probably explains the reduction in gas holdup with the increase in solids concentration shown in Figure 7. This reduction is more pronounced at lower solids concentration.

The effect of solids concentration on gas holdup is less significant at higher superficial slurry velocities. The typical effect of solids concentration on bubble diameter is illustrated in Figure 8. These bubble diameters, calculated by Nakanoh and Yoshida's (1980) method, increase with the increase in the solids concentration, and a decrease in gas velocity.

Figures 9 and 10 illustrate the effect of solids particle size on gas holdup for low (0.02 m/s) and high (0.102 m/s) values of slurry velocities, respectively. In the low liquid velocity region the data by Kato (1972) and Kato and Nishiwaki (1971) are presented and they match well with our experimental data. As can be seen, an increase in the solid particle size decreases the gas holdup.

2. Liquid Holdup

As shown in Figure 12, in the present study , liquid holdup decreased with an increase in the gas velocity and a decrease in slurry velocity. The effect of particle size on the liquid holdup was found to be more significant for the higher values of superficial gas velocities. However, Kim et al. (1972) reported a decrease in liquid holdup with the increase in both gas and liquid velocities. They also indicated that the presence of solids reduces the liquid holdup in three phase beds. For a fixed particle size and, slurry and gas velocity, liquid holdup in this work was found to be higher for lower solid concentration as shown by broken lines in Figure 12.

3. Solid Holdup

The solid holdup behaved in a way similar to liquid holdup. Some typical results for the solid holdup are shown in Figure 13. Although no explicit information on solid holdups were readily available from their work, solid concentration profile in the bubble column slurry reactors has been investigated by Cova (1966), Imafuku et al. (1968), Farkas and Leblond (1969), and Narayanan et al. (1969). Cova (1966) studied the effects of liquid velocity, gas velocity and liquid viscosity on the solid concentration profile and concluded that higher gas and liqui velocities and high viscosities all tend to give more uniform solid distributions. Imafuku et al. (1968), reported that the critical gas velocity for complete suspension of solid particles was affected by liquid flow near the gas distributor, this liquid flow being governed by the shape of the bottom of the column and the position of the gas distributor, solid particle size, liquid-phase surface tension and solid-liquid wettability on the critical solid holdup, which is defined as the maximum quantity of suspended solid particles in liquid within the column. In all these studies, density difference between the solid and liquid was higher and the ranges of liquid and gas velocities were smaller than those used in the present work. High fluid velocities, small particle sizes and small density differences between the solid and liquid were also believed to be major causes for the flat solid concentration axial profiles as can be seen from Table 5.

C. Dispersion Coefficient

In the present study the measured axial dispersion coefficient was found to be dependent upon gas and slurry velocities and the particle size, and relatively independent of solids concentration.

For an air-water system, the dispersion coefficient increased with an increase in superficial gas velocity and a decrease in superficial liquid velocity. A comparison of the present data with those published in the literature is shown in Figure 14. The heat dispersion coefficients obtained in this work for low gas and liquid velocities, are somewhat higher than those observed by Deckwer (1973), Baird and Rice (1975) and Towell and Ackerman (1972). At high gas velocities, the predicted values from those correlations agreed well with the present data. Deviations at low gas velocities may be attributed to the differences in the distributor devices used in those studies, as it is generally claimed. The effect of distributor design is more pronounced in the low gas velocity region where independent gas bubbles are formed.

The effects of phase velocities on the dispersion coefficient are similar to the ones observed for two-phase system. The present data for 3, 30 and 300 micron particles at the highest (.102 m/s) and the lowest (.034 m/s) slurry velocities, under the conditions of 10% solid concentration, are shown in Figures 15 and 16, respectively. The data are compared with those predicted from the correlations of Kato et al. (1972), Baird and Rice (1975) and Joshi (1980) along with the empirical correlation of this work. Kato et al. (1972) and Baird and Rice (1975) have neither taken the effect of solid concentration nor the particle size into account in their correlations. Therefore their correlations predicted identical values of dispersion coefficients regardless of the size of the particles and solid concentration. The bubble diameter required for the calculation of circulation velocities was estimated by Calderbank's (1967) and Nakanoh's (1980) methods. Although at lower slurry velocites the present experimental results can be predicted by other correlations, significant discrepancy at higher slurry velocities emphasizes the stronger effect of slurry velocities on dispersion coefficients obtained in this work relative to the ones mentioned. Although the effect of particle size cannot be seen clearly due to the scattered nature of these data, the empirical correlation, given by Equation 13, indicates some effect of particle size and it predicts lower values of dispersion coefficient at larger particle sizes. A typical correlation for 300μ particles at the highest slurry velocity is shown in Figure 15.

All the correlations used for the comparisons are summarized in Table 6. Based on the present data, we also conclude that the axial heat dispersion coefficient matches well with the mass dispersion coefficient (Figure 16). The present data were correlated by an empirical relation of the following form.

$$D_{SL} = A\ V_G^{\ B}\ V_{SL}^{\ C} \qquad (13)$$

Constants in Equation 13 along with the asymptotic standard deviations estimated for them are listed in Table 6. A comparison of the predicted and experimental results is presented in Figure 18 for all the dispersion data obtained in this work.

CONCLUSION

The fluid dynamics of three-phase cocurrent flow in a 15.2 cm diameter by 335 cm long column were studied. Gas, liquid and solid holdups in gas/liquid/solid systems were investigated both in the presence and absence of liquid flow. The variables studied were: particle size (3-300 microns), solids concentration (0-40%), gas velocity (3-30 cm/sec), and liquid velocity (0-10 cm/sec). It was found that gas holdups increased with gas velocity, but decreased with liquid velcoity, solids content, and particle size. Axial heat dispersion coefficients were also determined in the same system. The heat dispersion model represented the observed axial temperature profiles. The heat dispersion coefficients were found to increase with gas velcoity, to be independent of solids concentration. The heat dispersion coefficients were found to decrease with both increasing slurry velocity and increasing size of the solids over the ranges studied. The dispersion coefficients for both heat and mass were judged to be equivalent. Flow regime maps were also determined by relating drift flux and gas holdup. Most of the present data fell in the turbulent-bubbling or transition region. The effects of solids concentration, particle size, and slurry velocity on the prevailing flow regimes were found to be significant. An increase in values of all these parameters caused an earlier deviation from bubble flow regime.

The experimental data for flow regimes, gas holdup and dispersion coefficient were empirically correlated and are given in Equations 10, 12 and 13, respectively.

TABLE 1

Ranges of System Parameters

System Parameter	Range
Superficial Gas Velocity (V_{SG})	3.0 - 30.0 cm/sec
Superficial Liquid Velocity (V_{SL})	0.0 - 10.0 cm/sec
Solid Concentration	0 - 40% by weight
Average Particle Size	3 - 300μ
Type of Solids	Dried Mineral Residue from SRC-I Plant (Fort Lewis) (3μ)
	West Virginia and Pittsburgh Seam Coal (30μ)
	Kentucky coal (300μ)
Density of Solid Particles	Approximately 1.3 (g/cc)(for all types of solids used in this work)

TABLE 2

Settling Characteristics of the Coal-Water Slurry

Particle Size	Slurry Concentration (% wt)	Terminal Settling Velocity (cm/sec)	Hindered Settling Velocity (cm/sec)	Settling Regime
300	.105 .240 .360	1.254	.766 .380 .168	Intermediate K = 4.88 > 3.3
30	.40 .31 .25 .14	0.018	.0002 .0015 .0048	Stokes K = .488
3	.110 .215	0.0018	.000068 .000023	Stokes K = .0488

TABLE 3

Physical Properties of Coal-Water Slurry

Temperature = 25°C

Particle Size (μ)	Conc. wt%	$\frac{\epsilon_S}{\epsilon_S+\epsilon_L}$	ρ_{SL} (g/cc)	μ_{SL}* (CP)
300	10.5	.085	1.025	1.265
	24	.195	1.060	2.379
	36	.305	1.093	5.966
30	14	.115	1.034	1.480
	25	.215	1.044	2.720
	31	.250	1.077	3.498
	40	.340	1.107	7.547
3	11	.085	1.075	1.265
	21.5	.175	1.052	2.095
Air-Water	0	0	1.000	1.000

---

*Calculated from the relationship of Larosa (1979).

TABLE 4

Constants in Gas Holdup Equation 12 for Different Particle Sizes Along with the Asymptotic Standard Deviations of Parameter Estimates*

Particle Size (μ)	A	B	C	D
300	667.96 ± 36	4.75 ± .14	.408 ± .054	16644.05 ± 1663
30	1337.50 ± 1468	3.84 ± .038	.602 ± .018	9247.19 ± 5294
3	1315.92 ± 438	3.27 ± .011	.130 ± .029	8730.84 ± 3396
Air-Water	1521.23 ± 145	3.75 ± .10	.136 ± .023	0

---

*Dimensionless units.

TABLE 5

Solids Concentration Profiles (weight percent) for 300 Micron Particles

Average Solid Conc. (%wt)	V_{SL} (cm/sec)	V_G (cm/sec)	Distance from Distributor (cm)				
			23.5	145.4	206.4	267.3	Exit
10.5	0.00	2.1	.117	---	.107	---	--
		10.7	.120	---	.104	---	--
		23.8	.117	---	.112	---	--
	3.40	2.1	.110	---	.107	---	.106
		10.7	.105	---	.103	---	.096
		23.8	.107	---	.109	---	.109
	10.16	2.1	.100	---	.103	---	.111
		10.7	.106	---	.109	---	.099
		23.8	.105	---	.103	---	.103
24	0.00	2.1	.245	.246	---	---	---
		10.7	.253	.250	---	---	---
		23.8	.244	.244	---	---	---
	3.40	2.1	.222	.255	---	.243	.235
		10.7	.237	.245	---	.241	.230
		23.8	.256	.247	---	.239	.242
36	0.00	2.1	---	---	.367	---	---
		10.7	---	---	.364	---	---
		23.8	---	---	.359	---	---
	3.40	2.1	.370	---	.378	---	.355
		10.7	---	---	.362	---	---
		23.8	.368	---	.359	---	.357

TABLE 6

Constants for Heat Dispersion Coefficient Equation 13

dp(μ)	A	B	C
Air-Water	240.051 ± 34.27	.205 ± .032	- .235 ± .0620
3	179.591 ± 22.60	.336 ± .034	- .126 ± .049
30	285.320 ± 40.70	.262 ± .035	- .450 ± .060
300	165.933 ± 15.35	.249 ± .030	- .465 ± .033

TABLE 7

PROPOSED MODELS FOR LIQUID PHASE DISPERSION COEFFICIENTS

Reference	Model		
Deckwer et al. (1974)	$D_{SL} = 0.68\ D^{1.4}\ V_G^{0.33}$	D : m V_G : m/s D_{SL} : m²/s	Gas-Liquid
Baird and Rice (1975)	$D_{SL} = 0.35\ D^{1.33}\ (g\ (V_G \pm \frac{\varepsilon_G}{1-\varepsilon_G} V_L))^{1/3}$	D : m V_G, V_G : m/s g : m/s² D_{SL} : m²/s	Gas-Liquid
Joshi (1980)	$D_{SL} = 0.29\ D\ (V_C+V_L)$ where $V_C=1.31\ (gT\left[V_G+V_L - \frac{\rho_L V_L}{\varepsilon_L\rho_L+\varepsilon_S\rho_S} - \varepsilon_S\ (\frac{\rho_S}{\varepsilon_S\rho_S+\varepsilon_L\rho_L} - 1)\ V_S-\varepsilon_G V_{b\infty}\right])^{1/3}$ V_S = hindered settling velocity; obtainable from McCabe and Smith (1976) V_b = terminal rise velocity of bubbles; can be obtained by Grace et al. (1976) $d_{B\infty}$ = bubble diameter; can be obtained by (a) Nakanoh and Yoshida (1980) (b) Calderbank (1967)		Two and Three Phase
Kato et al. (1972)	$Pe = \frac{13\ Fr}{(1+8\ Fr^{.85})}$ where $Pe = \frac{V_G D}{D_{SL}}$ and $Fr = \frac{V_G}{\sqrt{gD}}$		Three Phase

TABLE 7
(Continued)

Towell and Ackermann (1972)	$D_{SL} = 73.5\ D^{1.5}\ V_G^{.5}$	D : ft V_G : ft/hr D_{SL} : ft^2/hr	Gas Liquid
Pilhofer et al. (1978)	$D_{SL} = (0.15 + 69\ V_G^{.77})\ (D^{1.25}\ \frac{0.001}{\mu_{SL}})$	D : m/s μ_{SL} : $\frac{kg}{m \cdot sec}$	Gas Liquid

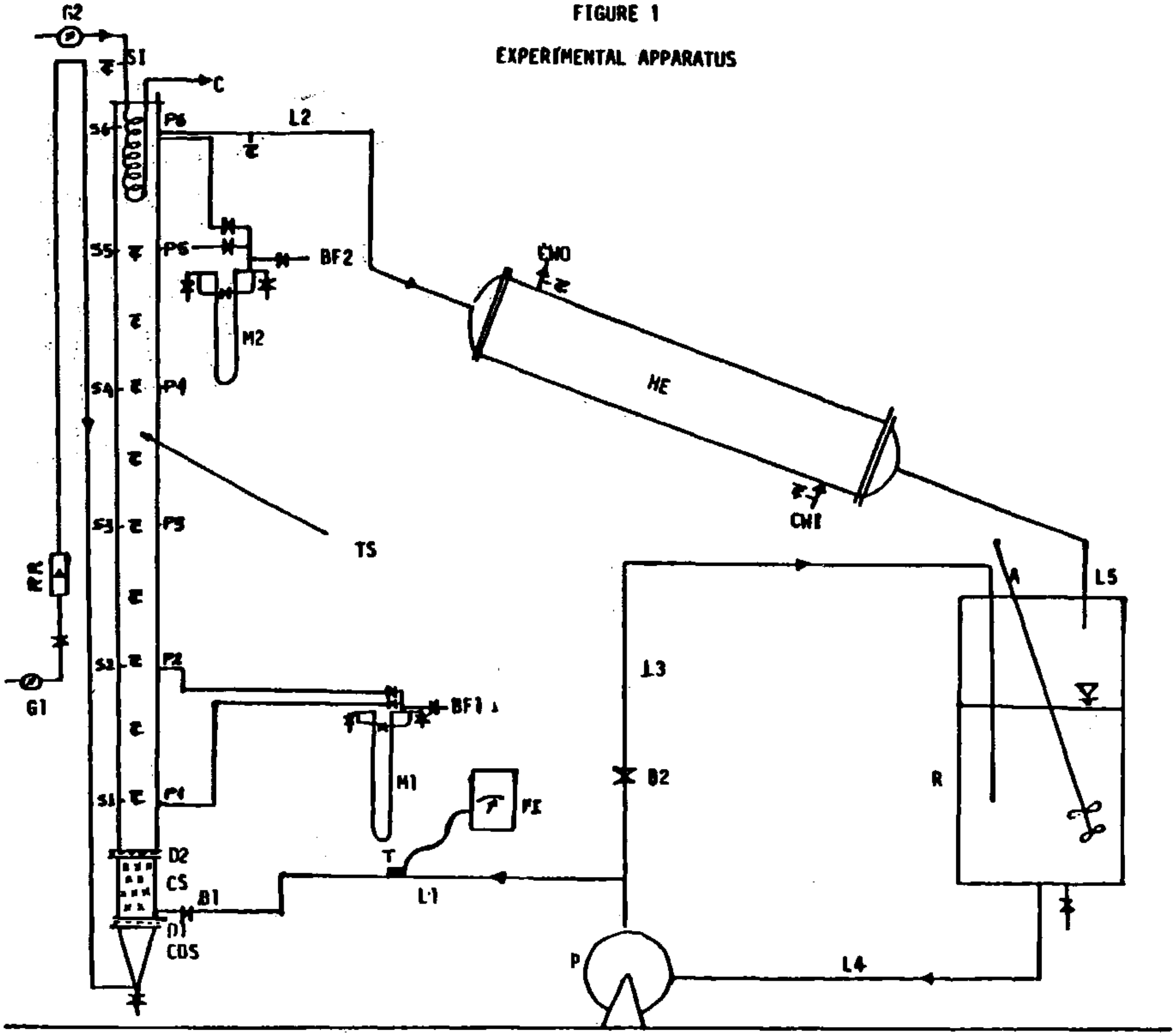
Note: 1) Baird and Rice: suggested not to be used for (a) high V_{SL}; (b) for large particle diameters, large particle settling velocities and high solid holdups

2) Pilhofer: indicates that increasing liquid viscosity decreases the dispersion coefficient for the liquid phase

3) Bubble diameters calculated by Nakanoh and Yoshida have two-fold higher values than the ones calculated by Calderbank's formula; also Nakanoh and Yoshida's equation predicts an increase in bubble diameter as concentration of slurry and liquid velocity increases whereas such a tendency cannot be obtained by using Calderbank's equation due to the lack of experimental data on surface tension.

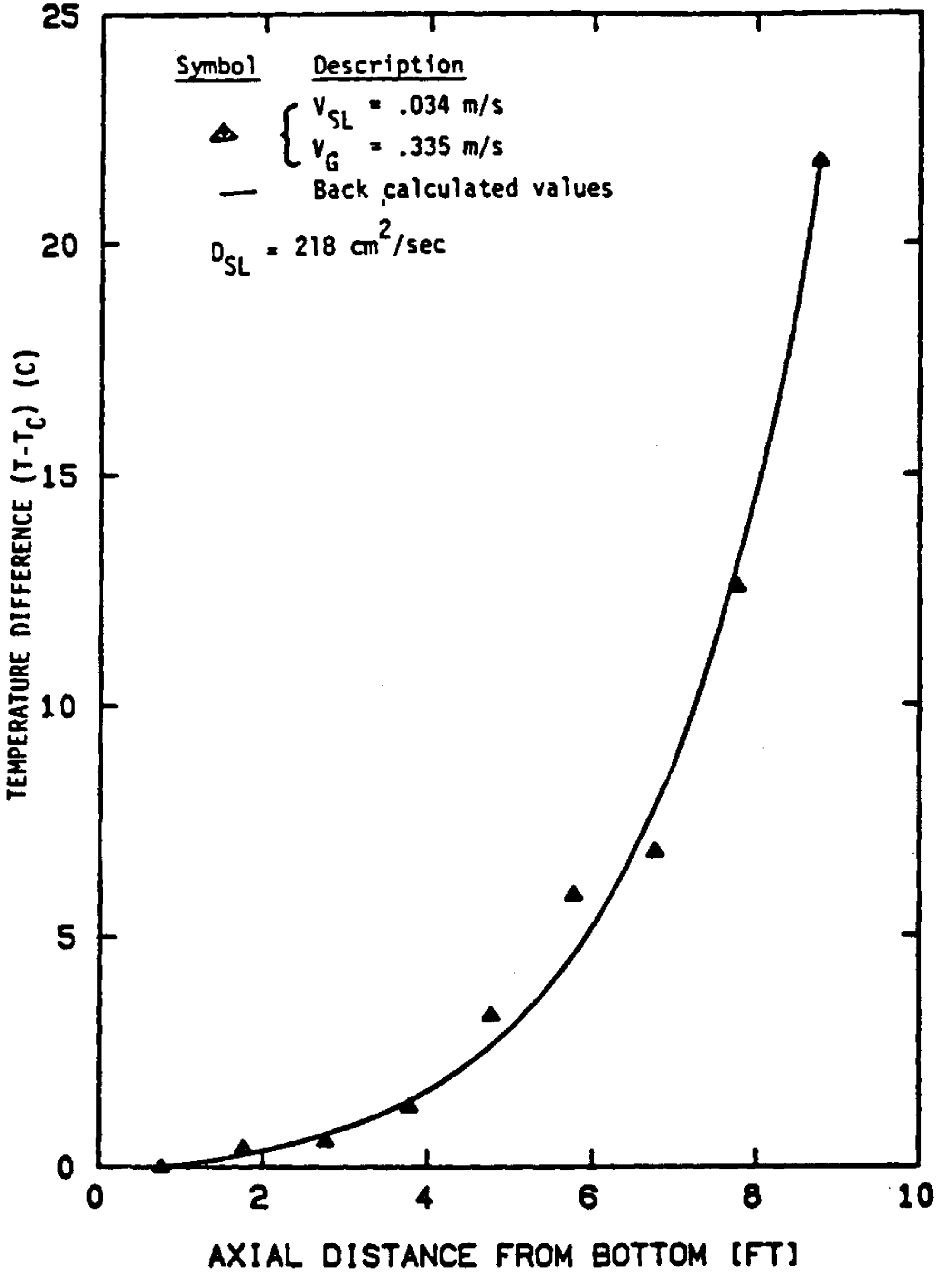
FIGURE 1

EXPERIMENTAL APPARATUS



DEFINITIONS FOR FIGURE 1

A	Agitator
B1, B2	Ball Valves
BF1, BF2	Backflushing System
C	Condensate
CBS	Conical Bottom Section
CS	Calming Section
CWI, CWO	Cooling Water Inlet and Outlet
D1, D2	Distributor Plates
G1, G2	Pressure Gauges
HE	Shell and Tube Heat Exchanger 1" OD SS tubes on the tube (slurry side) Heat transfer area 25 ft^2 14" OD, 6 ft long Black iron shell side (cooling water)
L1-L5	2" OD, Schedule 80, PVC pipes
M1, M2	Manometers
<u>P</u>	Galigher Horizontal Centrifugal Pump
P1-P6	Pressure Taps
R	Reservoir (550 liter, plastic)
RR	Gas Rotameter
SI	Steam Inlet
S1-S6	Sampling Taps
TS	Main Test Section
T, FI	Transducer and Flow Indicator (ultrasonic)
₵	Thermocouple Holes



TYPICAL TEMPRATURE DISTRIBUTION [30 MICRONS, 40% conc.]

FIGURE 2

FIGURE 3

Effect of Particle Size on Flow Regime Transition

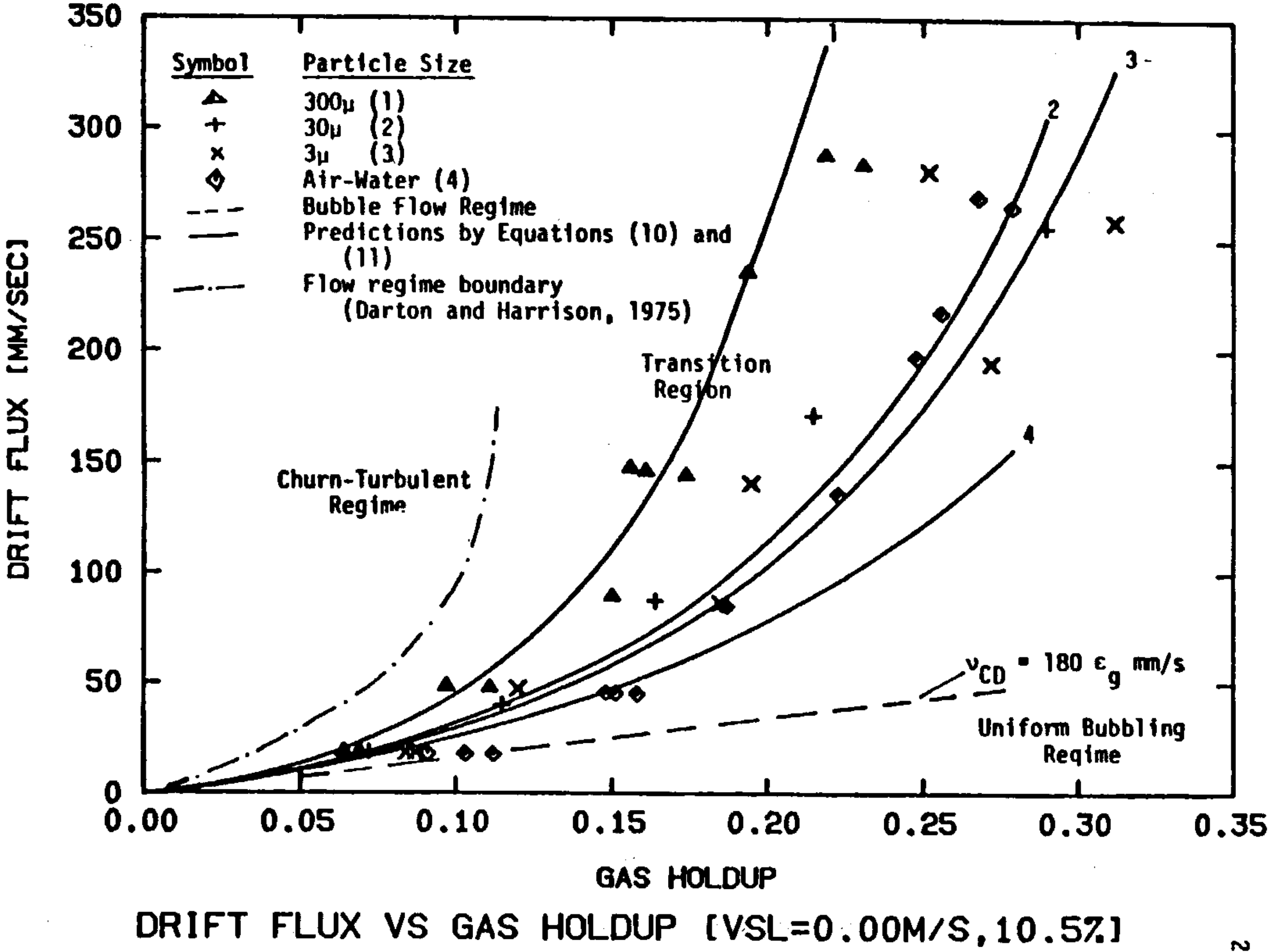


FIGURE 4

Effect of Concentration on Flow Regime Transition

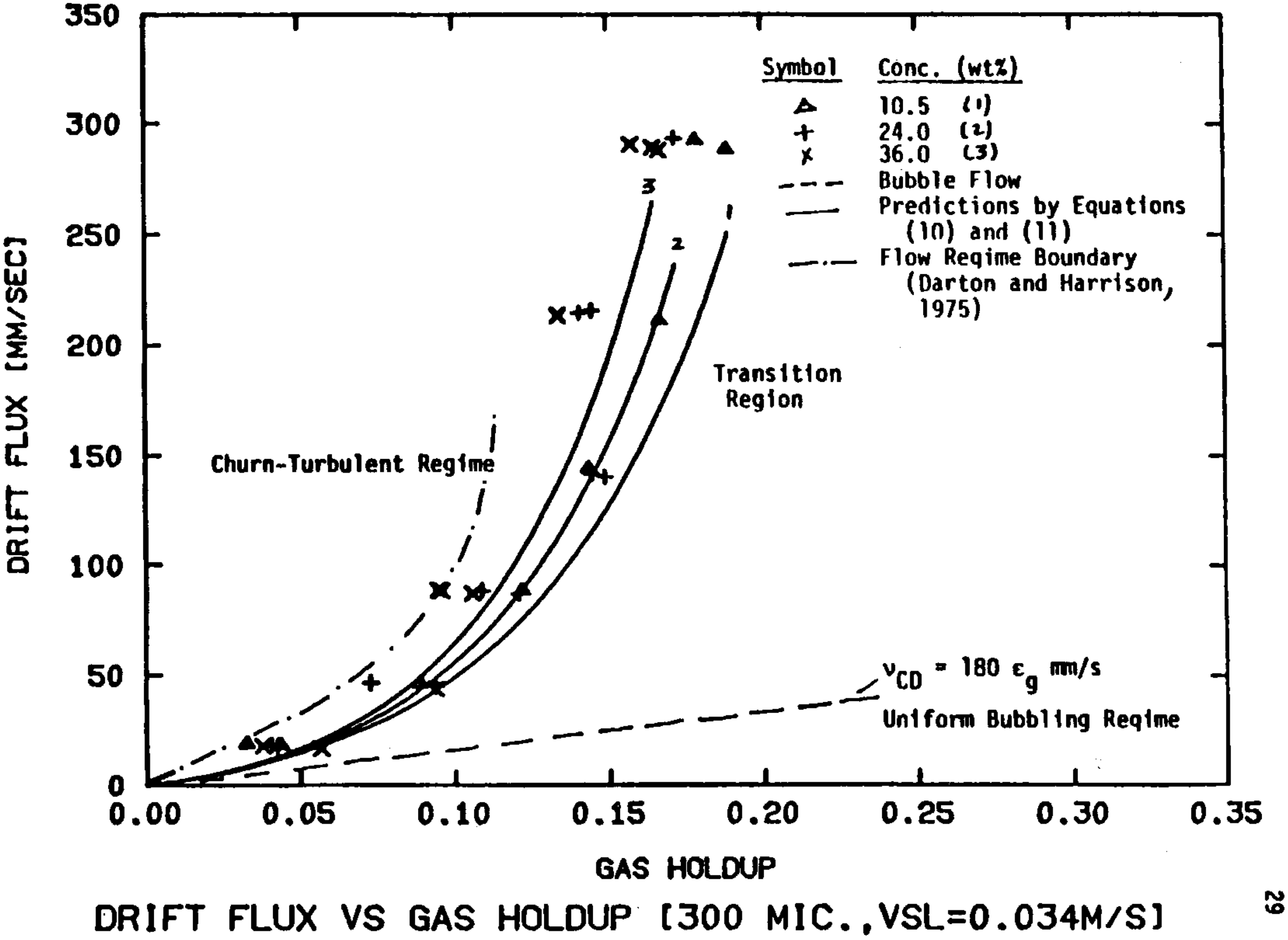
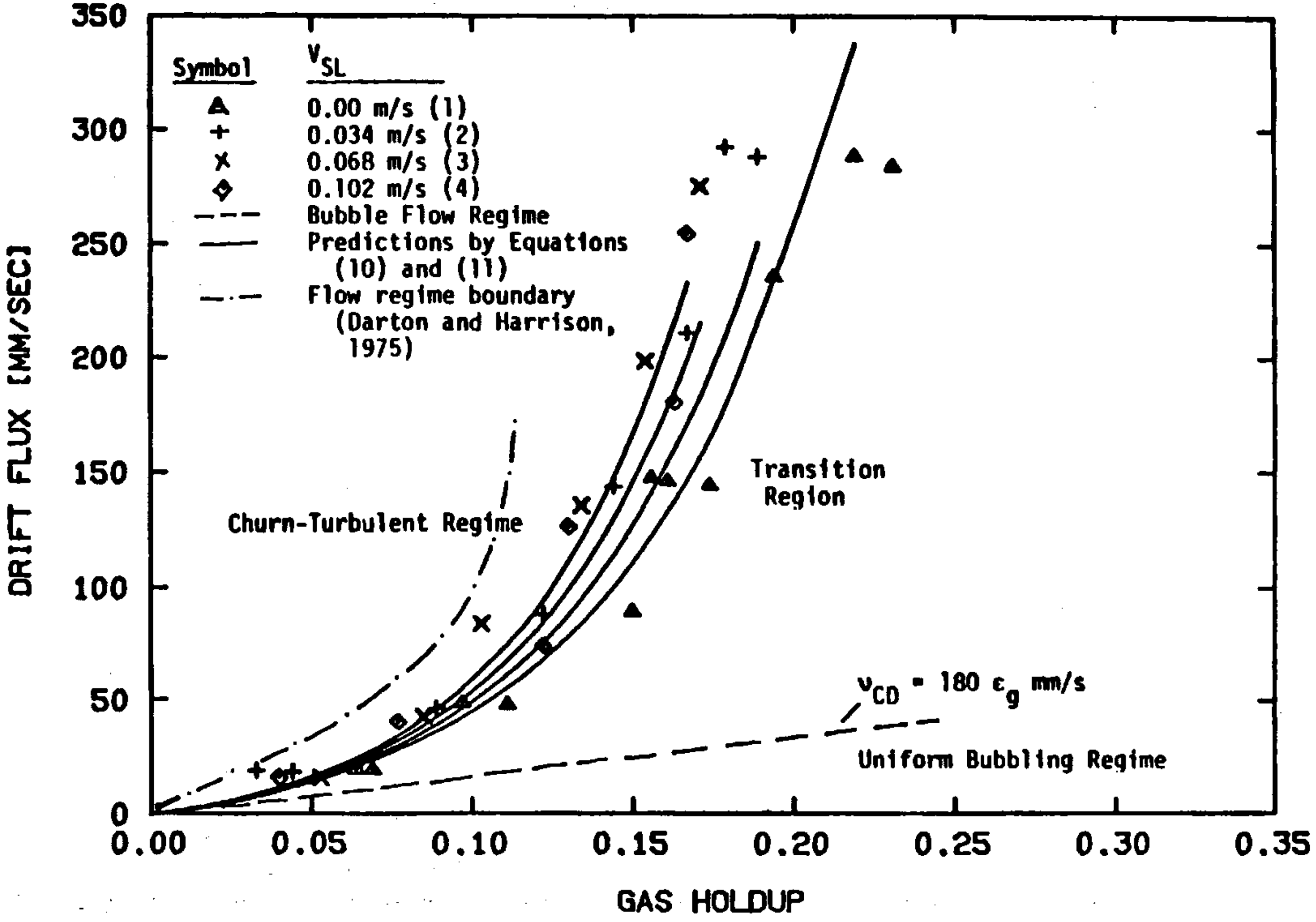


FIGURE 5

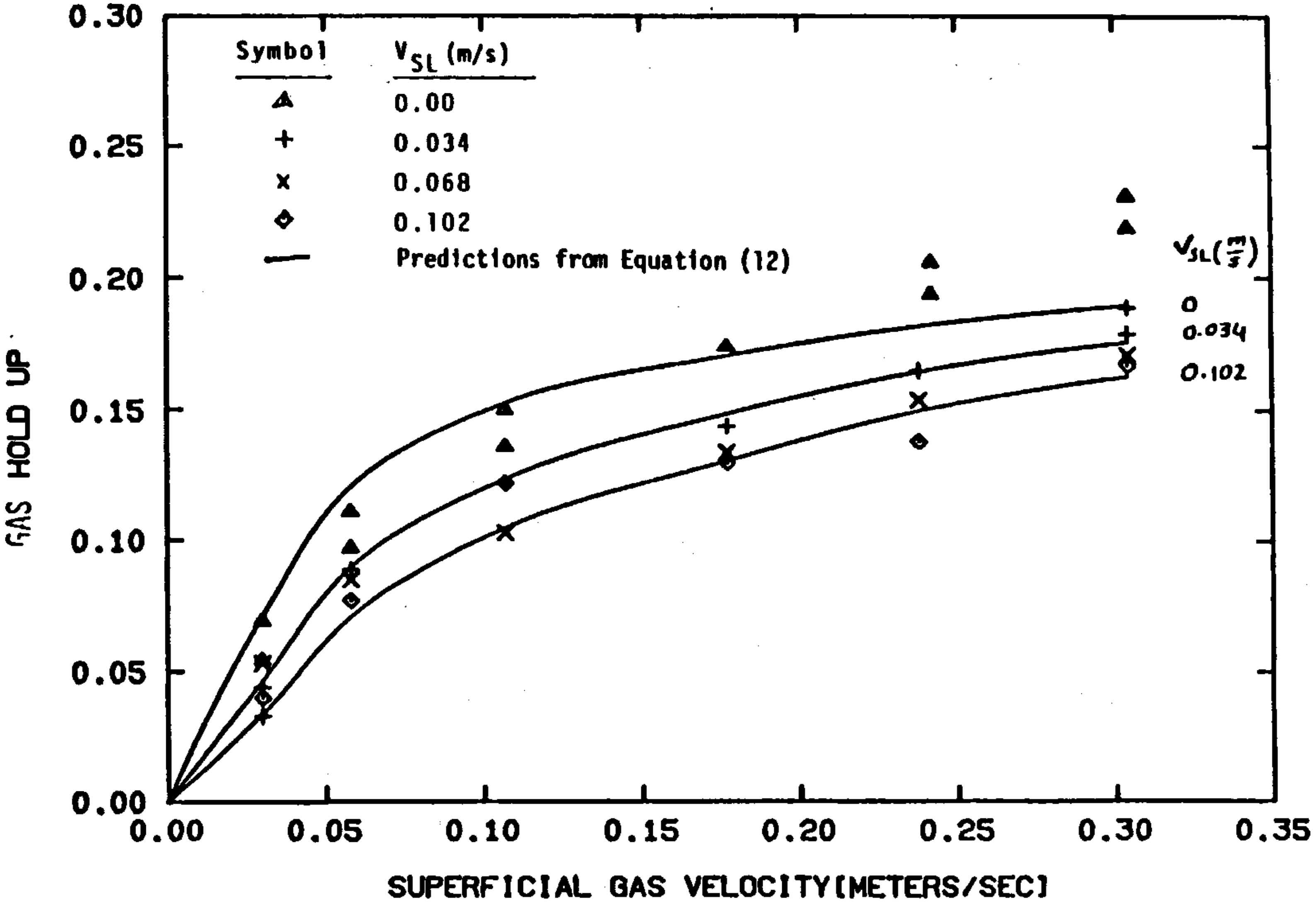
Effect of Slurry Velocity on Flow Regime Transitions



DRIFT FLUX VS GAS HOLDUP [300 MIC.,10.5%]

FIGURE 6

Effect of Gas and Liquid Velocities on Gas Holdup



HOLD UP VS SUPERFICIAL GAS VELOCITY [300 MICRON,10.5%]

FIGURE 7

Effect of Concentration on Gas Holdup

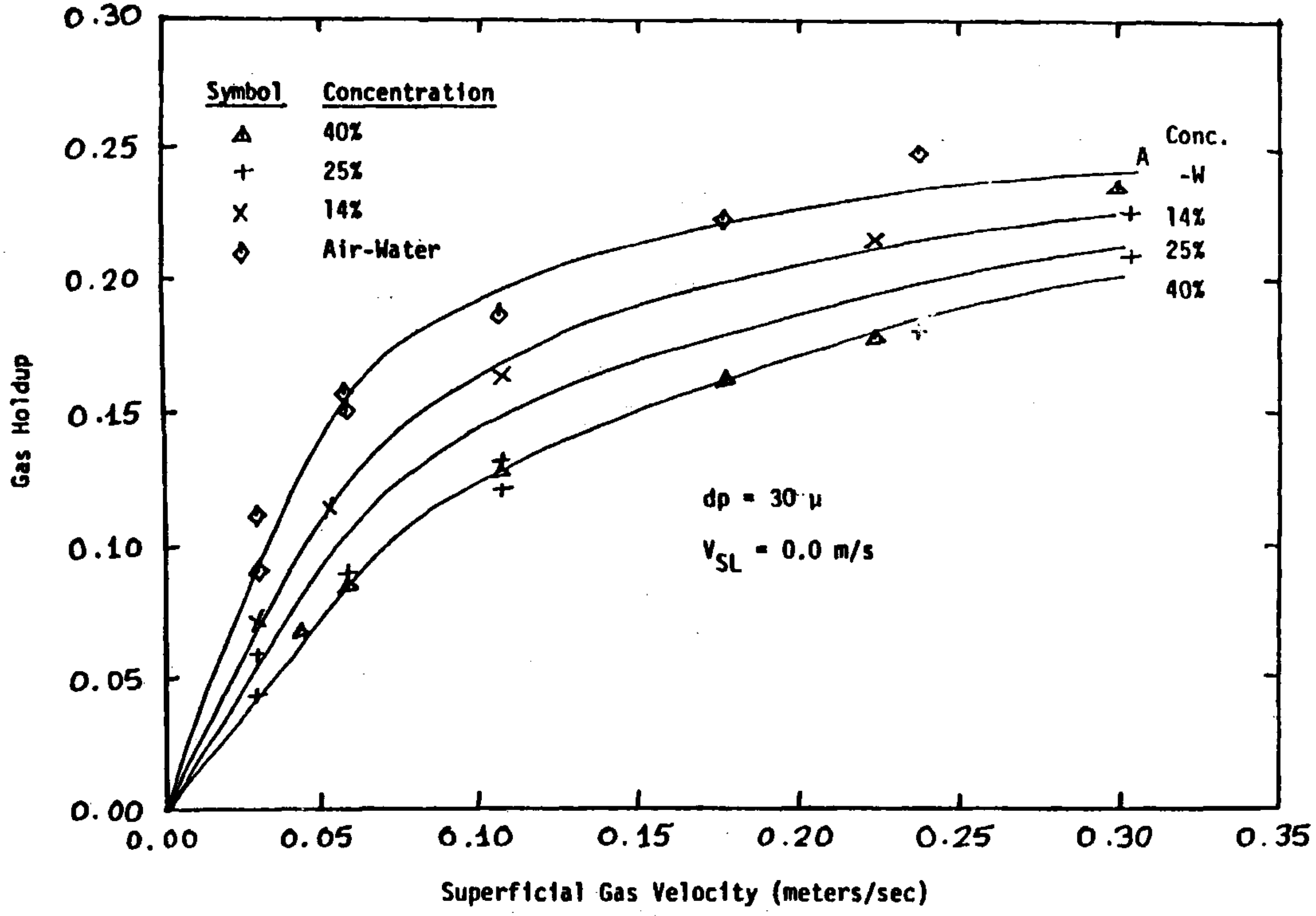
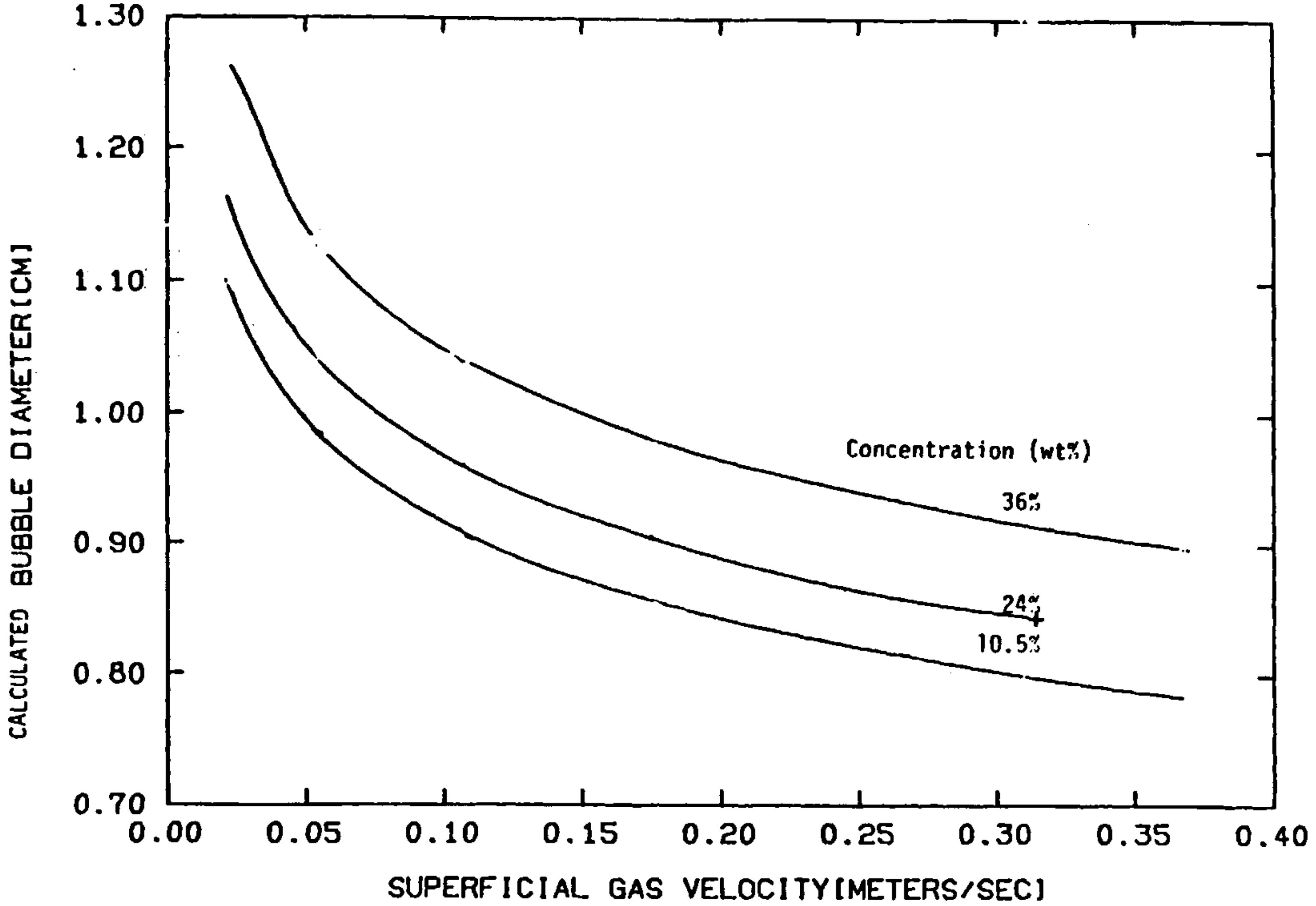


FIGURE 8

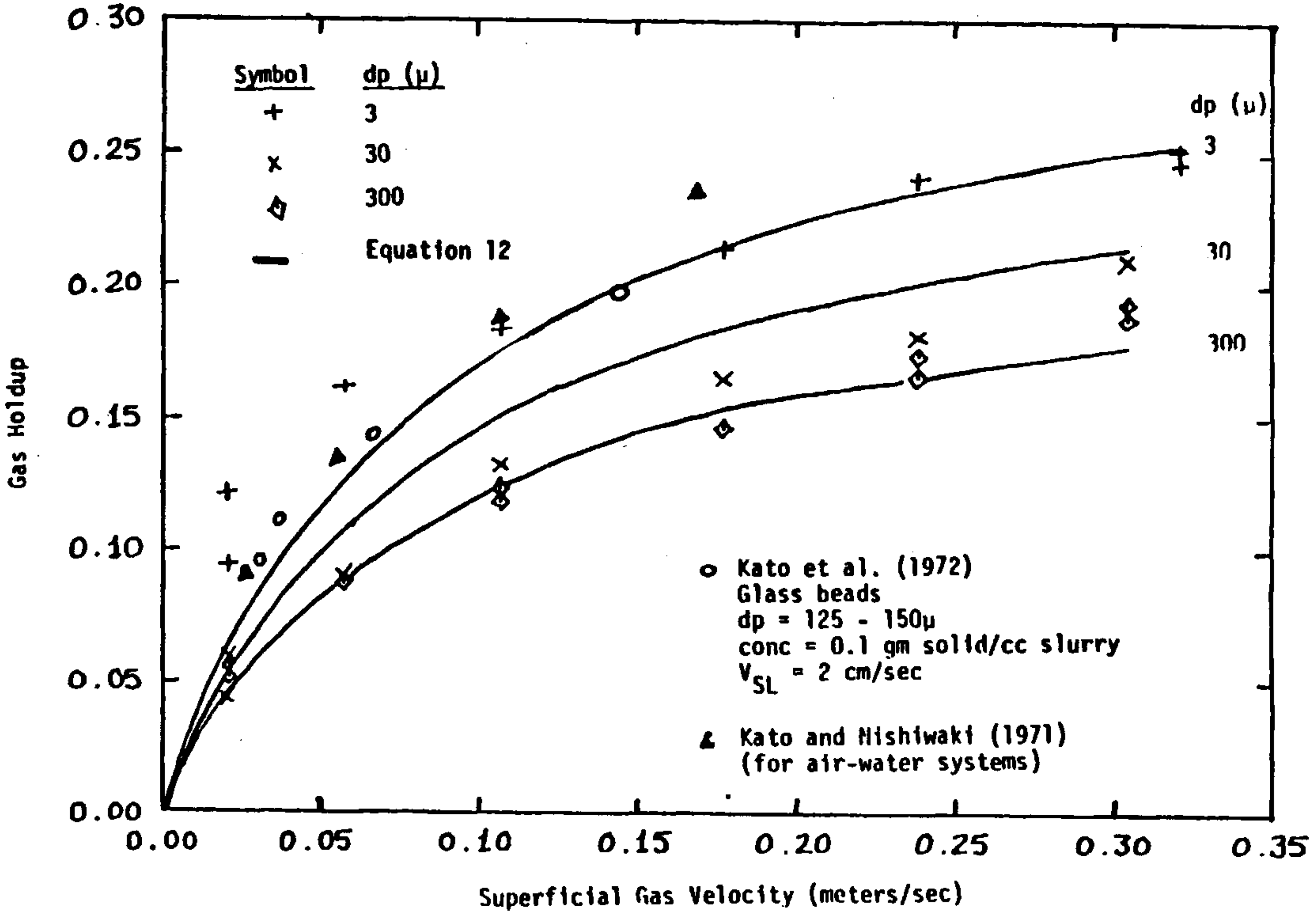
Effect of Solids Concentration on Bubble Size



BUBBLE DIAMETER VS SUPERFICIAL GAS VELOCITY[300 MIC.,VSL=0.034M/S]

FIGURE 9

Effect of Particle Size on Gas Holdup at Low Liquid Flow Rates



Gas Holdup vs Superficial Gas Velocity (V_{SL} = 0.00 m/s, 21%)

FIGURE 10

Effect of Particle Size on Gas Holdup at High Liquid Flow Rates

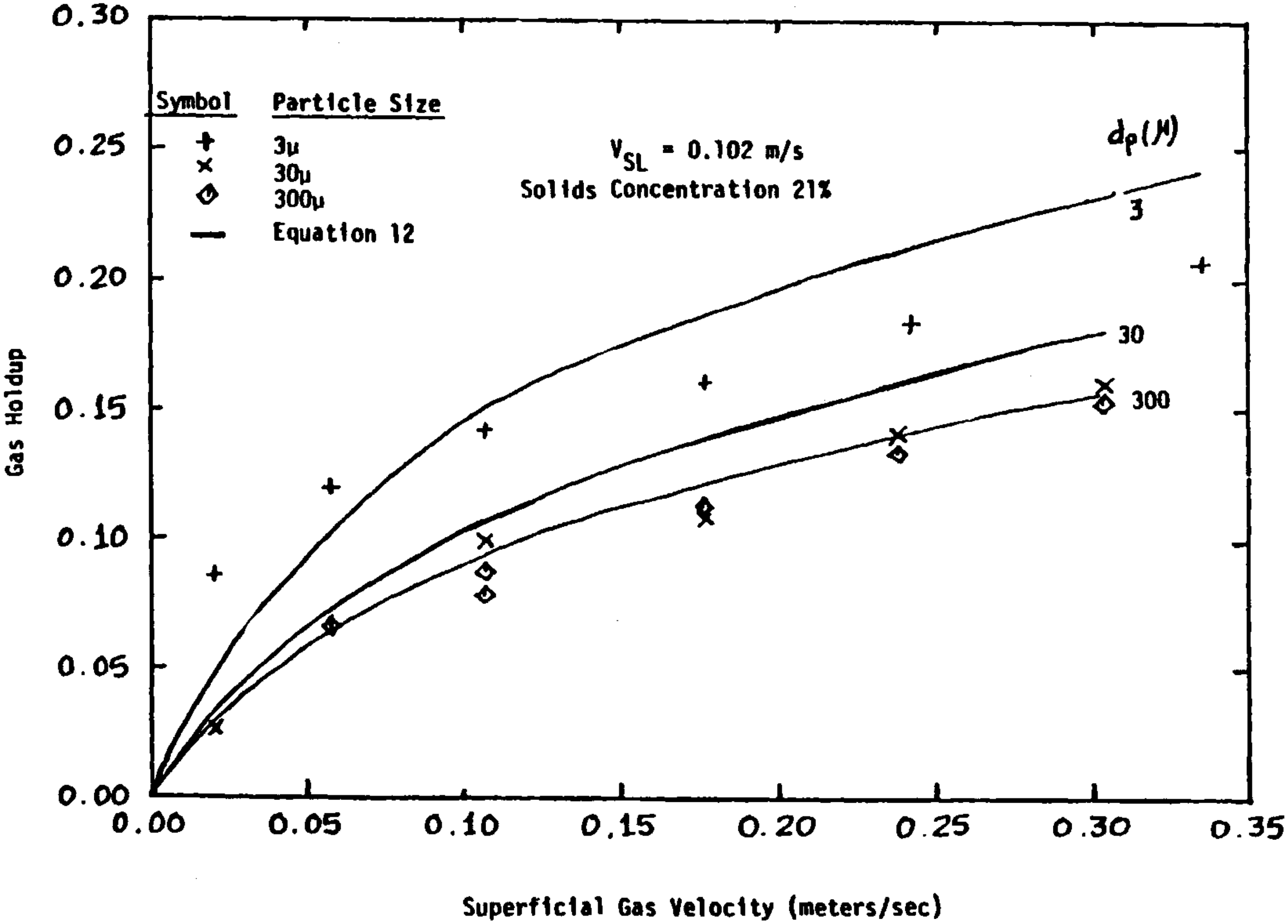
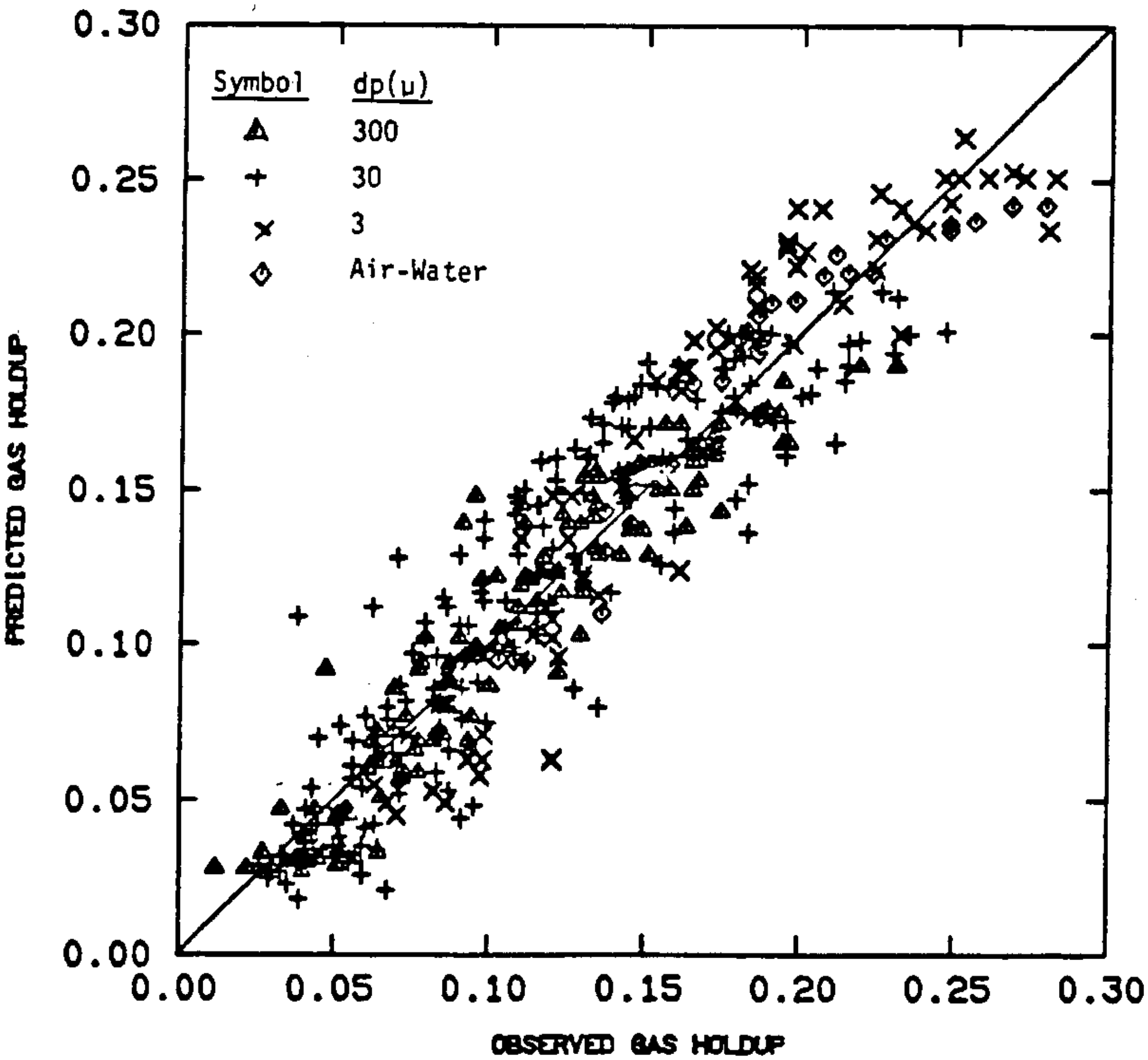
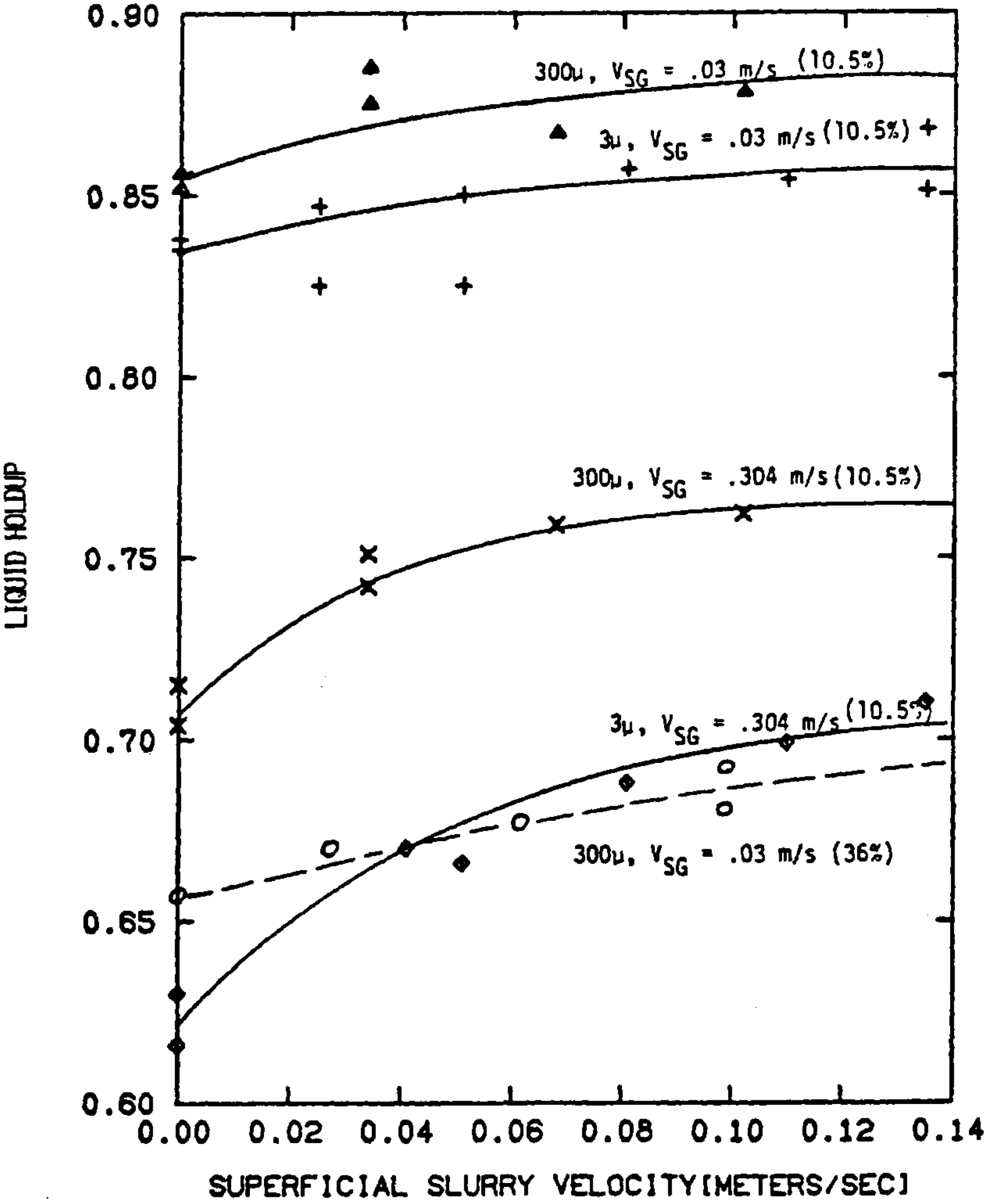


FIGURE 11

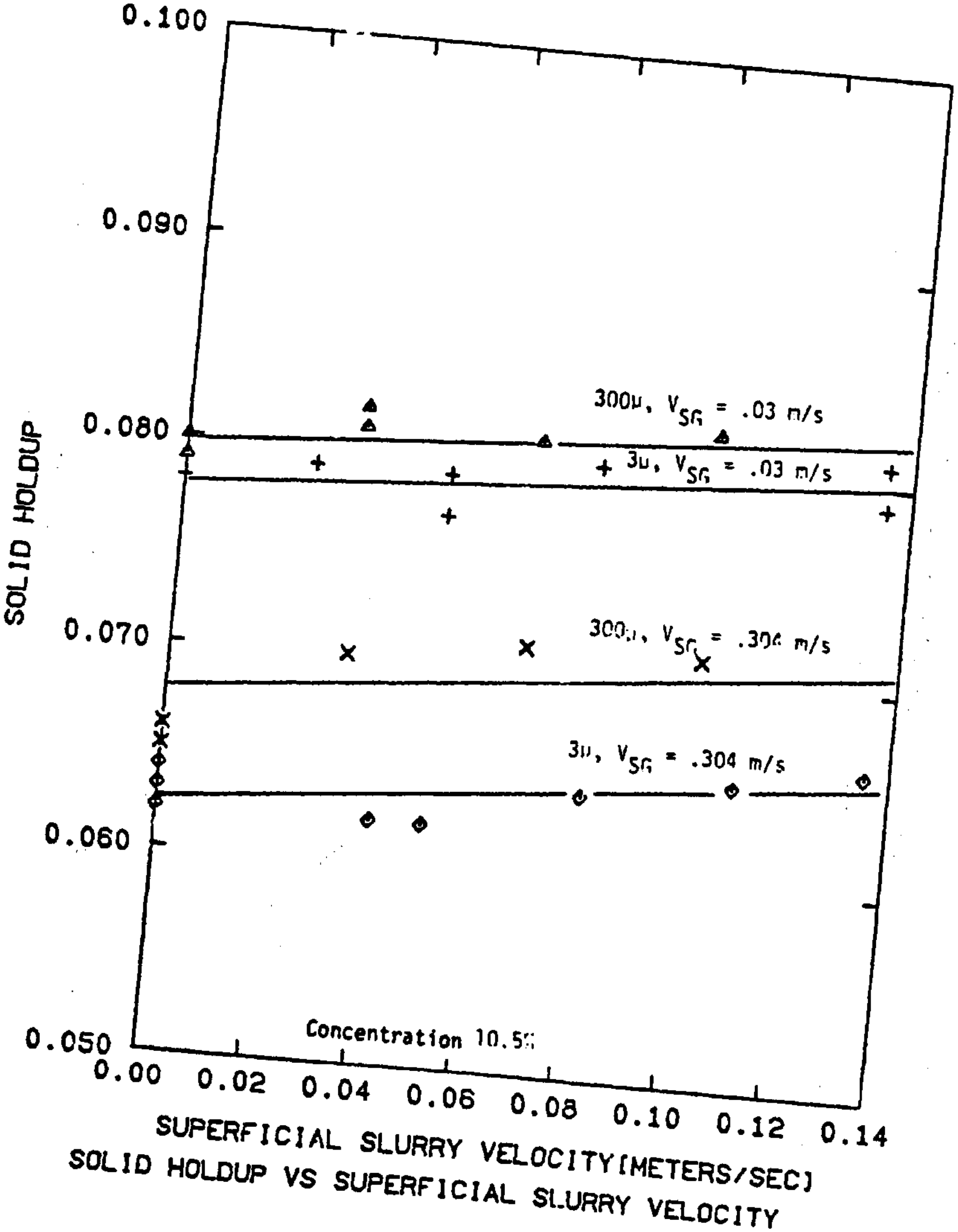


COMPARISON OF PREDICTED AND OBSERVED GAS HOLDUPS.



LIQUID HOLDUP VS SUPERFICIAL SLURRY VELOCITY

FIGURE 13



SOLID HOLDUP VS SUPERFICIAL SLURRY VELOCITY

FIGURE 14

Comparison of Air-Water Data with the Literature Values

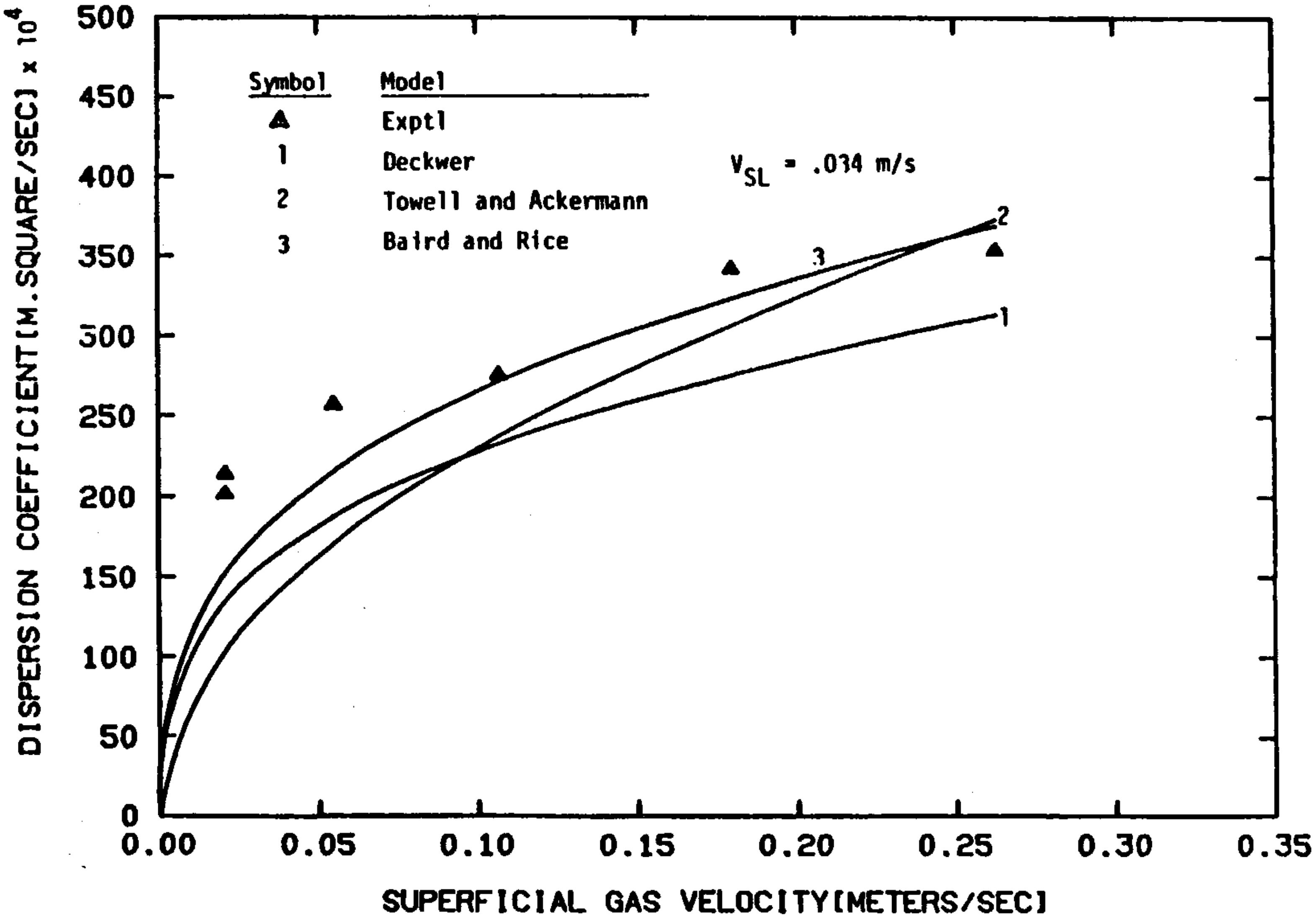


FIGURE 15

Comparison with Literature Correlations at High Slurry Flow Rates

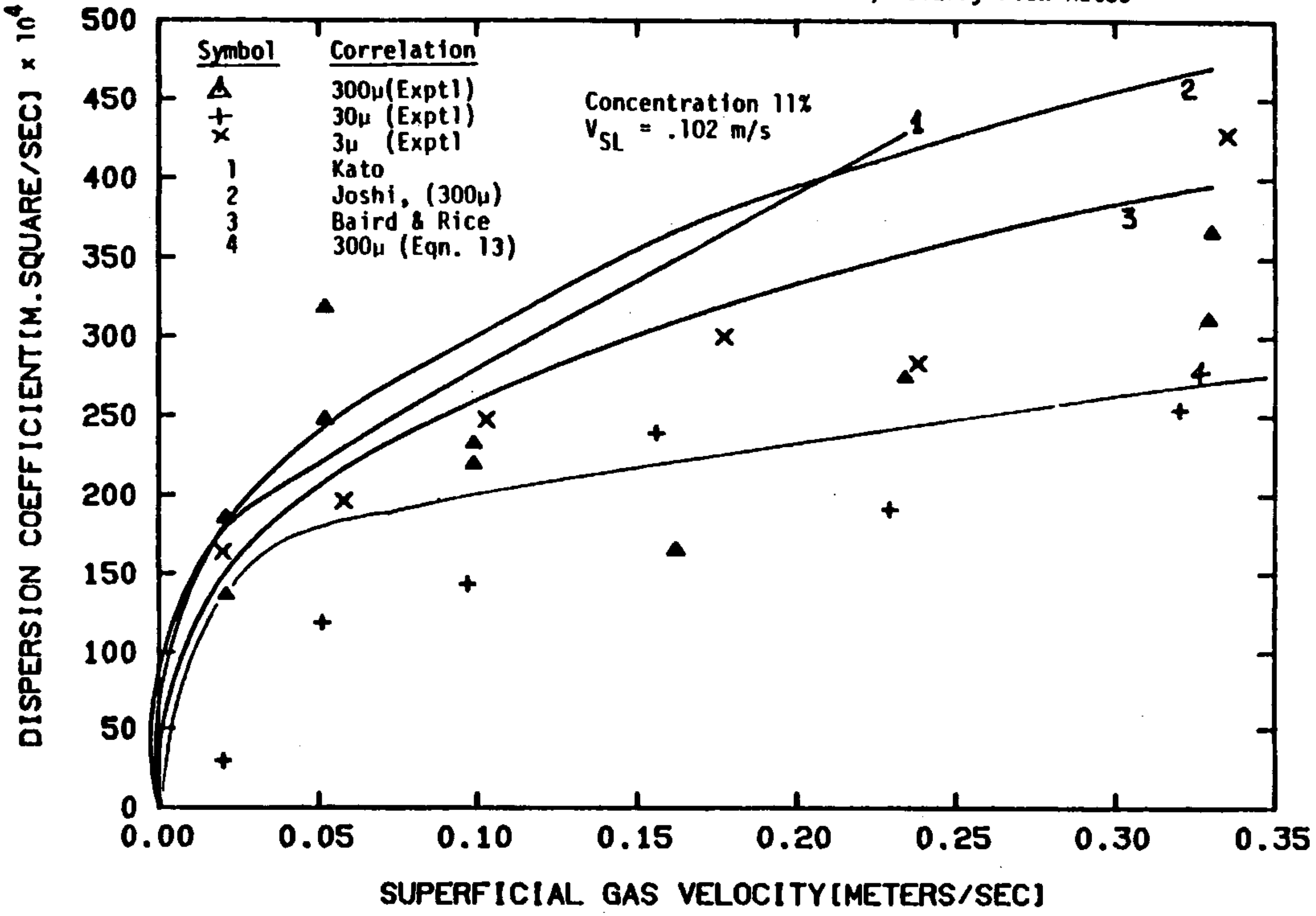


FIGURE 16

Comparison with Literature Correlations at Low Slurry Flow Rates

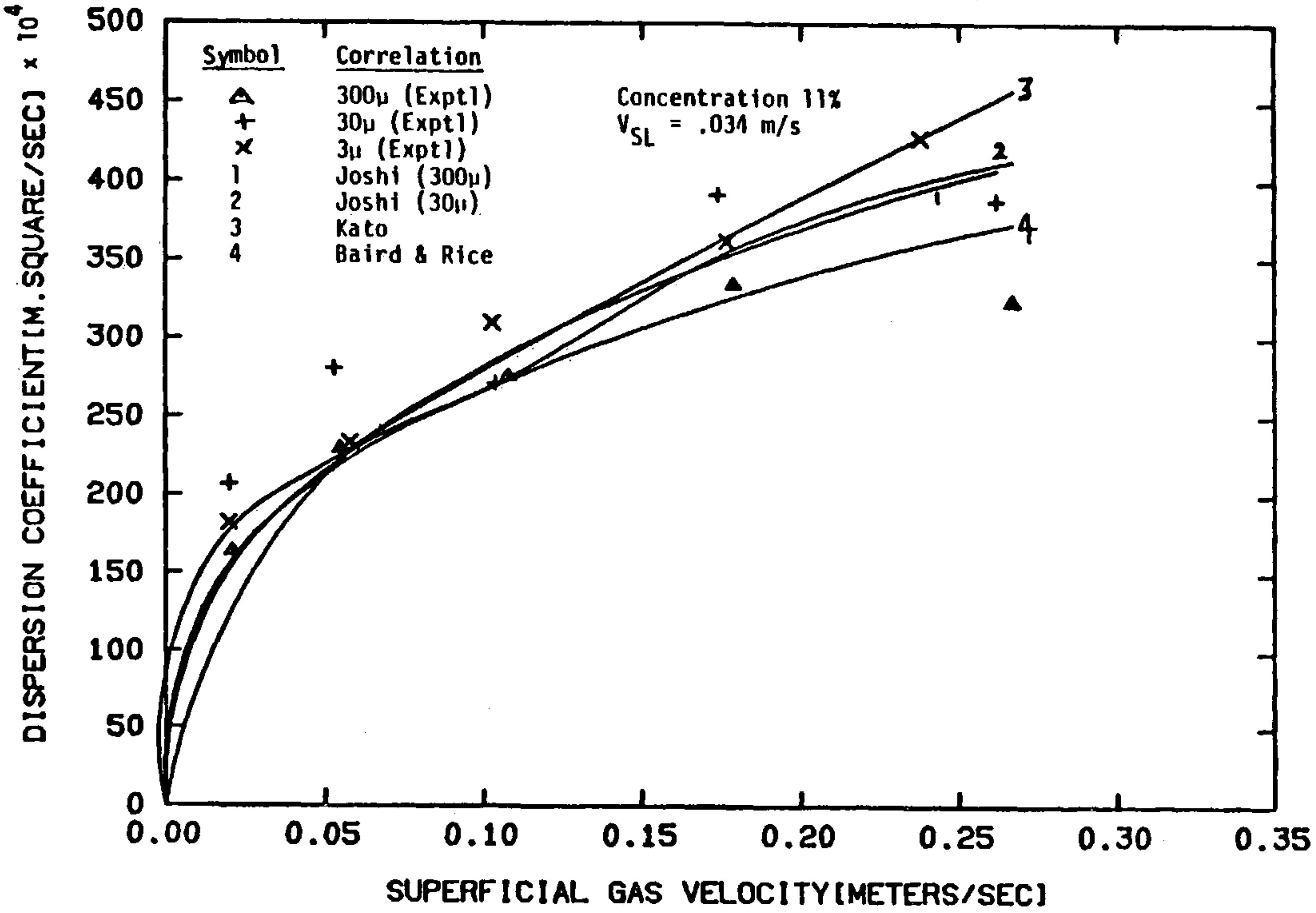


FIGURE 17

Effect of Concentration on Dispersion Coefficient

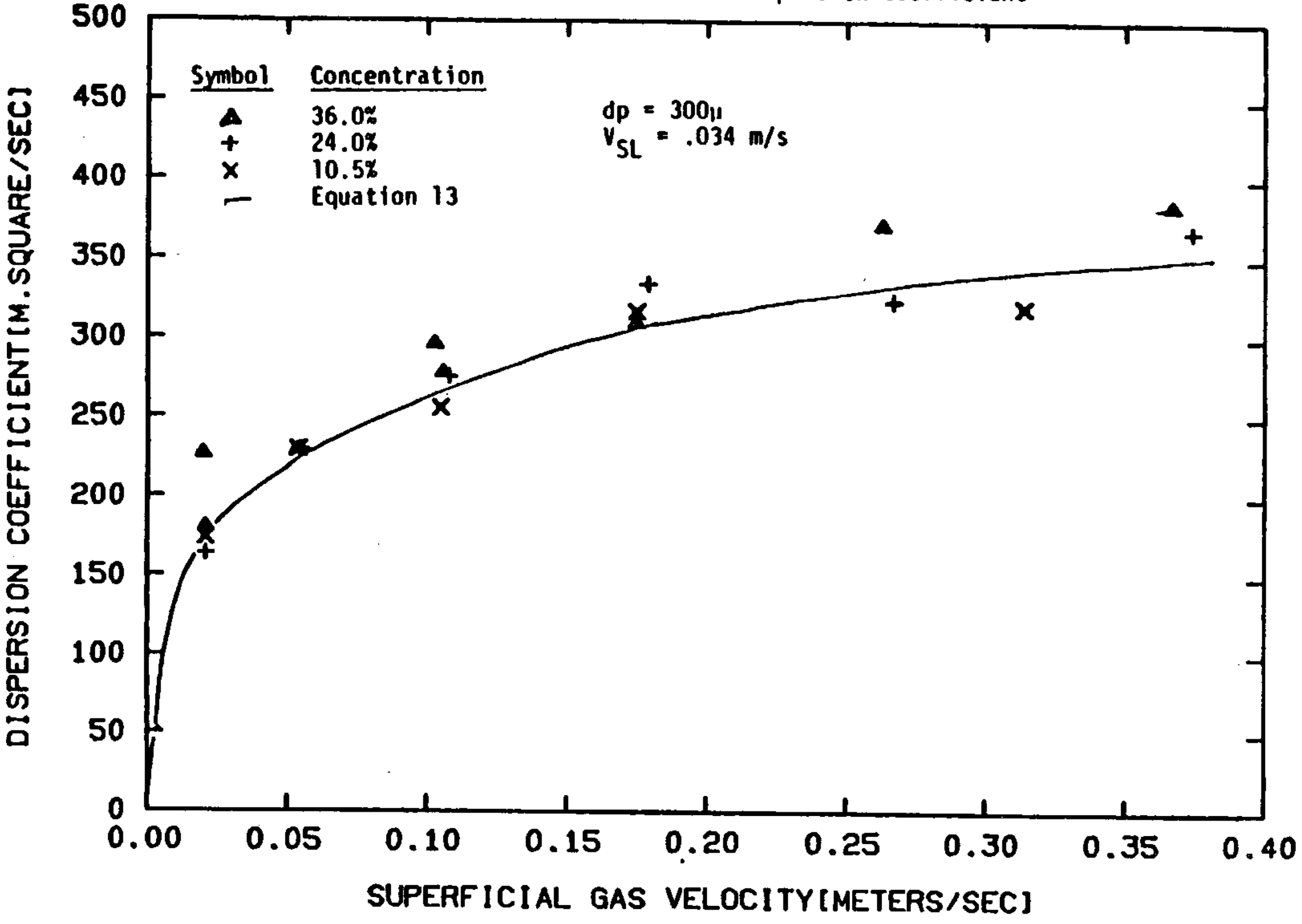
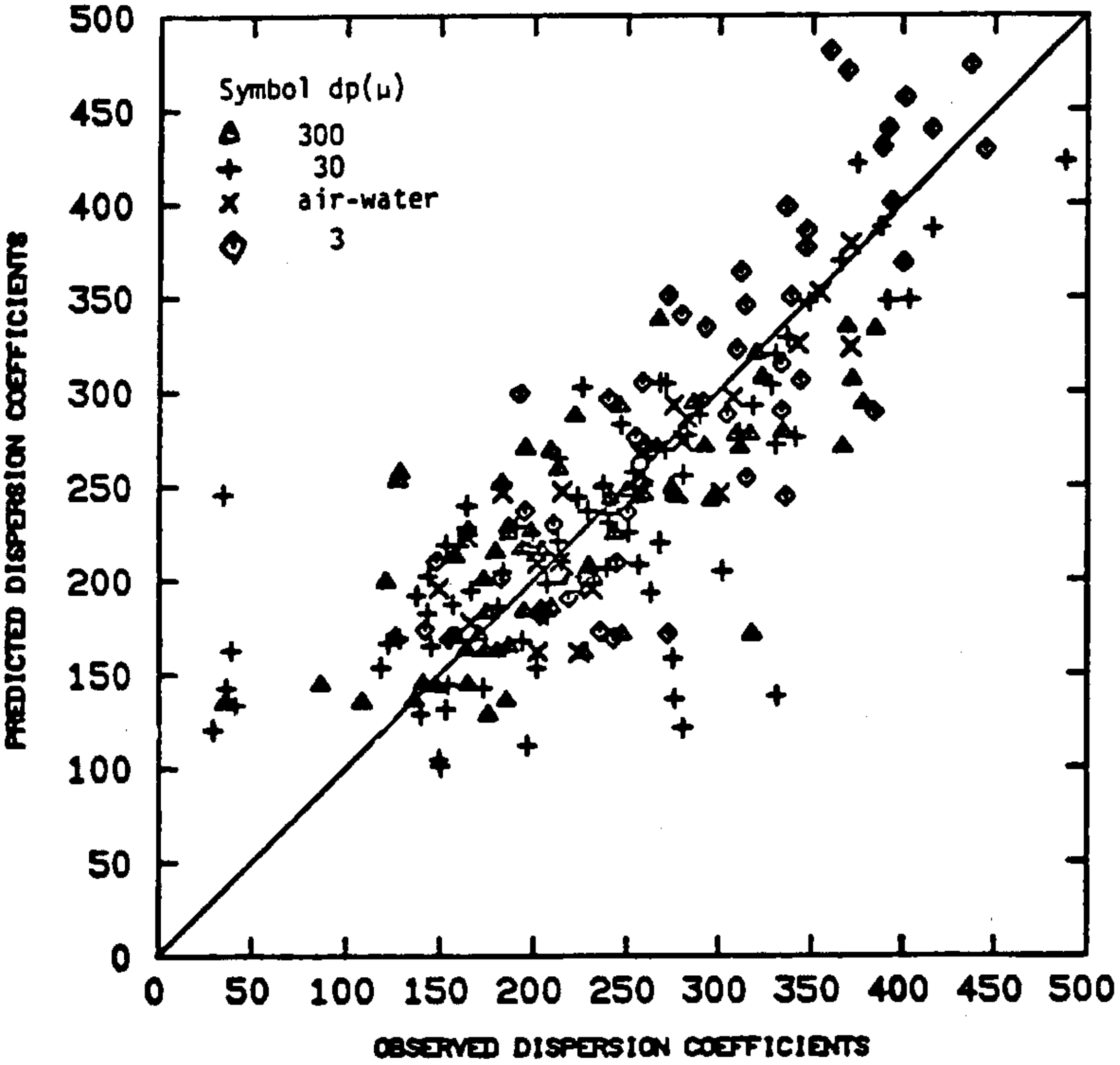


FIGURE 18



COMPARISON OF PREDICTED AND OBSERVED DISPERSION COEFFICIENTS.

NOMENCLATURE

CP_{SL}	specific heat of slurry, cal/g°C
dp	particle size; microns
D	column diameter, cm
D_{SL}	dispersion coefficient of slurry or liquid phase, cm^2/sec
g	gravitational acceleration, cm/sec^2
HH	hydrostatic head; com of water
$HH_{VG = 0}$	hydrostatic head in the absence of gas; cm of water
m	empricial constant defined by equation 11
Re_G, Re_{SL}	Reynolds number of gas and slurry phases respectively
Re_p	Reynolds number of particles based on the free settling velocity
T	temperature, °C
T_C, T_H	temperatures at the boundaries of cold side and hot side of the test section respectively, °C
U_{Bo}	terminal rise velocity of a single bubble in infinite medium; cm/sec
U_S	slip velocity; cm/sec
V_C	circulation velocity; cm/sec
V_G, V_L, V_{SL}	superficial velocities of gas, liquid and slurry phase respectively; cm/sec
Z	axial distance from the distributor plate; cm
Z_C, Z_H	distances of boundaries (cold side and hot side) from the distributor plate; cm

Greek Letters

$\epsilon_G, \epsilon_L, \epsilon_S, \epsilon_{SL}$	holdup of gas, liquid, solid and slurry phase, respectively
ρ_{SL}	density of slurry phase; g/cm^3

ν_{CD}	drift flux velocity; cm/sec
μ_{SL}	viscosity of slurry or liquid phase; $\frac{gm}{cm \cdot sec}$

Acknowledgment

The authors would like to thank Gulf Research and Development Company for the financial support to aid this research. We also express our thanks to Dr. A. A. Bishop and Dr. J. B. Joshi for helpful discussions and comments. Authors are grateful to Mr. Larry Herman for his help in constructing the equipment.

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☆U.S. GOVERNMENT PRINTING OFFICE: 1981-740-145/2780