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Direct conversion of coal into coal liquids is usually carried out in a reactor composed of gas, solid and liquid. Two types of coal liquefier reactors are examined for the optimization of coal conversion systems: three-phase fluidized bed and bubble column of slurry reactors. These two reactors have some similarities, i.e., both are three-phase reactors, both deal with mass transfer of gas species from gas phase into liquid and solid phases and both involve dissolution of solid phase. The main difference between them is that solid particles have much longer residence time in three-phase fluidized beds than in bubble columns.

The first part of this report is the literature survey of three-phase fluidization to clarify the hydrodynamic aspect of ebulated bed reactor design. The second part is the literature survey of the hydrodynamics of solid-liquid-gas system slurry reactors.

Part I

THREE-PHASE FLUIDIZATION FOR EBULATED BED REACTOR DESIGN

Three-phase fluidization is a process to allow good contact of gas, liquid and solids. The normal mode of operation is concurrent upward flow of gas and liquid which keeps solid particles in suspension. The properties and industrial applications of three-phase fluidization have been reviewed by Østergaard (1). One of the most important industrial applications of three-phase fluidization is the heterogeneous catalytic hydrogenation of residual oils or coal slurry for the removal of sulfur and the production of hydrocarbon distillates by hydrocracking.

The effectiveness of a three-phase fluidized bed used as a chemical reactor is determined largely by its hydrodynamic properties. The hydrodynamics of three-phase fluidization can be characterized by phase holdups, bubble behavior and dispersion of phases.

Phase Holdups

The holdup of a three-phase fluidized bed follows the relation:

$$h_G + h_L + h_S = 1$$

where h_G , h_L and h_S are the holdup of gas, liquid and solid, respectively.

The average solid holdup can be determined from bed expansion measurements. The holdups of gaseous and liquid phases may be determined from pressure drop measurements or from measurements of the mean residence time of the fluid phase.

A review of earlier studies on phase holdups, given by Østergaard (1), indicated that the phase holdups varied with gas velocity, liquid velocity,

particle size and amount of solids. Many authors (3, 4, 5, 6, 7) reported that the bed expansion (or the sum of the gas and liquid holdups) increased with increasing liquid or gas velocities. However, under certain conditions, the bed contracts with increasing gas velocity. This is a unique feature of three-phase fluidization. The contraction of a liquid-solid fluidized bed upon injection of gas bubbles was explained by Stewart and Davidson (5). They suggested that liquid was carried in the bubble wakes, which rises at a velocity considerably higher than the average liquid velocity in the bed. The velocity of liquid, other than those contained in the wakes, is consequently reduced because of conservation of mass. Hence the liquid voidage was reduced following the Richardson-Zaki type relationship (41). The bed contraction will be observed if the reduction of liquid voidage is greater than the holdup of gas in the bed. Epstein (2) derived a criterion for predicting either contraction or expansion caused by injection of gas bubbles into solid-liquid fluidized beds. This criterion was verified experimentally by Epstein and Nicks (28) for a variety of air-water-solid systems. Several studies (15, 23, 29, 30, 31) demonstrated that bed contraction would not occur in beds of large particles, e.g., glass spheres of sizes larger than 3 mm.

Adlington and Thompson (6) observed that the presence of solids had little influence on gas holdup at gas superficial velocity less than 1.5 cm/sec. At higher gas velocities the gas holdup was reduced by the presence of solids. Viswanathan et al. (11) found that for an air-water system, the gas holdup in beds of small particles (0.649 and 0.928 mm glass beads) was lower than the solid free system. However, the gas holdup in beds of large particles (4 mm glass beads) was higher than the

air-water system. Østergaard and Michelsen (10) measured the gas holdup in beds of 0.25, 1 and 6 mm glass particles. They found that h_G was proportional to U_{og}^n . U_{og} is the superficial gas velocity. n varies from 0.78 to 0.93 and has a value of 1.05 for a solid free system. They also found that in beds of 6 mm particles, the gas holdup decreased with increasing liquid flow rate, while in beds of 0.25 and 1 mm particles, the gas holdup increased with increasing liquid flow rate. Østergaard (1) found that for very small particles (40- through 60-mesh and 60- through 80-mesh), the gas holdup was independent of particle size and liquid velocity, but increased linearly with superficial gas velocity. Sherrand (9) observed that in beds of large or heavy particles (beds of 1.6 mm glass ballotini, 6 mm acrylic spheres and 12- through 14-mesh lead shot), the gas holdup decreased with increasing superficial liquid velocity, whereas in beds of small or light particles (beds of 12- through 14-mesh and 36- through 48-mesh glass ballotini) the gas holdup remained essentially independent of liquid velocity. Michelsen and Østergaard (33) studied the phase holdups in beds of three different particle sizes (6, 3 and 1 mm). The results showed that the liquid holdup increased with increasing liquid velocity and decreasing gas velocity. The gas and solid holdups did not change monotonically with gas and liquid velocities.

Kim et al. (23) studied the effects of viscosity and surface tension on phase holdups. The data showed that the liquid holdup as well as the sum of the liquid and gas holdups increased with increasing liquid viscosity for all particle sizes (1, 2.6 and 6 mm), whereas the effects of surface tension on the liquid holdup depended upon both the size and nature of solid particles. Armstrong et al. (32) showed that the

wettability of solids could significantly affect the phase holdups. The data indicated that at the same conditions nonwetable solids had slightly smaller solid holdup and much smaller gas holdup than wettable solids. Consequently, the liquid holdup would be higher for nonwetable solids.

Three models were proposed to describe three-phase fluidization. These models attempted to take into consideration the details of bed structures, such as bubble rising velocity and bubble wake volume. Often, the relations describing these details were empirical.

Ostergaard (8) proposed a model to describe three-phase fluidization based on the assumption that the bed consists of a liquid fluidized phase (ϵ_l), a bubble phase (ϵ_b) and a wake phase (ϵ_w). The liquid fluidized phase consists of the solids and the portion of liquid not contained in the wake phase. The wake phase moves at the gas velocity and has the same porosity as the liquid fluidized phase. Ostergaard also presented empirical equations relating the bubble velocity (U_b), ϵ_g , ϵ_w and ϵ_l to the superficial gas and liquid velocities.

Darton and Harrison (12) proposed a model based on the work of Stewart (13) and Efremov and Vankrushev (14). They proposed that the bubble wakes were particle-free and the liquid flux in the bubble wakes can be expressed by $\bar{K}U_g$. \bar{K} was the mean value of the ratio of liquid wake volume to bubble volume. Darton and Harrison developed an empirical correlation for \bar{K} as a function of U_l and U_g . The superficial liquid velocity and the liquid holdup in the particulate phase can be expressed as $(U_l - KU_g)/(1 - \epsilon_g - \bar{K}\epsilon_g)$ and $(\epsilon_l - \bar{K}\epsilon_g)/(1 - \epsilon_g - \bar{K}\epsilon_g)$, respectively. These expressions were used in conjunction with the Richardson-Zaki correlation to obtain a relation between ϵ_l and \bar{K} , U_g , U_l and ϵ_g . They

analyzed the data of Østergaard and Michelson (15) in terms of the drift flux. A plot of V_{CD} (drift flux) vs. ϵ_g revealed two regions: the ideal bubbly and the churn-turbulent. Using this plot and the relations mentioned above, the phase holdups can be calculated by an iterative procedure.

Bhatia and Epstein (16) proposed a generalized wake model. Four distinct phases in the bed, solid, liquid, gas and wake were proposed. The solid concentration in the wake phase can be expressed by X_k , a fraction of the solid concentration in the particulate liquid/solid fluidized phase. The value of X_k ranges from 0 to 1. The wake volume in the three-phase system was related to that in a gas/liquid two-phase system by an empirical function of solid holdup. The particulate liquid/solid phase was described by the Richardson-Zaki type model. Their model also took into consideration the possible existence of two-flow regimes (ideal bubbly and churn-turbulent) in the bed.

Data of various gas/liquid/solid combination in a fluidized bed were collected by many investigators. A summary of these experiments is listed in Table I. Empirical correlations for the volume fractions of phases are summarized and listed in Table II. These correlations relate the volume fractions to gas and liquid properties and superficial flow velocities.

Bubble Behavior

The size and the rising velocity of gas bubbles will influence the gas-liquid interfacial area, mass transfer coefficient and gas holdup. Darton and Harrison (17) studied the hydrodynamics of a single gas bubble. The bubble rising velocity was correlated by bubble diameter, minimum fluidization velocity, etc. A similar study was also reported by Massimilla et al. (18). However, significant wall effects may exist in their results.

Henriksen and Østergaard (19) indicated that the bubble rising velocity was proportional to the square root of the radius of the circular cap for large bubbles in a two-dimensional bed. Kim et al. (20) studied systems of 1 - 6 mm particles and found out both bubble size and bubble rising velocity increased with increasing gas velocity but are relatively insensitive to liquid velocity, viscosity and surface tension. At high gas velocity, the bubble characteristics were independent of particle size. The correlations suggested by Kim et al. are also listed in Table II.

The bubble coalescence or breakage in three-phase systems appears to depend on Rayleigh-Taylor instability, as for two-phase systems. Lee et al. (21) proposed a critical Weber number as a criterion for bubble stability. No breakup occurred for

$$W_e = \frac{\rho_s U_b^2 \bar{d}_s}{\sigma} \leq 3$$

Bubble breakup occurs in beds of large particles at high liquid flow rates and low gas flow rates. In the breakup regime the gas phase forms a uniform dispersion of small bubbles, and the extent of mixing in liquid is very low. Bubble coalescence takes place in beds of small particles at low liquid flow rates and high gas flow rates. The gas phase is characterized by a nonuniform bubble size distribution and the extent of mixing of liquid is high.

Dispersion of Fluids and Solid Phases

Very few data on gas-phase axial dispersion in a three-phase fluidized bed are available. Schügerl (22) and Michelsen and Østergaard (15) measured the gas-phase RTD in a three-phase concurrent upflow fluidized bed.

Schügerl reported that the intensity of mixing in both the gas and liquid phases decreased from the top to the bottom of the column. At low liquid velocities, the gas phase Peclet number increased with the gas flow rate, but at high liquid velocities, the Peclet number showed a maximum with respect to gas flow rate.

The liquid dispersion coefficients increase with gas flow rate. The particle size and liquid flow rate also influence the liquid phase dispersion coefficient. Kim et al. (23) correlated the data obtained by Michelsen and Østergaard (15) by dimensionless variables. Kato et al. (27) obtained another correlation by using dimensionless groups. All these correlations are listed in Table II.

Cova (24) measured solid phase axial concentration profiles at different gas and liquid flow rates. A theoretical model for axial concentration profiles expressed as a function of physical properties and operating conditions was presented. Kolbel et al. (25) found that the axial distribution of solids increased with decreasing gas velocity. The axial dispersion characteristics of solids was studied by Imafuku et al. (26) and Kato et al. (27). A wide range of variables was examined. The conclusions of these studies were: for small particles the dispersion coefficient of solids is the same as that of liquid in small diameter columns; for large particles and large columns the correlation can be revised to fit the conditions of large particles and large columns.

Table I. Summary of Data for Gas/Liquid/Solid Fluidization

Investigators	System	Particle Size	Experimental Unit
1. Razumov Manshillin Nemets [34]	Air/Water/Sand Air/Water/Slag Beads	0.493 to 1.27 mm	300 mm Diameter Column
2. Michelsen Østergaard [33]	Air/Water/Glass Beads	1, 3, 6 mm	5" Diameter Column
3. Dakshinamurty Subramanyam Rao [29, 30]	Air/Water/Various Spherical Particles Air/Kerosene/Various Spher- ical Particles	1.06 to 6.8 mm	56 mm Diameter Column
4. Viswanathan Kakal Murti [11]	Air/Water/Quartz Particles Air/Water/Glass Beads	0.928, 0.649 mm 4 mm	50.8 mm Diameter Column
5. Klm Baker Bergougnou [23]	Air/Water/Glass Beads Air/Water/Irregular Gravel	6 mm 2.6 mm	26 x 1" Rectangular Channel
6. Klm Baker Bergougnou [35]	Air-Water/Acetone-Glass Beads Air-Water/Acetone-Irregular Gravel Air-Sugar/Water-Glass Beads Air-Carboxymethyl Cellulose/ Water-Glass Beads Air-Sugar/Water-Irregular Gravel Air-Carboxymethyl Cellulose/ Water-Irregular Gravel	1, 6 mm 2.6 mm 1, 6 mm	26 x 1" Rectangular Channel

Table I (cont.)

Investigators	System	Particle Size	Experimental Unit
7. Blum Toman [36]	Nitrogen/Light Mineral Oil	See Table II	4" Diameter Column
8. Nemets Razumov Manshilin [37]	Air/Water/Sand Air/Heptane/Sand Air-Water/Glycerol-Sand	0.820 mm	90 mm Diameter Column
9. Bruce Revel-Chion [38]	Air/Water/Glass Spheres	2, 4, 6, 8 mm	46.3 mm Diameter Column
10. Østergaard Theisen [7]	Air/Water/Glass Ballotini	0.28 to 2.2 mm	2" and 4" Diameter Column

Table 11. Empirical Correlations for Three-Phase Beds

Authors	Correlation	Gas/Liquid	Solids (Dimension)	Column Diameter or Dimensions
Kim Baker Bergounghou [23]	$(c_1)U_g = 0 - c_1 = 0.0025 (Fr_1 \frac{\rho_g}{\rho_l})^{0.149} (Fr_B \frac{\rho_g}{\rho_s})^{0.181} Ro_1^{0.259} (\frac{Ro_1}{Ro_B})$ $(c_1)U_g = 0 = 0.409 (Fr_1 \rho_g / \rho_l)^{0.193} (Ro_1)^{0.074}$	Air/Water	Glass Beads (6 mm)	26" x 1"
Kim Baker Bergounghou [35]	$(c_1 + c_g) = 1.40 (Fr_1)^{0.17} (Wo)^{0.078} \text{ (Expanding Beds)}$ $(c_1 + c_g)U_g = 0 = 1.3 (Fr_1)^{0.126} (Wo)^{0.073} \exp [0.31 U_l/U_g]$ $(c_1)U_g = 0 \text{ (contracting beds)}$ $(c_1)U_g = 0 = 1.353 (Fr_1)^{0.208} (Ro_1)^{-0.1}$ $c_1 = 1.54 (Fr_1)^{0.234} (Fr_B)^{-0.086} (Ro_1)^{-0.082} (Wo)^{0.092}$	Air/Sugar Solutions Air/Carboxymethyl Cellulose Solution Air/Water Acetone	Glass Beads (6, 1 mm) Irregular Gravel (2.6 mm)	26" x 1"
Dakshinamurty Subrahmanyam Rao [29, 30]	$(c_g + c_1) = (K \frac{U_l}{U_c})^m (\frac{U_l U_g}{\sigma})^n$	Air/Water Air/Kerosene	Numerous Diameters 1.06 to 6.8 mm	56 mm Diameter

Table II. (cont.)

Authors	Correlation	Gas/Liquid	Solids (Dimension)	Column Diameter of Dimensions
	$n = 0.08$ $K = 2.12, m = 0.41, Re_t < 500$ $K = 2.65, m = 0.6, Re_t > 500$			
Blum Tomian [36]	$\frac{(c_g + \epsilon_1) - (\epsilon_1) U_g = 0}{1 - (\epsilon_1) U_g = 0} = f(U_g)$	Nitrogen/Light Mineral Oil	Cylinders Diameter Length 1/8" 1/8" 3/16" 3/32" 3/16" 3/16"	4" Diameter
Razumov Manshillin Nemets [34]	$\epsilon_s = 0.578 - 3.198 U_1 - 0.538 U_g$ $\epsilon_1 = 0.422 + 0.135 U_1/\bar{d} - 0.562 - 1.82 U_g$ $\epsilon_g = K(1 - \epsilon_s)^{2.09} (U_g/U_1)^{0.78}$	Air/Water	Sand, Slag Bends 0.49 to 1.27 mm	300 mm

Table II (cont.)

Authors	Correlation	Gas/Liquid	Solids (Dimension)	Column Diameter or Dimensions
Kim Baker Bergougnou (20)	$d_g = 13.4 U_z^{0.052} U_g^{0.248} \rho^{0.008} \sigma^{0.034}$ Error stand. = 7.2 mm $U_{br} = 83.1 U_z^{0.065} U_g^{0.339} \rho^{0.025} \sigma^{0.179}$ Error stand. = 164 mm/s $U_{br} = 18.0 d_g^{0.989}$ Error stand. = 178 mm/s		Glass Beads (1 mm, 6 mm) Irregular Gravel (2.6 mm)	660 mm x 25 mm
Rigby Van Blockland Verk Capes (39)	$[V_b - (\frac{\sigma}{\lambda}) + \frac{1}{\lambda}] (\frac{1-\epsilon}{\epsilon})^2 = 32.5 (\epsilon)^{1.53}$ Variance = 0.073 λ = average bubble length ϵ = 0.57 d_g		Glass (0.12, 0.29, 0.47 mm) Ottawa Sand (0.775 mm)	10 cm I.D.

Table II. (cont.)

Authors	Correlation	Gas/Liquid	Solids (Dimension)	Column Diameter or Dimensions
Kim et al. [23]	$(INU/H_o) - (INU/H_o) U_{og} = 0$ $= 0.068 (Fr_L \rho_s / \rho_L)^{0.128} (Fr_G \rho_s / \rho_G)^{0.168} (Re_L \cdot Re_G)^{0.120}$ $(\sigma' = 0.36)$ $INU = 2H/Po_L, Fr_L = U_{oL}^2 / d_p \cdot g, Fr_G = U_{oG}^2 / d_p \cdot g$ $Re_L = d_p U_{oL} \rho_L / \mu_L, Re_G = d_p U_{oG} \rho_G / \mu_G, Po_L = U_{oL} \cdot H / E_{zL}$ $(INU/H_o) U_{og} = 0 = 5.05 (Fr_L \rho_s / \rho_L)^{0.842} (Re_L)^{-0.450}$ $(\sigma' = 0.037)$			Rectangular
Kato [40]	$Po_L = U_{oG} d_p / E_{zL} = 13 (\sqrt{Fr_G} / (1 + 8(\sqrt{Fr_G})^{0.85}))$ $\sqrt{Fr_G} = U_{oG} / \sqrt{g d_p}$		63 - 177 μ m	6.6, 12.2, 21.4 cm Diameter
Kato et al. [27]	$Po_s = \frac{U_{oG} d_p}{H_{25}} = 13 (\sqrt{Fr_G})^{0.85} (1 + 0.009 Re_p (\sqrt{Fr_G})^{0.85}) / (1 + 8(\sqrt{Fr_G})^{0.85})$ $Re_p = d_p U_{oG} / \nu_L$		60-250 mesh 1 - 9 g/cm ³	5 - 22 cm Diameter

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Part II

SLURRY REACTOR FOR SOLID-LIQUID-GAS SYSTEMS

Flow Regime Maps

In developing a model for simulation of coal liquefaction processes, the nature of slurry-gas flow needs to be taken into consideration. Such a model should be more applicable to a wider range of operating conditions having a better theoretical base as opposed to just an empirical correlation.

Flow regime maps have been developed for gas-liquid systems in horizontal and vertical flows. Most of these studies were empirical except for the semi-theoretical flow models developed by Taitel and Dukler (1976) and Taitel et al. (1980) for transitions in horizontal and vertical flows, respectively. However, recent data revealed a number of deficiencies in these models. Weisman and co-workers (Weisman et al., 1978; 1981) then proposed an alternate set of correlations that was able to better fit the available data. Again the results were of an empirical nature. It has been suggested (Hetsroni, 1982) that perfect prediction methods may never emerge because of the indefinite nature of data interpretation. No standard experimental technique has been adopted in the collection of flow regime data. Nevertheless, the studies on flow regimes should be incorporated into the simulation models. The ability of models to predict flow regimes can be expected to improve when more accurate data becomes available.

The works of Taitel (1976; 1980) are reviewed. It is believed that their theoretical bases have wider applicability. The semi-theoretical models by Taitel et al. (1976; 1980) are recommended here for the time

being. Other works will be reviewed with the objectives of improving and generalizing the correlation for flow regime predictions.

Axial Dispersion Coefficients in Bubble Columns

Gas Axial Dispersion Coefficient

Gas phase dispersion is important in bubble columns of diameters greater than 0.5 m and especially at high conversion levels. There are relatively few correlations on gas dispersion coefficients and some are listed below:

*Towell and Ackermann (1972)	$D_g = 19.7D^2U_g$	(MKS units)
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*Pilhofer et al. (1978)	$D_g = 2.64 U_s^{3.56}$	$D < 100 \text{ mm}$
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	$D_g \sim D^{1.5}$	$D > 100 \text{ mm}$
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where	$U_s = U_g/\epsilon$	(MKS units)
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Field and Davidson (1980)	$D_g = 56.4D^{1.33} (U_g/\epsilon)^{3.56}$	(MKS units)
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**Mangartz and Pilhofer (1980)	$D_g = 5 \times 10^{-4} D^{1.5} (U_g/\epsilon)^3$	(cgs units)
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where	$D_g =$ gas dispersion coefficient	
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	$U_g =$ superficial velocity of gas	
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	$D =$ column diameter	
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	$\epsilon =$ gas holdup	
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*Obtained indirectly from Field and Davidson (1980)

**Obtained indirectly from Deckwer et al. (1980)

The correlation of Field and Davidson (1980) successfully represented the data of Pilhofer et al. (1978). It is applicable for a wide range of column diameter (experimental data ranges from 0.305 m to 5.2 m) with a mean deviation of 46 percent from data used.

Liquid Axial Dispersion Coefficient

Shah and Sharma (1978) presented a review on backmixing coefficients in gas-liquid reactors where gas is dispersed as bubbles in a continuous liquid phase. For vertical gas-sparged reactors the axial dispersion coefficient is essentially independent of liquid velocity and the liquid properties such as viscosity, surface tension, density, etc. A more recent work by Field and Davidson (1980) provides the following correlations:

$$D_L = 0.90^{1.5} [L(U_g - \epsilon U_s)]^{1/3} \quad U_g \gg U_L$$

$$D_L = 0.44 D^{1.55} g^{1/3} [U_g - \epsilon U_s - \epsilon U_L H/L(1 - \epsilon)]^{1/3} \quad U_g \text{ comparable to } U_L$$

where

$$U_s = U_b (1 - \epsilon)^{1.39} (1 + 2.55 \epsilon^3) \quad (\text{Lockett and Kirpatrick, 1975})$$

U_b = rise velocity of single bubble

U_s = slip velocity

L = distance from spargers to surface

ϵ = gas voidage

H = distance between liquid inlet and outlet

(MKS units)

Although this work includes U_L and L , the slip velocity is difficult to obtain for industrial systems.

Ying et al. (1980) also investigated this subject for a water-sand slurry/air system in 5- and 12-inch columns. He compared the correlations of several workers with his data. Although no general agreement can be found with regard to the effects of liquid properties, he concluded that the presence of sand reduced the liquid dispersion coefficient. However, the solid concentration and particle size have no effects on the liquid dispersion coefficient.

Solid Axial Dispersion Coefficient

For large particles and large columns, Kato et al. (1972) gave

$$Pe_s = \frac{U_g D}{D_s} = 15Fr_g(1 + 0.009 Re_p Fr_g^{-0.8}) / (1 + 8Fr_g^{0.85})$$

where $Fr_g = U_g / (gD)^{1/2}$ and $Re_p = d_p U_t \rho_L / \mu_L$

U_t = terminal velocity of a single particle

D_s = backmixing coefficients for the solid particles

g = gravitational acceleration

d_p = particle diameter

μ_L = viscosity of liquid

In the current investigation, the dispersion coefficients (gas, liquid, solid) will be examined in order to choose the best correlations applicable

to a different range of operating conditions. A generalized correlation to predict values for the dispersion coefficients will be formulated, if possible.

Gas Holdup

Ying et al. (1980) suggested that the correlations of Yoshida and Akita (1973) are adequate for gas-liquid-solid systems at high gas flow rates (greater than 6.1 cm/s for N₂/water/silica system). Other conclusions by the same workers include:

1. The gas holdup is reduced by the presence of solid particles at low superficial velocities.
2. At higher superficial velocities, there is very little difference between the holdups for gas-liquid and gas-liquid-solid systems.
3. Increasing the solid concentrations decreases the gas holdup.
4. The gas distributor and the liquid flow rate does not affect gas holdup.

Deckwer (1980) studied the gas holdup in an N₂/paraffin/Al₂O₃ powder system and showed that the correlation of Yoshida and Akita (1973) failed completely for gas velocities above 1 cm/s. Therefore, a more detailed examination is being done on gas-liquid-solid systems. One conclusion that can be made and agreed to by both groups is that the presence of solids decreases the gas holdup.

Interfacial Area and Bubble Size

The gas/liquid interfacial area can be estimated if the gas holdup and mean surface to volume bubble diameter is known by $a = 6\epsilon_G/d_b$. Some relationships for holdups and bubble diameters are given below.

Hughmark (1967)

$$d_b = 0.635 \left(\frac{\sigma}{72}\right)^{0.6} \left(\frac{1}{\rho_L}\right)^{0.2}$$

$$\epsilon_G = [2 + (0.55/U_G)(\rho_L \sigma / 72)^{1/3}]^{-1}$$

Akita and Yoshida (1973, 1972)

$$\frac{\epsilon_G}{(1 - \epsilon_G)^4} = 0.20 \left(\frac{D^2 \rho_L g}{\sigma}\right)^{1/8} \left(\frac{D^3 \rho_L^2 g}{\mu_L^2}\right)^{1/12} \left(\frac{U_G}{(Dg)^{1/2}}\right)$$

$$\frac{d_b}{D} = 26 \left(\frac{D^2 \rho_L g}{\sigma}\right)^{-0.5} \left(\frac{D^3 \rho_L^2 g}{\mu_L^2}\right)^{-0.12} \left(\frac{U_G}{(Dg)^{0.5}}\right)^{-0.12}$$

where

d_b ~ bubble diameter	g ~ acceleration constant
G ~ gas holdup	D ~ diameter of column
σ ~ surface tension	μ_L ~ viscosity of liquid
ρ_L ~ density of liquid	a ~ interfacial area
U_G ~ gas velocity	

(cgs units)

Heat Transfer in a Bubble Column

Summary of findings by past investigators:

1. The main effects on the heat transfer coefficient were from gas velocity and liquid phase properties.
2. Geometric sizes of the columns used did not affect heat transfer.

3. Different gas spargers did not lead to different heat transfer coefficients.
4. Bubble diameter has no influence on heat transfer.
5. Effect of the liquid flow rate on the heat transfer coefficient may be considered negligible.
6. Increasing solid concentration increases the heat transfer coefficients.
7. For small particles ($d_s < 5 \mu\text{m}$), the suspension can be treated as a homogeneous phase. The effects of solid concentration could be represented by the change in physiochemical properties (density, viscosity, heat capacity, thermal conductivity of slurry).

Two correlations are recommended.

Deckwer (1980) presented a semitheoretical correlation for gas velocities less than 10 cm/s. Hikita et al. (1981) presented a correlation for gas velocities reaching 0.34 m/s. However, since Deckwer's correlation was based upon theory, its applicability for general systems is recommended. Hikita's correlation should be considered whenever gas velocities are above 10 cm/s.

$$\text{Hikita et al.} \quad \frac{h_w}{\rho C_p U_G} \left(\frac{C_p \mu}{k} \right)^{2/3} = 0.411 \left(\frac{U_G \mu}{\sigma} \right)^{-0.851} \left(\frac{\mu^4}{\rho \sigma^3} \right)^{0.308}$$

$$\text{Range of Validity:} \quad 5.4 \times 10^{-4} < (U_G \mu / \sigma) < 7.6 \times 10^{-2}$$

$$4.9 < (C_p \mu / k) < 93$$

$$7.7 \times 10^{-2} < (\mu^4 g / \rho \sigma^3) < 1.6 \times 10^{-6}$$

$$\text{Deckwer} \quad \frac{h_w}{\rho C_p U_G} \left(\frac{C_p \mu}{k} \right)^{1/2} = 0.1 \left(\frac{U_G^3 \rho}{\mu g} \right)^{-0.25}$$

Range: $U_G < 10 \text{ cm/s}$

Work Plan in the Future

The collection of semitheoretical and empirical equations for holdups, dispersion coefficients, heat and mass transfer coefficients, bubble size, interfacial areas and flow regime determination will be critically reviewed. The range of applicability will be assessed along with other proposed equations. The result should yield a generalized system of equations for a hydrodynamic model. If a gas-liquid-solid system were given along with physical properties, a prediction of results should be possible. However, the accuracy would have to be verified by experimental data.

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