

Heat Transfer in Three-Phase Fluidized Beds

Wall-to-bed heat transfer coefficients in three-phase fluidized beds with glass beads and cylindrical γ -alumina particles were determined experimentally. In the present study, the relative increase in heat transfer coefficient was found equal to the relative decrease in liquid holdup in three-phase fluidized beds. Modified Stanton number has been correlated with modified Reynolds number.

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SCOPE

This study investigates the effects of fluid velocity, particle size, particle density, and particle shape on the wall-to-bed heat transfer coefficients in three-phase fluidized beds. Three-phase reactor applications are increasing in the chemical process industry. Classifications of the various reactor types are presented by Chiu (1982). The characteristics of three-phase slurry reactors have been reviewed by Chadhari and Ramachandran (1980), and those of trickle bed reactors by Satterfield (1975). Epstein (1979) and Shah (1979) have reviewed the characteristics of three-phase fluidized beds.

A three-phase fluidized bed as defined in the current study is a bed of particles fluidized by the cocurrent upward flow of gas and liquid. In operation, liquid and gas flow through the bed and maintain the expanded bed of catalyst particles in a continuous random motion. The liquid phase forms a continuous medium, and the gas is dispersed as discrete bubbles. Three-phase fluidized beds are distinguished by their high transport rates and good phase contact.

The principal applications of three-phase fluidized beds with cocurrent upward flow of gas and liquid are in catalytic reactors, gas absorbers and fermenters. Applications of three-phase fluidized beds in catalytic reaction include the heterogeneous catalytic hydrogenation of residual oil or coal slurry for the removal of sulfur, the production of hydrocarbon distillates by hydrocracking, and methanation of gases containing hydrogen and carbon oxides. The liquid phase in three-phase fluidized beds may be either a reactant or an inert used as a heat transfer medium. An example of the use of liquid phase as the heat transfer medium is Chem Systems three-phase fluidized bed used for the methanation of gases containing hydrogen and carbon oxides (Blum and Tomn, 1977). A number of industrial applications of three-phase fluidized beds have been discussed by Ostergaard (1968, 1971).

Recently, Li and Lin (1981) noted the application of three-phase fluidized beds to upgrading heavy crude. Potential ad-

vantages include minimal on stream catalyst replacement, fouling-free operation, minimum reactor temperature gradient, and the ability to use small catalyst to enhance reaction rates. Temperature uniformity has been demonstrated by van Driesen and Steward (1964) in a commercial desulfurization unit, and in H-oil and H-coal reactors by Li and Lin (1981). This factor is important in choosing the best reactor type for highly exothermic reactions. There are undoubtedly interrelationships among hydrodynamics, mixing, and transfer rates. Phase holdup is one of the important hydrodynamic phenomenon in three-phase fluidized beds and is reviewed elsewhere, Chiu (1982). Wall-to-bed heat transfer coefficients are essential for the estimation of the transfer surface required to remove heat from or to supply heat to a three-phase fluidized bed. The maintainance of a uniform temperature, by using such an exchange surface is necessary to avoid the local hot spots which may cause catalyst degradation and/or reduce yield by upsetting the product distribution.

In contrast to the large number of investigations of phase holdup in three-phase fluidized beds, very little has been published on heat transfer in three-phase fluidized beds. Published data are largely confined to systems with spherical particles. Previous work on wall-to-bed heat transfer and heat transfer from immersed surfaces in three-phase fluidized beds have established that heat transfer coefficients are generally higher than those in corresponding liquid fluidized beds. An exception was found by Kato et al. (1978).

The catalyst support is traditionally γ -alumina due to its inertness to most reacting species, structural stability at relatively high temperature, and large surface area. Therefore, γ -alumina was chosen as the solid particles in this study. To interpret the heat transfer phenomena, bed expansion, phase holdup, and pressure drop were also studied (Chiu, 1982). Complementary measurements were also made on gas-liquid flow beds, and two-phase liquid-solid fluidized beds.

CONCLUSIONS AND SIGNIFICANCE

Our conclusions concern heat transfer in three-phase fluidized beds, and apply to all sizes of spheres and to the different types of cylindrical particles studied unless otherwise noted. Heat transfer coefficients in three-phase fluidized beds are higher than those of comparable liquid fluidized beds. Heat transfer coefficients in three-phase fluidized beds are higher than those of comparable gas-liquid systems except for small spherical particle size situations at high gas velocity. Plots of heat transfer coefficients in three-phase fluidized beds vs. bed porosity or

liquid velocity pass through a maximum as the bed porosity or liquid velocity, respectively, is progressively increased. The bed porosity at which the maximum value occurs decreased with increasing particle size. The relative increase of heat transfer coefficient is equal to the relative decrease of liquid holdup. The heat transfer coefficient in both three-phase and liquid fluidized beds can be correlated by:

$$ST_{m3} = St_{m2} = C Re_{m2}^{-0.305} Pr^{-2/3}$$

The above correlation can be useful in reactor design. Heat transfer in three-phase fluidized beds is hypothesized to encounter a wall resistance and bed resistance in series. All of the aforementioned conclusions, are consistent with this hypothesis.

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BACKGROUND

Wall-to-bed heat transfer coefficient in a three-phase fluidized bed has been measured by Østergaard (1964), Viswanathan et al. (1965), and Kato et al. (1978, 1980). Heat transfer from an immersed surface in a three-phase fluidized bed has been reported by Baker et al. (1976, 1978). The investigators confined themselves to systems with spherical particles.

Østergaard (1964) found that the heat transfer coefficient was a strong function of gas velocity, especially for liquid velocity close to the minimum fluidization velocity but a rather weak function of liquid velocity. Viswanathan et al. (1965) found that heat transfer coefficient achieved maximum value with respect to the gas/liquid ratio and solid concentration. Also, the radial temperature profiles are flat except in the region very close to the wall. Unfortunately, their experiments were carried out at a constant bed height, which was achieved by simultaneously changing the water and air flow rates over relatively narrow ranges.

Baker et al. (1978) used a coaxial electric heater to study the effects of fluid velocities and particle size on the heat transfer coefficient. It was found that heat transfer coefficient increased with increasing superficial gas velocity for all liquid flow rates and particle sizes. For a given particle size, the heat transfer coefficient passed through a maximum value as bed porosity was increased. The bed porosity at which the maximum values occurred decreased as particle size increased. The heat transfer coefficient remained essentially independent of particle size when particle size was greater than approximately 0.3 cm. Their study proposed an empirical correlation for heat transfer coefficients.

died unless Heat transfer coefficients in a three-phase fluidized bed with 0.22-cm glass beads particles have been reported by Kato et al. (1978). It was found that heat transfer coefficients in three-phase fluidized beds are smaller than those in two-phase liquid-solid fluidized beds. This trend is just opposite to the findings of Baker et al. (1978) and of the present work. Kato et al. (1980) found that the rate of increase of heat transfer coefficient with respect to liquid velocity decreased gradually, and finally the heat transfer coefficient converged to the value found in a bubble column. They also

found that the effects of liquid velocity and properties of solid particles on the heat transfer coefficient are relatively small. Heat transfer coefficient in a three-phase fluidized bed increased with increasing gas velocity and decreased with increasing liquid viscosity. A summary of experimental conditions on previous heat transfer studies is presented in Table I.

HEAT TRANSFER EXPERIMENTS

The test section, a modification of the equipment of Patel and Simpson (1977), consisted of a vertical 5.08-cm I.D. copper tube, Figure 1. The pipe section in the upper part of the test section was made of Lucite so that bed expansion could be observed. The top of the pipe section protruded into a concentrically mounted outlet header. The test section was electrically heated with a Watlow band heater. The heater section (or test section) was itself divided into three parts: the upper heated section; main heater section; and the lower heated section. The main heater section (Figure 2) was inserted between the lower and upper parts and was separated from them by an insulating Micarta spacer. All data on heat transfer coefficients were taken with the main heater in operation. The lower heater section consisted of three heaters. The topmost heater rate 600 W, 240 V was 7.62 cm long and the Micarta spacer. The two lower heaters, each 15.24 cm long and rated at 1,200 W, 240 V abutted each other and the top heater. The upper heated section was fitted with a 300-W heater, 5.08 cm long.

Sixteen copper-constantan thermocouples were used to measure the wall temperature, liquid inlet and outlet temperatures, and room temperature. The liquid outlet temperature was also measured with calibrated thermometer. A potentiometer with range 0–1 mV was used to measure the emf values. Water inlet temperature was used as the reference temperature. On calibration it was found that the thermocouple constant was 0.04 mV/°C. The thermocouple constant was checked by comparing the temperature difference between the water inlet (or outlet) temperature and room temperature with the difference in emf values during each experiment. Each heater had its own 110 V AC autotransformer power supply.

TABLE 1. EXPERIMENTAL CONDITIONS OF PREVIOUS HEAT TRANSFER INVESTIGATION IN THREE-PHASE FLUIDIZED BEDS

Authors	Gas	Liquid	Solid	Heating Method	Variables Studied	D_t , cm	d_p , cm	$\frac{u_L}{cm/s}$	$\frac{u_g}{cm/s}$
Østergaard (1964)	Air	Water	Glass Beads	Steam Heated Wall	Fluid Velocity	7.62	0.05	0.784	0.0–13.5
Viswanathan (1965)	Air	Water	Quartz	Steam Heated Wall	Fluid Velocity Solid Concentration	5.1	0.0649 –0.0928		
Armstrong (1976)	Air	Water	Glass Beads	Coaxial Electric Heater	Fluid Velocity Particle Size	24.0	0.05 0.1, 0.3 0.5	0.8 4.0– 12.6	0.0– 23.7
Baker et al. (1978)	Air	Water	Glass Beads	Coaxial Electric Heater	Fluid Velocity Particle Size	24.0	0.5 0.1 0.3, 0.5	0.8 4.0– 12.6	0.0– 23.7
Kato et al. (1978)	Air	Water	Glass Beads	Nichrome Wire Heated Wall	Fluid Velocity	12.0	0.22	0.4– 6.0	0.0– 15.0
Kato et al. (1980)	Air	Water CMC Sol.	Glass Beads Alumina Sphere	Nichrome Wire Heated Wall	Fluid Velocity Particle Size Particle Density Column Dia. Liquid Viscosity	5.2 12.0	0.042 0.066 0.12 0.22 0.33	0.3– 15.0	0.0– 15.0
Muroyama et al. (1981)	Air	Water	Glass Beads, Activated Carbon Beads, Alumina Beads		Fluid Velocity Particle Size Particle Density	9.56	0.061– 0.69	—	—
Present work	Air	Water	Glass Beads, Cylindrical γ -Alumina	Electric Heated Wall	Fluid Velocity Particle Size Particle Density Particle Shape	5.08	0.1, 0.3, (0.3175×) (0.3175) (0.476×) (0.3175) (0.476×) (0.476)	6.3– 15.0	0.0– 13.7

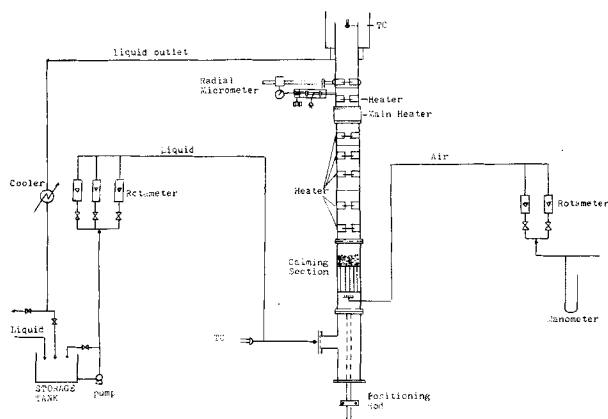


Figure 1. Flow diagram of heat transfer experiment.

The power to the heater was measured by calibrated wattmeters and checked against temperature rise of the flowing fluid. To minimize heat loss from heaters, glass wool insulation was placed over the entire test section. The power to the guard heater was adjusted to prevent heat losses from the main heater to the ambient. This design thermally isolated the main heater so that all of its power was dissipated in the bed. The radial temperature profiles were obtained by laterally moving a stiffened-and-sheathed thermocouple whose locations were measured with a dial indicator. Temperature reproducibility was $\pm 0.1^\circ\text{C}$ and positional reproducibility $\pm 0.05\text{ cm}$.

The temperature is measured at a position far from the fluids inlet (same level). Also, the heat capacity of the gas is small in comparison to the liquid. The gas temperature should, therefore, approach that of the liquid with reasonable accuracy. The maximum error for neglecting wall resistance was estimated to be 1.2%. Other procedures and safeguards employed for accuracy are discussed in detail elsewhere (Chiu, 1982). To avoid including any thermal entrance length effects, the temperature difference, $T_w - T_B$, was used rather than an integrated or logarithmic mean temperature difference to calculate the heat transfer coefficient. The coefficient was calculated from the following equations:

$$h = \frac{q_w}{A_s (T_w - T_B)} \quad (1)$$

and

$$q_w = mC_p (T_{\text{Bout}} - T_{\text{Bin}}) \quad (2)$$

Where T_B is the bulk stream temperature of the bed at the level of the main heater, and T_w is the wall temperature.

$T_w - T_B$ was calculated from

$$T_w - T_B = \frac{\int_0^R u(r) [T_w - T(r)] r dr}{\int_0^R u(r) r dr} \quad (3)$$

Cairns and Prausnitz (1959) and Latif and Richardson (1972) found

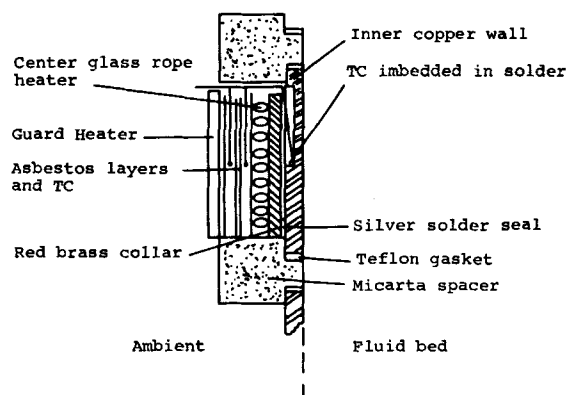


Figure 2. Detail of main heater.

that the velocity profile in liquid fluidized beds was almost flat. No information on velocity profile in three-phase fluidized beds is available in the literature. Turbulence within the bed has been shown to intensify upon the addition of gas. This tends to flatten the velocity profile. Therefore, it was reasonable and convenient to assume a flat velocity profile in the present study.

Particle Density and Size

Two sizes of glass beads and three of cylindrical γ -alumina were used in the liquid-solid and three-phase experiments. The particle diameters and length (for cylindrical particles) were measured using a micrometer. Fifty particles of each type were measured and the arithmetic average is the particle diameter and length given. For cylindrical particles, the equivalent diameter, d_e , is calculated from

$$d_e = (1.5 \times d_p^2 L)^{1/3} \quad (4)$$

where d_p is the particle diameter, and L is particle length. The shape factor ϕ_s , is calculated from

$$\phi_s = \frac{d_e^2}{d_p \left(L + \frac{1}{2} d_p \right)} \quad (5)$$

Since γ -alumina is a porous material, it is necessary to know the particle porosity, ϵ_p . The particle porosity measurements were described by Chiu (1982).

The density of the particles was measured using the standard water displacement technique. For cylindrical particles, there are three different definitions used for density, namely dry bulk density, ρ_d , wet bulk density, ρ_e , and true density, ρ_s . The various density measurements have been described by Chiu (1982). The relationship among the three solids densities found can be expressed as follows:

$$\rho_e = (1 - \epsilon_p) \rho_s + \epsilon_p \rho_L \quad (6)$$

$$\rho_d = (1 - \epsilon_p) \rho_s \quad (7)$$

A summary of particle properties is given in Table 2.

TABLE 2. PARTICLE PROPERTIES

Particle	d_e , cm	Diam., cm	Length, cm	ϕ_s	ρ_s , g/cm ³	ρ_d , g/cm ³	ϵ_p
Glass Beads	0.05	0.05		1.0	2.67	2.67	
	0.1	0.1		1.0	2.67	2.67	
	0.3	0.3		1.0	2.83	2.83	
γ -Alumina	0.349	0.3175 ^a (0.308) ^b	0.3175 (0.3)	0.876	3.3	1.479	0.552
	0.461	0.476 (0.467)	0.3175 (0.3)	0.857	3.285	1.475	0.551
	0.5329	0.476 (0.46)	0.476 (0.476)	0.873	3.14	1.43	0.545

^a Nominal particle size.

^b Actual particle size.

ϕ_s and d_e based on actual particle size.

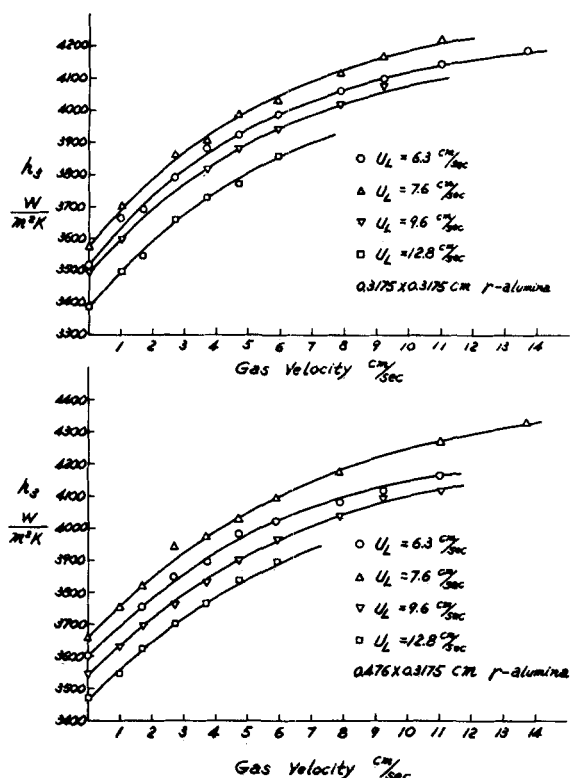


Figure 3. Heat transfer coefficients as a function of gas velocity.

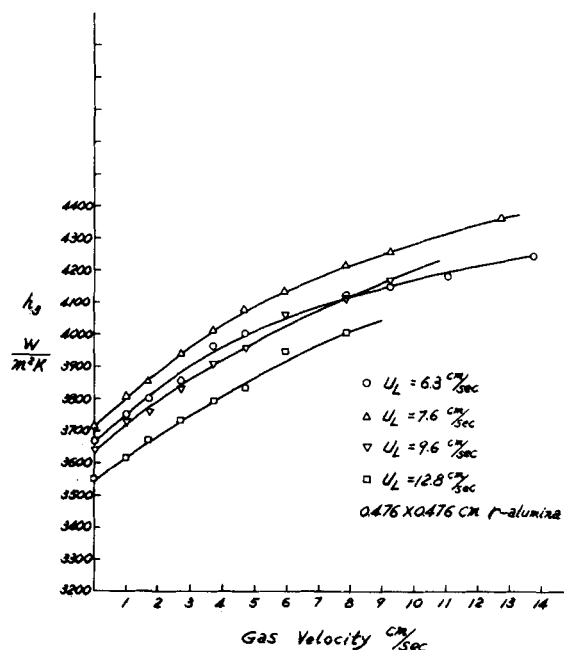


Figure 4. Heat transfer coefficients as a function of gas velocity.

trend was observed by Baker et al. (1978) for heat transfer from an immersed heater surface in a three-phase fluidized bed and for glass beads in the present system (Chiu, 1982). However, Kato et al. (1978) found that wall-to-bed heat transfer coefficients for three-phase systems were larger than those for liquid-solid systems in the range of $u_L < 1.6$ cm/s, but reported the converse to be true in the range of $u_L > 1.6$ cm/s. The magnitude of the initial increase of the heat transfer coefficients upon introducing gas at low velocity to a liquid fluidized bed depends on the liquid velocity and particle size. This initial jump in coefficient value will increase when either particle size or liquid velocity, is decreased. The heat transfer coefficients in three-phase fluidized beds approach asymptotic values at higher gas velocities.

Figure 5 presents the variation of heat transfer coefficient with liquid velocity at a given gas velocity. The heat transfer coefficient

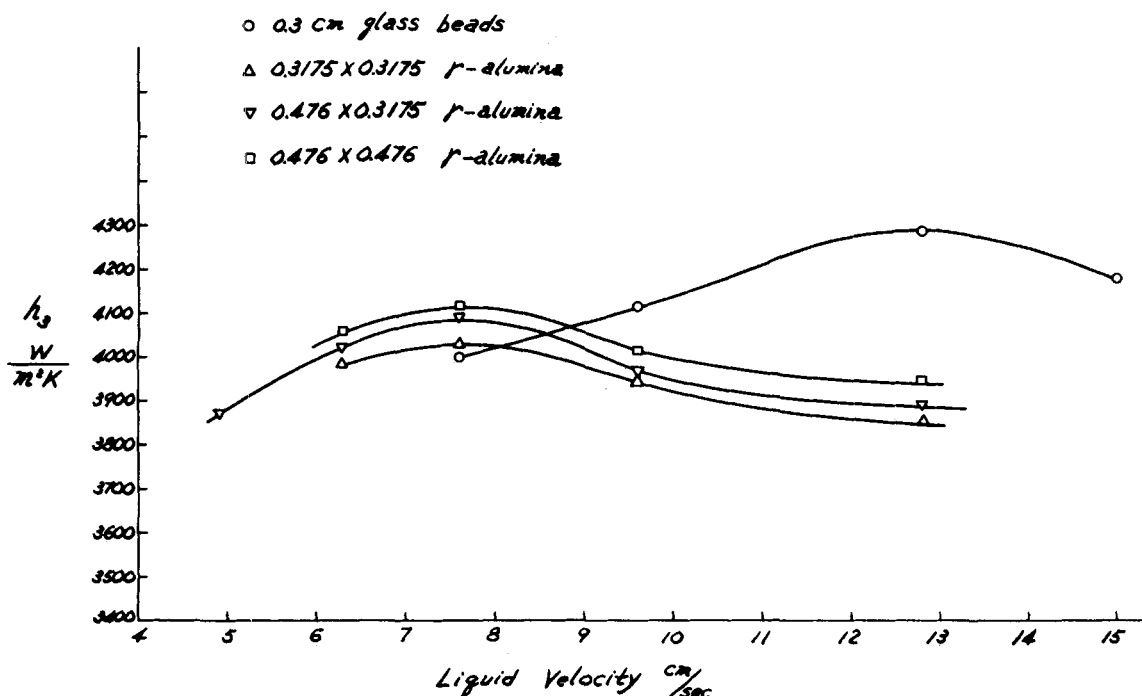


Figure 5. Heat transfer coefficients vs. liquid velocity @ $U_g = 5.92$ cm/s.

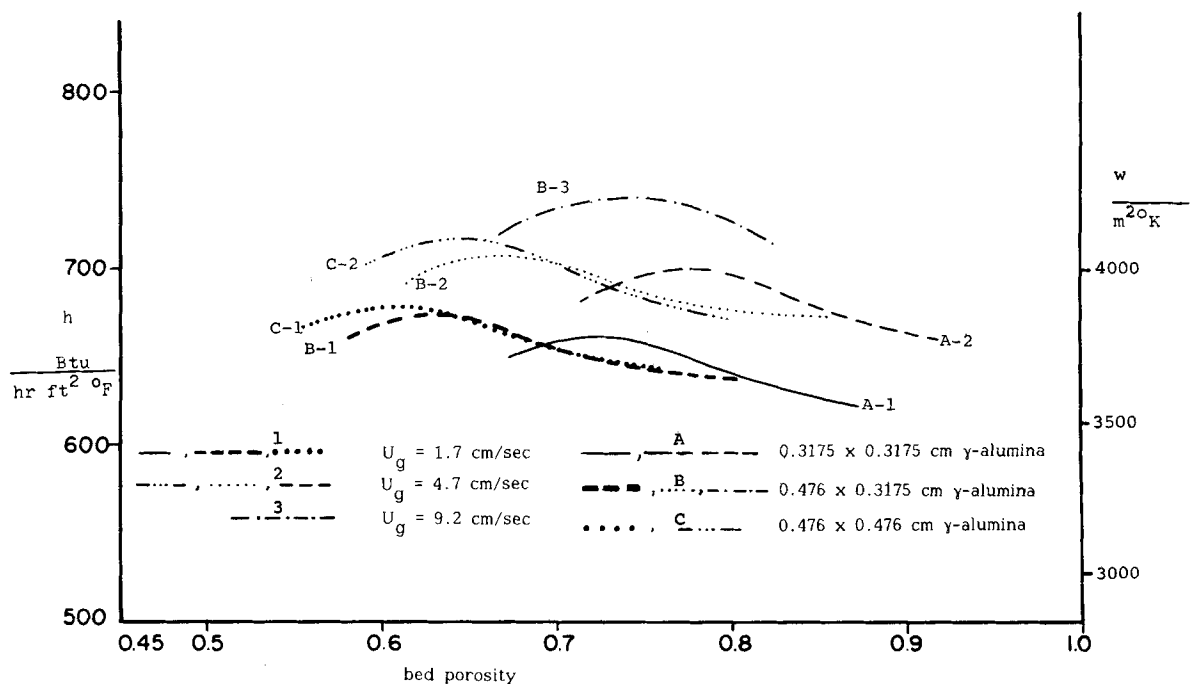


Figure 6. Heat transfer coefficient vs. bed porosity for cylindrical particles.

passes through a maximum value as the liquid velocity is progressively increased. The corresponding liquid velocity falls between 7 and 8 cm/s for all γ -alumina systems and is about 12.8 cm/s for 0.3 cm glass beads.

Figure 6 presents typical plots of heat transfer coefficient vs. bed porosity in three-phase fluidized beds using data obtained at various liquid velocities for a given gas velocity. It can be seen that the heat transfer coefficient passes through a maximum value as bed porosity is progressively increased at a given gas velocity. The bed porosity corresponding to the maximum heat transfer coefficient decreased with increasing particle size. At a given particle size, the effect of increasing gas velocity was to shift the maximum to higher bed porosities for large particles.

The effect of the presence of solid particles in gas-liquid systems on the heat transfer coefficient can be determined from the plot of heat transfer coefficient vs. particle size, Figure 7. It was observed that for a representative liquid velocity, 7.6 cm/s, heat transfer coefficients generally increased in the presence of solid particles at low gas velocities. However, the coefficient passes through a minimum value with progressive increase in particle size at 11-cm/s gas velocity (i.e., the coefficients in three-phase fluidized beds can be smaller than those in comparable gas-liquid systems). The same trend but for a heater surface coefficient vs. particle size was observed by Baker et al. (1978) at roughly the same liquid velocity. For large particles (greater than 3 mm), the heat transfer coefficients in three-phase fluidized beds are always larger than those in the comparable gas-liquid systems. Operating at well above u_{mf} , heat transfer coefficients in a three-phase fluidized bed

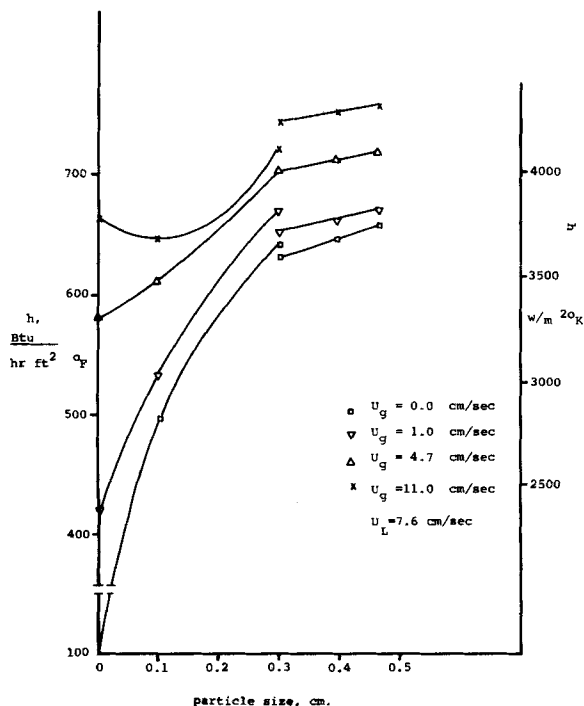


Figure 7. Heat transfer coefficients vs. particle size at typical liquid velocity.

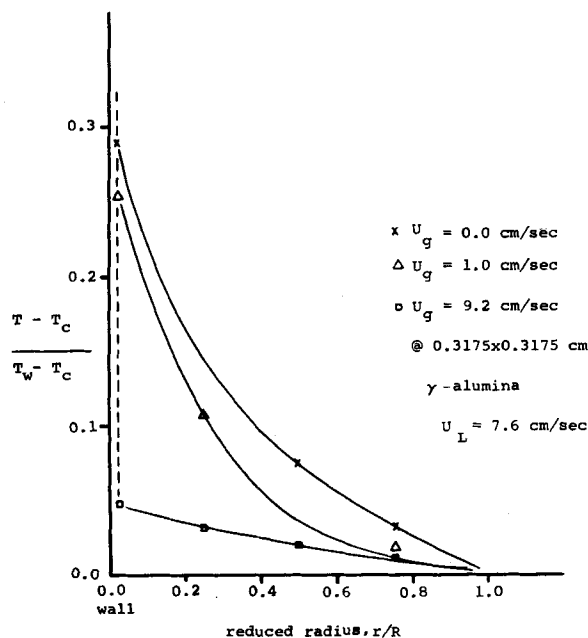


Figure 8. Comparison of radial temperature profiles in three-phase systems with liquid fluidized bed.

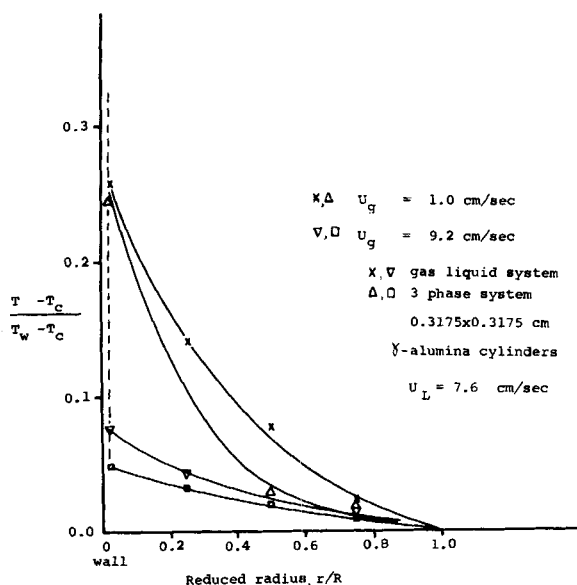


Figure 9. Comparison of radial temperature of three-phase systems with gas-liquid systems.

will be higher than a comparable gas-liquid situation, except for a small particle size, high gas velocity condition. The discontinuity in Figure 7 is possibly due to different particle density and/or shape.

The effect of gas velocity on the radial temperature profiles in three-phase fluidized beds is shown in Figure 8. It was observed that the temperature distributions flattened with increasing gas velocity. This indicates that radial mixing in three-phase fluidized beds increases with increasing gas velocity. A similar trend was observed by Vail et al. (1968) and El-Temtamy et al. (1979) for the radial mixing of liquid phase in three-phase fluidized beds. Figure 9 displays the effect of solid particle addition on the temperature profiles of a gas-liquid system, namely a flattening of the temperature profile. However, the effect is relatively insensitive at high gas velocities. A similar trend was observed by Vail et al. (1968) for radial mixing in three-phase fluidized beds.

A linear regression analysis was performed on the heat transfer data of the present three-phase fluidized-bed study. A 0.052 power dependence on the superficial gas velocity was found for cylindrical γ -alumina systems (correlation coefficient is 0.985). The effect of gas velocity on heat transfer coefficients in three-phase fluidized beds is less significant than in comparable gas-liquid systems, i.e., 0.19 power in the Chiu (1982) study.

For 0.1-cm glass beads, Chiu's coefficients were slightly higher than of Baker et al.'s (1978) correlation (and nearly identical to Baker's actual data) and slightly lower than those of Kato et al. (1980) at high gas velocities. For 0.3-cm glass beads, Baker et al. found that heat transfer coefficients at 9.2 cm/s liquid velocity were greater than those at 12.6 cm/s for any given gas velocity. However, Chiu's data followed the opposite trend. Possible reasons for the differences will be discussed later.

Data Correlation

Baker et al. (1978) proposed the following equation to represent experimental heat transfer from an immersed surface to a three-phase fluidized bed:

$$h_3 = 3,393 u_L^{0.07} u_g^{0.059} d_p^{0.106} \quad (8)$$

where h_3 is in $W/m^2 \cdot K$. It may be inferred from this correlation that heat transfer coefficients in three-phase fluidized beds increase with increasing liquid velocity at a given velocity. Also, the heat transfer coefficient should approach zero as gas velocity approaches zero. To incorporate both the observed trends and the limiting velocity case, a new correlation case is proposed in the present work. The J factor and modified Reynolds number approach used to

TABLE 3. COMPARISON OF MODIFIED STANTON NUMBER IN THREE-PHASE AND LIQUID FLUIDIZED BEDS

Particle Size Dia. or Dia. \times length, cm	Liquid Velocity cm/s	$Pr^{2/3} St_{m3}$	$Pr^{2/3} St_{m2}$	$\frac{St_{m2}}{St_{m3}}$
0.1	7.6	0.02412	0.025172	1.0437
0.3	7.6	0.0255	0.0247	0.968
Glass Beads	9.6	0.02248	0.02252	1.002
	12.8	0.0199	0.01961	0.985
	15.0	0.01767	0.01767	1.0
0.3175 \times 0.3175 γ -Alumina	6.3	0.03257	0.03148	0.967
	7.6	0.0295	0.0285	0.966
	9.6	0.0253	0.241	0.951
	12.8	0.020	0.01943	0.964
0.476 \times 0.3175 γ -Alumina	6.3	0.0308	0.0313	1.015
	7.6	0.0272	0.0282	1.037
	9.6	0.0238	0.0243	1.022
	12.8	0.02063	0.0202	0.980
0.476 \times 0.476 γ -Alumina	6.3	0.03228	0.0311	0.964
	7.6	0.0274	0.0283	1.033
	9.6	0.02361	0.02415	1.023
	12.8	0.01965	0.01988	1.012

correlate data on heat transfer in liquid fluidized beds was extended to three-phase systems by modifying the Stanton number and the Reynolds number. Liquid holdup is known to change when gas is introduced to the bed. The interstitial liquid velocity is defined by u_L/ϵ_{L3} , where ϵ_{L3} is the liquid holdup in three-phase systems.

The modified Stanton number in three-phase fluidized beds was found to be independent of the three-phase modified Reynolds number. The magnitude of the three-phase Stanton number depended on the superficial liquid velocity and decreased with liquid velocity. A comparison of modified Stanton number in three-phase systems with that in liquid fluidized beds (Table 3) implies that both are of almost identical magnitude. This may be due to the offsetting effects of an increase in heat transfer coefficient and a decrease in liquid holdup. The relationship found between the modified Stanton number and the modified Reynolds number in liquid fluidized beds (Chiu, 1982) may be extended to three phases. Thus,

$$Pr^{2/3} St_{m3} = St_{m2} Pr^{2/3} = C Re_{m2}^{-0.305} \quad (9)$$

and additionally,

$$\frac{h_3}{h_2} = \frac{\epsilon_{L2}}{\epsilon_{L3}} = \left(\frac{\epsilon_{L3}}{\epsilon_{L2}} \right)^{-1} \quad (10)$$

for various gas velocities within current study ranges at given liquid velocity, where St_{m2} and ϵ_{L2} are the modified Stanton number and liquid holdup, respectively in liquid fluidized beds. The value of C is found to be independent of particle size for cylindrical γ -alumina pellets and is equal to 0.1238. However, it is equal to 0.1038 and 0.088 for 0.1- and 0.3-cm glass beads, respectively. The data of Wasmund and Smith (1967) and Simpson (1973) on heat transfer in liquid fluidized beds were also correlated satisfactorily by Eq. 9 (Chiu, 1982). The plots of $Pr^{2/3} St_{m3}$ vs. Re_{m2} with slope (-0.305) are presented in Figure 10.

Equation 9 implies that the relative increase in heat transfer coefficient is equal to the relative decrease in liquid holdup, when comparing three-phase to two-phase fluidized beds. If u_g is zero, ϵ_{L3} is equal to ϵ_{L2} and St_{m3} is identical with St_{m2} . It is known that gas holdup increases with increasing gas velocity at the expense of liquid holdup in a three-phase fluidized bed. Hence, heat transfer coefficient is expected to increase with increasing gas velocity as a result of the decrease of liquid holdup. This is in agreement with observations of heat transfer. It was observed that the decrease in liquid holdup is smaller than the simultaneous increase in gas holdup for beds of large particles. This results in an increase in bed volume. However, for beds of small particles, the decrease in liquid holdup is greater than the simultaneous increase in gas holdup. This corresponds to a bed contraction.

The increase in heat transfer coefficient is smaller for beds of large particles due to the smaller decrease in liquid holdup and vice

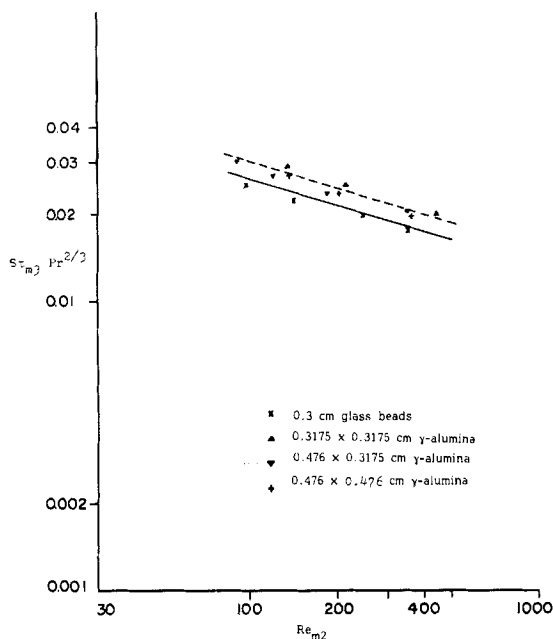


Figure 10. Correlation of $Pr^{2/3} St_{m3}$ with Re_{m2} .

versa (Chiu, 1982). This is in agreement with observed heat transfer behavior. Also, the liquid holdup increases with liquid velocity. Thus, the increase in heat transfer coefficient should be proportionally smaller (as a result of the decreases in the $\epsilon_{L2}/\epsilon_{L3}$ value) with increasing liquid velocity. Again, this is in agreement with the observed behavior of the coefficients. If the fluid velocities and properties, particle properties, and values of $\epsilon_{L2}/\epsilon_{L3}$ or ϵ_{L3} are known, heat transfer coefficients in a three-phase fluidized bed can be estimated using Eq. 9. The possible procedures are as follows:

- (1) Calculate ϵ_{L2} using experimental data or the correlation of Richardson and Zaki (1954).
- (2) Calculate h_2 using Eq. 9 for two-phase operation.
- (3) Calculate ϵ_{L3} or $\epsilon_{L2}/\epsilon_{L3}$ using either of two correlations proposed by Chiu (1982).
- (4) Calculate h_3 via Eq. 9 and various combinations of (1) through (3) above.

The more general equation of the two relations may be used to calculate the ratio $\epsilon_{L2}/\epsilon_{L3}$ for all particle sizes,

$$\frac{\epsilon_{L2}}{\epsilon_{L3}} = 1.184 u_L^{-0.088} u_g^{0.059} (d_p \phi_s)^{-0.058}$$

For large particles of all shapes studied, the average error of prediction is about 3.5%, based on a total of 102 data points and use of method (3) to calculate $\epsilon_{L2}/\epsilon_{L3}$ values.

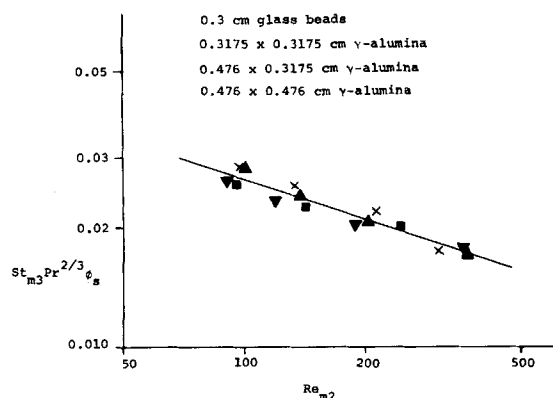


Figure 11. Generalized correlation of heat transfer coefficients in three-phase fluidized beds with large particles.

TABLE 4. COMPARISON OF ESTIMATED AND EXPERIMENTAL VALUES OF $St_{m3} Pr^{2/3} \phi_s$ FOR LARGE PARTICLES

Particle Size Dia. or Dia. X Length, cm	Liquid Velocity cm/s	Esti- mated	Experi- mental	Estimated/ Experimental
0.3 Glass Beads	7.6	0.0268	0.0255	1.052
	9.6	0.02357	0.02248	1.0485
	12.8	0.01964	0.0199	0.987
	15.0	0.01743	0.01767	0.987
0.3175 X 0.3175 γ-Alumina	6.3	0.0268	0.02853	0.939
	7.6	0.024	0.02584	0.93
	9.6	0.0206	0.02216	0.93
	12.8	0.018276	0.01752	1.043
0.476 X 0.3175 γ-Alumina	6.3	0.02745	0.0264	1.04
	7.6	0.025	0.0233	1.073
	9.6	0.0216	0.0204	1.059
	12.8	0.0175	0.01768	0.989
0.476 X 0.476 γ-Alumina	6.3	0.026265	0.0283	0.93
	7.6	0.02385	0.02398	0.995
	9.6	0.02092	0.02066	1.013
	12.8	0.0173	0.0172	1.006

For large particles, Eq. 11 is proposed for considering the effect of particle shape on the heat transfer coefficients in three-phase fluidized beds.

$$St_{m2} Pr^{2/3} \phi_s = St_{m3} Pr^{2/3} \phi_s = 0.1234 Re_{m2}^{-1/3} \quad (11)$$

Equation 11 had a correlation coefficient of 0.96 for 0.3-cm glass beads and all γ -alumina pellets, Figure 11. Table 4 presents the comparison of experimental values of $St_{m3} Pr^{2/3} \phi_s$ and those estimated using Eq. 11.

Transport Phenomena

If was observed that the intensity of solid particle motion increases when a gas at low velocity is added to a liquid fluidized bed, especially at conditions in which fluidization is not fully developed. The enhanced particle motion is probably accompanied by an increase in liquid turbulence and heat transfer coefficient. At high gas velocities, although the benefits of an increase in bubble rise velocity on bed turbulence are still felt, an opposing influence on bed transport, i.e., mutual hindrance of the bubbles, becomes more significant. Therefore, it is expected that the movement of solid particles as well as liquid turbulence and, hence, heat transport increases more gradually with increasing gas velocity in the high gas velocity regime.

The extent of increase in heat transfer coefficients with increasing gas velocity decreases with increasing particle size for spherical and cylindrical shapes.

The total energy loss from the solid phase probably occurs either by bubble dispersion or by transfer of kinetic energy to the continuous phase. The relative consumption associated with bubble dispersion might increase as particles size increases with a relative decrease in the energy associated with imparting turbulence to the continuous phase. Also the displacement of liquid and particles might increase as particle size decreases, since turbulence is enhanced by the relatively larger coalesced bubbles and wakes. El-Temtamy and Epstein (1978) pointed out that relative wake holdup in three-phase fluidized beds decreases with increasing gas velocity.

For a given liquid velocity, the magnitude of increase in heat transfer coefficients with gas velocity is smaller for 0.3-cm glass beads than those for γ -alumina systems even though particle size of all γ -alumina systems is larger than 0.3 cm. Since this runs counter to the size effect in particles of the same type, the phenomenon is probably due to the difference in particle density and/or particle shape. The γ -alumina systems are all easier to fluidize than are 0.3-cm glass beads (Chiu, 1982).

Introducing gas into a liquid fluidized bed increases the particle movement as well as liquid turbulence and results in increasing heat transfer coefficients. It is, therefore, probable that the presence of

a gas phase in a liquid fluidized bed does not change the mechanism of heat transfer. Also, the liquid turbulence which results from particle movement depends on the solid concentration. The solid concentration effect opposes the influence of turbulence on coefficients in three-phase fluidized beds, the coefficient is expected to pass through a maximum as liquid velocity is progressively increased. The porosity of three-phase fluidized beds is a strong function of liquid velocity rather than gas velocity. This partially accounts for the plot of heat transfer coefficient vs. bed porosity passing through a maximum as the bed porosity is progressively increased. Solid holdup in three-phase fluidized beds starts to decrease at high gas velocities for 0.1-cm glass beads. The effect of decreasing both solid holdups and wakes is to reduce the turbulence at the wall. This may explain why the heat transfer coefficients in three-phase fluidized beds with 0.1-cm glass beads are less than those gas-liquid systems at high gas velocities (Fig. 7). It is interesting to note that the gas velocity at which solid holdup in three-phase fluidized beds with 0.1-cm glass beads starts to decrease, decreases with decreasing liquid velocity. A similar trend was observed by Baker et al. (1978) for three-phase heat transfer. They pointed out that the gas velocity at which heat transfer in three-phase fluidized beds with 0.1-cm glass beads starts to be less than those in gas-liquid systems decreases with decreasing liquid velocity.

As described previously, a different trend was observed in the present work than that of Baker et al. for the heat transfer coefficients in three-phase beds with 0.3-cm glass beads at liquid velocity 9.2 and 12.6 cm/s, possibly because of the different heater geometry. The bed porosity at which maximum heat transfer coefficient occurred in two-phase liquid-solid fluidized beds for 0.3-cm glass beads is about 0.7 in the equipment used for this work. Wasmund's data using a heater similar to that of the present investigation predicts a porosity 0.72 for 0.284-cm glass beads as pointed out by Hamilton (1970). Baker et al. found that bed porosity was about 0.6 at the maximum heat transfer coefficient condition. The heat transfer coefficients of Kato et al. are higher than those of the current study. This results in part from the different data reduction process of heat transfer coefficients (Chiu, 1982).

Wall-to-bed heat transport in a three-phase fluidized bed may be thought to occur through two zones in series. From Figure 8, one may infer that the bed resistance decreases with increasing gas velocity and the wall resistance increases with increasing gas velocity. The decrease of bed resistance is due to the increase of turbulence in the bed when gas phase is admitted. The increase of wall resistance is probably due to the increase of gas-liquid packets which decrease the particle's ability to disperse the layer adjacent to the wall. However, particle motion is enhanced as gas velocity is progressively increased. The increase of heat transfer coefficient in a liquid fluidized bed by addition of a gas phase might result, because the magnitude of decrease of the bed resistance is greater than the magnitude of increase of the wall resistance. The smaller increase of wall resistance at low gas velocities might result from the competitive influences of a situation which increases both gas-liquid packet formation and particle motion. At high gas velocities, the decrease in bed resistance and increase in wall resistance are of comparable magnitude. This is possibly responsible for heat transfer coefficient in a three-phase fluidized bed increasing gradually, Figure 3 or 4, as gas velocity is increased.

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At the time of the final draft of this manuscript, an abstract of the work by Moroyama et al. (1981) was obtained. Although their work is not discussed herein, their operating parameters have been added to Table 1.

NOTATION

A_s	= heat transfer area, cm^2
C_{pL}	= specific heat of fluid, $\text{cal/g}^\circ\text{C}$
D_t	= column diameter, cm
d_o	= equivalent particle diameter, cm
d_p	= particle diameter, cm
h_2	= heat transfer coefficient in liquid-solid systems, $\text{W/m}^2\text{-K}$
h_3	= heat transfer coefficient in three-phase systems, $\text{W/m}^2\text{-K}$
k	= thermal conductivity of liquid, W/m-K
L	= particle length, cm
m	= mass flow rate, g/s
q_w	= heat transfer rate, W or cal/s
r	= radial distance, cm
S	= surface area of particles per unit volume, cm^{-1}
T_B	= bulk bed temperature, $^\circ\text{C}$
T_w	= wall temperature, $^\circ\text{C}$
u_g	= superficial gas velocity, cm/s
u_L	= superficial liquid velocity, cm/s
u_{mf}	= minimum fluidization velocity in liquid fluidized beds, cm/s
ρ_L	= density of liquid, g/cm^3
ρ_e	= effective particle density or wet, g/cm^3
ρ_s	= true particle density, g/cm^3
ρ_d	= dry bulk particle density, g/cm^3
μ_L	= viscosity of liquid, g/cm-s
ϵ_p	= particle porosity
ϵ_3	= bed porosity in three-phase systems
ϵ_{L2}	= liquid holdup in liquid-solid system
ϵ_{L3}	= liquid holdup in three-phase system
ϕ_s	= shape factor

Dimensionless Groups

Pr	= Prandtl number
Re_{m2}	= modified Reynolds number in liquid-solid systems
	$\frac{u_L \rho_L}{S(1 - \epsilon_{L2})\mu_L}$
Re_{m3}	= modified Reynolds number in three-phase systems
	$\frac{u_L \rho_L \epsilon_3}{S(1 - \epsilon_3)\mu_L \epsilon_{L3}}$
St_{m2}	= modified Stanton number in liquid-solid systems,
	$h_2 \epsilon_{L2} / C_{pL} \rho_L \mu_L$
St_{m3}	= modified Stanton number in three-phase systems, defined by
	$h_3 \epsilon_{L3} / C_{pL} \rho_L \mu_L$

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