Heat Transfer Characteristics in Two- and Three-Phase Slurry-Fluidized Beds

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Introduction

Slurry-fluidized bed reactors can be conveniently used for processes where easy addition and removal of catalysts are required with good temperature control. Examples are coal liquefaction processes, hydrodesulfurization of petroleum feed stocks, and microbial fermenters. To design and scale-up such reactors, essential data for hydrodynamic properties such as phase holdups, heat and mass transfer characteristics, and bubble properties are required.

In the past, Vasalos et al. (1980) measured phase holdups in coal slurry-catalyst-nitrogen gas slurry-fluidized beds. Also, Deckwer et al. (1980) have determined the gas phase holdup and heat and mass transfer coefficients in a simulated Fischer-Tropsch synthesis reactor. Recently, Kim and Choi (1984) proposed correlations for phase holdups and friction factors in a coal slurry-gas transport reactor. In three-phase (gas-solid-liquid) fluidized beds, heat transfer characteristics have been determined by Ostergaard (1964), Baker et al. (1978), Kato et al. (1981), Chiu and Ziegler (1983), and Kang et al. (1985).

In this study, the effects of gas and slurry phase velocity, particle size, and slurry viscosity on phase holdups and heat transfer coefficients in two- and three-phase slurry fluidized beds have been determined.

Experimental

The experimental apparatus and procedures were described in detail in a previous publication (Kang et al., 1985). Briefly, a three-phase system consists of air as the gas phase; 1.7, 3.0, 4.0, 6.0, or 8.0 mm glass beads with density of 2.5 g/cm^3 as the solid phase; and as the liquid phase water, kerosene, mineral oil, or the slurry, which is a pulverized bituminous coal (<100 mesh) in mineral oil (1.0, 6.0, and 10 wt.% of coal) or in kerosene (17.5 wt.% of coal).

The liquid phase properties are shown in Table 1. The liquid or slurry and gas phase holdups were determined from the knowledge of pressure drop, bed height, and fluid and solid properties (Kim et al., 1975; Kang et al., 1985).

Results and Discussion

Heat transfer in coal slurry-gas and coal slurry-solid systems

In beds of cocurrent flow of coal slurry and air, heat transfer coefficient increased with gas flow rate and decreased with slurry viscosity at a given slurry velocity (Kang et al., 1985). The heat transfer coefficient is about three to five times that of the bed of liquid flow alone. However, the rate of increase in h at higher gas velocity was insignificant since the major part of gas flow at a high velocity penetrates the liquid column in the form of large bubbles and slugs. These large bubbles and slugs are unable to transfer their energy efficiently into microscale eddies and only induce an overall circulation of liquid phase, thus not effectively improving the heat transfer.

The rate of increase in h with gas velocity was significant at low slurry velocity, but at higher slurry velocity it was insignificant. The same trend was observed in the beds of viscous CMC solutions (Kato et al., 1981; Kang et al., 1985) and of water (Baker et al., 1978; Chiu and Ziegler, 1983). It can perhaps be attributed to the bubble properties. In the lower slurry velocity range, the bubble size may grow with gas velocity, but the bubbles can disintegrate in the turbulence that is generated by the higher slurry flow rate. Thus these small bubbles and eddies can enhance the rate of heat transfer. The heat transfer coefficient decreases with slurry viscosity since the laminar sublayer thickness in turbulent flow increases with liquid viscosity and the energy dissipation by microscale eddies through viscous forces cannot increase without limit (Deckwer, 1980).

In the coal slurry-solid system, the heat transfer coefficient decreased with slurry viscosity, but increased with particle size. The change in the coefficient with slurry flow rate may imply that a maximum value of h exists at the optimum bed porosity

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Table 1. **Properties of Liquids and Slurry**

Liquid Phase	Coal Content wt. %	Density g/cm^3	Viscosity mPa · s	Surface Tension dyn/cm
Water	0	1.00	1.00	72.8
Mineral oil	0	0.83	14.00	32.4
Mineral oil	1	0.84	18.17	30.7
Mineral oil	6	0.86	23.54	29.1
Mineral oil	10	0.87	28.00	29.0
Kerosene	0	0.79	1.25	27.0
Kerosene	17.5	0.82	1.77	27.0

conditions, Figure 1. This maximum heat transfer coefficient with slurry or liquid velocity or phase holdup has been reported in previous studies (Kang et al., 1985; Baker et al, 1978; Chiu and Ziegler; 1983). In the present system, the maximum value of h is obtained at the bed porosity of around 0.6–0.7 in beds of different solid sizes. In general, the bed porosity at which the maximum h occurred decreased with particle size and increased with slurry viscosity. This may be due to the erosion of the thermal boundary layer and also due to the particle movement that is retarded by the increase of bed porosity or decrease of bed density.

The slurry holdup at which the maximum h occurred coincides with the maximum energy dissipation rate, as reported by Kang et al. (1985).

Heat transfer in coal slurry-gas-solid systems

It is known that when a solid particle is very small and has a specific gravity close to that of the surrounding liquid, its settling rate is minimal and the slurry behaves like a homogeneous liquid. Therefore, if the criteria for forming a homogeneous slurry are satisfied, the present coal slurry-gas-solid system can be regarded as a liquid-gas-solid three-phase system (Kim and Choi, 1984).

The addition of gas flow into the coal slurry-solid beds enhanced the heat transfer between the heater and the fluidizing medium, Figure 2. As can be seen from the figure, h_3/h_2 decreased with particle size and slurry flow rate for a given gas velocity. It can be expected that the increase in turbulence due to gas injection into the coal slurry-solid beds is comparatively smaller in beds of larger particles and of higher slurry flow rate. Also, h decreased with increase in slurry viscosity in beds of different solids. This decrease may be due to the increase in the boundary sublayer thickness of laminar flow around the heater, the decrease in turbulence, and the increase in viscous friction loss between the phases. The laminar sublayer thickness around the heater surface has been expressed (Kay and Nedderman, 1974) as

$$\delta = \frac{5\,\mu_\ell}{(\tfrac{1}{2} f \rho_\ell^2 U_c^2)} \tag{1}$$

in which f and Uc are a friction factor and the liquid circulating velocity, respectively. As can be seen in Eq. 1, the thickness increases with liquid viscosity but decreases with liquid flow rate.

Since the bed porosity is a function of fluid velocities, the heat transfer coefficient in a three-phase slurry bed is plotted as a function of bed porosity, as shown in Figure 3. As can be seen in the figure, the bed porosity at which the maximum h occurred decreased with increasing particle size but increased with slurry viscosity. The variation of h with bed porosity exhibited a maximum value at the bed porosity of 0.55–0.7 in all the experiments conducted. At the optimum bed porosity for the maximum h,

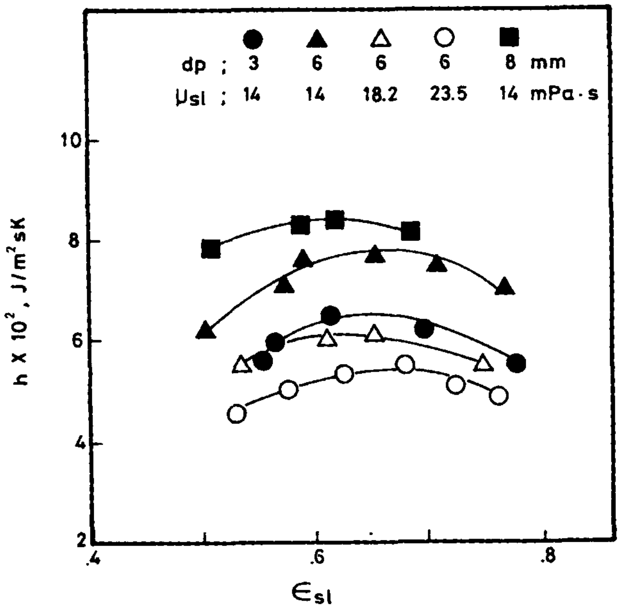

Figure 1. Effect of bed porosity on heat transfer coefficient in coal slurry-solid system.

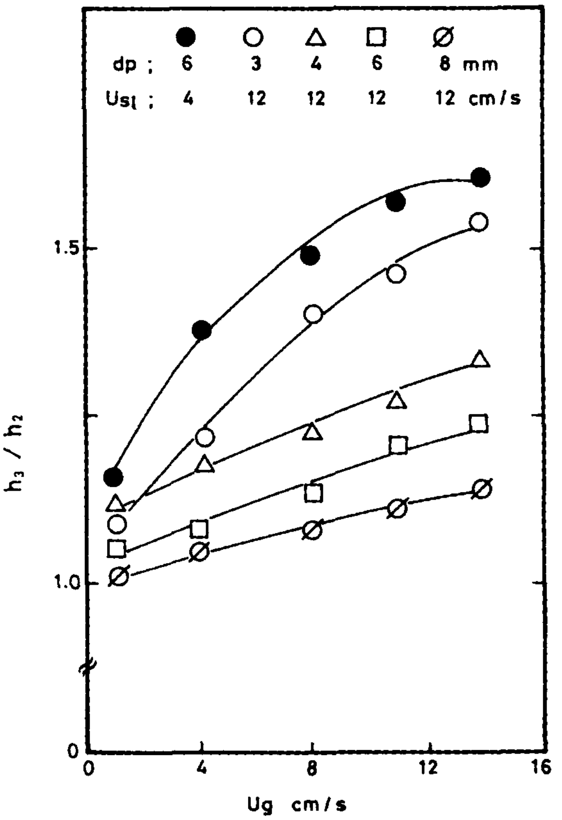

Figure 2. Effect of gas velocity on h_3/h_2 in three-phase slurry-fluidized beds.

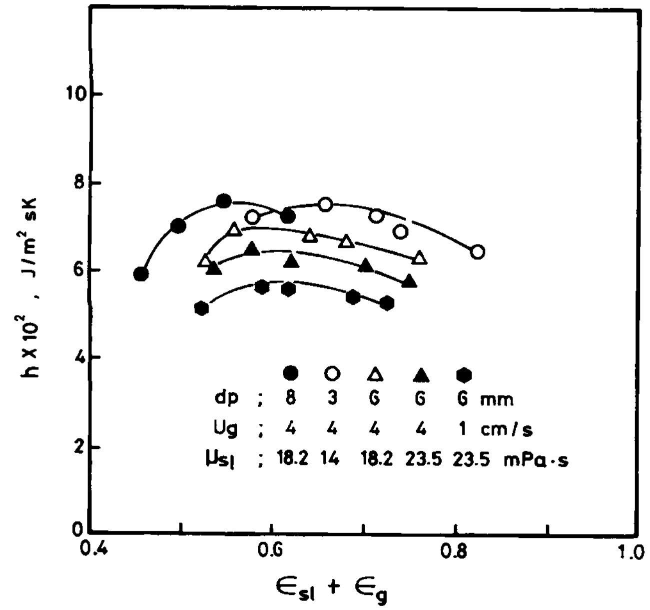

Figure 3. Effect of bed porosity on heat transfer coefficient in three-phase slurry-fluidized beds.

particle motion would change from circulation to random motion with increasing slurry flow rate. Therefore, the increase in h can be caused by the increase in turbulent random motion of individual particles, which may provide more contacting frequencies between the fluid element and the heater surface at the optimum bed porosity condition. This condition may coincide with the maximum energy consumption rate per unit mass of liquid phase.

In general, h increased with particle size in different viscosity solutions. As can be seen in Figure 4, h decreased gradually with slurry viscosity. This decrease may be due to the bubble properties and particle movement caused by variations in slurry viscosity. It has been reported that bubble size and rising velocity increased somewhat with liquid viscosity (Kim et al., 1977) due to the bubble coalescence and the decrease in gas holdup with liquid viscosity (Kim et al., 1975). This may have caused the decrease in h with increase in slurry viscosity.

At a given fluid velocity, the bed porosity increased with fluid viscosity due to an increase in drag on particles. Thus the space between particles will be widened with viscosity. However, the mobility or velocity of particles will be reduced because of the higher frictional losses between the phases. It may reduce h as viscosity increases.

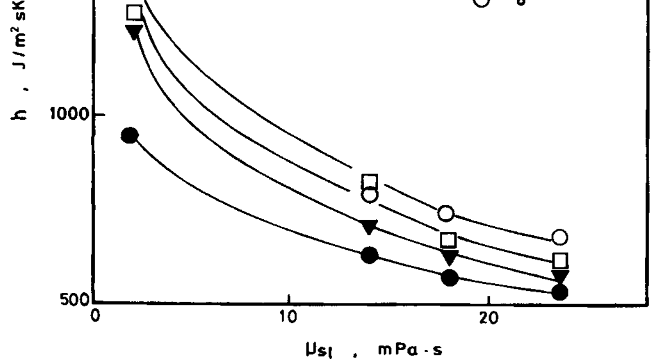

Figure 4. Effect of slurry viscosity on heat transfer coefficient in three-phase slurry-fluidized beds.

Semitheoretical correlations for heat transfer coefficients in two- and three-phase fluidized beds

It may be assumed that unsteady heat diffusion takes place in the liquid element adjacent to the heater wall. The rate of heat transfer is controlled by the rate of renewal of the liquid element, which may depend on the intensity of turbulence. With this assumption, the average heat flux during the contact time of the liquid eddy at the heat exchanger surface can be calculated (Deckwer, 1980). The contact time can also be related to the length and velocity scale of microscale eddies (Deckwer, 1980). The energy dissipation by the microscale eddies is locally isotropic and mainly governed by the viscous forces (Hinze, 1958). Therefore, the length and velocity scale of microscale eddies can be correlated with the kinematic viscosity and the energy dissipation rate per unit mass of fluid since the heat transfer in two- and three-phase fluidized beds may be attributed to an intense mixing of solid phase. However, the thermal conductivity and volumetric heat capacity are much higher for liquids than for gases. Thus the heat transfer by particle convection can be neglected in liquid-solid (Patel and Simpson, 1977; Wasmund and Smith, 1967) and three-phase fluidized beds (Suh et al., 1985).

In three-phase fluidized beds, the continuous splitting and recombination of the liquid stream around the bubbles and particles generates the radial flow of the liquid phase between the heater surface and the bed. The liquid elements will stay for a certain period of time at the surface and then leave it and enter the bulk fluid again. In the liquid elements adjacent to the wall, unsteady heat transfer diffusion may take place. The rate of heat transfer is controlled by the rate of renewal of the liquid element, which may depend on the intensity of turbulence. The details of mathematical derivations have been reported elsewhere (Suh et al., 1985) for three-phase fluidized beds.

The resulting heat transfer coefficient can be expressed as

$$h \propto [k_{sl}\, \rho_{sl}\, Cp_{sl}\, (P_v/\nu)^{1/2}]^{1/2} \qquad (2)$$

The energy dissipation rate per unit mass of liquid phase in three-phase slurry-fluidized beds may be given by the energy input rate (Suh et al., 1985):

$$Pv = [(U_{sl} + U_g)(\epsilon_s\rho_s + \epsilon_g\rho_g + \epsilon_{sl}\rho_{sl}) - U_{sl}\rho_{sl}]g/(\epsilon_{sl}\rho_{sl}) \qquad (3)$$

in which the terms for the solid phase can be neglected for a liquid-gas system and the terms for the gas phase can be neglected for a liquid-solid system. The heat transfer coefficients in two- and three-phase slurry beds can be obtained by substituting Eq.

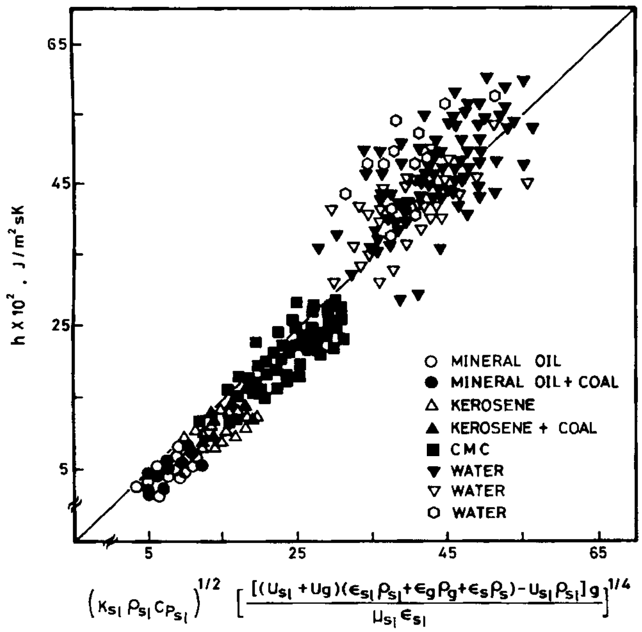

Figure 5. Comparison of observed and calculated values of heat transfer coefficient in three-phase slurry-fluidized beds.

3 into Eq. 2 as

$$h = C(k_{sl}\rho_{sl}\,Cp_{sl}\,\{[(U_{sl} + U_g)(\epsilon_g\rho_g + \epsilon_{sl}\rho_{sl} + \epsilon_s\rho_s) - U_{sl}\rho_{sl}]g/(\epsilon_{sl}\mu_{sl})\}^{1/2})^{1/2} \quad (4)$$

where C is the proportionality constant, which can be determined from the experimental data of h in two- and three-phase fluidized beds.

A multiple regression analysis on the present and previous (Baker et al., 1978; Kato et al., 1981; Kang et al., 1985) experimental data for h in two- and three-phase beds results in a C value of 0.0722 for 910 data points, with a correlation coefficient of 0.96. The goodness of fit between the experimental and calculated values of h in two- and three-phase fluidized beds is shown in Figure 5. The present semitheoretical correlation of h in two- and three-phase fluidized beds covers the range of variables $1.1 < U_{sl} < 12$ cm/s; $0.0 < Ug < 14.0$ cm/s; $0.0 < dp < 0.8$ cm; $0.18 < \epsilon_{sl} < 0.78$; $0.00 < \epsilon_s < 0.62$; $0.92 < Cp_{sl} < 4.18$ J/gK; $0.14 < k_{sl} < 0.59$ W/mK; and $1.0 < \mu_{sl} < 38.9$ mPa · s.

In summary, heat transfer coefficients in two- and three-phase slurry-fluidized beds can be successfully estimated from the knowledge of liquid or slurry properties and energy input rate in the given system based on the surface renewal model with isotropic turbulence theory.

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Notation

A = effective surface area of heater, m^2
Cp = heat capacity, $J/kg \cdot K$
d_p = particle diameter, m
g = gravitational acceleration, m/s^2
h = heat transfer coefficient, $W/m^2 \cdot K$
k = thermal conductivity, W/mK
Pv = energy dissipation rate, J/s
U = superficial velocity, m/s
W = power input to heater, J/s

Greek letters

ρ = density, kg/m^3
ν = kinematic viscosity, m/s
μ = viscosity, $Pa \cdot s$
ϵ = phase holdup
δ = laminar sublayer thickness, m

Subscripts

g = gas
ℓ = liquid
s = solid
sl = slurry
2 = two-phase
3 = three-phase

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