Bed-Immersed Object Heat Transfer in a Three-Phase Fluidized Bed

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Although correlations for the heat transfer coefficient from a vertical cylindrical heater to a three-phase fluidized bed have been developed by several investigators (Baker et al., 1978; Kang et al., 1983; Kato et al., 1984), little is known regarding the mechanism involved. Suh et al. (1985) employed Deckwer's (1980) mechanism of a gas-liquid bubble column with modification on the rate of total energy input to account for the heat transfer in three-phase fluidized beds. Although Suh et al. successfully correlated experimental data based on this mechanism, the physical implication governing the heat transfer processes in a three-phase fluidized bed remains to be examined. In addition, most heat transfer studies in three-phase fluidized beds have been limited to low gas holdup conditions. This study intends to analyze the mechanism of heat transfer from a submerged heater to a three-phase fluidized bed under high gas holdup conditions typical of ebullated bed reactors used in the hydrotreating of petroleum resids.

Experimental Method

Experiments were conducted in Plexiglas columns of 7.62 and 15.2 cm ID in which a cylindrical heater was coaxially located. Diagrams of the apparatus and the heater are presented in Figure 1. Air was used as the gas phase, while either water or a 0.5% tert-pentyl alcohol (TPA) aqueous solution was the liquid phase. Four different cylindrical catalyst particles were used in these experiments. Their physical properties are presented in Table 1.

Two chromel-alumel thermocouples were located along the column axis at 4 and 24 cm above the distributor; their measurements were averaged to give the bed bulk temperature T_b. Three thermocouples were mounted in 1.4 mm wells at 3 cm intervals in the copper section of the heater. These measurements were found to be within ±5% deviation and were averaged to give the heater surface temperature T_h. The heater was connected to a wattmeter and a voltage regulator to measure the steady state heat transfer rate Q. Heat transfer coefficients were calculated from

$$h = Q/A(T_h - T_b) \tag{1}$$

where A, the area for heat transfer, is 0.00397 m^2.

Results

Phase holdups in surfactant systems under the same operating conditions as the present experiments have been reported by Song et al., (1987). Typical variations in the heat transfer coefficient with the liquid velocity for surfactant three-phase fluidized systems are shown in Figure 2. As the liquid velocity increased, the heat transfer coefficient was found to increase sharply, reach a maximum, then decrease. This behavior is consistent with that for wall-to-bed or immersed surface-to-bed heat transfer previously reported (Kato et al., 1984; Kim et al., 1986; Chiu and Zeigler, 1983, 1985). An increase in gas flow at a given liquid flow rate increases h for each type of particle in both pure water and surfactant fluidized systems, as shown in Figure 3a. However, at high gas velocities, little effect is seen on h with further increases in U_g. A comparison of h vs. $u.c._\ell$ at a given gas velocity for pure water and surfactant systems is also shown, in Figure 3b. In Figure 3, h is considerably higher in the surfactant system than that in the pure water system for each type of particle studied.

The variation in h with bed voidage ($\epsilon_g + \epsilon_\ell$) shown in Figure 4 is similar to that shown in Figure 2 for the variation in h with the liquid velocity. In the small column, the maximum h occurred at a bed voidage of 0.7 to 0.75, while in the large column the maximum h occurred at bed voidages up to 0.9. Maximum heat transfer coefficients have been reported to occur at bed voidages varying from 0.6 for immersed surface-to-bed heat transfer (Kang et al., 1985; Kim et al., 1986) to about 0.8 for

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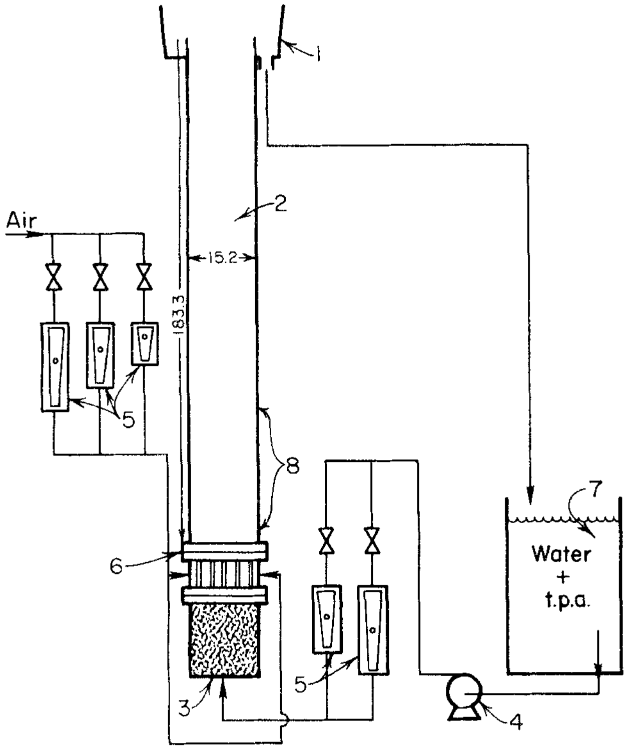

Figure 1a. Experimental apparatus.

Dimensions in cm

1. Disengagement section
2. Test section
3. Calming section
4. Liquid Pump
5. Rotameters
6. Distributor
7. Liquid reservoir
8. Thermocouples

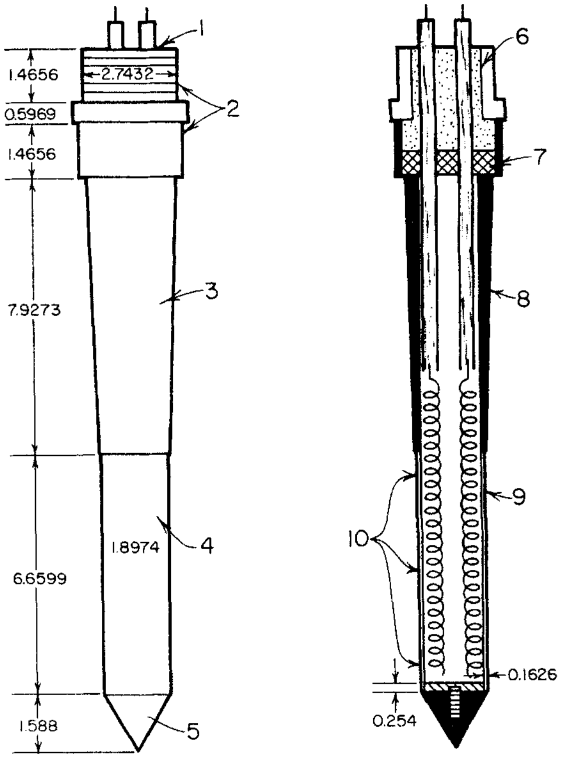

Figure 1b. Heating element.

Dimensions in cm

1. Connecting wire
2. Waterproof fitting
3. Insulated section
4. Heating section
5. Teflon cone
6. Insulating material
7. Ceramic cap
8. Resistances
9. Copper shell
10. Thermocouples

wall-to-bed heat transfer (Chiu and Ziegler, 1983, 1985) under low gas holdup conditions.

The heat transfer coefficient increases with the particle size within the range of the present study. No significant effect of column size on h is evident, similar to the results of Kato et al. (1984) for a nonsurfactant system.

Mechanism of Heat Transfer in Three-Phase Fluidized Beds

Surface renewal by particle movement

Based on Higbie's surface renewal concept and Kolmogoroff's theory of isotropic turbulence, Deckwer (1980) attributed the heat transfer in a gas-liquid bubble column to surface renewal by fluid microeddies:

$$h = c_1(k_\ell \rho_\ell C_{p\ell} S_e)^{1/2} = c_1[k_\ell \rho_\ell C_{p\ell}(P_v/\nu)^{1/2}]^{1/2} \quad (2)$$

In a three-phase fluidized bed system particle movement may play as important a role for the surface renewal as the contribution of the fluid eddies. When a particle moves in the fluidized bed, an adjacent liquid boundary layer also moves along with the particle. As the particle moves from the bulk phase to the vicinity of an immersed object surface, the liquid originally adjacent to the surface of the object is displaced by the particle and its associated liquid layer and is subject to a sudden change in temperature. In order to account for this independent renewal contribution of the particle movement, the heat transfer coefficient of a three-phase fluidized bed is modified by an additional term to Eq. 2 to yield

$$h = c_1(k_\ell \rho_\ell C_{p\ell} S_e)^{1/2} + c_2(k_\ell \rho_\ell C_{p\ell} S_p)^{1/2} \quad (3)$$

A similar concept of particle movement promoting heat transfer in liquid-solid fluidized beds was also proposed by Richardson et al. (1976); however, the particle movement contribution was not theoretically quantified. It should be noted that the second term in Eq. 3 cannot be regarded as mere particle convection. In fact, the particles are not directly involved and the ultimate heat transfer is still through the liquid phase, as only the liquid thermophysical properties appear in that term. It should also be noted that the particle renewal mechanism described here may not be applicable to a system with very fine particles, such as slurry bubble column systems, because the suspended fine particles may not have enough inertia to induce major particle collision effects. Deckwer et al. (1980) have indicated that Eq. 2 with the effective thermophysical properties of

Table 1. Physical Properties of the Solid Particles

	Type of Particle			
	B1	B2	A1	A2
Length, mm	2.941	5.187	2.982	4.079
Diam., mm	0.9398	0.9398	0.899	0.899
Wet density, kg/m^3	2,000	2,000	1,890	1,890

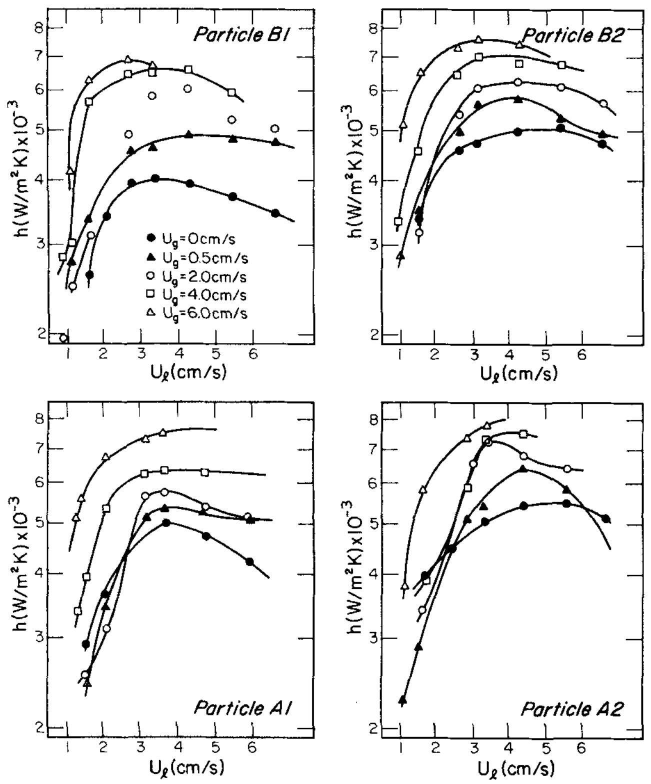

Figure 2. Effects of liquid velocity on heat transfer coefficient.

Surfactant fluidized system using a 7.62 cm ID column

the slurry in place of those of pure liquid could predict the heat transfer in slurry bubble columns fairly well.

The particle renewal frequency in Eq. 3 can be assumed to be proportional to a particle characteristic speed and inversely proportional to a particle characteristic length:

$$S_p \propto U_s/\ell_s \tag{4}$$

The particle's characteristic length can be expressed by

$$\ell_s \propto \phi_s d_{eq}/\epsilon_s^{1/3} \tag{5}$$

The particle average speed is assumed proportional to the liquid velocity in excess of the minimum fluidization velocity similar to that in the liquid-solid fluidized bed (Fan et al., 1985):

$$U_s \propto U_\ell - U_{\ell mf} \tag{6}$$

$U_{\ell mf}$ can be estimated from equations available in the literature (Song et al., 1987). Combining Eqs. 4, 5, and 6 yields the particle renewal frequency as:

$$S_p \propto \epsilon_s^{1/3}(U_\ell - U_{\ell mf})/(\phi_s d_{eq}) \tag{7}$$

Energy dissipation rate

Suh et al. (1985) assumed that energy dissipation in three-phase fluidized beds is due to the liquid phase alone in the form of microeddies. However, the presence of particles may consume energy in two different ways without involving fluid microeddies. First, particles move around in the fluidized bed and cause collisions to occur. As long as the collisions are not perfectly elastic, energy dissipation will always occur. Second, part of the input energy is also consumed in supporting the particles in the bed. In order to generate the drag force to counterbalance the buoyant weight of the particles, the liquid boundary layer prior to the boundary layer separation dissipates energy laminarly, which does not involve microeddies.

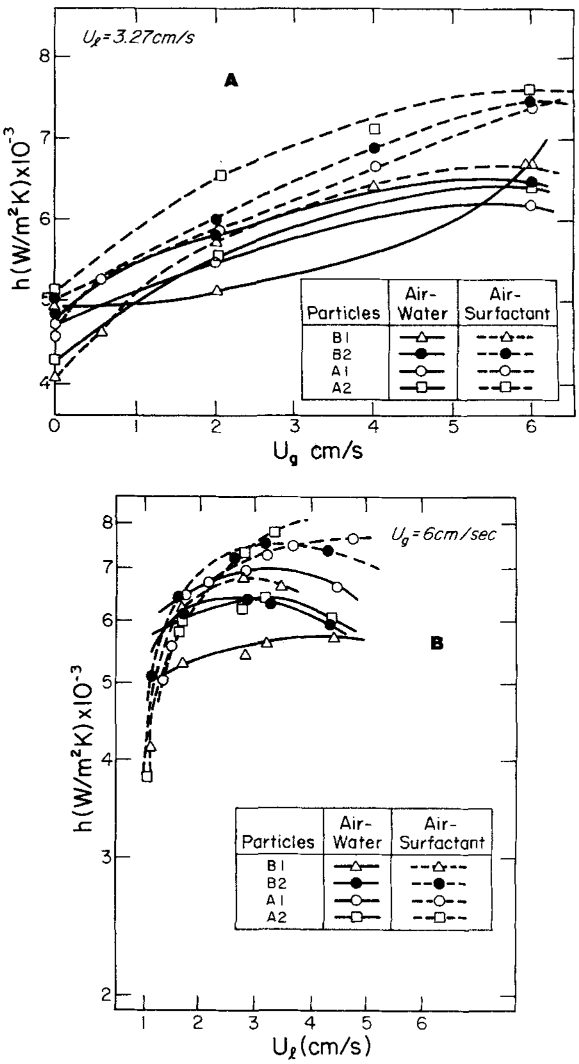

Figure 3. Comparison of effects on heat transfer coefficient between pure water and surfactant fluidized systems.

7.62 cm ID column

(a) Gas velocity effects; (b) Liquid velocity effects

To the first approximation, it is assumed that the energy input is dissipated uniformly by the mass of the system, regardless of phase. This leads to the energy dissipation rate per unit mass of the three-phase fluidized system as

$$P_v = \frac{[(U_\ell + U_g)(\epsilon_s\rho_s + \epsilon_\ell\rho_\ell + \epsilon_g\rho_g) - U_\ell\rho_\ell]g}{(\epsilon_s\rho_s + \epsilon_\ell\rho_\ell)} \tag{8}$$

The inclusion of the particle contribution to the energy dissipation in Eq. 8 is consistent with the inclusion of the particle con-

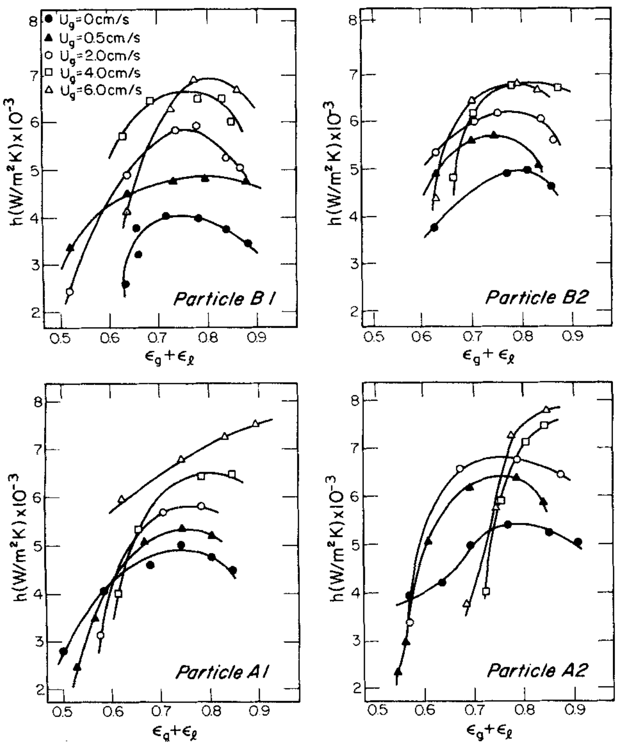

Figure 4. Variation of heat transfer coefficient with bed voidage.

Surfactant fluidized system using a 7.62 cm ID column

tribution to the surface renewal frequency in Eq. 3 since, as noted earlier, the particle surface renewal results from collisions followed by energy dissipation.

The present mechanism, Eqs. 3 and 8, can also be justified by the following experimental results. Consider the conditions in a fluidized body when gas and liquid are maintained at constant flow rates and the particle size is increased. The fluidized bed becomes denser with greater solid holdup and less liquid holdup. If the mechanism of Suh et al. (1985) is employed, one would expect that the heat transfer rate would increase exponentially by simply increasing the particle size. However, from the experimental results of Baker et al. (1978), Chiu and Ziegler (1983), and Kang et al. (1985), it is noted that the heat transfer coefficient in a three-phase fluidized bed either exhibits a minimum or increases initially and then levels off as particle size increases. This is due to the fact that most of the additional energy put into a denser bed with larger particles is consumed in supporting the particles, which does little to promote heat transfer. These experimental results can be correctly accounted for by the present mechanism of Eqs. 3 and 8.

Combining Eqs. 3, 7, and 8, the heat transfer coefficient in the three-phase fluidized bed becomes

$$h = c_1[\kappa_\ell \rho_\ell C_{p\ell}(P_v/\nu)^{1/2}]^{1/2} + c_2\{\kappa_\ell \rho_\ell C_{p\ell}[\epsilon_s^{1/3}(U_\ell - U_{\ell mf})/(\phi_s d_{eq})]\}^{1/2} \quad (9)$$

Correlation

The constant c_1 in Eq. 9 is set as 0.1 such that the equation reduces to that of Deckwer (1980) under the limiting condition of no particle content, i.e., gas-liquid bubble columns. A regression analysis of the present experimental data resulted in a c_2 value of 0.285. The correlation coefficient is 0.90 and the standard deviation is 568 $W/m^2 \cdot K$, which is about 11.3%. It is noted that both terms on the righthand side of Eq. 9 are of the same order of magnitude. This supports the original concept that the contributions of both the liquid eddies and particle movement to the heat transfer mechanism in three-phase fluidized beds are important.

Comparisons of the experimental data of Kang et al. (1985) for an air-water fluidized system using glass beads of 1.0, 1.7, 2.3, 3.0, and 6.0 mm dia. with the predictions of the same correlation were also made. It is noted that only the data of the 6 mm glass beads are close to the predictions, while data of other particles are overpredicted by about 100%. This may be due to the fact that the air-water system using 6 mm glass beads was primarily operated in the dispersed bubble regime (Fan et al., 1986) with small bubbles, similar to the condition in the surfactant system, while experiments using smaller particles primarily operated in the coalesced bubble regime (Fan et al., 1986) with much larger bubbles. Kang et al.'s data using glass beads of 1.0, 1.7, 2.3, and 3.0 mm dia. must be correlated by an entirely separate equation of the form of Eq. 9 with a different c_1 value of 0.071 and c_2 value of 0.01. The correlation coefficient is 0.88 and the standard deviation is 7.7%. The lower value of c_1 for the coalesced bubble regime indicates that the fluid eddy formation is not very efficient in a system of large bubbles, and it was in fact originally indicated by Deckwer (1980) that a c_1 value of 0.1 should only apply to systems with dispersed small bubbles. The lower value of c_2 indicates that the contribution of particle movement is much lower in the coalesced bubble regime. This may be due to the fact that in a coalesced bubble regime the bubble wake controls the particle movement in an up-and-down circulation, which may sharply reduce the radial motion essential to the collision between the particles and the heater.

Comparisons of the present experimental data with predictions based on the equations of Suh et al. (1985) and Kim et al. (1986) were also made. Consistent underpredictions of about 100% by both equations were found. This result is not surprising as most of the data correlated by these authors were from fluidized beds operated in the coalesced bubble regime, in which the heat transfer rates tend to be much lower than those in surfactant systems or systems operated in the dispersed bubble regime.

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Notation

A = heat transfer area of heater
c_1, c_2 = constants, Eq. 3
$C_{p\ell}$ = specific heat of liquid
d_{eq} = equivalent hydraulic diameter of particle
g = gravitational acceleration
h = heat transfer coefficient
$u.c._v$ = energy dissipation rate per unit mass of system
Q = heat transfer rate
S_e, S_p = renewal frequencies due to fluid eddies, particles
T_b, T_h = average bed temperature, heater surface temperature
U_g, U_ℓ = superficial velocities of gas, liquid

$U_{\ell mf}$ = minimum fluidization velocity
U_s = particle characteristic speed

Greek letters

κ_ℓ = liquid thermal conductivity
ℓ_s = particle's characteristic length
ϕ_s = particle sphericity
$\epsilon_g, \epsilon_\ell, \epsilon_s$ = gas, liquid, solid holdups
ν = kinematic viscosity of liquid
$\rho_g, \rho_\ell, \rho_s$ = gas, liquid, solid densities

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