Heat Transfer in Three-Phase Fluidization and Bubble-Columns with High Gas Holdups

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Bubble column and three-phase fluidized bed reactors have wide applications in biotechnological and petroleum processes (Deckwer, 1985; Fan, 1989). In such biotechnological processes as fermentation and waste water treatment, small bubbles of oxygen and/or nitrogen are introduced in the column to enhance oxygen transfer and to ensure the stability of immobilized cell particles. In addition, tiny bubbles are formed by such gases as CO_2, H_2, and CH_4, which are produced during the biological process due to the production of surface active compounds. The presence of these small bubbles causes an increase in the gas holdup of the system. High gas holdups are also characteristics of industrial processes such as coal liquefaction and hydrotreating of residual oils. Tarmy et al. (1984) studied high gas holdup behavior in pilot- and large-scale coal liquefaction reactors and reported gas holdup as high as 50 vol %. Such high gas holdups were attributed to the formation of small rigid bubbles with noncoalescing tendencies due to high-pressure and high-temperature operations. Blum and Toman (1977) reported a maximum gas holdup of 0.5 in a high-temperature (100 to 350°C, 6.9 MPa) three-phase fluidized bed methanator. Thus, good understanding of the transport properties of three-phase fluidized beds with high gas holdups is essential to the design, control and optimum operations of the commercial reactors employed in the above-mentioned processes.

High gas holdups in laboratory scale under ambient conditions can be obtained by the addition of surfactants in air-water and air-water-solid systems. Such high gas holdups in the presence of a surfactant simulate the conditions commonly encountered in coal liquefaction and petroleum resids hydrotreating operations. Levich (1962) reported that the presence of surfactants induces a noncoalescing tendency in the gas-liquid flow and reduces the bubble rise velocity significantly by creating surface tension gradients that cause tangential stresses along the bubble surface. The addition of surfactants to pure water systems, however, increases the system complexity by altering the average bubble size, gas holdup, and bubble-particle interactions. This causes the flow regime transition which affects the heat-transfer behavior of the system. In addition to the dependence on liquid properties, operating conditions for flow regime transition also depend on the gas distributor design, particle properties, and the column size.

Several researchers (such as Kelkar et al., 1983; Shah et al., 1985) reported the effects of surfactants on bubble column hydrodynamics. Kelkar et al. (1983) observed that the gas holdup in a bubble column with an aqueous alcohol solution decreased with increasing liquid velocity, but the effect diminished at high liquid velocities. The hydrodynamics of three-phase fluidized beds under high gas holdup conditions have been reported by Blum and Toman (1977), Tarmy et al. (1984), Morooka et al. (1986), Fan et al. (1987), Song et al. (1989), and Gorowara and Fan (1990). In annular three-phase fluidized beds, Fan et al. (1987) observed a higher gas holdup for *t*-pentanol than *n*-butanol, and found the surface tension and concentration effects to be insignificant. Song et al. (1989) continued the work of Fan et al. (1987) and studied the hydrodynamic characteristics of a simulated resids hydrotreating reactor containing cylindrical hydrotreating catalysts in an air-water (0.5% *t*-pentanol) system. In a slurry system using a 100 ppm solution of Triton X-405 surfactant, Morooka et al. (1986) observed a near linear increase in the gas holdup by increasing the gas velocity, reaching a maximum value of 0.7 at $U_g = 5$ cm/s, indicating homogeneous bubble flow. Beyond this limiting gas velocity, the holdup leveled off or even decreased.

Heat-transfer studies in three-phase fluidized beds have been reviewed recently by Kim and Laurent (1991). Past studies focused primarily on the measurements of time-averaged heat transfer from the column wall to bed (Chiu and Ziegler 1983; Muroyama et al., 1986) or on immersed heating objects to bed (Baker et al., 1978; Kato et al., 1984) in aqueous systems. Recently, Kumar et al. (1992) provided a mechanistic understanding of the heat transfer in bubbly-liquid and liquid-solid systems. They reported that the heat-transfer enhancement due to the passage of a single gas bubble is caused by the bubble wake which is primarily responsible for the rapid surface renewal of fluid on the heat-transfer surface. The bubble-wake-induced heat-transfer enhancement increases with the bubble size because of the larger bubble wake and stronger vortices. While much progress has been made in the understanding of heat transfer in aqueous systems, precise effects of liquid properties, especially the surface tension on heat transfer, remain largely unexplored. Saberian-Broudjenni et al. (1985) measured the heat-transfer coefficient in a three-phase fluidized

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bed with organic liquids and reported that the heat-transfer coefficient exhibits a negative correlation with the thermal conductivity of the liquid. This was attributed to the difficulty in the independent measurement of the influence of the physical properties, as the viscosity, surface tension, thermal conductivity and heat capacity of these organic liquids are varied simultaneously. Magiliotou et al. (1988) studied bed-immersed object heat transfer in a three-phase fluidized bed containing cylindrical catalyst particles with air and 0.5% *t*-pentanol aqueous solution as the gas and the liquid phases, respectively. They found that under the same operating conditions, the overall immersed object to bed heat-transfer coefficient was higher in surfactant fluidized systems than that in the pure water system. They attributed this to the increase in energy dissipation rate due to liquid microeddies. Since they used relatively large particles, the contribution of particle movement, in addition to the liquid eddies due to the bubble wake on heat-transfer enhancement, was also considered. On the other hand, Kumar et al. (1992) alluded that in three-phase fluidized beds containing relatively small spherical light particles, local time-averaged heat-transfer coefficients are higher in the coalesced flow regime compared to those in the dispersed flow regime due primarily to the enhanced bubble-wake-induced surface renewal caused by larger bubbles and their wake. The heat-transfer area in the study of Kumar et al. (1992) was 25 times smaller than that in the study of Magiliotou et al. (1988), and thus the heat-transfer measurements were local in nature solely dictated by the local hydrodynamics of the bed.

The purpose of this work is to investigate the heat transfer in a three-phase fluidized bed under high gas holdup conditions. The associated hydrodynamic behavior of the system is also studied. The high gas holdup condition is achieved by using a porous gas-liquid distributor with an aqueous solution of 0.1% *t*-pentanol and air as the fluidizing medium. The holdup effects on heat-transfer coefficients in a three-phase fluidized bed under finely dispersed bubble flow conditions, in which solid particles are fluidized by the emulsion of gas and liquid, are discussed on the basis of apparent physical properties of the emulsion phase.

Experimental Apparatus

A schematic diagram of the experimental apparatus is shown in Figure 1. The experiments are performed in a three-dimensional Plexiglas column, 150 cm high and 7.62 cm ID. Liquid enters the column through a packed layer of 6-mm glass beads and premixed with air via a perforated shell and tube distributor, and then redistributed into the bed through a PMMA (poly methyl methacrylate) porous plate with a nominal pore size of 30 μm. To study the effects of distributor design, some experiments are also carried out in the system using a perforated shell and tube distributor with a stainless steel wire cloth (200 mesh) placed above it to provide support for the particles. Air is used as the gas phase, while either water ($\rho_l = 997$ kg/m^3, $\mu_l = 8.93 \times 10^{-4}$ kg/m·s, $\sigma_l = 72.8$ dyne/cm) or a 0.1% *t*-pentanol aqueous solution ($\rho_l = 996$ kg/m^3, $\mu_l = 9.10 \times 10^{-4}$ kg/m·s, $\sigma_l = 58.0$ dyne/cm) is the liquid phase. The fluidized particles used are 2.5-mm Nylon beads ($\rho_p = 1{,}150$ kg/m^3, $\epsilon_{so} = 0.60$, $U_t = 7.37$ cm/s) and 0.76 mm glass beads ($\rho_p = 2{,}500$ kg/m^3, $\epsilon_{so} = 0.63$, $U_t = 11.80$ cm/s).

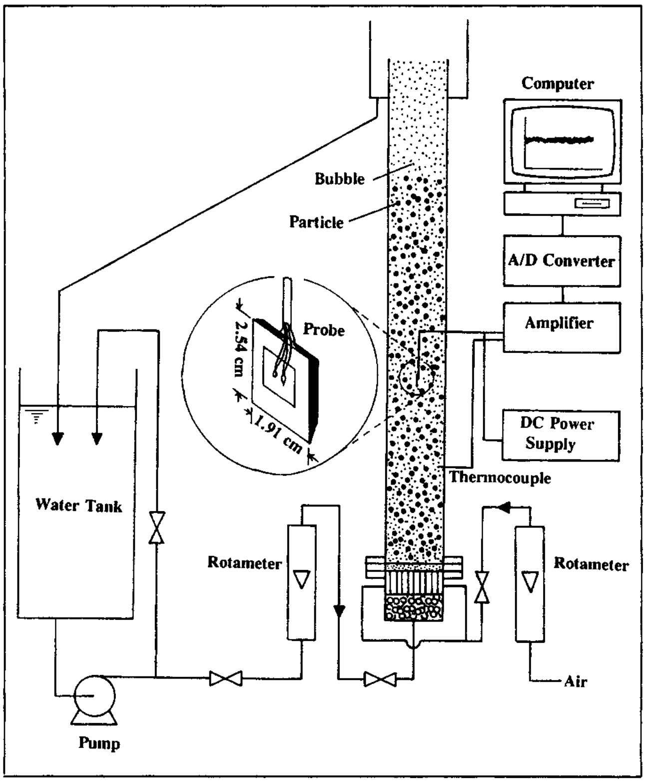

Figure 1. Experimental setup.

A special heat-transfer probe (shown in Figure 1) is employed to measure local heat-transfer coefficients in the bed. Details of the probe design and heat-transfer measurements are given by Kumar et al. (1992). The probe is located along the column axis with the aid of a support 52 cm above the distributor. Since the width of the probe at a horizontal section of the column is small (0.32 cm), the disturbance to the flow field in the vicinity of the probe is minimal and the measured instantaneous heat-transfer rate is a direct measure of the local instantaneous heat-transfer coefficient in the bed at that particular location. The local time-averaged heat-transfer coefficient is obtained by averaging the instantaneous heat-transfer coefficient data as:

$$h_{av} = \frac{1}{n} \sum_{i=1}^{n} \frac{Q}{\Delta T_i} \tag{1}$$

where

$\Delta T_i = (T_{si} - T_\infty)$

n = number of sampling points (2,048)

The overall gas holdup is determined by the quick-closing valve technique, that is, simultaneously cutting off the gas and liquid flows. Gorowara and Fan (1990) reported that for air-surfactant systems, the gas holdup in the freeboard region of a three-phase fluidized bed is closer to the gas holdup within the bed than that in a bubble column. Thus, in the present study, the overall gas holdup is assumed to be equivalent to that within the bed. The solids holdup of the bed is determined by noting the fluidized bed height visually.

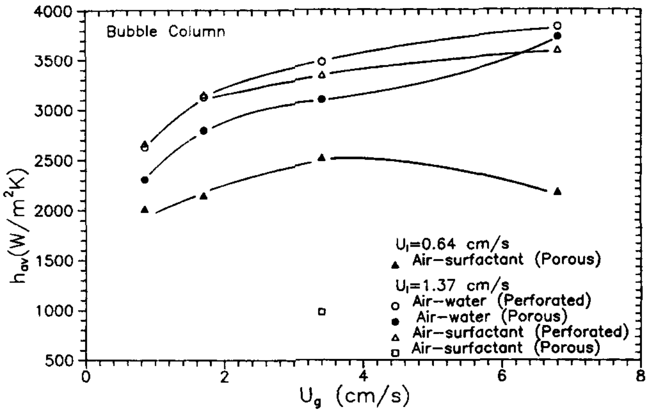

Figure 2. Effect of gas and liquid velocities and type of distributor on the heat-transfer behavior in bubble columns with water and aqueous surfactant solution.

Results and Discussion

Bubble columns

Typical heat-transfer data in bubble columns with air-water and air-surfactant are shown in Figure 2. The overall gas holdup of the system for each run is also studied, but is not shown here. Heat-transfer data in the air-water bubble columns with the perforated distributor are found to be in good agreement with the heat-transfer correlation proposed by Deckwer (1980). In a bubble column with the perforated distributor, coalesced bubble flow conditions prevail in the air-water system, while dispersed large bubble flow conditions prevail in the air-surfactant system. In a bubble column with the porous distributor, however, the flow condition in the air-water system is the dispersed bubble flow at $U_g = 0.85$ cm/s, and a transition from the dispersed to the coalesced flow takes place with an increase in the gas velocity. Under identical operating conditions, the local heat transfer in the coalesced bubble flow is significantly higher than that in the dispersed flow due to enhanced surface renewal caused by larger bubbles and their wake. Thus, the local heat-transfer coefficient in the bubble column (air-water) with the perforated distributor is larger than that with the porous distributor at low gas velocities, but at a high gas velocity ($U_g = 6.8$ cm/s), the heat-transfer coefficient ultimately approaches the value in the system with the porous distributor, signifying that the distributor effects on heat transfer become less pronounced at high gas velocities for the air-water system. The overall gas holdup in the air-water system with the porous distributor is also close to that obtained with the perforated distributor at $U_g = 6.8$ cm/s.

The heat-transfer coefficient in the air-water system is consistently higher than that in the air-surfactant system for $U_g > 1.7$ cm/s due to the presence of larger bubbles and their wake. The difference between heat-transfer coefficients in air-water and air-surfactant bubble columns with the perforated distributor is quite small at low gas velocities as the bubble size is only slightly smaller in the air-surfactant system, but at high gas velocities the difference becomes larger because of dispersed small bubbles in the air-surfactant system due to the coalescence inhibition property of surfactants. This difference, however, is quite significant for bubble columns with the porous distributor as shown in Figure 2, indicating pronounced effects of surfactants on heat transfer in the presence of the porous distributor. Similarly, the gas holdup of the air-surfactant system also shows a significant dependence on the distributor design. At $U_g = 6.8$ cm/s, the gas holdups in the air-surfactant system with the perforated and the porous distributor are 0.20 and 0.60, respectively. These gas holdups are comparable with the data reported by Fan et al. (1987) for the annular bubble column with 0.5 wt. % *t*-pentanol solution.

The effect of gas distributor design on the heat-transfer coefficient was also reported by Yoshitome et al. (1965) in bubble/slurry bubble columns. They found that columns with the perforated distributor have a higher heat-transfer coefficient at high gas velocities ($U_g > 5$ cm/s) than those with the porous distributor. It was also observed that the heat-transfer coefficient was independent of the type of the porous plate used, but depended strongly on the perforated-plate design. In the present study, the heat-transfer coefficient in the bubble column (air-surfactant) with the porous distributor exhibits a maximum value with respect to the gas velocity for a given liquid velocity ($U_l = 0.64$ cm/s). Increasing gas velocity intensifies turbulence, but changes the flow regime. For $U_g > 3.4$ cm/s, there is a significant increase in the gas holdup due to significantly smaller bubbles which causes the heat transfer to decrease. Figure 2 also shows a sharp drop in the local heat-transfer coefficient in the air-surfactant system due to a change in the liquid velocity from 0.64 to 1.37 cm/s. At $U_l = 0.64$ cm/s, bubble swarms are observed which intensify turbulence in the bed resulting in a higher heat transfer. At a higher liquid velocity ($U_l = 1.37$ cm/s), however, the bubbling characteristics become finely dispersed small bubble flow resulting in a sharp drop in the heat transfer.

Three-phase fluidization

To provide a better understanding of the heat transfer in three-phase fluidization with high gas holdups, it is imperative to study the phase holdups. A representative phase-holdup behavior in the presence of the aqueous surfactant solution in a three-phase fluidized bed with the porous distributor is shown in Figure 3. The holdups in beds containing 2.5-mm Nylon beads (NY2500) and 760-μm glass beads (GB760) exhibit similar trend. Gas holdups in the air-surfactant system with the porous distributor are greatly enhanced (shown in Figure 3) by the synergistic effects of the noncoalescing tendency of bubbles in the presence of an aqueous surfactant solution (0.1% *t*-pentanol) and generation of small bubbles from the porous distributor. The gas holdup increases with increasing gas velocity and attains a maximum value of 0.65 at $U_g = 6.8$ cm/s, as shown in Figure 3. The variation of gas holdup with U_g shows the similar behavior as observed in the bubble column. At a low liquid velocity ($U_l = 0.64$ cm/s for NY2500 and $U_l = 1.37$ cm/s for GB760), the gas holdup for the fluidized bed is smaller than that in the bubble column under similar operating conditions as the bed is under incipient fluidization state. Under such conditions, the bed is heterogenous and bubble swarms are periodically observed resulting in the reduced gas holdup. At high liquid velocities ($U_l \geq 1.37$ cm/s for NY2500 and $U_l \geq 2.82$ cm/s for GB760) in a fully fluidized bed, however, a finely dispersed bubble flow condition is ob-

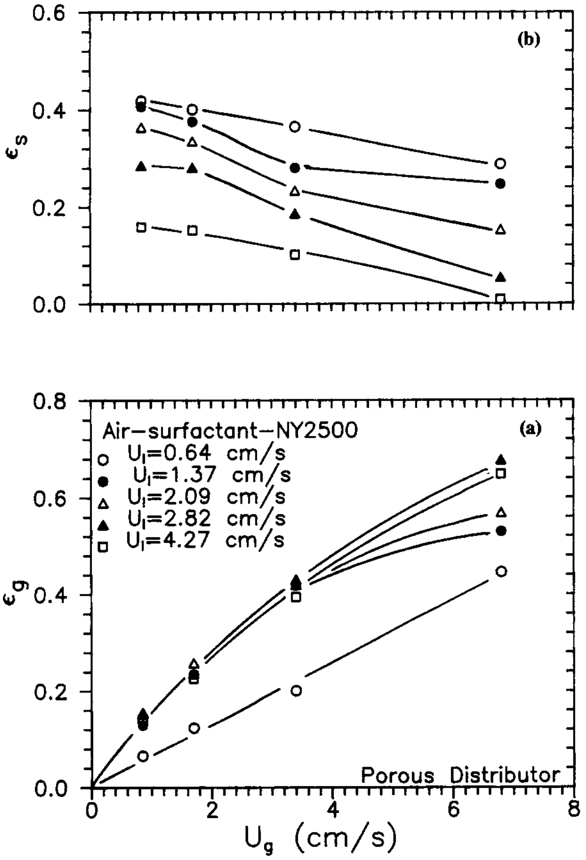

Figure 3. Effect of gas and liquid velocities on (a) gas holdup and (b) solids holdup in an air-surfactant fluidized bed of nylon beads with the porous distributor.

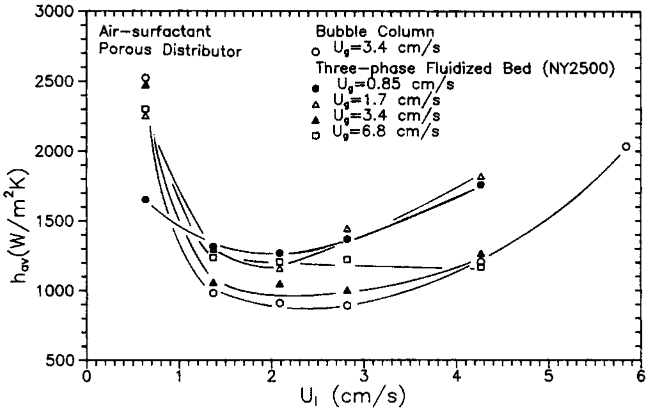

Figure 4. Variation of heat-transfer coefficient with the liquid velocity in an air-surfactant fluidized bed of nylon beads with the porous distributor.

tained and the freeboard region above the bed behaves like a gas-liquid emulsion. A clear boundary between the bed and the freeboard is observed similar to that in a liquid-solid fluidized bed. Similar fluidization characteristics are reported in nitrogen-kerosene-glass beads fluidized beds by Saberian-Broudjenni et al. (1987). They found that when particles of large size are fluidized at high liquid velocities and low gas velocities, a stable emulsion of fine gas bubbles is formed due to the foaming nature of kerosene. Under these conditions, the three-phase fluidized bed behaves like a liquid-solid fluidized bed and can be classified as a pseudo-homogenous fluidized bed (Saberian-Broudjenni et al., 1987). At a low liquid velocity (U_l=0.64 cm/s for NY2500 and U_l=1.37 cm/s for GB760), however, the bed becomes heterogenous and large bubbles are observed periodically. Under fully fluidized conditions in the air-surfactant bed with the porous distributor, the bubbles are finely dispersed and are much smaller than those in the dispersed flow regime in air-water fluidized beds. The bubbles are spherical in shape separated by thick liquid films and the gas-liquid emulsion is milky in appearance and behaves like a pseudo-homogeneous fluid. Song et al. (1989) also reported a pseudo-homogeneous nature of the fluid phase in the air-surfactant fluidized bed. Such a continuous phase is best described as *kugelschaum* or wet foam (Adamson, 1982). Sebba (1971) reported that the dispersion of finely divided *kugelschaum* (average diameter of gas bubbles = 20 μm) behaves like a liquid, but when the gas fraction exceeds 65%, the viscosity rises considerably due to the distortion of spherical gas bubbles. In the present study, the gas holdup may approach the critical fraction at high gas velocities ($U_g \geq 6.8$ cm/s).

In fully fluidized beds, little effect of liquid velocity on the gas holdup is observed in Figure 3 for $U_g < 4$ cm/s. The relatively high solids holdup in the bed at low liquid velocities (U_l = 1.37-2.09 cm/s for NY2500) prevents any significant increase in the gas holdup at a high gas velocity (U_g = 6.8 cm/s). This is verified by calculating the solid-free gas holdups ϵ_g^* [$=\epsilon_g/(\epsilon_g+\epsilon_l)$] and the values lie between 0.6 and 0.7 at U_g = 6.8 cm/s for intermediate liquid velocities. The gas holdup data obtained in the ebullated bed reactor by Tarmy et al. (1984) are comparable to the data reported in the present study. However, the gas holdups in the present study are higher than those reported previously (Fan et al., 1987; Song et al., 1989; Gorowara and Fan, 1990) due to the differences in distributor design. The maximum gas holdup obtained by Song et al. (1989) was 0.3 at U_g = 6 cm/s with the perforated distributor.

Figures 4 and 5 show the variation of heat-transfer coefficients with U_l under finely dispersed bubble flow condition for the air-surfactant fluidized beds of nylon beads and glass beads, respectively. Also shown are the heat-transfer coefficients in the bubble column under similar operating conditions for com-

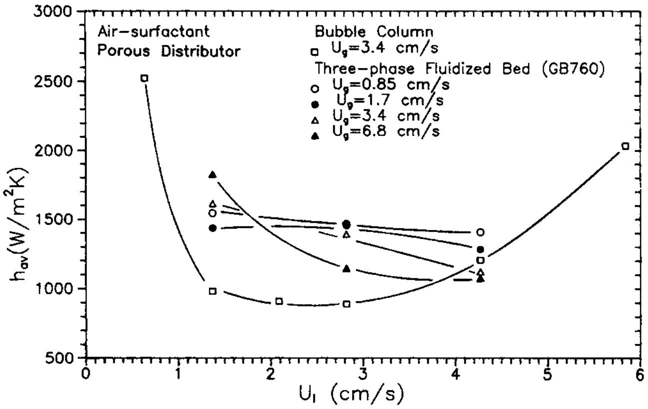

Figure 5. Variation of heat-transfer coefficient with the liquid velocity in an air-surfactant fluidized bed of glass beads with the porous distributor.

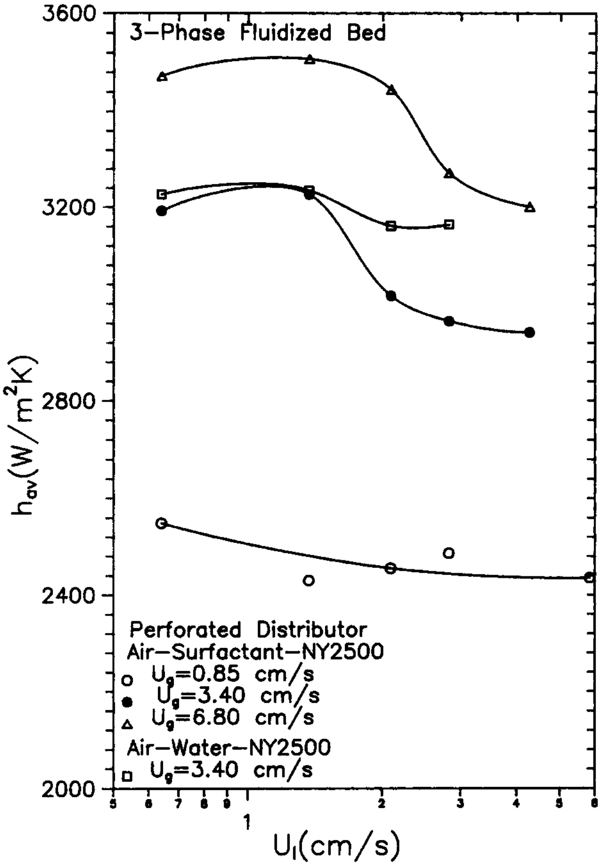

Figure 6. Variation of heat-transfer coefficient with the liquid velocity in an air-surfactant fluidized bed of nylon beads with the perforated distributor.

parison. The air-surfactant fluidized beds of nylon and glass beads with the porous distributor exhibit similar heat-transfer behavior. The heat-transfer coefficient initially decreases with the liquid velocity and then either increases or remains constant with U_l at higher liquid velocities. The initial decrease in the heat-transfer coefficient with increasing liquid velocity corresponds to the transition from incipient to fully fluidized state. Kato and Morooka (1982) reported similar heat-transfer behavior. They found that the heat-transfer coefficient between the wall and a three-phase fluidized bed exhibits a maximum value and then a minimum value as a function of the liquid velocity in the region from the packed bed to the fully fluidized bed. Under the incipient fluidization condition, the bed is heterogenous and is significantly influenced by the internal circulation of liquid induced by the gas flow resulting in higher heat-transfer coefficients. However, as U_l increases, the bed becomes pseudo-homogeneous due to finely dispersed bubbles, and an increase in the liquid velocity intensifies turbulence at low gas velocities resulting in higher heat-transfer coefficients. At $U_g = 6.8$ cm/s for $U_l > 1.37$ cm/s, the heat-transfer coefficient almost decreases with U_l due to the lower apparent thermal conductivity of the gas-liquid mixture as the gas holdup of the system becomes higher than 0.5.

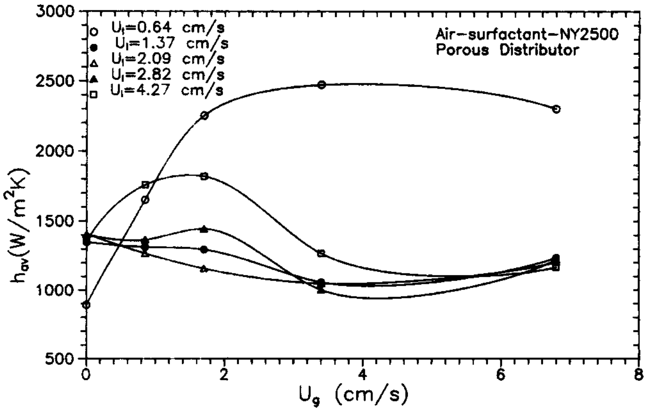

Figure 7. Variation of heat-transfer coefficient with the gas velocity in an air-surfactant fluidized bed of nylon beads with the porous distributor.

To understand distributor effects, heat-transfer data for air-surfactant fluidized beds with the perforated distributor are shown in Figure 6. At high U_l, the heat-transfer coefficient decreases with U_l due to a transition in the flow regime. Fan et al. (1987) reported that at low U_l and $U_g > 3$ cm/s in three-phase fluidized beds with the perforated distributor, the bubble size is a strong function of U_l: as the liquid velocity increases, the bubble size decreases indicating a transition from the dispersed large bubble to the dispersed small bubble regime. However, the bubble size in this system always remains larger than that in the system with the porous distributor. The comparison of Figures 4 and 6 shows that under identical gas and liquid velocities, the heat-transfer coefficient in the air-surfactant fluidized bed with the porous distributor consistently is much smaller than that in the bed with the perforated distributor. Due to relatively large bubbles in the air-surfactant fluidized bed with the perforated distributor, the bubble-wake-induced heat-transfer enhancement is higher. Thus, the distributor design plays an extremely important role in the transport phenomena in air-surfactant systems (bubble columns and three-phase fluidized beds). In air-water systems, bubble coalescence or breakage phenomena dominate, thus system properties associated with the bubble formation, such as gas distributor designs, become less important to the flow regime behavior. However, in air-surfactant systems systems where bubble coalescence is inhibited partially due to the formation of an electrically charged layer at the gas-liquid interface, the gas distributor design becomes more important since the initial bubble size distribution controls the bubble flow pattern and hence the flow regime behavior.

Figure 7 represents the effect of the gas velocity on the heat-transfer coefficient in an air-surfactant fluidized bed with the porous distributor. Under the incipient fluidization condition at $U_l = 0.64$ cm/s, the heat-transfer coefficient increases with the gas velocity and levels off at high gas velocities. An increase in U_g intensifies the liquid-phase turbulence and increases the heat-transfer coefficient, but several investigators observed that the increase tapered off at high U_g (Magiliotou et al., 1988). In the present study, under fully fluidized state the heat-transfer coefficient decreases with increasing gas velocity for $U_g > 2$ cm/s. Under the finely dispersed bubble flow condition at $U_g > 2$ cm/s, the gas fraction in the pseudo-homogeneous gas-liquid mixture significantly rises with the gas velocity; thus, the changes in the apparent thermal conductivity of the gas-liquid mixture must be considered to explain the heat-transfer behavior.

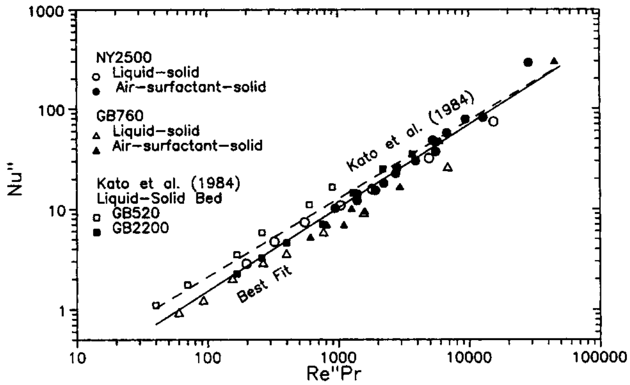

Figure 8. Relationship between *Nu″* and *Re″ Pr* for heat transfer in three-phase fluidized beds under finely dispersed bubble flow conditions.

Several heat-transfer correlations have been proposed to satisfactorily account for the heat-transfer behavior in three-phase fluidized beds. These, however, are obtained under coalesced and dispersed bubble flow conditions consisting of bubbles with the average size of more than several millimeters. The transport mechanism in these flow regimes is generally explained by the turbulence due to the bubble wake or the bubble motion (Fan and Tsuchiya, 1990; Kumar et al., 1992). Since the fluidization condition in this study, as discussed earlier, is quite different from the coalesced and dispersed flow conditions, it is evident that these heat-transfer correlations cannot satisfactorily describe the heat-transfer behavior in the present system. Also, the diverse heat-transfer behavior in different flow regimes in the surfactant system prohibits using a single unified correlation for the heat-transfer prediction in all regimes. The immersed object-to-bed heat-transfer correlations (Baker et al., 1978; Kato et al., 1984) for three-phase fluidized beds with water as the liquid phase may not be satisfactorily used for the heat-transfer prediction under finely dispersed bubble flow conditions commonly encountered in air-surfactant fluidized beds. As the air-surfactant fluidized bed under such flow conditions behaves like a liquid-solid fluidized bed, a correlation for the heat transfer can be obtained based on the similar form of the heat-transfer correlation proposed for liquid-solid fluidized beds. The correlation equation for heat transfer from immersed vertical heater to bed in a liquid-solid fluidized bed is generally of the form:

$$Nu' = \alpha (Re' Pr)^{\beta} \tag{2}$$

where

$$Nu' = h_{av} d_p \epsilon_l / k_l (1 - \epsilon_l)$$

$$Re' Pr = d_p U_l \rho_l C_{pl} / k_l (1 - \epsilon_l)$$

The values of α and β as given by Kato et al. (1984) are $\alpha = 0.058$ and $\beta = 0.78$.

Consider the apparent physical properties of the gas-liquid emulsion as defined below:

$$\rho_{app} = (\epsilon_g \rho_g + \epsilon_l \rho_l)/(\epsilon_g + \epsilon_l) \tag{3}$$

$$k_{app} = (\epsilon_g k_g + \epsilon_l k_l)/(\epsilon_g + \epsilon_l) \tag{4}$$

$$C_{papp} = (\epsilon_g \rho_g C_{pg} + \epsilon_l \rho_l C_{pl})/(\epsilon_g \rho_g + \epsilon_l \rho_l) \tag{5}$$

The Nusselt number and the product of Reynolds and Prandtl numbers for air-surfactant fluidized beds are given by:

$$Nu'' = h_{av} d_p (\epsilon_l + \epsilon_g)/k_{app}(1 - \epsilon_l - \epsilon_g) \tag{6}$$

$$Re'' Pr = d_p (U_l + U_g) \rho_{app} C_{papp}/k_{app}(1 - \epsilon_l - \epsilon_g) \tag{7}$$

Note that the terms in Eqs. 3 through 7 for air-surfactant fluidized beds reduce to the corresponding terms for liquid-solid fluidized beds as U_g becomes zero ($\epsilon_g = 0$). The heat-transfer data in the nondimensional form for air-surfactant fluidized beds under finely dispersed flow conditions are plotted in Figure 8 along with the heat-transfer data for liquid-solid fluidized beds reported by Kato et al. (1984). Also shown in Figure 8 are the liquid (water)-solid fluidized bed heat-transfer data for NY2500 and GB760 obtained in the present study. The heat-transfer correlation proposed by Kato et al. (1984) for a liquid-solid fluidized bed is represented by the dotted line. An excellent agreement was observed between local heat-transfer coefficients data in liquid-solid fluidized beds and the data for air-surfactant fluidized beds under finely dispersed bubble flow conditions. This confirms that for three-phase fluidized beds under high gas holdups, the gas-liquid mixture behaves as a pseudo-homogeneous phase, and by defining the pseudo-homogeneous properties of the mixture, the local heat transfer in the bed can be well predicted by the heat-transfer behavior in a liquid-solid fluidized bed. A regression analysis of all the experimental data is performed, and a better fit as represented by the solid line in the figure can be obtained by using $\alpha = 3.33 \times 10^{-2}$ and $\beta = 0.83$.

Conclusions

Time-averaged local heat-transfer coefficients in gas-liquid and gas-liquid-solid systems with high gas holdups are measured by employing a special heat-transfer probe. High gas holdups up to 70 vol % is obtained in air-surfactant systems (three-phase fluidized beds and bubble columns) with a porous gas-liquid distributor. Local heat transfer in such systems is governed mainly by the complex interplay between holdups and the flow regime which changes with the gas and liquid velocity and the distributor design. The heat transfer is considerably higher in the pure-water system than that in the surfactant system for each type of particles studied. Local heat-transfer data in air-surfactant fluidized beds with the porous distributor are in excellent agreement with the local heat-transfer coefficients in liquid-solid fluidized beds. Due to pseudo-homogeneous nature of the fluid-phase in air-surfactant fluidized beds under high gas holdup conditions, apparent physical properties of the gas-liquid emulsion can be used to represent the local heat-transfer behavior in the bed satisfactorily. The local heat-transfer data in three-phase surfactant fluidized beds under finely dispersed bubble flow conditions are well correlated by a heat-transfer correlation of the form used commonly for liquid-solid fluidized beds. Distributor design significantly affects the flow regime and thus the heat-transfer coefficient in air-surfactant fluidized beds and bubble columns. The heat-transfer coefficient is found to be considerably higher in the dispersed large bubble flow regime with the perforated distributor compared to that in the finely dispersed bubble flow regime with the porous distributor. Heat transfer in the air-surfactant fluidized bed exhibits a maximum with respect to U_g due primarily to changes in the apparent thermal conductivity of the gas-liquid emulsion with increasing gas holdups.

Acknowledgment

This work was supported by NSF Grant No. CTS-9200793.

Notation

C_{papp} = apparent heat capacity of gas-liquid emulsion, J/kg·K
C_{pg} = heat capacity of gas, J/kg·K
C_{pl} = heat capacity of liquid, J/kg·K
d_p = particle diameter, μm
h_{av} = time-averaged experimental local heat-transfer coefficient, W/m^2·K
k_{app} = apparent thermal conductivity of gas-liquid emulsion, W/m·K
k_g = thermal conductivity of gas, W/m·K
k_l = thermal conductivity of liquid, W/m·K
n = number of sampling points
Nu' = Nusselt number defined for liquid-solid fluidized beds
Nu'' = Nusselt number defined for three-phase fluidized beds
Pr = Prandtl number
Q = time-averaged heat flux, W/m^2
Re' = Reynolds number defined for liquid-solid fluidized beds
Re'' = Reynolds number defined for three-phase fluidized beds
T_{si} = instantaneous probe surface temperature, K
T_∞ = bulk temperature, K
U_g = superficial gas velocity, m/s
U_l = superficial liquid velocity, m/s
U_t = terminal velocity of particle in water, cm/s

Greek letters

α, β = coefficients for empirical correlation
ΔT_i = instantaneous temperature difference between probe surface and bulk, K
ϵ_g = gas holdup
ϵ_g^* = solid free gas holdup
ϵ_l = liquid holdup
ϵ_s = solids holdup
ϵ_{so} = packed bed solids fraction
μ_l = liquid viscosity at 25°C, kg/m·s
ρ_{app} = apparent density of gas-liquid emulsion, kg/m^3
ρ_g = gas density, kg/m^3
ρ_l = liquid density, kg/m^3
ρ_p = particle density, kg/m^3
σ_l = liquid surface tension, dynes/cm

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Manuscript received Oct. 19, 1992, and revision received Dec. 28, 1992.