Effect of Low Velocity Liquid Pulsations on Some Hydrodynamic Characteristics of Liquid and Gas-Liquid Fluidized Beds

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Introduction

Externally applied vibrations or pulsations have been extensively studied as a means for improving transfer characteristics in two-phase systems (Baird, 1966; Baird and Garstang, 1972; Kim and Baird, 1976; Battacharya and Harrison, 1976). Similar studies for three-phase systems, e.g., three-phase fluidized beds, have been limited to examining the effect on solid-liquid mass transfer of bubble cycling caused by vertical oscillation of a slurry column containing entrained gas bubbles (Lemcoff and Jameson, 1975). In three-phase fluidization, bubble coalescence predominates over bubble breakup for small-particle beds (e.g., $d_p < 3$ mm for sand fluidized by water and air), so that the desirably high particle surface areas and low fluidizing velocities associated with such beds are accompanied by lower gas holdups and hence inferior gas-liquid volumetric mass transfer coefficients, compared to those for large-particle beds (Epstein, 1981; Deckwer and Schumpe, 1983). It is therefore of some interest to determine whether the undesirable features of small-particle beds are somewhat mitigated by the imposition of deliberately imposed liquid pulsations. In the present preliminary study at low pulse velocities (<5.8 mm/s), measurements were made, with and without pulsing, of phase holdups for liquid-solid, gas-liquid, and gas-liquid-solid systems; of liquid phase axial dispersion for the gas-liquid and gas-liquid-solid systems; and of the hydrodynamic transition from a fixed to a fluidized bed for the liquid-solid system.

Experimental

In the three-phase experiments, closely sized beads of mean diameters 0.5 and 0.9 mm, and particle densities 2,472 and 2,475 kg/m^3, respectively, were fluidized by a cocurrent flow of metered air at approximately 1 atm (101.3 kPa) and metered water at 14°C in a plexiglass column 0.152 m in diameter and 4.5 m high. Thirty pressure taps were used to measure longitudinal pressure profiles by differential manometry, and, as described by El-Temtamy and Epstein (1980), to convert these to phase holdups where the solids were completely fluidized. The two-phase experiments were conducted in a similar manner, without the air flow in the case of the liquid-solid system and without the solids in the case of the gas-liquid system. The superficial velocities of the air and the water were each varied up to about 60 mm/s.

A Milton Roy diaphragm pump with a variable speed motor and an adjustable pulse volume was used to produce liquid pulses in the frequency range 1.67–3.00 Hz and amplitudes up to 1.93 mm. It should be noted that the resulting maximum pulsation intensity of 5.8 mm/s is more than an order of magnitude smaller than that attained in other studies (Lemcoff and Jameson, 1975; Kim and Baird, 1976). The pulsating water line together with the steady state water line entered the column calming section at its base.

A steady state tracer injection technique, similar to that employed by El-Temtamy et al. (1979), was used to study liquid phase axial dispersion, with saturated sodium chloride solution as tracer. Beckman conductivity cells in conjunction with "solumeters" were used to measure the axial concentration profiles upstream of the injection plane. From these profiles the axial dispersion coefficients were evaluated using the one-dimensional dispersion model:

$$C/C_o = \exp\left(-V_\ell z/D_z\right)$$

Results and Discussion

Minimum liquid fluidization

Before any gas was introduced to the system, fluidization of the solids was first achieved by increasing the liquid flow rate. Typical pulsed and unpulsed runs for the 0.9 mm glass beads are shown in Figure 1 as a plot of manometric pressure gradients vs. superficial water velocity. It is seen that the pressure gradient measured for the fixed bed under nonpulsed conditions was markedly increased by pulsing the liquid, with the result that minimum fluidization under pulsed conditions was achieved at a water rate 55% lower than for steady conditions.

A similar decrease in minimum fluidization velocity due to bed vibrations has been recorded for gas-fluidized beds (Bretsnajder et al., 1959). The pulsations or vibrations evidently cause a fixed bed to become more compact, thus increasing the interstitial fluid velocity and frictional pressure drop for a given superficial velocity, which gives rise to earlier fluidization and also accounts for the significantly increased value of solids holdup (i.e., decreased voidage) at minimum fluidization recorded in Figure 1.

Phase holdups

The coincidence of the steady and pulsed flow pressure gradients in Figure 1 after the steady minimum fluidization velocity is exceeded is an indication of the fact that the solids (and therefore the liquid) holdup has become the same for both the steady and pulsed fluidized beds.

Experimentally insignificant effects of liquid pulsing on phase holdups were similarly recorded over the entire range of conditions investigated, both for the gas-liquid system in the absence of solids and for the gas-liquid fluidization of the two solid particle sizes. This result, which is not really surprising in view of the low values of fA used in the present experiments, applies to experiments in which gas holdups were increased up to $\epsilon_g = 0.20$ and solids holdups down to $\epsilon_s = 0.20$ (i.e., $H = 0.6$–2.0 m). Baird and Garstang (1972) similarly reported for bubble columns that little difference in gas holdup between pulsed and steady operation could be observed as long as the bubble velocity was much larger than the liquid pulsation velocity fA, a condition which prevailed in all the present experiments involving gas flow. The observed variations of gas and solid holdups with gas and liquid velocities for the three-phase fluidized beds, and the effect of solids holdup on gas holdup, were similar to those previously reported (Bhatia and Epstein, 1974) for steady operation, and any recorded differences in phase holdups between the steady and the liquid-pulsed beds were within the range of experimental error.

Axial dispersion of liquid

Data for the air-water system at two liquid velocities and a range of gas velocities are shown in Figure 2. At the lower liquid velocity, the liquid pulsations usually produced a significant increase in the axial dispersion coefficient, D_z, while at the higher liquid velocity they yielded a more ambiguous decrease in this coefficient. The effect on D_z of different combinations of frequency and amplitude at a fixed liquid pulsation velocity, fA, and fixed liquid and gas velocities, is shown in Figure 3, along with the corresponding steady flow point. The value of D_z is markedly increased compared to the steady flow value when the lowest frequency pulsation is introduced, but thereafter it falls as f is further increased. A similar trend has been observed by Kim and Baird (1976) for the effect of reciprocation frequency on liquid phase axial dispersion for countercurrent air-water flow through a reciprocating plate column, at constant A (rather

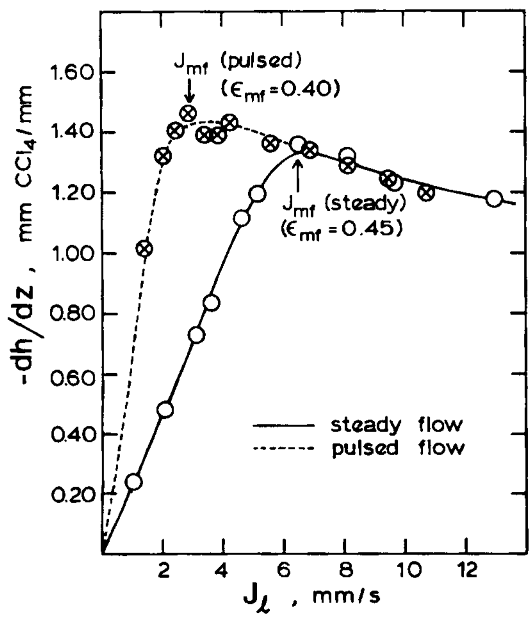

Figure 1. Effect of liquid pulsations on pressure gradient for water fluidization of 0.9 mm glass beads with static bed height H_s = 0.61 m.

O steady liquid flow; ⊗ pulsed liquid flow with f = 3.00 Hz and A = 0.55–1.44 mm.

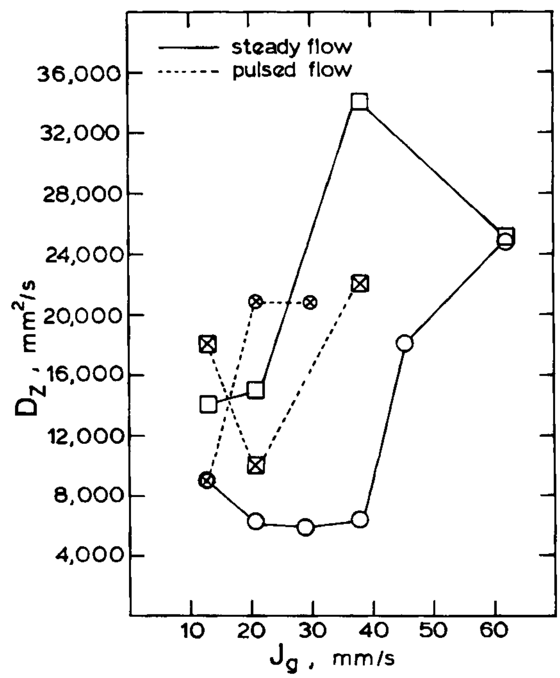

Figure 2. Effect of liquid pulsations on axial dispersion in the liquid phase of an air-water bubble column.

Open points, steady liquid flow; crossed points, pulsed liquid flow with f = 3.00 Hz and A = 1.44 mm. O ⊗ J_ℓ = 5.12 mm/s; □ ⊠ J_ℓ = 22.9 mm/s.

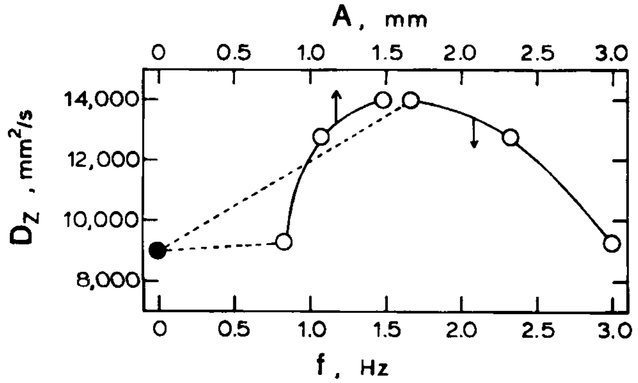

Figure 3. Effect of liquid pulsation frequency and amplitude on axial dispersion in the liquid phase of an air-water bubble column.

J_g = 13.0 mm/s, J_l = 5.12 mm/s, and pulse velocity fA = 2.50 mm/s. ● represents steady flow.

than at constant fA as in the present runs). These authors argued, after Rosen and Krylov (1974), that once pulsation is introduced, increasing the frequency increases the radial uniformity of the liquid phase and thus reduces the axial dispersion. It is of interest to note that when the D_z vs. f data of Figure 3 are replotted in the same figure as D_z vs. A, D_z then rises monotonically with A even when one includes the point at $A = 0$. Similar effects of f and A on continuous phase axial dispersivity, at constant but low values of fA, have been observed by Kim and Baird (1976) for both single-phase (water and aqueous solutions) and two-phase (kerosene dispersed in water) countercurrent flow through their reciprocating plate column.

Data for three-phase fluidization are shown in Figures 4a and 4b for 0.5 and 0.9 mm beads, respectively. As with the gas-liquid system, at the lower liquid velocities the liquid pulsations caused an increase in the axial dispersion coefficients above the corresponding steady state values, while at the highest liquid velocity this trend was reversed. The latter trend is also shown in Figure 5 by the sharp drop in D_z from steady to pulsed flow, for one of the high liquid velocity points of Figure 4b. However, the effect on D_z of varying f and A, keeping pulse velocity fA within a narrow range, was very different for the three-phase system than for the two-phase gas-liquid system, as can be seen by comparing the D_z vs. f points for pulsed flow in Figure 5 with those in Figure 3. In the three-phase system D_z rises rather than falls as f is increased, and when the same data are replotted in Figure 5 as D_z vs. A, D_z then changes monotonically with A even when one includes the points at $A = 0$, as it does also in Figure 3, but now it falls rather than rises as A increases. Presumably this different behavior is related to the bubble coalescence and decreased gas holdups caused by the presence of the small solid particles and is one more manifestation of the recent observation by Alvarez-Cuenca et al. (1984) that "three-phase fluidized beds are far more complex than bubble columns."

Implications for gas-liquid mass transfer

The rate of mass transfer from the bubbles of pure gas to the bulk of the liquid in three-phase fluidized beds depends on the gas-liquid interfacial area, the driving force in the liquid phase, and the corresponding mass transfer coefficient. The low liquid pulsation velocities of the present study appear to produce no

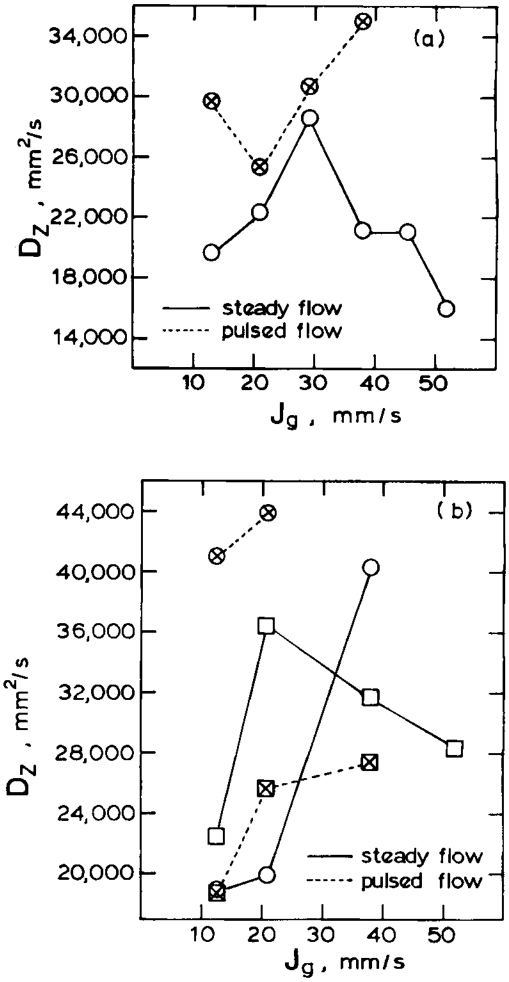

Figure 4. Effect of liquid pulsations on axial dispersion in the liquid phase of an air-water fluidized bed of glas ballotini.

Open points, steady liquid flow; crossed points, pulsed liquid flow with pulse velocity $fA \simeq 4$ mm/s.

(a) d_p = 0.50 mm; H_s = 0.71 m; f = 2.33 Hz; A = 1.64 mm; J_l = 30.0 mm/s. (b) d_p = 0.90 mm; H_s = 0.61 m; f = 3.00 Hz; A = 1.44 mm. ○ ⊗ J_l = 22.34 mm/s; □ ⊠ J_l = 44.8 mm/s.

significant changes in gas holdup and therefore in interfacial area. The increase of D_z on pulsing at the lower liquid velocities and its decrease on pulsing at the higher liquid velocities signify reduced and enlarged driving forces, respectively, for gas-liquid (and for that matter, also for particle-liquid and wall-liquid) mass transfer. The third effect, that of liquid pulsations on the mass transfer coefficient, can only be determined by making the appropriate mass transfer measurements.

Conclusions

1. The transition from a fixed to a fluidized bed for a liquid-solid system occurs at a lower minimum fluidization velocity and lower minimum fluidization voidage when the liquid is pulsed than when it is in steady flow.

2. For the low pulse velocities of the present study, no significant effects of liquid pulsation on phase holdups are discernible for a bubble column, a liquid fluidized bed, and a gas-liquid

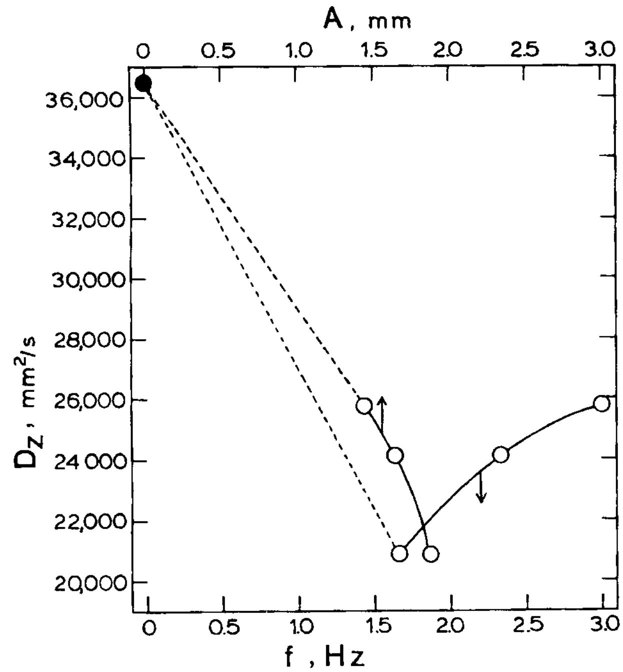

Figure 5. Effect of liquid pulsation frequency and amplitude on axial dispersion in the liquid phase of an air-water fluidized bed of 0.90 mm glass ballotini.

H_s = 0.61 m, J_g = 20.9 mm/s, J_ℓ = 44.8 mm/s, and pulse velocity = 3.1–4.3 mm/s. ● represents steady flow.

fluidized bed of solid particles smaller than 1 mm, with gas and liquid velocities each varying up to 60 mm/s.

3. The presence of small fluidized solid particles in a cocurrent stream of gas and liquid can reverse the effect of changing pulse frequency and amplitude on liquid phase axial dispersivity.

Acknowledgment

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Notation

A = liquid pulse amplitude = pulse volume/column cross-sectional area
C = tracer concentration upstream of the injection plane
C_o = mixed tracer concentration of the liquid leaving the column
D_z = liquid phase axial dispersion coefficient or axial dispersivity
d_p = solids particle diameter
f = liquid pulse frequency
H = bed height
H_s = static bed height
J_g = superficial gas velocity
J_ℓ = total superficial liquid velocity = $J_{\ell s} + fA$
$J_{\ell s}$ = steady superficial liquid velocity
J_{mf} = superficial liquid velocity at minimum fluidization
h = differential manometer reading, mm CCl_4 surrounded by water
V_ℓ = interstitial liquid velocity = J_ℓ/ϵ_ℓ
z = axial distance

Greek letters

ϵ = voidage = void volume fraction = $\epsilon_g + \epsilon_\ell$
ϵ_g = gas holdup
ϵ_ℓ = liquid holdup
ϵ_{mf} = voidage at minimum fluidization
ϵ_s = solids holdup = $1 - \epsilon$

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