The Mechanics of Vertical Moving Liquid-Liquid Fluidized Systems: II. Countercurrent Flow

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In a previous publication it has been shown that for the case of liquid droplets moving through a second quiescent liquid the holdup vs. slip-velocity ratio could be predicted on the basis of analogous rigid sphere fluidization data. The present paper extends these results to cover the countercurrent flow of both liquid phases. As long as the system is operated below the flooding point, the behavior of the spray column can be predicted from the corresponding solids fluidized data.

In a previous publication (3) a theory of fluidization was presented which postulated that the fractional or volume holdup of the dispersed phase for all types of vertical fluidized systems is a unique function of the slip velocity between the two phases. This theory has been verified for the ideal case of rigid spheres fluidized by a continuous fluid phase (*4, 5, 7*).

Weaver, Lapidus, and Elgin (*8*) extended these concepts to include the case of liquid droplets moving through a quiescent continuous liquid, that is a liquid liquid spray column with a stationary continuous phase. The operating curve of holdup vs. slip-velocity ratio (ratio of slip velocity to single particle terminal velocity) for liquid droplets fluidized in this manner was predicted on the basis of analogous rigid sphere fluidization data. As a result of this investigation it was concluded that the nonidealities of droplet oscillation, distortion, and internal circulation patterns associated with a liquid-in-liquid fluidized system could be accounted for by corrections to the ideal rigid sphere fluidized theory.

In the present paper Weaver's work is extended to cover the case of countercurrent flow of both liquid phases. The data show that, within the range of droplet sizes used, the operational behavior of liquid-in-liquid countercurrent fluidized systems can be predicted from the analogous solid-in-liquid systems.

In Weaver's investigation an unexplained deviation was found between the experimentally observed and theoretically predicted holdup for the case of high droplet flow rate. These deviations are here shown to be due to operation beyond the flooding point; below the flooding point the agreement between experiment and theory is excellent. Since the theoretical equations are not expected to hold above the flooding condition, these results remove the discrepancy noted.

THEORY

Since all of the theoretical equations and subsequent manipulation have been presented in previous publications, only the main features of Weaver's method are listed here. For predicting liquid-in-liquid column holdup, these include:

1. For the desired flow conditions of the disperse phase a Sauter mean diameter (volume-to-surface ratio) is determined.
2. By means of the Klee and Treybal correlation (*2*) the single droplet terminal velocity for the mean diameter drop is predicted.
3. From the Sauter mean diameter and fluid density the Zenz correlation (*9*) is used to predict the holdup vs. slip velocity ratio curve for an analogue model of fluidized rigid spheres.
4. The locus of points on this curve characterizes the nonflooded spray-column operation. An iteration procedure can be used to determine the holdup corresponding to the desired flow rates from the equation (*3*):

$$V_s = \frac{Q_d}{A(1-\epsilon)} + \frac{Q_c}{A\epsilon}$$

EXPERIMENTAL SYSTEM

The equipment of this investigation was essentially the same as that used by Weaver. The major change was the addition of a water recirculation system to accommodate flow of the continuous phase (Figure 1). Isobutanol was used as the dispersed phase and tap water as the continuous phase. Both liquids were continuously recirculated and presaturated with respect to each other.

The continuous water phase entered the top column section through a calming device and was withdrawn continuously from the bottom of the column below the isobutanol nozzles. The isobutanol entered the bottom entry section through an assembly of twelve sharp edged brass nozzles affixed to a single brass nozzle head which was packed with ¼-in. glass beads to prevent channeling. Nozzle diameters of both 1/16 and 1/8 in. were used. These lead to Sauter mean drop diameters of 0.053 and 0.075 in. respectively. Droplet coalescence occurred at the isobutanol water interface, which was maintained above the water entry head in order that the isobutanol phase might be continuously removed from the top of the column. Manual valve adjustment kept the interface level constant.

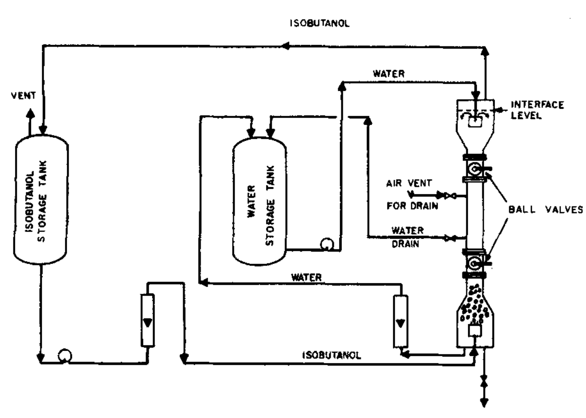

Fig. 1. Flow diagram for experimental apparatus.

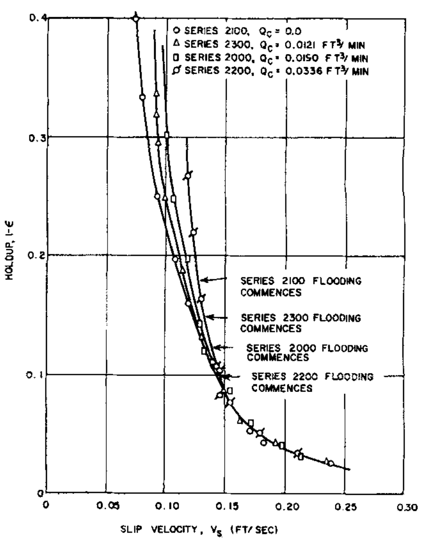

Fig. 2. Experimental holdup vs. slip velocity for isobutanol-water, 1/16-in. nozzles.

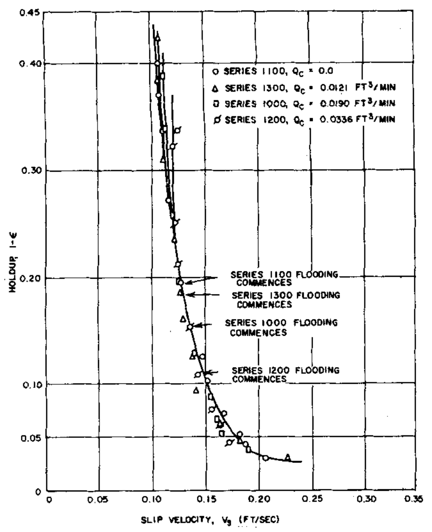

Fig. 3. Experimental holdup vs. slip velocity for isobutanol-water, 1/8-in. nozzles.

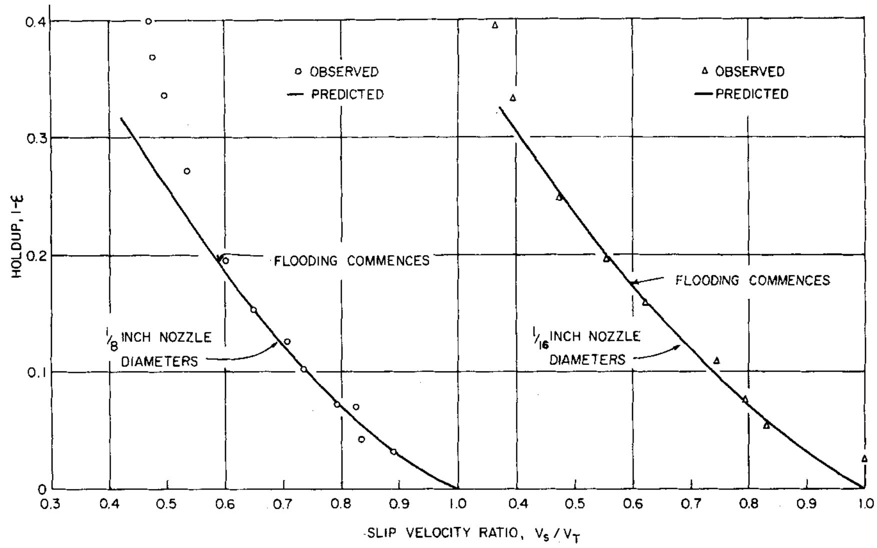

Fig. 5. Comparison of observed with predicted behavior, static continuous phase, 1/16- and 1/8-in. nozzles.

RESULTS AND DISCUSSION

The present study shows that Weaver's method for predicting spray column holdup also applies to the case in which the continuous phase is in countercurrent flow. The experimental curves for holdup as a function of slip velocity both for 1/16- and 1/8-in. nozzles are presented in Figures 2 and 3. Each of these curves is the result of a series of runs in which the dispersed phase flow rate was varied while a constant continuous phase flow rate was maintained.

It can be seen from these plots that each of the curves representing a given continuous phase volumetric flow rate begins to diverge gradually from the curve representing a quiescent continuous phase at a different holdup value. The holdup at which each curve diverges corresponds to that at the flooding point for that particular continuous phase flow rate. The observed flooding points as determined in separate runs are plotted in Figure 4. The higher the continuous phase flow rate the lower is the flooding value of the holdup and the corresponding dispersed phase flow rate; hence the deviations from the nonflooded holdup curve commence at different holdups for each continuous phase flow rate.

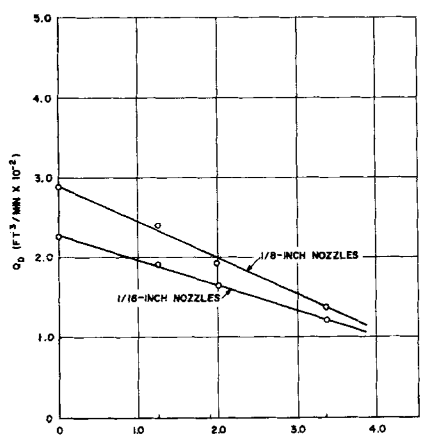

Fig. 4. Flooding points for countercurrent flow of isobutanol dispersed in water.

The Elgin-Lapidus generalized theory of fluidized systems does not extend beyond the flooding point, unless modified for a compound system comprised of two different fluidized beds in juxtaposition. Furthermore in industrial practice spray columns are not operated in a flooded condition. Therefore flow of the continuous phase in a spray column can be said to have no practical effect upon the relationship of holdup to slip velocity, that is no effect upon a nonflooded column. As a final result this implies that the conclusions of Weaver for a static continuous phase can be directly extended to cover the case of a moving continuous phase.

Flooding Phenomena

Under flooding conditions a second fluidized bed of droplets is formed in the flared dispersed phase entry section. The holdup of droplets in this feed section fluidized bed is larger than that in the main column section. As the flow rates are increased beyond the initial flooding point, this lower bed of droplets moves further into the conical end section toward the feed point. Eventually a pair of flow rates will be reached at which the bottom interface of the lower fluidized bed reaches the dispersed phase nozzles and causes a complete disruption of column stability by extreme turbulence in the feed section.

Incoming droplets of dispersed phase feed this second bed which in turn feeds the column. Dispersed-phase droplets entering a flooded column must therefore pass from a dense fluidized bed in the feed section before reaching the column length of constant area.

The main column section may be considered to be occupied by a bottom restrained fluidized bed in countercurrent flow, in which the bottom restraint is the denser bed of fluidized droplets in the lower feed section. Increased throughput beyond the flooding point is accompanied by an increase in holdup; however the resulting holdup is smaller than that which would correspond to one in a nonflooded spray column at the same feed rates. The dispersed-phase velocity through a flooded column must therefore be greater than it would be were the column not flooded with the same feed rates. Similarly the velocity of the continuous phase through a flooded spray column is less than would be expected if the column were not flooded at the same volumetric feed rates.

The breakup of the jet from the nozzles into drops may have been affected by impingement upon the bed of droplets in the flooded feed section and the turbulence induced by the presence of these fluidized droplets; however any small changes in the Sauter mean droplet diameter would produce a negligible effect on holdup compared with the observed effects.

Comparison with Rigid-Sphere Results

Weaver's method of plotting holdup vs. the slip-velocity ratio V_s/V_t can be used to predict spray-column holdup with a static continuous phase from the fluidized rigid-sphere model. The experimental holdup vs. slip-velocity ratio curves for a static continuous phase are compared in Figure 5 with the predicted curves obtained from the rigid-sphere model. Agreement between the experimental and predicted curves is good, when one considers the accuracy of the Klee and Treybal terminal velocity correlation (± 4.5%), the precision of the statistically determined Sauter mean diameter (± 8%), and the errors inherent in the Zenz correlation itself. It is possible that the use of an alternate correlation rather than that of Zenz for predicting the holdup from the analogue solid-fluidized system might increase the agreement between the experimental and predicted curves. Such an alternate correlation has been given by Richardson and Zaki (*6*).

A curve relating holdup to slip velocity can be constructed readily from the above predicted curve for the case of a spray column operating with a quiescent continuous phase. But, as demonstrated earlier, the relationship of holdup to slip velocity in a nonflooded spray column is independent of continuous-phase flow. Thus the holdup vs. slip-velocity curve for a nonflooded spray column with a quiescent or flowing continuous phase can be predicted from the solid-in-liquid analogue system.

NOTATION

V_s = slip velocity
V_t = single droplet terminal velocity
Q_d = volumetric flow rate of disperse phase
Q_c = volumetric flow rate of continuous phase
A = column cross-sectional area
ϵ = fraction void or fraction of column occupied by the continuous phase
$1-\epsilon$ = fraction holdup of disperse phase

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Fluid-Particle Mass Transfer in a Packed Bed

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Mass transfer coefficients have been measured for air flowing through a bed of naphthalene pellets. Results were obtained for ⅜-in. spheres and for ½, ⅜, and ¼-in. cylinders in a bed 4 in. in diameter and packed to heights varying from 5 to 10 in. Radial concentration profiles were obtained for some conditions, from which the radial variation of the mass transfer coefficient was determined. The point-values of k_g follow an equation of the form $k_g = a\ G^b$. From mixed outlet concentrations, values of $k_{g_{av}}$ and j_d, corrected for the effect of axial diffusion, have been obtained for all the pellet sizes as functions of mass velocity and Reynolds number, respectively.

Since it is known that the velocity in a packed bed varies with the radial position (*31*), it is of interest to determine how the mass transfer coefficient varies with radial position. The coefficient should depend on the velocity in a systematic way, and the results of the present work show that it does. Another investigator (*33*) found on the contrary that k_g was not a function of radial position. Local coefficients in the present work were measured at room temperature for a bed of ½-in. cylinders of naphthalene evaporating into air for a range of $D_p\ G_{av}/\mu$ from 840 to 9,900 for bed heights of about 4.5 and 6.0 in.

Although many studies of mass transfer in packed beds have been made (*1, 4, 10, 11, 14, 16, 17, 18, 19, 20, 21, 22, 25, 27, 28, 29, 30, 32, 33, 34, 36, 37*), apparently none have been made in which the effect of axial dispersion has been evaluated quantitatively. As a matter of fact Ergun (*13*) has asserted that many earlier studies on solid-gas systems actually were made on beds in which there was complete mixing. Therefore in the present study not only has the configuration of the bed been designed to reduce the importance of axial eddy diffusion, but the mass transfer coefficients have also been corrected for the

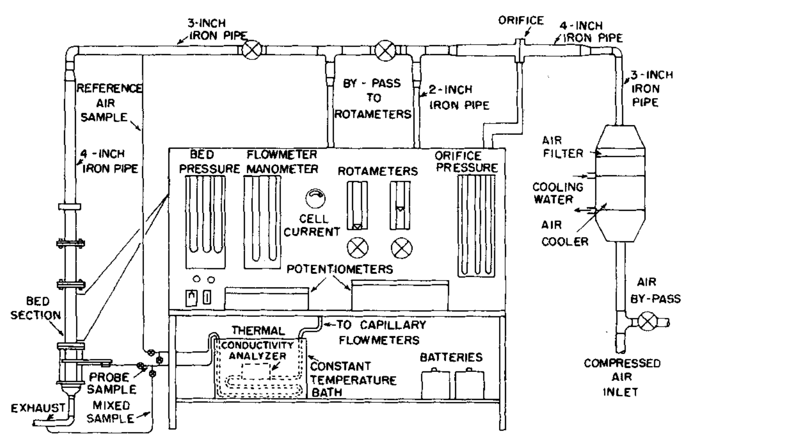

Fig. 1. Diagram of experimental apparatus.

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