may find it desirable to keep its spurious behavior well in mind particularly if potential applications are at low pressures.

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Wall-to-Bed Heat Transfer in Fluidized Beds

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A model has been proposed by Gabor (*1*) for relating the rate of wall-to-bed heat transfer in a fluidized bed to the particle residence time at the heat-transfer surface. In that model, a bed of spherical particles was approximated by a series of alternate slabs representing the solid and gas phases. The model was solved numerically by the method of Dusinberre (*2*).

The predictions of the alternate-slab model, using a gas slab thickness of $0.1d$ and a solid slab thickness of $2/3d$, were in good agreement with data by Botterill et al. (*3*), Harakas and Beatty (*4*), and Hampshire (*5*) on heat transfer from a wall to moving packed beds of glass and copper particles. However, existing data on heat transfer in fluidized beds are either incomplete or were not taken under applicable conditions for a satisfactory test of the model (*1*). Therefore, to test the applicability of the alternate-slab model to fluidized beds, measurements were made of heat-transfer coefficients as a function of particle residence time in fluidized beds.

EXPERIMENT

Heat transfer measurements made in two 18-in. high columns operated at the same conditions were compared. One column was semicylindrical, and the other was cylindrical. The semicylindrical column was constructed by cutting a 4-in. diam. brass tube in half lengthwise. A 1-in. thick glass plate coated with an electrically conducting mixture of antimony and tin oxides to reduce static charges in the fluidized bed was attached to form a flat face. A 1-in. by 1-in. heater surrounded by guard heaters was mounted in the back curved surface 6 in. above the sintered metal gas distributor plate at the bottom of the column. Air at the bed minimum fluidization velocity was introduced through the gas distributor. Bubbles were periodically injected from a tube mounted in the gas distributor plate adjacent to the midpoint of the base of the glass wall. A bubble that formed at the injection point was assumed to have a diameter equivalent to the dimension observed through the glass wall and was presumed to be a half-bubble whose radius dimension into the bed (perpendicular to the glass wall) was half that of the observed bubble diameter. The half bubble retained its identity as it rose along the axis of the column on the glass wall. Because of axial symmetry, the particle motion at the edges of the column (which was observed through the glass wall) was assumed to be identical to the particle motion within the bed at the heater surface. The effects of bubbling on both particle motion and heat transfer were studied simultaneously.

The cylindrical column was constructed from a 4-in. diameter stainless steel tube. This column had a 2-in. high by 1½-in. wide glass window that was interchangeable with a heater of the same dimensions. From an injection tube installed in the center of the base of the column, individual axisymmetric bubbles were injected into the bed. The cylindrical column had to be operated twice under identical conditions, once for measurement of heat-transfer rates and once for observation of particle movement. Comparison of the results for the two columns was used to check the validity of the assumption that particle behavior is axisymmetric in the semicylindrical column.

The particle movement was photographed with a Milliken Model DBM-5 motion picture camera at 120 frames/sec. The particles fluidized were two sizes of glass beads, 1.51×10^{-3} ft. and 1.17×10^{-3} ft. in diam., with colored glass particles as visual tracers and 1.93×10^{-3} ft. copper shot with 1.0 weight percent nickel particles as visual tracers. The alternate-slab model predicts that the addition of 1.0 weight percent nickel particles, which have the same density as copper particles, to the bed would affect the data by only 0.1 percent.

Bubbles of 1- to 1½-in. diam. were injected into both columns with the beds at minimum fluidization at frequencies of 45 to 90 per min. by means of a solenoid valve and an electric timer. To obtain frequencies higher than 90 per min., the gas for bubble formation was introduced at a uniform flow rate through the injector tube to form bubbles freely. These bubbles were not uniform in size but were formed at a higher frequency (~ 120 min.$^{-1}$) than could be achieved with the timer.

DETERMINATION OF PARTICLE RESIDENCE TIME

The particles along the wall moved upward during bubble formation at the injector tube. The particles then moved downward as the bubble rose up the column; after the bubble passed, they remained stationary until the next bubble formed. The particle residence time was determined by the following equation:

$$t = \frac{L(t_u + t_d + t_r)}{V_u t_u + V_d t_d} \tag{1}$$

Equation (1), which is based on the assumption that all particles above or below the heater are at bed temperature, is derived by dividing the inventory of particles at the heater surface by the rate of particle flow past the heater.

The particle velocities and residence times during the different stages of bubble injection, formation, and rising were determined by a frame-by-frame analysis of the motion picture film for 15 bubbles. For the freely formed bubbles, there was essentially no upward particle movement. The downward movement of particles into the heat-transfer region was at a higher velocity than the downward movement at the bottom of the heat-transfer region. Some particles were observed to leave in the direction approximately perpendicular to the heater surface; this accounted for the difference in downward particle velocities at the top and bottom of the heat-transfer region. If it is assumed that the velocity decreases linearly along the heater surface,

$$V_d = V_i - \frac{x}{L}(V_i - V_o) \qquad (2)$$

The time for a particle to move a distance x along the heat-transfer surface is

$$t_x = \int_0^x \frac{dx}{V_d} \qquad (3)$$

The average residence time for the particles is then:

$$t = \frac{\int_0^x t_x\,dx}{L} = \frac{L\,V_i}{(V_o - V_i)^2}\left[\frac{V_o}{V_i}\ln\frac{V_o}{V_i} - \frac{V_o}{V_i} + 1\right] \qquad (4)$$

Equation (4) was used to determine the average residence time of the fluidized particles for the freely bubbling bed from the observed particle velocities at the top and bottom of the heat-transfer region.

RESULTS

The data for the glass beads from the cylindrical and semicylindrical columns are compared with the alternate-slab model in Figure 1. Data for the copper particles are given in Figure 2. For both the glass and copper particles, the fluidized-bed data agree with the heat-transfer rates predicted by the alternate-slab model with about the same amount of scatter as the moving packed bed data. The data were reproducible within the range of scatter shown on Figures 1 and 2. These data verify that the alternate-slab model, which was originally compared with moving packed bed data, is applicable to fluidized beds. The moving packed bed data were in good agreement with the alternate-slab model over the entire range given in Figures 1 and 2.

The experimental work was also supplemented by extending the alternate-slab model in considering the effect of gas flow at the minimum fluidization velocity through the bed. The volumetric heat capacity for gas is so small compared to that of the solid particles that the inclusion of convective heat removal by the flowing gas did not alter the values of the heat transfer coefficient as a function of particle residence time. This modification of the model gave additional theoretical justification for the appropriateness to fluidization of the alternate-slab model which originally assumed a stagnant gas medium, which is more applicable to a moving packed bed situation.

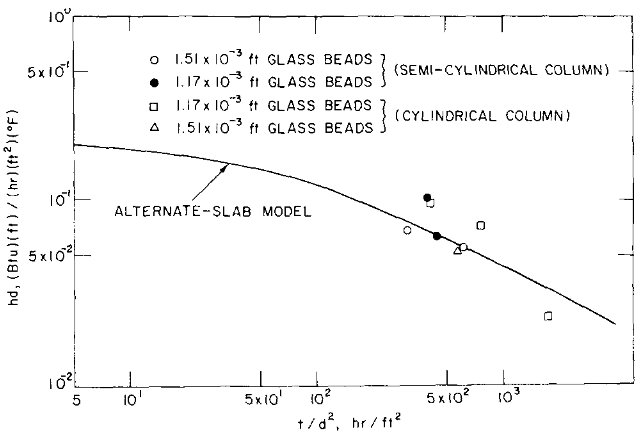

Fig. 1. Wall-to-bed heat transfer coefficients as a function of fluidized particle residence time for glass particles.

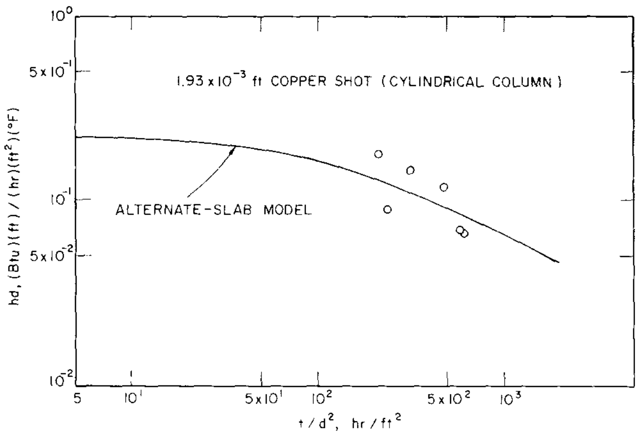

Fig. 2. Wall-to-bed heat transfer coefficients as a function of fluidized particle residence time for copper particles.

The data from the cylindrical and semicylindrical columns agreed, indicating the validity of the assumption that particle movement at the extremities of the flat face of the semicylindrical column is similar to particle movement near the heater surface. This demonstration of axisymmetric behavior should be of value in other areas of research in which it is desired to observe a cross section of a three-dimensional fluidized bubble.

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NOTATION

d = particle diameter, ft.
h = heat transfer coefficient, B.t.u./(hr.)(sq.ft.)(°F.)
L = height of heat transfer surface, ft.
t = particle residence time, hr.
t_d = time during which particles move downward, hr.
t_r = time during which particles are at rest, hr.
t_u = time during which particles move upward, hr.
t_x = time for particle to move distance of x, hr.
V_d = downward particle velocity, ft./hr.
V_i = particle velocity into heat transfer region, ft./hr.
V_o = particle velocity out of heat transfer region, ft./hr.
V_u = upward particle velocity, ft./hr.
x = coordinate along heater length, ft.

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