Although various assumptions are involved in deriving this solution, it does provide a quantitative evaluation of the heat-transfer rate for such a layer, as well as the effects of the individual parameters. Subsequently, it will be necessary to consider the process for other boundary conditions, and also to incorporate the actual setting, in particular, the lower effective layer density at the wall and the nonuniform component velocity distributions.

NOTATION

a, surface area of particles in unit volume; c, specific heat; d, particle size; r, current radius; r_0, channel radius; v, velocity; w, water equivalent; β, volume concentration of solid component; λ_g^*, effective thermal conductivity in bed; μ_n, root of characteristic equation; $\vartheta = t - t_w$, local excess temperature; $\theta = \vartheta/\vartheta_{g0}$, local dimensionless excess temperature. Indices: g, gas component; f, flow; w, wall; s, solid component; 0, inlet section (x = 0); mb, moving blown; mu, moving unblown.

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PARTICLE MIXING IN A FLUIDIZED BED CONTAINING BAFFLES

N. V. Kuzichkin, I. P. Mukhlenov, A. T. Bartov, and A. N. Prokopenko

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Results are given on the mixing of the solid phase in a fluidized bed containing baffles; it is shown that a diffusion model can be used to advantage in analyzing longitudinal particle transport.

Mixing in the solid phase in a fluidized bed is a major aspect of fluidization theory, since it provides information on the heat-transfer mechanism and data on the interaction between the gas and solid. Information has been published [1, 2] on the particle mixing in a fluidized bed containing baffles (a structured bed), but the data relate in the main to small laboratory columns and are unrelated to the bed structure. The longitudinal-mixing coefficient correlated with the hydrodynamic factors serving as the basic definitive parameter contains the fluidization number, which unfortunately does not adequately characterize the hydrodynamic setting in a fluidized bed [3, 11]. Here we present results on the solid mixing in larger models of diameters 0.15 and 0.55 m.

The baffles were, as in [10], bent pieces of mesh forming blocks. The specific volume of the baffles was about 2% of the total fluidized-bed volume. The cell size in the mesh was 5 × 5 mm and the wire diameter 1 mm.

Qualitative experiments were made with a planar model with the packing described in [4] to establish the general mode of particle mixing in such beds; stained parti-

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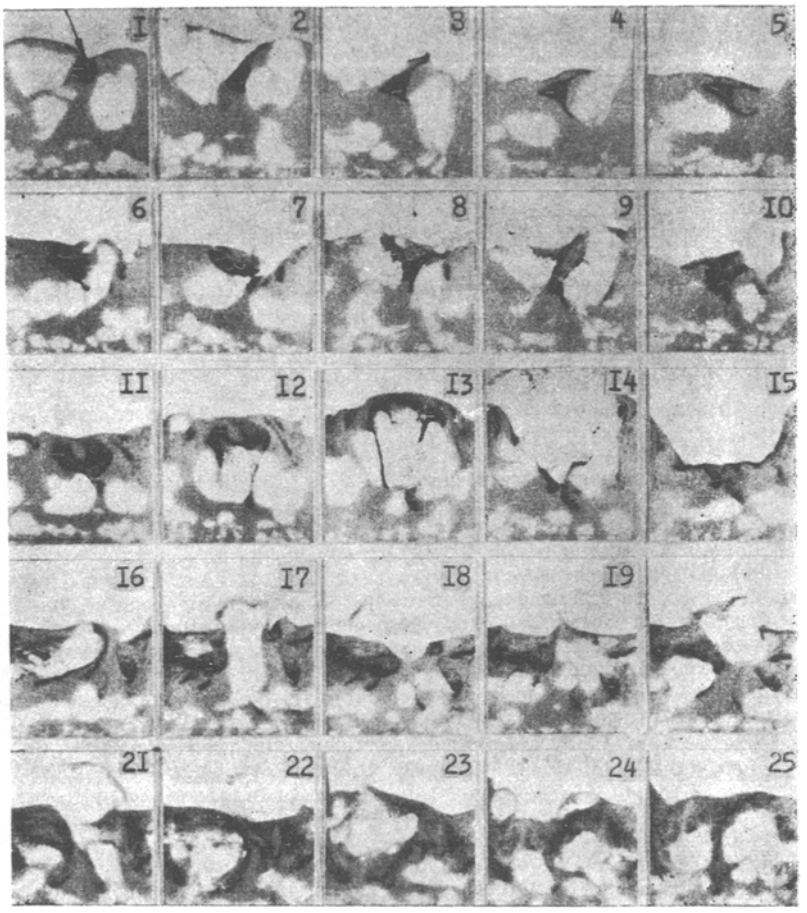

Fig. 1. Mixing of tracer particles in a free bed at $\Delta w = 0.5$ m/sec (recorded in transmitted light). Frame intervals: along lines 0.12 sec, between lines 0.5 sec.

cles of aluminosilicate were introduced at a point at the center of the bed. Cinephotography was used to examine the mixing of the tracer. The frames (Figs. 1 and 2) show that the mixing in a free bed is qualitatively different from that in a structured one. In the first case, the particles mix as unstable clumps, which do not retain their individuality. The random motion is accompanied by ordered transport, and the scales of the turbulent eddies are comparable with the scale of the flow as a whole. The large-scale circulation of particle clumps in a free bed is associated with the passage of gas bubbles, which restricts the applicability of the diffusion model to solid-phase transport, especially for beds of large diameter and small initial height.

In a bed containing baffles, the volume of labeled particles disperses slowly (Fig. 2), and the mixing mechanism is simplified, being close to diffusion type. This means that the equation for nonstationary diffusion can be applied to describe the longitudinal transport:

$$D_s \frac{\partial^2 C(z, \tau)}{\partial z^2} = \frac{\partial C(z, \tau)}{\partial \tau}. \quad (1)$$

The boundary and initial conditions are

$$C(z, 0) = 0 \text{ for } H > z > 0; \quad (2)$$

$$C(z, 0) = C_0 \text{ for } -l_0 < z < 0; \quad (3)$$

$$\frac{\partial C(-l_0, \tau)}{\partial z} = \frac{\partial C(H, \tau)}{\partial z} = 0; \quad (4)$$

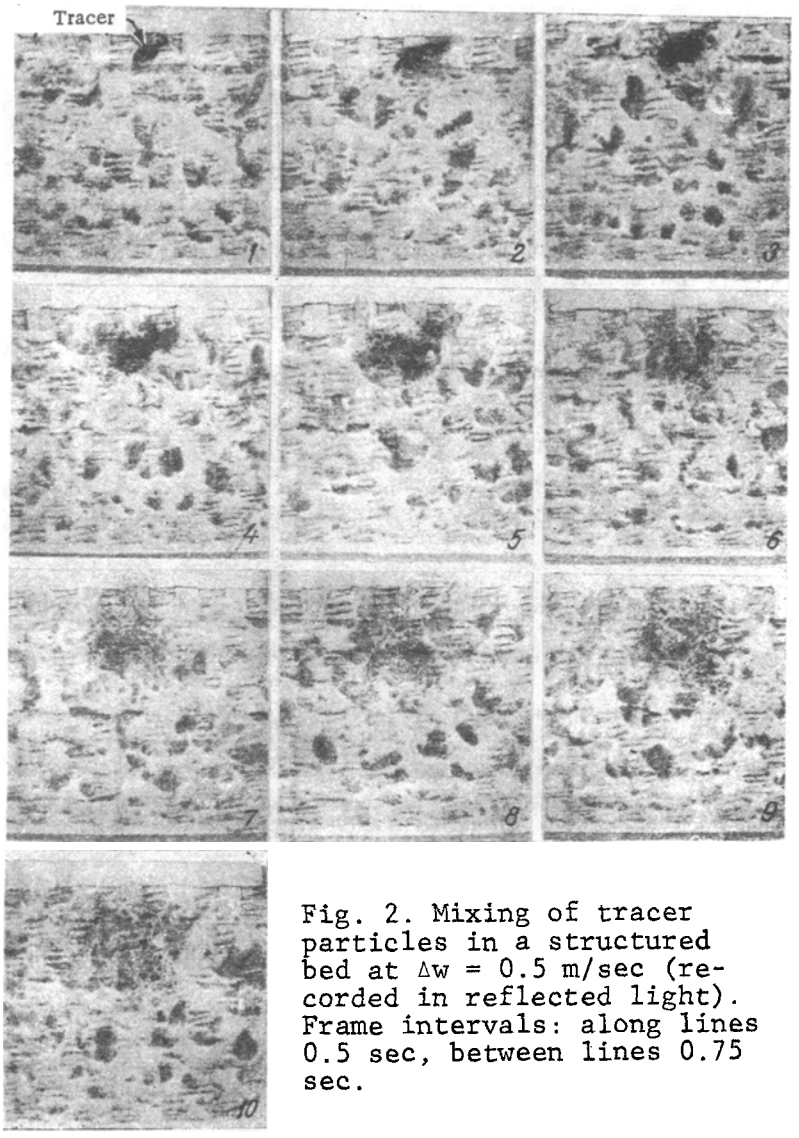

Fig. 2. Mixing of tracer particles in a structured bed at Δw = 0.5 m/sec (recorded in reflected light). Frame intervals: along lines 0.5 sec, between lines 0.75 sec.

$$C(-0, \tau) = C(+0, \tau) \quad \text{for } \tau > 0; \tag{5}$$

$$\frac{\partial C(-0, \tau)}{\partial z} = \frac{\partial C(+0, \tau)}{\partial z}. \tag{6}$$

The following is the solution to (1) with (2)-(6) when the height l_0 of the injected tracer is small by comparison with the depth of the bed:

$$\frac{C}{C_\infty}(\xi, \tau) = 1 + 2\sum_{n=1}^{\infty} \exp\left(-\frac{\pi^2 n^2 \tau}{H_1^2} D_s\right) \cos \pi n \xi. \tag{7}$$

The measurements were checked against (7) on the above models using grains of aluminosilicate and wear-resistant vanadium catalyst with mean diameters $\bar{d}$ of 0.25, 0.75, and 1.5 mm. The initial bed depths H_0 were 0.4 and 1.5 m. The carrier gas was air. The velocity difference $\Delta w = w - w_0$ was used as the definitive velocity; the choice of Δw for this purpose has been discussed previously [3, 10].

The tracer particles were injected into the upper part of the bed, and samples were taken at various intervals and heights; we used chemical labeling, as well as color labeling [5]. The chemical label was provided by particles of AV-17 and KU-1

TABLE 1. Longitudinal-Mixing Coefficients for Free and Structured Fluidized Beds

w_0, m/sec	$\Delta w = w - w_0$, m/sec	$\bar{d}$, mm	H_0, m	D_A, m	D_S, cm²/sec
		Structured bed			
0,4	0,15	1,5	1,5	0,15	40
0,40	0,20	1,5	1,5	0,15	50
0,4	0,30	1,5	1,5	0,15	75
0,4	0,40	1,5	1,5	0,15	105
0,4	0,60	1,5	1,5	0,15	140
0,4	0,80	1,5	1,5	0,15	185
0,4	0,20	1,5	0,4	0,15	45
0,4	0,20	1,5	0,4	0,55	48
0,40	0,60	1,5	0,4	0,55	155
0,25	0,10	0,75	1,5	0,15	14
0,25	0,30	0,75	1,5	0,15	40
0,25	0,40	0,75	1,5	0,15	52
0,25	0,60	0,75	1,5	0,15	75
0,25	0,80	0,75	1,5	0,15	105
0,25	0,40	0,75	0,4	0,55	58
0,025	0,10	0,25	1,5	0,15	12
0,025	0,20	0,25	1,5	0,15	40
0,025	0,40	0,25	1,5	0,15	65
		Free bed [7, 8]			
0,015	0,088	0,05—0,3	2,23	0,10	310
0,015	0,088	0,05—0,3	0,73	0,10	310
0,015	0,081	0,05—0,3	2,35	0,30	1150
0,015	0,186	0,05—0,3	2,07	0,30	1430
0,015	0,085	0,05—0,3	1,71	0,60	1425
0,015	0,186	0,05—0,3	1,69	0,60	1510
0,010	0,23	0,02—0,15	H=9 m	0,38	1350
0,010	0,23	0,02—0,15	H=9,6 m	1,50	4500

ion-exchange resins, which contained a higher proportion of the relevant component (carbon) in the particles. The carbon contents of the samples were determined by a standard method [6].

Figure 3 shows that the measurements agreed satisfactorily with the calculations.

The longitudinal-mixing coefficient is given in Table 1 for the various fractions at various gas speeds, initial depths, and equipment diameters.

It is clear that the longitudinal-mixing coefficient D_S for the solid is almost independent of the diameter of the equipment and the initial bed depth within the range of the hydrodynamic parameters used in the structured bed; however, the D_S for the structured bed were much lower than those for a free bed.

If one considers a bed of catalyst as a system consisting of sequential complete-mixing cells with reverse transfer of the solid phase between cells, we can derive a relation between the longitudinal-mixing coefficient, the transfer rate, and the cell length:

$$D_s = q \cdot \Delta x. \tag{8}$$

Then (8) is equivalent to a standard relation for the turbulent-diffusion coefficient, in particular, in terms of the mean speed of the turbulent pulsations and the displacement length for turbulent eddies. The speed of the pulsating motion of the solid and the path length are closely related to the passage of gas bubbles through the bed. It has been shown [9] that the distance traveled by a bunch of solid particles is directly proportional to the diameter of the latter; the scale and rate of the particle circulation are dependent on the diameter of a gas bubble and the rise rate.

One can therefore put

$$D_s = q \cdot \Delta x = f(a, \varphi, w_l). \tag{9}$$

A structured bed shows bubbles of size much smaller than those found in a free bed, and the motion of the bubbles is extremely unstable.

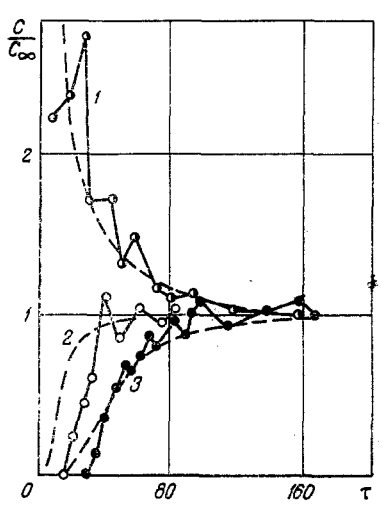

Fig. 3

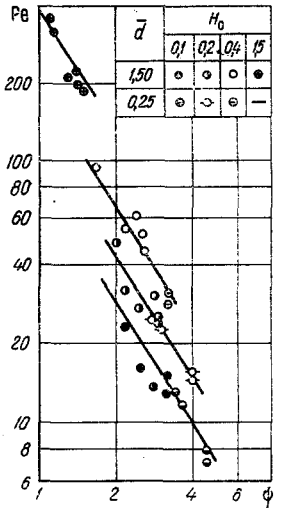

Fig. 4

Fig. 3. Comparison of values from (7) (broken line) with experiment (solid line) for $\Delta w = 0.6$ m/sec; $D_s = 140$ cm^2/sec; 1) $\xi = 0.075$; 2) 0.52; 3) 0.95.

Fig. 4. Relation of Pe to $\bar{d}$ (mm) and H_0 (m).

Also, the bubbles vanish at a higher frequency at the various levels in such a bed, which considerably reduces the mean path and bubble existence time [10]. Therefore, the scale of particle transport due to the bubbles is much less than the scale of the flow.

The spreads in bubble size and speed are only slightly dependent on the gas speed, particle diameter, and initial bed depth; they are dependent solely on the parameters characterizing the baffles [10], and so for this type of packing one can put

$$D_s = F(\varphi). \tag{10}$$

The following relationship was derived (Table 1) by processing the measurements for the ranges used in the parameters:

$$D_s = 8.5 \cdot 10^2 \varphi^{1.1}. \tag{11}$$

The discrepancies between theory and experiment did not exceed ±15%; the values for the proportion of the volume occupied by the gas were determined on a planar model using correction coefficients for the transport of the bulk model [10]. This value for the gas factor did not vary with the initial bed depth for either model.

The structural parameters such as the gas filling factor are very nonlinearly dependent on the particle diameter or Archimedes number, and the relationships are difficult to represent directly as equations [10], because φ is substantially affected by the surface properties of the catalyst (roughness, adhesion, and so on), in addition to effects from the particle size. It is then extremely difficult to produce even a correlation between the parameters for the longitudinal transport of the solid and the hydrodynamic factors. However, (11) provides sufficient information on the particle mixing.

The scale of the motion of the particle clumps can be estimated from the extent and frequency of the pressure-difference fluctuations; at large values of the relative pressure-difference deviation ψ and small mean pulsation frequencies ω one finds that the structure parameters vary with large amplitudes but low frequencies in a free fluidized bed [10], and, consequently, the bed has large-scale circulation of unstable clumps, and hence the description of the transport in terms of diffusion equations will deviate considerably from experiment. When baffles are added, ψ is

reduced but ω is increased, so the particle transport plays a lesser part. However, even in a structured bed one can observe fluctuations in the tracer concentration, but then the measurements deviate only slightly from the theoretical values.

It has previously been observed [10] that the turning points in $\psi = f(\Delta w)$ and $Pe = f(\Delta w)$ correspond to roughly the same range in the excess gas speed; this appears to be explained as follows: The mean relative deviation of the pressure difference can be considered as the proportion of the fluctuation energy consumed in displacing the solid, the proportion being relative to the total amount of energy entering the fluidized bed.

The convective component of Pe is proportional to the total energy consumption, while the diffusion component is proportional to the energy consumed in displacing the solid phase.

There should thus be a negative correlation between Pe and ψ; the experimental data (Fig. 4) were processed to give the equation

$$Pe = 320 H_0^{0.55} \cdot \psi^{-1.6}. \quad (12)$$

In (12),

$$Pe = \frac{wH}{(1-\varphi) D_s}$$

is the Peclet number.

Formula (12) gives larger errors than (11) in calculations on D_s, but (12) is more universal, since the effects of the baffle parameters on the bed structure are incorporated by the variation in ψ.

NOTATION

w_0, w, gas velocities referred to the whole bed section (at onset of fluidization and when working); H_0, H, bed heights at onset and when working; $\bar{d}$, mean particle diameter; D_A, apparatus diameter; w_l, bubble lift velocity; a, void dispersion (ratio of void volume to the surface separating voids and dense part of a bed); φ, gas filling (void volume as a fraction of total bed volume); z, tracer coordinate; C_0, C, C_∞, tracer concentrations (initial current, and for complete mixing); Δx, cell length; q, velocity of reverse mixing between cells; D_s coefficient of longitudinal (axial) particle mixing; l_0, initial height of introduced volume of tracer; $\xi = (z + l_0)/(H + l_0)$, dimensionless coordinate; $H_1 = H + l_0$; τ, time; ω, mean frequency of pressure pulsations; $\overline{\Delta p_i}$, mean pressure difference; $\overline{\Delta\Delta p_i}$, mean deviation of instantaneous pressure difference from $\overline{\Delta p_i}$; ψ, mean relative deviation of pressure drop pulsations; $Pe = wH/(1-\varphi)D_3$, Peclet number.

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