TURBULENT MIXING IN A THREE-PHASE FLUID BED

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Longitudinal-and radial-turbulent diffusion of the phases in a fluid bed are the main factors to which the smoothing-out of the temperature profile in the effective reaction zone of the reactor is to be ascribed. For several technological processes which are used in refining oil and are characterized by a large heat effect, an isotropic temperature profile is quite important and outweighs the main drawback of a fluid bed system; this drawback is that if the reaction order exceeds zero, the volume of the reaction zone must be larger because the concentration profile is smoothed out.

The effects of the various hydrodynamical factors on the anisotropy of the temperature field under operation conditions, may be estimated from quantitative experimental data on longitudinal and radial mixing in media physically similar to the actual media used in the process. Consequently, even a study on longitudinal and radial mixing in cold models should be considered to yield valuable practical information which, in particular, is applicable to three-phase fluid beds; the important technological characteristics of a three-phase fluid bed have been investigated insufficiently so far.

Experimental study on the hydrodynamic regularities in a three-phase fluid bed (solid particles-liquid–gas) has also an independent aim: clarification of the conditions under which such a system exists and elucidation of the mechanism in the actual process of hydrocracking of heavy hydrocarbon fractions.

In a three-phase fluid bed consisting of catalyst, liquid hydrocarbons, and gas, only the liquid phase remains continuously distributed. This phase is the most representative for the mass and heat balance, since the heat and products liberated from the catalyst pores migrate fast into the bulk liquid phase. For this reason we studied first the longitudinal diffusion of the liquid phase.

In the last few years several papers on longitudinal diffusion have been published [1-7]. These papers deal mainly with single- and two-phase flows. Data on longitudinal mixing in three-phase fluid beds are lacking in the literature.

The present paper reports data on longitudinal diffusion of the liquid phase, not only in a three-phase gas-liquid-solid particles system, but also in a two-phase gas-liquid system; the latter system can be considered a limiting case, being a system with an infinitely low concentration of the solid phase.

The observations in these studies were made mainly in forcedly supplied flowing phases. The mean residence time of the liquid in the observation zone equaled a few seconds; therefore, reliable samples from different points in the observation zone could be taken only if an indicator was admixed continuously and uniformly, and the process studied was stationary. For this reason we used a diffusion model and assumed that the variation of the indicator concentration with the height in the observation zone obeyed a logarithmic law. The experimental data were evaluated by means of the equation

$$\log \frac{c}{c_0} = Ax + B, \tag{1}$$

where c denotes the indicator concentration at one of the four levels where samples were taken; c_0 the indicator concentration in the water supplied to the reactor; x the distance between the level where samples were taken and the inlet point of the indicator. According to Eq. (1), a plot of log c/c_0 versus x will be a straight line; A will be the slope of this line, and B the intercept cut from the ordinate axis; since log c/c_0 is negative, the ordinates of the experimental points are negative.

Authors who studied longitudinal mixing [1, 2] proposed the equation

$$\ln \frac{c}{c_0} = \frac{v}{E} x + B', \tag{2}$$

All-Union Scientific Research Institute of Oil Industry. Translated from Khimiya i Tekhnologiya Topliv i Masel, No. 12, pp. 4-8, December, 1967.

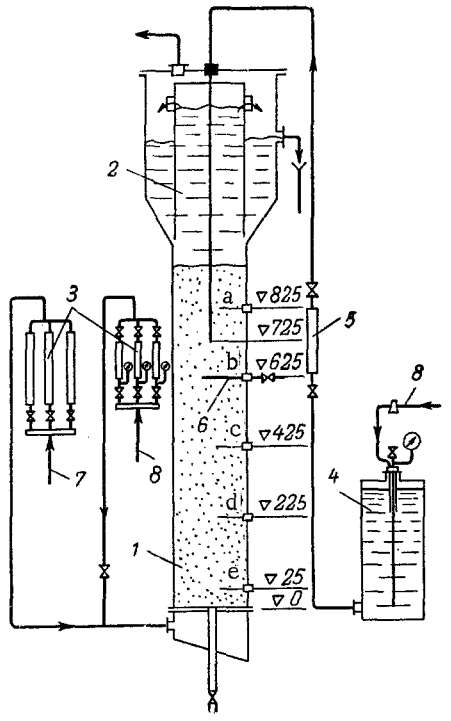

Fig. 1. Diagram of the equipment: 1) reactor model; 2) separator; 3) rotameters; 4) vessel containing the indicator; 5) rotameter measuring the discharge of indicator; 6) samples (a, b, c, d, e); 7) water; 8) air.

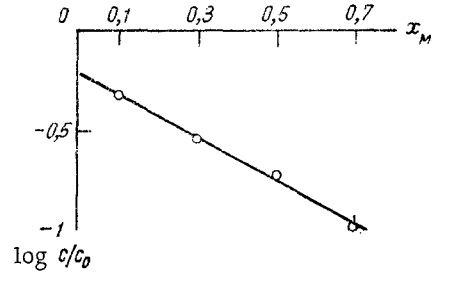

Fig. 2. Relative concentration as a function of the vertical distance (v_g = 42 mm/sec, v_l = 17.4 mm/sec, free cross-sectional area of the grate 0.26%).

where v denotes the velocity of the fluidizing current, m/sec; E the coefficient of longitudinal mixing, m^2/sec; B' a constant. Equation (2), exactly as (1), may be represented by a straight line.

The experimental data were evaluated by applying the simpler Eq. (1), which does not contain parameter v; this was done because the linear velocities in the observation zone were only very roughly known; nevertheless, combining Eqs. (1) and (2), we may calculate the E/v ratio from the formula:

$$\frac{E}{v} = \frac{0.434x}{B - \log\frac{c}{c_0}}. \qquad (3)$$

This ratio, which was called the mixing length by Reman [3], satisfactorily characterizes longitudinal mixing in fluidized two-phase beds consisting of gas and solid particles.

It was interesting to find out whether the mixing length can be used for characterizing longitudinal mixing in a three-phase fluid bed. Analysis of Eq. (2) shows that, whatever the value of B, the slope of the straight line represented by (2) decreases with increasing E/v ratio, ln c/c_0 tends towards its minimum value B, and c tends to its maximum value. Consequently, parameter E/v approximately characterizes the intensity of longitudinal mixing in a three-phase fluid bed; the larger E/v, the more intensive this mixing under given hydrodynamic conditions; E/v is always positive; in the case of ideal prop-flow E/v = 0.

Figure 1 shows a diagram of the equipment used in our study on longitudinal mixing of the liquid phase in a three-phase fluid bed.

The equipment consists of the transparent reactor model 1 measuring 146 mm in diameter and 1500 mm in height; the upper part ends in separator 2 which keeps the solid phase in the reactor. The lower part of the reactor model is provided with an interchangeable distributing grate. The experiments were done with two distributing grates whose free cross-sectional area equaled 1.03 and 0.26% of the cross-sectional area of the reactor; the grates contained 14 and 55 apertures of 2 mm diameter, respectively. The velocities of the liquid-gas current leaving the apertures of the distributing grates ranged from 3 to 47 m/sec.

If the velocity of the current entering the observation zone from the grate apertures is high, the inertia forces are considerable and cause a scattering of the measured data. The mean deviation from the logarithmic law equaled 15%; that in the two-phase system was only 9% because the inertia forces in the latter system are weaker.

The gas phase used was air, the liquid phase was water from the supply line, and the solid phase consisted of fused sediment spheres; the true particle density was 2700 kg/m^3, and the mean particle diameter calculated from the data given below equaled 0.87 mm.

Sieve mesh, mm	Left on the sieve, weight %
1.0	27.7
0.85	35.6
0.63	33.6
0.4	2.6

TABLE 1. Length of Longitudinal Mixing as a Function of the Velocity of the Flowing Phases

v_l, mm/sec	Length of longitudinal mixing, m					
	In the three-phase system at v_g			In the two-phase system at v_g		
	15–17 mm/sec	43–44 mm/sec	94–99 mm/sec	14–15 mm/sec	42–44 mm/sec	89–96 mm/sec
17,4	0,40	0,52	0,61	0,25	0,96	1,68
30,7	0,36	0,44	0,49	0,50	0,78	0,99

TABLE 2. Length of Longitudinal Mixing as a Function of the Free Cross-Sectional Area of the Grate at $v_l = 17$ mm/sec (referred to the free cross-section of the apparatus)

Number of phases	Length of longitudinal mixing, m			
	Grate with 0.26% free cross-sectional area, v_g		Grate with 1.03% free cross-sectional area	
	41–42 mm/sec	89–94 mm/sec	42–43 mm/sec	94–98 mm/sec
3	0,52	0,61	0,40	0,48
2	0,96	1,68	1,03	1,07

20.5 kg sediment spheres were poured into the reactor model, and the height of the fixed bed thus formed equaled 700 mm (referred to the grate plane). Water was supplied via rotameter 3, causing the bed to expand to a given degree. Then, a given amount of air ensuring stable operation of the three-phase fluid bed was supplied via the same rotameter. The range of the water and air velocities inside the reactor model was so chosen that no solid phase was entrained from the observation zone.

After the fixed hydraulic regime of the reactor had been established, the indicator was supplied at a constant rate from vessel 4, via rotameter 5 to the inlet 725 mm above the grate.

After the indicator concentration at the exit from the observation zone had become constant, liquid samples were taken by means of the special samplers 6 at five different levels; one was above and four below the indicator inlet. Five samples were taken at each level; the arithmetical mean of the five samples was used in the calculations.

The concentration of the indicator (sodium chloride) was determined by titrating the chloride ions. At the start of every run the chloride content of the water from the supply line was determined, and the concentrations of the samples were corrected for this blank value.

As can be seen in Fig. 2, the experimental data on the three-phase systems fairly well obey Eq. (1), which confirms that the diffusion model is applicable.

Table 1 gives data on the mixing length in three- and two-phase systems at identical discharge rates of gas and liquid through the grate with a free cross-sectional area of 0.26%. As can be seen in Table 1, an increase of the gas discharge rate at constant liquid discharge rate intensifies longitudinal mixing; this intensification is generally observed in practice when liquids are mixed by means of gas currents. At identical flow rates of gas and liquid, the presence of a solid phase depresses longitudinal mixing of the liquid phase. If the discharge rate of the liquid is increased and that of the gas kept constant, the effect of longitudinal mixing weakens noticeably.

The energy needed for intensive mixing of the interacting phases in the reactor is provided only by the kinetic energies of the liquid and gas currents leaving the apertures of the distributing grate. To study the effect of the kinetic energy of the current on longitudinal mixing, we, therefore, did experiments with the two distributing grates mentioned above (Table 2).

Table 2 shows that at the gas and liquid velocities examined the intensity of longitudinal mixing increases slightly with the velocity of the phases in the apertures of the distributing grate; the length of longitudinal mixing in three-phase systems rises only 30-40% when the velocity of the current is increased fourfold. At the other gas and liquid velocities applied, similar observations were made.

According as the reactor sizes are increased, quantitative estimate of radial mixing in the reactors becomes ever more necessary. In this paper the ratio of the difference between the maximum and minimum indicator concentration to the logarithmic mean of the five indicator concentrations at the same sampling level is taken as the criterion of radial mixing at this level. The sum of the radial mixing criteria taken over all five sampling

TABLE 3. Characteristic of Radial Mixing as a Function of the Velocity of the Flowing Phases (grate with 0.26% free cross-sectional area)

v_l, mm/sec	Characteristic of radial mixing in					
	Three-phase system at v_g			Two-phase system at v_g		
	14–17 mm/sec	39–43 mm/sec	89–99 mm/sec	14–15 mm/sec	41–44 mm/sec	89–96 mm/sec
17,4	139	108	84	72	53	51
30,7	291	144	111	156	109	92

TABLE 4. Characteristic of Radial Mixing as a Function of the Free Cross-Sectional Area of the Grate (v_l = 17.4 mm/sec)

Free cross-sectional area, %	Characteristic of radial mixing at v_g	
	42-43 mm/sec	89-98 mm/sec
0.26	108	84
1.03	307	117

levels is a general parameter characterizing the intensity of radial mixing in our case. The smaller this parameter (the closer to zero), the better radial mixing will be, and the more uniform the concentration field in a given cross section. The dependence of radial mixing on the velocities of the flowing phases is shown in Tables 3 and 4. Radial mixing in three-phase systems increases with the gas velocity. If the free cross-sectional area of the distributing grate is reduced, the radial concentration profile is markedly levelled out.

The mean indicator concentration found by integrating the function $c_m = f(x)$, without making any assumption on the shape of this function and without introducing coefficients, may be used as a criterion characterizing the intensity of longitudinal mixing. The numerical values of this criterion differ from the E/v ratio, but the regularities are not essentially different when the former criterion is used.

SUMMARY

Our study on longitudinal and radial mixing in the liquid phase of a three-phase fluid bed shows that:

1) the concentration profile of the indicator in the observation zone approximately satisfies a logarithmic law;

2) acceleration of the gas velocity intensifies longitudinal mixing;

3) acceleration of the liquid velocity reduces the effect of longitudinal mixing

4) the presence of a dispersed solid phase depresses the effect of longitudinal mixing;

5) a rise of the kinetic energy of the flows entering the working zone of the reactor model promotes longitudinal mixing, but the increase in the length of longitudinal mixing is not proportional to the velocity of the current;

6) the concentration profile of the indicator in a cross section through the three-phase bed becomes less uniform with increasing discharge rate of the liquid, and more uniform with increasing discharge rate of the gas;

7) the concentration profile of the indicator in a cross section through the three-phase fluid bed becomes more uniform when the free cross sectional area of the grate is increased.

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