DETERMINATION OF THE MIXING COEFFICIENT IN A FLUIDIZED BED REACTOR

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A method is presented for determination of the mixing coefficient in a fluidized bed by means of a mathematical model of the process. A numerical example is given for the case of an ore-drying process.

In fluidized beds intensive particle mixing takes place. As shown by experiment, in the fluidized beds of large industrial reactors the rates of chemical reactions or changes in other process parameters are commensurate with the rate of mixing of the particles, so that in calculating the yield it is not sufficient to consider the reactor as a lumped capacity; on the contrary, it is necessary to take into account the finite particle mixing time.

To allow for this factor certain authors [1] have proposed to introduce a mixing coefficient (by analogy with the diffusion coefficient in the theory of liquids and gases) and assume that Fick's equation

$$\Phi = -D \operatorname{grad} \rho \tag{1}$$

is applicable to a fluidized bed.

The main difficulty here is that there have been very few studies of the mixing coefficient. The available data [1] are fragmentary and inconsistent. Moreover, there is a lack of data on the mixing coefficients for large-scale (industrial) apparatus.

Knowledge of this coefficient is necessary to set up mathematical models of fluidized-bed processes in which the particle mixing time cannot be disregarded.

On the other hand, using a mathematical model of the process, it is possible to determine an unknown mixing coefficient, if the necessary experimental data are available.

The mixing coefficient can be determined by periodically injecting labeled particles into the bed and measuring their concentration along the bed at specified intervals of time.

In a continuous technological process characterized by the variation of some processing coordinate (for example, the amount of reduced or residual sulfur in the process of roasting sulfide ores, the relative moisture content in drying processes, etc.; we will henceforth call this coordinate the degree of processing) there is no need to use labeled particles to find the mixing coefficient. In such processes, under steady-state conditions, the value of the mixing coefficient determines the distribution along the bed of the degree of processing, averaged in a given section over all the particles or the individual fractions. The shape of this curve also depends on the kinetics of the process.

Thus, we may draw the following conclusion. If the kinetics of the process have been sufficiently studied and, with the aid of a mathematical model, the relation between the shape of the distribution curve for the averaged, in the above sense, degree of processing and the mixing coefficient has been found, then, by determining experimentally the distribution of the degree of processing along the length of the furnace, we can uniquely assign to each such distribution a corresponding mixing coefficient.

In references [2, 3] mathematical models of thermochemical processes in a fluidized bed were presented in general form; these take the form of equations for the probability density function $\rho(r, d, m, t)$.

For a horizontal reactor with total mixing along the vertical coordinate the equation is written in the form

$$\frac{\partial \rho}{\partial t} = D \frac{\partial^2 \rho}{\partial x^2} + \frac{\partial \rho}{\partial m} \frac{dm}{dt} - \alpha\rho. \tag{2}$$

The value of dm/dt is determined by the kinetics of the process and may depend on temperature. In this case Eq. (2) must be supplemented by the heat balance equation.

In deriving Eq. (2) variation in the radius of the particles during the process was neglected.

By solving Eq. (2), i. e., obtaining the function $\rho(x, d, m, t)$, we can find the mean value of the degree of processing at any section along the length of the furnace $\overline{m}(x, d, t)$:

$$\overline{m} = \int m \rho \, dm / \int \rho \, dm. \qquad (3)$$

Expression (3) gives the relation between the distribution of the averaged degree of processing along the length of the furnace and the mixing coefficient, which can now be found after determining experimentally the relation $\overline{m}(x, d)$ in static regimes.

By varying the conditions of the experiment, it is possible to study the dependence of the coefficient D on different parameters.

In constructing the mathematical model it is possible to use another approach, which in a number of cases is more fruitful. The degree of processing m and particle density ρ are considered as functions of the coordinates x and t, and instead of (2) two equations are written, from which both functions are determined:

$$\frac{\partial \rho}{\partial t} = D \frac{\partial^2 \rho}{\partial x^2} - \alpha \rho; \qquad (4)$$

$$\frac{\partial (m \rho)}{\partial t} = D \frac{\partial^2 (m \rho)}{\partial x^2} + q_s - \alpha m \rho. \qquad (5)$$

This approach is more convenient in the sense that solution of equations (4) and (5) immediately gives the averaged degree of processing in the coordinates x, d, t.

Thus by determining for a specific process the source density, carryover factor, and the boundary conditions needed to supplement equations (4), (5), one acquires the possibility of determining the mixing coefficient D (d) from a comparison of the experimental curves m (x, d) with the theoretical.

Below we present an example of determination of the mixing coefficient using a mathematical model of the drying process for a polydisperse material in a fluidized bed. The degree of processing is the moisture content of the particles relative to the weight of dry material, which we will denote by w.

The experimental investigations necessary for refining the mathematical model of the drying process and recording the moisture content distribution along the length of the oven were carried out on industrial equipment. Material was charged at one end of the oven and discharged at the other, i. e., the drying oven represented a typical system with distributed parameters. Discharge was over a baffle plate 280 mm high.

The basic parameters of the drying process were: solid throughput 1400-2000 kg/hr; mass flow rate of heat transfer agent 2340-3150 kg/hr. m^2; velocity of heat transfer agent referred to free cross section of oven 1. 0-1. 4 nm/sec; initial moisture content of material 8-10%; final moisture content 0. 5-2. 0%; initial temperature of heat transfer agent 200-300°C; mean diameter of processed granules of copper-zinc concentrate 0. 5-4 mm.

In constructing the mathematical model it was assumed that all the particles belonged to one of three fractions containing only particles of equal size. To fraction I we assigned particles between 0. 5 and 1. 5 mm in diameter (d_1 = 1 mm), to fraction II, those between 1. 5 and 2. 5 mm (d_2 = 2 mm), and to fraction III, those between 2. 5 and 4 mm (d_3 = 3 mm).

The carryover factor α was equal to $5 \cdot 10^{-3}$ $\sec^{-1}$ for fraction I; for fractions II and III $\alpha = 0$.

The boundary conditions were determined by the experimental conditions and the process of charging and discharging. In our case of continuous operation the boundary condition at x = 0 must be given in the form of an input flux

$$D \left. \frac{\partial \rho}{\partial x} \right|_{x=0} = \Phi. \qquad (6)$$

The presence of an overflow baffle at the second boundary determines the boundary conditi [illegible] = l

$$\rho |_{x=l} = \rho_0. \qquad (7)$$

The question of source density reduces essentially to a question of drying rate, i. e., to the question: how much heat is absorbed by a granule in unit time and how much of this heat goes into the evaporation of water?

The quantity of heat absorbed by a granule in unit time is determined by the heat transfer coefficient, and the question of how much heat goes into the evaporation of moisture is equivalent to the question of the drying kinetics of the granules.

Two cases are possible [4]: 1) the drying rate is limited by the amount of heat supplied in unit time to unit surface area of the granule; the temperature of the material is equal to the wet-bulb temperature at which evaporation occurs (external problem); 2) the drying rate is limited by diffusion of water to the evaporating surface; the temperature of the material is not equal to the wet-bulb temperature, but gradually increases; the evaporating surface sinks into the granule (internal problem).

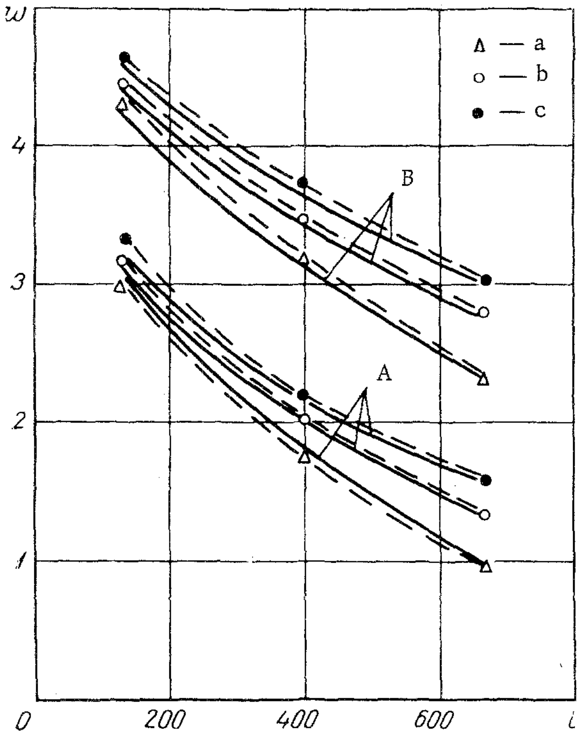

Fig. 1. Distrubution of moisture content in different fractions of material along the length of the oven. (T_0 – 250°C; fraction I – 25%, fraction II – 35%, fraction III – 40%; $G_m = 0.435$ kg/sec – A and 0.522 kg/sec – B): a) Fraction I; b) II; c) III. Continuous curve – theoretical; broken curve – experimental.

Observations of the temperature of the fluidized bed during the experiments indicated that at final moisture contents of not less than 0.5-1.5%, which satisfies the technological requirements of the process, drying of the granules may be treated as an external problem.

Investigations showed that with a bed height of 280-300 mm, the heat transfer agent is able to give up all its heat, and leaves the bed at a temperature equal to the wet-bulb temperature.

Under these conditions the source density is given by the following expression:

$$q_s = -\frac{G_h c_h (T_0 - T_{wb})}{r_v a(d)} \; \frac{d^2 \rho}{\sum\limits_d d^2 \rho} .$$

It is assumed that the heat is divided among the fractions in proportion to their total surface in the bed.

If account is taken of the heat expended on heating material entering the bed at a temperature T_{m0} to the wet-bulb temperature T_{wb} and if it is assumed, for simplicity, that during heating there is no evaporation of moisture from the material, then the source density must be determined as follows:

$$q_s = \begin{cases} 0; & x \leq x' \\ -\dfrac{G_h c_h (T_0 - T_{wb})}{r_v a(d)} \; \dfrac{d^2 \rho}{\sum\limits_d d^2 \rho} ; & x > x', \end{cases} \tag{8}$$

where

$$x' = \frac{(c_m + c'_w w_{in})(T_{wb} - T_{m0}) G_m}{G_h c_h (T_0 - T_{wb})} .$$

Thus, substituting the values found for the carryover factor and source density (8) in Eqs. (4), (5) and supplementing them with boundary conditions (6), (7), we obtain a mathematical model of the drying process for a polydisperse material which is completely determinate except for the mixing coefficient.

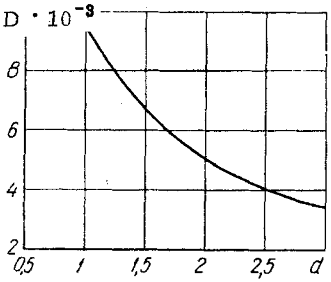

Fig. 2. Mixing coefficient as a function of particle size.

We solve Eqs. (4), (5) under static conditions with different mixing coefficients D(d), and by comparing the obtained curves w(x, d) with the corresponding experimental ones we determine the mixing coefficient.

For specific fluidization conditions (v = 1.04 nm/sec, d = 2.25-2.6 mm) the following values of the mixing coefficient were obtained (Fig. 1): for fraction I, $D = 9.2 \cdot 10^{-3}$; for II, $D = 4.95 \cdot 10^{-3}$; for III, $D = 3.54 \cdot 10^{-3}$ m^2/sec.

Figure 1 also shows the agreement between the experimental curves for the moisture content distribution in different fractions of the material along the length of the oven and theoretical curves obtained using the calculated mixing coefficients for other conditions.

The obtained dependence of mixing coefficient on particle size is presented in Fig. 2. In the case examined this dependence may be expressed approximately as $D \sim 1/d$.

NOTATION

Φ – solid particle flux; D – mixing coefficient; ρ – particle density; m – degree of processing; d – particle diameter; $\alpha(d)$ – weight of individual particle; $r = \{x, y, z\}$ –radius-vector of point in fluidized bed; x – horizontal coordinate; t – time; α – carryover factor; w –particle moisture content; G_h – mass flow rate of heat transfer agent; C_h – specific heat of heat transfer agent; T_0 – initial temperature of heat carrier; T_{wb} – wet-bulb temperature; r_v – specific heat of vaporization; c_m – specific heat of dry material; c'_w – specific heat of water; G_m – mass flow rate of material; v – velocity of heat transfer agent referred to free cross section of oven; q_s – corresponding source density.

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