

COEFFICIENTS OF MASS TRANSFER BETWEEN A  
FLUIDIZED BED AND THE SURFACE OF A  
CAPILLARY-POROUS BODY

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Results are presented from a theoretical determination of coefficients of mass transfer between a fluidized bed of porous particles and a capillary-porous body.

In the production of ceramics made by hot pressing, the most important stage – and the one which most determines the quality of the finished product – is the removal of the process (usually paraffin) binder [1]. Conducting this operation in a fluidized bed of porous particles is very promising, since the normal (for such a bed) rate of mass transfer between the dispersed material and the washed surface of the submerged body makes it possible to shorten the process considerably [2, 3].

When products are placed in a fluidized bed in which the temperature is below the temperature at which the binder is rapidly evaporated, the latter is removed mainly through capillary impregnation of porous particles of the bed which come briefly into contact with the product surface. Obviously, the kinetics of the process as a whole will be limited, first, by the rate of penetration of the binder from the product into the particles and, second, by the rate of replacement of the particles at the product surface and the transfer of these particles to the core of the bed.

The kinetics of the first of the above processes was examined in [4] for ceramics with a pronounced bimodal pore distribution according to size. Here, the following expression was obtained for the flow of a mass of liquid binder from a unit area of the surface of the body to the dispersed material of the bed:

$$i \approx \rho \varepsilon_p^2 \varepsilon_b \varepsilon'_b r (1-f) n F (\tau/8\mu)^{1/2} (p_p - p_b)^{1/2}. \quad (1)$$

In this equation, the capillary potentials of the particles and the material of the product are expressed in standard fashion through the characteristic radii of the pores

$$p_p = \frac{2 \sigma \cos \theta_p}{r}, \quad p_b = \frac{2 \sigma \cos \theta_b}{R_j}, \quad (2)$$

where the values of  $j = 1, 2$  pertain to pores corresponding to the two maximums in the above-noted bimodal distribution.

At the beginning of the first stage of the process, when the product contains a substantial quantity of liquid, the latter is actually removed only from the surface of fine capillaries. Thus, in Eq. (2), here we have  $j = 1$  (for the sake of definiteness, it is assumed that  $R_1 > R_2$ ). Meanwhile, in the broad capillaries the minisci are receding deeper into the porous body. At the end of the first stage, the formation of minisci in the narrow pores has been completed and they too have begun to recede from the surface of the body. At this moment, the potential on the product surface is equal to the capillary potential of the narrow pores and  $j = 2$  [4].

The flow  $i$  in Eq. (1) depends not only on the parameters characterizing the physical properties of the investigated materials and the pore structure of the particles and product, but also on quantities determined by gasdynamic conditions close to the surface of the submerged product. The following estimates have been proposed for these quantities as well as for  $F$  in [4] (see also [5])

$$n \approx \frac{3}{8} \frac{\rho_d}{\pi a^3} v, \quad \tau \approx \frac{a}{v}, \quad \rho_d = 1 - \varepsilon_d, \quad F = \frac{a^2}{4}. \quad (3)$$

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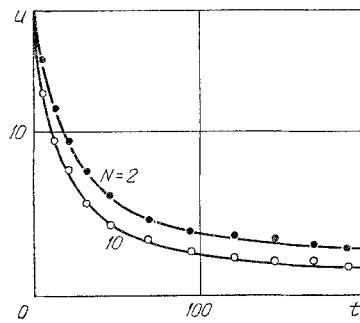


Fig. 1

Fig. 1. Kinetic curves of the process of paraffin removal from ceramic disks in a fluidized bed of alumina particles at 80°C and different values of  $N$  (numbers next to curves); the content of paraffin in the specimen is measured in percentages of the weight of the dry specimen;  $t$ , min;  $u$ , %.

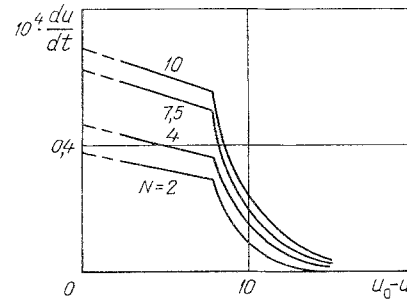


Fig. 2

Fig. 2. Dependence of rate of paraffin removal on total amount of paraffin removed at different  $N$  (numbers next to curves).  $10^4 \cdot du/dt$ ,  $\text{sec}^{-1}$ ;  $u_0 - u$ , %.

The rate of replacement of particles at the surface and the rate of their removal to the core of the bed are determined by the values of these quantities, which in turn are dependent on the intensity of the pseudo-turbulent pulsations of particles in the bed. We have the following expression, based on the theory in [6], for the mean square component of the fluctuation velocity of the particles in the direction normal to the surface:

$$v^2 \approx 0,8 \frac{\rho_d}{\varepsilon_d} \left(1 - \frac{\rho_d}{\rho_m}\right)^{1/2} \frac{1 + 2\rho_d + 0.0466 N \text{Re}_m}{\rho_d + 0.0233 N \text{Re}_m} I_2 a g, \quad (4)$$

$$\rho_d = 1 - \varepsilon_d, \quad \rho_m = 1 - \varepsilon_m,$$

where  $I_2$  is a coefficient which is also dependent on  $\varepsilon_d$  and  $N \text{Re}_m$ , the expression for which is presented in [6].

Equations (1)-(4) determine the coefficient of the mass transfer of liquid binder between the dispersed material of the bed and the surface of the body submerged in the bed:

$$\beta = \frac{i}{(\rho_p - \rho_b)^{1/2}}. \quad (5)$$

In particular, by assigning values to the various physical parameters which affect the process, we can in principle construct corresponding relations describing the dependence of coefficient  $\beta$  on the main process parameter - the number of fluidizations  $N$ .

The chief difficulty in using Eq. (4) to theoretically determine coefficient (5) is the lack of accurate information on the dependence of the porosity  $\varepsilon_d$  of the dense phase of the bed ( $\varepsilon_d + \rho_d = 1$ ) on the number of fluidizations and the physical parameters. If the two-phase theory were rigorously proven, then we would have  $\varepsilon_d = \varepsilon_m$  and  $v = 0$ . In calculations below, we use the empirical estimate  $\varepsilon_d \approx 0.45$ , which follows for the beds of alumina particles that will be examined (with  $\varepsilon_m \approx 0.40$ ) from the results obtained in [7] and from the theoretical dependence of  $\varepsilon_d$  on  $N$  and  $\text{Re}_m$  for low beds obtained in [8]. As for the quantity  $f$  in Eq. (1), it is assumed that it is slightly dependent on  $N$  in the interval of values of  $N$  which is of interest and is equal to approximately 0.2 [7, 9]; in case the theory from [8] is used, this quantity may be evaluated on the basis of this very theory.

In order to check the validity of the proposed simple model of the process of "external" transport of a liquid binder, we set up a series of experiments involving the removal of liquid paraffin from pores in semi-finished steatite ceramics (disks 6 cm in diameter and 0.5 cm thick) in a fluidized bed of porous alumina particles. The bed was heated beforehand to 60-110°C (the results presented below pertain to a bed with a temperature of 80°C). The conditions of the experiments corresponded to the following parameter values:  $\varepsilon_p = 0.4$ ;  $\varepsilon_b = 0.5$ ;  $\varepsilon_b^1 = 0.25$ ;  $r = 2 \cdot 10^{-6}$  cm;  $a = 5 \cdot 10^{-3}$  cm;  $\rho = 820$  kg/m<sup>3</sup>;  $\mu = 1.05 \cdot 10^{-2}$  N·sec/m<sup>2</sup>;  $R_2 = 0.5 \cdot 10^{-5}$  cm;  $R_1 = 10^{-3}$  cm. These values correspond to the theoretical formula below, which follows from (1)-(5):

$$\beta = 4.73 \cdot 10^{-5} (1 - f) (1 - \varepsilon_d) v^{1/2}, \quad (6)$$

where the coefficient  $4.73 \cdot 10^{-5}$  has the dimension of  $(\text{kg} \cdot \text{sec})^{0.5} / \text{m}^2$ .

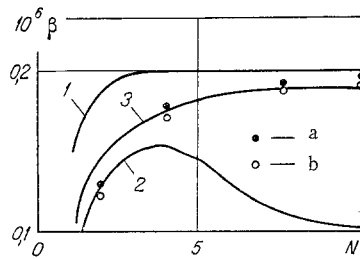


Fig. 3

Fig. 3. Mass-transfer coefficient in relation to  $N$ : a, b) values calculated from experimental data corresponding respectively to the beginning of the first and second periods of the process; 1-3) theory (explanations in text);  $10^6 \beta$ ,  $(\text{kg} \cdot \text{sec})^{0.5}/\text{m}^2$ .

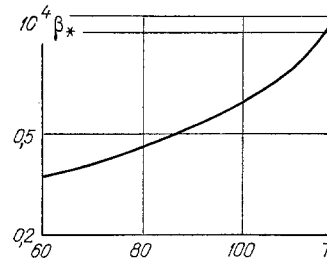


Fig. 4

Fig. 4. Dependence of corrected mass-transfer coefficients  $\beta_*$  on bed temperature  $T$ ,  $^{\circ}\text{C}$ ;  $10^4 \beta_*$ ,  $\text{kg}^{0.5}/\text{m}^{1.5}$ .

In order to record kinetic curves of the process, the ceramic disks were periodically removed from the apparatus and analyzed for the content of residual binder. The initial kinetic curves – examples of which are shown in Fig. 1 – were also plotted in the coordinates removed paraffin – rate of paraffin removal. These modified curves are shown in Fig. 2. It is apparent that the process of removal of the liquid actually occurs in two stages, as was noted in [4].

Equation (2) for  $p_b$  is valid with  $j = 1$  at the very beginning of the process. This formula is also valid at the moment of the transition to the second stage, but with  $j = 2$ . The experimental determination obviously requires the use of the quantity  $du/dt$ , which is determined at the moments of time just noted. The coefficient  $\beta$  can be calculated from the experiment by using the formula

$$\beta = \frac{\rho_0 V}{S} (p_p - p_b)^{-1/2} \left| \frac{du}{dt} \right|, \quad (7)$$

which was obtained from definition (5) and the obvious expression

$$i = \frac{\rho_0 V}{S} \left| \frac{du}{dt} \right| \quad (8)$$

for the mass flow of a liquid from a unit area of a submerged surface. For the ceramic used,  $\rho_0 = 2100 \text{ kg/m}^3$ .

The points in Fig. 3 represent values of the mass-transfer coefficient when determined according to Eq. (7) at different fluidization rates. Also shown are the curves  $\beta = \beta(N)$  plotted in accordance with the model presented above. Curve 1 was constructed on the basis of an empirical assumption that the porosity of the dense phase is independent of the number of fluidizations and is roughly equal to 0.45 (with a porosity of 0.40 at the initial moment of fluidization), and that  $f \approx 0.2$ . (This approximation may be regarded as corresponding to a certain generalization of orthodox two-phase fluidization theory – in accordance with one of the postulates of the two-phase theory, the value of  $\varepsilon_d$  does not change with an increase in  $N$  within a certain interval  $N > 1$ , but in contrast to its other postulate  $\varepsilon_d$  is not equal to  $\varepsilon_m$ .) Curve 2 corresponds to a uniform expansion of the bed with an increase in  $N$  (not seen in the tests described here). Finally, the more realistic curve 3 was plotted for an inhomogeneous fluidized bed with variables  $\varepsilon_d$  and  $f$  evaluated in accordance with the results of the theory in [8]. Here, for the mean porosity of the bed as a whole (with allowance for the bubble phase), we used the empirical result obtained in [7]. This result, valid for a bed of alumina particles, expresses porosity in the form of a function of  $N$  and the Archimedes criterion. As might be expected, the last curve describes the test data considerably better than do the other curves.

The results of a comparison of the theory and the experiment in Fig. 3 allow us to recommend the above model, based on the results in [4, 6], for engineering calculations pertinent to the process of removing a processing binder from porous products. A similar conclusion follows from analyzing tests conducted at other bed temperatures. Figure 4 shows the temperature dependence of the quantity

$$\beta_* = \frac{\beta}{(1-f)(1-\varepsilon_d)v^{1/2}}, \quad (9)$$

determined for the interval of values of  $N$  in which  $\beta \approx \text{const}$ . We should also note that the good agreement between the theory and the experiment may be regarded as a certain indirect proof of the adequacy of the theoretical conclusions in [8].

In conclusion, let us make several remarks concerning limitations with respect to the validity of the model. First of all, we completely ignored removal of the liquid due to evaporation and convection diffusion of the vapors into the bed. These factors were examined in [10]. This means that the temperature of the bed should not be too close to the temperature at which there is substantial evaporation of the liquid in the pores. We also ignored the formation of "crusts" of particles adhering to the surface of the product being treated. These crusts may play a dual role: on the one hand, their formation promotes more rapid removal of liquid from the product pores; on the other hand, they create additional resistance to the transfer of liquid to the particles in the fluidized state. It is possible that this factor was in some measure responsible for the discrepancy between the theory and the test data at low values of  $N$ , when the action of the fluidized particles may not have been sufficient to remove the crust.

Finally, the above model is based on the same representations regarding the approach of "packets" of the dense phase to the product surface as are used in the packet model of external heat transfer [11]. In actuality, there is a zone near the surface of a body submerged in a fluidized bed characterized by intensified breakthrough of the fluidized medium. This zone has special properties – properties which are different from the properties of the bed itself or its dense phase [12]. This circumstance ultimately leads to a change in the value of  $f$ . However, in contrast to the consequences suggested, e.g., by the results in [5], the existence of this zone does not fundamentally alter the description of the basic physical mechanism of the transfer process.

#### NOTATION

$a$ , particle radius;  $F$ , area of contact of particles with the surface of the body;  $f$ , percentage of area of surface of product in contact with the bubble phase;  $g$ , acceleration due to gravity;  $i$ , flow of liquid mass from a unit area of the surface;  $N$ , number of fluidizations;  $n$ , number of particles coming into contact with a surface of unit area per unit of time;  $p_p$ ,  $p_b$ , capillary potentials of particles and product;  $R_2$ ,  $R_1$ , radii of narrow and broad pores inside the product;  $r$ , radius of capillaries in the particles;  $S$ , area of the surface being treated;  $T$ , temperature of the bed;  $t$ , time of treatment;  $u$ , percentage content of liquid in the specimen;  $V$ , volume of the product being treated,  $v$ , mean square component of the fluctuation velocity of the particles in the direction normal to the surface;  $\beta$ ,  $\beta_*$ , standard and corrected mass-transfer coefficients determined from (5) and (9);  $\varepsilon_b$ ,  $\varepsilon_p'$ ,  $\varepsilon_p$ , porosities of product determined for all and for only the small pores and the porosity of the material of the particles;  $\varepsilon_d$ ,  $\varepsilon_m$ , porosity of the dense phase and the porosity of the bed in the state of minimum fluidization;  $\theta_b$ ,  $\theta_p$ , angles of wetting of the materials of the product and particles, respectively, by the liquid binder;  $\mu$ ,  $\rho$ , viscosity and density of the liquid;  $\rho_0$ , density of the dry product;  $\sigma$ , surface tension coefficient of the liquid;  $\tau$ , characteristic time of contact of particles with the surface;  $Re_m$ , Reynolds number corresponding to particle radius and minimum-bed-fluidization velocity [6].

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CALCULATING RADIANT HEAT EXCHANGE  
BETWEEN A FLUIDIZED BED AND A SURFACE

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Radiant heat transfer in a nonisothermal fluidized bed is calculated.

The contribution of radiant heat transfer becomes important in heat exchange between a high-temperature fluidized bed and a surface [1]. The radiant flux is usually calculated using a formula which is valid for two gray isothermal surfaces [2]:

$$q_r = \sigma \varepsilon_{cr} \left[ \left( \frac{T_{bc}}{100} \right)^4 - \left( \frac{T_{ss}}{100} \right)^4 \right], \quad (1)$$

where

$$\varepsilon_{cr} = \frac{1}{\varepsilon_{ss}} + \frac{1}{\varepsilon_{bc}} - 1.$$

A certain average steady-state temperature profile is formed in the process of heat transfer close to the submerged body. The effect of this profile is taken into account by replacing  $\varepsilon_{bc}$  with the effective value of the degree of blackness of the fluidized bed [3]:

$$\varepsilon_e = \varepsilon_e(T_{ss}, T_{bc}, \varepsilon_p). \quad (2)$$

There are presently no methods which make it possible, after assigning values to  $\varepsilon_p$ ,  $T_{ss}$ , and  $T_{bc}$ , to calculate the temperature profile between the surface and the core of the bed, the function  $\varepsilon_e$ , and radiant flux without resorting to special measurements of the intensity of the radiation from the bed [3].

This article proposes the calculation of these characteristics on the basis of the model described in [4]. The nonisothermal zone of the bed between the surface and the bed core is represented by a set of  $N$  parallel translucent isothermal (since the thermal resistance is concentrated mainly in the gas interlayers [1]) planes with reflection coefficients  $r$  and transmission factors  $\tau$  (Fig. 1). The surface submerged in the bed is represented in the model by the 0-th plane, with reflection coefficient  $r_{ss}$  and temperature  $T_{ss}$ . The bed core is represented by the  $N + 1$ -st plane, with the parameters  $r_{bc}$ ,  $T_{bc}$ . It is assumed that the thickness of the bed is sufficiently great so that  $\tau_{bc} = 0$ . The coefficients  $r$ ,  $\tau$ , and  $r_{bc}$  were computed for an assigned blackness of the fluidized particles from equations of [4].

As a first approximation, let us examine the simple case whereby energy is transmitted in a system of  $N$  translucent planes by radiation alone. Under steady-state conditions, the energy balance equation for the  $i$ -th plane will have the form

$$2 \varepsilon_i \sigma \left( \frac{T_i}{100} \right)^4 = \varepsilon_i \sum_{k=0}^{N+1} (q_{ik}^- + q_{ik}^+)^{inc}. \quad (3)$$

Having expressed  $q_{ik}^{\pm inc}$  through the characteristics of the model, we can form the following system of equations relative to the temperature  $T_k$ :

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