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STATEMENT OF PROBLEMS ON HEAT AND MASS TRANSFER IN PROCESSES IN A DISPERSED SOLID PHASE

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The statement of problems on heat and mass transfer in dispersed processes in chemical engineering is analyzed. It is shown that the evaluation of the governing simplifying assumptions, the detailed description of experiments done, and the discussion of the obtained results are the important points of heat and mass transfer studies.

The statement of the problem is the main aspect of a study that predetermines the final result.

In experimental studies of chemical engineering processes, making measurements of the main parameters is most important, especially of those that are somewhat difficult to measure. So, for example, direct measurements of the surface temperature of particles that exchange heat with the vertical gas flow in tube-dryers, where irregular polydispersed particles are accelerated in a complicated manner by the gas flow and simultaneously rotate, are not successful [1]. Under such conditions the velocity of flow past the particles usually varies over a wide range in the direction of the dispersed material. Interpretation of the result on coefficients of heat transfer from the particle surface to the gas flow requires specific simplifying assumptions on the particle surface temperature, the velocity difference of the gas and particles, and the mean temperature difference of dispersed and solid phases, which were made in processing the obtained experimental data. Even the determination of the total interphase surface of irregular polydispersed particles is not an entirely trivial problem. In any case, whenever possible, a more detailed description of the experimental conditions, the assumptions made and the measurements is quite necessary to correctly evaluate and interpret the results obtained in one or another work.

Furthermore, the statement of the problem that will be mathematically described and then analyzed (analytically or numerically) is of great importance. Even if the description of one or another process that is arrived at is adopted just to obtain dimensionless groups, as is done in processing experimental data by the similarity method, a physical defect in the initial description will usually result in an incorrect interpretation of the data in this case, as well. On this point, the inclusion of the boundary conditions of the process under study into the main differential equation governing it is insufficiently grounded. Physically, this replacement of the condition at the boundary of the region, where the process develops (e.g., heat or mass flux), by a source (heat or mass) at each point of the region under consideration is not entirely justified. Therefore, the substantiation here must be very convincing [2, 3].

Generally speaking, the evaluation of the assumptions adopted is obligatory or, at least, desirable. However, in the majority of engineering works this important point for the problem statement is usually absent. In general, it is better to make such estimates using a more detailed model for the process under consideration; then how valid this is will depend only on the reliability of the model that is available [4]. Unfortunately, more often than not, a more detailed model of the complex process is not available, and the simplifying assumptions to be adopted must be evaluated by analyzing the references or one's own preliminary data. In any case, our assertion that in the majority of cases, the experiments conducted, taken together, are evidence in favor of the adequacy of model representations adopted in the work is insufficiently convincing.

Not always do we have a clear concept of the equivalent diameter assumed for irregular particles. This is especially characteristic of complex heat and mass transfer processes where flow past polydispersed particles or their suspension in continuous flow produces a strong effect (fluidization and pneumatic transportation at heat and mass,

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etc.). Indeed, the lag of the internal fields of potentials of heat and mass transfer depends on the particles to the power 2 being in the Fourier number, the intensity of external heat and mass transfer is governed by the particle size to a lesser degree than by the flow condition, and the particle suspension conditions are related by the Archimedes number entering the size to the power 3. These and other contingencies make the problem on using one particular (e.g., volume-equal) equivalent diameter of irregular particles poorly substantiated. Possibly, it would be more valid to use not one but several sizes, each of which would be representative of the corresponding physical aspect of the complex process. Polydispersed material fraction averaging of such sizes would yield a better substantiated weight coefficient. Needless to say, it is difficult to operate with several sizes of one and the same object.

Processing dispersed particles in the continuous flow by heat and mass transfer (processes of thermal drying, adsorption, heterogeneous catalysis, crystallization, solution, etc.) entails certain difficulties in correctly interpreting the experimental data on particle kinetics. The problem is that the phase contact surface of fine dispersed materials is very developed, as a result of which transfer potential values (temperature, concentration) in a solid phase vary substantially in the direction of the phase displacement. Hence, only the very first particle bed contacts a solid phase that has a constant temperature (or concentration) while all the other particle beds contact a solid phase whose potential values vary in the phase direction and in time. These variations are functions of the heat and mass transfer process under consideration. In such a situation, it is not simple to decide to what temperature (concentration) of the solid phase the curve obtained for the kinetics of development of the dispersed material should be referred. For coarse particles (of the order of 3 or more millimeters) the solution would be to deal with a bed one grain thick, in passing through which the solid phase practically does not change its parameters. For fluidized bed conditions there are some artificial methods to stabilize the mean temperature of a drying agent, which contacts the particles of the real-height fluidized bed [5], e.g., the kinetic drying curve is to be obtained. However, the smaller the particles, the more difficult it is to keep a constant temperature of a drying agent near the particle surface during the entire kinetic experiment.

In the majority of continuous dispersed processes, each particle contacts the solid phase, so that the temperature (concentration) in the solid phase near the particle surface would vary in time. Such a variation is most typical of moving dispersed bed apparatuses. For regular particles analytical solutions to the problems on the diffusional propagation of heat or concentration inside particles exist, which consider a constantly varying value of the appropriate transfer potential in a continuous medium. In what follows, the variation of the potential external to the particles is not an independently assigned time function but is formed by interaction of the phases, i.e., it must also be obtained from the solution of the problem [6, 7]. The solutions are tedious, slowly converging series, and the problem on combined convective drying heat and mass transfer under moving bed conditions requires some extra simplifying assumption, implying that a linear dependence exists between the equilibrium values of the parameters of the particles and the drying agent [7]. In more real cases, when the heat and mass transfer processes inside capillary-porous particles cannot be satisfactorily discribed by the simple laws of heat conduction or diffusion, and the kinetics of heat and mass transfer of a particular polydispersed material composed of irregular particles has been preliminarily obtained in experiments with constant external potentials, it is difficult to allow for a constant change of external conditions in a real apparatus because it is not easy to estimate the response the particles entering, at each time moment, a medium with a different temperature and concentration to such a change of heat and mass transfer kinetics. It is obvious only that the more coarse the particles, the greater is the lag of their internal potential transfer properties, and that the more rapidly the external conditions vary, the more significant their influence on the rate of internal heat and mass transfer processes.

In some papers submitted to the Session "Heat and Mass Transfer in Chemical Processes" [8-10, etc.] as well as in many published articles "optimization" of one or another process using the obtained experimental data or the developed model concepts was vaguely mentioned. Unfortunately, such mention of optimization is not informative and frequently does not indicate optimization in the current sense of this term. This means that no optimization has been done, and only several possible versions of the process under study have been analyzed and compared in some particular parameters. We would like to see a more rigorous statement and analysis of the problems on optimizing chemical engineering processes based, first of all, on a more precise feasible (or any particular) concept of the optimality criterion adopted, e.g., the expenses incurred, and on well-developed methods for searching for an extremum of a numerical value of the optimality criterion.

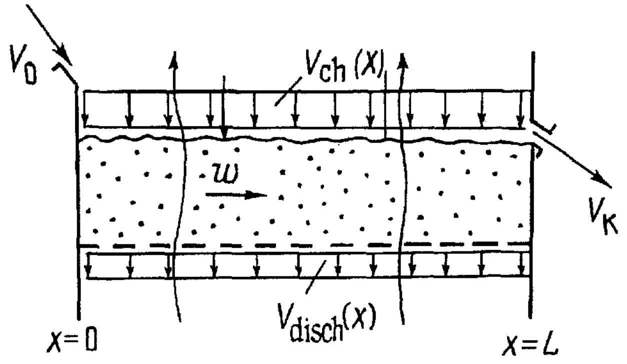

Fig. 1. Flow pattern of dispersed and solid phases in a fluidized bed apparatus.

Continuous interactions of flows of continuous (usually gas) and dispersed phases can be arranged so that the flows cross, and the flow of suspended dispersed material is characterized by a noticeable mixing of particles in their resultant direction (Fig. 1). Such a mixing influences the final interphase interaction to a considerable extent (during drying, final moisture content of the dispersed material); it is random in nature and, more often than not, it is described by the diffusional model with a constant quasidiffusional mixing coefficient. In this case, the differential equation for the moisture content distribution u(x) of the dispersed material in the x-direction of its mass mixing with the mean flowrate velocity w is written in the following traditional form:

$$\frac{d}{dx}\left(D\frac{du}{dx}\right)-\frac{d}{dx}(wu)=\left|\frac{du}{d\tau}\right|. \tag{1}$$

For D, w = const the equation reduces to

$$D\frac{d^2u}{dx^2}-w\frac{du}{dx}-\left|\frac{du}{d\tau}\right|=0, \tag{2}$$

where D is the quasidiffusion coefficient of the dispersed material; $|du/d\tau|$ is the drying rate of individual particles of the monodispersed material or a representative sample of the polydispersed product determined in separate kinetic experiments at different temperatures and velocities of the drying agent [9]. In integrating equation (2) the quantity $|du/d\tau|$ is taken as a known parameter that either has a constant value (period of constant drying rate for the particle) or is definitely dependent on a desired value of the moisture content of the material (period of linearly decreasing drying rate).

To find the integration constants of equation (2) two boundary conditions must be assigned, which are usually written in the form of the so-called Danckwerts conditions taken from the methods for analyzing the operation of catalytic reactors with a fixed dispersed catalyst bed and a continuous reacting mixture flow filtrating through this bed. It is assumed that diffusional longitudinal mixing can occur in the continuous mixture flow and that a stepwise concentration decrease in the flow exists at the bed entrance (x = 0) due to the abrupt onset of diffusional transfer of the desired component (Fig. 2). This allows one to write a boundary condition for the balance of the desired component in the continuous flow at x = 0

$$wu_0=wu|_{x=0}-D\left.\frac{du}{dx}\right|_{x=0}. \tag{3}$$

Following the same logic, the boundary condition for the balance of the desired component at the bed exit (x = L) is assumed to be in the form of the following relation:

$$-D\left.\frac{du}{dx}\right|_{x=L}+wu|_{x=L}=wu_{\text{к}}, \tag{4}$$

where u_k is the moisture content of the dispersed flow at the apparatus exit. Condition (4) corresponds to a stepwise increase of the moisture content of the material flow at the coordinate x = L (Fig. 2). While a concentration change

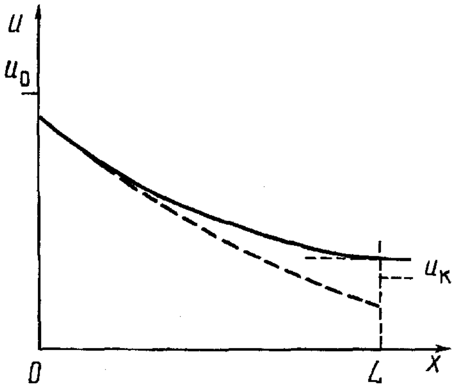

Fig. 2. Distribution of moisture content of dispersed material in the direction of the material flow: solid line, boundary conditions (3) and (5); dashed line, conditions (3) and (4).

at the entrance is not particularly excluded, a positive change at the entrance is considered to be physically impossible, and condition (4) is replaced by the of the homogeneous condition second kind

$$\left.\frac{du}{dx}\right|_{x=L} = 0. \tag{5}$$

It should be noted that the physical meaning of condition (5) by itself raises doubts since it is difficult to imagine that the dispersed flow must "know" that it should reach a certain coordinate x = L with a zero longitudinal moisture content gradient. Condition (5) may usually be considered to be approximately satisfied for chemical reactors with a large catalyst bed and a relatively high degree of conversion of the chemical reaction in the continuous reacting mixture flow (the extent of reaction completion asymptotically tends to zero); however, for some processes where the main point is associated with the completion of the dispersed flow (drying, adsorption, extraction from solid particles, crystallization or dissolving), the extent of process completion falls far short of approaching the limiting one, especially at the exit from a bed of small length in the direction of the dispersed flow (for continuous crystallization, e.g., such limiting completion cannot exist in general). Therefore, for the processes, whose technological meaning consists in treating the dispersed phase, boundary conditions (3) and (5) can be used only to a first approximation for not very large quantities L, w and for sufficiently rapid kinetics of the process. The simplicity and, to some extent, the traditional representation of conditions (3) and (5) allow them to be used for analyzing separate nontrivial modes of interactions of crossflows of the drying agent and the moist dispersed material. So, it is known, e.g., that for the point feed of the lumpy moist material into a continuous drying suspended bed apparatus this material can be deposited near the feed place. This may be avoided if the charging feed is arranged along the apparatus. When the dispersed product is fed at several places along the apparatus, it is assumed that for a constant height of the entire fluidized bed its velocity along the apparatus will be increased jumpwise with growing total volume flowrate of the dispersed material along the apparatus. The mathematical description of such a process corresponds to differential equation (2) written for each individual bed section at different values of material velocity. At the places of the point feed of the moist material the conditions for continuity of the longitudinal distribution of the moisture content of the material and for merge moisture flow in a dispersed product, allowing for its local feed, are considered valid. Conditions (3) and (5) are assumed valid at x = 0 and x = L. The distribution of the moisture content of the material in the direction of the material can be obtained in a relatively simple form [10] but the cumbersomeness of the algebraic determination of the integration constants rapidly increases with increase of the number of sections.

The charging of moist material can be continuously distributed along the apparatus (see Fig. 1). The bed discharging can be continuous, too. In such a case, the directed displacement velocity of the dispersed material suspended in the bed will increase continuously:

$$w(x) = w_0 + \frac{1}{S}\int_0^x (V_{\text{ch}} - V_{\text{disch}})\,dx, \tag{6}$$

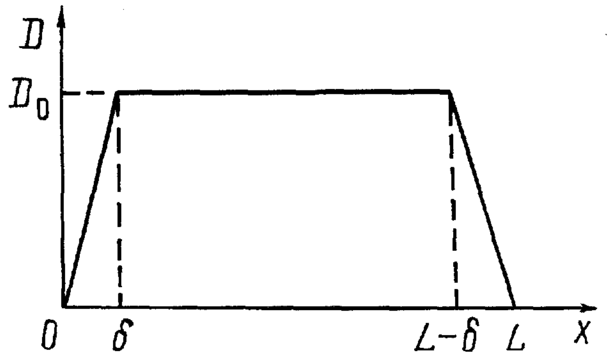

Fig. 3. Distribution assumed for the quasidiffusion coefficient in the dispersed flow direction.

where $V_{ch}(x)$ and $V_{disch}(x)$ are the volume charging and discharging of dispersed material based on unit length in the direction of the material, $m^3/(s \cdot m)$; S is the bed cross section normal to the material flow direction, m^2; $w_0 = V_0/S$ is the linear material flow velocity due to its point charging V_0, m^3/s, in the cross-section $x = 0$. A similar situation can take place when the volume of particles to be dried can noticeably change due to shrinkage or abrasion accompanied by the entrainment of the dust formed.

A variable velocity of the dispersed material reduces differential equation (1) to equations having an integration variable-dependent coefficient. This considerably narrows the possibility of obtaining analytical solutions in a closed form. For uniform material charging (discharging) along the bed or for a linear rate of change of the particle volume due to abrasion (shrinkage) Eq. (1) can be integrated in the form of Bessel or Whittaker functions: two integration constants are found from boundary conditions (3) and (5). After the moisture content distribution u(x) along the apparatus has been found, the mean moisturecontent of the discharged product is determined as:

$$\bar{u} = \frac{V_{\text{к}} u|_{x=L} + \int\limits_0^L (V_{ch} - V_{disch})\, u(x)\, dx}{V_{\text{к}} + \int\limits_0^L (V_{ch} - V_{disch})\, dx}. \tag{7}$$

As mentioned above, Danckwerts boundary conditions (3) and (5) are open to criticism from the viewpoint of their physical meaning. In one of the works submitted to Session No. 11 of the Second International Forum on Heat and Mass Transfer an attempt was made to analyze the problem on continuous drying of dispersed material in a cross spouting bed where the quasidiffusion coefficient was taken to be varying along the apparatus and was considered to be some known function of the longitudinal coordinate: D(x). An explicit form of D(x) was chosen, proceeding from the following considerations. Along the main suspended (spouting) bed, the value of the particle diffusion coefficient is apparently approximately constant: $D = D_0$, which is supported by the available experimental measurements [11]. In the immediate vicinity of the vertical end walls of the apparatus (at $x = 0$ and $x = L$), the random displacement of material particles or separate groups (aggregates) of particles is damped by the solid walls, and the values of D must tend to zero on the surfaces themselves ($x = 0$ and $x = L$). A linear decrease of D is both the most natural and the simplest case to be analyzed when the particles approach the end walls of the apparatus in some wall zone of thickness δ approximately equal to 10-15 particle sizes (Fig. 3).

Within the framework of assuming a parametric piecewise-continuous behavior of D(x) and constant values of the linear velocity w of the dispersed material along the coordinate x, the mathematical description of the continuous drying process corresponds to a system of three differential equations: for the central zone ($\delta \le x \le L - \delta$) it is equation (2) with constant coefficients (D_0 and w), and for the first ($0 \le x \le \delta$) and third ($L - \delta \le x \le L$) zones the second-order differential equation has a linearly varying coefficient ($D_0 x/\delta$ and $D_0(L-x)/\delta$). The conjugation conditions at the section boundaries for the moisture content fields and the moisture flows along the dispersed material can be written. Such a system of equations can be solved in a closed form in terms of the Whittaker and Kummer functions, and the experimental measurements performed for moisture content distributions of a number of dispersed materials (sodium sulfate-sulfite, some polymer materials, and carbon-containing dispersed products) have shown, first, the adequacy of the diffusional model for material mixing in the long spouting bed apparatus and, second, a noticeable influence of the nonconstant longitudinal material mixing coefficient on the final moisture content of the discharged material only at relatively small values of L and high particle drying rates $|du/d\tau|$.

In those cases, when it is not the moisture content distribution of the dispersed material along the suspended bed apparatus that is of interest, but only the mean value of the moisture content of the discharged product at $x = L$, the problem can be analyzed by the macrokinetic method in terms of the density distribution of particles during their residence time within the drying zone and by using the experimentally found drying curve of a representative portion of dispersed material: $\bar{u} = \int_0^\infty \rho(\tau)u(\tau)d\tau$. In the general case, the explicit form of the function $\rho(\tau)$ can be determined experimentally, e.g., by the well-known tracer method. The influence function corresponding to the diffusional propagation of a pulse-induced mark in an unbounded medium, with consideration of directed mixing of the medium with the velocity w is used as one of the approximations [12]. Then

$$\bar{u} = \frac{1}{2\sqrt{\pi D}} \int_0^\infty \frac{u(\tau)}{\sqrt{\tau}} \exp\left[-\frac{(L - w\tau)^2}{4D\tau}\right] d\tau, \tag{8}$$

where D is the constant coefficient of quasidiffusional mixing of a dispersed material; the value of the rapidly converging integral of the drying curve $u(\tau)$ known from the kinetic experiments is calculated by any approximate method.

It should be noted, however, that the macrokinetic method requires the use of the drying curve corresponding to the mean temperature of the drying agent along the fluidized bed. This temperature can be found only by using successive approximations. In this case, kinetic relation (8) and the heat and mass balance equations are solved simultaneously.

In conclusion, it should be noted that the statement of a particular problem, the estimation of the governing simplifying assumptions, the detailed description of the experiments conducted, as well as the discussion of the obtained results are natural and important stages of the problem on heat and mass transfer processes.

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