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MODEL OF COAL COMBUSTION IN A FLUIDIZED BED AND ITS EXPERIMENTAL IDENTIFICATION

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The author has formulated a system of one-dimensional steady-state differential equations for the balance of oxidizer, fuel and energy in the diffusion approximation. The model of coal combustion in a fluidized bed is identified from the experimental data, and the unknown parameters of the model describing the rate of oxidation of fuel and the intensity of gas and fuel transfer in the bed are determined.

Fluidization technology opens up possibility of efficient use of a wide range of solid fuels under conditions of increasing ecological requirements to protect the atmosphere. An obstacle to its further development is lack of understanding of the laws of coal combustion in a fluidized bed. The existing combustion models are based mainly on two-phase description of the hydrodynamics of the fluidized system [1-3]. In recent years attempts have been made to determine how large the factors may be, which has led to the development of very cumbersome models [1, 2] that are complex to analyze and especially to relate with the available experimental information on the system examined. It therefore seems desirable to use simpler and physically based models. These contain several unknown parameters which are then determined by comparing the results of theory and experiment (identification of the model) [4]. A simplified model of this kind was proposed in our work [5], where we formulated the equations of oxidizer balance in the continuum and discrete phases of a fluidized bed and the balance equation for the energy of the burning fuel particles. This incomplete model (it lacks the fuel balance equation) was matched or identified using the experimental values of carbon dioxide content in the fuel gases. With it one can analyze a number of regime parameters of the system, assuming that the concentration is constant over the height of the fluidized bed. With a generalized combustion model, based on the full system of fuel balance equations, one can study the fluidized system in greater detail.

We write the full system of balance equations:

for the fuel

$$C'' - m\xi \frac{B}{k} \varphi \text{У} C = 0, \tag{1}$$

for the oxidizer in the continuum phase

$$k\varepsilon \text{У}'' - \frac{\gamma}{N-1} \text{У}' - \xi\varphi B C \text{У} + P\text{П} (\text{У}_\text{п} - \text{У}) = 0, \tag{2}$$

for the oxidizer in the discrete phase

$$\frac{N-\gamma}{N-1} \text{У}_\text{п}' + P\text{П} (\text{У}_\text{п} - \text{У}) = 0 \tag{3}$$

and the energy equation for the hot fuel particles in the bed

$$(C\theta)'' + q\varphi m\xi B \text{У} C - \Phi B \frac{\varphi}{k} \theta C = 0. \tag{4}$$

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In this system of equations the rate of fuel combustion referenced to unit volume of the bed is varied in direct proportion to the product of oxidizer concentration and fuel concentration, since the number of particles (their area) describes the fuel combustion area. By the same reasoning the energy of unit volume of the bed is given by the product of the excess temperature of the hot particle and the fuel concentration. Therefore the energy flux due to the unordered motion (i.e., mixing) of the hot particles in the bed is proportional to the gradient of energy concentrated in unit volume of the bed.

We now introduce the following notation:

$$B=\frac{6K_*H}{d(U-U_0)},\quad m=\frac{\rho_*}{\rho_p(1-\varepsilon)},\quad \theta=\frac{T_p-T_b}{T_b-T_0}, \tag{5}$$

$$Pe=\frac{W}{U-U_0},\quad \Pi=\frac{\beta_\Pi H}{U-U_0},\quad q=\frac{Q_p^u}{c_4(T_b-T_0)}, \tag{6}$$

$$\Phi=\frac{\mathrm{Nu}\lambda/d_i+4\varepsilon_{re}C_0T_b^3}{K_*\rho_4c_4(1-\varepsilon)}. \tag{7}$$

The dimensionless group of Eq. (7) describes the convective-conductive and radiative heat transfer to the hot fuel particle surface in the fluidized bed. The group B is the dimensionless rate of combustion of the particle, and the ratio Π gives the gas transfer between the continuum and the discrete phases in the fluidized bed.

Since combustion of fuel in the furnace of a fluidized bed is accomplished with increased excess of air (greater than the values of Eqs. (1) and (2)), it is natural to postulate that oxidation of carbon is accomplished fully at the fuel particle surface, with formation of carbon dioxide. We shall use the known kinetics of this process [6], and compute the added impedance due to transfer of oxygen to the combustion surface [7], so that the relative rate of combustion of the carbon particle can be written as follows:

$$\frac{1}{\xi}=\frac{K_*d_i}{\mathrm{Sh}\,D}+\exp\frac{E}{RT_b}\left[\frac{1}{1+\theta(1-T_0/T_b)}-\frac{T_b}{T_*}\right]. \tag{8}$$

The parameters appearing in the dimensionless group were found from the empirical correlations of [8-10]:

$$\mathrm{Nu}=0{,}41\,\mathrm{Ar}^{0,3}\left(\frac{d_i}{d}\right)^{0,2}\left(\frac{\rho_4}{\rho_i}\right)^{0,07},$$
$$\mathrm{Sh}=2{,}78\,(\mathrm{Re\,Sc})^{1/3}\left(\frac{\rho_4}{\rho_i}\right)^{0,15}\left(\frac{d_i}{d}\right)^{0,13}, \tag{9}$$
$$\varepsilon=\varepsilon_0(1-0{,}5\,\mathrm{Fr}^{0,27}).$$

It was established experimentally that the stochastic motion of the gas is analogous to the motion of material in the fluidized bed and that the energy spectra of the velocity fluctuations of the particles and gas are similar [11, 12]. This gives a basis for asserting that in the diffusion approximation the stationary transfer of gas and material in a fluidized bed can be described by a single effective mixing coefficient which we will determine from the empirical correlation obtained by generalizing the experiments on mixing of the solid phase [13]:

$$D_T=D_g=k(U-U_0). \tag{10}$$

The gas transfer between the continuum and the discrete (bubble) phases in a fluidized bed, which is described by an exchange transfer coefficient, was determined from the empirical relation [14]

$$\beta_\Pi=520\left(\frac{U^2\rho}{gH\rho_i}\right)^{0,53}\mathrm{Ar}^{-0,15}. \tag{11}$$

TABLE 1. Parameters for the Model of Coal Combustion in a Fluidized Bed

Experiment	$E, \frac{kJ}{mole}$	φ	k	P	n	Errors σ_c	Errors σ_x
ITMO* [5]	109,2	1,75	0,107	5,09	2,1	0,74	0,995
ITMO	118	6,26	0,08	4,58	2,3	0,51	0,685
TsKTI	139	5,8	0,07	4,46	2,3	0,62	0,71

*Simplified model [5].

In the assumed model the gas filters through the continuum phase with a velocity greater than that necessary to begin fluidization ($\gamma > 1$), which generates the appropriate transfer of oxidizer (specified by the model) in the continuum (emulsion) phase with velocity γ and transfer in the discrete phase with velocity $N-\gamma$:

$$\gamma = (1 - nN)/(n-1). \tag{12}$$

The boundary conditions of the problem are:

$$x = 0,\ \frac{dY}{dx} = \frac{\gamma}{k(N-1)}(Y - Y_0),\ Y_{\text{п}} = 0{,}21,\ \frac{d(C\theta)}{dx} = 0; \tag{13}$$

$$x = 1,\ \frac{dC}{dx} = \frac{\text{Pe}}{k}(1 - C),\ \frac{dY}{dx} = 0,\ \frac{d(C\theta)}{dx} = -\frac{\text{Pe}}{k}(1 - C\theta). \tag{14}$$

At the lower boundary of the bed in the discrete phase (bubbles) air arrives with atmospheric oxygen content ($Y_0 = 0.21$), while the oxygen concentration in the continuum phase is summed according to the Dankwerst condition. Also, there is no diffusion flux of material and energy (hot particles) through the lower boundary. The fuel reaches the top surface of the bed and generates a fixed flux of mass and a negative energy flux, given by boundary conditions of the third kind.

This problem of fuel combustion in a fluidized bed, Eqs. (1)-(4), (13) and (14) was reduced to a system of seven differential equations of first order and was solved numerically by a matrix marching method. The solution of the problem depends on a number of unknown parameters: the activation energy of the combustion reaction (E), the mixing coefficient (k) from Eq. (10), the mean relative size of the hot particle ($1/\phi$), the coefficient (n) describing the velocity of filtering through the continuum phase (γ), and a coefficient giving inter-phase gas transfer (P), found from the existing experimental data. We used tests on fuel combustion on the laboratory equipment of ITMO, Academy of Sciences of the BSSR with a column of diameter 140 mm [15, 16], and also test data from the DKVr-6.5-13 boiler with a fluidized bed furnace, obtained by NPO of TsKTI [17]. Theory and experiment were compared according to the mean fuel concentration in the fluidized bed, since this quantity was determined with the least accuracy. As a target function we chose the square of the difference of the experimental and theoretical values (mean over the height of the bed):

$$F = \sum_{e=1}^{M} (C_{\text{e}}^{l} - C_{\text{p}}^{l})^2. \tag{15}$$

The unknown model parameters were found from the condition of a minimum of the target function, when the deviation of the theoretical values from the experimental is a minimum. The problem was solved numerically by the compressive multiboundary method. Thus the identification of the coal combustion model in a fluidized bed was accomplished.

The computed results are shown in Table 1. It shows the model parameters computed for both series of experiments, and also earlier estimates using a simplified model [5]. In addition, Table 1 shows estimates of the rms deviation of the test points from the theoretical dependence (σ_c) and the total error in determining the unknown model parameters (σ_x):

$$\sigma_c = \sqrt{F/(M-1)},\ \sigma_x = \sqrt{F/(M-N_x)}. \tag{16}$$

To picture the values obtained we use an estimate of the accuracy of determining the hot particle concentration in the fluidized bed. A total of 4-5 samples were taken from the ITMO Acad. Sci. BSSR facility under steady combustion conditions and the fuel content was analyzed. The ratio of the rms deviation to the mean was 0.3. Therefore, the accuracy of measuring the fuel concentration in the fluidized bed is close to the scatter of points relative to the theorical relation, $\sigma_c = 0.51$. This near agreement of these estimates justifies our assertion that the model developed adequately describes the process of solid fuel combustion in a fluidized bed.

The model parameter values presented were based on two theoretical models derived from the same body of experimental data: one estimate was made with the simplified model [5], and the other with the full model. The values obtained practically coincide, which evidently indicates that the technique is valid. A noticeable difference is observed only in the mean relative diameter of the hot particle. Probably the estimate $\phi = 6$ is close to the actual value, since under combustion in a fluidized bed the particles experience cracking, pulverizing and abrasion.

One should not pay attention to the fact that the estimates of model parameters agree, although they were obtained from very different experiments: in a laboratory facility and in the furnace of an industrial boiler complex. Probably the fuel oxidation reaction occurring at high intensity is accomplished within a small height range of the bed near the gas distribution grid, and therefore the progress of the process depends only a little on the horizontal scale of the facility.

We now turn to the remaining model parameters. The activation energy of fuel combustion agrees roughly with the literature data for Donets deposit coal [6], which was used in the experiments. The mixing coefficient k (Eq. (10)) was determined quite accurately. It does not differ decisively from the earlier value of 0.1 from tests of solid phase mixing [13]. The values of P and n differ noticeably from the predicted values. As was noted above, the coal combustion reaction occurs in the near-grid zone where the bed hydrodynamics is developing. Air emerging from the grid apertures penetrates into the bed of disperse material and suspends the latter. Gas bubbles (the discrete phase) are generated, and an emulsion of mixed particles (continuum phase). In this zone the flow of the fluid differs from the predictions of two-phase fluidization theory. Firstly, in the continuum phase the gas is filtered with a velocity exceeding by a factor of two the value for the theoretical onset of fluidization, and therefore the coefficient n is not zero. Secondly, the exchange between the discrete and the continuum phases is intensified and the coefficient P is noticeably greater than 1.

Thus, the model developed for fuel combustion in a fluidized bed adequately describes the system and can be used to seek optimal conditions and design parameters of furnaces of power equipment.

NOTATION

C, C_p, C_e, fuel concentration in the fluidized bed: ambient, mean and experimental; C_0, emission coefficient of a perfect black body; D, D_T, D_g, coefficient of molecular diffusion, and of effective diffusion for the material and the gas in the fluidized bed; d, d_i, diameter of fuel particles and inert bed material; F, target function; H, bed height; K*, T*, constants in the combustion kinetics equation (K* is rate, T* is characteristic temperature); k, mixing coefficient; M, number of test points; n, coefficient giving the gas filtration velocity in the continuum phase; N, fluidization number; N_x, number of unknown model parameters; Q_p^H, lower heat of combustion of the working mass of the fuel; P, coefficient describing gas exchange between the continuum and the discrete phases; R, universal gas constant; T_p, T_0, T_b, temperatures of the hot particle, of the fuel supplied to the bed, and of the bed, respectively; U, U_0, velocity of filtration and velocity at the onset of fluidization; x, vertical coordinate; Y_0, Y, Y_π, concentrations of oxygen in air, of the continuum and of the discrete phases; W, rate of supply of fuel; α, heat transfer coefficient for the hot particle with the bed; β, β_π, mass transfer coefficients for the particle surface and between the continuum and discrete phases; γ, relative filtration velocity in the continuum phase; ε_0, ε, porosity of the charge and the fluidized bed; ε_{re}, reduced emissivity of the system of hot particle plus fluidized bed; θ, relative overheat temperature of the hot particle; λ, thermal conductivity of the gas; ν, viscosity of the gas; ξ, relative combustion rate; ρ_p, ρ, ρ_i, $\rho*$, density of the particles of coal, gas, inert matter and gaseous carbon; σ_c, σ_x, rms errors of determining the fuel concentration and the total error of determining the model parameters; $1/\phi$, mean relative diameter of a hot particle. Dimensionless numbers: $Nu = \alpha d_i/\lambda$; $Sh = \beta d_i/D$; $Fr = (U-U_0)^2/gH$; $Sc = \nu/D$; $Ar = \frac{gd_i^3}{\nu^2}\frac{\rho_i - \rho}{\rho}$.

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CHOICE AND OPTIMIZATION OF THE PARAMETERS OF A POROUS-SUBLIMATION COOLER

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The parameters of a porous-sublimation cooler (PSC) such that the temperature of the cooling surface a) does not exceed a prescribed value, b) is minimum, and c) does not exceed a prescribed value with minimum mass of the PSC are determined.

In the process of porous-sublimation cooling [1-3] heat is removed from the surface being cooled (Fig. 1) along a porous framework and is transferred to a solid refrigerant residing in the pores of the framework; this process results in sublimation of the refrigerant. Compared with other methods of sublimation cooling [4] a PSC substantially simplifies the construction of the cryosublimation systems, increases their reliability, and significantly improves heat transfer between the object being cooled and the refrigerant [2, 5]. When cryogenic systems are cooled it is usually necessary to choose the parameters of the PSC so that the temperature of the surface being cooled is minimum or does not exceed a prescribed value. For space cryogenic technology it is especially important to provide these conditions with minimum mass of the PSC.

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