MATHEMATICAL MODEL OF COAL COMBUSTION IN A FURNACE WITH A FLUIDIZED BED

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The combustion of solid fuel in a fluidized bed is a complex process which can be regarded as a physicochemical system (PCS) with an inherent structural hierarchy [1]. The study of a PCS from the viewpoint of systems analysis [2] consists of independent modeling of the processes at each level of the hierarchy and the establishment of links between these processes. It is possible to distinguish five characteristic levels which can be described by special mathematical systems.

The processes of the first level are the phenomena of heterogeneous and homogeneous kinetics, volatilization, etc. occurring at the ionic and molecular level. At the second level, we examine the growth and movement of bubbles, mass and heat transfer between phases in the layer, and heat exchange with the heating surface. The third level of the hierarchy includes processes initiated by fields of both internal and external forces (the hydrodynamics of the bed, circulation of the solid and gas phases, particle distribution and entrainment, etc.). The fourth level of the hierarchy used in structural analysis of the system takes into account processes occurring at the flow-chart level and ensuring a connection between the FFB (furnace with fluidized bed) and other system elements.

The fifth level is the effect of the given production scheme on the environment.

The boundaries of the above-described levels are conditional. The levels affect one another.

Investigators have developed a large number of mathematical models which with varying degrees of accuracy describe actual processes taking place in fluidized-bed furnaces. The most interesting mathematical models in the domestic literature are those surveyed in [3]. However, these models do not offer a detailed description of coal combustion processes in an FFB or combustion kinetics in particular, which would make it possible to evaluate the temperature of the bed, its fractional composition during the combustion of polyfractional fuels, or any other specified distribution.

The goal of the present study is to describe the first three levels of physicochemical processes taking place in an FFB, establish links between these levels, and describe a mathematical model of coal combustion in an FFB.

The proposed mathematical model of the FFB consists of the system of energy and balance equations of six gaseous components: O_2, CO_2, H_2O, H_2, CO, N_2 for each of the four phases of the fluidized bed: 1) the bubble phase, representing the gas which bypasses the bed; 2) the dense (emulsion) phase, consisting of a mixture of gas and particles of the SO_2 sorbent; 3) the bubble-train phase, including gas and entrained solid particles; 4) the phase of burning coal particles.

Being the basis for calculation of chemical underfiring and bed temperature, this system of equations — together with the equation of the size distribution of burning coal particles in the bed — makes it possible to calculate the percentage of carbon in the bed, coal combustion efficiency, and mechanical underfiring.

Structutally, the model is an optimum elaboration of existing models from the viewpoint of levels of the hierarchy of physicochemical processes (Fig. 1).

An important difference between the present model and previous models is the expanded description of combustion kinetics, allowance for several important factors in the balance equations, and the possibility of calculating the combustion of polyfractional fuels of any specified distribution. The physical motivation for allowing for given quantities in the balance equations is given as a result of analysis of the PCS at three levels of the hierarchy.

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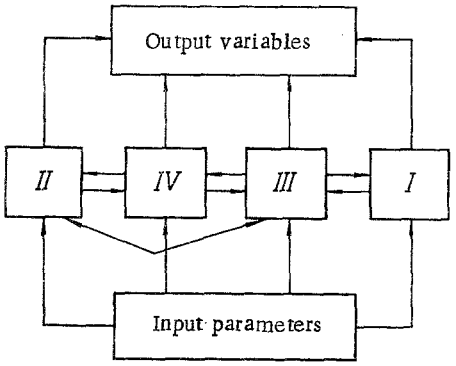

Fig. 1. Systemic-structural representation of a model of coal combustion in a furance with a fluidized bed: I) bubbles, ideal displacement, mass balance of O_2, CO_2, H_2O, CO, H_2, and N_2, energy balance: II) total mixing, mass balance of O_2, CO_2, H_2O, H_2, CO, and N_2, energy balance; III) train, ideal displacement, mass balance of O_2, CO_2, H_2O, H_2, CO, and N_2, energy balance; IV) burning coal particles; mass balance of O_2, CO_2, H_2O, H_2, and CO for each particle, energy balance, particle distribution in the bed.

Let us briefly examine the processes being modeled in the FFB.

STUDY OF PHYSICAL PROCESSES AT THE MICROSCOPIC LEVEL. KINETICS

The combustion of a coal particle occurs in two stages: 1) the escape of volatile matter and its combustion; 2) oxidation of the coal residue in the form of coke, resulting in the formation of CO and CO_2. It is assumed on the basis of theoretical and experimental studies [4, 5] that the volatile matter escapes instantaneously and the coke burns in the flame of the volatiles and air oxygen.

It is believe that coke particles undergo combustion by the Nusselt mechanism [6], whereby oxygen diffusing through the boundary layer of the particle reacts with the coke, simultaneously forming CO and CO_2. The combustion regime changes in relation to the temperature and size of the particle and the concentration of gaseous reactants. Here, both the rates of chemical reaction and diffusion through the boundary layer are considered. The change in the coke particle was examined in accordance with the model of segregated ash (AS) [7]. Depending on the grade of coal and the ash fraction in the particle, the AS representation can be replaced by the "progressive shell" model (SP) [7]. Let us write the reactions being considered in coke combustion:

$$C_{(S)} + \frac{1}{2} O_{2(G)} \xrightarrow{k_1} CO_{(G)}, \tag{1}$$

$$C_{(S)} + O_{2(G)} \xrightarrow{k_2} CO_{2(G)}, \tag{2}$$

$$C_{(S)} + CO_{2(G)} \xrightarrow{k_3} 2CO_{(G)}, \tag{3}$$

$$C_{(S)} + H_2O_{(G)} \xrightarrow{k_4} CO_{(G)} + H_{2(G)}, \tag{4}$$

$$CO_{(G)} + \frac{1}{2} O_{2(G)} \xrightarrow{k_5} CO_{2(G)}, \tag{5}$$

$$CO_{(G)} + H_2O_{(G)} \underset{k_6}{\overset{k_6}{\rightleftarrows}} CO_{2(G)} + H_{2(G)}. \tag{6}$$

The first four reactions are heterogeneous and occur on the particle surface. The homogeneous reactions for examine in the gas phase.

As already noted, processes at the third level lead to a change in concentrations and temperatures in the system.

The concentrations of the gaseous components O_2, CO_2, H_2O, H_2, CO on the particle surface and in the core of the flow are linked as follows:

$$K_{IS}(C_{je} - C_{js}) + \sum_{i=1}^{4} \alpha_{ij} R_i = 0, \quad i = 1 - 4; \quad j = 1 - 5. \tag{7}$$

Equation (7) is solved for all particles having a radius from r_{min} to r_{max} and located in the bed. The system of energy equations describing the heating of coke particles of the radii from r_{min} to r_{max}:

$$4\pi r^2 \sum_{i=1}^{4} R_i(-\Delta H_i) + 4\pi r^2 (T_e - T_s) H_{IS} + 4\pi r^2 \varepsilon_s \sigma (T_e^4 - T_s^4) = 0. \quad (8)$$

The Sherwood number, entering into the mass-transfer coefficient k_{IS}, changes with radius for particles in the emulsion phase in the minimal-fluidization state. The number of changes in accordance with Eq. (VII.1) in [8]. For particles located in the train, it is equal to two. The second term of Eq. (8) represents convective heat transfer between the gases and the burning particle:

$$H_{IS} = \mathrm{Nu}_S \frac{\lambda_g}{2r}. \quad (9)$$

The Nusselt number for particles of the emulsion phase is determined from Eq. (VII.4.1) in [8]. For particles of the train, $\mathrm{Nu}_S = 2.0$. Proceeding on the basis of the coke-combustion kinetics adopted here, the carbon mass balance in the particle is written as follows:

$$\frac{\partial}{\partial t}\left(\frac{4}{3}\pi r^3 \rho_S \frac{W_f}{W_f + W_a}\right) = -M_S 4\pi r^2 \sum_{i=1} \alpha_i^c R_i. \quad (10)$$

Equations (7), (8), and (10) are solved separately for the train and emulsion phases.

PHYSICOCHEMICAL PROCESSES AT THE MACROSCOPIC LEVEL. BUBBLE GROWTH AND MOVEMENT IN THE BED

When they bypass the bed, the bubbles grow as a result of a pressure change and coalescence. Relations for the change in bubble diameter in an FB were proposed in [9-11]. In our model, we used the Mori–Wen relation [10] for FB's in pilot units.

Expressions for the relative rate of rise of the bubbles, with allowance for the effect of the reactor walls, were borrowed from [12]:

$$u_b = u_0 - (1 - \delta - \alpha\delta)\varepsilon_{mf} u_e + 0.711 (gD_b)^{1/2} \quad \text{at} \quad D_b/D_t < 0.5, \quad (11)$$

$$u_b = u_0 - (1 - \delta - \alpha\delta)\varepsilon_{mf} u_e + 0.35 (gD_t)^{1/2} \quad \text{at} \quad D_b/D_t > 0.5.$$

The coefficients of mass and heat transfer between the phases of the FB were determined with allowance for the bubble–(cloud + train)–emulsion transfer mechanism in accordance with the model of Kunii and Levenshpil' [8]:

$$K_{bc} = \frac{4.5 u_{mf}}{D_b} + \frac{5.85 D^{1/2} g^{1/4}}{D_b^{5/4}}, \quad (12)$$

$$K_{ce} \cong 6.78 (\varepsilon_{mf} D u_b / D_b^3)^{1/2}, \quad (13)$$

$$H_{bc} = 4.5 u_{mf} \rho_g C_{pg} / D_b + 5.85 (\lambda_g \rho_g C_{pg})^{1/2} g^{1/4} / D_b^{5/4}, \quad (14)$$

$$H_{ce} \cong 6.78 (\rho_g C_{pg} \lambda_g)^{1/2} (\varepsilon_{mf} u_b / D_b^3)^{1/2}. \quad (15)$$

PHYSICOCHEMICAL PROCESSES AT THE THIRD LEVEL OF THE HIERARCHY. CIRCULATION OF THE SOLID PHASE IN THE BED

It is assumed that the solid phase (particles of filler and coal) are transported by the bubble train [13]. As was noted in [13], circulation of the solid and gas phases is possible under the bed at critical fuel-feed velocities. In the present model, this possibility is considered in the initial conditions of the balance equations and in the expression for the velocity of the solid phase in the emulsion

$$u_e + u_s = \frac{u_{mf}}{\varepsilon_{mf}}. \quad (16)$$

In accordance with the two-phase theory of fluidization [11], the fraction of the gas flow which enters the emulsion phase is

$$n = \{[1 - \delta(1+\alpha)]\, u_e \varepsilon_{mf}\}/u_0. \tag{17}$$

BALANCE EQUATIONS FOR THE PHASES OF AN FFB

In comformity with the chosen gas distribution among the phases, the flow through the bubble phase

$$q_b(z) = \rho_g(z) A_t u_b(z) \delta(z). \tag{18}$$

An ideal displacement regime in the bubble phase yields the equation for the molar balance of the concentration of the j-th component and temperature

$$\frac{d}{dz}[q_b(z) y_{jb}] = A_t \delta_b(z) K_{bcj}(z)(C_{jc} - C_{jb}) + \delta_b(z) A_t(\alpha_{5j} R_5^b + \alpha_{6j} R_6^b), \tag{19}$$

$$\frac{d}{dz}[q_b(z) C_{pg}(z) T_b] = A_t \delta(z) H_{bc}(z)(T_c - T_b) + A_t \delta(z)[(-\Delta H_5) R_5^b + (-\Delta H_6) R_6^b]. \tag{20}$$

The gas flow thorugh the train phase

$$q_c(z) = \rho_g(z) A_t u_b(z) \delta(z) \alpha \varepsilon_{mf}. \tag{21}$$

Having assumed that the ideal displacement regime exists in the train, with allowance for the heterogeneous reactions we obtain the following expressions for the balances of concentration and temperature:

$$\frac{d}{dz}[q_c(z) y_{jc}] = A_t \delta(z) K_{bcj}(z)(C_{jb} - C_{jc}) + A_t \delta(z) K_{cej}(z)(C_{ej} - C_{cj}) +$$

$$+ A_t \alpha \varepsilon_{mf} \delta(z) \sum_{i=5}^{6} \alpha_{ij} R_i^c + \frac{1}{H_b} \frac{\alpha\delta}{1-\delta} \int_0^\infty \frac{f_c W P_b(r)}{4/3\pi r^3 \rho_s} R_{jc}(r,\ C_{jm})\, dr, \tag{22}$$

$$\frac{d}{dz}[q_c(z) c_{pg}(z) T_c] = A_t \delta(z) H_{bc}(z)\ (T_b - T_c) + A_t \delta(z)(T_e - T_c) H_{ce} +$$

$$+ \alpha \varepsilon_{mf} \delta(z) \sum_{i=5}^{6} (-\Delta H_i) R_i^c + \frac{\alpha\delta}{1-\delta} \frac{1}{H_b} \left[\int_0^\infty \sigma_s W f_c H_{Is}(T_s - T_c) P_b(r)\, dr + \int_0^\infty \sigma_s W f_c \sigma_b (T_s^4 - T_c^4) P_b(r)\, dr \right]. \tag{23}$$

The initial conditions for Eqs. (19)-(23) will be written for two different cases of motion of the gas flows in the emulsion phase.

The emulsion phase consists of bed filler, coke particles, and gas. In accordance with the two-phase theory of fluidization, the fraction cavities in the emulsion is equal to ε_{mf}. The gas flow in the emulsion phase

$$g_e = \bar{\rho}_g (1 - \delta(1+\alpha)) \varepsilon_{mf} u_e A_t. \tag{24}$$

With the assumption that complete mixing occurs, we write the following system of heat and mass conversion equations:

$$0 = q_e(y_{ej}^f - y_{ej}) + \underset{I}{\int_0^{H_b}} \underset{II}{A_t \delta(z) K_{ce}(z)(C_{jc}(z) - C_{je})\, dz} + \left(\frac{1-\delta-\alpha\delta}{1-\delta}\right) \times$$

$$\times \underset{III}{\int_0^\infty K_{Is}(C_{js}(r) - C_{je}) \frac{3}{r\rho_s} W f_c P_b(r)\, dr} + \underset{IV}{V_e \sum_{i=5}^{6} \alpha_{ij} R_i^e} + \underset{V}{\frac{F_0 W_j}{M_j}} - \underset{VI}{F_0 W_v \xi_{jv}}; \tag{25}$$

$$0 = \underset{VII}{q_e C_{pg}(T^f - T^e)} + \underset{VIII}{\int_0^{H_b} A_t \delta(z) H_{ce}(z)(T_c - T_e)\, dz} + \int_0^\infty H_{Is}(T_s(r) - T_e) \times$$

$$\times \underset{IX}{\frac{3}{r\rho_s} W \cdot f_c P_b(r)\, dr} + \underset{X}{V_e \sum_{i=5}^{6} (-\Delta H_i) R_i^e} + \underset{XI}{F_s W_v (-\Delta H_v)} +$$

$$+ \underset{XII}{\varepsilon_s \sigma \int_0^\infty \frac{3}{r\rho_s} W \cdot f_c P_b(r)(T_s^4(r) - T_c^4)\, dr} + \underset{XIII}{H_{ta} A_t (T_{st} - T_e)} - \underset{XIV}{\int_0^\infty F_s C_{ps}(T_s(r) - T_{s0}) P_0(r)\, dr}. \tag{26}$$

The Roman numerals in Eqs. (25), (26) denote terms which describe the following processes: I – the change in the molar fractions of substance j between entry into the emulsion phase and exit from the bed; II – overall mass transfer between the emulsion and the plane (train-shell); III – mass transfer between the gas in the flow core and on the surface of a burning particle; IV – change in component j due to homogeneous reactions; V – the amount of the j-th component introduced with the fuel; VI – the amount of the j-th component altered with the escape of volatilities; VII – the analog of (I) for temperatures; VIII – the analog of (II) for temperature; IX) the analog of (III) for temperature; X - heat release as a result of homogeneous reactions; XI – heat release as a result of combustion of volatiles; XII – radiative heat transfer between the bed filler and coke; XIII – heat exchange with the heating surface; XIV – heat losses due to the heating of newly arrived coke particles.

In the case of gas circulation, $u_e < 0$:

$$y_{je}^{f} = \frac{y_{jb}^{H} + \alpha\varepsilon_{mf} y_{jc}^{H}}{1 + \alpha\varepsilon_{mf}}, \tag{27}$$

$$T^{f} = \frac{T_b^H + \alpha\varepsilon_{mf} T_c^H}{1 + \alpha\varepsilon_{mf}}. \tag{28}$$

The initial conditions for (19)-(23):

$$z = 0:\ C_{jb}^{0} = C_{jc}^{0} = \frac{1}{(u_0 - (1 - \delta(1+\alpha)\dot{\varepsilon}_{mf} u_e))}\left[u_0 C_j^0 - (1 - \delta - \alpha\delta)\varepsilon_{mf} u_e C_{je}\frac{T_e}{T_0}\right], \tag{29}$$

$$T_b^0 = T_c^0 = \frac{1}{u_0 - (1 - \delta - \alpha\delta)\varepsilon_{mf} u_e}[u_0 T^0 - (1 - \delta - \alpha\delta)\varepsilon_{mf} u_e T_e]. \tag{30}$$

In the absence of gas circulation, $u_e > 0$:

$$y_j^f = y_{je}^0, \tag{31}$$

$$T^f = T^0. \tag{32}$$

The initial conditions for Eqs. (19)-(23):

$$z = 0:\ C_{jb}^0 = C_j^0,\ C_{jc}^0 = C_j^0,\ T_b^0 = T_c^0 = T_0. \tag{33}$$

PARTICLE DISTRIBUTION IN THE BED ACCORDING TO SIZE

The number of particles of radius r which enter the reactor is equal to the number of particles removed from the reactor minus the number of burning particles. In mathematical variables:

$$\frac{F_0 P_0(r)}{(4/3)\pi r^3 \rho_f} - \frac{F_1 P_1(r)}{(4/3)\pi r^3 \rho_s} - \frac{f_c W P_b(r) K(r)(1 - \eta(r))}{(4/3)\pi r^3 \rho_s} - \frac{d}{dr}\left[R(r)\frac{f_c W P_b(r)}{(4/3)\pi r^3 \rho_s}\right] = 0, \tag{34}$$

$$\rho_s = \rho_f (1 - W_v). \tag{35}$$

The resulting equation for the coke-particle probability density function in the bed $P_b(r)$ takes the form

$$\frac{d}{dr}\chi(r) + \left\{\frac{F_0/f_c + K(r)(1 - \eta(r)) + \frac{d}{dr}R(r)}{R(r)} - \frac{3}{r}\right\}\chi(r) = \frac{F_0(1 - W_v)P_0(r)}{R(r)W}, \tag{36}$$

where $\chi(r) = f_c P_b(r)$.

The initial condition for (36)

$$\lim_{r\to\infty}\chi(r) = 0. \tag{37}$$

The entrainment constant K(r) is calculated from the correlation obtained in [14] for pilot FFB's. The entrainment from the furnace is determined by the expression

$$F_2 = Wf_c \int_0^\infty K(r) P_b(r)\, dr. \tag{38}$$

With allowance for reflux and entrainment, the efficiency of coal combustion in the furnace is determined as

$$\eta_e = \frac{F_0 - F_1 - F_2}{F_0}. \tag{39}$$

MATHEMATICAL MODEL OF AN AFFB

Thus, the mathematical model includes a system of nonlinear differential equations which describe the material and energy balances in each of the four phases of the bed. We reduced the system to dimensionless form and we developed a program pack and corresponding software.

The adequacy of the model was checked against experimental data supplied by tye Altai Polytechnic Institute (API) (monofractions) and the Kazakh Scientific-Research Institute of Power Engineering (KazNIIÉ) (polyfractions). The material in each case was Ékibastuzsk coal. Figure 2 shows experimental and calculated values of mechanical underfiring and bed temperature. The calculated correlation coefficients, equal to 0.84 (KazNIIÉ data) and 0.94 (API data) with confidence intervals of 0.33-0.97 and 0.74-0.98, indicate the presence of a physical connection between the experimental and theoretical results. The adequacy of the model is further supported by results obtained for Donetsk anthracite. Hwere, we performed a fatorial experiment using a two-way table and involving second-order interactions, We calculated the Fisher criterion for experimental and theoretical data on combustion efficiency, temperature, and gas composition. The resulting values (4.0, 1.4, and 2.5) were less than the tabulated value of 4.7, with P = 0.05 and degrees of freedom $f_1 = 14$ and $f_2 = 5$. This is further evidence of the adequacy of the model is regard to the experiment.

Numerical calculations were also performed with the data in [15]. The input parameters of the model are shown below:

composition of the coal: $W^P = 20\%$; $A^C = 17.2\%$, $C^P = 53.2\%$, $H^P = 3.7\%$; $O^P = 17.6\%$; $S^P = 0.2\%$; $V^g = 36\%$;

cross section of the bed $A_t = 3.3\ m^2$,·

weighted-mean size of filler particles $d_p = 1.5$ mm;

fractional composition of the coal: 98% particles 0-2.29 mm in size (Rosen–Rammler distribution);

weight of bed filler 3700 kg;

area of heating surface in the bed 15 m^2;

coefficient of heat transfer with the heat surface $A_{st} = 170\ W/(m^2 \cdot °K)$;

temperature of vapor $T_{st} = 570°K$;

velocity of air at bed inlet $u_0 = 1.5$ m/sec;

rate of fuel feed $F_0 = 0.15$ kg/sec.

With a velocity of 1.5 m/sec, a fuel fractional composition with a characterisic Rosen–Rommler distribution, and a bed weight of 3700 kg, calculated output parameters such as fuel combustion efficiency (97%), carbon fraction in the bed (0.5%), bed temperature (about 1100°K), and gas concentrations at the bed outlet agree with the experimental data in [15] to within 1.5-2%.

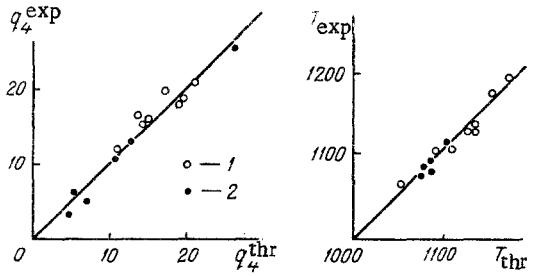

Fig. 2. Correlation between results of calculations with the mathematical model and experimental data: 1) data from KazNIIÉ (Ékibastuzsk coal, polyfraction); 2) data from APO (Ékibastuzsk coal, monofraction); $\beta = 0.95$; $r' = 0.84$ ($0.33 < r' < 0.97$); $r^* = 0.94$ ($0.74 < r^* < 0.98$). T, °K; q, %.

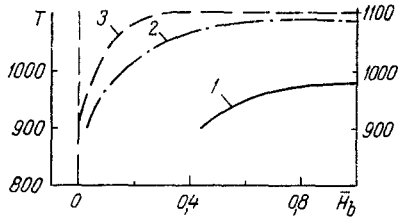

Fig. 3

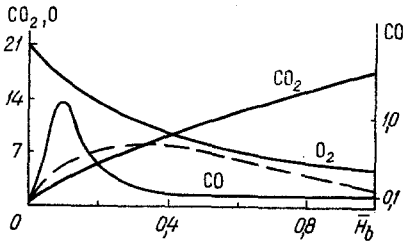

Fig. 4

Fig. 3. Temperature profiles of bubble and cloud-train in the FFB: 1) bubble; 2) bed; 3) cloud-train.

Fig. 4. Change in the CO_2, O_2, and CO concentrations over the bed height in the bubble phase (%); dashed curve — CO in the bed.

Figure 3 shows the change in the temperature of the phases over the height of the bed. Due to the relatively high velocity (2.5-3 m/sec), there is not sufficient time for the bubble to be heated to the temperature of the emulsion. The calculation shows a rapid increase in the concentration of CO in the bubble and the cloud near the distributing grate (up to 1.3% in the cloud). There is then a sharp drop in CO concentration due to the predominance of the CO combustion rate over mass transfer with the emulsion phase (Fig. 4). Such a character of change was also seen in the experimental curves in [16]. The model calculates the temperature of a burning coal particle as a function of its radius. With an increase in air and fuel consumption, combustion for a large spectrum of particle radii moves to the diffusion regime.

Thus, the proposed mathematical model can be used to calculate such important AFFB characteristics as mechanical and chemical underfiring and the temperature and fractional composition of the bed. An original algorithm was developed which makes it possible to reduce the time required to solve the complex system of equations to 1.5 min of processor time per iteration, which in turn makes it possible to eventually use the model to solve optimization problems.

NOTATION

A_t, cross section of furnace, m^3; C_{je}, C_{js}, C_{jb}, C_{jc}, concentrations of the j-th gaseous component in the emulsion, on the particle surface, in the bubble phase, and in the cloud-train, kmole/m^3, C_{pg}, heat capacity of the mixture; D_j, diffusion coefficient of the j-th component of the mixture, m^2/sec; D_b, apparent diameter of bubble, m; F_o, coal feed to the bed, kg/sec; F_1, coal reflux, kg/sec; F_2, entrainment, kg/sec; f_c, fraction of coal in the bed; g, acceleration due to gravity, m/sec^2; H_b, height of expanded bed, m; ΔH_v, heat of combustion of volatiles, J/kg; ΔH_i, heat of reactions, j/kmole; H_{IS}, gas–particle heat-transfer coefficient, W/kmole; H_{ta}, bed–heating-surface heat-transfer coefficient, W/m^2·°K); H_{bc}, bubble–cloud-train heat transfer coefficient, W/(m^3·°K); H_{ce}, train–emulsion heat-transfer coefficient, W/(m^3·°K); K(r), entrainment constant, sec^{-1}; K_{IS}, gas–particle mass-transfer coefficient, m/sec; K_{ce}, cloud–emulsion heat-transfer coefficient, sec^{-1}; Nu_S, Nusselt number; $P_o(r)$, $P_b(r)$, $P_2(r)$, particle-size density function in the feed, bed, and product, m^{-1}; R_i, rates of reactions, kmole/(m^2·sec); r, particle radius, m; R_{ji}, number of moles of reactant j, changing per unit of time due to heterogeneous reactions i over the particle radius r, kmole/sec; q_b, q_c, q_e, gas flows through the bubble phase, train, and emulsion, kg/sec; T_e, T_b, T_c, T_s, temperatures of the emulsion, bubbles, train, and burning coal particle, °K; t, time; u , u_{mf}, u_b, u_e, u_s, velocity of air delivery under the bed, minimum fluidization velocity, bubble velocity, gas velocity in the emulsion phase, and particle velocity in the emulsion phase, m/sec; W^p, moisture content of coal; W_f, fraction of solid coal in a coal particle; W_a, fraction of ash in a coke particle; W_v, fraction of volatile matter in coal; W, weight of bed, kg; y_{jb}, y_{jc}, y_{je}, molar fractions of the j-th component in the bubbles, train, and emulsion; α_i, α_{ij}, stoichiometric coefficients of carbon and the j-th component in the i-th reaction; σ, Stefan–Boltzmann constant, W/(m^2·°K^4); ε_s, radiative heat transfer attenuation factor; ρ_g, ρ_s, ρ_f, densities of gas and coal, kg/m^3; ε_{mf}, porosity of bed in the fluidized state; λ_g, thermal coductivity of gas, W/(m·°K); δ, fraction of bubbles in the bed; α, fraction of train-cloud in volume of bed; ξ_v, stoichiometric coefficient of oxygen in the volatiles combustion reaction, kmole/kg; η_e, coal combustion efficiency; η(r), efficiency of capture of entrained material; K_{bc}, bubble–cloud mass-transfer coefficient, sec^{-1}.

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INFLUENCE OF POROSITY ON THE DYNAMIC PROPERTIES OF A LOOSE GRANULAR LAYER

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Processes accompanying the sudden retardation of a layer with its different porosity are analyzed by using a perfected dynamic model of a loose granular medium. The results are compared with systematic experiments.

In technological processes associated with the reworking of granular media, it is usual to deal with different loose layers whose density change occurs principally because of structural deformations. In intensive methods of reworking, when the layers are subjected to some kind of mobile state, the operating density of the granular mass is less than or equal to the critical.* Under the eposidic nature of the displacements, the density of the fill increases and can approach the maximal.† Theoretical investigations of the dynamics of such systems are complex [7]. Hence in laboratories it is often obligatory, in substance, to conduct full-scale tests on equipment simulating industrial specimens. In a number of cases they can be replaeed by models. In particular, certain dynamic properties of fluidized and moving layers are simulated in an impact-disturbed stationary loose fill with an equivalent structure [3].

In order to determine the dynamic properties of a granular layer it is necessary to know its force characteristic. Idealizing the phenomenon, we represent the compression curve of a

*The critical τ_{cr} is understood to be the bulk density at which a zero-th dilatancy effect is assured [1, p. 194; 2, p. 84]. The so-called closest value to it is observed in so-called "compact" moving layers and fluidization at rest [3, 4, p. 489].

†The maximal bulk density τ_{max} delimits loose and compact granular layers; a change in the compactness of the latter occurs not under structural but elastic-plastic deformations of grain pressure [5, p. 19; 6, p. 26].

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