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MECHANISM OF COAL COMBUSTION IN A FLUIDIZED LAYER OF COARSELY DISPERSED MATERIAL

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A system of one-dimensional steady-state balance equations is formulated for fuel, oxidizer, and energy. An approximate solution is obtained and compared to experiments on coal combustion in a fluidized layer.

In connection with the wide use of coal and other low-grade solid fuels, much promise has been shown by the process of combustion in fluidized bed furnaces, in which relatively coarse coal particles burn in a suspended state within a relatively low height fluidized layer of nonburning material. The system is inhomogeneous and consists of gas bubbles and a continuous phase (the solid particles suspended in the gas), which makes the character of coal combustion in a fluidized bed unique.

In recent years a number of two-phase models of coal combustion have been proposed [1, 2], based on information on gas bubble motion in a fluidized bed [3] and the heterogeneous reaction rate for combustion of residual coke [4].

The concepts of the two-phase model, which are valid for fluidized layers of fine particles of relatively low height, such as those used in heterogeneous catalytic reactors, are not completely applicable to shallow layers of relatively coarse particles used in low power furnaces. In the latter case the rate of bubble motion is less than the filtration velocity, and thus gas traversal through the layer with no contact with the solid material, so-called "bypass," is impossible. Moreover, in this layers including a zone near the grate, gas bubbles are generated, interphase exchange is quite intense, and traversal of gas streams through the layer without interaction with the dispersed particles is improbable.

Below we will propose a model for the process of combustion of a solid fuel in a fluidized bed of noncombustible material in the form of two interpenetrating continua: a gas

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and a solid phase.

We will write one-dimensional balance equations for the fuel, oxidizer, and energy for the steady-state process of fluidized bed combustion. We assume that the temperature, maintained by heat removal surfaces immersed in the layer, remains constant over layer height.

We introduce dimensionless coordinates and variables

$$\theta = \frac{T_b - T_b}{T_b - T_0}$$

and write the following equations:

fuel balance

$$C'' - \frac{m}{k} B\varphi\zeta y C = 0, \quad (1)$$

oxygen balance

$$\varepsilon k y'' - \frac{N}{N-1} y' - B\varphi\zeta y C = 0, \quad (2)$$

combusting fuel energy balance

$$(C\theta)'' + q \frac{m}{k} B\varphi\zeta y C - \frac{\Phi}{k} B\varphi C\theta = 0. \quad (3)$$

The fuel together with other particles participates in the chaotic circulation motion of the dispersed material. Under steady-state conditions the diffusion transport of fuel caused by this motion can be characterized by an effective diffusion coefficient. The latter can be estimated with the empirical correlation [5]

$$D_s = k(U - U_0)H. \quad (4)$$

The first term in Eq. (1) describes the diffusion flow of fuel, while the second reflects the rate of fuel "loss," i.e., the combustion rate. Since volatiles are rapidly removed from the fuel, it is basically a coke residue (carbon) which burns in the bed. Its combustion rate is directly proportional to the oxygen concentration and the burning particle surface available, i.e., the fuel concentration in the layer.

The rate of coke particle oxidation can be described by the expression [4]

$$K = K_* \exp \frac{E}{R} \left(\frac{1}{T_p} - \frac{1}{T_*} \right), \quad (5)$$

while the net combustion rate with consideration of resistance to oxygen transport to the burning particle surface will be equal to [6]

$$\frac{1}{\zeta} = \frac{K_* d}{Sh D} + \exp \frac{E}{RT_b} \left[\frac{1}{\theta(1 - T_0/T_b) + 1} - \frac{T_b}{T_*} \right]. \quad (6)$$

The intensity of oxygen transport to the combustion surface is determined by the correlation of [7], which describes mass exchange on a particle surface in a fluidized bed:

$$Sh = 2.78 Re_0^{1/3} Sc^{1/3} \left(\frac{d_i}{d} \right)^{0.13} \left(\frac{\rho_p}{\rho_i} \right)^{0.15}. \quad (7)$$

We will now consider the oxygen balance equation. The gas in the fluidized bed participates in two motions. On the one hand, the gas enters into the chaotic circulation mixing of the dispersed material, while on the other, it shifts upward relative to the particles at the filtration velocity. The first and second terms of Eq. (2) consider both these factors. Since, as the experiments of [8, 9] have shown, fluctuations in the velocities of gas and solid phase are similar to each other, and their spectra are analogous, the chaotic motion of the gas is similar to the stochastic displacement of the solid material.

Therefore, we can use Eq. (4), obtained for the solid material, to describe the diffusion transport of the gas. The third term in Eq. (2) characterizes loss of oxidizer, expended in fuel combustion.

We will now turn to the energy balance equation (3). The burning fuel particles have a higher temperature than the surrounding bed material. The heat liberated during combustion (second term) is transferred to the surrounding particles, as reflected by the third term. The first term of Eq. (3) describes diffusion transport of hot particles. Thus, the unique features of fuel combustion in a fluidized bed depend on the following dimensionless complexes:

$$B = \frac{6K_*H}{d(U - U_0)} \quad (8)$$

the relative combustion rate;

$$m = \frac{\rho_*}{\rho_p(1 - \varepsilon)} \quad (9)$$

density ratio of gaseous and solid carbon;

$$\xi = \frac{\bar{K}}{K_*} \quad (10)$$

relative net combustion rate;

$$\Phi = \frac{Nu\lambda}{K_*\rho_p c_p d(1 - \varepsilon)} \quad (11)$$

relative heat liberation of burning particles to the fluidized bed;

$$q = \frac{Q_p^h}{c_p(T_p - T_0)} \quad (12)$$

relative heat of combustion of the coal.

We will now formulate the boundary conditions for the problem:

$$x = 0 \quad C' = 0, \quad (13)$$

$$x = 0 \quad (C\theta)' = 0, \quad (14)$$

$$x = 0 \quad y' = \frac{N}{k(N-1)}(y - y_0), \quad (15)$$

$$x = 1 \quad (C\theta)' = -\frac{Pe}{k}, \quad (16)$$

$$x = 1 \quad C' = \frac{Pe}{k}, \quad (17)$$

$$x = 1 \quad y' = 0. \quad (18)$$

Conditions (13) and (14) reflect the absence of a diffusion flow of material through the lower boundary ($x = 0$); Eqs. (16), (17) characterize the organized flow of material, and thus, heat, through the upper boundary ($x = 1$); Eq. (15) reflects the change in oxidizer concentration n the gas traversing the lower boundary ($x = 0$), caused by gas mixing in the bed (the Dunquist condition).

It is difficult to solve this system of nonlinear equations, so we will linearize and seek an approximate analytical solution. To do this we make the following assumption: due to intense mixing of material in the fluidized bed the fuel concentration and relative combustion rate in the bed are practically constant over layer height.

Now, solving Eq. (2) with boundary conditions (15), (18), we determine the oxygen concentration profile over bed height:

$$y = y_0 \frac{N}{k(N-1)} \frac{\exp \lambda_2 x - (\lambda_2/\lambda_1) \exp (\lambda_2 - \lambda_1 + \lambda_1 x)}{\frac{N}{k(N-1)} - \frac{\lambda_2}{\lambda_1} \left(\lambda_1 - \frac{N}{k(N-1)} \right) \exp (\lambda_2 - \lambda_1)}, \quad (19)$$

where

$$\lambda_{1,2} = \frac{N}{2k\varepsilon(N-1)} \left(1 \pm \sqrt{1 + 4Bk\varphi\zeta^0 C^0 \varepsilon \frac{N-1}{N}} \right). \quad (20)$$

The oxygen concentration in the stack gas at exit from the bed can be determined from the balance condition and expressed in terms of the excess air coefficient, which reflects the ratio between the oxidizer supplied and that required for complete combustion of the fuel:

$$y = y_0 \frac{\alpha - 1}{\alpha}. \quad (21)$$

The second term under the radical is small (less than 0.5), and therefore, $\lambda_1 \gg \lambda_2$. In this case Eq. (19), defining the oxygen concentration in the stack gas leaving the layer, simplifies:

$$\frac{y}{y_0} = \frac{\alpha - 1}{\alpha} = \exp \lambda_2. \quad (22)$$

With Eq. (20) we estimate the mean carbon particle combustion rate in the fluidized bed

$$\zeta^0 = \frac{\lambda_2}{\varphi BC^0} \left(\lambda_2 \varepsilon k - \frac{N}{N-1} \right). \quad (23)$$

On the other hand, the local combustion rate depends on the local temperature head in accordance with Eq. (6). We find this temperature increase by solving differential equation (3) with boundary conditions (14), (16), with consideration of the height dependence of oxidizer concentration, Eq. (19). After transformations we have

$$\theta = L_1 [-\exp \lambda_2 x + L_2 \exp \sqrt{\Phi_1} x + L_3 \exp (-\sqrt{\Phi_1} x)] + L_4 [\exp \sqrt{\Phi_1} x - \exp (-\sqrt{\Phi_1} x)], \quad (24)$$

where

$$L_1 = \frac{\varphi B m \zeta^0 q y_0}{k(\lambda_2^2 - \Phi_1)}; \quad L_2 = \frac{\lambda_2}{(\lambda_2^2 - \Phi_1) \sqrt{\Phi_1}} \cdot \frac{\exp(-\sqrt{\Phi_1}) - \exp \lambda_2}{\exp(-\sqrt{\Phi_1}) - \exp \sqrt{\Phi_1}};$$

$$L_3 = \frac{\lambda_2}{(\lambda_2^2 - \Phi_1) \sqrt{\Phi_1}} \cdot \frac{\exp \sqrt{\Phi_1} - \exp \lambda_2}{\exp(-\sqrt{\Phi_1}) - \exp \sqrt{\Phi_1}}; \quad L_4 = \frac{Pe}{kC^0 \sqrt{\Phi_1}} \frac{1}{\exp(-\sqrt{\Phi_1}) - \exp \sqrt{\Phi_1}}; \quad \Phi_1 = \Phi B \frac{\varphi}{k}.$$

We now substitute in Eq. (6) the expression for the temperature increase, Eq. (24), then integrate Eq. (6) to find the mean relative combustion rate. Then equating the expression obtained to the mean combustion rate of Eq. (23), we arrive at

$$\zeta^0 = \int_0^1 \zeta dx = \frac{\lambda_2}{\varphi BC^0} \left(\lambda_2 \varepsilon k - \frac{N}{N-1} \right). \quad (25)$$

Now we solve transcendental equation (25) numerically by the dichotomy method and find the value of λ_2 . We then find the main unknowns defining the process of coal combustion in the fluidized bed: the mean relative carbon particle combustion rate (Eq. (25)); the temperature increase or burning particle temperature (Eq. (24)); the relative oxygen concentration in the stack gas or excess air coefficient (Eq. (22)). Further, we can clarify the effect on these factors of the fundamental system parameters: fuel characteristics (d , ρ , E , Q_p^h), layer material characteristics (ρ_i , d_i), regime parameters (U , T_b , C^0), and layer height.

Thus the problem of coal combustion in a fluidized bed has been reduced to solution of transcendental equation (25). The procedure used for finding approximate solutions with use of auxiliary equation (6) has permitted us to consider the nonlinear dependence of coal combustion rate on temperature.

Having access to this approximate solution of the problem of coal combustion in a fluidized bed, we can consider the converse problem and attempt to determine the coefficients in the original system of differential equations for experimental data. These coefficients E , ϕ , k reflect the specifics of the system under consideration: activation energy, which defines the rate of the carbon fuel combustion reaction; the mean relative hot coal particle diameter, i.e., the combustion surface; and the intensity of mixing in the fuel-oxidizer system.

We will use experimental data on combustion of a 2.5-mm fraction coal in a fluidized layer of magnesite (fraction with mean size of 1 mm) [10]. The coal (WP = 3.7%, AP = 23.6%, $Q_p^h = 23.2$ J/kg) with low volatile content (14.73%) was burned in a column 146 mm in diameter. Experiments were performed with magnesite packed to heights of 0.1-0.3 m, air filtration rates from 1.3 to 1.8 m/sec at bed temperatures in the range 780-900°C. The experiments determined the mean concentration of fuel (coal) in the layer (C^0), bed temperature (T_b), and fuel and air expenditure. Gas analysis was used to determine the excess air coefficient. Experimental conditions and measurement techniques were described in greater detail in [10].

The experimentally determined oxygen concentration in the stack gas, or the excess air coefficient, will serve as a sufficiently complete characteristic of coal combustion in the fluidized bed. Therefore we seek the unknown coefficients by attempting to minimize the divergence between the experimental and calculated excess air coefficient for the entire mass of experimental data:

$$F = \sum_{i=1}^n (\alpha_i - \alpha_{i0})^2 = \min. \quad (26)$$

The minimum of this function was found numerically on a computer using an unconditional minimization program. Heat-exchange intensity and layer expansion were determined from previously proposed empirical expressions [11, 12]

$$Nu = 0.41 Ar^{0.3} \left(\frac{d_i}{d} \right)^{0.2} \left(\frac{\rho_p}{\rho_i} \right)^{0.07}, \quad (27)$$

$$\varepsilon = \varepsilon_0 (1 + 0.5Fr^{0.27}). \quad (28)$$

Using the above procedure, 10 experiments on coal combustion in a fluidized bed were generalized, with excess air coefficients ranging from 1.3 to 3.3. By processing the experimental data three coefficients defining the system were determined:

$$E = 118 \text{ kJ/mole}, \quad \phi = 1.88, \quad k = 0.12, \quad (29)$$

The mean-square deviation of the calculated and experimental excess air coefficients was 0.84. The maximum uncertainty in calculating the three coefficients was 3, 21, and 16%, respectively. For coal with a low volatile content the activation energy is approximately equal to 138 kJ/mole, which differs from the value obtained by no more than 17%. The value found for the second coefficient is practically the same as the predicted value. If coal combustion occurs at a constant rate, then the change in volume of the burning particle occurs in proportion to its surface area, and hence, the diameter changes linearly with time from unity to zero, and the mean relative diameter will be 0.5, i.e., $\phi = 2$. The value found for the third coefficient also coincides with the value obtained in [5], which studied mixing of material in a fluidized bed under cold conditions.

Thus, the agreement of the predicted values of E , ϕ , and k with experimentally determined values indicates the adequacy of the proposed model of coal combustion is a low height fluidized bed of coarsely dispersed material.

NOTATION

B, coefficient; C, C⁰, coal concentration in fluidized bed and mean value thereof; c_p, specific heat of coal; D, gas diffusion coefficient; D_s, solid phase diffusion coefficient; E, activation energy for coal combustion; H, layer height; k, coefficient; K_x, coal combustion rate constant; K, K, coal combustion rate and net rate; m, coefficient; Q_p^h, heat capacity of coal; R, universal gas constant; T_p, T₀, T_b, temperatures of hot particle, entering cold coal, and bed; T_x, characteristic temperature of coal combustion reaction; U, U₀, filtration velocity and velocity for onset of fluidization; x, x', coordinate; y, y₀, oxygen concentration in gas, current and initial values; W, coal supply rate; α, excess air coefficient; β, mass exchange coefficient; ε, ε₀, porosity of fluidized bed and initial value thereof; θ, θ̄, relative particle temperature increase and mean value thereof; λ, thermal conductivity coefficient of gas; ν, gas viscosity; ξ, ξ⁰, relative combustion rate and mean value thereof; ρ_p, ρ_i, ρ, densities of coal particle, inert material, and gas; reciprocal of relative mean coal particle size; ρ_p^{*}, density of gaseous carbon. Dimensionless complexes: Pe = W/(U - U₀); Re₀ = (U₀d_i)/ν; Sh = βd_i/D; Fr = (U - U₀)²/gH; Ar = (d_i³ g/ν²)(ρ_i/ρ - 1).

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