STABLE-COMBUSTION TEMPERATURE REGIME OF COAL IN A FLUIDIZED BED

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The system of energy, fuel, and oxidizer balance equations is analyzed and the temperature of the fluidized layer determined for ignition of the fuel and at the moment the ash component begins to soften.

Furnaces using a fluidized bed which permit efficient combustion of a solid fuel must operate with varying temperature of the bed in accordance with changes in thermal loading. The upper limit of permissible bed temperature is that at which the ash component of the fuel begins to soften [1], while the lower limit is defined by conditions required for stable combustion of the fuel, i.e., the ignition temperature [2]. Conditions for ignition of solid fuel were described in the well known studies [2-4].

The goal of the present study is to define fluidized bed parameters for fuel ignition and softening of the ash component by analyzing the system of energy, fuel, and oxidizer balance equations.

In fluidized bed furnaces the fuel is scattered in a hot layer of nonburning material. Within the layer it rapidly (in less than 10 sec) loses its volatile components, thus forming a coke residue.

The coke residue is set in motion by the chaotically moving nonfuel material, the mixing time of which is quite short (10-30 sec) in comparison to the burnup time of the coke residue (several min).

The oxidizer gas filtering through the layer moves in an orderly manner upward from below, while a fluctuation component is superposed on this motion by the chaotic motion of the solid phase. The fluidized bed contains only a small quantity of coal (less than 2%), so that the hydrodynamics of the system are determined by the nonfuel material and are independent of the fuel combustion process. This physical pattern of solid phase, gas, and fuel motion permits formulation of a system of one-dimensional steady-state balance equations:

for the fuel

$$C'' - \frac{m}{k}\nu\xi BYC = 0, \tag{1}$$

for the oxidizer (atmospheric oxygen)

$$Y'' - \frac{N}{\varepsilon k(N-1)}Y - \frac{\gamma\xi}{\varepsilon k}BYC = 0 \tag{2}$$

and for the energy

$$(C\Theta)'' + \frac{\nu q m\xi}{k}BYC - \frac{\Phi}{k}BC\Theta = 0. \tag{3}$$

Here the following dimensionless complexes characterizing the system have been introduced:

$$B = \frac{6K_*H}{d\varphi(U - U_0)};\quad m = \frac{\rho_*}{\rho_p(1-\varepsilon)},\quad \Theta = \frac{T_p - T_{cl}}{T_{cl} - T_0},$$

$$\Phi = \frac{\mathrm{Nu}\,\lambda}{K_*\rho_p c_p d_i(1-\varepsilon)},\quad q = \frac{Q_1 + 0{,}56\,Q_2}{(T_{cl} - T_0)c_p},$$

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together with the relative net coal combustion rate for particles on the surface of which there simultaneously occur oxidation reactions with formation of carbon monoxide (j = 2) and dioxide (j = 1) [5]:

$$\frac{1}{\xi}=\frac{K_* d_i}{\mathrm{Sh}\,D}+\sum_{j=1}^{2}\exp\frac{E_j}{RT_l}\left[\frac{1}{\Theta(1-T_0/T_l)+1}-\frac{T_l}{T_*}\right], \tag{4}$$

where 0.56 is the ratio of the reaction rates for j = 2, j = 1 at T_p = 1300°K. The system of differential equations (1)-(3) considers diffusion transport of fuel, oxidizer, and energy (first term in Eq. (1)-(3)), organized oxidizer transport (second term in Eq. (2)), fuel and oxidizer loss (second term in Eq. (1) and third term in Eq. (2)), as well as heat liberation by the hot particles and heat loss from them to the surrounding nonfuel particles forming the fluidized bed (second and third terms of Eq. (3)). Cold fuel enters through the upper boundary, while air is supplied through the lower boundary. There is no fuel or energy transport through the lower boundary. The boundary conditions then take on the following form:

$$x=0,\quad C'=0,\quad (C\Theta)'=0,\quad Y'=\frac{N}{k(N-1)}(Y-Y_0), \tag{5}$$

$$x=1,\quad C'=\mathrm{Pe}/k,\quad (C\Theta)'=-\frac{\mathrm{Pe}}{k},\quad Y'=0. \tag{6}$$

In the fluidized bed fuel mixing occurs quite intensely, so that fuel concentration may be assumed constant over height. Combustion of coal particles in the fluidized bed takes place at elevated air excesses and occurs primarily in the diffusion region, so that we take the combustion rate as constant and equal to its mean value. Under such conditions system (1)-(3) becomes linear, allowing determination of the dependence of oxidizer concentration on layer height [6]

$$Y=Y_0\exp\lambda_2 x \tag{7}$$

and its mean value for the layer as a whole

$$\overline{Y}=Y_0(\exp\lambda_2-1):\lambda_2, \tag{8}$$

where

$$\lambda_2=\frac{N}{2\varepsilon k(N-1)}\left[1-\sqrt{1+4\gamma\varepsilon k\xi B\overline{C}\left(\frac{N-1}{N}\right)^2}\right]. \tag{9}$$

The second term beneath the radical in this expression is markedly less than unity. Expanding the function in a Taylor series and limiting ourselves to linear terms, we obtain

$$\lambda_2=-\gamma\overline{\xi}\,\frac{N-1}{N}B\overline{C}. \tag{10}$$

Since the fuel concentration and relative gas density change little over bed height,

$$\int_0^1 mCY\,dx=\overline{m}\,\overline{C}\int_0^1 Y\,dx=\overline{m}\,\overline{C}\,\overline{Y}.$$

Using this relationship, we integrate fuel balance equation (1) with consideration of boundary conditions (5), (6):

$$\frac{\mathrm{Pe}}{k}-\overline{m}\,\overline{\xi}\frac{\nu}{k}B\overline{C}\,\overline{Y}=0, \tag{11}$$

whence we express the mean fuel concentration as

$$\overline{C}=\frac{\mathrm{Pe}}{\nu\overline{m}\,\overline{\xi}B\overline{Y}}. \tag{12}$$

We will now solve Eqs. (8), (10), and (12) simultaneously and determine the mean oxidizer and fuel concentrations in terms of the original system characteristics:

$$\lambda_2=-\frac{N-1}{N}\,\frac{\gamma\,\mathrm{Pe}}{\nu\overline{m}\,\overline{Y}}, \tag{13}$$

$$\overline{Y}=-\frac{N-1}{N}\,\frac{\gamma}{\nu m}\,\mathrm{Pe}\left[\ln\left(1-\frac{N-1}{N}\,\frac{\gamma\,\mathrm{Pe}}{\nu\overline{m}Y_0}\right)\right]^{-1}, \tag{14}$$

$$\bar{C} = \frac{N}{N-1}\,(\gamma\bar{\xi}B)^{-1}\ln\left(1 - \frac{N-1}{N}\,\frac{\gamma\,\mathrm{Pe}}{\nu\bar{m}Y_0}\right). \qquad (15)$$

We now integrate energy balance equation (3) with boundary conditions (5), (6)

$$-\frac{\mathrm{Pe}}{\bar{C}} + q\nu\bar{m}\bar{\xi}B\bar{Y} - \Phi B\bar{\Theta} = 0. \qquad (16)$$

Using Eqs. (14), (15), after transformations we obtain

$$\bar{\Theta} = \gamma\bar{\xi}\,\mathrm{Pe}\,\frac{N-1}{N}\,(1-q)\left[\Phi\ln\left(1 - \frac{N-1}{N}\,\frac{\gamma\,\mathrm{Pe}}{\nu\bar{m}Y_0}\right)\right]^{-1}. \qquad (17)$$

Since the fuel is intensely mixed in the fluidized bed, the temperature of the hot particles does not vary greatly over bed height. Therefore, we will define the mean fuel combustion rate in the bed with Eq. (4), using the mean particle temperature. We now introduce notation for all coefficients which are independent of particle temperature:

$$L_1 = \frac{N-1}{N}\,\mathrm{Pe}\,\frac{\gamma}{\nu m_0 Y_0}, \quad L_2 = \frac{N-1}{N}\,\frac{\gamma d_i K_* \rho_\mathrm{p} c_\mathrm{p}(1-\varepsilon)\,\mathrm{Pe}}{\mathrm{Nu}\,\lambda},$$

$$L_3 = \frac{K_* d_i}{\mathrm{Sh}\,D}, \quad L_4 = -\frac{E}{RT_*}, \quad L_5 = \frac{E}{RT_0}, \quad L_6 = \frac{Q_2 + 0{,}56\,Q_1}{c_\mathrm{p}T_0}.$$

We rewrite energy balance equation (17):

$$\Theta_{cl} = \frac{Z\,[L_3 + L_4\exp(L_5/Z)\ln(1 - L_1 Z) + L_2(1 + L_6)}{L_2 + [L_3 + L_4\exp(L_5/Z)]\ln(1 - L_1 Z)}, \qquad (18)$$

where

$$Z = \bar{T}_\mathrm{p}/T_0; \quad \Theta_l = T_l/T_0; \quad m_0 = \frac{0{,}536}{\rho_\mathrm{p}(1-\varepsilon)}\,\frac{273}{T_0};$$

$$\bar{m} = m_0/Z; \quad \bar{\Theta} = (Z - \Theta_l) : (\Theta_l - 1).$$

We will use Semenov's thermal ignition theory to determine the fuel ignition temperature. If we construct the dependence of heat liberation and heat loss upon temperature of the hot coke, the moment of particle ignition corresponds to the point of tangency of the heat liberation and loss curves [3, 4]. We determine the curve tangency point and the corresponding coke particle ignition temperature by taking the derivative with respect to Z of energy balance equation (18). After transformations, the carbon ignition condition can be described by the following equation

$$(Z - \Theta)^{-1} = \frac{L_4 L_5\exp(L_5/Z)}{Z^2[L_3 + L_4\exp(L_5/Z)]} + \frac{L_1}{(1 - L_1 Z)\ln(1 - L_1 Z)}. \qquad (19)$$

The layer temperature, which is defined by Eq. (18), appears in this expression. The problem of determining the ignition temperature of a carbon particle in a fluidized bed and the corresponding bed temperature has thus been reduced to solution of transcendental equation (19).

The problem was solved numerically by the dichotomy method for a concrete example: T_0 = 293°K, N = 4, d_i = 1 mm, d = 2.5 mm, K_* = 100 m/sec, T_* = 2600°K. Calculations revealed that the major parameter determining bed temperature for coal ignition in the fluidized bed is the activation energy. Thus, for example, for activation energies of 117.2, 125.6, and 134 kJ/mole the coal ignition temperature was 714.9, 768.7, and 822.3°C, respectively.

Change in other system parameters by almost an order of magnitude had little effect on the layer ignition temperature. Since the activation energy of coals varies from 117.2 to 134 kJ/mole, the minimum stable fluidized bed temperature at which combustion is obtained lies in the interval 715-820°C, which agrees approximately with experimental data. Thus, the minimum fluidized bed temperature at which stable combustion was observed in our experiments (V^f = 14%) comprised 775-780°C. According to other data, for Donets coal (V^f = 4-5%) this temperature equals 720-740°C [7].

The data of [8] indicate the bed temperature at which stable combustion of coke screenings (V^f = 3.1%) occurred was 818-830°C, while for coal combustion (V^f = 20.8%) the corresponding temperature was 737°C.

Thus, the lower temperature limit for a fluidized bed has been determined. The upper temperature limit of the bed is determined by the inadmissibility of agglomeration (sintering), i.e., achievement of a hot particle temperature at which the ash component begins to soften. If the temperature at which ash softening commences is known, then we can find the maximum permissible operating temperature for a fluidized bed furnace from Eq. (8). Thus, in the authors' experiments on combustion of Donets coal [1] with an ash softening point from 1473 to 1523°K, the upper temperature limit for the fluidized bed proved to be in the interval 1173-1197°K. Calculation of this temperature with Eq. (8) gives a range of 1159-1260°K, which agrees satisfactorily with experiment.

Thus, the operating temperature range for a fluidized bed furnace and its dependence on fuel characteristics and system parameters have been determined.

NOTATION

C, fuel concentration in fluidized bed; c_p, specific heat of fuel particle; D, gas diffusion coefficient; d, d_i, diameter of fuel and inert particles; E, activation energy; H, bed height; K_*, carbon combustion rate constant; N, fluidization number; k, effective dimensionless mixing coefficient; M_C, molecular mass of carbon; Q_1, Q_2, heats of reaction for formation of carbon dioxide and monoxide; R, universal gas constant; T_p, T_0, T_ℓ, temperatures of hot particle, incoming cold coal, and layer; T_*, characteristic temperature of coal combustion reaction; U and U_0, filtration and fluidization onset rates; Y, oxygen concentration in the gas; w, coal supply rate; α, heat exchange coefficient for hot coal particle with bed; β, mass exchange coefficient; γ, carbon content of fuel; ε, porosity of fluidized bed; λ, thermal conductivity coefficient of gas; ξ, relative combustion rate; ρ_p, ρ, ρ_*, densities of coal particle, inert material, and gaseous carbon; ν, molar ratio of carbon and oxygen for simultaneous occurrence of both carbon oxidation reactions; φ, mean relative diameter of hot fuel particle; $Nu = \alpha d_i/\lambda$; $Sh = \beta d_i/D$; $Pe = w d_i/D$.

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