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GAS FORMATION FROM A SOLID FUEL BURNING IN A FLUIDIZED BED

V. A. Munts, A. P. Baskakov,
and A. A. Ashikhmin

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A method is given for calculating combustion performance, which incorporates various factors in boiler processes.

There are numerous model concepts [1-3] on fluidized-bed combustion and gasification, but most of them are too complicated for engineering calculations such as [1] or do not incorporate major features [2, 3].

The following reactions are usually considered [4] in describing gasification in carbon combustion:

$$2C + O_2 = 2CO, \tag{1}$$

$$C + O_2 = CO_2, \tag{2}$$

$$CO_2 + C = 2CO, \tag{3}$$

$$2CO + O_2 = 2CO_2. \tag{4}$$

Unfortunately, the true rate constants in (1)-(3) are unknown for most coals, with most of the data [5] relating to (2), with simultaneous allowance for (1) and (4). Carbon monoxide reacts with oxygen in the boundary layers around the coal particles, which is fairly complicated because the process is nonisothermal, and also because the combustion function is nonlinear. Mathematical analysis applied to a single particle shows [6] that the composition of the primary oxides formed by carbon reacting with oxygen is unimportant at temperatures exceeding the CO ignition point (about 700-750°C), since the carbon monoxide combines with oxygen at the surface and becomes the dioxide. We therefore neglect (1) and (4) individually on the basis that only (2) and (3) occur at the surface. The distributions for the concentrations of oxygen C, the monoxide C_1, and the dioxide C_2 over the combustion chamber height in the steady state are therefore defined [4] by

$$v\frac{\partial C}{\partial x} = -\alpha SC, \tag{5}$$

$$v\frac{\partial C_2}{\partial x} = -\alpha_2 SC_2 + \alpha SC, \tag{6}$$

$$C_0 = C + (1+\beta^*)C_2 + (0{,}605+\beta^*)C_1. \tag{7}$$

Here β^* is a fuel characteristic dependent on the composition [7].

The specific oxygen flux j to a single carbon particle is

$$j = \alpha^* C_{\text{p}}. \tag{8}$$

The overall rate constant for the reaction of a carbon particle is determined by the constant for the reaction rate (in general, allowance must be made for reaction in the pores) and the oxygen diffusion from the outer phase via the boundary layer to the burning particle:

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$$\alpha^* = \frac{1}{\dfrac{1}{k} + \dfrac{\delta}{\mathrm{Sh}\,D}}. \tag{9}$$

Here the Sherwood number defining the oxygen transport through the boundary layer is [8]

$$\mathrm{Sh} = 0{,}24\,\mathrm{Ar}^{0,333}(\delta/\delta_i)^{0,5}. \tag{10}$$

The amount of oxygen taken up by all particles in unit volume in the bed is $J = jS$ and is proportional to the specific reaction surface S, which is related to the relative fuel mass concentration by

$$S = \frac{6\rho(1-\varepsilon_0)\varphi z}{\delta\rho_c}. \tag{11}$$

The specific reaction surface is independent of the coordinate because of the vigorous mixing, and it can be calculated from the mean particle size in the range from the maximal in the ash to the value governed by the supply rate for a polydisperse fuel. With a major contribution from the fuel to the bed, the mean particle density is given by $\rho = \rho_i(1 - z) + \rho_c z$.

On the other hand, the oxygen flux consumed in reaction in unit volume passes from bubbles on account of mass transfer with the dense phase:

$$J = \beta(C_b - C_p). \tag{12}$$

The effective reaction rate constant is α, which is referred to the mean oxygen concentration in the bubbles in a given section, so $J = \alpha SC_b$. We equate the total oxygen flux written from (8) and (9) to the flux between the bubbles and the dense phase as in (12) and exclude the unknown concentration C_p to get the effective rate constant as

$$\alpha = \frac{1}{\dfrac{1}{\alpha^*} + \dfrac{S}{\beta}} = \frac{1}{\dfrac{1}{k} + \dfrac{\delta}{\mathrm{Sh}\,D} + \dfrac{S}{\beta}}. \tag{13}$$

This corresponds to the limiting two-phase stage, i.e., when all the gas passes through the bed as bubbles. That state is characteristic of a fluidized bed containing small inert particles at a relatively low fluidization speed.

Similar arguments for (3) give the total carbon dioxide consumption factor as

$$\alpha_2 = \frac{1}{\dfrac{1}{k_2} + \dfrac{\delta}{\mathrm{Sh}\,D} + \dfrac{S}{\beta}}. \tag{14}$$

At moderate fuel concentrations, such as are characteristic of combustion conditions, the combustion-product volume is virtually equal to the volume of air at the same temperature supplied. At the high fuel concentrations characteristic of gasification, the gas volume in air blowing may increase by not more than 20% because two moles of CO are formed from one mole of O_2. Mass-transfer coefficient correlations do not describe the process very accurately, so the gas speed as a function of position in the bed can be taken as constant even for gasification.

The mass-transfer coefficient β is dependent on the bubble diameter, which increases with height [9]:

$$d_b = 0{,}53\,(v - u_0)^{0,4} H^{0,7}. \tag{15}$$

The local mass-transfer coefficient β_l is [9]

$$\beta_l = \frac{6K(v - u_0)}{d_b(v - u_0 + 0{,}71\sqrt{gd_b})}. \tag{16}$$

The mean integral β is obtained by integrating (16) over the height on the basis of (15) and is used in (13) and (14), while K is given for sufficiently large particles by [10]

$$K = 0{,}75u_0 + 0{,}975D^{0,5}(g/d_b)^{0,25}. \tag{17}$$

The solution to (5) and (6) is

$$C = C_0 \exp\left(-\frac{\alpha SH}{v}\right), \tag{18}$$

$$C_2 = \frac{C_0}{1 - \dfrac{\alpha_2}{\alpha}}\left[\exp\left(-\frac{\alpha_2 SH}{v}\right) - \exp\left(-\frac{\alpha SH}{v}\right)\right]. \tag{19}$$

Then (7) defines C_1.

Another case of more practical interest is fluidization in a coarse material with flow bubbles, where the gas infiltration rate through the dense phase exceeds the bubble rise rate [10]. Then the gas passing between the particles penetrates the bubbles and passes onward into the dense phase. There is no bubble bypass transport in such a system, and the gas concentrations in the dense phase and the bubbles in a given section become the same. This is equivalent to infinitely large β in (13) and (14), i.e., the third term in the denominator vanishes there.

That case corresponds also to gas formation in a bed of small particles blown with a speed close to the critical value or in a bed of any particles if the fluidization rate greatly exceeds the critical value, i.e., there is a fluidization in turbulent or forced modes. Then the mass transfer from the bubbles to the dense phase is unimportant, i.e., the third term in the denominator in (13) and (14) becomes zero. The solution to (5) and (6) is again given by (18) and (19) with the substitution $\beta = \infty$ for the α appearing there. Here by C, C_1 and C_2 one understands not the oxygen, monoxide, and dioxide concentrations in the bubbles, as in the first case, but the values averaged over the bed.

In intermediate fluidization states, the reactant concentrations in the bubbles and dense phase vary in different ways with the height, and a fairly cumbersome solution occurs [3]. On the other hand, it is difficult to measure how the gas concentrations vary with height, since the proportions of the gases entering the sampling tube from the dense phase and the bubbles do not correspond to the flow-rate proportions. One can obtain a representative combustion-product samples averaged over the flow rates only at the exit from the bed after the gases emerging from the dense phase and the bubbles have mixed. Therefore, the reactant concentrations at the exit from the bed will for simplicity be derived also from (18) and (19) by substitution for α and α_2 calculated from interpolation formulas:

$$\alpha = \frac{1}{\dfrac{1}{k} + \dfrac{\delta}{\mathrm{Sh}\, D} + \dfrac{S}{\beta}\xi^2}; \quad \alpha_2 = \frac{1}{\dfrac{1}{k_2} + \dfrac{\delta}{\mathrm{Sh}\, D} + \dfrac{S}{\beta}\xi^2}. \tag{20}$$

The proportion of the gas passing through the bed as bubbles in the bubble fluidization mode is given by $\xi = 1 - (u_0/v)$; $\xi = 0$ in the turbulent or transit-bubble state. It is found that the agreement with experiment is better if quadratic interpolation with respect to ξ is used rather than simple. We took the activation energies for the oxidation of the coal by oxygen and the reduction of the dioxide by carbon for brown coals [5] as 92.1 and 146.5 kJ/mol, with k_0 as 3000 and 35,000 m/sec correspondingly.

The scheme was tested on a laboratory system by a method described previously [11]. Narrow fractions of Irsha–Boroda brown coal were used, mean particle sizes 2.5, 4, and 5.15 mm, composition: $W^r = 7.11\%$, $A^r = 13.39\%$, $V_g = 48.89\%$, $S^g = 0.39\%$, $O^g = 22.26$, $C^g = 71.26$, $H^g = 4.88$, $N^g = 1.21\%$. The bed material was a 120 μm electrocorundum fraction. The mass change in the fuel in the bed during combustion was recorded, and the products were analyzed. With fixed height, bed temperature, and fluidization speed, we varied the fuel concentration. The combustion products were analyzed after the volatiles had been released, when the stack gas composition remained almost constant. The gases were sampled only at the exit from the apparatus.

Figures 1-3 show the measurements and gas-component concentration calculations for the exit ($T = 1013$ K, $v = 0.35$ m/sec, $\delta_i = 120$ μm). The measurements are best compared with calculations based on the dimensionless combustion function for the fuel in unit bed area:

$$\Psi(z) = 0{,}5\left(1 - 3{,}76\frac{C - 0{,}5C_1}{100 - C_1 - C_2 - C}\right), \tag{21}$$

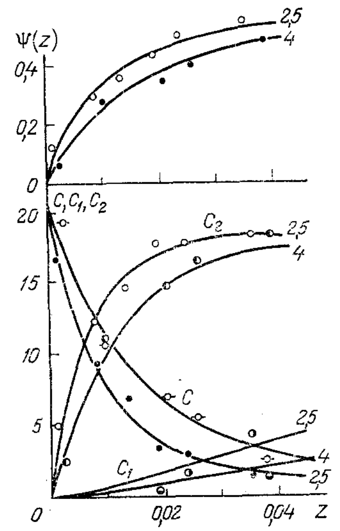

Fig. 1

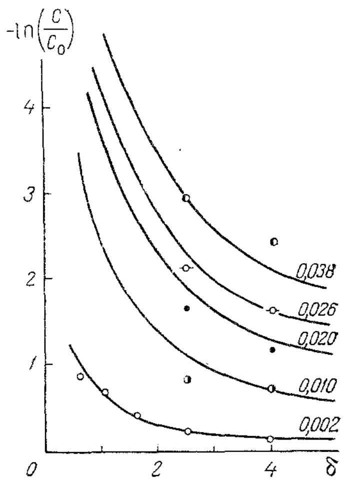

Fig. 2

Fig. 1. Variations in the dimensionless combustion function and in the product composition, vol. %, for low fuel concentrations ($z < 0(05)$); the numbers on the curves are the coal-particle sizes in mm, with the curves from theory and the points from experiment.

Fig. 2. Variation in oxygen concentration in the combustion products with coal-particle size, the numbers on the curves being the fuel concentration in the bed, with the curves derived from theory and the points from experiment; δ in mm.

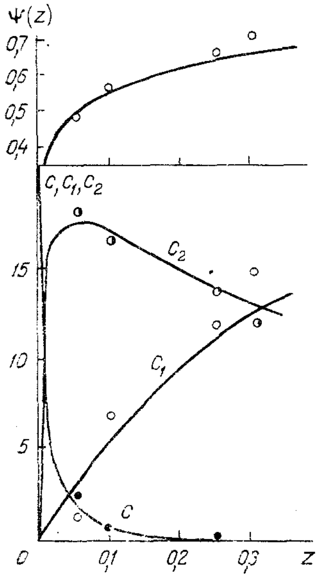

Fig. 3. Variation in dimensionless combustion function and in product composition, vol. %, for an elevated fuel concentration in the bed ($z < 0.4$, $\delta = 4$ mm), curves from theory, points from experiment.

which was proposed in [12] as a general dimensionless parameter.

Figures 1-3 show that this fairly simple method for calculating carbon combustion in a fluidized bed (curves) agrees satisfactorily with experiment (points) and provides for qualitative performance analysis as the parameters vary.

For low carbon concentrations z in the bed, the factors governing the burning rate are the reaction kinetics and the mass transfer with the oxygen in the dense phase. In that range ($z < 0.01$), there is a substantial increase in the oxygen consumption as the fuel content increases (Fig. 1). At higher z, there is an increasing effect on the combustion rate from mass transfer between the bubbles and the dense phase, and for $z \geq 0.02$-0.03, the increase in performance is due to the dioxide being reduced to CO. If the carbon concentration is raised to levels characteristic of gasification (Fig. 3), the carbon monoxide content in the gas increases considerably. There is a nonlinear relation between the combustion function and the carbon concentration, which shows that mass transfer between phases influences coal gasification in fluidized beds.

The coal-particle size has a considerable effect on the burning performance. The measurements and calculations (Fig. 2) show that the performance increases substantially as the size is reduced.

This scheme for calculating solid-fuel combustion in a fluidized bed is quite simple and can be used in engineering calculations. For example, it readily enables one to establish the effects on the performance from factors such as bed height and temperature, fluidization speed, and so on for various types of solid fuel.

NOTATION

C, C_1, C_2, and C_0, concentrations of oxygen, carbon monoxide, carbon dioxide, and oxygen in the blast correspondingly; C_p and C_b, oxygen concentrations in the dense phase and in the bubbles correspondingly; d_b, bubble diameter; g, acceleration due to gravity; D, effective diffusion coefficient; H, bed depth; J, oxygen flux to all carbon particles; k and k_2, rate constants; u_0, initial fluidization speed; S, reactive surface per unit volume; v, gas speed reckoned on the free volume; z, relative fuel concentration; α and α_2, overall consumption coefficients for oxygen in (2) and carbon dioxide in (3); δ and δ_i, sizes of coal and inert particles correspondingly; β^*, fuel characteristic dependent on composition; β and β_1, mean integral and local mass-transfer coefficients; ρ, ρ_i, and ρ_c, mean densities of bed particles, inert particles, and coke particles correspondingly; ε_0, bed porosity at fluidization limit; φ, bed expansion; ξ, fraction of oxygen passing through in bubbles; Ar and Sh Archimedes and Sherwood numbers.

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