INTERNATIONAL JOURNAL OF ENERGY RESEARCH, VOL. 15, 185–201 (1991)

# SPOUTED AND SPOUT-FLUID BED COMBUSTORS 1: DEVOLATILIZATION AND COMBUSTION OF COAL VOLATILES

S. TIA, S. C. BHATTACHARYA AND P. WIBULSWAS
*Energy Technology Division, Asian Institute of Technology, GPO Box 2754, Bangkok 10501, Thailand*

## SUMMARY

The devolatilization and volatile combustion of a single coal particle in spouted and spout-fluid beds have been studied. The results showed that the flame extinction time increases with the particle diameter, and decreases with the bed temperature. When the bed temperature and the air flow rate were fixed, the operation modes (spouted or spout-fluid bed) showed less effect on the mean flame extinction time. A mathematical model of the spouted bed mode for preignition and postignition periods has also been developed assuming the devolatilization rate to be controlled by heat transfer and multireaction pyrolysis kinetics based on volatile products. Ignition, heat transfer back from the volatile flame to the particle surface, variation in flame temperature, and the hydrodynamics of SB are taken into account. The model predictions, with some adjusting parameters, were in good agreement with experimental results.

KEY WORDS Coal devolatilization Spouted bed combustor Spout-fluid bed combustor

## INTRODUCTION

Normally, low grade coals consist of a substantial amount of volatiles which may produce up to 50% of the total heat released in the combustor. Knowledge of devolatilization and volatiles combustion rates as well as the location where the combustion takes place is, therefore, very important for the design and operation of the combustor. Experimental and theoretical studies on the devolatilization and combustion of coal volatiles in fluidized beds (FB) have been reported elsewhere (Atimtay, 1980; Stubington and Davidson, 1981; Pillai, 1981; La Nauze, 1982; Agarwal *et al.*, 1984, 1987; Prins *et al.*, 1989). Recent studies on spouted bed (SB) and spout-fluid bed (SFB) combustion have been focused on char combustion (Zhao *et al.*, 1987*b*) and combustion efficiency (Lim *et al.*, 1988), but the devolatilization and volatile combustion have received less attention. The work reported here is a first step towards quantitative studies on the devolatilization and combustion of coal volatiles in SB and SFB. Since SB and SFB are alternatives of FB with similar overall properties, the background on volatile combustion in FB is therefore essential.

Experimental studies on coal volatile combustion in an FB have correlated the devolatilization time empirically by a power law as a function of the initial coal particle diameter at various bed conditions (Pillai, 1981). The bed temperature has recently been included in the correlation by Ekinci *et al.* (1988). The modelling of single coal particle devolatilization under oxidizing conditions in FB was carried out by Borghi *et al.* (1977), and Agarwal *et al.* (1986, 1987). The burning of volatiles, in both the models, was assumed to be in the form of an enveloping diffusion flame, and the devolatilization rate was controlled by heat transfer and pyrolysis kinetics (Anthony *et al.*, 1975). Agarwal *et al.* (1986, 1987) also included the effect of bed hydrodynamics by considering residence times of fuel particle in bubble and emulsion phases in their model. Park *et al.* (1981) proposed a plume model for volatile combustion in an industrial FB combustor with under-bed feeding. The presence of volatile plume has been experimentally verified in a 4 $MW_{th}$ FB combustor by De Kok *et al.* (1985). Prins *et al.* (1989) reported from their two-dimensional FB experiment that above the bed surface and within the bubble interior coal volatiles burn with a yellow diffusion flame at the particle surface. A small blue, premixed, detached flame at the bed surface, above the coal particle situated a few centimetres below the bed surface in the emulsion phase, was also observed. Volatile flames have not been

0363–907X/91/030185–17$08.50

*Received 14 June 1990*
*Revised 20 September 1990*

observed in the dense phase of FB owing to the quenching effect of bed particles. They also found that the batch experiments gave longer flame extinction times than in the case of a single particle. As claimed by the authors, the results of a batch of experiments are determined by the last flames of just those few coal particles whose devolatilization time period happens to be relatively long. Their observed flame extinction times for a single coal particle were much lower than the predicted results from Agarwal's model (1986). For SB and SFB, Zhao *et al.* (1987*a*) reported the volatile combustion as a diffusion flame in the spout, fountain, and bed surface.

In this paper the model proposed by Agarwal *et al.* (1987) is applied for devolatilization and volatile combustion of a single coal particle in SB. The analysis is divided into preignition and postignition periods, in which the coal volatiles evolution rate is assumed to be controlled by heat transfer and multireaction pyrolysis kinetics proposed by Suuberg *et al.* (1978). Ignition, heat transfer back from the volatile flame to the particle surface, variation in flame temperature, and the hydrodynamics of SB are taken into account. Results from the model prediction are compared with experimental results.

## MODEL FORMULATION

The combustion behaviour of coal particles in a spouted bed has been reported by Tia (1990). The experimental result showed that when a single coal particle was dropped into the bed the heterogeneous ignition of fine particles resulting from abrasion between itself and inert sand particles was clearly observed. After 2–4 s, volatile combustion starts in the form of a diffusion flame surrounding the particle as it travels from the spout to fountain region and then falls back to the bed surface in the annulus region. It moves downwards into the annulus with the inert sand and is entrained again to the spout region to complete the cyclic movement. Inside the annulus regions, where observation is difficult, the volatile flame envelope is expected to be extinguished owing to the high heat loss rate from flame front to the surrounding sand particles. Fragmentation of coal particles during the combustion of volatiles is also observed.

Considering the combustion behaviour described above, the theoretical model for devolatilization and volatile combustion of a single coal particle in SB is developed based on the following assumptions:

(*a*) In an SB, a coal particle cyclically moves from the annulus to the spout and the fountain, and then falls back again to the annulus. It is assumed that the mean cycle time of a given coal particle size is constant and equal for each circulation.

(*b*) The process of devolatilization, which has been assumed to be thermally neutral (Agarwal *et al.*, 1984; Devanathan and Saxena, 1986), takes place at constant volume with spherical symmetry and no secondary reaction. The devolatilization rate is assumed to be controlled by the devolatilization kinetics and heat transfer (both to and through the particle) during both preignition and postignition stages.

(*c*) Quasisteady state exists in the gas phase, and the transport processes and chemical reactions take place within the gas film, whose thickness is a function of the ambient gas velocity and temperature.

(*d*) The gas-phase oxidation of the volatiles can be represented by an overall kinetic expression that is assumed to be a one-step irreversible chemical reaction, subject to second-order Arrhenius kinetics. In addition, the effects of chemical changes and energy release in the gas phase, prior to ignition, on both the gas and solid phases are negligible.

(*e*) Owing to the quenching effect of sand particles in the annulus, the ignition of volatiles can occur only in the spout and the fountain after the ignition criterion is satisfied. After that (postignition period), the burning of coal volatiles, which takes place within the gas-film boundary layer, as an enveloping diffusion flame with flame sheet approximation, occurs in both the spout and the fountain regions; while in the annulus region the flame is assumed to be extinguished.

(*f*) The volatile flame radius is approximately equal to the sum of the coal particle radius and the halfwidth of the oxygen mass-transfer boundary layer thickness (Agarwal, 1986).

(*g*) The volatile surface flux at which the volatile flame is extinguished, estimated from a single particle model of coal devolatilization in the hot oxidizing gas stream (Tia, 1990), is used as the flame extinction criterion. This criterion was suggested by Essenhigh *et al.* (1989) for coal ignition.

(*h*) The heterogeneous reaction between coal and oxygen is neglected during the preignition stage owing to the low surface temperature. In addition a coal particle, prior to ignition, is assumed to be a nonporous solid.

(*i*) The effects of volatile flame on radiative heat transfer between the particle and ambient are ignored.

(*j*) Thermophysical properties of air and flue gas are used for the gas phase in the preignition and postignition stages, respectively.

*(i) Preignition period*

A coal particle, after being dropped into the bed from the top, starts to travel from the upper surface of the annulus downwards, and then gets entrained into the spout and the fountain, the residence times in the three regions being $t_a$, $t_s$ and $t_f$, respectively. During circulation, it is heated up and the volatiles begin to evolve.

*Annulus region*

The temperature profiles inside the coal particle can be written as

$$\frac{\partial}{\partial t}(\rho_p C_{pp} T_p) = \frac{k_p}{r^2}\frac{\partial}{\partial r}\left(r^2 \frac{\partial T_p}{\partial r}\right) \tag{1}$$

with the following boundary conditions:

$$T_p = T_0; \quad 0 < r < R, \quad t = 0 \tag{2}$$

$$\frac{\partial T_p}{\partial r} = 0; \quad r = 0, \quad t \geqslant 0 \tag{3}$$

$$k_p \frac{\partial T_p}{\partial r} = h_t(T_b - T_p); \quad r = R, \quad (n-1)t_c < t < (n-1)t_c + t_a \tag{4}$$

$$h_t = h_{cc} + h_r \tag{5}$$

where $\rho$ = density, $C_p$ = specific heat, $k$ = thermal conductivity, $r$ = radial position, $R$ = particle radius, $t$ = time, $T$ = temperature, $n$ = number of circulations (1, 2, 3, . . .), and $h_t$, $h_{cc}$, and $h_r$ are the total heat transfer coefficient, the combined conduction and convection heat transfer coefficient between bed and coal particle, and the radiation heat transfer coefficient, respectively. Subscripts 0, $b$, and $p$ refer to the initial condition, the bed, and the coal particle, respectively.

The value of $h_{cc}$ is estimated from Broughton and Howard (1983)

$$h_{cc} = 2k_E/D_p + 0{\cdot}016/\sqrt{D_i} \quad (\mathrm{W/cm^2K}) \tag{6}$$

where $k_E$ is the effective thermal conductivity of the packed bed, and $D_i$ and $D_p$ are the diameter of the inert sand and coal particles, respectively. The value of $k_E$ can be estimated from Xavier and Davidson (1985)

$$k_E = k_E^0 + k_E^t \tag{7}$$

$$k_E^0/k_g = (k_p/k_g)^m \tag{8}$$

$$m = 0{\cdot}28 - 0{\cdot}757 \log \varepsilon_a - 0{\cdot}057 \log(k_p/k_g) \tag{9}$$

$$k_E^t/k_g = 0{\cdot}1\,\mathrm{Re}_a \mathrm{Pr} \tag{10}$$

where $k_E^0$ = effective thermal conductivity of packed bed with motionless fluid, $k_E^t$ = turbulent eddy conduction, $k_g$ = gas thermal conductivity, $\mathrm{Re}_a$ = coal-particle Reynolds number in the annulus, Pr = Prandtl number, and $\varepsilon_a$ = annulus voidage.

The value of $h_r$ is estimated from Baskakov (1985)

$$h_r = 7{\cdot}3\sigma\varepsilon_p\varepsilon_i T_R^3 \tag{11}$$

where $\sigma$ = Stefan–Boltzmann constant, and $\varepsilon_p$ and $\varepsilon_i$ are the emissivities of coal and inert sand particles, respectively.

The devolatilization rate $m_v$ is calculated from the multireaction kinetics (Suuberg *et al.*, 1978) by considering the rates of the different species evolved in the different infinitesimal volume elements of the particle in the form of spherical shells, and can be written as

$$m_v = \sum_{m=1}^{M} \sum_{i=1}^{n} \left(\frac{dV_i}{dt}\right)_m = \sum_{m=1}^{M} \sum_{i=1}^{n} \left[k_{ov,i}(V_i^* - V_i)\exp\left\{-\frac{E_i}{R_g T_p(t)}\right\}\right]_m \tag{12}$$

where $V_i$ = amount of volatile type $i$ evolved, and $V_i$ becomes $V_i^*$ when $t$ approaches infinity, $k_{ov,i}$ = frequency factor, $E_i$ = activation energy, and $R_g$ is the universal gas constant.

*Spout and fountain regions*

In the spout and fountain regions, where the ignition of volatiles can occur, the solid phase (coal particle) is also governed by equations (1) and (12). Neglecting the effect of inert sand on the coal particles, the energy equation for the gas phase is

$$\frac{1}{r^2}\frac{\partial}{\partial r}\left(k_g r^2 \frac{\partial T_g}{\partial r}\right) - \frac{m_v}{4\pi r^2}\frac{\partial (C_{pg} T_g)}{\partial r} = 0; \quad R < r < R + \delta_{th} \tag{13}$$

and the quasisteady mass transfer equations for the gas phase are

$$\frac{1}{r^2}\frac{\partial}{\partial r}\left(\rho_g D_v r^2 \frac{\partial Y_v}{\partial r}\right) - \frac{m_v}{4\pi r^2}\frac{\partial Y_v}{\partial r} = 0; \quad R < r < R + \delta_v \tag{14}$$

$$\frac{1}{r^2}\frac{\partial}{\partial r}\left(\rho_g D_o r^2 \frac{\partial Y_o}{\partial r}\right) - \frac{m_v}{4\pi r^2}\frac{\partial Y_o}{\partial r} = 0; \quad R < r < R + \delta_o \tag{15}$$

The initial conditions for the spout and the fountain regions, respectively, are

$$T_p = T_p(r), \quad \rho_p = \rho_p(r); \quad 0 < r < R, \quad t = (n-1)t_c + t_a \tag{16}$$

$$T_p = T_p(r), \quad \rho_p = \rho_p(r); \quad 0 < r < R, \quad t = (n-1)t_c + t_a + t_s \tag{17}$$

The boundary conditions for both the spout and the fountain regions are the same. These are as follows:

$$\frac{\partial T_p}{\partial r} = 0; \quad r = 0, \quad t \geqslant 0 \tag{18}$$

$$k_p \frac{\partial T_p}{\partial r} = k_g \frac{\partial T_g}{\partial r} + \sigma\varepsilon_p (T_b^4 - T_R^4); \quad r = R, \quad t > 0 \tag{19}$$

$$T_g = T_R; \quad r = R, \quad t \geqslant 0 \tag{20}$$

$$T_g = T_b; \quad r = R + \delta_{th}, \quad t \geqslant 0 \tag{21}$$

$$Y_v = Y_{v,R}; \quad r = R, \quad t \geqslant 0 \tag{22}$$

$$Y_v = 0; \quad r = R + \delta_v, \quad t \geqslant 0 \tag{23}$$

$$Y_o = Y_{o,\infty}; \quad r = R + \delta_o, \quad t \geqslant 0 \tag{24}$$

$$\rho_g D_o \frac{\partial Y_o}{\partial r} = \frac{m_v}{4\pi R^2} Y_{o,R}; \quad r = R, \quad t \geqslant 0 \tag{25}$$

$$\frac{m_v}{4\pi R^2}(1 - Y_{v,R}) = -\rho_g D_v \frac{\partial Y_v}{\partial r}; \quad r = R, \quad t \geqslant 0 \tag{26}$$

where $D$ = mass diffusivity, $\delta$ = boundary layer thickness, $Y$ = mass fraction, $m_v$ = volatile evolution rate, and subscripts $\infty$, $g$, $th$, $o$, $R$, and $v$ refer to mainstream, gas phase, thermal, oxygen, coal surface and coal volatiles, respectively. Note that heat conduction from the ambient to the particle surface, $k_g \partial T_g/\partial r$ in equation (19), can be estimated from the gas-phase temperature profiles obtained by solving equation (14) and differentiating with respect to $r$. However, no volatile is evolved at the beginning of the preignition period, and so this heat conduction term can be evaluated as for simple steady state heat conduction of a concentric sphere, which is equal to $k_g(T_b - T_R)/\{R(1 - R/(R + \delta_{th}))\}$.

The boundary layer thickness $\delta_j(j = th, o, v)$, is estimated from Sangiovanni (1978)

$$\delta_j/R = 2/(\mathrm{Nu}_i - 2); \quad j = th, o, v, \quad i = h, m \tag{27}$$

where $\mathrm{Nu}_i$ = Nusselt number which can be calculated from the correlation developed by Ranz and Marshall (1952) under force convection as follows:

$$\mathrm{Nu}_h = 2 + 0{\cdot}69\,\mathrm{Re}^{1/2}\,\mathrm{Pr}^{1/3} \tag{28}$$

$$\mathrm{Nu}_m = 2 + 0{\cdot}69\,\mathrm{Re}^{1/2}\,\mathrm{Sc}^{1/3} \tag{29}$$

where $\mathrm{Nu}_h = 2Rh/k_g$, $\mathrm{Nu}_m = 2Rk_d/\rho_g D_j$, Re = Reynolds number, $\mathrm{Pr} = C_{pg}\mu_g/k_g$, $\mathrm{Sc} = \mu_g/\rho_g D_j$, $k_d$ = mass transfer coefficient, $h$ = heat transfer coefficient, and $\mu_g$ = gas viscosity. For the present investigation, the gas phase properties are evaluated at the mean temperature of ambient and surface.

Note that the particle Reynolds number in the annulus ($\mathrm{Re}_a$), the spout ($\mathrm{Re}_s$) and the fountain ($\mathrm{Re}_f$) are, respectively.

$$\mathrm{Re}_a = \frac{D_p(u_a + v_a)}{\nu_g} \tag{30a}$$

$$\mathrm{Re}_s = \frac{D_p(u_s - v_s)}{\nu_g} \tag{30b}$$

$$\mathrm{Re}_f = \frac{D_p u_f}{\nu_g} \tag{30c}$$

where $\nu_g$ is the kinematic viscosity of air or gas. The mean interstitial gas velocities $u_s$, $u_f$ and $u_a$ in the spout, fountain and annulus regions, respectively, and the mean velocities $v_a$ and $v_s$ of fuel particles in the annulus and the spout, respectively, can be estimated from the hydrodynamic correlations of the spouted bed as shown in the following Section.

The ignition criterion selected for this study depends on the chemical reaction rate between the volatiles and oxygen in the gas phase as proposed by Kashiwagi and Summerfield (1973).

$$\int_R^{R+\delta_v} k_{0,vc}\,\rho_g(r)\,Y_o(r)\,Y_v(r)\exp\{-E_v/R_g T_g(r)\}\,dr \geqslant C^* \qquad \mathrm{g/cm^2\,s} \tag{31}$$

where $C^*$ = critical value for gas-phase ignition, $k_{0,vc}$ = frequency factor ($\mathrm{s}^{-1}$), $E_v$ = activation energy for volatiles combustion, and the unit of $\rho_g$ is in grammes per cubic centimetre. This criterion is close to the experimental criterion of a minimum detectable light intensity seen by photomultipliers viewing the boundary layer (Kashiwagi and Summerfield, 1973). The values of $T_g(r)$, $Y_v(r)$, and $Y_o(r)$ can be estimated from equations (13)–(15) with equations (16)–(26) as boundary conditions.

The model for this period can be represented schematically as shown in Figure 1*a*.

*(b) Postignition period*

Suppose that the ignition of volatiles happens at the $(n + 1)$th circulation in either the spout or the fountain. The analysis for each region is as follows:

*Spout and fountain regions*

In these regions, where volatiles burn as an enveloping diffusion flame, the temperature variation and devolatilization rate of coal particle are also governed by equations (1) and (12) with the following initial and

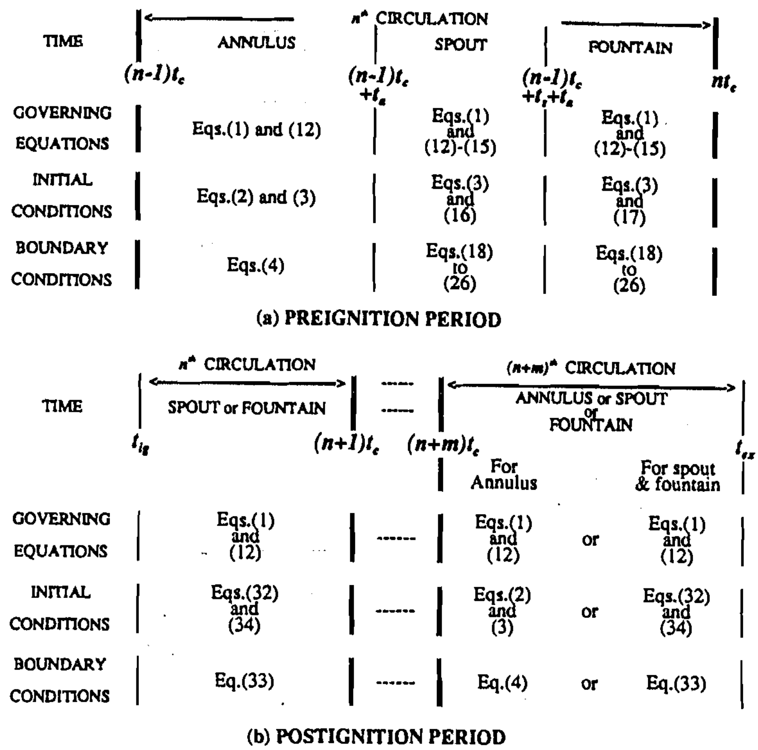



Figure 1. Schematic representation of the model: (*a*) preignition, and (*b*) postignition periods

boundary conditions:

$$\rho_p = \rho_p(r), \quad T_p = T_p(r); \quad 0 < r < R, \quad t = t_{ig} \tag{32}$$

$$k_p \frac{\partial T_p}{\partial r} = k_g \frac{\partial T_g}{\partial r} + \sigma\varepsilon_p(T_b^4 - T_R^4) + \frac{\sigma\varepsilon_p(T_F^4 - T_R^4)}{(1 + \varepsilon_p(R/R_F)^2(1/\varepsilon_F - 1))}; \quad r = R, \quad t > t_{ig} \tag{33}$$

$$\frac{\partial T_p}{\partial r} = 0; \quad r = 0, \quad t \geqslant t_{ig} \tag{34}$$

where $t_{ig}$ = ignition time delay, and $T_F$ and $R_F$ are the flame temperature and radius, respectively. Initial temperature and density in equation (32) can be calculated from preignition period with $t = t_{ig}$.

The heat conduction from flame front to the particle surface in equation (33) can be evaluated by solving equation (13) with $R < r < R_F$, and differentiating the result to yield:

$$k_g \frac{\partial T_g}{\partial r} = \frac{k_g(T_F - T_R)\beta_{th}}{R\{\exp(\beta_{th}(1 - R/R_F)) - 1\}}; \quad r = R \tag{35}$$

where $\beta_{th} = m_v/4\pi R\rho_g\alpha_g$ and $\alpha_g$ = thermal diffusivity.

The flame radius $R_F$ is calculated from Agarwal (1986), see assumption *f* above.

$$R_F = R + \delta_0/2 \tag{36}$$

and has different values in the spout and in the fountain owing to the difference in boundary layer thickness or the gas velocity.

The flame temperature is estimated from the energy balance equation. A part of the heat released from gas-phase volatile combustion preheats the volatile fuel itself and combustion air, and the remainder is transferred to the particle and the ambient. It is assumed that the heat of combustion of volatiles is ependent of temperature, and that the specific heats of all gases are equal. Neglecting the thermal dissociation effect, the energy balance can be written as

$$m_v H_v = m_v C_{pg}(T_F - T_R) + \frac{m_v \gamma}{Y_{0,\infty}} C_{pg}(T_F - T_b) + Q_{l,cp} + Q_{l,ca} + Q_{l,ra} + Q_{l,rp} \tag{37}$$

where $H_v$ = heat combustion of gaseous volatiles per unit mass; $Q_{l,cp}$ and $Q_{l,ca}$ are the conduction heat losses from the flame to the particle and to ambient, respectively; and $Q_{l,ra}$ and $Q_{l,rp}$ are the radiation heat losses from the flame to ambient and to the particle. $Q_{l,cp}$ is equal to the right-hand side of equation (35) multiplied by the particle surface area, and $Q_{l,ca}$, $Q_{l,ra}$ and $Q_{l,rp}$ are given by

$$Q_{l,ca} = \frac{4\pi(R + \delta_{th})k_g(T_F - T_b)\beta_{th}^+}{1 - \exp(\beta_{th}^+(1 - (R + \delta_{th})/R_F))}; \quad r = R + \delta_{th} \tag{38}$$

$$Q_{l,ra} = 4\pi R_F^2 \sigma \varepsilon_F (T_F^4 - T_b^4) \tag{39}$$

$$Q_{l,rp} = \frac{4\pi R^2 \sigma \varepsilon_p (T_F^4 - T_R^4)}{1 + \varepsilon_p (R/R_F)^2 (1/\varepsilon_F - 1)} \tag{40}$$

where $\beta_{th}^+ = m_v \gamma / 4\pi(R + \delta_{th})\rho_g \alpha_g Y_{o,\infty}$, and $\gamma$ = stoichiometric oxygen fuel weight ratio.

Note that $H_v$, $\gamma$, and $R_F$ are time-dependent variables and depend on the evolved volatile species which can be determined, for each species, by means of equation (12).

*Annulus region*

For the annulus where no volatile flame was assumed, the governing equations and boundary conditions are the same as for the preignition period.

The schematic representation of the model for the postignition period is shown in Figure 1*b*. The number of particle circulations goes on from $n + 1$ to $n + 2 \ldots . n + m$ until the volatile flux at the coal surface is lower than a critical value, at which point the volatile flame extinction occurs. This means that the extinction criterion can be achieved in any of the three regions of the SB.

(c) *Residence times of coal particle in different regions of spouted bed*

From Figure 2, the fuel particle has a cyclical movement with the travelling length of $Z$ from the bed surface. Assuming an isothermal bed and constant spout diameter along the bed height $H$, the mean particle cycle

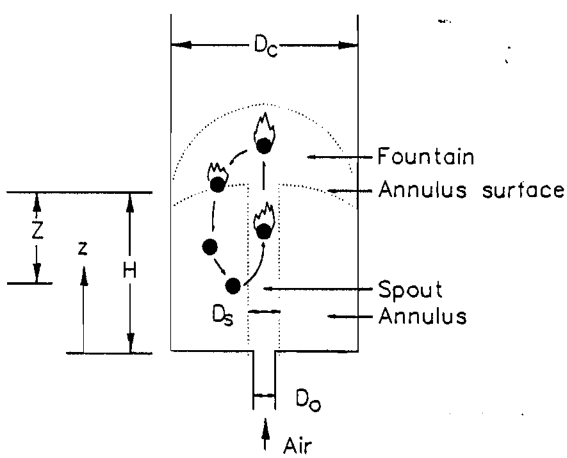



Figure 2. Cyclical movement of coal particle with the travelling length of $Z$ from the bed surface

time $t_c$ is defined as

$$t_c = t_a + t_s + t_f \tag{41}$$

and the ratio of mean particle residence time in annulus $t_a$ to annulus and spout $t_a + t_s$ is

$$t_a/(t_a + t_s) = t_a/(t_c - t_f) = \frac{Z/v_a}{Z/v_a + Z/v_s} = v_s/(v_s + v_a) \tag{42}$$

where $t_f$ is the mean particle residence time in fountain, and $v_a$ and $v_s$ are the average particle velocity in annulus and spout, respectively.

In the spout region, Lim (1975) used the relation of voidage and velocity proposed by Richardson and Zaki (1954) to determine the particle velocity, which can be written as

$$(u_s - v_s)/U_i = \varepsilon_s^{n-1} \tag{43}$$

$$\log U_i = \log U_t - D_p/D_s \tag{44}$$

$$u_s = U_s/\varepsilon_s \tag{45}$$

$$U_s = (UD_c - U_a(D_c^2 - D_s^2))/D_s^2 \tag{46}$$

$$U_a = \int_0^H U_{az}\,dz/H \tag{47}$$

where $u_s$ = average interstitial gas velocity in the spout, $\varepsilon_s$ = mean spout voidage, $U_i$ = terminal velocity of particle at $\varepsilon_s = 1$, $U_t$ = terminal free-falling velocity, $U_s$ = average superficial gas velocity in the spout, $U$ = superficial gas velocity, $U_a$ = average superficial gas velocity in the annulus, $U_{az}$ = superficial gas velocity in the annulus at any height $z$, $D_p$ = coal particle diameter, $D_s$ = mean spout diameter, and $n$ = a constant dependent on $D_p/D_s$ and the particle Reynolds number at $U_t$.

The mean interstitial gas velocity $u_f$ in the fountain region can be estimated from Grace and Mathur (1978):

$$u_f = U/\varepsilon_f \tag{48}$$

Assuming constant voidage $\varepsilon_a$ in the annulus, the particle velocity in the annulus can be estimated from a solid mass balance between the spout and the annulus as follows:

$$v_a = (1 - \varepsilon_s)v_s D_s^2/\{(D_c^2 - D_s^2)(1 - \varepsilon_a)\} \tag{49}$$

where $D_c$ is the column diameter.

The approximate value of the mean particle residence time in the fountain region is (Grace and Mathur, 1978)

$$t_f = 2v_{sH}/g \tag{50}$$

and, from equation (43),

$$v_{sH} = U_s/\varepsilon_{sH} - U_i\varepsilon_{sH}^{n-1} \tag{51}$$

where the subscript $H$ denotes the value at the bed surface.

Using the correlations and valves shown in Tables 1 and 2 with the air properties, the values of $u_s$, $v_s$, and $v_a$ can be estimated from equation (43)–(49). If $t_c$ is known, the particle residence times in each region can be determined from equations (41)–(42), (50), and (51).

The mean probabilities of a particle being in the fountain ($p_f$), the annulus ($p_a$), and the spout ($p_s$) during one circulation are

$$p_f = t_f/t_c \tag{52a}$$

$$p_a = t_a/t_c \tag{52b}$$

$$p_s = 1 - p_f - p_a \tag{52c}$$

Table 1. Hydrodynamic parameters of spouted bed

| Parameter | Reference |
|---|---|
| Superficial annulus gas velocity<br>$U_{az} = 0.88U_{mf}\{1-(1-z/H_m)^3\}$ | Piccinini *et al.* (1979) |
| Minimum fluidization velocity<br>$U_{mf} = \mu\{(31.6^2 + 0.0425\mathrm{Ar})^{0.5} - 31.6\}/\rho_g D_i$ | Thonglimp (1981) |
| Terminal free-falling velocity<br>$U_t = \left(\frac{4(\rho_p-\rho_g)^2 g^2}{225\rho_g\mu}\right)^{1/3} D_p;\quad 0.4 < \mathrm{Re}_p < 500$<br>$= \left(\frac{3.1g(\rho_p-\rho_g)D_p}{\rho_g}\right)^{0.5};\quad 500 < \mathrm{Re}_p < 2\times10^5$ | Kunii and Levenspiel (1969) |
| Minimum spouting velocity, cm/s<br>$U_{ms} = 75.8;\quad T_b = 1073\ \mathrm{K}$<br>$= 72.5;\quad T_b = 973\ \mathrm{K}$ | Experiment |
| Maximum spoutable bed height<br>$H_m = \frac{568b^2D_c^2}{\mathrm{Ar}D_i}\left(\frac{D_i}{D_0}\right)^{2/3}(\sqrt{1+35.9\times10^{-6}\mathrm{Ar}}-1)^2$<br>$b = 1.11$ | Epstein and Grace (1984) |
| Mean spout diameter, cm<br>$D_s = 470G^{0.4333}D_c^{0.5832}\mu^{0.1334}(\rho_b\rho_g g)^{-0.2834}$ | Wu *et al.* (1987) |
| $n$ in equations (43) and (51)<br>$n = (4.4 + 18D_p/D_s)\mathrm{Re}_t^{-0.03};\quad 0.2 < \mathrm{Re}_t < 1$<br>$= (4.4 + 18D_p/D_s)\mathrm{Re}_t^{-0.01};\quad 1 < \mathrm{Re}_t < 200$<br>$= 4.4\mathrm{Re}_t^{-0.1};\quad 200 < \mathrm{Re}_t < 500$<br>$= 2.4;\quad \mathrm{Re}_t > 500$ | Richardson and Zaki (1954) |
| Mean spout voidage | $\varepsilon_s = 0.85$ |
| Spout voidage at bed surface | $\varepsilon_{sH} = 0.7$ |
| Mean annulus voidage | $\varepsilon_a = 0.45$ |
| Mean fountain voidage | $\varepsilon_f = 0.65$ |

In the above relations, $\mu$ = gas viscosity, Ar = Archimedes number = $gD_i^3(\rho_i - \rho_g)\rho_g/\mu^2$, $D_0$ = inlet orifice diameter, $D_i$ = inert particle diameter, $G$ = mass velocity of gas, $\mathrm{Re}_t$ = Reynold number at $U_t$, $\rho_p$ = apparent coal density, $\rho_g$ = gas density, and $\rho_i$, $\rho_b$ = apparent and bulk density of inert particle, respectively.

## SOLUTION PROCEDURE

The calculation steps are shown in Figure 3. A time step of 0·01 s and 20 spherical shells were used for solving the equations by explicit finite-difference methods.

## EXPERIMENTAL DETAILS

The details of the experimental apparatus have been described elsewhere (Tia *et al.*, 1991). The combustor was made from stainless-steel pipe of 108 mm internal diameter and 800 mm high, heated externally by a resistance heater under automatic control (see Figure 4).

Table 2. Approximate analysis of lignite and properties of sand particle

| | |
|---|---|
| *Lignite* | |
| Volatile, % | 47·58 |
| Fixed carbon, % | 47·03 |
| Ash, % | 5·39 |
| Apparent density, g/cm$^3$ | 1·27 |
| *Sand* | |
| Diameter, mm | 2–2·39 |
| Bulk density, g/cm$^3$ | 1·45 |
| True density, g/cm$^3$ | 2·63 |

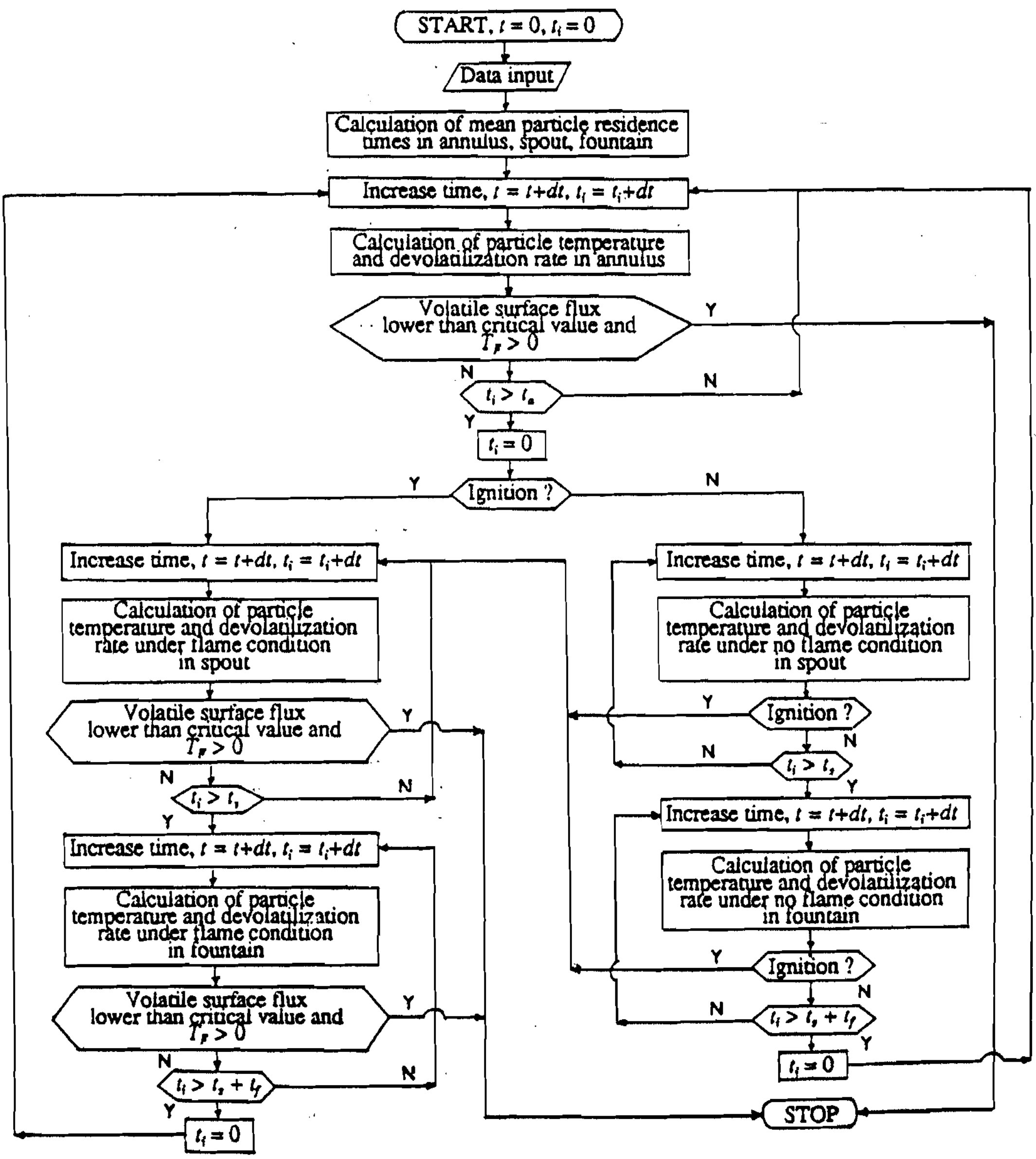



Figure 3. Calculation steps for determination of volatile flame extinction time in spouted bed combustor

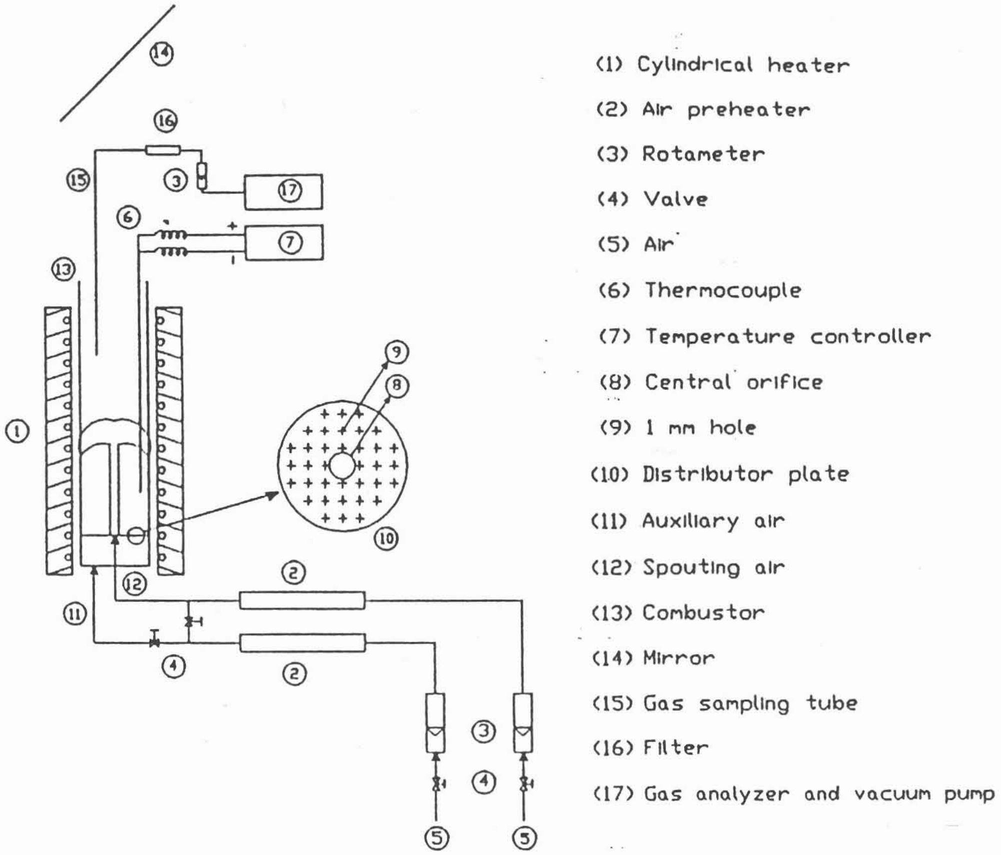



Figure 4. Schematic diagram of experimental equipment

Sand of size 2·19 mm was used as the inert bed material with a static bed height of 80 mm, and its properties are shown in Table 2. Lignite samples were prepared by crushing lumps, and screening to different size fractions. All the samples were oven dried for 24 h before testing. The properties of lignite samples are also shown in Table 2.

The bed was preheated under the SFB mode to about 700 K, and then 2–3 g of 1–2 mm lignite particles were dropped batchwise into the bed from above to increase its temperature, while the operation was changed to the desired mode at the selected ratio of superficial velocity to the minimum superficial spouting velocity ($U/U_{ms} = 1{\cdot}25$). As soon as the desired bed temperature was achieved and the inventory of lignite had burned out, a single lignite particle of known size was dropped into the bed from the top. The flame extinction time, defined as the time from the beginning to the time when the extinction of the volatile flame occurs, was measured by a stopwatch. The data for the samples that broke into fragments were rejected. Several readings were taken for each particle size in order to get reproducible results. The fluctuation of bed temperature during each experiment was within $\pm 20$ K.

For comparison, batch experiments were performed with 2 g samples at the selected bed condition. The effect of the air flow rate on the coal burnout time, which has been reported to be insignificant (Zhao *et al.*, 1987*b*), has not been studied.

## RESULTS AND DISCUSSION

### (a) Experimental finding

Because the devolatilization rate is controlled by the rate of heat transfer to and within the particle, the mean flame extinction time for all the conditions studied here therefore increased with the initial coal particle

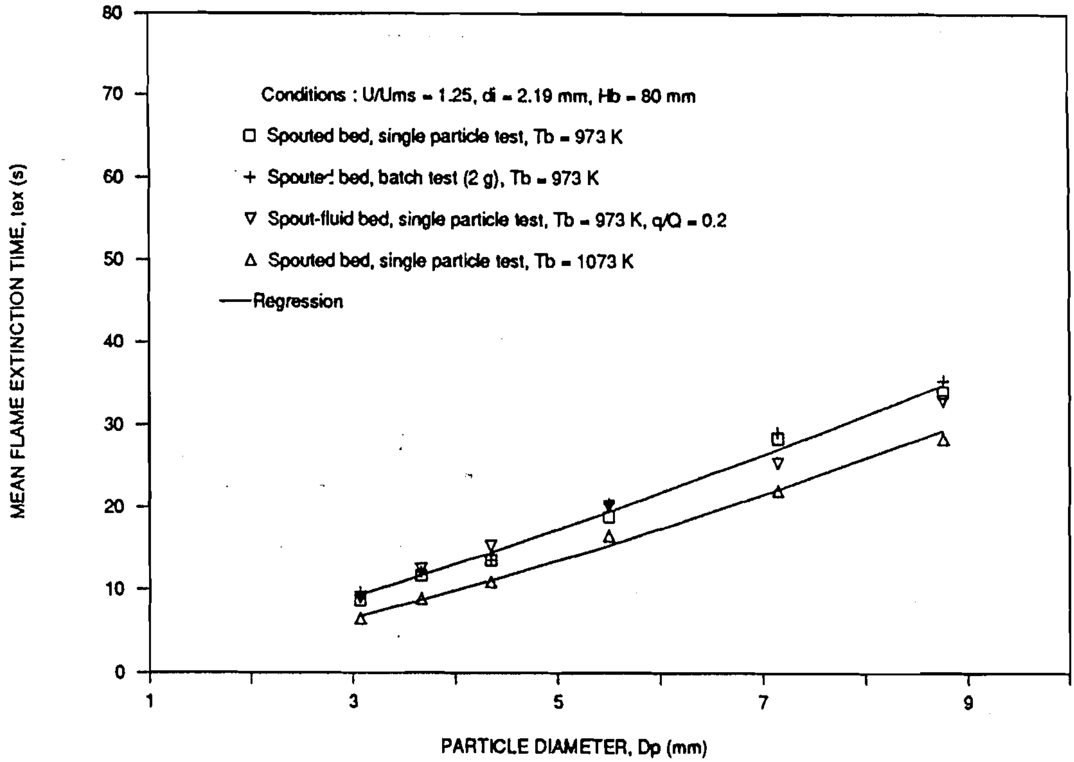



Figure 5. Mean flame extinction times as a function of initial particle diameter

diameter as shown in Figure 5. When the bed temperature and air flow rate are kept constant, the change in operation mode (SB to SFB) or from single particle to batch test (2 g mass) have little effect on the mean flame extinction time. But with an increase in bed temperature from 973 to 1073 K with other parameters fixed, for all particle sizes studied, the mean flame extinction time decreased significantly.

The empirical correlation between extinction time and particle diameter is of the form (Pillai, 1981)

$$t_{ex} = aD_{po}^{n} \tag{53}$$

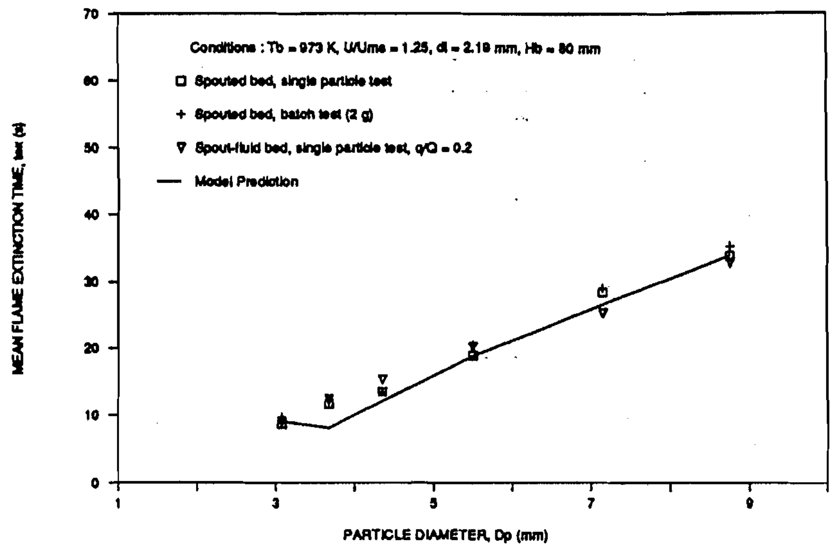
where $t_{ex}$ = flame extinction time in seconds, $D_{p0}$ = initial coal particle diameter in millimetres, and $a$ and $n$ are constants. Since in this case the mean flame extinction times for different operation modes, for a single particle, as well as for a batch test at 973 K bed temperature, were very close (Figure 5), and so all the data were combined for regression analysis. The analysis results gave values of $n$, and $a$ as 1·26 and 2·29 for 973 K and 1·39 and 1·43 for 1073 K bed temperatures, respectively. Correlation results are shown in Figure 5 by the solid lines. The values of $n$ are in line with those reported in the literature (Pillai, 1981; Ekinci *et al.*, 1988).

*(b) Model prediction*

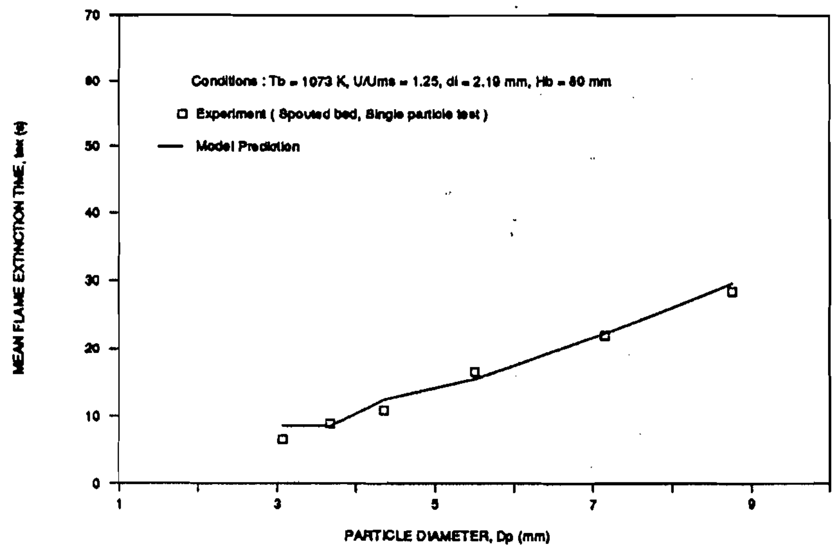
Since the bed and operating condition in this study are the same as those used in the study of the mean cycle times of coal and carbon particles in spouted and spout-fluid bed combustors performed by Tia (1990), the correlation of mean cycle time $t_c = 2{\cdot}078 \exp(4{\cdot}138/D_p)$, in conjunction with the hydrodynamic correlations shown in Table 1 were therefore used for the estimation of coal particle residence time in each region of the spouted bed as described in the preceding Section. For the conditions under which the experiment was performed, only the constant in the spout diameter correlation proposed by Wu *et al.* (1987) was adjusted (Table 1) in order to get a reasonable velocity of coal particle in the spout. The details of calculation have been described elsewhere (Tia, 1990).

Multireaction kinetic parameters of lignite pyrolysis proposed by Suuberg *et al.* (1978) were used in this study with a modification of the ultimate yield of volatile species (Tia, 1990).

Using the calculated particle residence time, pyrolysis kinetics, and the parameters shown in Tables 2 and 3 with the critical volatile surface flux of $7{\cdot}87 \times 10^{-4}$ and $8{\cdot}73 \times 10^{-4}$ g/cm$^2$ s$^1$ for 973 K and 1073 K bed temperature, respectively (calculated from a single particle model of devolatilization and combustion of coal volatiles in a hot oxidizing gas stream proposed by Tia (1990), at the extinction time for a 5·5 mm particle




(a)




(b)

Figure 6. Comparison of mean flame extinction times from model prediction and experiment at bed temperature (*a*) 973 K, and (*b*) 1073 K

Table 3. Coal properties and kinetic parameters of volatile combustion used in the model

| | | |
|---|---|---|
| Lignite thermal conductivity $k_p^{(a)}$ | $= 0{\cdot}0025$ | W/cm K |
| Apparent lignite density $\rho_{p0}$ | $= 1{\cdot}27$ | g/cm$^3$ |
| Critical reaction flux $C^*$ | $= 0{\cdot}0015$ | g/cm$^2$ s |
| Frequency factor of volatile combustion $k_{0,vc} \times 10^{7\,(b)}$ | $= 1{\cdot}868 - 0{\cdot}21\mathrm{Dp} + 0{\cdot}00904\mathrm{Dp}^2$ | s$^{-1}$ |
| Activation energy of volatile combustion $E_v^{(c)}$ | $= 66{\cdot}88$ | kJ/gmole |
| Lignite specific heat $C_{pp}^{(d)}$ | $= 0{\cdot}87 + 0{\cdot}00255(T - 273)$ | J/g K |
| Lignite emissivity $\varepsilon_p$ | $= 1$ | |
| Flame emissivity $\varepsilon_F$ | $= 0{\cdot}3$ | |
| Sand emissivity $\varepsilon_i$ | $= 0{\cdot}5$ | |

(*a*) Ragland and Yang (1985); (*b*) Tia (1990); (*c*) Biswas and Essenhigh (1972), and Rybak *et al.* (1987); (*d*) Bliek *et al.* (1985).

size, which is the mean size of the range used here), the model predictions are compared with experimental results in Figure 6. Good agreement is obtained for both the bed temperatures. Note that the value of $C^*$ was determined by adjustment until the ignition time obtained from the model was close to the experimental value for SB at $T_b = 973$ K, $D_p = 8{\cdot}76$ mm, and was then fixed and used for the other conditions. However, the calculated flame extinction time of 3·07 mm particle was slightly longer than that of the 3·76 mm particle. With the extinction criterion used here, this result is possible due to the relatively longer cycle time of the 3·07 mm particle, and hence longer time spent in the annulus (Tia, 1990). Inside this region, the devolatilization rate is low compared with those of the spout or fountain regions where it is enhanced by the volatile flame, and the flame extinction time, as a result, can be longer. It is also found from experiment that, for the two smallest sizes, the flame extinction occurred within one circulation, and no volatile flame is observed if the particle spends a relatively long time in the annulus. Therefore, the estimation of mean extinction time from visual observation is not suitable for the small particles.

No attempt has been made to develop a model for an SFB owing to the lack of hydrodynamic data.

Note that, from calculated results, the particle surface temperature increases rapidly in all the three regions of the SB during the preignition period. However, for a short time after ignition, this temperature can be higher than the bed temperature when the particle is in the spout and fountain regions owing to high volatile flame temperature. Therefore, when it falls back to the annulus where the flame extinguishes, the surface temperature decreases owing to the heat loss from the particle to the inert sand. The event repeats until flame extinction occurs.

## CONCLUSIONS

The devolatilization and volatile combustion of Thai lignite in an SB and an SFB have been studied. The flame extinction time increases with the particle diameter, and decreases when the bed temperature increases. This shows that heat transfer controls the devolatilization rate. In the present study, the difference in experimental techniques (batch or single particle) and operation modes (SB or SFB) showed less effect on the mean flame extinction time when the bed temperature and the air flow rate were fixed. The power-law correlation between the mean flame extinction time and the initial coal diameter gave the exponent values of 1·26 and 1·39 for 973 K and 1073 K bed temperatures, respectively. The results from the model prediction are in good agreement with experimental data.

## NOMENCLATURE

| | |
|---|---|
| $A$ | = cross-sectional area, cm$^{-2}$ |
| $a$ | = constant |
| $b$ | = constant |
| Ar | = Archimedes number |
| $C^*$ | = critical volatile evolution flux for gas phase ignition, g/s cm$^2$ |

$C_p$ = specific heat, J/g K
$D$ = diameter or diffusivity, cm(mm) or $cm^2/s$
$E$ = activation energy, Jg/mole
$G$ = mass velocity of gas, g/s cm
$g$ = acceleration of gravity, $cm/s^2$
$H$ = heat of combustion or bed height, J/g or cm
$h$ = heat transfer coefficient, $W/cm^2 K$
$k$ = chemical reaction rate constant or thermal conductivity, $s^{-1}$ or W/cm K
$k_d$ = mass transfer coefficient, cm/s
$k_0$ = frequency or pre-exponential factor, $s^{-1}$
$m_v$ = volatile evolution rate, g/s
Nu = Nusselt number
$n$ = power index, or integer value
Pr = Prandtl number
$p$ = probability
$Q$ = gas flow rate, $cm^3/s$
$Q_{l,cp}$ = conduction heat loss rate from flame front to particle, W
$Q_{l,ca}$ = conduction heat loss rate from flame front to ambient, W
$Q_{l,rp}$ = radiation heat loss rate from flame front to particle, W
$Q_{l,ra}$ = radiation heat loss rate from flame front to ambient, W
$q$ = gas flow rate at annulus inlet, $cm^3/s$
$R$ = particle radius, cm (mm)
Re = Reynolds number
$R_g$ = universal gas constant, Jg/mole K
$r$ = radial direction
Sc = Schmidt number
$T$ = temperature, K
$t$ = time, $s^{-1}$
$U$ = superficial gas velocity, cm/s
$U_i$ = terminal velocity of particle at unity voidage, cm/s
$u$ = interstitial gas velocity, cm/s
$V$ = coal volatiles, g
$v$ = particle velocity, cm/s
$Z$ = average travelling length of coal or carbon particle from bed surface into the annulus, cm
$z$ = Cartesian co-ordinate
$\rho$ = density, $g/cm^3$
$\sigma$ = Stefan–Boltzmann constant, $J/K^4 cm^2$
$\gamma$ = stoichiometric oxygen fuel weight ratio
$\nu$ = kinematic viscosity, $cm^2/s$
$\mu$ = dynamic viscosity, g/cm s
$\alpha$ = thermal diffusivity, $cm^2/s$
$\delta$ = boundary layer thickness, cm
$\varepsilon$ = emissivity or void fraction

*Subscripts*

$a$ = average value for annulus region
$b$ = bed
$c$ = cycle, or column
$cc$ = combined conduction and convection between bed and coal particle
$E$ = effective

*ex* = extinction
*F* = flame
*f* = average value for fountain region
*g* = gas phase
*H* = at bed surface
*i* = inert particle or *i*th component
*ig* = ignition
*m* = maximum condition or number of spherical shell
*mf* = minimum fluidizing condition
*ms* = minimum spouting condition
*o* = oxygen, or bed inlet orifice
*p* = particle
*R* = particle surface
*s* = average value for spout region
*t* = terminal free-falling condition
*th* = thermal
*v* = volatile
*vc* = volatile combustion
*z* = at any bed height
0 = initial condition
$\infty$ = mainstream

*Superscripts*

0 = motionless fluid
*t* = turbulent eddy
* = ultimate yield

## REFERENCES


Agarwal, P. K., Genetti, W. E. and Lee, Y. Y. (1984). 'Model for coal devolatilization in fluidized beds', *Fuel*, **63**, 1157

Agarwal, P. K. (1986). 'A single particle model for the evolution and combustion of coal volatiles', *Fuel*, **65**, 803

Agarwal, P. K. and Wildegger-Gaissmaier, A. E. (1987). 'Combustion of coal volatiles in gas fluidized beds', *Chem. Eng. Res. and Design*, **65**, 431

Anthony, D. B., Howard, J. B., Hottel, H. C. and Meissner, H. P. (1975). 'Rapid devolatilization of pulverized coal', Proceedings of 15th Symp. (Int.) Combustion, Combustion Institute, Pittsburgh, 1303

Atimtay, A. (1980). 'Combustion of volatile matter in fluidized beds', *in* Grace, I. R. and Matsen, J. M. (Eds.) *Fluidization*, Plenum Press, New York, 159

Baskakov, A. P. (1985). 'Heat transfer in fluidized beds: radiative heat transfer in fluidized beds', *in* Davidson, J. F. *et al.* (Eds.), *Fluidization, 2nd Edn.*, Academic Press, London, 465

Biswas, B. K. and Essenhigh, R. H. (1972). 'The problem of smoke formation and control', AIChE Symposium Series 126, **68**, 207

Bliek, A., van Poelje, W. M., van Swaaij, W. P. M. and van Beckum, F. P. H. (1985). 'Effects of intraparticle heat and mass transfer during devolatilization of a single coal particle', *AIChE J.*, **31**, 1666

Borghi, G., Sarofim, A. F. and Beer, J. M. (1977). 'A model of coal devolatilization and combustion in fluidized beds', 70th Annual AIChE Meeting, New York, USA

Broughton, J. and Howard, J. R. (1983). 'Combustion of coal in fluidized beds', *in* Howard, J. R. (Ed.) *Fluidized beds: combustion and applications*, Applied Science Publ; London, 37

De Kok, J. J., Nieuwesteeg, M. W. C. M. A. and van Swaaji, W. P. M. (1985). 'An experimental evaluation of the plume model for atmospheric fluidized-bed (coal) combustors', in Proceedings of the 8th International conference on fluidized bed combustion, Houston, Texas, **1**, 105

Devanathan, N. and Saxena, S. C. (1986). 'A model for the devolatilization of large coal particles', *Chem. Eng. Sci.*, **41**, 2442

Ekinci, E., Yalkin, G., Atakul, H. and Erdem-Senatalar, A. (1988). 'The combustion of volatiles from turkish coals in a fluidised bed', *J. Inst. Energy*, **61**, 189

Epstein, N. and Grace, J. R. (1984). 'Spouting of particulate solids', *in* Fayed, M. E. and Otten, L. (Eds.), *Handbook of powder science and technology*, Van Nostrand Reinhold, New York, 507

Essenhigh, R. H., Misra, M. K. and Shaw, D. W. (1989). 'Ignition of coal particles: a review', *Combustion and Flame*, **77**, 3

Grace, J. R. and Mathur, K. B. (1978). 'Height and structure of the fountain region above spouted beds', *Canadian J. Chem. Eng.*, **56**, 533

Kashiwagi, T. and Summerfield, M. (1973). 'Ignition and flame spreading over a solid fuel: non-similar theory for hot oxidizing boundary layer', Proceedings of 14th International Symposium on Combustion, Combustion Institute, Pittsburgh, USA, 1235
Kunii, D. and Levenspiel, O. (1969). *Fluidization engineering*, Wiley, New York, USA
La Nauze, R. D. (1982). 'Coal devolatilization in fluidized bed combustors', *Fuel*, **61**, 771
Lim, C. J. (1975). Ph.D. thesis, University of British Columbia, Vancouver, Canada
Lim, C. J., Watkinson, A. P., Khoe, G. K., Low, S., Epstein, N. and Grace, J. R. (1988). 'Spouted, fluidized and spout-fluid bed combustion of bituminous coals', *Fuel*, **67**, 1211
Park, D., Levenspiel, O. and Fitzgerald, T. J. (1981). 'Plume model for large particles in fluidized bed combustors', *Fuel*, **60**, 295
Piccinini, N., Grace, J. R. and Mathur, K. B. (1979). 'Vapour phase chemical reaction in spouted beds: verification of theory', *Chem. Eng. Sci.*, **34**, 1257
Pillai, K. K. (1981). 'The influence of coal type on devolatilization and combustion in fluidized beds', *J. Inst. Energy*, **54**, 142
Prins, W., Siemons, R. and van Swaaij, W. P. M. (1989). 'Devolatilization and ignition of coal particles in a two-dimensional fluidized bed', *Combustion and Flame*, **75**, 57
Ragland, K. W. and Yang, J. Y. (1985). 'Combustion of millimeter sized coal particles in convective flow', *Combustion and Flame*, **60**, 285
Ranz, W. E. and Marshall, Jr., W. R. (1952). 'Evaporation from drop—Parts I and II', *Chem. Eng. Progress*, **48**, 141 and 173
Richardson, J. F. and Zaki, W. N. (1954). 'Sedimentation and fluidization', *Trans. Inst. Chem. Engrs.*, **32**, 35
Rybak, W., Kordylewski, W. and Karcz, H. (1987). 'The kinetics of combustion of coal volatiles', *Fuel*, **66**, 139
Sangiovanni, J. J. (1978). 'A model for the nonsteady ignition and combustion of the fuel droplet', *in* Gould, R. F. (Ed.), *Advances in chemistry series*, No. 166, Am. Chem. Soc., Washington, DC, 27
Stubington, J. F. and Davidson, J. F. (1981). 'Gas-phase combustion in fluidized beds', *AIChE J.*, **27**, 59
Suuberg, E. M., Peters, W. A. and Howard, J. B. (1978). 'Product composition and kinetics of lignite pyrolysis', *Industrial Engineering Chemistry: Process Design and Development*, **17**, 37
Thomglimp, V. (1981). *Docteur-Ingenieur* thesis, Institut National Polytechnique, Toulouse, France
Tia, S. (1990). 'Combustion of lignite in spouted and spout-fluid beds', Ph.D. thesis, Energy Technology Division, Asian Institute of Technology, Bangkok, Thailand
Tia, S., Bhattacharya, S. C. and Wibulswas, P. (1991). 'Combustion behaviour of coal and carbon in spouted and spout-fluid beds', *Int. J. Energy Res.*, **15**, 249–255
Wu, S. W. M., Lim, C. J. and Epstein, N. (1987). 'Hydrodynamics of spouted beds at elevated temperatures', *Chem. Eng. Communication*, **62**, 251
Xavier, A. M. and Davidson, J. F. (1985). 'Heat transfer in fluidized beds: convective heat transfer in fluidized beds', *in* Davidson, J. F. *et al.* (Eds.) *Fluidization, 2nd Edn.*, Academic Press, London, 437
Zhao, J., Lim, C. J. and Grace, J. R. (1987*a*). 'Flow regimes and combustion behavior in coal-burning spouted and spout-fluid bed', *Chem. Eng. Sci.*, **42**, 2865
Zhao, J., Lim, C. J. and Grace, J. R. (1987*b*). 'Coal burnout times in spouted and spout-fluid beds', *Chem. Eng. Res. and Design*, **65**, 426