ESTIMATION OF BURNING RATES IN SOLID WASTE COMBUSTION FURNACE

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ABSTRACT

For a mass burning furnace, one of the key areas in design is the determination of burning intensity or area burning rate. An independent estimation of this factor would require the knowledge of chemical kinetics of waste combustion, supply and distribution of combustion air, fluid dynamical mixing of combustion air with the burning bed, and physical and chemical characteristics of the refuse as well as operational factors which might affect the combustion processes in the furnaces.

In this paper an attempt was made to estimate the refuse burning rates in the furnaces by calculating the oxygen distribution in the combustion chamber. A model based on the fundamental principles of conservation of mass, energy and momentum balance is established for this purpose. Gas compositions in the fuel bed are assumed to be proportional to the amount of oxygen present in the bed. Parameters like bed height, porosity of the bed, and underfire air rates are found to be important influencing factors in controlling the burning rates.

INTRODUCTION

Solid waste disposal by means of incineration will become one of the dominant methods of waste disposal in Taiwan. Incineration is gaining world-wide acceptance among industrialized nations as a practical and feasible technical solution to the mounting problem of waste generation. Despite the fact that the technology of incineration has been practiced for nearly a century, the recent advancements of fluid and thermal sciences have not been playing significant roles in the design of waste combustion facilities. The incineration technology evolves more as an art than a science through years of practices and experience. Even though this was the way this technology was evolved, the growing concerns of pollutant emissions and ever more stringent environmental regulations are pressing for more fundamental understanding of the processes of solid waste combustion. A direct benefit of this would be a better designed facility for cleaner and more efficient solid waste incineration.

Combustion of a heterogeneous mixture of refuse materials, in broad terms, involves drying and heating of the solids to reaction temperature, pyrolysis and release of combustible volatiles, cracking and oxidation of chars or carbon particles, and gas phase combustion reactions. Although theoretical treatment of the complex processes of refuse combustion is, so far, incomplete, the empirical art of combustion engineering has become so advanced that direct combustion of solid waste is accepted as a well proven element of waste management system.

In principle, control and manipulation of solid waste combustion processes involves the fundamental considerations of the 3 "Ts" of combustion: time, temperature, and turbulence or mixing. In practice, this entails proper design of the furnace grate to promote the mixing of solids and combustion air, selection of combustion air supply and distribution systems to provide adequate proportions of combustion air and maintaining the temperatures needed for efficient combustion, and properly sizing the combustion chamber and fuel grate to allow enough residence time for complete combustion.

Issues of combustion air manipulation include percentage of excess air, distribution of grate air and proper division of underfire and overfire air, and control of combustion air temperature. Excess air influences the combustion flame temperature as well as the combustion efficiency and airborne emissions. Distribution of combustion air are known to have significant influences on system performance. Improper combustion air distribution may promote reducing conditions in the combustion chamber which have the detrimental consequences of increasing metal wastage and lowering the ash fusion temperatures. The distribution combustion air also influences the emissions of CO, toxic and nontoxic organics, particulates, and NO_x.

Traditionally, sizing of the combustion chamber is done empirically by selecting the rates of heat releases. Selection of grate heat release and volumetric heat release rates could have a significant influence on combustion gas residence time and turbulent mixing of gases. Generally, various heat release rate selections reflect the differences in characteristics of the grate design, methods of combustion air manipulation and the specific nature of the waste material.

Fundamentally, the heat release rates could be determined from the basics of combustion mechanism and the thermo-fluid characteristics of the combustion system. So far, attempts with this approach (Kanury and Blackshear, 1970; Daw and Krishnan, 1982) are limited to relatively simple fuels like coal and wood. Efforts to develop fundamental design equations were made by Essenhigh and Kuo (1969). A central issue identified in their work was the need for an independent evaluation of the combustion intensity, or alternatively, the grate burning rates from the first principles. The work reported in this paper is an attempt to establish a model to estimate the solid waste burning intensity in a combustion chamber from the basic principles of heat and mass transfer processes.

RATES OF HEAT RELEASE

The rates of heat release determine the shape and size of the combustion chamber, and many of the parameters associated with combustion residence time and mixing of the gaseous products. The throughput of the combustion system can be related to the volumetric heat release rate by a design equation developed by Essenhigh (1968)

$$B_G = C_1 \left(\frac{I_v}{Q_{hhv}} \right)^{1/3} B_C^{2/3} \tag{1}$$

where B_G is the grate burning rate, an area burning rate (e.g. kg/m^{2}hr), B_C is the furnace combustion capacity (e.g. kg/hr). I_v is the average volumetric heat release rate (e.g. kcal/m^{3}hr) and Q_{hhv} is the higher heating value of the wastes. C_1 is a shape factor. This equation is nothing more than a dimensional relation of the grate loading, B_G, and the combustion intensity I_v. Sizing of the combustion chamber requires the knowledge of both I_v and B_G. An independent evaluation of either one of these quantity is needed.

Historically, the designers select either the combustion heat release intensity or the grate loading based on experience. For example, empirically, the grate burning rates may relate to the combustion rates by George and Williamson (1969),

$$B_G = K_1 \log B_C + K_2 \tag{2}$$

where K_1 and K_2 are empirical constants with values dependent on the waste type (or alternatively, the moisture content or the heating value). On the other hand, if the furnace volume is approximated as the product of multiplying grate area, A_G, with a characteristic furnace interior height H, Equation (1) can be reduced to

$$B_G = C^2 B_C \tag{3}$$

where C is essentially the inverse of the furnace height. This gives a linear relation of grate area burning rate and the system capacity, instead of a logarithmic relation as indicated by Equation (2). Figures 1 and 2 are collections of manufacturer quoted grate burning rates varying with the system capacity. These data indicate that for smaller units (typically the fixed grate multi-chamber systems with capacities under 1000 kg/hr) where mixing of the solids and the gases is not optimum, the area burning rates are much lower than those obtainable in large central units which usually have better grate design and superior air distribution systems. Analyzing the data in Figures 1 and 2 reveals that, basically, there are two types of grate burning rate correlations. One type of furnace designs are based on essentially constant area burning rates irrespective of the system capacity, and varying only with the waste type. The other designs are based on grate burning rates which are increasing with the system capacity. Due to better mixing and more effective stoking, large capacity furnaces are sized with much higher area burning rates.

Figure 3 is the plot of the same sets of manufacturer's data to reveal the furnace shape factors of different designs. The group of data with shape facor close to 4 are those systems designed for burning high heat value polymers or rubber/plastic wastes. These systems are typically cylindrical in shape and have relatively high area burning rates. Strong overfire jets are used in these units.

Alternatively, assuming that the combustion processes within the furnace occurs in two different zones, the solid bed and the overbed regions, Essenhigh and Kuo (1969) developed an equation to calculate the area burning rates

$$B_G = \frac{0.175 f_{rcs} G_{air}}{(1 - V)(1 - M - A)} \tag{4}$$

where G_{air} is mass flux of air through the bed, and f_{rcs} is known as the relative carbon saturation factor. A and M

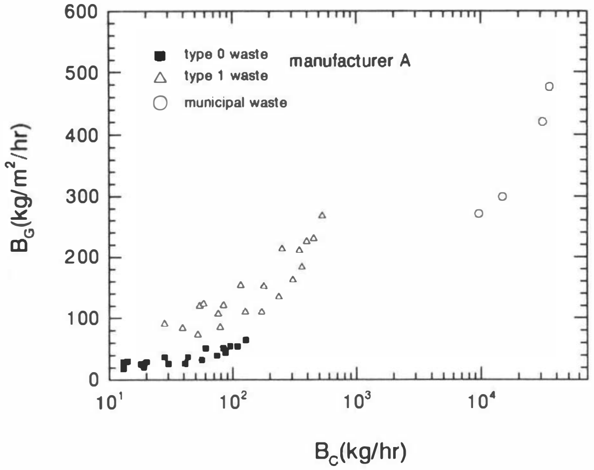

FIG. 1 MANUFACTURER QUOTED GRATE BURNING RATES

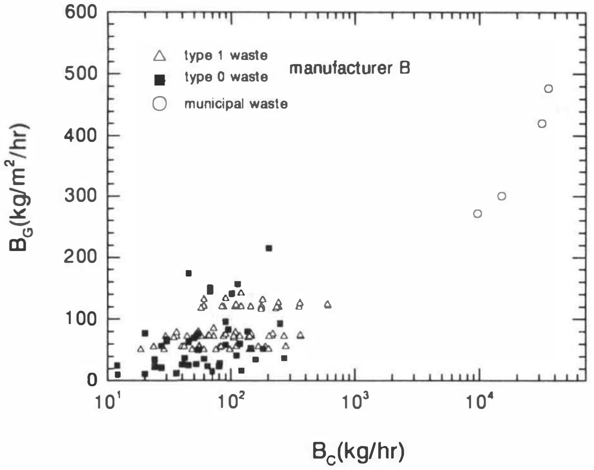

FIG. 2 MANUFACTURER QUOTED GRATE BURNING RATES

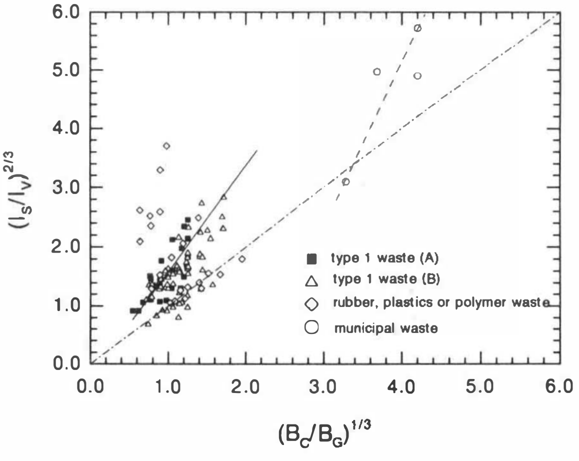

FIG. 3 FURNACE SHAPE FACTOR $C = V^{2/3}/A = (l_S/l_V)^{2/3}(B_C/B_G)^{-1/3}$

stand for the moisture and ash fraction of the wastes, and V is the volatile fraction of the combustible wastes. If the bed is deep enough allowing maximum mass saturation of oxygen by carbon, meaning the chars in the bed are completely converted to CO, then $f_{rcs} = 1$. On the other hand, if the bed is shallow, all the carbons are converted to CO_2, then f_{rcs} is 0.5. In general, f_{rcs} is related to the stoichiometrics of the bed reaction

$$f_{rcs} = \frac{1}{2 - r_{co}} \tag{5}$$

where r_{co} is the fraction of carbon that has been converted to CO. If the grate air supply is equal to the stoichiometric air requirement for complete combustion, the area burning rates would be inversely proportional to the higher heating value of the refuse.

$$B_G = \frac{740}{Q_{hhv}} G_{air} \tag{6}$$

The units of Q_{hhv} are Kcal/kg.

While these design equations are useful when empirical information regarding f_{rcs} and G_{air} are readily available, it would be desirable if the empirical aspects of the equations can be further supplemented with more fundamental considerations of the chemical and physical mechanisms of the burning processes.

BURNING RATE ESTIMATION

Burning of solid wastes in a furnace are often modeled as a solid fuel bed where the primary air enters the base of the grate, passing through a solid bed zone and the secondary air (or the overfire air) feeds over the top of the solid bed. In reality, the combustion processes could be considered as a mixture of burning, gasification and pyrolysis with no clear boundary between zones. In general, it can be assumed that the actual combustion occurs on the material surfaces. The complete processes take place through a sequence of events like, heating and drying of the refuse, transfer of oxygen from the bulk gas through a stagnant layer of gas adjacent to the surfaces of the solids, absorption and reaction of the oxygen with the solids, and diffusion of the reaction products. Any one of these events may be the principal rate controlling factor. The rates of reaction could either be kinetically controlled which depend on the Arrhenius term, exp $(-E/RT)$, or diffusionally controlled where the reaction rates are dependent on how fast the oxygen can diffuse to the solid surfaces.

When the reactions are kinetically controlled, the rate of reaction can be expressed as

$$\frac{dC_{O_2}}{dt} = -a_p X_{O_2}^n G_{st,O_2} \exp\left(-\frac{E}{RT}\right) \tag{7}$$

a_p is the surface area of the refuse per unit volume and n is the order of reaction. G_{st,O_2} is the mass flux of oxygen.

On the other hand, for a diffusion controlled reaction, the rate of reaction is given by

$$\frac{dC_{\mathrm{O}_2}}{dt} = -a_p G_{sd,\mathrm{O}_2} X_{\mathrm{O}_2}^n \left(\frac{1}{a_r}\right) \ln(1 - a_r X_{\mathrm{O}_2}) \tag{8}$$

where X_{O_2} is the mole fraction of oxygen in the bulk gas leaving the bed and a_r is a numerical factor arising principally because of the change of gaseous moles in reaction. Similarly, G_{sd,O_2} is the diffusional mass flux of oxygen.

From mass balance consideration, the gasification rates of the solid bed are related to the oxygen consumption rates as the supplied air filtering through the fuel bed. On a molar basis the solid bed gasification rate is

$$\frac{dN_W}{dt} = \frac{N_{air}}{g}\left(1 - a_1 \frac{x_{\mathrm{O}_2,b}}{1 - a_r X_{\mathrm{O}_2,b}}\right) \tag{9}$$

where $a_1 = 4.76 - a_r$ and g is the stoichiometric air/fuel mole ratio. $X_{\mathrm{O}_2,b}$ is the mole fraction of oxygen in the bulk gas leaving the solid bed zone. N_w and N_{air} are moles of refuse and air supply respectively.

Comparison of Eq. (9) and Eq. (4) would give

$$f_{rcs} = f_r \frac{5.72}{g}\left(\frac{\overline{M}_W}{\overline{M}_{air}}\right)(1 - V)(1 - M - A) \tag{10}$$

where

$$f_r = 1 - a_1 \frac{X_{\mathrm{O}_2,b}}{1 - a_r X_{\mathrm{O}_2,b}} \tag{11}$$

$\overline{M}_W$ and $\overline{M}_{air}$ are molecular weights of the waste and the air respectively. Equation (9) is equivalent to Eq. (4) except for an oxygen consumption factor f_r. Basically, f_r is dependent upon the rates of reaction and air supplied to the fuel bed.

BED OXYGEN DISTRIBUTION

Assuming the oxygen concentration in the solid bed region is varying one-dimensionally along the depth, y, of the solid waste bed, for diffusion controlled reaction, the oxygen distribution in the bed region can be expressed as

$$\frac{dX_{\mathrm{O}_2}}{dy^*} = k_r \frac{X_{\mathrm{O}_2}^n}{a_r}(1 - a_r X_{\mathrm{O}_2}) \ln(1 - a_r X_{\mathrm{O}_2}) \tag{12}$$

and

$$\frac{dX_{\mathrm{O}_2}}{dy^*} = -k_r X_{\mathrm{O}_2}^{n+1} \qquad \text{if} \quad a_r = 0 \tag{13}$$

$y^* = y/L$, where L is the bed depth and k_r is a dimensionless reaction rate factor.

$$k_r = a_p L \frac{j_D}{Sc^{2/3}} \tag{14}$$

where j_D is the Chilton-Colburn j factor for oxygen transfer and Sc is the Schmidt number. While there is no well established correlation for j_D applicable to solid bed combustion, an estimation of j_D based on the mass transfer correlation for packed beds indicates that j_D is around 0.1. For a bed of spherical solids of diameter d, if $L/d = 10$ and the void fraction is 0.4, then $a_p L = 36$, with Sc on the order of 1, k_r is 3.6. Thus K_r is likely to be on the order of 10.

For reactions controlled by chemical kinetics, k_r is temperature dependent.

$$\frac{dx_{\mathrm{O}_2}}{dy^*} = K_t x_{\mathrm{O}_2}^n (1 - a_r x_{\mathrm{O}_2}) \exp\left(-\frac{E}{RT}\right) \tag{15}$$

where $k_t = (a_p L)(V_{s,\mathrm{O}_2}/V_g) \cdot V_{s,\mathrm{O}_2} = G_{st,\mathrm{O}_2}/\rho_{\mathrm{O}_2}$ is a refuse surface oxygen transfer velocity and V_g is the average gas velocity through the refuse bed. In this case, K_r is defined as

$$K_r = K_t \exp\left(-\frac{E}{RT}\right) \tag{16}$$

If E/R is around 10^4 K, for a bed of $a_p L$ on the order of 10 and V_g of 1 m/sec, then V_{s,O_2} would be on the order of 10^4, if k_r is to be comparable with the diffusionally controlled rates at 1000 K. Beyond 1000 K, the reactions are expected to be controlled by diffusion processes.

The one-dimensional bed temperature distribution would be

$$\frac{dT^*}{dy^*} = R_f(q_r^* - n_w R_m h_{fg}^*) \tag{17}$$

and

$$q_r^* = \frac{q_r}{C_{pg} T_o}; \qquad h_{fg}^* = \frac{h_{fg}}{C_{pg} T_o}$$

q^*r is the net heat transfer to the strip of bed under consideration. In this work only the chemical heat release is considered. In reality, other heat transfer mechanisms, like radiative and conductive heat transfers may be significant and should be included in the analysis. R_f is the mass ratio of the rate of waste gasification to the product gas flow. n_w is moles of moisture per mole of waste in the combustion products. R_m is the molecular weight ratio of water and the refuse.

RESULTS AND DISCUSSION

In the fuel bed three possible forms of carbon-oxygen reactions are considered, namely,

$$C + 0.5(1 - r_c)O_2 = r_c C + (1 - r_c)CO \qquad (18a)$$

$$C + (1 - 0.5r_{CO})O_2 = r_{CO}CO + (1 - r_{CO})CO_2 \qquad (18b)$$

$$C + O_2 = CO_2 \qquad (18c)$$

The occurrences of these reactions are assumed to be dependent upon the amount of oxygen present in the fuel bed. The values of r_c and r_{CO} are linearly proportional to the available oxygen in the bed.

The water-gas shift is known to have a significant effect on off-gas composition. The following forms of water-gas reaction are assumed,

$$C + r_h H_2O + 0.5(1 - r_c)O_2$$

$$= (r_c - r_h)C + (1 - r_c + r_h)CO = r_h H_2 \qquad (19a)$$

$$C + r_h H_2O + (1 - 0.5r_{CO})O_2$$

$$= (r_{CO} - r_h)CO + (1 - r_{CO} + r_h)CO_2 + r_h H_2 \qquad (19b)$$

r_h is the fraction of H_2O involved in the water-gas reaction. Here r_h is assumed to be linearly proportional to r_c or r_{CO}. For example, if the chemical formula of the refuse is expressed as $C_aH_bO_c \cdot f(H_2O)$, then

$$r_h = \frac{r_{CO}\left(\frac{b}{2} + f\right)}{\left(a + \frac{b}{2} + f\right)} \qquad (20)$$

Based on this simplified reaction model, gas composition distribution in the fuel bed can be calculated.

Assuming the waste is given by the chemical formula of cellulose, $C_6H_{10}O_5$, Figure 4 shows the variations of gas composition and temperature through the bed at different K_r, without consideration of water-gas shift effect, based on diffusion controlled reaction and assuming $n = 1$. There is a marked difference in temperature distribution when K_r changes from 10 to 100. The oxygen depletes much faster through the bed when the reaction rates are high. This is why temperature peaks near the bottom of the bed. Figure 5 shows the effect of water-gas shift on the bed gas composition and temperature distribution. The water-gas shift reaction has a significant effect on the bed temperature distribution and the CO distribution.

Thring (1952) calculated the fuel bed gas composition based on a three-zone reaction model and diffusionally controlled combustion with $n = 0$. When the reaction rates are high, this model gives a trend of gas composition variation along the depth of the fuel bed similar to the work of Thring (1952). A distinct different feature of this model is

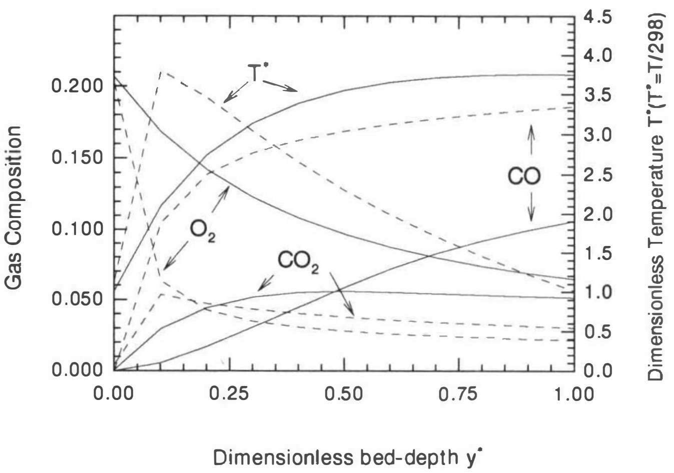

FIG. 4 GAS COMPOSITION AND TEMPERATURE DISTRIBUTION IN THE FUEL BED (—— k_r = 10, ------ k_r = 100)

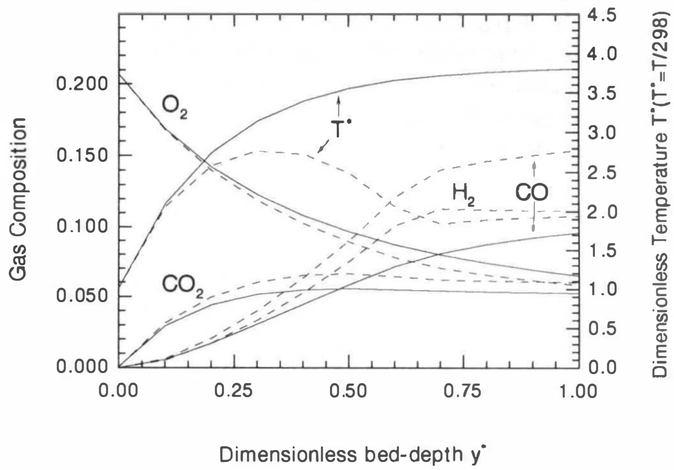

FIG. 5 EFFECT OF WATER-GAS SHIFT ON BED GAS COMPOSITION (k_r = 10, —— WITHOUT WATER-GAS SHIFT, ------ WITH WATER-GAS SHIFT)

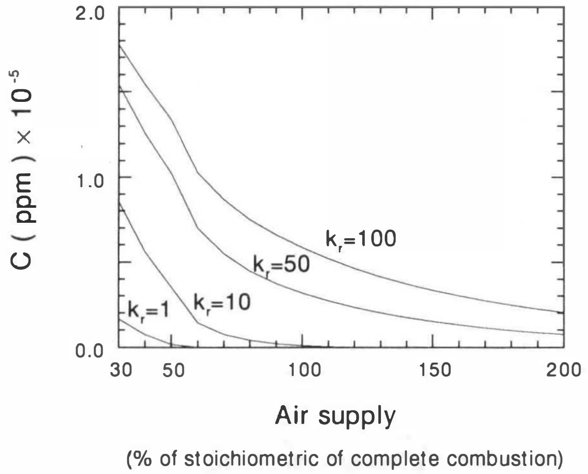

FIG. 6 SOOT FORMATION AT VARIOUS AMOUNTS OF AIR SUPPLY

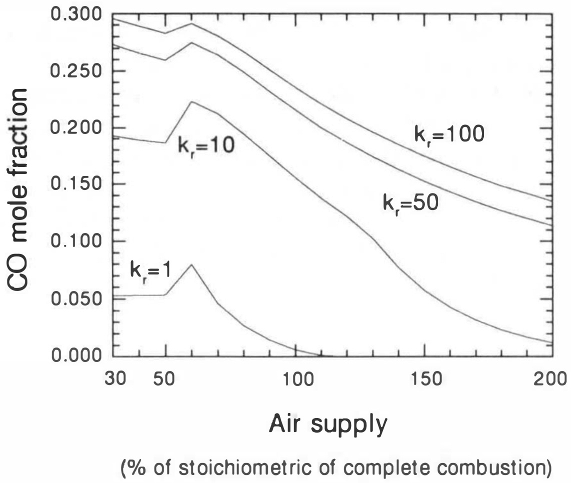

FIG. 7 CO FORMATION AT VARIOUS AMOUNTS OF AIR SUPPLY

the allowance for soot formation. The fuel bed chemistry is strongly dependent on the rates of oxygen depletion through the bed. When the reaction rates are high, the oxygen is quickly depleted in the initial stage of penetration, leaving most of the bed in an oxygen starving condition, promoting the formation of soot and carbon monoxide. On the other hand, at low reaction rates, when the air supply is sufficient, the bed consists of, predominantly, the products of complete combustion.

Figures 6 and 7 are the predicted formations of soot and carbon monoxide at various amounts of air supply for different reaction rates. The finite reaction rates pose a limit on the amounts of supplied oxygen that can be consumed through the bed. A large K_r would mean either a thick bed or a high reaction rate. The oxygen consumption factors f_r calculated according to the one-dimensional model at different reaction constants for various air supplies are shown in Figure 8. When the air supply is very much below the complete combustion requirement, the reactions in the bed favor the formation of soot and carbon monoxide. This would further reduce the oxygen requirement of the bed gasification reaction. Thus f_r is the combined results of diffusion limitation and reaction stoichiometrics.

In practice, underfire air supply is in the range of 0.02 to 0.5 Nm^3/m^2-sec, which is equivalent to 3.2 to 80 moles/m^2-hr. Assuming the fuel is cellulose, with stoichiometric air requirement of 28 moles of air/mole fuel, the area gasification rate predicted by the model is shown in Figure 9. The range of grate loadings shown in Figure 9 are consistent with those quoted by the incinerator manufacturers, if K_r is on the order of 10. Apparently, the capability of low air supply in sustaining a relatively high grate loading, comparing to what could be achieved with stoichiometric reaction, is due to the incomplete burnout of the solid bed which requires much less air than stoichiometric burning. This creates additional burdens on the

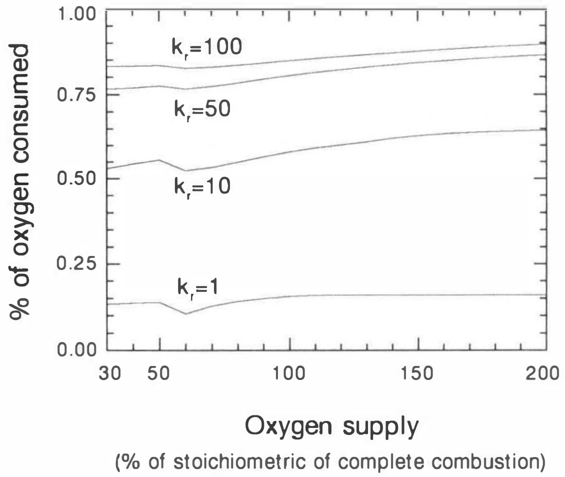

FIG. 8 FRACTION OF SUPPLIED OXYGEN CONSUMED IN REACTION

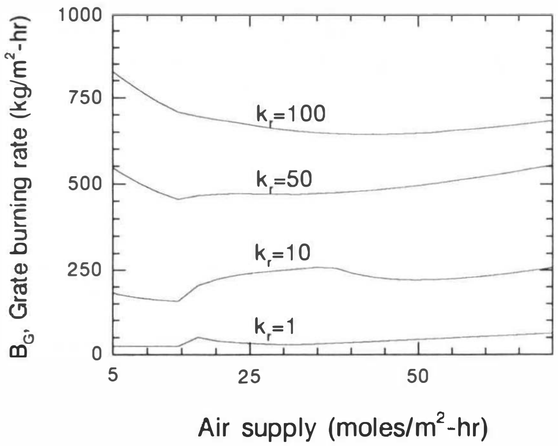

FIG. 9 GRATE AREA BURNING RATES OF CELLULOSE COMBUSTION

overfire zone and the secondary combustion chamber to complete the rest of the combustion processes.

CONCLUSIONS

The industrial selections of grate area burning rate (or grate loading) cover a wide range depending upon the system capacity and type of grate design. Small incinerators with fixed grates tend to have much smaller grate loading capability than those obtainable in large central facilities. Presumably this is due to the poor capability of the fixed-grate system in promoting the mixing of combustion air with the refuse bed. There is no apparent correlation among different manufacturers in the design selections of grate loading. The general trend is an increasing of grate loading with the system capacity. This trend generally reflects the improvement in grate design and better combustion air management of the large capacity systems.

The bed chemistry model adopted in this work is able to generate different profiles of bed gas composition and reaction temperature by choosing the order of reaction n and reaction rate factor K_r. Parametric studies of the model indicate that the reaction rate factors should be in the range of 1 to 10 for first order reaction ($n = 1$), if predictions of burning rates are to be consistent with what have been observed in actual burning test.

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