IMPORTANCE OF FRAGMENTATION ON THE STEADY STATE COMBUSTION OF WOOD CHAR IN A BUBBLING FLUIDIZED BED

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Abstract. *A simple mathematical model for the analysis of the steady state behavior of a bubbling fluidized bed burner is presented, with the main intention of evaluating the importance of the primary fragmentation of fuel particles on the performance of this type of burners. This model has pedagogical advantages because of its simplicity and easiness of application to the analysis of realistic situations. The model is based upon the classical models used for the study of batch combustion processes in fluidized bed reactors. Experimental data from studies of fluidized bed combustion of portuguese vegetable chars are used to support the analysis of the performance of a 1 m diameter fluidized bed combustor.*

Keywords*: biomass, combustion, fluidized bed, fragmentation, mathematical model*

1. INTRODUCTION

Wood is the oldest solar energy conveyor known by mankind (Obernberger, 1998; Strehler, 2000; McKendry, 2002; Speight, 2008). More than 15 % of the primary energy consumed by mankind, comes from the combustion of wood or ligneous residues (Bhattacharya, 1998), but with low combustion efficiency. The interest on bioenergy is raising and European Union targets are rather ambitious (Grassi, 1999).

In a country like Portugal fast growing plant species dedicated to combustion can lead to an exploitation rate of about 80 t/ha/year. Alternatively, the forest can be used for other ends and the cleaning waste of about 2 to 3 t/ha/year can be used as fuel (Nuñez-Regueira *et al.*, 1997; Nuñez-Regueira *et al.*, 1999; Nuñez-Regueira *et al.*, 2001), but it is necessary to avoid soil depletion of minerals and nutrients (EEA, 2006). The main attractiveness of using biomass as fuel is the achievement of a closed carbon cycle (El Bassam, 1998; Klass, 1998).

In developed countries the production costs of electricity and thermal energy by means of biomass are not competitive when compared with fossil fuels, the transportation can represent up to 70 % of the biomass cost (McKendry, 2002). Small power plants placed close to the biomass sources will attractive, it will be cheaper to transport electricity than bulk biomass. Because the wood and forestry wastes come from a restricted area, the physical, chemical and morphological characteristics of the fuel will be relatively narrow, which is advantageous in terms of fuel preparation, regulation and control of the combustion process, increasing the quality of combustion and conversion efficiency of the plant. In developing countries biomass is frequently the most important source of primary energy supplying 35 % of their energy needs (Bhattacharya, 1998; Dermibas, 2004). In developed countries the increase of the efficiency of biomass utilization is the main target. In the USA the efficiency of a power plant stays around 20 - 25 %, while forecasts for the maximum efficiency of biomass gasification plants can go up to 43 % (Descamps et al., 2008).

In fluidized bed combustion, fuel particles below 25 mm burn inside a bed of inert particles of 0.5 to 1 mm diameter (Oka, 2004). The solid fuel sent to the bed is rapidly heated up to the bed operating temperature. This provokes a strong devolatilization of the fuel particles and the majority of the volatiles burn above the free surface of the bubbling bed, while the charcoal solid core burns inside the bed (Avedesian and Davidson, 1973; Ross and Davidson, 1981; Pinho and Guedes de Carvalho, 1984; Hayhurst, 1991; Guedes de Carvalho *et al.*, 1991; Mota *et al.*, 1994; Hayhurst and Parmar, 1998; Ribeiro and Pinho, 2004; Ribeiro and Pinho, 2007). With low bed temperature combustion (800 to 900 ºC), low levels of NO_x emissions are obtained. They have a relatively wide operating ratio and burn low calorific value fuels as biomass with high combustion efficiency. This makes the fluidized bed combustion a reference in the combustion of coal and biomass (Oka, 2004).

2. STEADY STATE COMBUSTION OF COKE OR CHAR PARTICLES IN A FLUIDIZED BED

In continuously working furnaces, the fuel supply is in steady state regime. As the fuel concentration inside the bubbling bed is very small, many authors consider that the individual behaviour of a burning particle is not affected through interaction with other fuel particles and that the steady state combustion can be considered as a combustion of a sequence of batches of solid particles. There are more or less elaborated models to account for this steady state combustion process, but such models are very complex and present pedagogical limitations (Yang, 2003; Oka, 2004).

A simple mathematical model for the steady state combustion is proposed as an evolution of a model for the combustion of batches of coke or char particles in bubbling fluidized bed. Experimental data obtained in laboratory studies can thus be applied to the steady state combustion, and the importance of fragmentation can be easily quantified.

The fluidized bed reactor is considered an isothermal reactor and the burning particles are at bed temperature. For coke particles burning in a bubbling fluidized bed at 930 ºC, Roscoe *et al*. (1980) verified that particles could be burning around 130 to 160 ºC above the bed temperature. This was obtained from visual analysis of the particles floating at the bed surface; it may not be representative of the overall behaviour of the particles under combustion. Khraisha (2005) considers that particles burn at bed temperature, while Komatina *et al*. (2006) consider that the type of coal, the batch size and the O_2 concentration all are important to define the temperature of the particles.

A solid particle in a combustion process takes an elemental time dt to suffer an elemental size reduction d(d),

$$-\frac{\rho_c}{M_c}\frac{\pi d^2}{2}\frac{\mathrm{d}}{\mathrm{d}t}(d) = 2\,\pi\,d^{\,2} K\,C_p \tag{1}$$

The overall resistance to combustion is given by $\frac{1}{K} = \frac{\varphi\,d}{Sh\,D_g} + \frac{2}{k_c}$, (Avedesian and Davidson 1973; Ross and Davidson, 1981; Pinho and Guedes de Carvalho, 1984; Guedes de Carvalho *et al*., 1991; Mota, *et al*. ,1994). This particle belongs to a flow of particles continuously introduced into the bed, which compete among them for the available oxygen; to account for this competition a multiplying factor ($0 < \eta \le 1$) is used (Annamalai, 1995),

$$\mathrm{d}t = -\frac{\rho_c}{4\,M_c \eta\,C_p}\left[\frac{\varphi\,d}{Sh\,D_g} + \frac{2}{k_c}\right]\mathrm{d}(d) \tag{2}$$

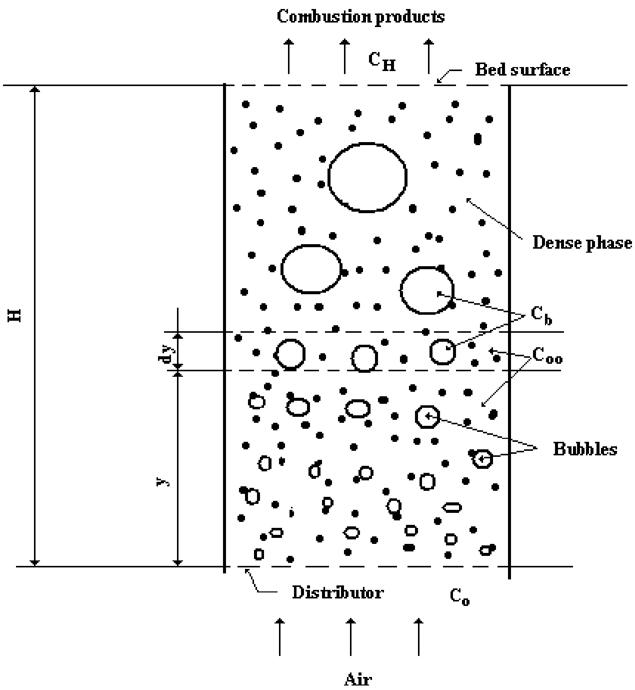

Figure 1. Schematic representation of the fluidized bed reactor

Figure 1 shows the O_2 balance inside the fluidized bed. Oxygen enters the bed through the fluidizing air and is distributed into the dense and the bubble phase (Davidson and Harrison, 1963). In the dense phase the O_2 concentration is uniform and equal to C_p. For the bubble phase, the oxygen concentration is given by

$$C_b = C_p + \left(C_o - C_p\right)\exp\left(-\frac{X}{H}y\right) \tag{3}$$

where X is the bubble cross flow factor (Hovmand *et al*., 1971). The dense phase is a uniformly mixed flow reactor, while the bubble phase is a plug flow reactor (Dobre and Marçano, 2007). The particles burn in the dense phase. Making an oxygen balance in a slice of fluidized bed with a thickness of dy, as the O_2 concentration in the dense phase is constant, it is only necessary to carry out the oxygen balance of the bubble phase,

$$\begin{aligned}&\left(U - U_{mf}\right) A_t \left(C_o - C_p\right)\left[exp\left(-\frac{X}{H} y\right) - exp\left(-\frac{X}{H}\left(y + \mathrm{d}y\right)\right)\right] = \\ &= -\dot{N}_p \frac{\rho_c}{2\, M_c} \pi\, d^{\,2} \mathrm{d}(d)\end{aligned} \tag{4}$$

where $\dot{N}_p$ is the number of particles introduced in the bed per unit of time, assuming that the solid mass flow is composed of particles of uniform size and that their number stays constant during the combustion process,

$$\frac{6\, \dot{m}_{cy}}{\rho_c \pi\, d_{iy}^{\,3}} = \frac{6\, \dot{m}_c}{\rho_c \pi\, d_i^{\,3}} = \dot{N}_p \tag{5}$$

and d_{ic} is the diameter of the fuel particles at the inlet of the bed slice under analysis.

The following system of differential equations,

$$\mathrm{d}t = -\frac{\rho_c}{4\, M_c \eta\, C_p}\left[\frac{\varphi\, d}{Sh\, D_g} + \frac{2}{k_c}\right]\mathrm{d}(d) \tag{6}$$

$$\begin{aligned}&\left(U - U_{mf}\right) A_t \left(C_o - C_p\right)\left[exp\left(-\frac{X}{H} y\right) - exp\left(-\frac{X}{H}\left(y + \mathrm{d}y\right)\right)\right] = \\ &= -\dot{N}_p \frac{\rho_c}{2\, M_c} \pi\, d^{\,2} \mathrm{d}(d)\end{aligned} \tag{7}$$

has no analytical solution. To follow an analytical path, an approximation can be adopted,

$$\exp\left[-\frac{X}{H}\left(y + \mathrm{d}y\right)\right] - \exp\left[-\frac{X}{H} y\right] \approx -\left[\exp\left(-\frac{X}{H} y\right)\frac{X}{H}\mathrm{d}y\right] \tag{8}$$

The system of equations now becomes,

$$\mathrm{d}t = -\frac{\rho_c}{4\, M_c \eta\, C_p}\left[\frac{\varphi\, d}{Sh\, D_g} + \frac{2}{k_c}\right]\mathrm{d}(d) \tag{9}$$

$$\mathrm{d}y = -\frac{\dot{N}_p \dfrac{\rho_c}{2\, M_c} \pi\, d^{\,2} \mathrm{d}(d)}{\left(U - U_{mf}\right) A_t \left(C_o - C_p\right) \exp\left(-\dfrac{X}{H} y\right)\dfrac{X}{H}} \tag{10}$$

Integrating Eq. (10) for the burned mass fraction f gives the following result,

$$y_f = -\frac{H}{X} ln\left[1 - \frac{f\, \dot{m}_c}{M_c\left(U - U_{mf}\right) A_t \left(C_o - C_p\right)}\right] \tag{11}$$

In the integration it is assumed that both the oxygen concentration at the bed inlet C_o, as well as in the dense phase C_p, do not change with time, according to the operational conditions for steady state combustion.

Integrating now Eq. (9) it is obtained that,

$$t_f = \frac{\rho_c}{4\, M_c \eta\, C_p}\left\{\frac{\varphi\left[1 - \left(1 - f\right)^{2/3}\right] d_i^{\,2}}{2\, Sh\, D_g} + \frac{2\left[1 - \left(1 - f\right)^{1/3}\right] d_i}{k_c}\right\} \tag{12}$$

This equation gives the combustion time of mass fraction f of a particle belonging to a mass flow of carbon particles being introduced inside the fluidized bed at the rate of $\dot{m}_c$. If the combustion of a mass fraction f takes place inside the bubbling bed, then Eq. (11) with $y_f = H$ is solved in order to get the adequate value of C_p,

$$C_p = C_o - \frac{f\ \dot{m}_c}{M_c\left(U - U_{mf}\right)A_t\left[1 - \exp\left(-X\right)\right]} \tag{13}$$

Then Eq. (13) is introduced into Eq. (12) to obtain,

$$t_f = \frac{\rho_c\left\{\dfrac{\varphi\left[1-\left(1-f\right)^{2/3}\right]d_i^{\,2}}{2\ Sh\ D_g} + \dfrac{2\left[1-\left(1-f\right)^{1/3}d_i\right]}{k_c}\right\}}{4\ M_c\eta\left\{C_o - \dfrac{f\ \dot{m}_c}{M_c\left(U - U_{mf}\right)A_t\left[1-\exp\left(-X\right)\right]}\right\}} \tag{14}$$

Looking at Eq. (14), for the steady state combustion regime, the bed hydrodynamics is superimposed on the kinetic and on the diffusion terms. The mass of a char particle, the mass of a batch of particles or the mass flow rate of char being introduced inside the bed will always represent the equivalent carbon value. Thus the density of the particles is corrected, by taking into account the corresponding mass fraction of carbon content obtained from the proximate analysis of the chars and then particle diameters are kept unchanged.

3. CARBON INVENTORY OF THE BED AND THE AVERAGE PARTICLE DIAMETER

When a given mass flow rate of carbon is introduced into the bed, it is assumed that all the particles have the same initial diameter d_i. Thus, all the particles existing inside the bed with the same generic diameter d will all have the same life time. The introduction of a mass flow rate of carbon into the bed means the introduction of a given number of particles per second $\dot{N}_c$, all with the same initial diameter d_i,

$$\dot{m}_c = \dot{N}_c\rho_c\frac{\pi\ d_i^{\,3}}{6} \tag{15}$$

and during the life time of these particles t_{stc}, obtained from Eq. (14) with $f = 1$, many other particles will be entering the bed at the rate of $\dot{N}_c$. The total number of particles existing inside the bed during such time will be of,

$$N_{ct} = \dot{N}_c t_{stc} \tag{16}$$

N_{ct} is the instantaneous number of fuel particles inside the bed and the corresponding mass of carbon will be

$$m_{ct} = N_{ct}\rho_c\frac{\pi\ d_{eq}^{\,3}}{6} \tag{17}$$

Because the bed is working in steady state regime, the rate of fuel supply is the same as the carbon combustion rate,

$$\dot{m}_c = \frac{2\ \pi\ d_{eq}^{\,2}C_pM_cN_{ct}}{\dfrac{\varphi\ d_{eq}}{Sh\ D_g} + \dfrac{2}{k_c}} \tag{18}$$

4. INFLUENCE OF PARTICLE FRAGMENTATION ON THE BURNER PERFORMANCE

Because of the thermal shock the char particles suffer at the moment of their introduction in the fluidized bed, they undergo a fragmentation. A batch of particles with a carbon mass of m_c is composed by N_c particles with an initial diameter d_i. When particles enter the bed, the batch is fragmented into n fractions of different initial diameters, so that their new after-breakage initial diameter is connected with the pre-breakage initial diameter by (Pinho, 2006),

$$d_{ji} = \alpha_{ji} d_i \tag{19}$$

where α_{ji} is a size reduction factor for the new particles belonging to the size fraction *j* (1<*j*<*n*). At the feeding instant, combustion has not yet started although the particles have already suffered fragmentation and so the initial mass of the batch is still the same and equal to the summation of the masses corresponding to the new size fractions formed after the particle breakage,

$$m_c = \sum_{j=1}^{n} m_{ji} \Rightarrow \rho_c \frac{\pi\, d_i^3}{6} N_c = \sum_j \rho_c \frac{\pi\, d_{ji}^3}{6} N_j \tag{20}$$

where N_j is the number of particles belonging to a given size fraction *j* if all the particles keep the same density ρ_c and d_{ji} is the post-breakage initial diameter of the particles of the size fraction *j*. As $d_{ji} = \alpha_{ji} d_i$ Eq. (20) changes to

$$\sum_j \alpha_{ji}^3 \frac{N_j}{N_c} = 1 \tag{21}$$

Particles now have different sizes, corresponding to the different resulting fractions after fragmentation and the corresponding burning velocities will also be different. In a given instant after particles are introduced inside the bed, there is a unique burned fraction for each one of the new size fractions,

$$d_j = \left(1 - f_j\right)^{1/3} \alpha_{ji} d_i \tag{22}$$

However if fragmentation was ignored, then for any time instant after the introduction of the particles in the bed, their average diameter d_s would be given by,

$$\rho_c \frac{\pi\, d_s^3}{6} N_c = \sum_j \rho_c \frac{\pi\, d_j^3}{6} N_j \tag{23}$$

and in consequence,

$$\sum_j f_j \alpha_{ji}^3 \frac{N_j}{N_c} = f \tag{24}$$

According to Rangel and Pinho (2010), the after breakage diameter ratio for each one of the obtained size fractions remains constant during combustion, $\alpha_{ji} = \alpha_j(t) = \text{const}$. So it is possible to correlate the average burned mass fraction for the whole batch, with the burned mass fraction for each one of the now existing after breakage size fractions. The mean effective diameter of the batch particles after fragmentation *d*, is not only a function of the burned fraction, but also of the fragmentation process itself and even a function of the initial diameter prior to the fragmentation process (Pinho, 2006)

$$d = d_i \frac{\left(1 - f\right)^{1/3}}{\left(\sum_j \frac{N_j}{N_c}\right)^{1/3}} \tag{25}$$

The quotient is the fragmentation factor,

$$\sigma = \sum_j \frac{N_j}{N_c} \tag{26}$$

which is the ratio between the total number of particles of a batch after suffering fragmentation by the initial number of particles of the same batch. The existence of fragmentation means that $\sigma > 1$. For pine nut char there is only primary fragmentation (Rangel and Pinho, 2010) and the fragmentation factor (Zhang *et al*., 2002; Pinho, 2006) had a value of

1.6. The mean initial diameter of a particle d_{in}, after the thermal shock and subsequent fragmentation once being introduced into the bed, will be given by,

$$d_{in} = \frac{d_i}{(\sigma)^{1/3}} \tag{27}$$

5. ANALYSIS OF THE PERFORMANCE OF A BURNER

To evaluate the steady state combustion of biomass in a fluidized bed, data obtained from a study of the bubbling fluidized bed combustion of wood chars (Moreira, 2008) will be used. The reaction rate constant for the heterogeneous phase reaction for commercial pine nut and cork oak chars, will then be given by

$$k_c = 1667{,}4 \, \exp\left[{-82{,}2 \times 10^{-6}} \Big/ {\bar{R} \, T_p} \right] \tag{28}$$

which is valid for the temperature range of 620 to 800 ºC.

The situation under analysis refers to the steady state operation of a 1 m internal diameter fluidized bed reactor with 0.3 m of static bed height operating according to the conditions presented in Tab. 1 and burning pine nut char. Other basic assumptions adopted for the present development are:

- The bed confining walls are refractory and adiabatic and the bed does not lose heat through radiation;
- Losses through unburned fine particles elutriated from the bed are neglected. In practical terms this approach makes sense as fluidized bed burners are usually equipped with cyclonic separation systems in order to collect unburned solid particles reintroducing them into the bed;
- The primary fragmentation of the fuel particles is initially neglected;
- Fuel particles burn at the uniform bed temperature of 800 ºC;
- The inlet air flow is at 25 ºC and 1 atm.

The CO formed at the surface of the particles will lately burn away from them, but inside the bed (Guedes de Carvalho *et al.*; 1991, Mota *et al.*, 1994; Fennell *et al.*, 2007). Figure 1 shows the performance of the fluidized bed burner. The mass flow rates of fluidizing air and fuel increase in order to keep the bed operating temperature at 800 ºC. The thermal power released during the combustion process goes from 92 to 208 kW, the fuel mass flow rate raises from 10.1 to 22.8 kg/h, while the mass flow rate of the fluidizing air goes from 400 until 900 kg/h. Figure 1 shows that the ratio U/U_{mf} changes from 5.7 to 12.7, and the maximum value is still far away from the ratio U_t/U_{mf} corresponding to the bed sand particles which is 26.4, the elutriation of the bed sand particles is negligible. The same cannot be said about the fuel particles because of their lower density they will be easily dragged out of the bed in the exhaust gases flow. In real systems solid unburned particles are captured by cyclonic systems and are reintroduced back in the bed. As such in a simple approach this can be done by ignoring this elutriation, capture and reintroducing sequence as if fuel particles are never elutriated away from the bed.

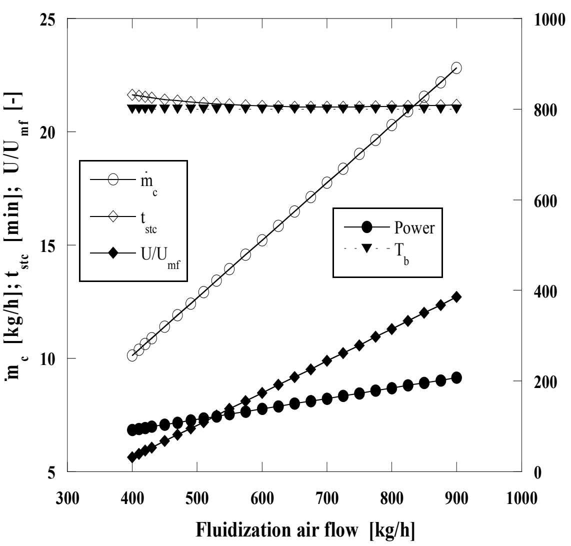

Figure 1. Behaviour of a bubbling fluidized bed burner with steady state combustion of pine nut char.

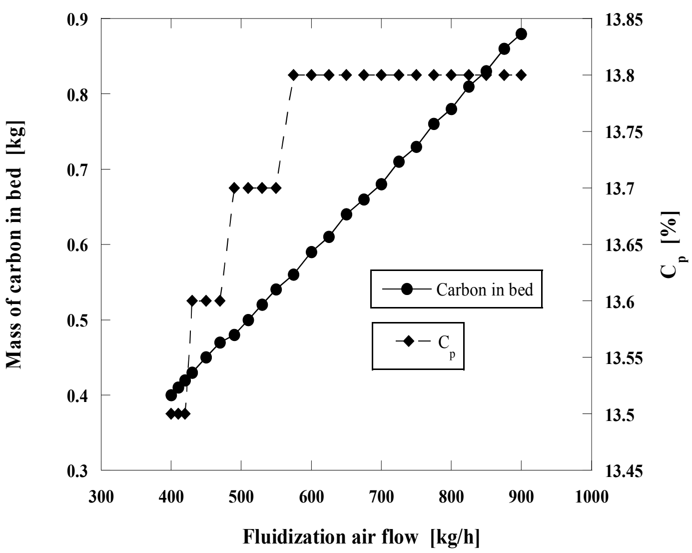

Figure 2. Bed carbon inventory and molar fraction of O_2 in the dense phase.

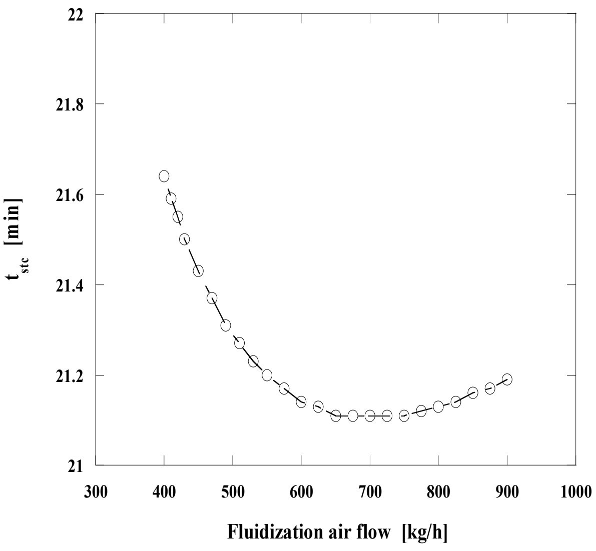

Figure 3. Evolution of the burning time of a particle in the steady the performance of the burner is very limited state conditions.

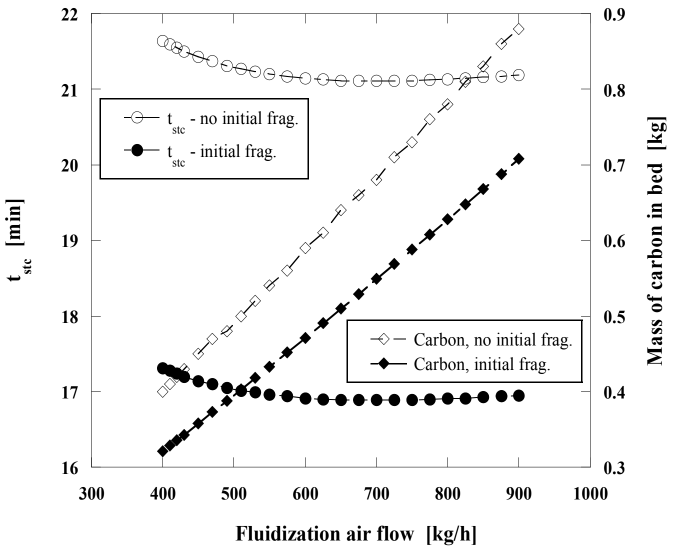

Figure 4. Influence of the initial fragmentation ratio on the particle burning time and on the bed carbon inventory.

Table 1. Properties of the fluidized bed under analysis.

Designation	Value	Unit
Bed diameter	1000	mm
Static bed height	300	mm
Sand density	3000	kg/m^3
Bed particle diameter	300	μm
Bed particle sphericity	0.77	-
Sherwood number	1.5	-
Number of distributor orifices	3000	-
Inter-particle competition factor	0.7	-
Mass flow rate of combustion air	300, 400, 500, 600	kg/h
Initial diameter of the char particles	5	mm

The bed carbon inventory, Fig. 2, shows how its amount increases although the same happens with the molar concentration of oxygen in the dense phase. The amount of fuel available in a given time instant is still a very small fraction of the mass of bed sand particles, 332 kg.

The particle combustion time, Fig. 3, follows a very interesting path for the range of operating conditions considered. For this particle size, type of char and bed temperature, diffusion has some supremacy on the reaction control and at the beginning of the combustion regime being analysed, the combustion time diminishes when the air flow rate increases, even if the carbon inventory in the bed is also increasing. From 650 kg/h to around 750 kg/h of fluidizing air mass flow, the combustion time is almost constant. In this intermediate region the combustion reaction is being controlled by the kinetics and the combustion time does not change because there is no change on the bed temperature. In the final part, for air mass flows above 750 kg/h, the importance of the increase of the carbon inventory in the bed is dominant and it raises the importance of bed hydrodynamics.

Figure 4 presents a comparison of the particle combustion time for steady state combustion as well as the bed carbon inventory for two situations, absence and existence of primary fragmentation of the particles. These two situations are for the operational conditions of Tab. 1, for an initial diameter of 5 mm when there is fragmentation with a ratio of $\sigma = 1.6$, leading to a pre-combustion initial diameter of 4.28 mm, as obtained through Eq. (27). Although the particle life time diminishes with the reduction of the average diameter of the particles, the flow of particles being fed into the bed stays constant and thus, in global terms, the average performance of the burner stays unchanged. The average diameter of the carbon particles inside the bed goes from 2.39 mm to 2.05 mm and besides this there is a reduction of the instantaneous amount of carbon in the bed. As the mass of carbon kept in a given time instant inside the bubbling bed is much smaller than the mass of inert composing the bed, the operating conditions of the bubbling fluidized bed reactor stay the same in broad terms, the hydrodynamic bed conditions are unchanged and are preponderant against kinetic and diffusion effects.

If the char was more fragile and the particle size reduction due to fragmentation was stronger a new burner operating condition could be achieved. However studies carried out with chars made with Portuguese woods (Rangel and Pinho, 2010) demonstrated that this breakage is primary, presents fragmentation ratios of the order of 1.6 and in practical terms the influence of this phenomenon on the performance of the burner is very limited.

6. CONCLUSIONS

Through the theoretical study of the steady state combustion of char particles in a bubbling fluidized bed reactor it was possible to asses the importance of the effect of the fragmentation of the fuel particles in the burner performance. The study was supported by a simple mathematical model for steady state combustion based on a model previously developed for batch combustion and used experimental data obtained in laboratory scale studies. The fuel under analysis was a vegetable char, nut pine char, of Portuguese origin.

It was found that the influence of the primary fragmentation of the fuel particles on the performance of the burner may have some importance.

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