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# CONSIDERATIONS ABOUT THE STEADY STATE COMBUSTION OF WOOD CHAR IN A BUBBLING FLUIDIZED BED REACTOR

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***Abstract.*** *A theoretical study on the performance of steady state bubbling fluidized bed burners is presented using a simple mathematical model. The proposed model has pedagogical and practical advantages due to its simplicity. The calculations whose results are plotted in several graphics were carried out based on data obtained in laboratory scale experiments. The experiments carried out with wood chars and the model allows a thorough evaluation of physical and chemical phenomena taking place inside the reactor, as well as a fast approach to the pre-design phase, before going towards more complex and time consuming numerical modeling. In the first part of the paper the steady state modeling is compared with the combustion of successive batches of char particles. Afterwards, the performance of a 1 m diameter bed operating from 700 to 800 ºC is shown. The results show the importance of fuel type, bed particle size and operating temperature upon the overall bed performance.*

***Keywords****: Biomass, wood char, fluidized bed, combustion, modeling*

## 1. INTRODUCTION

Fluidized bed burners are flexible in terms of fuel quality and have a relatively fast response to load variations. In these burners, fuel particles below 25 mm burn inside a bed of inert particles commonly of 0.5 to 1 mm diameter (Oka, 2004). The amount of fuel particles is a small mass fraction of the bed, around 1 %. The solid fuel is rapidly heated up to the bed operating temperature, which provokes a strong devolatilization of the fuel particles. The majority of the volatiles burn above the free surface of the bubbling bed, while the charcoal solid core of the particle 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). For low temperature combustion (800 to 900 ºC), low levels of $NO_x$ emissions are obtained. Other advantages are the high heat transfer coefficients that can be obtained if the heat transfer surfaces are placed inside the bubbling bed (Botteril, 1975), a relatively wide operating ratio and its ability to burn different fuels with high combustion efficiency, namely low calorific value fuels as biomass (Scala and Chirone, 2006).

The development of a simple mathematical model for the steady state combustion of biomass in a bubbling fluidized bed burner is interesting, as it will allow a quick understanding of the performance of this type of burners.

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

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

Here the results from a simple mathematical model for the steady state combustion are discussed. This model is based upon experimental data obtained in laboratory studies on the combustion of batches of coke or char particles. The model is simple enough to allow the students to rapidly grasp the relative importance of the different phenomena taking place during the combustion process. The fluidized bed reactor is considered an isothermal reactor and the burning particles are at bed temperature. This supposition can lead to some errors, for coke particles burning in a bubbling fluidized bed at 930 ºC Roscoe *et al*. (1980) verified experimentally that particles could be burning at temperatures 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 be possible that they are not representative of the overall behavior of the particles that compose the majority of the batch under combustion. There are some contradictions among several authors that more recently looked at this subject (Adánez *et al*., 2001 and Oka 2004). For example, 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 evolution of the temperature of the particles during the combustion process.

A single solid particle undergoing a combustion process, Fig. 1, takes an elemental time d$t$ to suffer an elemental reduction of its diameter d($d$),

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

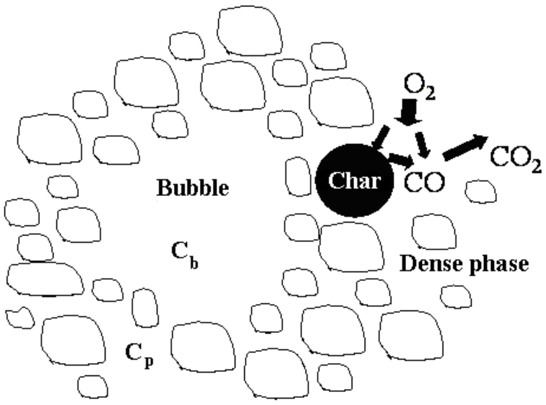


Figure 1. Schematic representation of the combustion of a carbon particle.

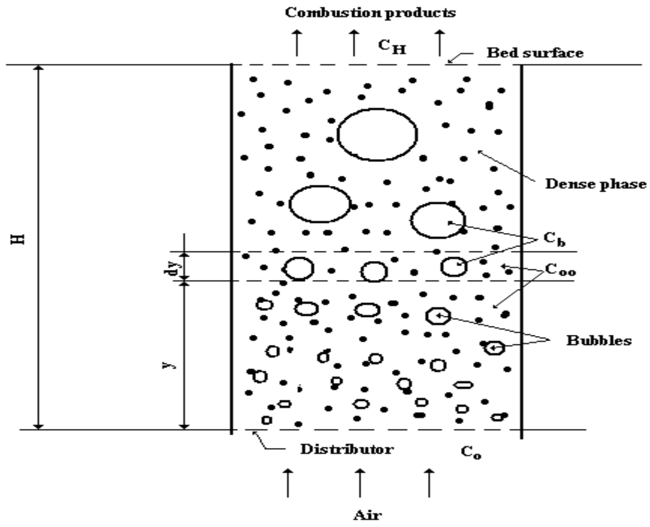


Figure 2. Schematic representation of the fluidized bed reactor.

The overall combustion resistance for a single particle is $\frac{1}{K}=\frac{\varphi\, d}{Sh\, D_g}+\frac{2}{k_c}$ (Pinho and Guedes de Carvalho, 1984; Guedes de Carvalho *et al*., 1991; Mota, *et al*. 1994). This particle belongs to a flow of particles that is continuously introduced into the bed where they compete for the available oxygen. To account for such competition a multiplying factor ( $0 < \eta \leq 1$ ), for the $O_2$ concentration inside the dense phase of the bed $C_p$, is used, (Annamalai, 1995),

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

where $\varphi$ is a parameter about the combustion model, $Sh$ is the Sherwood number and $D_g$ is the oxygen diffusivity.

Figure 2 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). 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 and their burning time is dependent on the $O_2$ concentration which is uniform and equal to $C_p$. Making an oxygen balance in a slice of fluidized bed with a thickness of d$y$ at level $y$, the $O_2$ inlet minus the outlet will be consumed in the combustion of the particles remaining inside it. As the $O_2$ concentration in the dense phase is constant and equal to $C_p$, it is only necessary to carry out the oxygen balance of the bubble phase,

$$\begin{aligned}&\left(U - U_{mf}\right)A_t\left(C_o - C_p\right)\left[exp\left(-\frac{X}{H}y\right) - exp\left(-\frac{X}{H}\left(y + \mathrm{d}y\right)\right)\right] = \\ &= -\dot{N}_p\frac{\rho_c}{2\, M_c}\pi\, d^{\,2}\mathrm{d}(d)\end{aligned} \tag{3}$$

$X$ is the number of times the bubble volume is swept during its ascension in the bed (Hovmand *et al*., 1971), $d_{iy}$ is the diameter of the fuel particles at the inlet of the bed slice under analysis and $\dot{N}_p$ 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. So, all the particles entering the bed at a given time instant burn at the same combustion ratio.

$$\frac{6\, \dot{m}_{cy}}{\rho_c \pi\, d_{iy}^{\;3}} = \frac{6\, \dot{m}_c}{\rho_c \pi\, d_i^{\;3}} = \dot{N}_p \tag{4}$$

To allow an analytical treatment, the following approximation can be adopted (Pinho, 2010),

$$\exp\left[-\frac{X}{H}(y+\mathrm{d}y)\right]-\exp\left[-\frac{X}{H}y\right]\approx-\left[\exp\left(-\frac{X}{H}y\right)\frac{X}{H}\mathrm{d}y\right] \tag{5}$$

giving the following system of equations,

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

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

that can now be integrated along the bed and for the combustion period, to give (Pinho, 2010),

$$t_f=\frac{\rho_c\left\{\frac{\varphi\left[1-(1-f)^{2/3}\right]d_i^2}{2\,Sh\,D_g}+\frac{2\left[1-(1-f)^{1/3}d_i\right]}{k_c}\right\}}{4M_c\eta\left\{C_o-\frac{f\,\dot{m}_c}{M_c(U-U_{mf})A_t\left[1-\exp(-X)\right]}\right\}} \tag{8}$$

In this development and in the subsequent analysis, the mass of a coal particle, or the mass of a batch of particles, or the mass flow rate of coal being introduced inside the bed, will always represent the equivalent carbon value. Then, 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 thus 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 $d_i$ and all the particles existing inside the bed with the same generic diameter $d$ will all have the same life time. The mass flow rate of carbon fed into the bed corresponds to the introduction of $\dot{N}_c$ (particles/s),

$$\dot{m}_c=\dot{N}_c\rho_c\frac{\pi d_i^3}{6} \tag{9}$$

During the life time of these particles $t_{stc}$, obtained from equation (8) with a burned fraction of $f$ = 1, many other particles will be entering the bed at the rate of $\dot{N}_c$, so that the total number of particles existing inside the bed during such time will be,

$$N_{ct}=\dot{N}_c t_{stc} \tag{10}$$

and the total mass of carbon remaining inside the bed will be given by

$$m_{ct}=N_{ct}\rho_c\frac{\pi d_{eq}^3}{6} \tag{11}$$

The carbon combustion rate is calculated through

$$\dot{m}_c=\frac{2\pi d_{eq}^2 C_p M_c N_{ct}}{\frac{\varphi d_{eq}}{Sh\,D_g}+\frac{2}{k_c}} \tag{12}$$

## 4. ANALYSIS OF SOME PARTICULAR SITUATIONS

This model will now be applied to the determination of the steady state combustion rate of biomass char in fluidized bed. The reaction rate constant $k_c$, for the heterogeneous phase reaction $CO + \frac{1}{2}O_2 \rightarrow CO_2$, for commercial chars made from nut pine and cork oak, is given by (Moreira, 2008) and is valid for the temperature range of 620 to 800 ºC.

$$k_c = 1667{,}4 \; \exp\left[-82{,}2 \times 10^{-6} \Big/ \bar{R}\,T_p\right] \tag{13}$$

In equation (13) $T_p$ is the particle temperature and $\bar{R}$ is the universal gas constant.

### 4.1 The steady state combustion and the combustion of successive batches

As the steady state combustion can be approximated by the combustion of successive batches thrown into the bed, it is convenient to compare the combustion results for successive batches with an equivalent steady state situation. The ratio between the carbon mass of the batch and the corresponding combustion time gives the mass flow rate of carbon supplied and will be the carbon bed inflow under steady state operating conditions. Figure 3 concerns particles with an initial diameter of 2 mm burning at 800 ºC, while Fig. 4 concerns 5 mm particles burning at 750 ºC. Table 1 shows the characteristics of the fluidized bed. In this first analysis it is not considered a mass flow of carbon enough to guarantee the maintenance of an adequate bed temperature. It is assumed a supplementary heating support to assure the maintenance of the required bed temperature, like in laboratory experiments.

Through the comparison between the batch combustion time $t_{bc}$ and the corresponding steady state combustion time $t_{stc}$, it is possible to verify that the divergence of values is about 5 to 9 %, assuming an inter-particle competition factor for the available oxygen of $\eta = 1$ in the steady state combustion regime, Fig. 3. In the situation considered in Fig. 4 the difference between the two combustion times is smaller, of the order of 1 to 2 %.

Table 1. Characteristics of the fluidized bed for comparison between batch and steady state combustion.

| Designation | Value | Unit |
|---|---|---|
| Bed diameter | 80 | mm |
| Bed static height | 200 | mm |
| Bed voidage at incipient fluidization | 0.53 | - |
| Bed particle diameter | 300 | μm |
| Bed particle sphericity | 0.77 | - |
| Pressure inside the bed | 1 | atm |
| Volume fraction of $O_2$ at the bed inlet | 21 | % |
| Sherwood number | 1.5 | - |
| Number of orifices in the distributor | 101 | - |

It is clear that the model for the steady state combustion allows a clear approach to the particle life time of a particle burning inside the bed, when considering as a reference basis the combustion of successive batches of particles.

### 4.2 Steady state combustion of chars made from pine nut and cork

Having done the comparison between the steady state and the batch combustion it is now possible to analyze the influence of the different parameters upon the steady state combustion performance. The first point to consider is the importance of the inter particle competition for the available oxygen through the evaluation of the combustion of 60 g/h of pine nut char particles with 2 mm initial diameter in a bed operating at 750 ºC. In Fig. 5 it can be seen that between $\eta$ = 0.6 and 1 there is a 65 % reduction on the combustion time of a particle, using the $\eta$ = 0.6 situation as the reference.

Analyzing now the influence of the bed temperature, keeping the other operating conditions constant (60 g/h of pine nut char, 2 mm initial diameter of fuel particles and $U/U_mf = 7$), the curves shown in Fig. 6 were obtained, whereas in Fig. 7 burning times for particles with initial diameters between 1 and 5 mm, carbon mass flow rates of 60 and 120 g/h and for a bed temperature of 800 ºC, are shown.

The importance of the superficial velocity ratio $U/U_{mf}$ is shown in Fig. 8 for combustion at 750 ºC and $\eta$ = 0.7. The influence of this velocity ratio $U/U_{mf}$ is enhanced with the fuel particle diameter indicating that for smaller particles the combustion reaction is kinetically controlled while diffusion starts to dominate for larger diameters.

In Fig. 9 the pine nut and cork oak chars are compared. The cork oak char is more porous, more reactive, and presents a higher intrinsic area available for the heterogeneous phase reaction, leading to shorter combustion times.

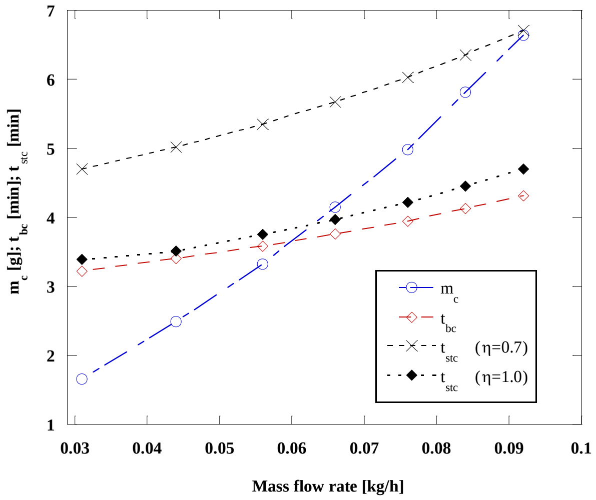


Figure 3. Combustion of batches and steady state combustion of pine nut char particles with 2 mm initial diameter, at 800 ºC and for $U/U_{mf}$ = 7.

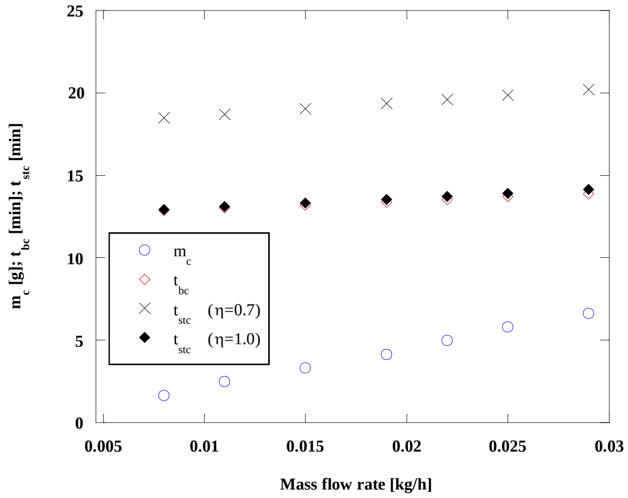


Figure 4. Combustion of batches and steady state combustion of pine nut char particles with 5 mm initial diameter, at 750 ºC and for $U/U_{mf}$ = 7.

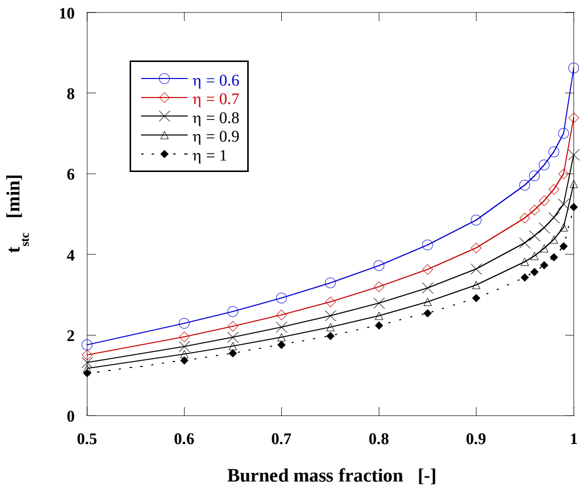


Figure 5. Particle burning time for steady state combustion of a mass flow rate of 60 g/h of pine nut char with 2 mm, for $U/U_{mf}$ = 7 and $T_{bed}$ = 750 ºC.

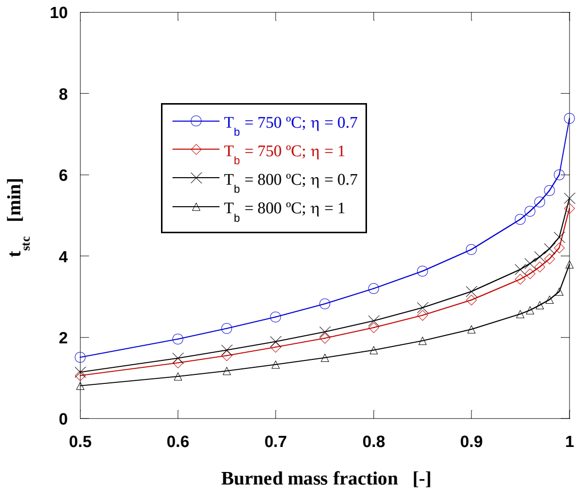


Figure 6. Particle burning time for a mass flow rate of 60 g/h of pine nut char at 750 and 800 ºC ($d_i$ = 2 mm and $U/U_{mf}$ = 7).

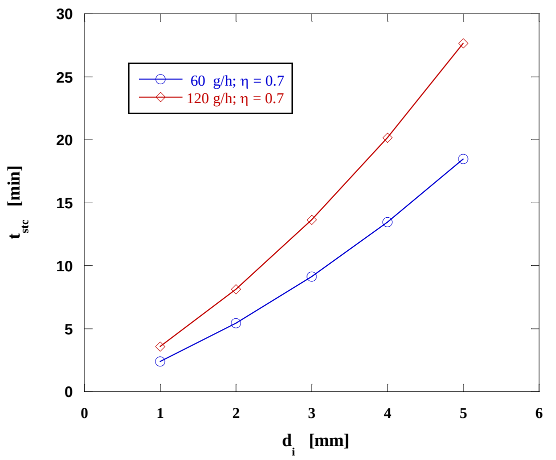


Figure 7. Particle burning time for mass flow rates of 60 and 120 g/h of pine nut char at 800 ºC and $U/U_{mf}$ = 7.

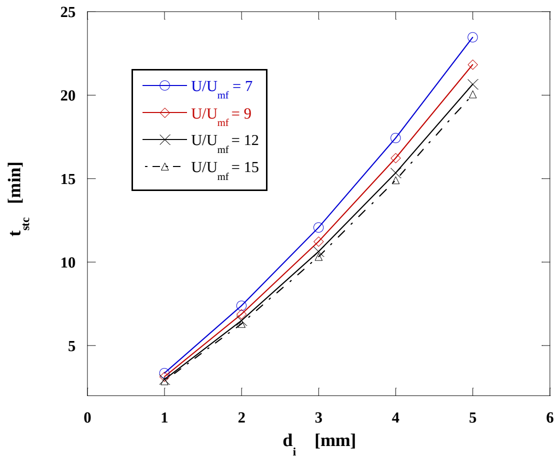


Figure 8. Importance of $U/U_{mf}$ on the particle burning time for steady sate combustion. Mass flow rate of pine nut char 60 g/h, $\eta$ = 0.7 at 750 ºC

### 4.3 Performance of a fluidized bed burner with 1 m diameter

Next is the analysis of the behavior of a larger bubbling fluidized bed burner working in steady state and with self-sustained combustion of pine nut char. The dimensions and main properties of this reactor are presented in Table 2. For a self-sustained combustion, the amount of thermal energy released by the combustion reaction taking place inside the bubbling bed must be accounted for, to assure a proper combustion temperature. Main assumptions adopted were:

- The bed confining walls are adiabatic and there are no radiation losses from the bed free surface;
- Losses through unburned elutriated particles bed are neglected;
- Fuel particles burn at bed temperature and their primary fragmentation is neglected;
- The inlet air flow is at 25 ºC and 1 atm.

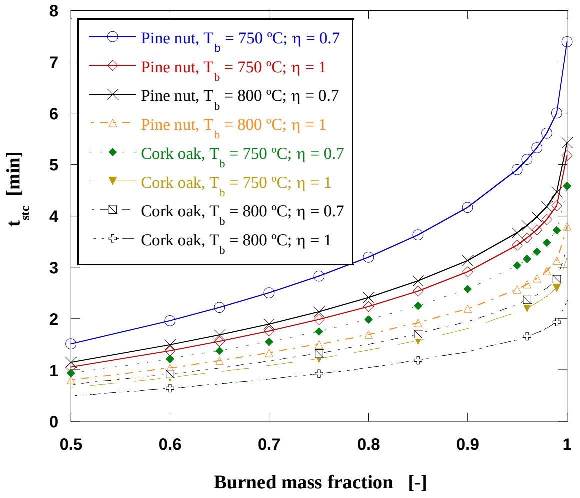


Figure 9. Steady state combustion of pine nut and cork oak char. Mass flow rate of 60 g/h, $d_i$ = 2 mm and $U/U_{mf}$ = 7.

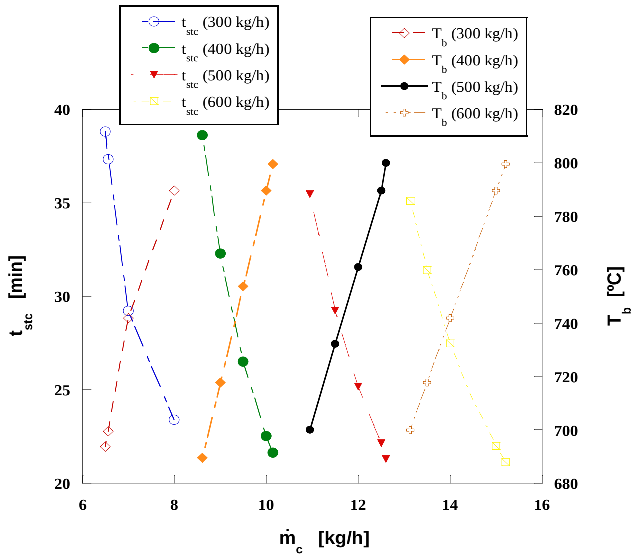


Figure 10. Bed diameter 1 m. Static bed height 0.3 m. Bed particle size 300 μm. Particle initial diameter 5 mm.

Table 2. Dimensions and main properties of the 1 m diameter reactor.

| 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 |

In Fig. 10, results from the application of this model to the four mass flow rates of combustion air, presented in Tab. 2, are plotted. The burner is fed with ambient air (21 % (v/v) $O_2$) and works at atmospheric pressure. It is assumed that the CO formed by the heterogeneous phase reaction 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). The results in Fig. 10 show the particle burning time in a steady state combustion process $t_{stc}$, and the evolution of the bed temperature as a function of the nut pine char mass flow rate (effective carbon mass flow rate). Figure 11, presents the curves corresponding to the thermal power being released through the char particles combustion process, assuming a calorific value for carbon of 32,794 kJ/kg. The evolution of $U/U_{mf}$ is also plotted in the figure. Finally Fig. 12, presents the instantaneous inventory of carbon mass in the bed as well as the evolution of the oxygen volume fraction inside the dense phase of the bed, whose values change from 13 to 14.7 %. All these tendencies are plotted as a function of the equivalent carbon mass flow rate fed into the bed.

These calculations result in four well defined equivalent carbon mass flow rates, 6.5 to 8 kg/h, 8.6 to 10.1 kg/h, 11 to 12.6 kg/h and finally 13.1 to 15.2 kg/h, for a constant mass flow rate of fluidizing air for each fuel flow range, showing the dependence between the bed operating conditions and the combustion regime.

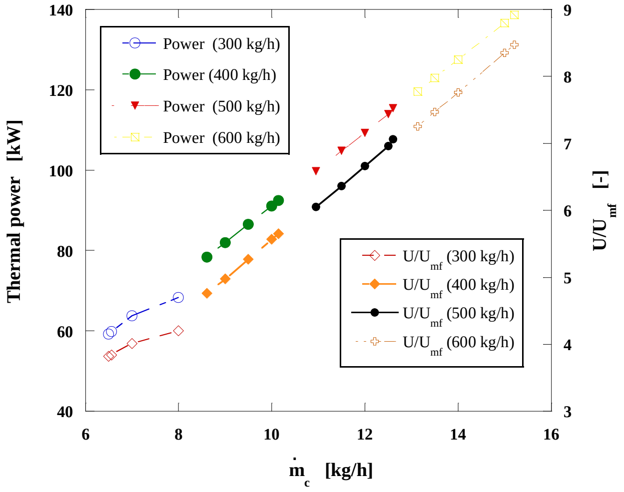


Figure 11. Thermal power output of the 1 m bed. Particle initial diameter 5 mm.

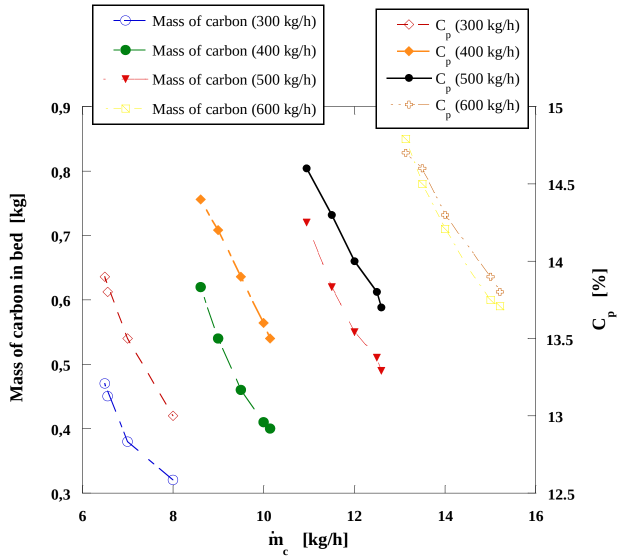


Figure 12. Carbon inventory of the 1 m bed. Particle initial diameter 5 mm.

The mass flow rates of 300, 400, 500 and 6000 kg/h refer to the air flow rates necessary to establish the four operating regimes previously referred. According to the carbon inventory in the bed, its value is always below to1 kg for all analyzed situations, while the total mass of the bed sand is of 332 kg, which means that the carbon mass fraction inside the bed is always lower than 0.3%.

In Fig. 13 results of the bed behavior in a situation closer to reality are presented, because in practical terms the control systems adjust the fuel and air flows in order to keep the bubbling bed temperature inside close limits previously defined by the operators. For this case the working conditions are in general those of Tab. 2, although with some slight changes. Figure 13 shows the performance of the fluidized bed burning nut pine char in steady state regime. Particles have an initial diameter of 2 mm and the bed temperature is of 750 ºC. The mass flow rates of fluidizing air and fuel increase in order to keep the bed operating temperature at 750 ºC. The thermal power released during the combustion process goes from 86 to 194 kW, when the carbon equivalent mass flow rate raises from 9.4 to 21.3 kg/h, while the mass flow rate of the fluidizing air goes from 400 until 900 kg/h.

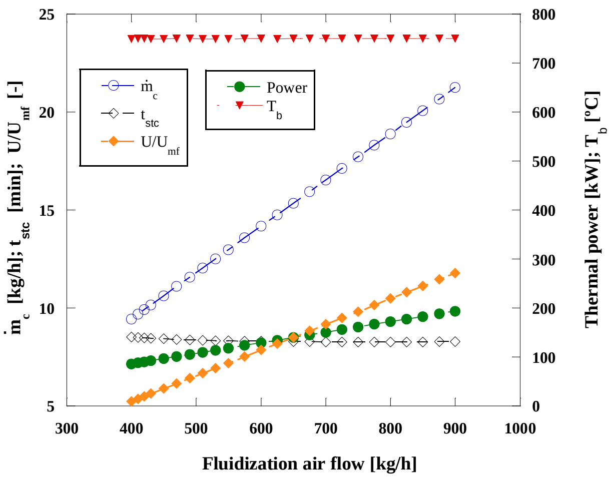


Figure 13 – Performance of the 1 m bed burning pine nut char. Initial diameter of char particles, 2 mm. Bed temperature, 750 ºC.

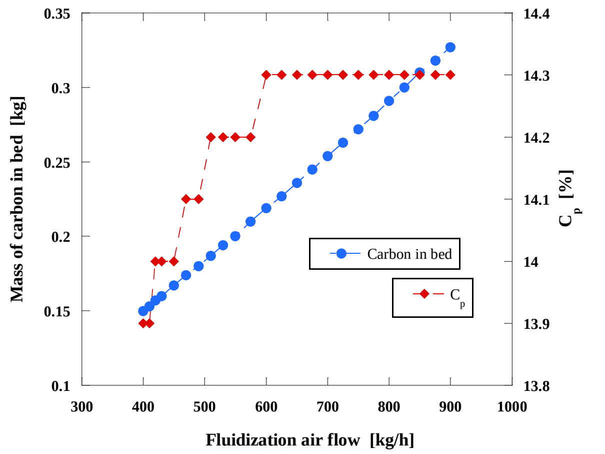


Figure 14 – Bed carbon inventory and molar fraction of $O_2$ in the dense phase. Operating conditions are those of Figure 13.

There is a minor reduction of the particle burning time because for this initial size the combustion is kinetically controlled and as the bed temperature stays constant the particle combustion time barely changes. The ratio $U/U_{mf}$ changes from 5.2 to 11.8, meaning that 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, but such event was not accounted for. The bed carbon inventory, Fig. 14, increases. 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 bed mass.

## 5. CONCLUSIONS

Results from simple mathematical model for the steady state combustion of biomass char in bubbling fluidized bed burner are presented and discussed. The validation of the model is carried out through the comparison of the steady state combustion with data concerning the burning of batches of wood char in a laboratory scale combustor and extending these to the combustion of a sequence batches.

The importance of fuel particle diameter and type, as well as of the bed operating conditions, can be easily detected and quantified. The model was used to analyze the performance of a 1 m diameter and 0.3 m height bed, burning pine nut and cork oak chars, in the 750 to 800 ºC operating range and for four different air and fuel mass flow rates.

The main use and justification of this very simple model is for teaching purposes and for the pre-design of fluidized bed burners.

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