ABSTRACT

Circulating fluidised bed combustion (CFBC) is receiving wide research attention in view its potential as an economic and environmentally acceptable technology for burning low-grade coals, biomass and organic wastes, and thereby mixtures of them. Designs of the existing fluidised bed boilers for biomass combustion are mainly based on experience from coal combustion because the mechanism of combustion of biomass in fluidised beds is still not well understood. A good understanding of the combustion and pollutant formation processes and the modelling of the combustor can greatly avoid costly upsets of the plants.

In this paper the performance of CFBC burning coal and biomass mixtures was analysed. Experimental results were obtained from the combustion of two kind of coals with a
forest residue (Pine bark) in two CFB pilot plants (0.1 and 0.3MWth). The effect of the main operating conditions on carbon combustion efficiency was analysed. Moreover, a mathematical model to predict the behaviour of the co-combustion of coal and biomass wastes in CFB boilers has been developed and validated. The developed model can predict the different gas concentrations along the riser (O2, CO, CH4, etc.), and the carbon combustion efficiency. The experimental results of carbon combustion efficiencies were compared with those predicted by the model and a good correlation was found for all the conditions used.

Keywords: co-firing, circulating fluidised beds, mathematical modelling.

1. INTRODUCTION

The use of biomass as an energy source has primarily addressed direct combustion, pyrolysis or fermentation for alcohol production. Until recently, there have been few studies concerning the co-firing of coal/biomass blends for energy generation [1]. Some typical biomass fuels in co-firing studies are cattle manure, sawdust, sewage sludge, wood chips, straw and refuse-derived fuels. Biomass fuels are considered environmentally friendly for several reasons. There is no net increase in CO2 because of burning a biomass fuel. Therefore blending coal with biomass fuels can reduce fossil-based CO2 emissions. Co-firing brings additional greenhouse gas mitigation by avoiding CH4 release from the otherwise landfilled biomass residues (sewage sludge, manure, etc.). The alkaline ash from biomass also captures some of the SO2 produced during combustion and therefore the net SO2 emissions can also be reduced by co-firing. In addition the fuel nitrogen content in biomass is in many cases much lower than in coals and is mainly converted to ammonia during combustion. Hence co-firing can also result
in lower NO\(_x\) levels. Blending can also result in the utilization of less-expensive fuels with a reduction in fuel costs.

There are several works dealing with the effect of biomass addition on the gas emissions [2-14]. They concluded that the levels of pollutants decreased with increasing the amount of biomass fuel added. Nordin [3] optimised the sulphur retention when co-combustion of coal and biomass fuels in a fluidised bed using statistical experimental designs for operating variables. The most influential factors were found to be the load, the primary air and the total airflow. Dayton et al. [8] investigated the interactions between biomass feedstock and coal to address the issues of gaseous emissions when co-combusting these fuels. The results revealed the synergetic effects of co-firing for CIH, KCl and NaCl, but the amounts of NO and SO\(_2\) detected suggested that any decrease was the result of diluting the N or S present in the fuel blend. However, they stated that any advantages of larger-than-expected SO\(_2\) reductions because of S capture by the biomass ash constituents might occur in large-scale systems. Recently, Armesto et al. [14] have carried out the co-combustion of a coal and an olive oil industry residue (foot cake) in a bubbling fluidised bed pilot plant to study the effect of some operating conditions on the emissions and combustion efficiencies. They found that the share of waste in the mixture (10-25%) has not any effect on combustion efficiency, although the effect of the waste in SO\(_2\) emissions is important due to the calcium and potassium content of the biomass.

The circulating fluidised bed technology was first used for combustion of coal because of its unique ability to handle low-quality, high-sulphur coals. In forest-rich countries, CFB combustion has increased its market share of biomass combustion during recent years. Extensive experimental investigations have been performed to date on the feasibility and performance of the CFBC of different alternative fuels. One of the first works on co-combustion was that of Leckner and Karlsson [2] who measured
experimental emissions of NO, N₂O, SO₂ and CO from combustion of mixtures of bituminous coal and wood in a CFB. They concluded that emissions from the combustion of mixtures are approximately proportional to the mixing ratio of the fuels and to the emissions properties of the respective fuels. Hein and Bemtgen [7] studied the co-combustion of different biomass with coal into different combustion techniques in a variety of scale pilot plants and large-scale power stations. They found that CFBs could be designed to handle the size of wood chips and that effect of biomass addition on the SO₂ emissions was significantly positive for all FB facilities. Werther et al. [9] and Amand et al. [10] recommended that in a CFB combustion system, an operation with higher excess air in comparison with the pure coal combustion conditions might become necessary.

There is currently a focus on developing models of CFB for burning biomass and waste material. The objectives of these models are to be able to predict the behaviour with respect to the combustion efficiency, fouling problems and pollutant emissions performance of different fuels or mixtures in commercial scale fluidised bed combustors. Combustion modelling for coal/biomass blends is a complex problem that involves gas and particle phases along with the chemical reactions. Most mathematical models consist of sub-models for fluid-mechanics, particle dispersion, fuel devolatilisation, gaseous combustion, heterogeneous char reaction, and pollutant formation. Existing coal combustion models should be modified to include the effects of biomass co-firing on the overall combustion behaviour. The problem in blend combustion is that two chemically different fuels are involved (biomass is much more reactive and has higher volatile and moisture contents than coal). There are few modelling studies on blend combustion in the literature because co-firing is a developing technology still in the testing phase. Sami et
al. [1] made a revision of the modelling effort on co-firing and found some models but only for pulverised or swirls burners.

Based on previous works on mathematical modelling of CFB coal combustion [15] and on biomass combustion [16,17], a mathematical model is developed here for the combustion of coal and biomass mixtures in circulating fluidised bed combustors. The model was validated with the experimental results found in tests carried out in two different CFBC pilot plants.

2. EXPERIMENTAL

2.1. Experimental facilities

Figures 1 and 2 show a schematic diagram of the two experimental set-ups used in this work. In the pilot plant of 300 kWth (CIEMAT, Spain) the riser was a cylinder of 200-mm id. and 6.5 m high. It is covered inside with refractory ceramic. Solid recirculation is carried out using the cyclone, return leg and J-valve. They are lined with refractory ceramic as well. The connection between the riser and the return leg is made with a J-valve. The pilot plant is also equipped with a combustion air preheater, which may be used for faster pre-heating during the star-up. Bed material was sand with a particle size between 0.3-0.5 mm. The secondary air was introduced through the wall at 1.5 m above the distributor plate.

The feeding system has two fuel hoppers mounted on a balance. The biomass and the coal are fed simultaneously with these systems to a third screw feeder system leading to the boiler. The rotation speed of this screw feeder is kept constant and high. Coal and biomass mass flows are controlled by separate. The plant is instrumented for measurement of pressures, temperatures and gas flow rates. All process variables are recorded in continuos form and processed by the control system. A sample is
continuously extracted from flue gases before bag filter and sent to the on-line flue gas analysers (O₂, CO₂, CO, N₂O, NOₓ, CH₄, SO₂, and HCl). The sample must be relatively clean and dry before entering the analysers so the sample is filtered and condensed.

The VTT’s pilot plant riser (100 kWth) is a cylinder of 170 mm id. and 8 m high. Bed material was sand with a particle size between 0.1-0.3 mm. Mean gas velocity in the reactor was 2.3 m/s. The share of primary air was 50%. Secondary air was fed through the uppermost port (2 m). Fuels were fed from separate vessels and mixed in a screw feeder. Temperature in the riser was kept constant at 870°C during the experiment. Dried flue gas was analysed with traditional on-line analysers (O₂, CO₂, CO, NO, SO₂) and wet and hot (180°C) flue gas with FTIR (CO₂, CO, NO, NO₂, CH₄, SO₂, H₂O, HCl). In this pilot plant, the gas concentrations were measured inside the riser at different heights in some experiments to assist in the model validation.

2.2. Fuel characteristics

The fuels used were a South African sub-bituminous coal (SA) from Fortum’s Meri-Pori power plant in Finland, a high sulphur content lignite (LT) from Teruel (Spain) and pine bark (PB) from UPM-Kymmene, Rauma Mills in Finland. Table 1 shows the analyses of the Finnish pine bark. The main characteristics of the pine bark are its high volatile matter and its low ash and sulphur contents. Initially, the moisture content of pine bark was 37% (wt.). In the CIEMAT pilot plant, the moisture content of the biomass decreased from 37% to 11% during the storage and grinding. Table 1 also shows the proximate and ultimate analysis of the coals. Figure 3 shows the particle size distributions of the different fuels, measured by sieving, used in the two pilot plants. Although the fuels used in the two pilot plants are the same, they have different particle size distributions because of the different milling systems of the installations. The pine bark particle size used is less than 3 mm in VTT pilot plant, however it ranges to 30 mm in the CIEMAT combustor.
South African coal was used with two different particle size distributions in the CIEMAT pilot plant, SA and SA2. This second distribution (SA2) was obtained removing the solid fraction below 417 μm of the first distribution (SA) by sieving.

3. RESULTS

Different operating variables were analysed in the CIEMAT pilot plant: the share of pine bark in fuel blend ($F_{\text{biomass}}=0-100$ %wt), combustor temperature ($T=800-900^\circ\text{C}$), fluidisation velocity ($u=4-6$ m/s), excess air ($\text{exc}=18-25$ %), secondary air/total air ratio ($\text{sec. air}=10-35$ %) and particle size distribution of the feed (SA and SA2). Moreover, influence of fuel type on combustion efficiency was studied burning a sub-bituminous and a lignite coal.

In the VTT pilot plant the effect of the share of pine bark in fuel blend ($F_{\text{biomass}}=0-100$%wt) on combustion efficiency was studied burning a sub-bituminous coal.

During the experimental work, steady state was maintained unless for three hours. At the end of the steady state the different solid streams (bed drained and bag filter) were weighed and analysed for unburned carbon content. To avoid analysis errors due to the low C concentration these solids samples were concentrated. Solid samples were leached with HCl increasing the organic C and decreasing the C analysis errors. Carbon analysis were made in a Carlo Erba CHN-O analyser. The carbon combustion efficiency ($E_c$) was calculated considering the C feed in and the C losses in the different solid streams (drainage and cyclone) by eq. (1). The contribution of the gas phase, mainly as CO (100 - 400 ppms by volume), to the total unburnt carbon losses were considered.

$$E_c(\%) = \frac{F_A x_{c,A} - (F_D x_{c,D} + F_C x_{c,C} + F_{\text{gas}} x_{CO})}{F_A x_{c,A}} \times 100$$ (1)
Figures 4-9 show the effect of the operating conditions on the experimental carbon combustion efficiencies ($E_c$). Figure 4 shows the effect of the percentage of biomass added in the feed on the $E_c$ for both pilot plants and two kinds of coals. As can be seen, for both coals the carbon combustion efficiency increased when the percentage of biomass increased. These results were expected because the particle size distributions of the coals had great amount of fines, being part of these fine particles lost in the cyclone system. It can also be appreciated the higher $E_c$ found for the lignite compared to the South African coal, due to the high reactivity of this coal.

Figures 5 and 6 show the effect of the combustor temperature on $E_c$ and the carbon concentration in the bottom region when using SA coal mixed with 60% wt of PB and LT coal mixed with 50% wt of PB. An increase in the combustor temperature increased the carbon combustion efficiency and decreased the carbon concentration due to the increase in the reaction rates. In all cases the carbon concentration was low, and as it was expected the char concentration burning the lignite was lower than burning the South African coal due to the higher reactivity of the lignite.

Figure 7 shows $E_c$ obtained with the sub-bituminous coal when working at different linear gas velocities keeping constant the temperature and the excess air. An increase in the linear gas velocity gave a decrease on $E_c$ because the solid circulation flowrate increased when gas velocity increased and so the flowrate of solid losses by the cyclone increased. This variable mainly act on the mean residence time of char particles in the bed, decreasing the residence time with increasing the gas velocity.

To analyse the effect of the particle size distribution of the fuel on $E_c$, the SA sub-bituminous coal was sieved to obtain a different particle size distribution with a lower amount of fine particles. Figure 8 shows the $E_c$ obtained with this new coal particle size distribution when working at different linear gas velocities. An increase of $E_c$ for all
gas velocities were obtained due to this new coal particle size distribution had fewer amounts of fine particles. The fine particles that can not be recovered by the cyclone are the main lost of unburned char particles in the system, and so on the Ec. We can also observe in this figure that the effect of gas velocity was higher with the distribution with greater proportion of fine particles.

Figure 8 shows Ec obtained as a function of excess air. An increase of excess air gave an increase in the mean oxygen concentration in the bed, thus increasing the carbon combustion efficiency. The introduction of a part of the combustion air as secondary air generates a zone in the lower region of the combustor with low oxygen concentration. As seen in Figure 9 an increase of the percentage of secondary air gave a small decrease of the efficiency. These results can be explained taking into account the fact that an increase in the percentage of secondary air produces a decrease in the oxygen concentration in the lower part of the combustor and therefore a decrease in the combustion rate.

4. MATHEMATICAL MODEL

The carbon combustion efficiencies in CFBC depends on bed temperature, gas velocity, excess air, feed particle size distribution and fuel reactivity. The great number of operation variables make a systematic experimentation very costly, as the costs in a pilot plant are relatively high. Therefore, in order to simulate and optimise the reactor, a global model of the system is necessary. In this work a mathematical model was developed for the combustion of coal and biomass mixtures in circulating fluidised bed combustors integrating hydrodynamic, devolatilisation and combustion kinetic submodels. The main hypotheses used in constructing the main submodels are discussed below. A CFB furnace in steady state and isothermal at a macroscopic level was considered.
4.1. Hydrodynamics. The hydrodynamic characteristics of the CFB were modelled taking into account the works of Johnson et al. [18], Johnson and Leckner [19] and Pallares et al. [20]. The riser was divided into three different zones: bottom, characterised by a dense bed, similarly to a bubbling bed; splash with a predominant homogeneous particle clustering flow; and transport zone with a core-annulus structure. In the splash and transport zones, the vertical distribution of solids was determined with an exponential decay model. The solid concentration was assumed to be the sum of the contribution from a cluster phase and a dispersed phase:

\[
\rho = (\rho_b - \rho_{d,b}) \exp[-a(h - H_b)] + \rho_{\text{exit}} \exp[K(H_0 - h)]
\]  

(2)

\[
\rho_{d,b} = \rho_{\text{exit}} \exp[K(H_0 - H_b)]
\]

(3)

\[
a = 4u_1 / u
\]

(4)

\[
K = 0.23 / (u - u_1)
\]

(5)

The solution of the hydrodynamic model gives, at each riser height, mean voidage, annulus and core voidages, core radius, upward solids flow in the core, downward solids flow in the annulus and external circulation solid flux.

4.2. Devolatilisation of biomass/coal. The model developed by de Diego et al. [16] was used to calculate the volatile generation rate of pine bark and coal particles. In this model the drying and pyrolysis of biomass/coal particles was assumed to be a coupled process controlled by the kinetics of devolatilisation as well as the heat transfer to and through the particles. The particles were assumed to be spherical and characterised by an equivalent particle diameter and a shape factor. The kinetic rate of volatiles was described using a distributed activation energy model with the kinetic parameters \(k_0\), \(E_0\) and \(\sigma_E\) shown in Table 2. The volatiles generated during the devolatilisation were considered as a mixture of \(\text{H}_2\text{O}\), \(\text{CO}\), \(\text{CO}_2\), \(\text{H}_2\), \(\text{CH}_4\), \(\text{C}_2\text{H}_4\) and \(\text{C}_3\text{H}_8\). The excess of C was considered as
elementary C, which is instantaneously oxidised to CO. In addition, to know the
generation of volatiles in the different locations inside the boiler it was necessary to know
the distribution of the devolatilising particles along the riser, both biomass and coal
particles. Two kinds of particles were assumed: large non-elutriable particles which
devolatilise uniformly in the bottom and splash regions and fine particles of elutriable
size which can devolatilise along all the combustor. To determine the distribution of
elutriable particles along the combustion chamber, age population balances of
devolatilising particles were developed for biomass and coal [17]. It has to be pointed out
that primary fragmentation of the biomass and coal particles has been included in the
model. The original particle was divided into a number of particles, \( N_p \), whose volumes
added had the same volume as the one of the initial particle.

The system of equations was solved for each elutriable particle size interval and fuel type.
By coupling the age distributions of devolatilising biomass/coal particles with the model
of drying and devolatilisation of biomass/coal, the volatile generation rates in the
different regions of the riser were obtained. In the bottom and splash regions the total
generation rate of volatiles at each height was determined by the sum of the rates for
volatiles generated from the elutriable and non-elutriable particles (biomass and coal).

4.3. Volatile combustion. The following chemical reactions with their corresponding
reaction rates are considered for volatile combustion.

\[
\begin{align*}
(-r_1) &= \text{CH}_4 + \frac{3}{2}\text{O}_2 \rightarrow \text{CO} + 2\text{H}_2\text{O} \\
(-r_2) &= \text{C}_3\text{H}_8 \rightarrow \frac{3}{2}\text{C}_2\text{H}_4 + \text{H}_2 \\
(-r_3) &= \text{C}_2\text{H}_4 + \text{O}_2 \rightarrow 2\text{CO} + 2\text{H}_2 \\
(-r_4) &= \text{H}_2 + \frac{1}{2}\text{O}_2 \rightarrow \text{H}_2\text{O}
\end{align*}
\]
Hydrocarbons are oxidised in two steps with CO as the intermediate reaction product. In the bottom zone, volatile combustion is modelled following Srinivasian’s [21] assumptions, that is, propane pyrolysis and hydrogen oxidation occurs in the emulsion phase whereas CO, ethylene and methane oxidation occur only in the bubble phase. The kinetic constants for these reactions were taken from Dryer and Glassman [22], Hautman et al. [23], and van der Vaart [24].

Mass balances for the oxygen and the \( n \) different volatiles were developed for the different regions inside the riser [17]. In the bottom bed, a set of \( 2n+2 \) ordinary first-order differential equations was obtained for the bubble and emulsion phases. Similarly, for the splash and core of the transport region, a set of \( n+1 \) equations was obtained assuming plug flow of gas. These mass balance equations allow us to determine the oxygen and volatiles concentrations along the combustor. This system of equations was solved by a Runge-Kutta method starting from the distributor plate and was coupled with the char population balances to fit the oxygen balance.

4.4. Char combustion. To enforce mass balances and determine carbon combustion efficiencies in a CFB with shrinking particles, it is necessary to develop population balances of char particles in the different zones of the CFB (bottom, splash, and transport zones). Secondary fragmentation has been taken into account for modelling purposes. The effects of secondary fragmentation are included in terms of a fragmentation rate constant \( (k_f) \), and a distribution function \( (P_f) \) of fragments. The relative radius \( \gamma \), controls the mass distribution of the fragments. A value of 0.13 for the relative radius and \( k_f = 3.25 \times 10^{-6}/t_{\text{mother}} \) has been taken from Thunman [25]. For discrete particle size distributions, the
population balances of char particles in the bottom and splash zones, involves the following system of equations [15]:

\[
P_{j}(r_{f}) \Delta r_{f} = \frac{W_{el,d}}{W_{el}} = \frac{F_{i}^{*} \Delta r_{f} + W_{el,i+1} r_{shrink(f)}(r_{f+1}) \frac{\Delta r_{f}}{\Delta r_{f+1}} + \sum P_{j}(r_{f}) W_{el,j} k_{f} r_{f}}{F_{3} \Delta r_{f} + W_{el} r_{shrink(f)}(r_{f}) + 3W_{el} r_{shrink(f)}(r_{f}) \Delta r_{f} / r_{f} + W_{el,j} k_{f} r_{f}}
\]

(11)

where \(F_{i}^{*} = F_{0,i} + F_{1,i} + F_{2,i}\)

The population balances of char particles burning in each compartment \(j\) of the transport region involves the following expression for the core region:

\[
P_{3,j}(r_{f}) \Delta r_{f} = \frac{F_{3,i-j} \Delta r_{f} + W_{cc,i-j} r_{shrink}(r_{f+1}) \frac{\Delta r_{f}}{\Delta r_{f+1}} + \sum P_{j}(r_{f}) W_{cc,j} k_{f} r_{f}}{F_{3,i} \Delta r_{f} + W_{cc,i} r_{shrink}(r_{f}) + T_{cc,i} \Delta r_{f} + 3W_{cc,f} r_{shrink}(r_{f}) \Delta r_{f} / r_{f} + W_{cc,j} k_{f} r_{f}}
\]

(12)

The solution of the population balances in the bottom+splash and transport regions allows for the determination of the carbon flow rates in all of the process streams [15].

4.5. Shrinking Rates. For the solution of char population balances, it is necessary to know the individual shrinking rate \(r_{shrink}(r_{f})\) of the char particles. Assuming the shrinking unreacted particle model, with mixed control by chemical reaction and mass transfer in the gas film and with a first order reaction, the shrinking rates of char particles are given by the expression:

\[
r_{shrink}(r_{f}) = \left( - \frac{dr}{dt} \right) = \frac{12C_{O_{2}}}{j_{c} \rho_{c} (1/k_{c} + d_{f}/ShD_{g})}
\]

(13)

The term \(C_{O_{2}}\) indicates the effective oxygen concentration seen by the char particles burning at any point of the combustion chamber. This concentration depends directly on the hypotheses of the type of gas flow in the combustion chamber and on the devolatilisation and volatile combustion considered. Therefore, the application of
equations to solve the population balances is not direct, and these balances must be solved at the same time as the oxygen profiles in the combustion chamber. In the bottom and splash regions of the riser, the Sherwood number ($Sh$) was calculated with the equation proposed by Palchonok et al. [26] for dense fluidised beds. In the core of the transport region, the equation proposed by Chakraborty and Howard [27] was assumed to calculate $Sh$.

Although the bed was considered isothermal, the temperature of the char particle was higher than the bed temperature. The char surface temperature was calculated by simultaneously solving the heat balance for a particle that transfers heat to the medium by both convection and radiation, together with the reaction rate [28].

The kinetic constants for pine bark and coal particles combustion were previously determined [29] and were shown in Table 2.

\[ k_c = k_a T_i \exp(-E_a / RT_i) \] (14)

The char shrinking rate for each particle size was calculated as a weighted mean of the individual values of the shrinking rates of the char coming from the biomass and the char formed from coal:

\[ -\frac{dr}{dt} = F_{\text{biomass},i} \left(-\frac{dr}{dt}\right)_{\text{biomass},i} + F_{\text{coal},i} \left(-\frac{dr}{dt}\right)_{\text{coal},i} \] (15)

Combustion of char was assumed to produce a mixture of CO and CO$_2$. The CO/CO$_2$ distribution was calculated according with the model proposed by Hannes [30].

The solution of the mathematical model implies the simultaneous convergence of the char particle population balances and the oxygen profile in the riser. As the oxygen concentration at each height depends on the char and volatile combustion, the mass balances for char, oxygen and volatiles were simultaneously solved [17]. The model
predicts the oxygen, CO, methane, propane, ethylene, hydrogen and water vapour concentration profiles along the different regions of the riser. The char concentration and char particle size distributions in the bottom, splash, core and annulus regions, the heat generation rates along the height of the combustion chamber and the carbon combustion efficiency are also predicted.

4.6 Modelling results. The model developed was validated with the experimental results obtained at the CFB boilers burning mixtures of pine bark and coals. Figures 4-9 show the model predictions when an operating condition was varied and the other variables were kept constant. As can be seen in those figures the predicted efficiencies and carbon concentrations are in good agreement with the experimental ones found in both installations and with both kind of coals.

Figure 10 shows a comparison between the experimental and predicted oxygen concentrations along the riser measured in the VTT pilot plant. As can be seen, higher oxygen concentrations along the combustion chamber have been measured than the predicted ones. Above the secondary air inlet (2 m) the experimental profile shows an increase of the oxygen concentration during one meter. This effect can be attributed to an insufficient penetration depth and mixing of the secondary air flow. Obviously these three-dimensional effects can not been explained by a one-dimensional model for the gas phase.

Finally, Figure 11 shows a comparison between the experimental \( E_c \) and those predicted by the model, including all the experimental results obtained in both pilot plants. In general, it can be observed a good agreement in the whole range of operating conditions used taking into account that the model does not use any adjustable parameters.

Once the model was validated it can be used to optimise the CFB co-firing of coal and biomass mixtures.
5. CONCLUSIONS

The performance of two CFB combustors (0.3 and 0.1 MWth) burning sub-bituminous and lignite coals mixed with pine bark was analysed. The effect of the main operating conditions, such as combustion temperature, percentage of biomass in the feed, air velocity, excess air, coal type, percentage of secondary air, and particle size distribution of coal on the combustion efficiency were studied.

A mathematical model for the prediction of carbon combustion efficiencies in circulating fluidised beds burning coal and biomass mixtures is presented. A good agreement between the carbon combustion efficiencies predicted by the model and the ones obtained in the two pilot plants was found without using any adjustable parameters.

ACKNOWLEDGMENTS

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NOMENCLATURE

a  Decay constant (m⁻¹)
C  Gas concentration (kmol m⁻³)
dₚ  Particle diameter (m)
Dₔ  Diffusivity (m² s⁻¹)
Eₐ  Apparent activation energy (J mol⁻¹)
E₀  Mean of the activation energy distribution (J mol⁻¹)
F₀₋₇  Carbon flow rates (kg s⁻¹)
F₀₋₇  Biomass flow rates (kg s⁻¹)
Fₐ  Fuel feed flow rate (kg s⁻¹)
F₈  Solids flow rate recovered in bag filter (kg s⁻¹)
FₐD  Solid drained flow rate (kg s⁻¹)
Fₐgas  Total flow rate of combustion gases (kg s⁻¹)
h  Height (m)
H₀  Height of the riser (m)
H₀  Height of the bottom region (m)
jₙC  Carbon fraction in the char
k₀  Backflow ratio
kₐ  Apparent kinetic constant for surface reaction (m s⁻¹)
kₐ  Fragmentation rate constant (m s⁻¹)
kₐ  Pre-exponential factor for surface reaction (m s⁻¹ K⁻¹)
kₐ  Pre-exponential factor for devolatilization reaction (s⁻¹)
K  Transport region decay constant (m⁻¹)
Nₔ  Number of fragments generated from a particle
Pₐ₋₇(r)  Normalized size distribution function of the char stream (m⁻¹)
Pₙ  Size distribution function on a mass basis of the fragments (m⁻¹)
r  Char particle radius (m)
r₀  Mean radius of particles in population i (m)
rₐmother  Radius of the mother particle (m)
rₐshrink(rₖ)  Shrinking rate of char particles of size rₖ (m s⁻¹)
(-rₖ)  Combustion rate of gas i (kmol m⁻³ s⁻¹)
R  Gas constant (J mol⁻¹ K⁻¹)
Sh  Sherwood number
t  Time (s)
Tₙ  Char surface temperature (K)
Tₙ  Flowrate of transmitted carbon from the core to the annulus (kg s⁻¹)
u  Superficial gas velocity (m s⁻¹)
u₀  Single particle terminal velocity (m s⁻¹)
W  Mass (kg)
xₐA  carbon fraction in the fuel feed
xₐC  carbon fraction in the recovered solids bag filter
xₐD  carbon fraction in the drained solids feed
xₐG  carbon fraction in the gas outlet

Greek symbols
Δrₖ  Size interval of population i (m)
γ  Relative fragment size
ρₐ  Average density of char (kg m⁻³)
ρ  Solids concentration (kg m⁻³)
ρₐ  Solids concentration in the bottom region (kg m⁻³)
ρₐb  Solids concentration due to the dispersed phase in the upper portion of the bottom region (kg m⁻³)
\( \rho_{\text{exit}} \) Solids concentration at the gas outlet (kg m\(^{-3}\))
\( \sigma_E \) Standard deviation in the activation energy distribution function (J mol\(^{-1}\))

**Subscripts**
- \( \text{cc} \) Carbon in the core
- \( \text{cl} \) Carbon in the bottom and splash regions
- \( i \) Relative to the population of char with average radius \( r_i \)
- \( j \) Relative to the differential element at height \( h \) in the transport region
- \( O_2 \) Oxygen
REFERENCES


### Table 1. Fuel composition

<table>
<thead>
<tr>
<th>Proximate analysis (wt % ar)</th>
<th>Pine Bark</th>
<th>South African Coal</th>
<th>Teruel Lignite</th>
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<tr>
<td>Moisture</td>
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<td>6.6</td>
<td>11.0</td>
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<td>3.6</td>
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<table>
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<th>Ultimate analysis (wt % daf)</th>
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<th>South African Coal</th>
<th>Teruel Lignite</th>
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<tbody>
<tr>
<td>C</td>
<td>46.2</td>
<td>67.7</td>
<td>49.2</td>
</tr>
<tr>
<td>H</td>
<td>4.9</td>
<td>3.8</td>
<td>4.6</td>
</tr>
<tr>
<td>N</td>
<td>0.5</td>
<td>1.8</td>
<td>0.6</td>
</tr>
<tr>
<td>S</td>
<td>0.02</td>
<td>0.5</td>
<td>6.5</td>
</tr>
<tr>
<td>LHV (MJ/kg daf)</td>
<td>19.9</td>
<td>27.6</td>
<td>20.6</td>
</tr>
</tbody>
</table>
Table 2. Kinetic parameters of the devolatilization model and the char combustion rates.

<table>
<thead>
<tr>
<th>Devolatilization rate</th>
<th>South African Coal</th>
<th>Teruel Lignite</th>
<th>Pine Bark</th>
</tr>
</thead>
<tbody>
<tr>
<td>$k_0$ ($s^{-1}$)</td>
<td>$10^{13}$</td>
<td>$10^{13}$</td>
<td>$10^{13}$</td>
</tr>
<tr>
<td>$E_o$ (kJ/mol)</td>
<td>235</td>
<td>235</td>
<td>205</td>
</tr>
<tr>
<td>$\sigma_E$ (kJ/mol)</td>
<td>35</td>
<td>35</td>
<td>25</td>
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</tbody>
</table>

<table>
<thead>
<tr>
<th>Char comb. rate</th>
<th>South African Coal</th>
<th>Teruel Lignite</th>
<th>Pine Bark</th>
</tr>
</thead>
<tbody>
<tr>
<td>$K_a$ (m/sK)</td>
<td>1.93</td>
<td>3.50</td>
<td>0.82</td>
</tr>
<tr>
<td>$E_a$ (kJ/mol)</td>
<td>72</td>
<td>71</td>
<td>66</td>
</tr>
</tbody>
</table>
Captions to illustrations

**Figure 1.** Schematic diagram of CIEMAT’s 300 kW CFB reactor.

**Figure 2.** Schematic diagram of VTT’s 100 kW CFB reactor.

**Figure 3.** Cumulative particle size distributions of the fuels. CIEMAT facility: 1-PB, 2-LT, 3-SA, 4-SA2. VTT facility: 5-PB, 6-SA.

**Figure 4.** Effect of percentage of biomass added in the fuel on the carbon combustion efficiency. CIEMAT facility (■LT/PB, ● SA/PB): T = 850°C, u = 5 ms\(^{-1}\), excess air = 25%, secondary air = 24 %. VTT facility (O SA/PB): T = 850°C, u = 2.3 ms\(^{-1}\), excess air = 30%, secondary air = 40%. Model predictions (______).

**Figure 5.** Effect of combustor temperature on the carbon combustion efficiency: ■LT/PB (F\(_{\text{biomass}}\) = 50 %wt), ● SA/PB (F\(_{\text{biomass}}\) = 60 %wt), u = 5ms\(^{-1}\), excess air = 25%, secondary air = 24%. Model predictions (______).

**Figure 6.** Effect of combustor temperature on the carbon concentration in bottom region: ■LT/PB (F\(_{\text{biomass}}\) = 50 %wt), ● SA/PB (F\(_{\text{biomass}}\) = 60 %wt), u = 5ms\(^{-1}\), excess air = 25%, secondary air = 24%. Model predictions (______).

**Figure 7.** Effect of linear gas velocity on the carbon combustion efficiency using two different particle size distributions: ● SA/PB (F\(_{\text{biomass}}\) = 60 %wt) ▲ SA2/PB (F\(_{\text{biomass}}\) = 60 %wt), T= 850°C, excess air= 25 %, secondary air= 24 %. Model predictions (______).

**Figure 8.** Effect of excess air on the carbon combustion efficiency: ●SA/PB (F\(_{\text{biomass}}\) = 60 %wt): T = 850°C, u = 5ms\(^{-1}\), secondary air = 24 %. Model predictions (______).

**Figure 9.** Effect of percentage of secondary air on the carbon combustion efficiency: ● SA/PB (F\(_{\text{biomass}}\) = 60 %wt): T = 850°C, u = 5ms\(^{-1}\), excess air = 25%. Model predictions (______).
**Figure 10.** Oxygen concentration profiles versus riser height in the VTT combustion chamber: ♦ SA/PB ($F_{\text{biomass}} = 46$ %wt.), $T = 850^\circ$C, $u = 2.3 \text{ms}^{-1}$, excess air = 30%. Model predictions (___).

**Figure 11.** Comparison between predicted by the model and experimental carbon combustion efficiencies. CIEMAT facility: ■ LT/PB, ● SA/PB, ▲ SA2/PB. VTT facility: ○ SA/PB.
Circulating fluidised bed co-combustion of coal and biomass, Gayan et al.

Figure 1
Circulating fluidised bed co-combustion of coal and biomass, Gayan et al.

Figure 2
Circulating fluidised bed co-combustion of coal and biomass, Gayan et al.

Figure 3
Figure 4
Circulating fluidised bed co-combustion of coal and biomass, Gayan et al.

Figure 5
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Figure 6
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Figure 7

![Graph showing the relationship between Ec (%) and u (m/s)]
Circulating fluidised bed co-combustion of coal and biomass, Gayan et al.

Figure 8
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Figure 9
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Figure 10
Circulating fluidised bed co-combustion of coal and biomass, Gayan et al.

Figure 11