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Effective velocity of biomass particles in a fluidized bed reactor

Master's Thesis - MSc programme in Innovative Sustainable Energy Engineering

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Department of Space, Earth and Environment
CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden 2019

MASTER’S THESIS 2019

Effective VELOCITY OF BIOMASS PARTICLES IN A FLUIDIZED BED REACTOR

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ABSTRACT

The growing greenhouse gas emissions (GHG) in the EU are produced mainly by the consumption of energy. For this reason, using more sustainable energy systems will contribute to reduce these emissions and then help to mitigate the climate change and the air pollution.

The CFB (circulating fluidized bed) technology has the capability of burning a wide range of fuels meeting SO_x and NO_x limits, which makes it an attractive energy system to achieve a successful efficiency of the combustion. However, to ensure an efficient conversion, it is necessary to achieve an effective mixing of the bed material and the fuel particles (controlled in axial and lateral directions). This control is not easy due to the chaotic behavior of the solid particles inside the furnace.

The present work investigates the behavior of the mixing of biomass particles (wood chips and wood pellets) when they circulate in a CFB when there is interaction with the inert particles and when there is not. For doing this investigation, the terminal velocities of each kind of biomass that were taken in the laboratory and the current existing models (by MATLAB) were compared.

Keywords: Terminal velocity, Wood chips, Wood pellets, circulating fluidized bed (CFB), particles interaction

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NOMENCLATURE

Abbreviations

BFB	Bubbling fluidized bed
CFB	Circulating fluidized bed
CO	Carbon Monoxide
CO_2	Carbon Dioxide
FBC	Fluidized bed combustion
GHG	Greenhouse gas
IEA	International Energy Agency
MPT	Magnetic particle tracking
NOX	Nitrogen oxides
PDF	Probability density function
SO_X	Sulphur oxides

Greek symbols

ρ_a	Density of active particles
ρ_i	Density of inert particles
ρ_g	Density of fluidization gas
$\rho_{b,c}$	Density of bed material in the core region
ε_c	Void of the dense phase
ϕ_s	Sphericity
τ	Constant coefficient
μ_g	Viscosity of fluidization gas

Latin symbols

a	Constant coefficient
$C_{D,g}$	Drag coefficient of the gas
d_a	Diameter of the active particle

d_p	Diameter of the particle
d_p^*	Dimensionless particle size
d_i	Diameter of the inert particle
d_{eff}	Effective diameter
d_{sph}	Equivalent spherical diameter
$F_{D,g}$	Drag force of the gas
G	Gravity number
h	Pellets length
K	Constant coefficient
m_a	Active particle mass
m_i	Inert particle mass
Δp	Pressure drop
$Re_{t,a}$	Active particle Reynolds number
t	Accumulation time
u_t	Terminal velocity
$U_{t,a}$	Active terminal velocity
Ut,i	Inert terminal velocity
u^*	Dimensionless terminal velocity
v_a	Active terminal velocity
V_{pellet}	Pellets volume
V_{sphere}	Sphere volume
X_A	Mass fraction
X_∞	Constant mass fraction
Δz	Height drop

1 INTRODUCTION

The share of renewable energy in the power sector, which has been rising quickly since the end of the 2000's, is expected to increase from 25% in 2017 to 85% by 2050, mostly due to the growth in solar and wind generation [1].

However, despite the strong growth in renewables, the overall share of fossil fuels in global energy demand in 2017 remains at 81%, a level that has been stable for more than three decades [2].

This demand of fossil fuels is mainly the responsible of the CO_2 emission rise, where energy-related emissions achieved historic highs of 32.5 gigatons (Gt) in 2017 [2] (Figure 1.1). According to the International Energy Agency (IEA) this energy use and CO_2 emissions will increase by a third by 2020 and nearly double by 2050 [3].

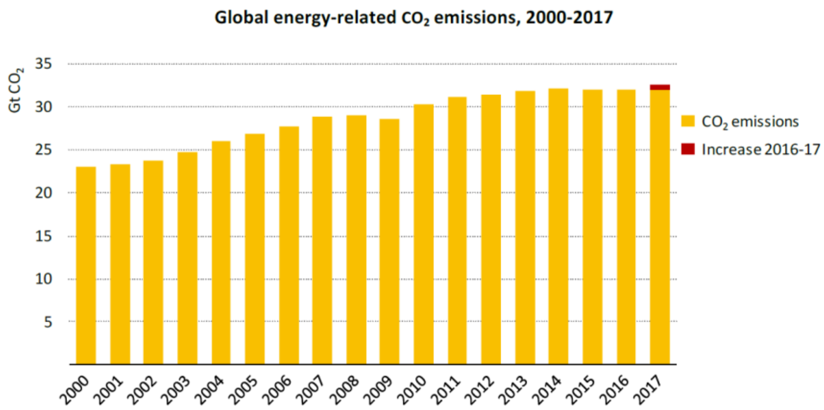


Figure 1.1. Global energy-related CO2 emissions and the 1.4% of emissions rose in 2017 [2]

Nowadays, energy accounts for 80% of all greenhouse gas (GHG) emission in the EU, which is the root of climate change and most air pollution [4]. For this reason, the decisions made today on the sustainability of future energy systems will influence our ability to mitigate climate change in the next three or four decades [3].

To stop with the climate change and to decrees our dependency on fossil fuel conversion, a clean and efficient combustion of renewable fuels is essential.

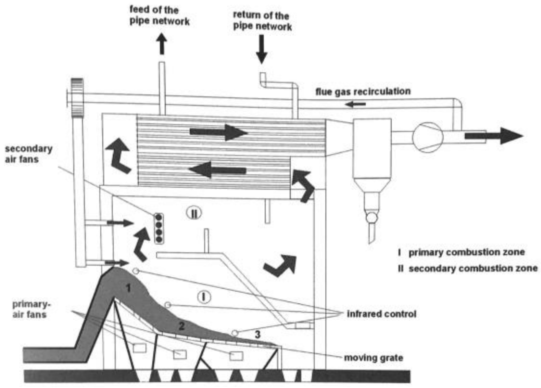
1.1 Background

As it has been said above, there is a high dependence of the fossil fuels as the primary energy supply. The fossil fuels include: coal, oil and natural gas. However, renewable energies are taking more place each year, and there is an increasing desire on using biomass as it produces negligible amounts of N_2O and is a CO_2-neutral energy resource. Moreover, the combustion of waste avoids landfilling [5].

There are several types of biomass (sawdust, wood pellets, bark, fiberboard, different types of waste, wood chips, etc.) which cover a wide range of fuels with different properties varying the combustion performance. The properties that differ from each biomass are the water content, bulk density, the gross calorific value and the particle size [6].

According to the differences in solid fuels, the combustion can be done in different units, such as pulverized coal, grate fired and fluidized bed units. To get a clean and efficient combustion, it is essential to choose the best technology which is more flexibility with these different kinds of biomass.

Pulverized coal units have the main purpose to increase the surface area of exposure to the combustion process, which results in faster and efficient combustion. The choice of this system depends on the boiler size, type of coal available, cost of coal, type of load, the load factor and availability of trained personnel [7]. This system has the advantages of having the highest combustion efficiency due to the high rate of heat release, easy control of the combustion, fast response to load changes, etc. However, air pollution from emissions of fine particles of git and dirt is an issue [7]. Another problem is the emission of nitrogen oxides, due to significant fractions of chemically bound nitrogen in the fuel [8], as well as thermal NO_x from high combustion temperatures. Further, pulverized coal units have a poor fuel flexibility, which is limited to different types of coal [7]. Eventually, the efficiency achieved with this system depends on whether hard or brown coal is used and on the coal quality [9].



Grate fired furnaces (

Figure 1.2) are mostly used as a conversion method to obtain heat and power from biomass. This system can deal with a big variety of biomass fuel types and is flexible regarding fuel size and moisture content. [10]

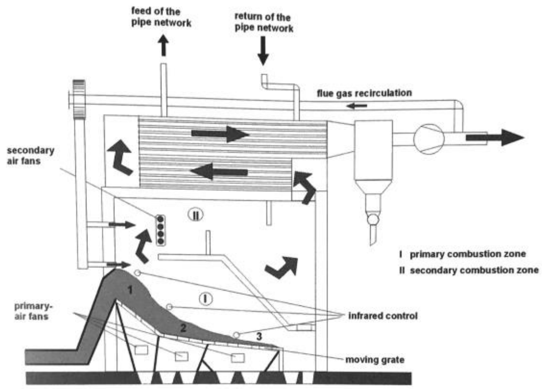


Figure 1.2. Schematic of a biomass grate furnace.[11]

Moreover, fixed grates are available for small scale combustion systems (<1.0 MW_{th})[6], and special grate constructions are able to cope with woody and herbaceous biomass fuel mixtures, e.g. vibrating grates (Figure 1.3) and rotating grates (Figure 1.4) [10]. Such grates have to ensure a mixture of the fuel across the grate.

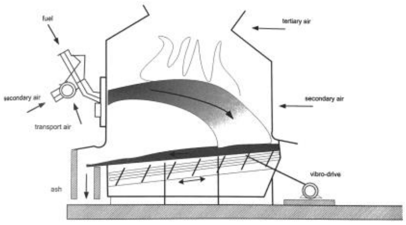


Figure 1.3. Vibrating grate fed by spreader stokers[11]

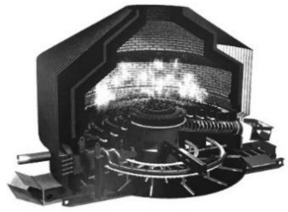


Figure 1.4. Underfeed rotating grate[11]

Finally, fluidized bed combustion (FBC) is a technology that includes two main combustion systems, bubbling fluidized bed (BFB) and circulating fluidized bed (CFB), and has been applied since 1960 for combustion of municipal and industrial wastes. This technology consists of a cylindrical vessel with a perforated bottom plate filled with a suspension bed of hot, inert and granular material (commonly silica sand or dolomite).

Firstly, the combustion air enters the furnace from below through the air distribution plate and fluidizes the bed. Following this state, the intense heat transfer and mixing provides good conditions for complete combustion with low excess air demands.

Due to the good mixing achieved, FCB plants can deal flexibly with various fuel mixtures (e.g. mixtures of different kinds of woody biomass fuels can be burned) but are limited when it comes to fuel particles size and impurities contained in the fuel.

According to the advantages and disadvantages from each technology, there are two main considerations that involve the choice of a boiler in an energy project. These two considerations to take account are: the availability of fuel and the local environmental regulations [12].

The FCB technology has the advantage of being fuel flexible and capable to burn low-grade fuels. In particular CFB with a high specific heat transfer capacity are an attractive option to generate steam for power generation in units greater than 30 MW_{th} meeting SO_x and NO_x limits [12].

Existing coal-fired CFB power plants can reduce carbon emission directly by partially submitting coal with carbon neutral biomass.

For these reasons, CFB is most suitable to achieve an efficient and carbon neutral energy conversion.

1.2 Problem statement

Woody biomass is a fuel which is not available in large quantities in the most countries, depending on the countries forest growth. Considering this situation and the increased tax of oil and coal, Sweden is well conditioned for the use of biomass, due to their large forest industry and no fossil fuel resources [13].

The main factors to obtain efficient combustion are the effective mixing of the bed material and the fuel particles. This mixing of biomass must be controlled in axial and lateral directions to ensure an efficient conversion.

However, this control is not easy due to the solid particles are floating in chaotic movements, and practically little work has been done in literature combining investigated the particles behavior inside the bed by both modelling and experiments. Until now, there are some papers that describes the fuel particles behavior without the interaction of the bed material in a FB, but not when the fuel particles interact with the bed material [14].

1.3 Aim

The aim of this master thesis is to investigate the behavior of the mixing of biomass particles (wood chips and wood pellets) in a circulating fluidized bed combining an experimental and modelling approach. The behavior is studied when the biomass particles flow inside the bed without any interactions and with the interaction of the bed material particles, which is going to be glass particles. The main reasons to choose this bed material is due to it does not produce fines and its sphericity is approximately 1.

Understanding how the terminal velocity affects the particles inside the bed is going to be crucial for this study. To achieve this goal, the project will include some further development of existing experimental methods and experiments in a scaled unit at the Energy Technology laboratory. These experimental results will help to see how truthful the existing models are.

1.4 Limitations

The possibilities of this study must be limited to carry out the work successfully. In order to limit the study, the experimental project part will be focus on cold conditions with a constant temperature and pressure to see the biomass particles behavior.

Another limitation is going to be the reactor scale. The scale will influence at the same time the bed geometry, which is going to be smaller and narrow and it will affect the particles behavior.

Moreover, such being a master’s thesis, the time will be another limitation to take account, in which the time spending in the experimental part will be adapted to this limitation.

2 METHODOLOGY

This thesis is based on theoretical concepts, experimental analysis and model implementation (Figure 2.1).

One of the main parts of the thesis is the comprehensive synthesis of different concepts and existing models. All the scientific databases consulted are Scopus, Science Direct, Google Scholar, aided by Chalmers Library database to have free access to this information and several books and reports.

On the other hand, an experimental part is needed as a source of new data about the behavior of the particles inside the bed.

The studied fuels in the experimental part are different kinds of biomass, like wood chips and wood pellets, which behavior is more complex and unknown.

To carry out this part of the thesis and to see how effective the energy conversion is, it is needed the access to newly developed and advanced measurement systems. Chalmers division of Energy technology content a scaled circulating fluidized bed connected to different transducers which measures the pressure along the reactor with the program LabView.

Using these measurement systems and analyzing the results taken in the laboratory, can give an idea about the behavior of the particles and how it could be for industrial scale applications.

Finally, a part of modelling is needed to achieve the aim of this work. This part is simulated by MATLAB to show which terminal velocity is suitable for the particles with different parameters (sphericity, particle size, fraction void, etc.). After these simulations, is important to verify if these terminal velocities obtained with MATLAB matches with the results of the experimental part.

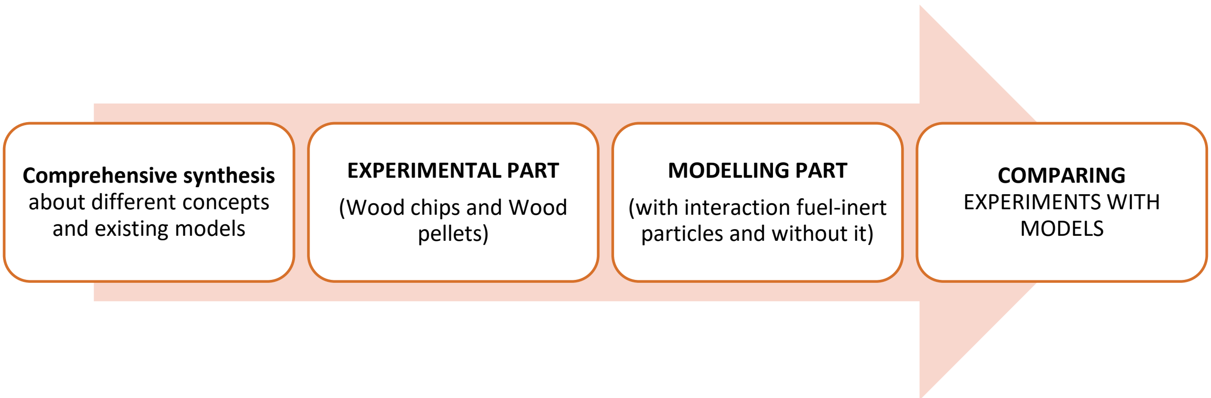


Figure 2.1. Scheme of the methodology followed in the project

3 THEORY

3.1 Fluid-bed technologies

Fluid-bed technologies can be divided in three different conditions (Figure 3.1).

The first one is when at low velocities the air does not exert much forces on the particles and they remain in place. This condition is called *fixed bed*.

When the air flow is increased further, the bed becomes less uniform, bubbles of air start to form, and the bed becomes turbulent. This is called a *bubbling fluidized-bed (BFB)*. The volume occupied by the air/solids mixture increases [15].

A further increase in air flow causes the particles to become entrained and elutriate the bed into the furnace. The solids are caught and separated from the air by a cyclone and returned to the bed, so that they circulate around in a loop. This condition is defined as a *circulating fluidized-bed (CFB)* [15], which is the technology used in this project.

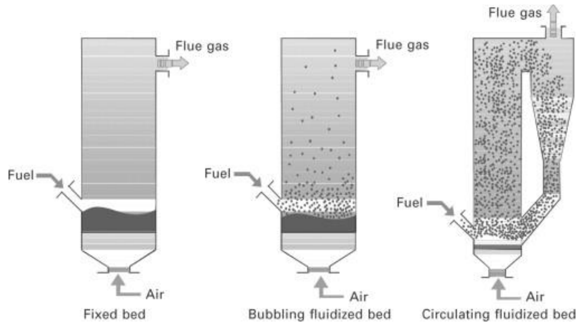


Figure 3.1. Schematics of fixed, bubbling and circulating FBC [16].

3.1.1 Circulating fluidized bed

The high fluidization velocity in CFBs increases the gas-solid mixing and allows for a higher burning rate [12]. The efficiency of a CFB (98-100 %) is mainly due hot unburned char particles are collected by the cyclone and continuously recirculation to the bottom of the furnace without cooling, increasing the residence time and allowing for complete conversion [17]. The only combustion loss is due to the escape of very fine char particles with the air flow out of the cyclone.

Inside the furnace the fine solids are transported at a velocity exceeding the terminal velocity. Nevertheless, the particles do not enter immediately into the vertical pneumatic transport system, but solids are found to move up and down in the form of aggregates, causing high degree of refluxing, which ensures uniformity of temperature in the furnace [12].

A schematic of this process is shown below in Figure 3.2, where the fuel is generally injected into to the lower section of the furnace. The fuel burns while mixed with the hot bed solids. While converting the fuel particles fall into smaller pieces, eventually following the motion of the bed solids.

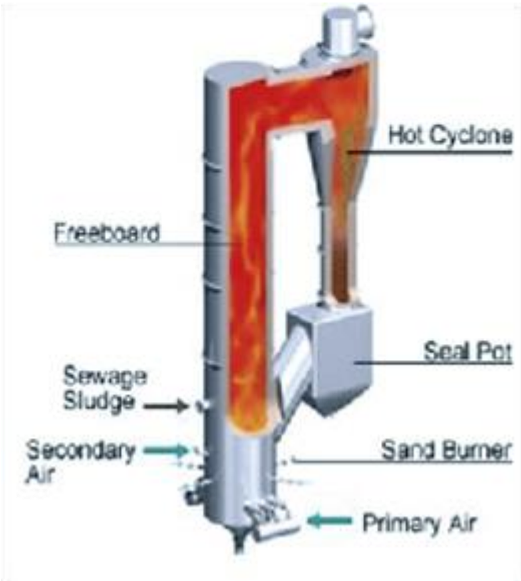


Figure 3.2. Process schematic for Circulating fluidized bed.[18]

The phenomenon that makes the fuel particles motion complex is the different terminal velocity of each particle when they interact with the inert particles. For this reason, studying the terminal velocity that makes particles recirculate will help to understand their behavior.

This behavior, in biomass case, differs in their different combustion stages.

3.2 Combustion stages

The process of biomass combustion involves a number of physical and chemical aspects of high complexity. The nature of this process depends on the fuel properties and the combustion application [10].

Biomass fuels generally have high moisture contents, which is between 50-65% by weight in fresh woody biomass [19]. The water starts to evaporate when the fuel is introduced into a hot environment and the conversion of the solid particle experiences four different stages: heating and drying, devolatilization (pyrolysis), combustion of gases, and char burning [20]. These stages are shown in Figure 3.3.

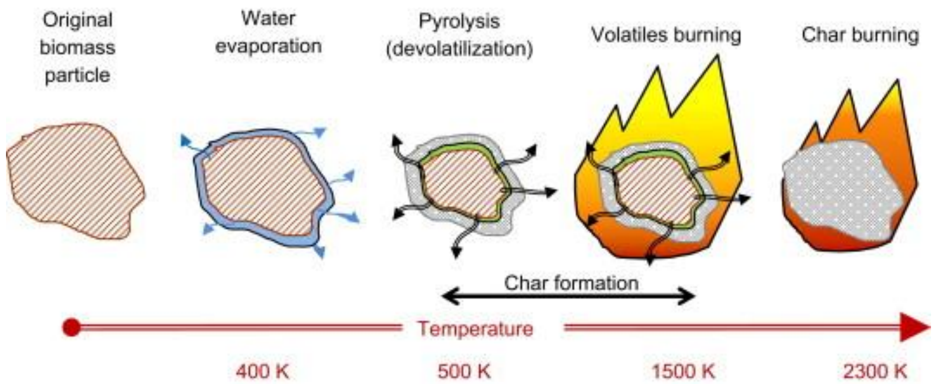


Figure 3.3. Stages in combustion of a solid biomass particle [21].

Drying and pyrolysis/gasification will always be in the first steps in a solid fuel combustion process. The importance of these stages will depend on the combustion technology implemented, the fuel properties and the combustion process conditions [10].

Each stage has its own characteristics that are defined a continuation:

I. Heating and drying is the stage where the moisture will evaporated at low temperatures (<100^{o}C)[10], so the heater dries the moisture from the particles' surface in a hot atmosphere by radiation or convection [20]. Burning these fuel particles generally constitutes around 1-3% by weight of the solids in the fluidized bed [12]. This process usually occurs very rapidly and much energy is required to evaporate the wet wood [10]. Consequently, moisture content is a very important fuel variable.
II. Pyrolysis can be defined as thermal degradation (devolatilization) in the absence of an externally supplied oxidizing agent. During this stage, volatile gases and tars are released from the particle and are combusted in the presence of oxygen [20].
III. Combustion of gases can ideally be defined as a complete oxidation of the fuel. The hot gases from the combustion may be used for direct heating purposes in small combustion units, for the water heating in small central heating boilers, to heat water in a boiler for electricity generation in large units, as a source of process heat, or for water heating in larger central heating systems [10].
IV. Char burning is the stage that generally starts after the release of volatile and it takes the longest time. The char reacts with oxygen to form mainly CO and CO_2. After the combustion is completed, an inorganic ash remains [12].

To obtain a successful combustion of solid fuel three main parameters have to be controlled: time, temperature and turbulences ("3 Ts of combustion").

Figure 3.4 shows several conversion stages differentiated with time and increasing temperature. As can be seen, as the temperature exceeds 900^{o}C within the furnace and is directly proportional to the residence time. If there is insufficient time in the furnace, the combustion reaction cannot proceed to completion and temperature declines.

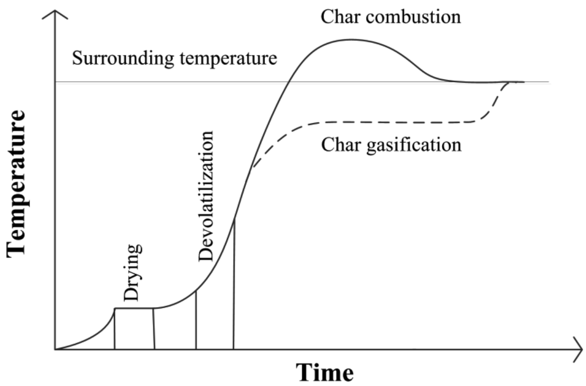


Figure 3.4. Different stages of biomass combustion versus temperature and time[22]

Due to with the complexity of a biomass particle, it is interesting to study its behavior inside the furnace of a fluidized bed when this one experiences all the combustion stages.

3.3 Particle terminal velocity

For understanding the fluidization phenomenon, it is important to know the basic hydrodynamic properties of a single particle called the free fall or terminal velocity, which relates to a balance of forces acting on a particle-gravity, buoyancy force and hydrodynamic resistance of a particle during motion [5].

This velocity determines the upper limit of the velocity range in which it is possible to maintain a fluidized state of a bed of particulate solids, as in a static state. It is important to know this terminal velocity while the FCB process is in elaboration, especially for the analysis of energy losses associated with unburned particles that are removed from the furnace [5].Moreover, the value of the terminal velocity helps the understanding and analyzing of the ash behavior in the furnace and why some particles leave during the recirculation in CFB.

The forces that act on a single spherical particle in an infinite space within the gravity field during free fall and without interactions with the bed material, are shown in Figure 3.5:

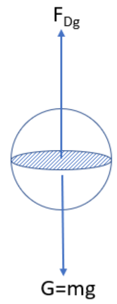


Figure 3.5. Representation of the acting forces on a particle without the interaction with the bed material

Where $F_{D,g}$ is the drag force of the gas and G is the gravity force. Taking these considerations, the forces balance can be expressed as the following:

$$\sum F = F_{Dg} - G = 0 \tag{1}$$

Where F_{Dg} and G are defined as:

$$F_{Dg} = C_{D,g} \frac{\pi {d_a}^2}{4} \rho_g \frac{{U_{t,a}}^2}{2} \tag{2}$$

$$G = m_a * g \tag{3}$$

Where $C_{D,g}$ is the drag coefficient of the gas, which value can be from different expressions. This affects the final value of the terminal velocity.

Depending on whether the sphericity, $\boldsymbol{\phi}_{s,}$ or the voidage of the dense phase, ε_c, is used, different expression describing $C_{D,g}$ exist.

When using the voidage of the dense phase, the drag coefficient of the gas, $C_{D,g}$, is expressed by the following equation:

$$C_{D,g} = \frac{\left(\frac{24}{R_{e_{t,a}}} + 0.44\right)}{{\varepsilon_c}^{4.75}} \tag{4}$$

In this case, when there is not interaction with the bed material particles, ε_c is 1 or lower.

On the other hand, if it is the sphericity what is taken into account, $C_{D,g}$ can be expressed as:

$$C_D = \frac{24}{Re_{t,a}} * \left[1 + \left(81716e^{-5.0748\Phi_s}\right) * Re_{t,a}{}^{0.0964+0.5565\Phi_s}\right] + \frac{73{,}69 * \left(e^{-5.0748\Phi_s}\right) * Re_{t,a}}{Re_{t,a} + 5.378e^{6.2122\Phi_s}} \tag{5}$$

If the sphericity, $\boldsymbol{\phi}_s$, is 1 the drag coefficient can be represented by the following equation:

$$C_{D,g} = \frac{\left(\frac{24}{Re_{t,a}} + 0.44\right)}{\varepsilon_c{}^{4.75}} \tag{6}$$

In all of the expressions of CD,g, the Reynolds, $Re_{t,a}$, is represented by Eq. (7):

$$R_e = \frac{\rho_a * v_a * d_a}{\mu_g} \tag{7}$$

If the fuel particle interacts with the bed material inside the furnace, a new force is added in the same way as F_{Dg}, this new force is the drag force of the solid (inert particle), F_{Ds}. The representation of this phenomenon is show in the Figure 3.6.

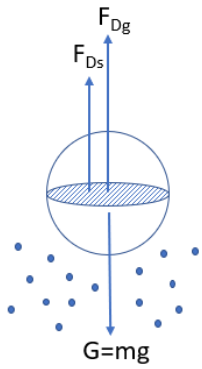


Figure 3.6. Representation of the acting forces in a particle with interaction with the bed material

The active char particle generally has different size and density from that of inert bed material. This results in a velocity difference between the active and the inert particles. Considering the model of Nowak et al. (1996)[23], in the situation showed above the interaction takes place between the spherical active particles with the parameters d_a, ρ_a and v_a and the surrounding particles, inert particles, with the uniform parameters di, ρ_i and v_i, describes a quasi-steady state motion of a coarse active article during the upward flow of a dilute, uniform suspension of gas and inert particles.

As a result of the additional momentum transfer, the velocity of the active particle increases compared to the case of a single-size particle flow. This results in a modification of Eq. (1), as it can be seen below:

$$\sum F = F_{Dg} + F_{Ds} - G = 0 \tag{8}$$

Where F_{Ds} is defined in Eq. (9):

$$F_{Ds} = \frac{2}{1+\frac{m_a}{m_i}} \frac{\pi(d_a+d_i)^2}{4} \rho_{b,c} \frac{(U_{t,a}-U_{t,i})|U_{t,a}-U_{t,i}|}{2} \tag{9}$$

And then the final force balance is shown in Eq. (10):

$$\underbrace{\frac{2}{1+\frac{m_a}{m_i}} \frac{\pi(d_a+d_i)^2}{4} \rho_{b,c} \frac{(U_{t,a}-U_{t,i})|U_{t,a}-U_{t,i}|}{2}}_{\text{1st term}} + \underbrace{C_{D,g} \frac{\pi {d_a}^2}{4} \rho_g \frac{U_{t,a}|U_{t,a}|}{2}}_{\text{2nd term}} - g m_a = 0 \tag{10}$$

In Eq. (10) there are two different terms: the first one describes the interaction of the fuel particles with the inert particles, and the second one the forces when they do not interact with the bed material.

In the first term we can differentiate the drag coefficient of the solid phase, $\frac{2}{1+\frac{m_a}{m_i}}$, and the particle area were the effective collision diameter take place, $(d_a+d_i)^2$. Further, the difference in velocity between the active particle and the inert particle of density $\rho_{b,c}$ [24] is used.

The second term describes the drag coefficient of the gas, where the drag coefficient, $C_{D,g}$ is calculated as it is shown in the Eq. (6). As it was said before, this coefficient interferes in the final terminal velocity obtained.

The concept of the terminal velocity or the free-fall velocity describes the situation when the particle of size d_p falls through a fluid. This steady-state can be estimated by Eq. (11):

Where C_D is the experimentally determined drag coefficient as it is shown in the Eq. (5), found by Haider and Levenspiel [25].

$$u_t = \left[\frac{4 * d_p * (\rho_s - \rho_g) * g}{3 * \rho_g * C_D}\right]^{1/2} \tag{11}$$

To calculate the terminal velocity, u_t, in this expression, the particle diameter, d_p, and the physical properties of the particle and the gas is needed. This can be related with by introducing a dimensionless particle size, d_p*, and a dimensionless terminal velocity, u_t*. These parameters are defined as the following:

$$d_p^* = d_p * \left[\frac{\rho_g * (\rho_s - \rho_g) * g}{{\mu_g}^2}\right]^{1/3} \tag{12}$$

$$u_t^* = u_t * \left[\frac{\rho_g^2}{\mu_g(\rho_s - \rho_g)g}\right]^{1/3} \tag{13}$$

Further, Eq. (14) can be used to directly estimate the terminal velocity of particles. This is an approximation of Haider and Levenspiel [25], using the equation suggested by Turton and Clark [26]:

$$u_t^* = \left[\frac{18}{\left(d_p^*\right)^2} + \frac{2.335 - 1.744\Phi_s}{\left(d_p^*\right)^{0.5}}\right]^{-1}, \qquad 0.5 < \Phi_s < 1 \tag{14}$$

Finally, if the terminal velocity is isolated from Eq. (13), equation (15) is obtained:

$$u_t = u_t^* * \left[\frac{\mu_g\left(\rho_s - \rho_g\right)g}{\rho_g^2}\right] \tag{15}$$

For both pellets and wood chips, the parameters needed for the model above are the sphericity (***ϕ***s), the density ($\boldsymbol{\rho}_a$), the particle diameter (d_p) and the voidage of dense phase in the top of the furnace (ε_c).

In order to find the voidage of the dense phase (ε_c) the solid concentration (C_s) in the top of the furnace versus the pressure difference between two pressure taps was obtained. The Bernoulli equation then defines the solids concentration depended on the furnace height:

$$C_s = \frac{dp}{dz} * \frac{1}{g} \tag{16}$$

What this equation defines is that when the furnace drop pressure decreases the solid concentration decrease, that means that in the top of the furnace this concentration is going to be small and then the ε_c bigger.

In order to find the voidage of the dense phase (ε_c) the following equation [27] can be used:

$$\frac{\Delta p}{\Delta z} = (1 - \varepsilon_c) * \left(\rho_{b,c} - \rho_g\right) * g \tag{17}$$

The pressure difference can then be obtained experimentally from two pressure taps with a known height difference Δz.

When modelling the terminal velocity of wood chips with the equations (12), (14) and (15) the particle diameter, d_p, is unknown. Therefore, a distribution of the terminal velocity of a work done by Nikku et al.[14] is used and translated into a diameter distribution. Once the wood chips d_p is taken for each sphericity, the u_t with interaction and without interaction with the bed material can be found using the models described above.

On the other hand, pellets have a more homogenous shape and their diameter (d_p) can be approximated with the effective diameter (d_{eff}) represented by the following equation [27],:

$$d_{eff} = \phi_s * d_{sph} \tag{18}$$

Where ***ϕ***s is the sphericity of the particle and d_{sph} is the equivalent spherical diameter, which is the diameter of the sphere having the same volume as the particle studied [27]. In this case the particles are cylindrical, so that d_{sph} can be defined as the following:

$$V_{pellet} = \pi * \left(\frac{d_p}{2}\right)^2 * h \tag{19}$$

$$V_{sphere} = \frac{4}{3} * \pi * \left(\frac{d_{sph}}{2}\right)^3 \tag{20}$$

$$V_{sphere} = V_{pellet} \tag{21}$$

$$d_{phr} = 2 * \left[\frac{{d_p}^2}{4} * h * \frac{3}{4}\right]^{1/3} \tag{22}$$

The wood pellets length (h) was found by measuring the length of each pellet in a random sample of the original batch of pellets used for the experiments.

The sphericity can be defined as the following [27], in which the sphericity can be isolated for the wood pellets case:

$$\emptyset_s = \left(\frac{surface\ of\ sphere}{surface\ of\ particle}\right)_{of\ same\ volume} \tag{23}$$

$$0 = {\emptyset_s}^2 * \left(h + \emptyset_s * \frac{d_{sph}}{2}\right) - d_{sph} \tag{24}$$

Taking the considerations above, the corresponding parameters for each material that are going to be use in the models are the following ones (Table 3.1):

Table 3.1. Wood chips and wood pellets properties

	d (mm)	h (mm)	d_{sph} (mm)	d_{eff} (mm)	$\boldsymbol{\rho}$ (kg/m^3)	$\boldsymbol{\phi}$s	$\boldsymbol{\varepsilon}_c$
Wood chips	Depending on the $\boldsymbol{\phi}$s	-	-	-	500 [28]	0.1:0.05:1	0.4576
Wood pellets	7	4:2:16	6.65-1.05	6.4-7.7	950 [28]	0.96-0.73	0.8546
Glass beads	0.075-0.12	-	-	-	2600	1	-

4 EXPERIMENTAL SETUP

4.1 Reactor's geometry and assumptions

To carry out the experiments a cold-flow model of a chemical looping combustion unit, consisting of an air reactor, a cyclone, a fuel reactor and a pot-seal, is used. The unit is operated as a CFB, this means with high fluidization in the air reactor and just enough fluidization in the fuel reactor and pot-seal to allow bed material to recirculate to the air reactor.

There are 22 pressure taps placed along the fluidized bed, as it can be shown in Figure 4.1, which have been calibrated prior to the experiments. Most of the taps are placed along the main reactor and each pressure tap has access to the chamber for pressure measurements at different heights of the bed. Mostly all the transducers measure the differential pressure between two pressure taps, except the first tap (Figure 4.1) and the taps connected to the cyclone and fuel reactor, whose transducers measure the pressure difference from the tap to the atmosphere.

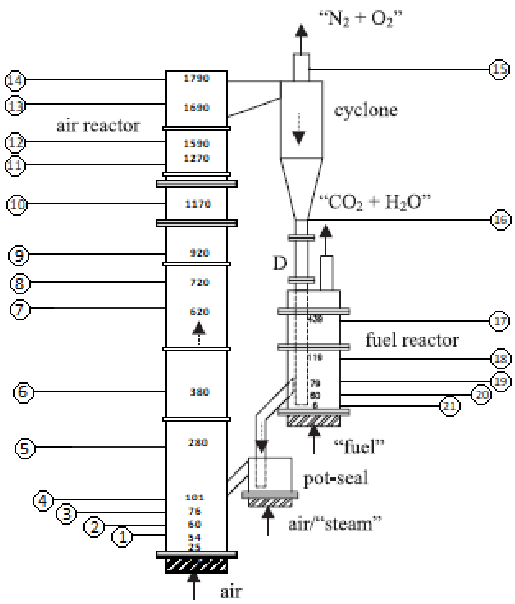


Figure 4.1. Principal layout of the model modifying the height pressure transducers position (represented by numbers) from [29] *D: downcomer*

Before starting with the experiments for checking the fluidization behavior, a test run with only bed material circulating has been done. For this the conditions described in a previous work has been used [29]. In Figure 4.2 it can be seen how the pressure drop decreases with the height of the furnace.

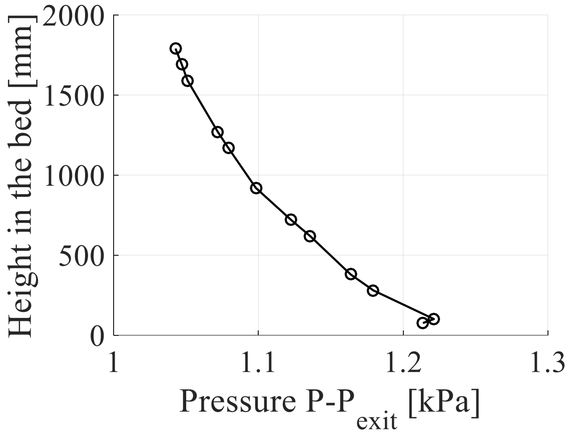


Figure 4.2. Pressure distribution along the furnace.

4.2 Material description

As it has ben said above, the investigated fuels were wood pellets and wood chips.

The wood chips are studied when they are dried (Figure 4.3). To study this kind of biomass the chips where preselected with two sieve sizes. This is to ensure that the chips fit into the TGA for pyrolysis. In order to maintain the same behavior of the wood in all the stages the portion of the finer sieve size is about 31,5% of the sample and the coarser sieve is about 68,5%.

Figure 4.3. Dried wood chips

As in wood chips case, the wood pellets (Figure 4.4) have been dried too. In this case, due to all the pellets have approximately the same size, the quantity taken was about the same volume that has been taken in the wood chips case.

Figure 4.4. Dried pellets

Regarding the bed material, it was glass beads with a diameter of 75-120 μm. The main reasons to choose this bed material is due to it does not produce fines and its sphericity is approximately 1.

The following table shows some properties of the fuels used:

	Total weight (g)	Drying temperature (°C)	Moisture content (%)
Wood chips	139.26	130	41.3
Wood pellets	422.93	105	7.14

Table 4.1. Properties of the materials

4.3 Process description

The process to study the different samples in the CFB was the following:

1. Fill the furnace with the bed material and the fuel particles that are going to be studied. The amount of bed material in the fuel reactor is given by the position of the overflow pipe, whereas the amount of bed material in the riser will be a function of total solids inventory and fluidization conditions [29].
2. Recirculate the bed material with the lowest velocity saving the pressure results taken of each tap with the program Labview. The estimated time taken in each recirculation is approximately 3 minutes (180 seconds).
3. If there are not fuel particles gotten in the extension arm after the cyclone that has a net to collect the particles (Figure 4.5), the velocity should be increase in the next recirculation.

Figure 4.5. Extension arm after the cyclone with a net that collect the particles

4. If fuel particles are collected, the velocity is maintained, and the fuel particles collected in the extension arm to be weighted. That velocity is going to be the terminal velocity of the fuel particles.
5. The process ends when no more fuel particles are collected due to the small sample that it is still inside the reactor or when the quantity of particles taken is constant for the same velocity over a longer period.

4.3.1 Wood chips

For the wood chips the process was repeated 93 times starting at a flow of 1600 l/min (0.94 m/s) until there were very little fuel particles found in the net after the cyclone.

The particles were of the both sizes, that were introduced in the furnace, meaning that in this case the terminal velocity was the same for sizes.

After 85 recirculations the portion taken was very low, so that the flow was increased to 1800L/min (1.06 m/s). When the flow was increasing the portion taken increased, but after five recirculations the portion taken was approximately 0 again. For this reason, although there were still wood chips of the initial sample in the furnace, the study of the sample finalized and it was approximated that the total mass of the sample can be taken at one velocity, 0.94 m/s (1600 L/min).

In total 110 g were taken from a sample of 139.26g (≈79%).

The mass fraction vs. time accumulation can be seen in Figure 4.6. It can be seen that the mass fraction increases as a negative exponential equation until a maximum valuen, meaning that at this maximum the number of total mass taken will be constant over time.

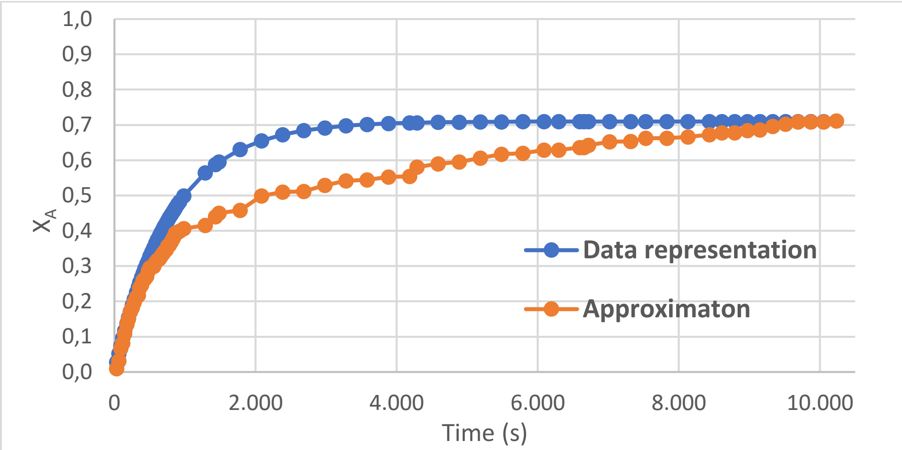


Figure 4.6. Mass dried wood chip fraction representation vs. accumulated time

The data taken from the experiment, orange line in Figure 4.6, was approximated with

the following equation (orange line):

$$X_A = K \cdot \left(1 - e^{\frac{-t}{\tau}}\right) \tag{25}$$

Where X_A is the mass fraction, K is the constant value of mass fraction that will not change with the time, t is the accumulation time and τ is a constant coefficient with value 813.

After representing the mass fraction vs. time accumulation of this sample, the data was divided into two parts and represented again as the mass fraction vs. accumulation time, Figure 4.7.

This division was done because of the inconvenient of the bed material accumulation in the cyclone. This accumulation makes that in high velocities the simulation had to be stopped after 30 seconds due to the issue that the whole cyclone becomes full of bed material. To solve this problem, a pressure tap was inserted in the middle of the cyclone from which flows air. Thanks to this tap the bed material could continue its recirculation during the three minutes of the simulation.

The division shows the data before and after the introduction of the pressure tap at the cyclone.

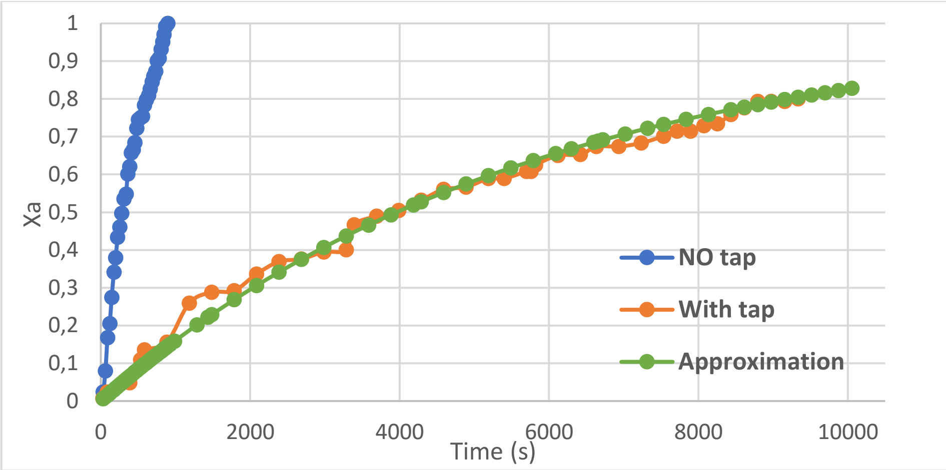


Figure 4.7. Mass fraction representation vs. time accumulation with and without the tap pressure in the cyclone

Where the green line in the Figure 4.7 is again an approximation that follows Eq. (25). In this case the value τ is 5714.3.

Considering these results, the air tap is used for all the following experiments.

4.3.2 Wood pellets

In the case of the wood pellets, at the air flow of 1600 L/min (0.94 m/s) a small amount of pellets can be found on the other side of the bed. However, after approximately 10 recirculations the quantity of pellets decreases considerably to almost 0 (Figure 4.8), so that the air flow was increased to 1800 L/min (1.06 m/s) (Figure 4.9).

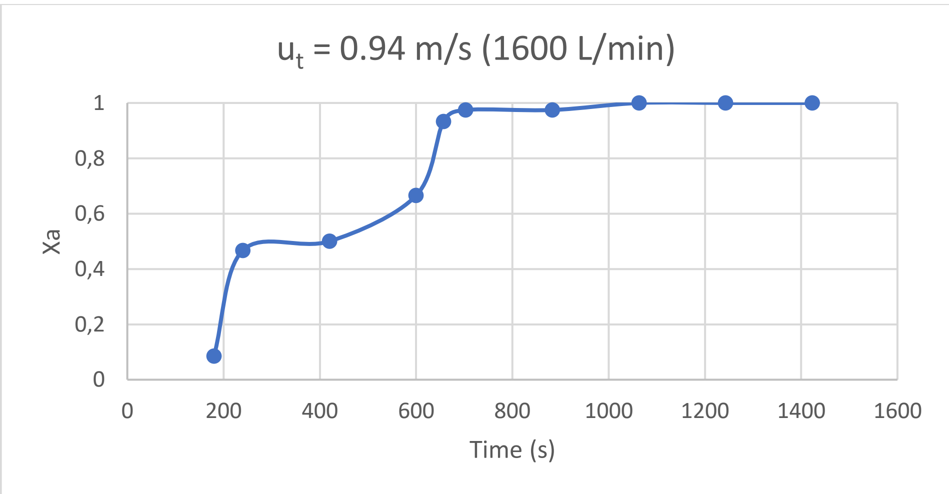


Figure 4.8. Mass dried wood pellets fraction representation vs. time accumulation when the u_t is 0.94 m/s

However, this velocity increasement did not affect the quantity of pellets exiting the furnace, and this quantity is still less that the quantity of wood chips that was taken. This observation can be related to the shape and weight of the pellets. The pellets characteristics make them harder to elevate, due to wood pellets are denser and more homogeneous in physical properties than wood chips [30].

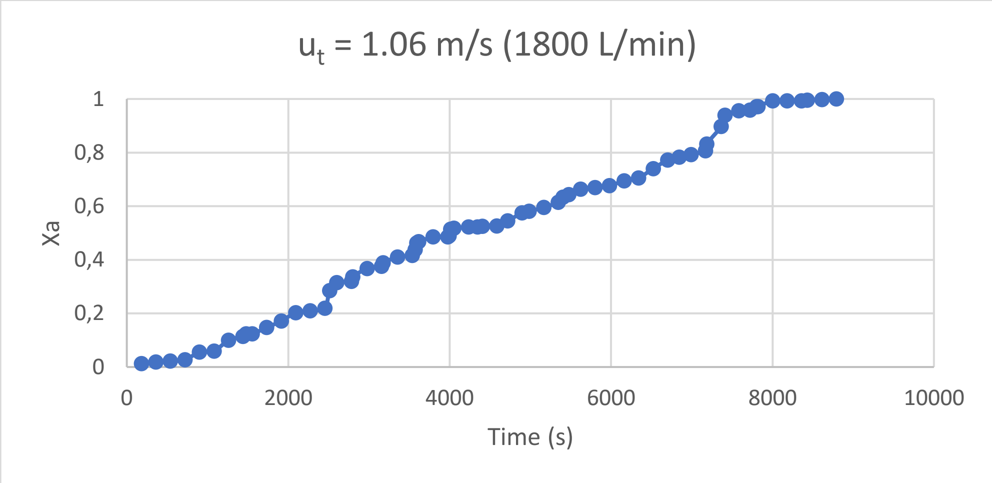


Figure 4.9. Mass dried wood pellets fraction representation vs. time accumulation when the ut is 1.06 m/s

After some recirculations the quantity of pellets left was the same and the decision was to increase again the velocity to 1.18 m/s (2000 L/min). At this point, the pellets that left did not converge to a point. For this reason, the results where extrapolated to find the point in where the quantity of mass that leave from the furnace was constant (Figure 4.10). This extrapolation line, orange line in the Figure 4.10, was defined by Eq. (26):

$$X_A = X_\infty - X_\infty * e^{-a*t} \qquad (26)$$

Where X_∞ is the mass fraction in which the exponential line converges to a constant value of 0.000264 in this case, and t is the accumulation time.

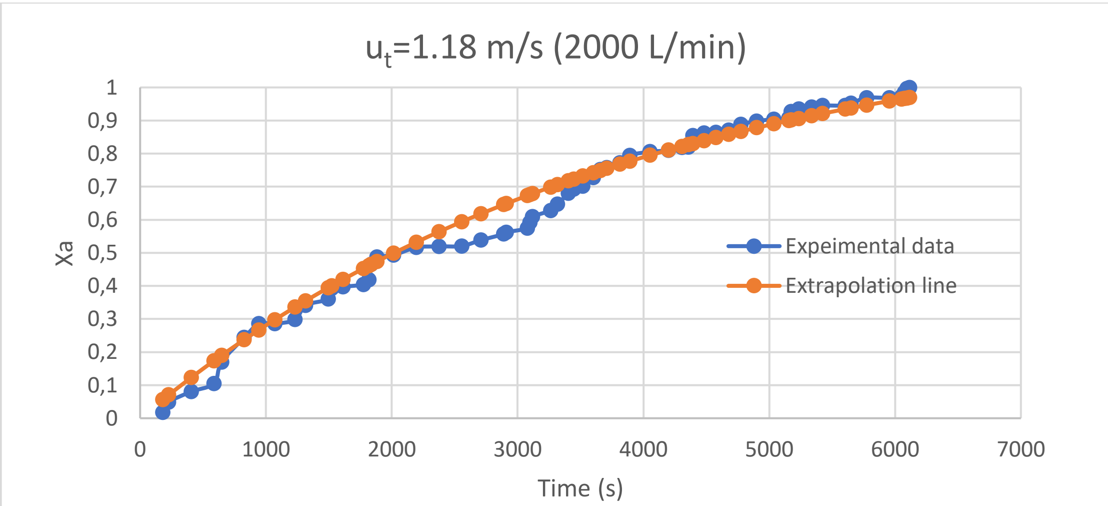


Figure 4.10. Mass dried wood pellets fraction representation vs. time accumulation when the u_t is 1.18 m/s and its extrapolation line

Finally, with the extrapolation the mass of pellets left (X_∞) at the velocity of 1.18 m/s was around 189g.

Summarizing all the quantities elutriated in each velocity (5.5 g at 0.94 m/s, 81 g at 1.06 m/s and 189 g at 1.18 m/s) it was 275.5 g of a total sample of dried pellets of 422.93g. That means that approximately the 65% of the sample had left with a maximum velocity of 1.18 m/s.

4.4 Laboratory issues

During the different experiments done in the laboratory, some issues appeared in connection with the equipment. The most significant ones that could affect the results are: the slugging phenomenon caused by the static electricity, small leakages, the bed material lost to the filter due the accumulation of this bed material in the cyclone and the accumulation of the bed material in the arm that connects the fuel reactor and the particle-seal.

In a given bed, the size of the bubble increases as the fluidizing velocity or the bed height is increased. Taking this into consideration, if the bed is small in cross section and deep, the bubble may increase to a size comparable to the diameter or width of the reactor. When this happens, the bubble passes through the bed as a slug, producing the slugging [12]. This phenomenon affects the recirculation of the bed material in the bed, as it can be seen in Figure 4.11, due to that, it takes time for the bed material to flow down and the experiment has to be stopped eventually.

Figure 4.11. Slugging phenomenon in the cyclone

On the other hand, there is bed material that flows to the filter. This happens because the bed material is accumulated in the cyclone at high velocities, due to the scale of the cyclone, and it cannot do its common dynamic flow.

Regarding the leakages of bed material, a main leakage can be found at the gas exit of the cyclone to the filter which can affect the pressure result in the reactor.

Finally, the bed material accumulated in the arm between the fuel reactor and the particle seal, Figure 4.12, results in less bed material available in the air reactor, hence, there is less bed material to interact with the biomass that is still in the furnace. The particles found in the net of the extension arm are den much less.

Figure 4.12. Bed material accumulated in the arm between the fuel reactor ant the post-seal

5 RESULTS, ANALYSIS AND DISCUSSIONS

First of all, the concentration in the reactor using the pressure drop between the different tap along the reactor was analyzed. This pressure information taken can be shown in the following figures for both cases, wood chips (Figure 5.1) and wood pellets (Figure 5.2):

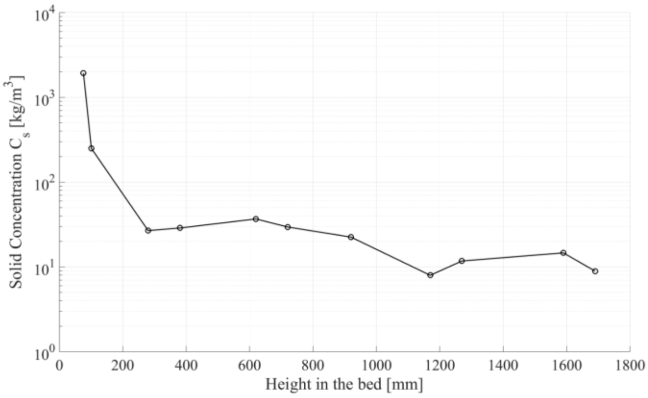


Figure 5.1. Solid concentration vs. furnace height whit the dried wood chips sample

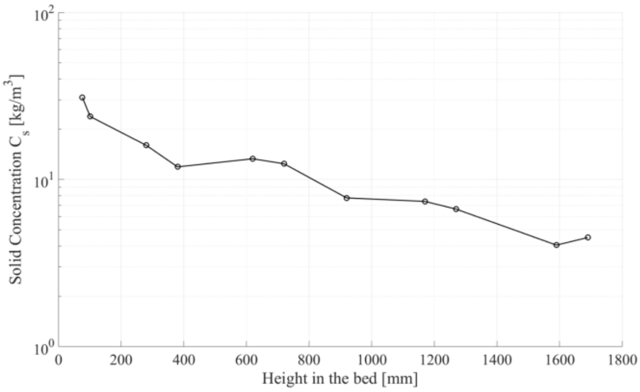


Figure 5.2. Solid concentration vs. furnace height whit the dried wood pellets sample

Both figures represent what was define in the Eq. (16), in which the solid concentration decreases when the furnace height increases. Further this means, that in the top of the furnace, the voidage of the dense phase is going to be high (due to the low solid concentration) compared to the bottom of the furnace.

5.1 Wood chips

The following scheme (Figure 5.3) visualizes how the results of the wood chips were obtained in the experimental and model part:

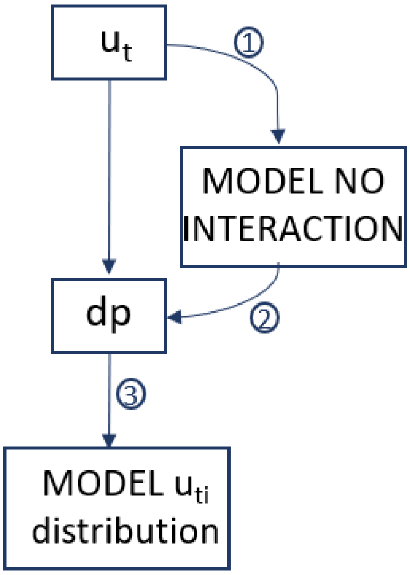


Figure 5.3. Schema of the process followed to analysed and discuss the model and experimental part

Due to the complex and heterogenous geometry of the wood chips, the particle diameter distribution cannot be obtained directly from observing the particles. Instead the first step is to obtain the diameter of this fuel particles from the experimental terminal velocities from a work done by Nikku et al.[14] when there is no interaction with bed material. The distribution of these terminal velocities is shown in Figure 5.4.

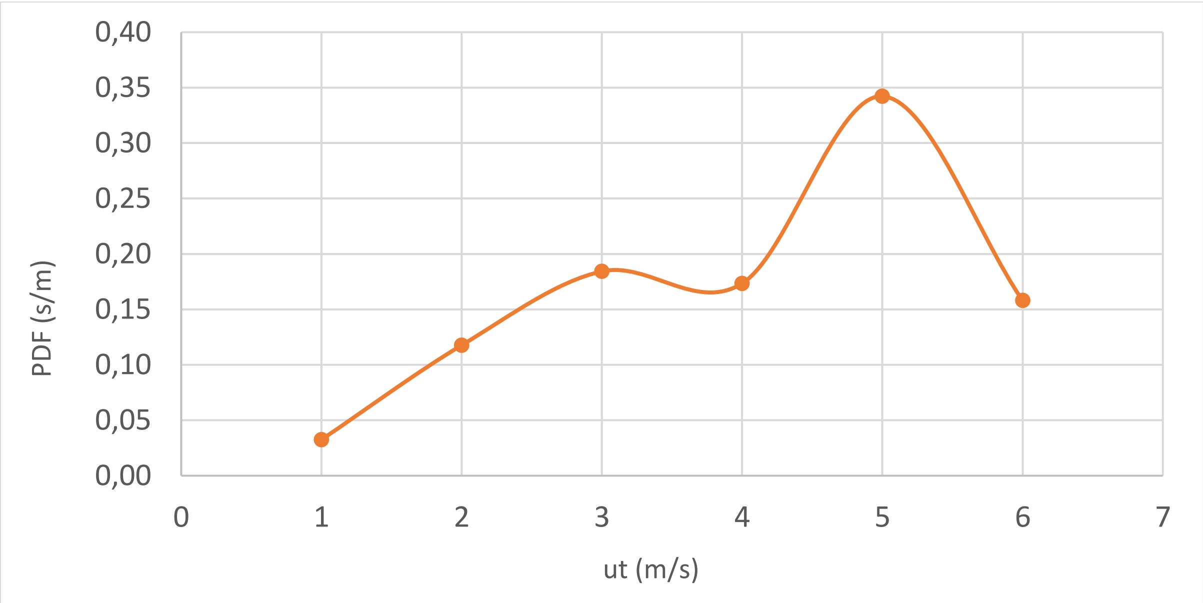


Figure 5.4. Terminal velocity distribution obtained from the work done by Nikku et al.[14]

Once the terminal velocities with no interactions are obtained, the diameter is calculated with the model (step 1 in Figure 5.3), considering different cases of sphericity, as it can be seen in Figure 5.5, where there are different ranges of diameter per sphericity ranging from 0.1 to 1.

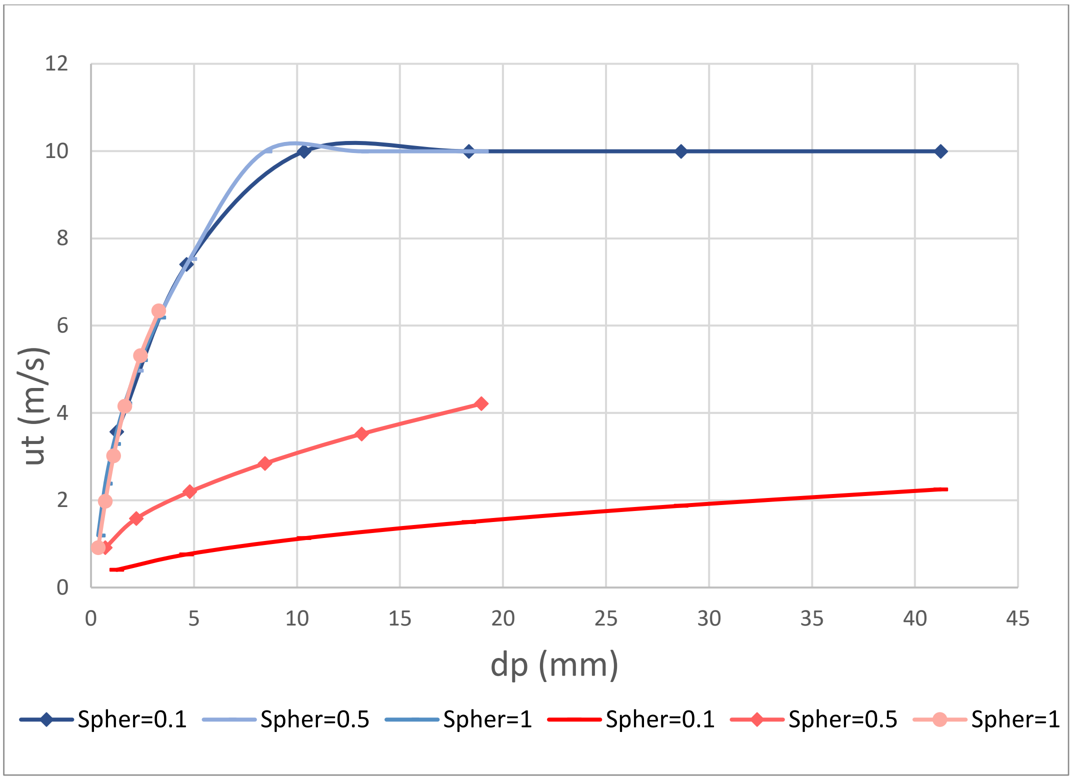


Figure 5.5. ut vs. dp when there is NO interaction

In Figure 5.5 it can be differentiated between blue and red lines. The red lines are obtained using Eq. (5) for calculating the drag coefficient C_D. This equation considers the sphericity and for this reason, there is a line represented for each sphericity. Moreover, when the sphericity is higher, the range of d_p is higher and the range of terminal velocities is lower.

In the case in which the equation used for calculating the C_D was Eq. (6), the blue lines in the Figure 5.5, the sphericity does not have relevance and the ranges obtained for each sphericity are mostly the same.

After the size of the wood chips is obtained for each sphericity, the distribution of the particles' diameters can be calculated, as can be seen in Figure 5.6.

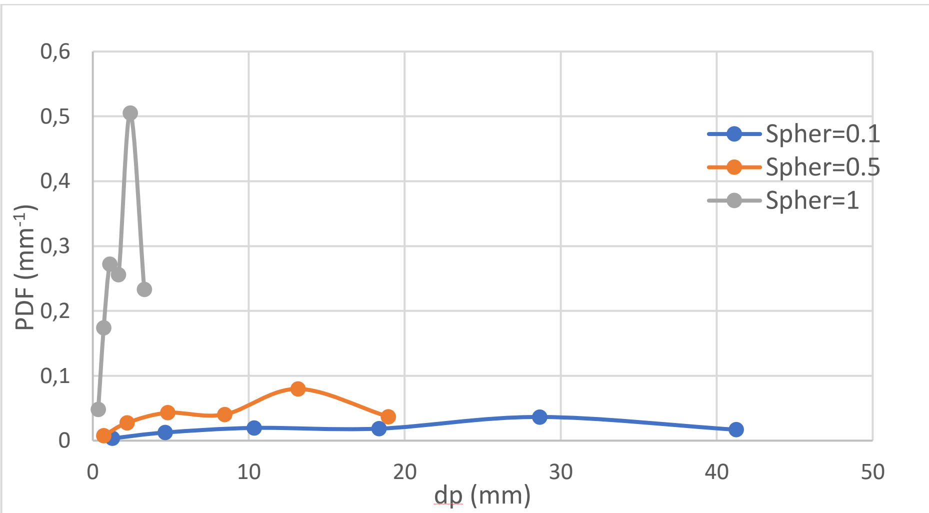


Figure 5.6. Wood chips diameter distribution

This diameter distribution can now be used to see the behavior of the wood chips terminal velocity, using the model when there is interaction, Figure 5.7.

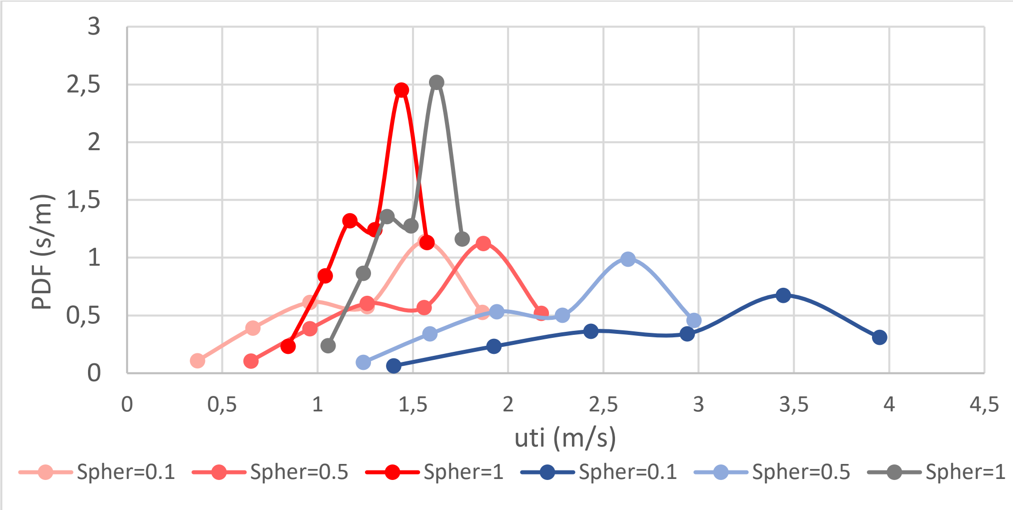


Figure 5.7. Wood chips terminal velocity using the model when there is interaction when the sphericity change. Representation of the terminal velocity calculating the CD with different equations.

Figure 5.7 shows that when the C_D is calculated with the equation Eq. (5) the range of terminal velocity is narrower, but in both cases, using the Eq. (5) or Eq. (6), due to the range of terminal velocity is narrow it can be considered that the size does not has influence in the final terminal velocity of the wood chips.

After obtaining the distribution of terminal velocities when there is interaction, the comparison between the model and the experimental results can be done (Figure 5.8).

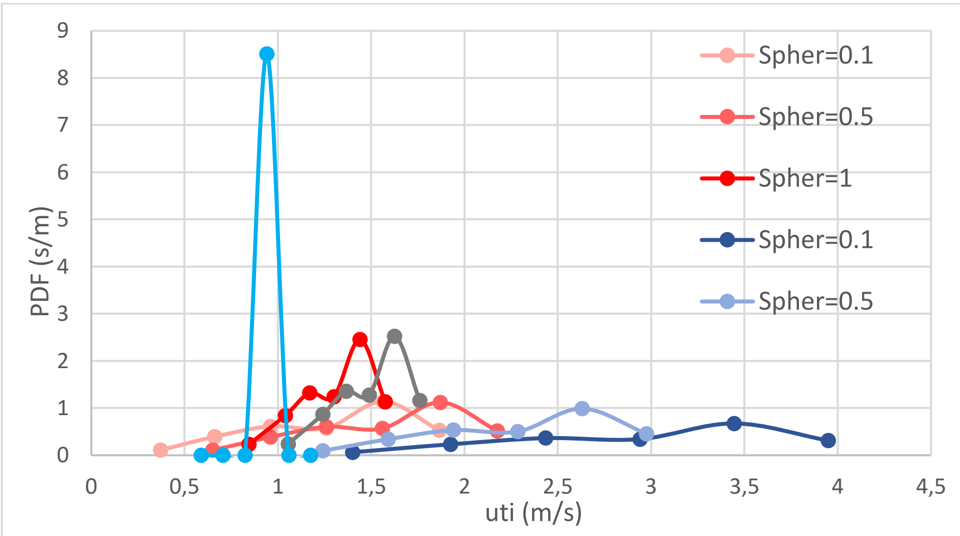


Figure 5.8. Comparation between model and experimental results

This comparison (Figure 5.8) shows that when the equation used for calculating the C_D is the Eq. (5) the magnitude of terminal velocities is closer to the experimental one. Moreover, the model gives results very different from the experimental result. However, what is seen is that in both cases (experimental and model result) the terminal velocity does not depend on the size of the wood chips due to the range of terminal velocities obtained is narrow.

Finally, and to see the effect of the inert particles, Figure 5.9 shows the comparison between the experimental result of terminal velocity when there is interaction and when there is none.

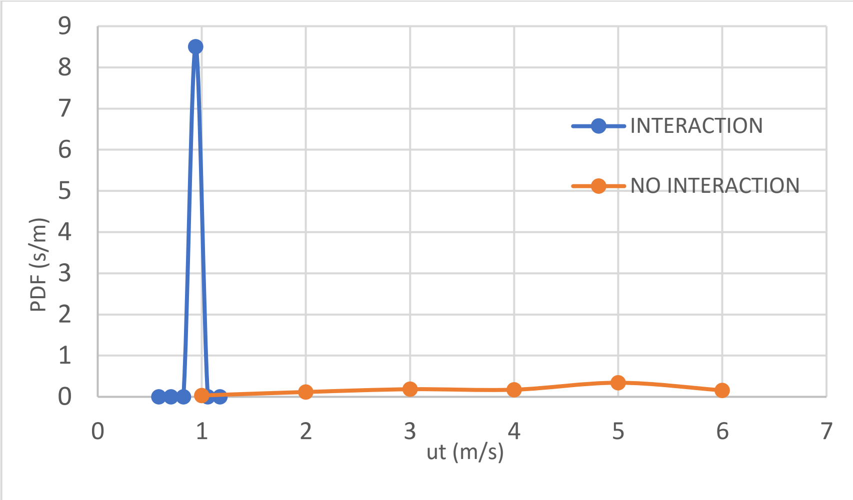


Figure 5.9. Comparation between experimental terminal velocities results when there is interaction and when there is no

This experimental comparison (Figure 5.9) shows that when the bed material interacts with the fuel particles in a circulating fluidized bed the terminal velocity of all the fuel particles (all the sizes) is mostly the same, and they can be left from the furnace at the same velocity. Contrary, when there is not bed material that interacts with the fuel particles the terminal velocity change for each fuel particle size.

5.2 Wood pellets

The wood pellets have a more uniform geometry, so that the pellets length (h) can be found by measuring the length of each pellet in a random sample of the original batch used for the experiments. Their length distribution can be seen in Figure 5.10.

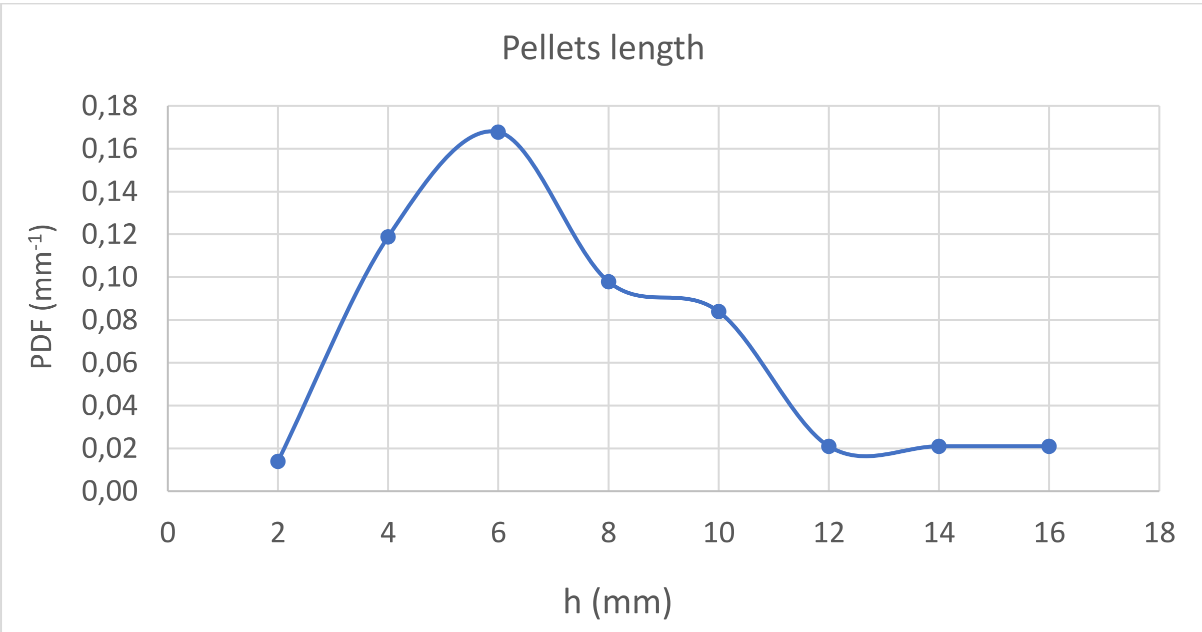


Figure 5.10. Wood pellets length distribution from a random sample.

As can be seen in Figure 5.10, the length that predominates in the sample is around 4-8 mm, which represents 31% of the sample.

Once the particle diameter (d_p) is obtained, the representation of the terminal velocity distribution considering interaction can with the bed material can be done with the model (Figure 5.11).

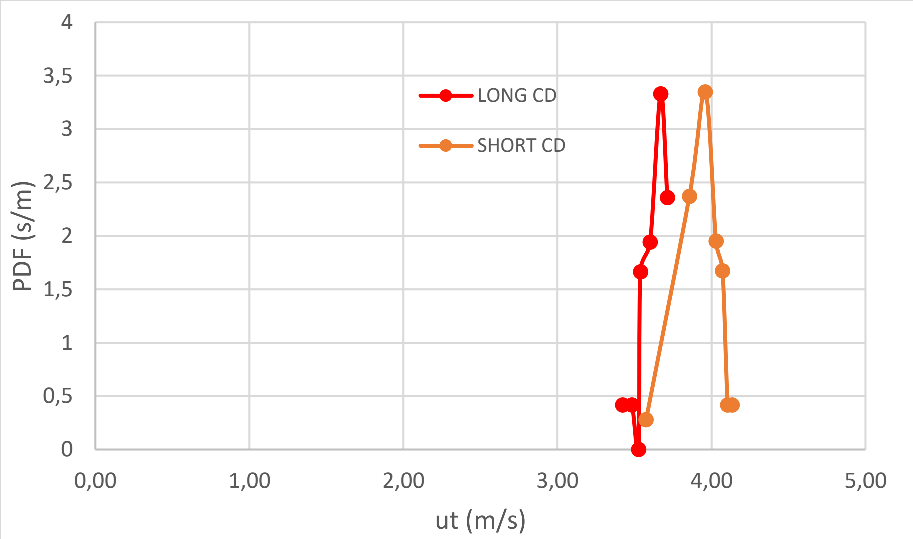


Figure 5.11. Model terminal velocity distribution of the wood chips when there is interaction

The Figure 5.11 shows that mostly all the pellets have the same terminal velocity being independent of their size. This observation can be seen due to that the range of the terminal velocity is very narrow. Moreover, when the sphericity is considered (using the Eq. (5)) the range is narrower then when it is not considered.

After observing the behavior of the wood pellets with the model the results can be compared with the experimental results. In the Figure 5.12 this comparation can be seen.

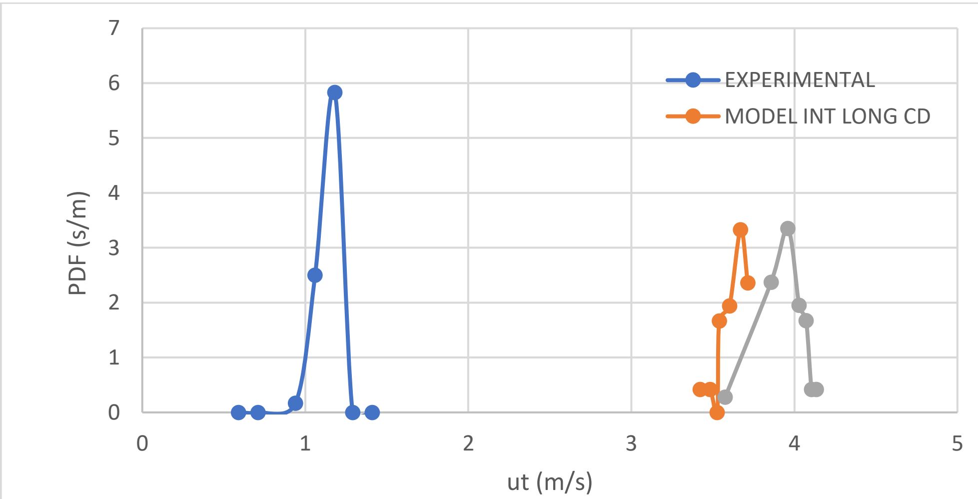


Figure 5.12. Comparation of the wood pellets experimental and model results

The Figure 5.12 shows that, as it happened in the wood chips case, the model results are far from the experimental ones, overestimating the terminal velocity. However, in both methods the range of the terminal velocity is narrow, what means that mostly all the pellets of the sample have the same terminal velocity independently of their size.

Finally, and to see how the relevant is the action of the bed material in the results, the comparation between the model results when there is interaction and no is necessary (Figure 5.13).

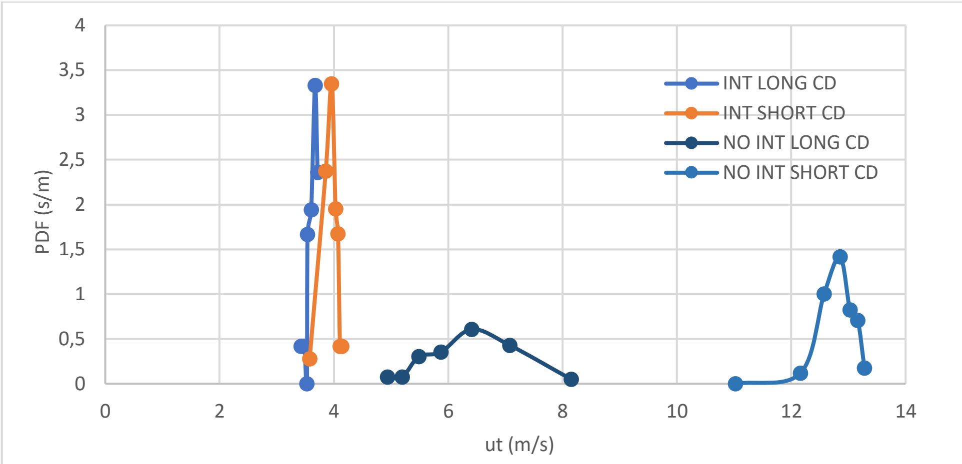


Figure 5.13. Comparation between the wood pellets model results when there is interaction and when no.

In the Figure 5.13 it can be seen that without interaction the size is relevant for the terminal velocity, due to there is a wide range of terminal velocities (one terminal velocity per size). Moreover, when the equation used for the calculation of the C_D is the Eq. (5) the range of terminal velocities is lower than when the equation used is Eq. (6).

Whereas, when there is interaction with the bed material, as it was found, the size does not influence the final terminal velocity.

6 CONCLUSIONS

Firstly, after doing the simulations in the laboratory, in where the fuel material interacts with the inert material, it has been seen that in both cases (wood chips and wood pellets) mostly the whole sample left from the furnace at the same terminal velocity. In the wood pellets case, the terminal velocity was higher than for the wood chips. This is because the shape and density have an important influence on the particle terminal velocity, hence the pellets characteristics make them harder to elevate.

Throughout the course of the thesis and comparing these conclusions of the experimental part with the model, it can be concluded that the model represents qualitatively the same particles behaviour that in the experiments. This means that the model represents that mostly the whole fuel particles sample has the same terminal velocity being independent of the particle size when there is interaction between fuel particle and the bed material in both samples.

When there is not interaction, in the wood chips sample, both methods (experimental and model) gives the same results regarding that in both methods they give a wide range of terminal velocities depending on the fuel particle size. When there is not interaction the particle size gets an important role.

However, when comparing the results of both methods quantitively, the model results give higher levels of terminal velocities compared with the results taken in the laboratory. The model is overestimating compared with the values found experimentally.

6.1 Future development

In order to provide a more detailed study on the fuel particles (wood chips and wood pellets) behaviour in a circulating fluidized bed, i the experiments in the different combustion stages of each fuel needs to be carried out. This can contribute to understand how the changes in density and size affects the behaviour of each particle.

Further, experiments when there is no interaction with the inert particles are needed, to be able to compare with the experiments with interaction in each combustion stage directly, instead of taking the data from literature.

Finally, to see how the fuel particles behaviour change when using different bed materials is of interest. By doing experiments with bed materials with different particle densities and sizes could tell how they influence the fuel particles behaviour when there is interaction with the bed material.

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