FLUID-DYNAMICS OF A COLD MODEL OF A FLUIDIZED BED GASIFICATION SYSTEM WITH REDUCED TAR CONTENT

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ABSTRACT: A new gasification concept for biomass/waste gasification for electricity production for small/medium application has been developed. The gasification system is aimed at improving the two major drawbacks in fluidized beds (FB), i.e. the low char conversion and the high tar content in the gas. The system comprises a FB devolatilizer, an air/steam reformer of the gas coming from the devolatilizer, and a chemical quench of the unconverted char removed from the FB. The thermal control of the various zones of the system, a key use for design, depends on the movement of the gas through the system and the fuel conversion behavior. This works presents the fluid-dynamics study of the concept, dealing with the design and operation of a cold model. Based on the results a pilot plant has been designed.
Keywords: gasification, biomass, fluidized bed, economics, electricity

1 INTRODUCTION

Gasification is an important route for conversion of wastes and biomass materials to useful gaseous products: fuel gas for direct firing in thermal applications, such as kilns and boilers, co-firing in existing coal-fired boilers, gas for engines, turbines and fuel cells generating electricity, as well as raw gas for production of fuels or chemicals [1].

Gasification of biomass and waste in fluidized bed offers advantages, since fluidized beds are capable of being scaled up to medium and large scale, overcoming limitations found in smaller scale, fixed-bed designs. On the other hand, the bed temperature is limited in order to avoid bed agglomeration and the gasification efficiency of a fluidized bed (FB) may be limited if part of the fuel energy remains in unconverted char. The two types, bubbling (BFBG) and circulating (CFBG), differ in the sense that the latter type is always built with recirculation of particles [2]. Recycling of fines leads to a greater efficiency of carbon conversion by increasing the residence time of particles. However, CFB is more expensive than BFB so it is not a feasible solution for small-medium scale systems. In both types of FB, if the temperature is not high enough in the gasifier, the tar in the product gas can make the process unsuitable from a technical and economical point of view [1].

A project is being developed at BEGUS (Bioenergy Group at the University of Seville) for the development of a gasification technology for small/medium application that makes it possible to overtake the two major drawbacks in fluidised beds (FB), namely: (1) The low carbon conversion achieved under practical operation conditions, due to the poor char conversion, and (2) the high tar concentration in the outlet gas. In addition the system is focused on processing difficult wastes, having a high ash content and where the nature of the ash limit the temperature of the gasifier.

2 PRINCIPLE OF THE NEW GASIFICATION SYSTEM

The new gasification concept is based on a three stage gasification process: a FB devolatilizer, a air/steam reformer of the gas coming from the devolatilizer, and a chemical quench of the unconverted char discharged from the devolatilization FB.

The system is based on the movement of the gas and solid from a bubbling fluidized bed and a seal. The operational principle of the system is outlined in Fig.1, where the direction of flow of solids and gas is indicated at different sections of the system. A control valve is included to control the residence time of the solid in different parts of the system, through the control of inventory in the loop.

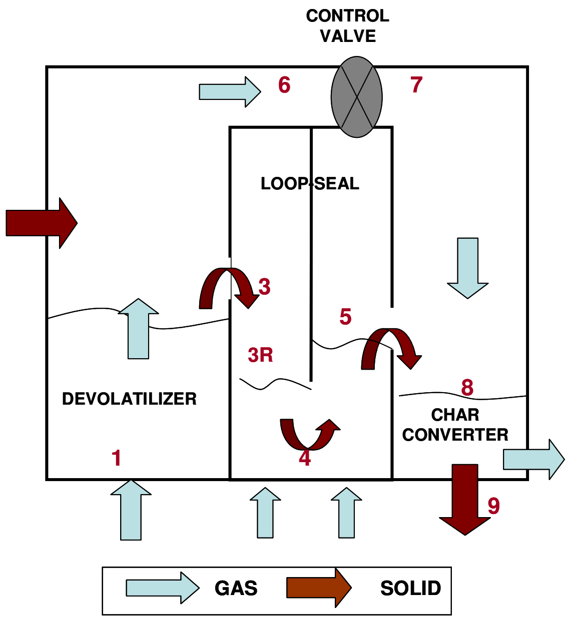

Figure 1: Operational principle of the new gasification system.

3 EXPERIMENTAL

To understand the conversion process under different operating solid and gas flow rates, the fluid-dynamics has been studied. A cold model has been constructed and operated. The cold model has been constructed based on a 2 MW_e reference plant processing dry sewage sludge (DSS). The chemical composition and main properties of

DSS used to scale-down the process is presented in Table I.

Table I: Properties of DSS

Composition of DSS (% dry)	
C	30.88
H	4.36
N	4.76
S	1.24
O	15.61
Ash	43.15
Physical properties of DSS	
Porosity	0.42
Sphericity	0.95
Average size (mm)	2
Density (kg/m^3)	800
LHV (MJ/kg dry DSS)	12

The scale-down process has been made by applying the fluid-dynamics similarity given in [3]:

$$\left(\frac{\rho_s}{\rho_g}\right), \left(\frac{G_s}{\rho_s \cdot u_o}\right), \left(\frac{\rho_s \cdot d_p \cdot u_o}{\mu}\right), \left(\frac{\rho_g \cdot L \cdot u_o}{\mu}\right), \left(\frac{u_o^2}{g \cdot L}\right), \left(\frac{L}{D}\right), (\varphi_s), PSD, BG$$

Table II shows the main relations used for the scale down.

Solids density	$\left.\frac{\rho_g}{\rho_s}\right)_m = \left.\frac{\rho_g}{\rho_s}\right)_c$
Chrasteristic length	$\left(\frac{L_m}{L_c}\right) = \left(\frac{\mu_{c(m)}}{\mu_{c(c)}}\right)^{2/3}$
Gas velocity	$\frac{u_{o(m)}}{u_{o(c)}} = \left(\frac{L_m}{L_c}\right)^{1/2}$
Residence time	$\frac{t_{r(m)}}{t_{r(c)}} = \left(\frac{\mu_{c(m)}}{\mu_{c(c)}}\right)^{1/3}$

Figure 2 shows a picture of the cold rig, indicating the main parts of the system. The main dimensions and design variables of the reactor, loop seal and fixed bed both of the cold model and reference plant is presented in Table III. Table IV presents the properties of the solids used in the test.

The experiments have been carried out with a continuous feed of solid and gas. The gas is fed in three different points (one in the reactor and two in the loop-seal). Apart from the type of particle, There are three manipulated variables: flow rate of gas to reactor and loop-seal as well as the position of the valve. The solid used was bauxita with particle size in the range of 500-800 µm and the reactor gas flow rate was set at 60 Nm^3/h. Two series of tests were carried out: one with a low flow rate in the loop seal and another with a high flow rate. In each test the valve is positioned in a given position but it is varied from one test to another in the same serie of tests.

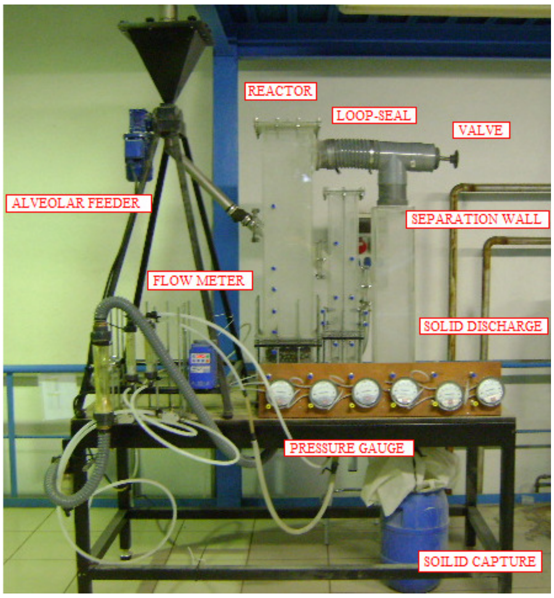

Figure 2: Cold model.

Table III Main geometric dimensions of the cold model

Cold model dimension (m)	
L_{Re}	0.22
Z_{Re}	0.22
h_{Re}	0.9
$h_{bed(Re)}$	0.32
L_{LS}	0.11
Z_{LS}	0.08
h_{LS}	0.6
$h_{bed(LS)}$	0.18
$h_{w(LS)}$	0.025 – 0.05 (*)
L_{CC}	0.16
Z_{CC}	0.16

** This height is variable*

Table IV Properties of the solid used in the tests (bauxite)

Bauxite physical properties	
porosity	0.4
sphericity	0.95
Average size (m)	0.0005
Density (kg/m^3)	3100

4 MODEL

A model was developed to determine the solids mass in the reactor and in the loop seal. In this later the solid is distributed between the downcomer (where the solids upflows) and standpipe (where the gas and solid upflow). For each of the three FBs, the following method is used to estimate the mass inventory.

The bubble fraction of the bed is calculated by two methods: In the first method the bubble fraction of the bed is estimated by [2]:

$$\delta_b = \frac{1}{1+\dfrac{u_b}{u_o - u_{mf} - u_{tf}}}$$

To calculate u_b (7) the average bubble diameter is estimated by [4,5,6].

$$u_b = 0,711 \cdot \left(g \cdot d_b\right)^{0,5}$$

$$d_b = \left(0,54 \cdot \left(u_o - u_{mf}\right)^{0,4} \cdot \left(h_{bed} + 4 \cdot A_b^{0,5}\right)^{0,8}\right) \cdot g^{-0,2}$$

The determination of u_{tf} was made by three different methods. The correlations used for each are shown in Table V [7,8]. The simplest approach taking u_{tf} =0 as in the original two-phase flow was also used.

Table V shows the main relations used for the scale down.

Johnsson [7]	$u_{tf} = (1-\chi)\cdot\left(u_o - u_{mf}\right)$ $\chi = f_2 \cdot \left(h_{bed} + 4 \cdot A_b^{0,5}\right)^{0,4}$ $f_2 = \left(0,26 + 0,7 \cdot 10^{\left[-3,3 \cdot d_p\right]}\right) \cdot \left(0,15 + u_o - u_{mf}\right)^{-0,33}$
Zijerveld [8]	$u_{tf} = (1-\chi)\cdot\left(u_o - u_{mf}\right)$ $\chi = 1,45 \cdot Ar^{-0,18}$
Two phase teory *(TPT)*	$u_{tf} = 0$

The second method for the determination of the bed bubble fraction was based on the estimation of expansion factor R: ($\delta = 1 - 1/R$). Two different correlations were used [9,10]:

$$R = 0.5482\, d_p^{-0.129}\left(u_0 - u_{mf}\right)^{0.111}$$

$$R = 1 + \frac{\left(14.31(u_o - u_{mf})^{0.738} d_p^{1.006} \rho_s^{0.376}\right)}{\left(\rho_g^{0.126} u_{mf}^{0.937}\right)} \quad \left(D_t > 0.0635\right)$$

$$R = 1 + \frac{\left(1.032(u_o - u_{mf})^{0.57} \rho_g^{0.083}\right)}{\left(\rho_s^{0.166} u_{mf}^{0.063} D_t^{0.445}\right)} \quad \left(D_t < 0.0635\right)$$

Once the bubble fraction in the bed is estimated, the bed porosity and the mass inventory of the bed are calculated by:

$$\varepsilon_b = \delta_b + \left(1 - \delta_b\right) \cdot \varepsilon_{mf}$$

$$W = \frac{\Delta P_b \cdot A}{g}$$

the pressure drop being determined by:

$$\Delta P_{bed} = \left(1 - \varepsilon_b\right) \cdot \rho_s \cdot g \cdot h_b$$

Other variables for the model are taken from the correlations presented in Table 2 in [2].

5 DISCUSSION

The bed porosity of the reactor predicted by the model using the five methods discussed above is presented in Figure 3. The experimental porosity determined by pressure measurements is also shown. As seen, all the methods give reasonable agreement except to the TPT. This is a reasonable result since, at the gas velocities investigated in this work, the throughflow is expected to be significant so that the TPT model is far from being correct.

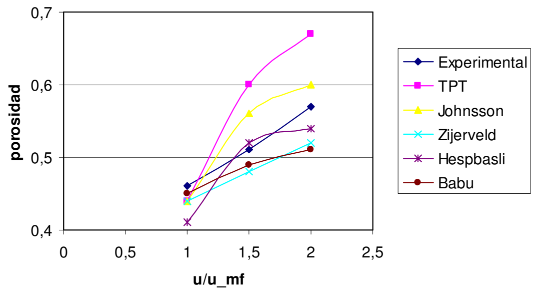

Figure 3: Porosity to bauxita 500-800 μm.

The experimental "pressure-loops" for two different runs are shown in Figures 4 and 5. In the figures the theoretical values calculated by previous model (applying the Babu correlation) is also shown. Figure 4 shows the loop for low flow in the loop-seal whereas the Figure 5 shows the loop for high flow in loop-seal. It is seen that the model reproduces the measurements reasonably.

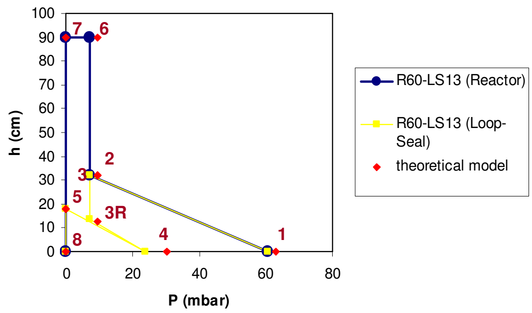

Figure 4: Pressure-loop experimental and theoretical value for low flow in loop-seal.

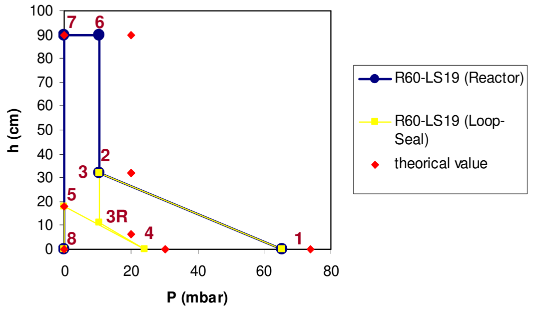

Figure 5: Pressure-loop experimental and theoretical

6 CONCLUSIONS

A new gasification system for biomass and wastes was presented. The development, construction and preliminary operation of a cold model were described. A model was developed to understand the movement of gas and solids, predicting the distribution of mass of solid between the bed and loop seal. Fuel conversion tests of DSS have also been conducted (presented also in this conference). Both the fluid-dynamics and the particle conversion have been the basis for the design of a pilot plant that will be constructed to demonstrate this new gasification concept.

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8 ACKNOWLEDGEMENTS

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9 NOMENCLATURE

A Area, m^2
A_b Area of distribuidor plate per orifice, m^2
Ar Archimedes number
D_t bed diameter
d_p average particle size, m
d_b bubble diameter, m
f_2 empirical dimensionless factor
g gravity acceleration, m/s^2
h_b height of the bed, m
$h_{w(LS)}$ wall height (grap between the both chambers of loop-seal)
R bed expansion ratio, dimensionless
t_r residence time
u_{burb} free bubble velocity, m/s
u_{mf} minimum fluidization velocity, m/s
u_{tf} throughflow velocity, m/s
u_o superficial velocity, m/s
W bed solid mass, kg
Z width

Greek letter

ε_{bed} bed porosity
ε_{mf} minimum fluidization porosity
δ_b bubble void fraction
ΔP_{bed} bed pressure drop, Pa
μ_c kinematic viscosity, m^2/s
ρ_s solid density, kg/m^3
ρ_g gas density, kg/m^3
χ dimensionless size, mass concentration

Subscripts

c commercial bed
b bubble, bed
CC char converter
L *large*
m scale model
Re reactor
LS loop-seal