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Halima Noubli

*CEA, LITEN, DTBH/LTB, France*

Sylvie Valin

*CEA, LITEN, DTBH/LTB, France*

Bertrand Spindler

*CEA, LITEN, DTBH/LTB, France*

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# DEVELOPMENT OF A MODELING TOOL REPRESENTING BIOMASS GASIFICATION IN A DUAL FLUIDISED BED UNIT

Halima Noubli\*, Sylvie Valin and Bertrand Spindler  
CEA, LITEN, DTBH/LTB;

17 Rue des Martyrs, F-38054; Grenoble, France

\*T: 33-4-3878-6885; F: 33-4-3878-5251; E: [halima.noubli@cea.fr](mailto:halima.noubli@cea.fr)

## ABSTRACT

This study is performed in the frame of the French GAYA Project for the development of a efficient and sustainable 2<sup>nd</sup> generation production pathway of Substitute Natural Gas (Bio-SNG). The aim of our study is to develop a 1D modeling tool of a dual fluidized bed biomass gasifier.

## 1. INTRODUCTION

This study is performed in the frame of the GAYA project, which aims to demonstrate the feasibility of bio-SNG (also called biomethane) production by biomass gasification and methanation. The GAYA project is coordinated by GDF SUEZ and subsidized by the French Environment and Energy Management Agency (ADEME). The technology of gasification selected in the project is that of FICFB (Fast Internal Circulating Fluidized Bed), characterized by two interconnected fluidized beds (a dense fluidized bed for the gasifier, and an entrained fluidized bed for the combustor) between which circulates a fluidizing media. The objective of one of the tasks of the GAYA project is to develop modeling tools for the gasification step. Six french partners (industrial: GDF SUEZ, academic: LGC<sup>1</sup>, LRGP<sup>2</sup>, RAPSODEE<sup>3</sup>, research centers: CEA<sup>4</sup>, CIRAD<sup>5</sup>) are involved in this task. CEA is in charge, with the support of the LGC who first initiated this work (1, 2), of developing a 1D modeling tool of the gasifier and the combustor, and the connections between the two reactors.

As a first step, we already developed a 0D modeling tool (3) to simulate the mass and energy flows between the two parts of the system. The second step is the development of 1D tools to describe the gasifier and the combustor, and finally the connection of the two. This paper presents the gasifier 1D modeling tool, in which bed hydrodynamic models are coupled with models describing biomass conversion phenomena.

## 2. PRINCIPLE AND DESCRIPTION OF THE DUAL FLUIDISED BED

The technology of gasification is that of FICFB characterized by two interconnected fluidized beds (a gasifier and a combustor) between which circulates a fluidizing media (olivine). Steam is used as a gasification agent. The FICFB process, developed by TU Vienna, is described by Hofbauer et al. (4).

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<sup>1</sup> Laboratoire de Génie Chimique, Toulouse

<sup>2</sup> Laboratoire Réactions et Génie des Procédés, Nancy

<sup>3</sup> Centre de Recherche d'Albi en Génie des Procédés des Solides Divisés, de l'Energie et de l'Environnement.

<sup>4</sup> Commissariat à l'Energie Atomique et aux énergies alternatives, Grenoble.

<sup>5</sup> Centre de Coopération Internationale en Recherche Agronomique pour le Développement, Montpellier.

This technology is successfully demonstrated at the 8 MW plant in Güssing - Austria (5), with combined production of heat and power. The principle of dual fluidized bed is presented in figure 1. The reactor on the left is a gasifier, with a dense bed at the bottom followed by a disengaging zone (or freeboard). The reactor on the right is a combustor, with a dense bed (at the bottom) and a transported bed above. Biomass is fed into the gasifier, which operates at about 850°C. The biomass is dried and devolatilized in the gasifier. The pyrolysis residue is partially gasified with steam. The residue (char) and the bed material are carried into the combustion zone which operates around 950°C. The combustor heats the bed material by air combustion of the carbonaceous char. Solids and combustion gases are then separated at the top of combustor by a cyclone. The bed material is reinjected into the gasifier, thus bringing the heat required for gasification. In some cases, an additional fuel is necessary to ensure energetic equilibrium of the system. A part of syngas at the outlet of the gasifier can then be recycled into the combustor.

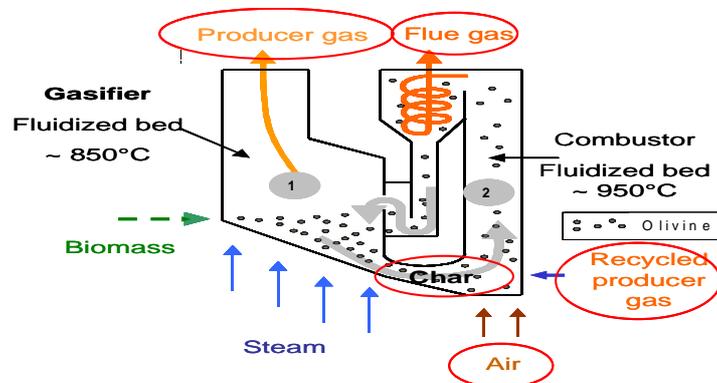


Figure 1 : Scheme of a dual fluidised bed installation, Pfeifer et al.,(6)

### 3. DESCRIPTION OF THE 1D MODELLING TOOL

In this study, only the model of the dense bed of the gasifier is presented. The disengaging zone, where gas phase reactions can also occur, will subsequently be added. The hydrodynamic description of the dense bed is based on the "bubbling bed model", developed by Kunii and Levenspiel for intermediate sized particles (7). Similarly to the 0D model (3), biomass conversion is represented as a succession of several phenomena: biomass drying, biomass devolatilisation (or pyrolysis), char steam gasification. Water-gas shift is also considered. The dense bed is divided into cells. Within each cell the local hydrodynamic, kinetic and thermodynamic variables are evaluated.

#### 3.1. BED HYDRODYNAMICS

The model is based on the following assumptions:

- The gas is divided into two phases: "bubble" phase and "emulsion" phase;
- The bed is represented as a succession of compartments or cells, with their height equal to the mean bubble diameter in the cell (Figure 2);
- The bubbles are considered to grow continuously along the bed;
- The emulsion phase is at minimum fluidization conditions;
- The gas in the bubbles is perfectly mixed;

- The solid particles (olivine and biomass) are supposed to be perfectly mixed within the dense bed, which implies that the biomass pyrolysis products (pyrolysis gases and char) are uniformly distributed among the dense bed ;
- The volume fraction of solids (char and fluidizing agent) in the bubbles is equal to 0.005;
- The gas contained in the bubble/emulsion phase flows from one cell to the other;
- The temperature is considered to be uniform due to the high mixing.

An interchange of gas by diffusion is considered between the bubble phase and the emulsion phase. Moreover, heterogeneous reaction in the emulsion (char gasification) leads to gas creation, although emulsion is supposed to remain at minimum fluidization conditions. So, a specific mass flow rate from emulsion to bubble was introduced, which corresponds to this gas excess (8, 9).

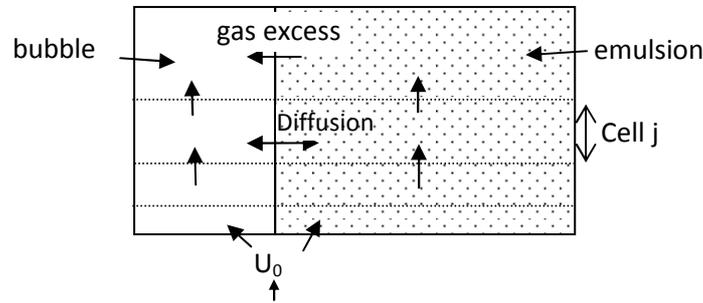


Figure 2 : Dense bed as represented in the model

In order to calculate the hydrodynamic parameters, we need to determine a number of fluid variables associated with the fluidization process. All the correlations used for calculations in this study are shown together in table 1.

**Table 1- Semi-empirical fluid-dynamic correlations used in the model**

Name of variable, Symbol	Correlations	Unit	Ref.
Maximum velocity of rise of a single bubble, $U_{b,\infty}$	$U_{b,\infty} = 0.711 (g d_b)^{0.5}$	m/s	(10)
Velocity of rise of bubbles, $U_b$	$U_b = U_{b,\infty} + (U_0 - U_{mf})$	m/s	(10)
Bubble diameter, $d_b$	$d_b = d_{bm} + (d_{b0} - d_{bm}) e^{-0.3.z/dt}$	m	(11)
Initial diameter of bubble, $d_0$	$d_0 = 1.3 / g^{0.2} ((U_0 - U_{mf}) / N_{or})^{0.4}$	m	(11)
Maximum diameter of bubbles, $d_{bm}$	$d_{bm} = 0.65 \times (\pi / 4 . d_t^2 (U_0 - U_{mf}))^{0.4}$	m	(11)
Porosity at minimum fluidization, $\varepsilon_{mf}$	$\varepsilon_{mf} = 0.586 \phi^{0.72} Ar^{-0.029} \left( \frac{\rho_G}{\rho_{FA}} \right)^{0.021}$	----	(12)
Minimum fluidization velocity, $U_{mf}$	$U_{mf} = \left( \sqrt{(31.62^2 + 0.0425Ar)} - 31.62 \right) \frac{\mu_G}{d_{FA} \rho_G}$	m/s	(7)

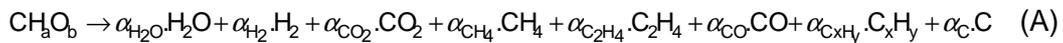
Physical properties - viscosity, density, and liquid and vapour heat capacities of each component - were evaluated using correlations from the DIPPR (13).

### 3.2. CHEMICAL REACTIONS

The simulation tool includes three chemical reactions: biomass devolatilization, char gasification and water gas shift (WGS) reaction. At the present time, the correlations used were found in the literature. They will be further modified when specific data from the GAYA project are available.

#### 3.2.1. BIOMASS DEVOLATILIZATION

Biomass devolatilization is represented as a one step reaction, with repartition coefficients deduced from the experimental work of Hémati on wood particles (14) The coefficients  $\alpha_i$  are function of temperature.



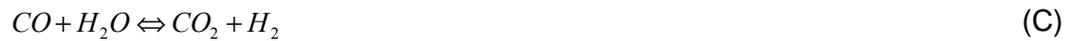
#### 3.2.2. CHAR GASIFICATION

The kinetic model of Barrio (15) is used to represent char steam gasification (Table 2). Gasification advancement depends on char residence time. A char mean residence time in dense bed is fixed in the input data of the modeling tool.



#### 3.2.3. WATER GAS SHIFT

The kinetic data for the WGS reversible reaction are presented in table 2.



**Table 2- Reaction kinetics**

Reaction parameters	Char steam gasification (B)	Water gas shift (C)	
		Forward	Reverse
r	$\frac{k m(char) P_{H_2O}^{0.57}}{M(char)}$	$k [CO]^1 [H_2O]^1$	$k [CO_2]^1 [H_2]^1$
k	$2.62 \times 10^8 \times \exp\left(\frac{-237000}{T R}\right)$	$2.78 \times \exp\left(\frac{-12560}{T R}\right)$	$\frac{2.78 \times \exp\left(\frac{-12560}{T R}\right)}{0.0265} \times \exp\left(\frac{-3958.5}{T}\right)$
Unit of k	$s^{-1} \cdot bar^{-0.57}$	$m^3 \cdot mol^{-1} \cdot s^{-1}$	$m^3 \cdot mol^{-1} \cdot s^{-1}$
References	(15)	(16)	(17)

### 3.3. MASS BALANCE

Each product mass balance is calculated in the two phases considering the gas flows in and out of the adjacent cells, the interchange of gas taking place between bubble and emulsion phases, and the chemical reactions:

$$Q_{in-b}(i, j) + Q_{Prod-b}(i, j) = Q_{Diff}(i, j) + Q_{out-b}(i, j) - Q_{Trans}(i, j) \quad (\text{Bubble phase})$$

$$Q_{in-e}(i, j) + Q_{Prod-e}(i, j) = Q_{out-e}(i, j) - Q_{Diff}(i, j) + Q_{Trans}(i, j) \text{ (Emulsion phase)}$$

With:  $Q_{in-b}(i, j)$  and  $Q_{out-b}(i, j)$  respectively the molar flow rate of component  $i$  entering / leaving the bubble phase in the compartment  $j$  (mol/s).

$Q_{in-e}(i, j)$  and  $Q_{out-e}(i, j)$  respectively the molar flow rate of component  $i$  entering / leaving the emulsion phase in the compartment  $j$  (mol/s).

$Q_{Prod-k}(i, j)$  the molar flow rate of component  $i$  produced by the reaction and transferred out of the phase  $k$  (bubble or emulsion) in the compartment  $j$ .

$Q_{Diff}(i, j)$  the molar flow rate of component  $i$  transferred from bubble to emulsion by diffusion in the compartment  $j$ .

$Q_{Trans}(i, j)$  the transverse molar flow rate of component  $i$  produced by the heterogeneous reactions in emulsion and entering the bubble phase in the compartment  $j$ .

The gas interchange between the bubble and emulsion phases by diffusion ( $Q_{Diff}$ ) is determined considering the interchange coefficient  $K_{be}$  ( $s^{-1}$ ), determined by Kobayashi correlation (18) :

$$K_{be} = \frac{0.11}{d_b} \quad (D)$$

### 3.5 RESOLUTION

The modeling tool is named GABRIEL. It is developed in FORTRAN 90. The mass balances in the two phases of each cell are solved sequentially by a Newton method. The derivatives are calculated analytically. To get a solution for the process, an iterative algorithm is used. The problem is reduced to the determination of the unknowns: molar flow rates of each component out of the bubble phase and emulsion phase, and char flow rate. The algorithm proceeds from bottom to top in the bed: the hydrodynamic parameters and reaction rates are evaluated inside each cell using the output results of the previous cell as the input. After convergence, the model yields the gas phase concentration profiles and the solid char flow rate leaving the reactor. An iterative process is superimposed because the height (or volume) of the bed is not known a priori and the pyrolysis products are uniformly distributed in the bed volume. The algorithm used to simulate the dense bed is given in figure 3.

### 4. SIMULATION RESULTS

The feasibility of the simulating tool is tested for a pilot plant of about 600 kWth. The 1D gasifier modelling tool gives the yields of the biomass gasification products –  $H_2$ ,  $CO$ ,  $CO_2$ ,  $CH_4$ ,  $C_2H_4$ ,  $H_2O$ , tar and char, the solid residue – as a function of the operating conditions of the gasifier. Examples of results with gas concentration profiles in the gasifier are given for a fixed char residence time and char diameter. Table 3 shows the typical plant operation.

**Table 3: Input parameters of the calculations**

Input parameters	Value	Unit
Biomass to Steam mass flow rates ratio	2.6	kg/kg
Gasification temperature	870	°C
Char residence time in the bed	260	s

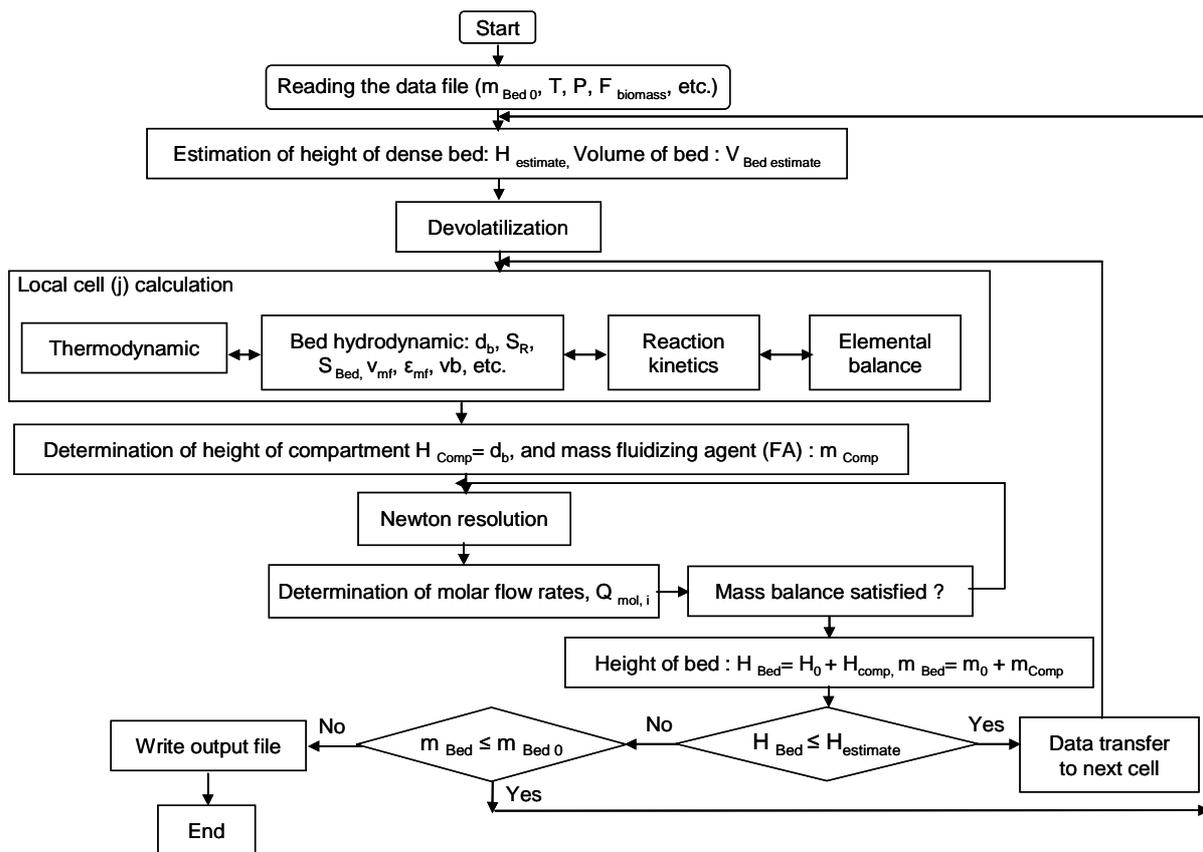


Figure 3 : Algorithm for the model

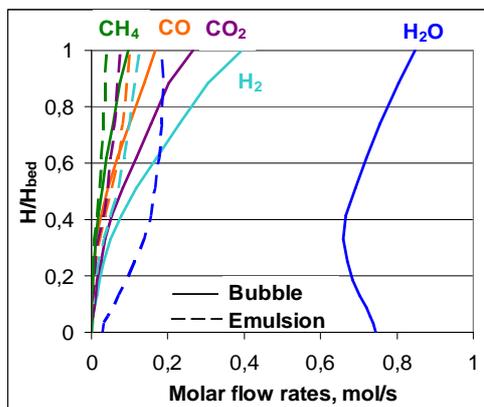


Figure 4 : Gas flow rates in bubble and emulsion phases along the dense bed

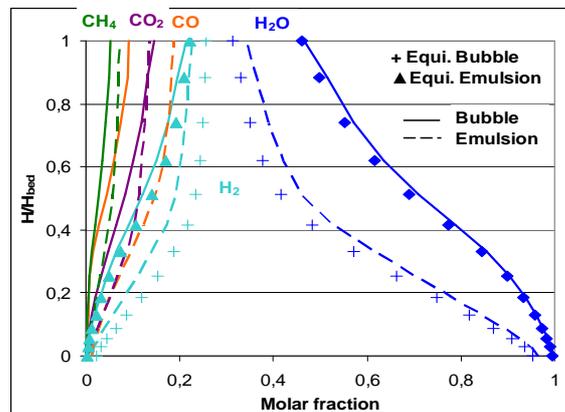


Figure 5: Gas composition in bubble and emulsion phases

Figure 4 and 5 respectively show the predicted gas flow rates passing through the bubbles and emulsion, and gas composition in bubble and emulsion phases versus the local height in the dense bed. Note that the gas flowrates in bubbles and emulsion at the top of the bed (Figure 4) correspond to what is released in the freeboard. Figure 5 shows that gas composition in the emulsion and bubble phases differ, as biomass pyrolysis products are considered to be produced only in emulsion phase, and as transfer to bubble phase is limited. The CO, CO<sub>2</sub>, H<sub>2</sub>

and H<sub>2</sub>O contents differ slightly from those calculated at WGS thermodynamic equilibrium (Figure 5).

## 5. CONCLUSION

A numerical calculation code was developed to evaluate the performance of a dual fluidized bed. First feasibility calculations in a reference case were performed. In this model, bubble-emulsion model was used to model a fluidized bed. Correlations from literature were used to represent biomass conversion reactions. In a near future, a sensibility analysis will be performed to identify which parameters have a significant influence on the results. The influence of the operating parameters will also be studied (biomass composition, biomass feeding rate, steam flowrate, temperature, diameter of char, char residence time in the dense bed, number of orifices in the distributor, dimensions of the installation, etc.). These results will help to find the optimal conditions for the operation of the gasifier. Moreover, semi-empiric laws based on specific experiments, developed by GAYA project partners, will be integrated. The simulation tool will be validated with the experimental results obtained in the lab-scale dual fluidised bed designed by the LGC, and in the pilot plant platform designed by GDF SUEZ.

### Nomenclature

#### Symbols

Ar	Archimede number, dimensionless, $Ar = d_{FA}^3 \cdot \rho_G \cdot (\rho_{FA} - \rho_G) \cdot g / \mu_G^2$
F <sub>0</sub>	Bubble to emulsion mass transfer coefficient, s <sup>-1</sup>
d <sub>b</sub>	Bubble diameter, m
dt	bed diameter, m
g	Gravity constant, cm.s <sup>-2</sup>
[i]	Concentration of species i, mol.m <sup>-3</sup>
m <sub>(char)</sub>	mass of char, kg
M <sub>(char)</sub>	molar mass of char, mol.kg <sup>-1</sup>
N <sub>or</sub>	Number of orifices by area unit, m <sup>-2</sup>
P	Pressure, Pa
Q	Molar flow rate, mol.s <sup>-1</sup>
r	Reaction rate, dependant
R	Universal constant, J.mol <sup>-1</sup> .K <sup>-1</sup>
Re <sub>mf</sub>	Reynolds number, dimensionless, $Re_{mf} = d_{FA} \cdot U_{mf} \cdot \rho_G / \mu$
T	Temperature, °C
U	Velocity, m.s <sup>-1</sup>
z	Distance to distributor, m

#### Greek letters

ε <sub>mf</sub>	Porosity at minimum fluidization, dimensionless
ρ	Density, kg/m <sup>3</sup>
μ	Viscosity, Pa.s <sup>-1</sup>
Φ	Ratio of shape, dimensionless

#### Subscripts

b	Bubble
Diff	Transferred from bubble to emulsion by diffusion
e	Emulsion
FA	Fluidizing Agent
G	Gas
in	Inlet, entering the cell
m	Maximal
mf	Minimum of fluidization

$\infty$	Maximum
out	Outlet, leaving the cell
Prod	Produced
Trans	Transverse flow from emulsion to bubble
0	Initial

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