DEVELOPMENT OF A UNISIM DESIGN MODEL FOR EVALUATING THE PERFORMANCE OF A DFB BIOMASS GASIFIER INTEGRATED WITH A ROTARY DRYER

Nargess Puladian
Jingge Li
Shusheng Pang
Chemical and Process Engineering Department
University of Canterbury,
Christchurch
Nargess.Puladian@pg.Canterbury.ac.nz

ABSTRACT

Dual fluidized bed (DFB) biomass steam gasification system has been proven to be the most promising technology to produce hydrogen rich syngas for synthesis of liquid fuels, and production of pure hydrogen and renewable chemicals. For efficient gasification of woody biomass, the biomass is required to be dried from an initial moisture content of 100% to 150% to a final moisture content of 15% to 20%. Integration and optimisation of the DFB gasifier with a biomass dryer is important in the system design and operation. In this work, a model has been developed in UniSim Design for the integrated biomass drying and gasification system to analyse the gasifier performance and system energy and exergy efficiencies.

In the integrated model, the drying and gasification operations were modelled by creating user-defined functions in UniSim Design. Mass balance, heat balance and heat transfer coefficient as well as dryer configuration were considered in the dryer model. The gasification process is considered to consist of three stages including initial pyrolysis, subsequent gasification reactions of char and volatile gas and those of resultant gases and steam. In addition, char-air reactions were also modelled in the circulating fluidised bed reactor which is part of the DFB gasification system to provide heat for the biomass steam gasification. Five factors are introduced in the modelling of the gasification process including carbonic and methane formation ratios for pyrolysis reactions, a steam participation factor for char–gas reaction, a table of empirical equilibrium constant for shift reaction and a factor for tar formation. This model has been validated with experimental data.

A parametric study was performed to investigate the effects of gasification temperature, steam to biomass ratio on char yield, gas yield and the $\text{H}_2/\text{CO}$ ratio in the producer gas. In addition, excess fuel demand and thermodynamic efficiency of the system were examined.

INTRODUCTION

Dual fluidised bed steam gasification has been proven to be the promising technology for converting biomass into hydrogen rich syngas for liquid fuel synthesis or for production of pure hydrogen for fuel cell application. A pilot scale dual fluidised bed (DFB) gasifier has been constructed at the University of Canterbury for biomass gasification following the pioneer work at the Vienna University of Technology (Saw and Pang, 2012, Hofbauer et al., 2002, Hofbauer et al., 2003). The DFB gasifier consists of a bubbling fluidised bed (BFB) gasification reactor fluidised with steam as the gasification agent and a circulating fluidised bed (CFB) combustion reactor fluidised
with air for char combustion. This process produces a hydrogen-rich producer gas with much higher caloric value of approximately 13 MJ/Nm³ compared with that from air gasification, which is approximately 5 MJ/Nm³ (Saw and Pang, 2012). It is also able to produce a syngas with the desired ratio of H₂/CO suitable for Fischer-Tropsch synthesis of transport fuels (Saw et al., 2012).

For efficient gasification in a fluidised bed, feedstock of woody biomass is normally in the form of pellets or chips with particle size of 2–30mm and moisture content of less than 20% on oven dry base (Hofbauer et al., 2002, Van der Drift et al., 2001). However, green wood material usually has high moisture content of 100–150% (Li and Pang, 2009). Drying of the feedstock is necessary as a pre-treatment procedure in an industrial gasification plant for which rotary dryer is identified to be most suitable (Xu and Pang, 2008).

Both the gasification and wood drying consume energy. In the DFB gasification system, the energy is self generated in the CFB reactor. While a proportion of the flue gas from the CFB reactor can be utilised for the biomass drying. Integration of the DFB gasifier with the rotary dryer is important in the operation optimisation. However, no publication has been found for analysis of energy and exergy flow in the integrated system including DFB gasifier and the biomass dryer. The other integrated systems including entrained flow gasifiers and fluidized bed gasifiers were studied by Bechtel (1998), Tijmensen (2002) and Swanson et al. (2010). The dryer was included in the work of Bechtel (1998) and Swanson et al. (2010) but the drying medium in these studies was the air heated by the energy recovered from other parts of the plant. Besides, these studies focused on economic analysis rather than energy consumption and process performance.

In the present study, the rotary dryer and the DFB gasifier are modelled using UniSim Design software to analyse the flows and efficiencies of energy and exergy in an integrated biomass drying and biomass gasification system. The model is then employed to investigate the performance and to optimise the gasifier operation.

For modelling the drying system, there are generally two approaches. One is based on detail mass and heat transfer both between the drying medium and the solid material and within the solid material (Marinos-Kouris et al., 1998). Another one considers the drying as a heating system and is based on the overall mass and heat balance of the system employing empirical equations or parameters (Xu and Pang, 2008).

There are generally three different approaches for modelling of the gasification system: thermodynamic equilibrium models, kinetic rate models and neural network models (Ngo et al., 2011). Choosing of the suitable model depends on the application and the extent of the study.

The thermodynamic equilibrium models are the simplest and give an estimation of the producer gas composition by considering elemental balances and equilibrium constants in relevant reactions. This method was employed successfully for coal gasification although attempts have also been made for biomass gasification (Schuster et al., 2001, Rutherford, 2006, Jarunthammachote and Dutta, 2007). In recent work of Ngo et al. (2011), quasi equilibrium model was employed which makes it possible for detail parametric studies of the gasification process.

Employing the commercial computer programs for modelling and simulation of chemical and petrochemical plants is very common these days. By the aid of these simulators, even the very complex plants can be divided into several unit operations. And each of these unit operations can then be modelled by a combination of the
simulation library models and user models in order to capture the performance of the actual process equipment. Each of these process simulation programs has its own unique characteristics however they share many common features such as type of reactors, columns, heat exchangers and thermodynamic packages (Towler and Sinnott, 2008). Among various available simulators Aspen Plus (Aspen Technology Inc.) and UniSim Design (Honeywell Inc.) are widely used. UniSim Design is based on the Hysys software that was originally developed by Hyprotech Ltd. and now is owned and licensed by Honeywell. UniSim Design is more popular in chemical plants and oil refineries while Aspen Plus is used in wider ranges of industries. These two simulators have both advantages and disadvantages.

Both Aspen Plus and UniSim Design have very powerful data bases. In UniSim Design, all the models and components are inside the simulator and the flowsheet is obvious and it is very easy for checking and tracking (user friendly). Aspen Plus can deal with very complex industrial processes. But, Unisim Design will allow you to do something otherwise it is impossible in the flowsheet such as water flows bellow its freezing point and pressure increasing through plug flow reactor or heat exchangers. And in simulating amine plants, Aspen Plus is more powerful than Unisim Design because packed columns is supported by the amine package and you can enter the kinetics as well. However, the graphic and interface in Aspen Plus is weaker than UniSim Design, the kinetics and models must be defined for the simulation and in some cases it is hard to converge (Wilcox, 2012).

Aspen Plus was used by many researchers such as Ramzan et al. (2011) and Raju et al. (2009) for simulating the special gasification processes including CFB and entrained flow gasifier for both coal and biomass. The methodologies employed in all these studies are very similar. The gasification process was divided into several steps namely pyrolysis, carbon-gas reactions and steam-gas reactions. Similarly, Aspen Technology Inc. itself has added a simple built-in model for coal gasification to Aspen Plus consisting of RStoic, RGibbs and REquil applicable for an IGCC model (aspentech, 2008). According to the description of the model, it covers the commercially available entrained flow gasifier. However, some researchers like Biagini et al. (2009) and Lang et al. (2011) preferred to develop their own model by a combination of user models and available library models since this approach makes the model more flexible for modification to become more consistent with the experiment and reality.

In conclusion, there is no special advantage for simulating the gasification process by Aspen Plus over UniSim Design. The complexity of the system, the ultimate goal of the simulation and the cost dictate which simulator should be used. In this project, it was decided that UniSim Design which has already been available in our department is satisfying for the simulation of the biomass gasification plant. Its user friendly interface, its objective nature and its ability to interact with MS Excel spreadsheet for exporting and importing the data are expected to make the simulation faster and more applicable.

MODEL ESTABLISHMENT

Process

The process diagram is shown in Figure 1. The wood chips enter the rotary dryer where they are dried to around 17% (od). The dried biomass then goes to the gasifier. The hot flue gas from the gasifier CFB reactor is used firstly for air preheating and steam generation, both of which are used in the DFB gasification system, and then for the biomass drying.
The drying of solid biomass occurs in three steps; initial heat up, fast and constant rate drying and falling rate drying. The first or initial step is when the sensible heat is transferred from the drying medium to the wet material in which the wet solid substance is heated to the gas wet bulb temperature (Perry and Green, 1999).

In the fast and constant drying rate period, liquid moisture of wet biomass evaporates on the biomass particle surface and the drying rate is controlled by the heat transferred from the drying medium to the evaporation surface (Perry and Green, 1999).

In the falling rate period, the moisture content of the biomass is less than its critical moisture content and the drying rate decreases with the biomass moisture content due to the increase in resistance for moisture transfer within the material. In this period, the rate of drying is controlled by both the diffusion of moisture from the surface to the gas and moisture movement within solid (Perry and Green, 1999).

However, drying was conventionally regarded as a heat transferred controlled process and thus the dryers were designed on the basis of heat transfer considerations (Ngo et al., 2011).

In this study, a user defined unit operation was established in the UniSim Design software for simulating the biomass drying in a rotary dryer. The input parameters of the drying unit operation are the flue gas as drying medium and wood chips as feedstock whereas the target biomass final moisture content is the outlet parameter. As illustrated in Figure 2, the model predicts the appropriate conditions of the inlet drying medium such as flow rate and drying temperature with a target to achieve best economic performance of the dryer. The basis of the calculations is described bellow.
The overall mass balance of the dryer is established based on the moisture flow and phase changes. The water lost by wood is gained by the gas phase. By assuming the moisture content of the outlet biomass, the moisture content of the exit gas and mass flow of outlet streams can be calculated from the following mass balance equation:

\[ M_2(X_2 - Y_2) = M_1(X_1 - X_2) \]  

The co-current configuration is selected to prevent ignition at the biomass exit where the drying biomass is in contact with hot gas. In establishing the energy balance equation, the heat lost by the drying medium (gasifier hot flue gas) is used on the following items:

- Heating the wet biomass and water to the wet bulb temperature in the heat up period:
  \[ M_2C_{p\text{wood}}(T_w - T_{\text{wz}}) + M_2X_2C_{p\text{lvw}}(T_w - T_{\text{zlw}}) \]
- Vaporization of liquid moisture:
  \[ M_2(X_2 - X_1) \Delta H_{\text{vrv}} \]
- Heating the biomass to the final solid temperature:
  \[ M_2C_{p\text{wood}}(T_{s_f} - T_w) \]
- Heating the remaining moisture in the biomass to the final solid temperature:
  \[ M_2X_2C_{p\text{lvw}}(T_{s_f} - T_w) \]
- Heating the water vapor to the gas final temperature:
  \[ M_2(X_1 - X_2)C_{p\text{lvw}}(T_{\text{gat}} - T_w) \]

The heat load of the system is calculated from heat and mass balances. In determination of the dryer dimensions, the heat transfer coefficient of the dryer was calculated using the correlation proposed by Saeman and Mitchell (1954).

**DFB GASIFIER**

An overview of the model established in UniSim Design for simulating the performance of DFB gasifier is shown in Figure 3. For establishing the model for the DFB gasifier system in UniSim Design, the whole process is divided into four steps including three steps in the bubbling fluidised bed reactor (initial pyrolysis; char-gas reactions and steam-gas shift reaction) and one step in the circulating fluidised bed reactor (combustion of char and supplementary fuel). The three steps in the BFB gasification reactor were modelled by a quasi three stage equilibrium model. For the pyrolysis step, a macro code in visual basic language was written in a user defined unit operation converting the woody biomass to pyrolysis gas and char at the operation temperature. For char-gas reactions, a Gibbs reactor was designed to calculate the equilibrium of
char–gas reactions with limited amount of steam. The limited steam-gas shift equilibrium reaction also occurred in the designed equilibrium reactor.

The un-reacted char followed into the CFB reactor in which the char and supplementary fuel (LPG) was burnt. For modelling of char combustion step, a conversion reactor was defined.

**Figure 3. An overview of the DFB gasifier model in UniSim Design.**

**Gasification Kinetics**

In the gasification process, a series of complex and competing reactions occurs each of which may have different significance in the gasifier (Franco et al., 2003). The process of biomass gasification can be divided into three steps, namely; initial pyrolysis, char-gas reactions and steam-gas reactions (Sadaka et al., 2002b, Ngo et al., 2011, Sadaka et al., 2002a, Nguyen et al., 2010, Franco et al., 2003). In the initial pyrolysis, the biomass feed is devolatilised to gases and char. For the char–gas reactions, Boudouard reaction, primary and secondary steam-gas reactions are involved which are followed by the steam-gas shift reaction in the third step.

According to Franco et al. (2003), the steam-gas shift reaction is more dominant at lower temperatures which is favourable for the reaction equilibrium. At higher temperatures, say 830–900°C, the steam-gas reactions and Boudouard reaction are more prevalent. The reaction constant of the steam-gas shift reaction was determined from experiments and compared with theoretical equilibrium constants as reported by Herguido et al. (1992) and Wei et al. (2007) (Figure 4).

**Figure 4. The theoretical and experimental shift reaction constant at different temperatures.**
The average value of steam-gas shift reaction constants presented by Herguido et al. (1992) and Wei et al. (2007) was used in the present modelling in UniSim Design. This is regarded as the first experimental factor employed in the DFB model.

**The Contribution of Steam**

The chemical equilibrium model for coal gasification was studied by Yoshida et al. (2008). They found that char–gas reactions did not reach the equilibrium in actual system. The reason was thought to be the limited contribution of steam in char–gas reactions. They suggested that for a more applicable equilibrium model, a temperature dependent correlation should be introduced in a reaction model to reflect steam participation in char gasification reaction. The formula represented by Yoshida et al. (2008) for steam participation showed that the influence of steam participation had a weak correlation with gasification temperature. Other factors influencing steam contribution was also investigated which included feed reactivity, feed particle size, system’s design and residence time. The empirical correlation which reflects the diffusion rate of steam on char particles was employed by Nguyen et al. (2010) for the system's design and residence time. The empirical correlation which reflects the contribution was also investigated which included feed reactivity, feed particle size, system’s design and residence time. The empirical correlation which reflects the diffusion rate of steam on char particles was employed by Nguyen et al. (2010) for the coal gasification as given in Eq.(2) in which \( n_{\text{H}_2\text{O,con}} \) is the number of moles of steam participating in the steam-gas reactions and \( n_{\text{H}_2\text{O}} \) is the total steam mole number available in the system.

\[
\beta = \frac{n_{\text{H}_2\text{O,con}}}{n_{\text{H}_2\text{O}}} = 51.4 \exp\left(-\frac{7542.8}{T}\right), \quad T \text{ in K}
\]  

(2)

**The Calculation Procedure**

In the modelling of initial pyrolysis, the first step is to find the composition of pyrolysis gases which consist of H\(_2\), CO, CO\(_2\), H\(_2\)O, and CH\(_4\). For simplification, traces of tars and other light hydrocarbons are represented by CH\(_4\) in the model. There are three elemental balance equations for C, H, and O and two empirical factors of CH\(_4\)/H\(_2\) (\(\Phi_{\text{CH}_4}\)) and CO/CO\(_2\) (\(\Phi_{\text{CO}}\)). The Arhenius equations for \(\Phi_{\text{CH}_4}\) and \(\Phi_{\text{CO}}\) were derived from data presented by Fagbemi et al. (2001) as listed in Table 1 for pyrolysis of wood at different temperatures. \(\Phi_{\text{CO}}\) and \(\Phi_{\text{CH}_4}\) are the third and fourth experimental factors employed in DFB model.

Therefore, the concentrations of the five major components of the gases during initial biomass pyrolysis step have been calculated. The three chemical element balance equations are given in Eqs. (3) – (5) in which \( n_{\text{CO}} \), \( n_{\text{CO}_2} \), \( n_{\text{CH}_4} \), \( n_{\text{H}_2} \) and \( n_{\text{H}_2\text{O}} \) stand for the moles of the corresponding components in the pyrolysis gas and \( n_{\text{H}} \), \( n_{\text{C}} \) and \( n_{\text{O}} \) are the moles of hydrogen, carbon and oxygen in the feed biomass.

\[
\begin{align*}
&n_{\text{CO}} + n_{\text{CO}_2} + n_{\text{H}_2\text{O}} = n_{\text{R}} \\
&4n_{\text{CH}_4} + 2n_{\text{H}_2} + 2n_{\text{H}_2\text{O}} = n_{\text{R}} \\
&n_{\text{CO}} + 2n_{\text{CO}_2} = n_{\text{O}}
\end{align*}
\]  

(3) (4) (5)

**Table 1. Values of \(\Phi_{\text{CO}}\) and \(\Phi_{\text{CH}_4}\) at different gasification temperatures for woody biomass (Fagbemi et al., 2001)**

<table>
<thead>
<tr>
<th>Temperature, °C</th>
<th>400</th>
<th>500</th>
<th>600</th>
<th>700</th>
<th>800</th>
<th>900</th>
</tr>
</thead>
<tbody>
<tr>
<td>(\Phi_{\text{CO}})</td>
<td>0.659</td>
<td>1.085</td>
<td>1.824</td>
<td>2.653</td>
<td>5.516</td>
<td>10.7</td>
</tr>
<tr>
<td>(\Phi_{\text{CH}_4})</td>
<td>7.154</td>
<td>1.684</td>
<td>1.528</td>
<td>1.069</td>
<td>0.683</td>
<td>0.478</td>
</tr>
</tbody>
</table>
The total carbon content of the gases in left hand side in Eq.(3) should be equal to the difference between the amount of carbon in ultimate analysis and the amount of fixed carbon in proximate analysis. The fixed carbon content of biomass is associated with the char production in pyrolysis step although some carbon may react in char–gas reactions, and therefore, the char production of the gasification process in the BFB reactor was determined after char–gas reaction phase.

For calculating the tar content in gasification gas, an empirical relation for tar content of producer gas as a function of gasification temperature was estimated from experimental data reported by Herguido et al. (1992). Eq.(6) is the fifth experimental factor employed in the DFB model.

\[
\text{Tar(Wt\%)} = -0.02245 \times T(\text{K}) + 27.6
\]  

(6)

RESULTS AND DISCUSSION

Validation of the Gasification Model with Experiment

The developed biomass gasification model has been validated by comparing the simulation results of producer gas composition with experimental data obtained from DFB gasifier reported in literature (Koppatz et al., 2011). The bed material in BFB gasifier was olivine which did not show significant catalytic effect. The results are shown in Figure 5 (a) for H₂/CO ratio and in Figure 5 (b) for individual gas species for gasification temperatures from 750°C to 950°C. From Figure 5 (a), close agreement between the simulated and experimentally measured ratio of H₂/CO can be observed. However, in Figure 5 (b), remarkable discrepancies have been found between the simulation results and the experimental data for the individual gas species. The H₂ content from simulation is about 18.6% higher than the experimental data at round 750°C while the discrepancy decreases to 14.8% at 850°C. Similarly, for CO there is about 24% discrepancy between the simulation results and the experimental data. For CO₂ and CH₄, the simulation results are consistently lower than the experimental data. However, the model predicted trends for all of the gas species are consistent with experimental data.

![Figure 5. Comparison of model simulated results from the present study and the experimental data reported by Koppatz et al. (2011) at steam to biomass ratio of 0.84 (a). H₂/CO ratio, (b). individual gas species.](image)

Parametric Analysis

The developed model in the present study has been used for sensitivity studies to investigate the operation conditions on the gasification performance including char yield, gas yield and H₂/CO ratio. The gas yield is defined as the volume of producer gas
(db) per kg of inlet biomass (od). The char yield is the percentage of the fixed carbon remained in the system after the gasification reactions in the BFB reactor. The effects of gasification temperature on the gasifier performance are shown in Figure 6 (a) for steam to biomass ratio of 0.67 (db). It has been found that $\text{H}_2/\text{CO}$ and gas yield increase significantly with the gasification temperature while char yield is almost constant. The effects of steam to biomass ratio (S/B) on the gasification performance are shown Figure 6 (b) at gasification temperature of 750°C. It can be seen that S/B ratio has more significant effect in increasing the $\text{H}_2/\text{CO}$ ratio than the gasification temperature. The char yield decreases dramatically with the increase in S/B ratio. Both temperature and steam to biomass ratio have a similar effect on gas yield. The combined effects of S/B ratio and the gasification temperature on the $\text{H}_2/\text{CO}$ ratio are shown in Figure 7(a) from which the highest $\text{H}_2/\text{CO}$ can be achieved with high S/B ratio and at high gasification temperature. In steady operation of the gasification system, supplementary fuel (LPG) is normally required to maintain the required operation temperature. The amount of LPG requirement is affected by both the S/B ratio and the gasification temperature as illustrated in Figure 7 (b). It can be seen that supplementary LPG requirement of the system also increases with the gasification temperature and the S/B ratio.

Figure 6. (a). Simulated effects of gasification temperature on gasifier performance at S/B ratio of 0.67, (b). Effects of S/B ratio on gasifier performance at T=750°C.

Figure 7. (a). Effects of S/B ratio and gasification temperature on $\text{H}_2/\text{CO}$ ratio in producer gas, (b). Effects of S/B ratio and gasification temperature on LPG requirement.
Thermodynamic Analysis for the Integrated Drying and Gasification

The drying operation significantly affects the energy efficiency of the system. With increase in the moisture content of the feedstock, the temperature of the drying medium which is the mixture of fresh air and flue gas after preheating the air and generation of steam for the gasification system has to be increased (Figure 8).

The other important factor affecting the energy efficiency is the amount of air supplied to the CFB reactor. Increase in the flow rate means more energy is required from the flue gas and thus resulting in reduced energy efficiency of the system. On the other hand, with increase in air flow rate to the CFB reactor, more gas is available for drying thus the drying temperature can be lowered. The effect of feed moisture content and air supplied to the system on the gas inlet temperature of the dryer can be seen in Figure 8 (a). By integrating the flue gas heat in the system for producing the steam requirement of the gasifier, the energy efficiency of the system can be improved. However, more energy is needed for drying with increase in the feed moisture content; less energy is available for generation of steam for the system as seen in Figure 8 b. At the extreme case, when the feed moisture content increases to 150%, no steam can be generated and all the steam needed has to be imported.

Figure 8. (a). The required drying temperature as a function of feed moisture content and air/biomass ratio. (b). The percentage of the steam generated from flue gas heat recovery for different feed moisture content.

The energy and exergy efficiency of a 100MW system has been simulated using the developed model and the results are shown in Figure 9 (a) for effects of moisture content on the efficiencies and in Figure 9 (b) for corresponding heat loss from the dryer. It has been observed that with increase in the feedstock moisture content, the energy efficiency of the integrated system decreases dramatically. In contrast the exergy efficiency shows a slight decrease. The dryer’s heat loss by the exhaust gas has similar trend to the exergy loss although the exergy loss is less significant with the increase in feedstock moisture content. The explanation of the less exergy loss is that the exhaust gas temperature is low thus the exergy possessed by the exhaust gas is low. The heat loss from the dryer’s exhaust gas is the main component of the energy loss in the system while for the exergy loss the irreversibility of gasifier is the main cause of exergy loss. The energy and exergy efficiencies of the gasifier, itself, are 90% and 75% respectively.
It has been found that the trends of the simulation results are consistent with the energy and exergy efficiencies for a commercial biomass DFB gasification plant. Studying the effect of various operation parameters on the gasifier performance and on gasification reactor are modeled by a quasi three stage equilibrium model with five reactor (combustion of char and supplementary fuel). The three steps in the BFB gasification reactor are modeled by a quasi three stage equilibrium model with five experimental factors. The char-air reactions in the CFB reactor are modeled based on complete combustion while supplementary fuel is added when needed to achieve the required gasification temperature. The approach results in a user-friendly model for studying the effect of various operation parameters on the gasifier performance and on the energy and exergy efficiencies for a commercial biomass DFB gasification plant.

The developed model has been validated by comparing the simulation results of gas composition and H₂/CO ratio with the experimental data recently published in literature. It has been found that the trends of the simulation results are consistent with the experimental data with close agreement for H₂/CO ratio. Further work is being undertaken to validate the developed model using experimental data obtained from this research team. In addition, downstream processes of gas cleaning and Fischer-Tropsch synthesis of liquid fuel using the clean gas from the gasification plant will be added.

CONCLUSION

In the present study, an integrated model for integrated biomass drying and biomass steam gasification in a DFB gasification system has been developed in UniSim Design. The mass and heat balances have been established and heat transfer correlation was employed from the literature. The gasification system was divided into four steps including three steps in bubbling fluidized bed reactor (initial pyrolysis; char-gas reactions and steam-gas shift reaction) and one step in the circulating fluidized bed reactor (combustion of char and supplementary fuel). The three steps in the BFB gasification reactor are modeled by a quasi three stage equilibrium model with five experimental factors. The char-air reactions in the CFB reactor are modeled based on complete combustion while supplementary fuel is added when needed to achieve the required gasification temperature. The approach results in a user-friendly model for studying the effect of various operation parameters on the gasifier performance and on the energy and exergy efficiencies for a commercial biomass DFB gasification plant.

The developed model has been validated by comparing the simulation results of gas composition and H₂/CO ratio with the experimental data recently published in literature. It has been found that the trends of the simulation results are consistent with the experimental data with close agreement for H₂/CO ratio. Further work is being undertaken to validate the developed model using experimental data obtained from this research team. In addition, downstream processes of gas cleaning and Fischer-Tropsch synthesis of liquid fuel using the clean gas from the gasification plant will be added.

NOMENCLATURE

- $M_b$: The inlet mass flow of biomass (dry), kg/hr
- $M_g$: The inlet mass flow of gas (dry), kg/hr
- $Y_i$: The moisture content of inlet gas, wt% (dry)
- $Y_o$: The moisture content of outlet gas, wt% (dry)
- $X_i$: The moisture content of inlet feed, wt% (dry)
- $X_o$: The moisture content of dried biomass, wt% (dry)
- $T_{bi}$: Biomass inlet temperature, °C
- $T_{fd}$: Solid final temperature, °C
- $T_{fgi}$: Hot flue gas inlet temperature, °C
$T_{\text{g, out}} = \text{Gas final temperature, } ^\circ\text{C}$
$T_{\text{w}} = \text{Wet bulb temperature, } ^\circ\text{C}$
$C_{p_g} = \text{Hot gas heat capacity, } \frac{\text{kJ}}{\text{Kg, } ^\circ\text{C}}$
$C_{p_{\text{wood, l}}, C_{p_{\text{water, v}}} = \text{Specific heat of biomass, liquid & water vapor, } \frac{\text{kJ}}{\text{Kg, } ^\circ\text{C}}$
$\Delta H_{\text{vap}} = \text{Heat of vaporization, } \frac{\text{kJ}}{\text{Kg}}$

REFERENCES


SWANSON, R. M., SATRIO, J. A. & BROWN, R. C. 2010. Techno-Economic Analysis of Biofuels Production Based on Gasification. NREL.


