ADDIS ABABA UNIVERSITY

ADDIS ABABA INSTITUTE OF TECHNOLOGY

SCHOOL OF CHEMICAL AND BIO-ENGINEERING

ENVIRONMENTAL ENGINEERING STREAM

BIOMASS GASIFIER PROCESS SIMULATION FOR SUSTAIBALE ENERGY PRODUCTION FROM DIFFERENT BIOMASS FEEDSTOCK

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.

November, 2017

Addis Ababa, Ethiopia

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A Thesis Submitted to Addis Ababa University, Institute of Technology, School of Chemical and Bioengineering in partial fulfillment for the requirement of Master of Science in Environmental Engineering

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November, 2017

DECLARATION

This is to certify that the thesis prepared by Seada Mohammed, entitled biomass gasifier process simulation for sustainable energy production from different biomass feedstock submitted to Addis Ababa University in partial fulfillment for the requirement of the Degree of Master of Science in Environmental Engineering complies with the regulations of the university and meets the accepted standards with respect to originality and quality.

Signed by the Examining Committee

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ABSTRACT

With the gradual depletion of coal and petroleum resources, biomass is being perceived as a self-sustainable source of energy production. It is cheap and readily available as well. Gasification is one of the potential sources for production of clean and eco-friendly fuel. ASPEN PLUS simulator is a strong tool for investigating the behavior of a process and it can be readily used to access various aspects like feasibility of an operation, effect of operating parameters on the performance of a gasifier. In this paper, steady state simulation model of air gasification has been developed in Aspen plus for fixed bed biomass gasifier using rice husk, coffee husk and saw dust as a fuel and the analysis of fuel samples was done through ultimate and proximate properties of the biomass. Aspen plus is selected as a simulation tool due to its higher capability of handling solid feed using physical models. The fixed bed gasifier used here is an updraft type one with drying, Pyrolysis, gasification, and combustion stages. The gasifier has been modeled based on thermodynamic equilibrium model by Gibbs free energy minimization in four stages. In the first stage moisture content of biomass feed is reduce through drying. In the second stage biomass is decomposed into its elements by specifying yield distribution. In the third and fourth stages gasification and combustion reactions has been modeled using Gibbs free energy minimization approach. Simulation result include; sensitivity analysis of Air to fuel ratio, gasification temperature, gasification pressure, and, steam to biomass ratio have been varied over wide range and the effect of these parameters on syngas composition and lower heating value has been investigated. The sensitivity analysis results indicated that increase in air to fuel ratio decreases the heating value of the producer gas. Temperature increases the production of CO and H_2 and enhances the heating value of the producer gas. Higher pressure reduces H_2 rich syngas hence decreases the lower heating value of the producer gases. Steam as a gasifying agent favors hydrogen production which results in increase of hydrogen content in the syngas, while increase of the steam to biomass ratio had negative effects on lower heating value of syngas. Based on the simulation results, maximum lower heating value of syngas was found at the gasification temperature of 800ºC, steam flow rate of 0.3 kg/hr, and pressure of 1bar was obtained at air flow rate of 0.5 kg/hr for 1.5 kg/hr of each feed stocks.

***Key words**: Biomass gasification, Sensitivity analysis, Aspen Plus*

ACKNOWLEDGEMENTS

I feel immense pleasure and privilege to express my deep sense of graduation and feel thankful towards all those people who have helped, inspired and encouraged me during the preparation of this paper. I would like to thank Dr. Ing, Abubaker Yimam and Dr. Ing. Hundessa Dessalegn for them support and advice during the course of my work. Last but not the least; I would like to thank whole heartedly my graduate class mates whose love and unconditional support, both on academic and personal front enabled me to see the light of this day.

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1. INTRODUCTION

Biomass resources include wood from plantation forests, residues from agricultural or forest production, and organic waste by-products from industry, domesticated animals, and human activities [1]. Biomass, which is made up of a wide variety of municipal solid waste, green waste and agricultural residues are found in large quantity and is a proven natural resource for renewable energy. The energy can be recovered from solid wastes through various technologies is a major contributor to renewable energy [2]. Biomass is also considered as a valuable energy alternative to fossil fuels. Because, dependency of traditional fossil fuels (oil, coal and natural gas) has led to energy crisis and environmental problems, i.e. fossil fuel exhaustion and pollutant emission. Carbon dioxide is the main greenhouse gas and a major part of CO_2 emission is combustion of fossil fuels. Also combustion of fossil fuels produces toxic gases, such as SO_2, NO_x and other pollutants, causing global warming and acid rain. It may be converted to a variety of usable forms of energy such as syngas, biogas and liquid transportation bio-fuels [3].

Now it is a major issue to explore and exploit new sources of energy that are renewable and eco-friendly. Renewable energy sources have a minimum environmental impact than conventional energy sources. The potential of renewable energy for urban and rural development, liquid fuel replacement and greenhouse gas reduction are current concerns all over the world. The renewable energy sources such as solar, wind, hydro wave, geothermal and biomass offer attractive prospects because they are unlimited and cheap [4].

Two main ways of converting biomass energy (solid fuel) into bio-fuels and bio-power are biochemical conversion and thermo-chemical conversion processes. Biochemical conversions convert the biomass into liquid or gaseous fuels by fermentation or anaerobic digestion. Fermentation of the biomass (starch and cellulose) produces primarily ethanol. Anaerobic digestion leads to the production of gaseous fuel primarily containing methane.

Thermo chemical conversion technologies include combustion, gasification and pyrolysis. While combustion of biomass is the most direct and technically easiest process, the overall efficiency of generating heat from biomass energy is low. Gasification has many advantages over combustion. It can use low-value feed stocks and convert them not only into electricity, but also into transportation fuels. In the upcoming years, it will serve as a major technology for complementing the energy needs of the world [5]. Use of advanced

technologies such as gas turbines and fuel cells with the syngas generated from gasification results in increased efficiency [6]. For complete combustion of solid fuels, excess air is needed, and high combustion temperatures generate more NO_X and other emissions, as compared with the combustion of products by gasification. During combined cycles for combined heat and power generation, contaminants in the syngas such as sulfur and nitrogen species and trace elements are removed efficiently resulting in much lower emissions [7]. Moreover, liquid and gaseous fuels are more of interest because of their ease of handling and operations, and their applications as transportation fuels.

Huber *et al*. [8] provide an excellent review of the possible biomass conversion strategies for the production biofuels. The main biofuel production pathways are depicted in figure 1.1.

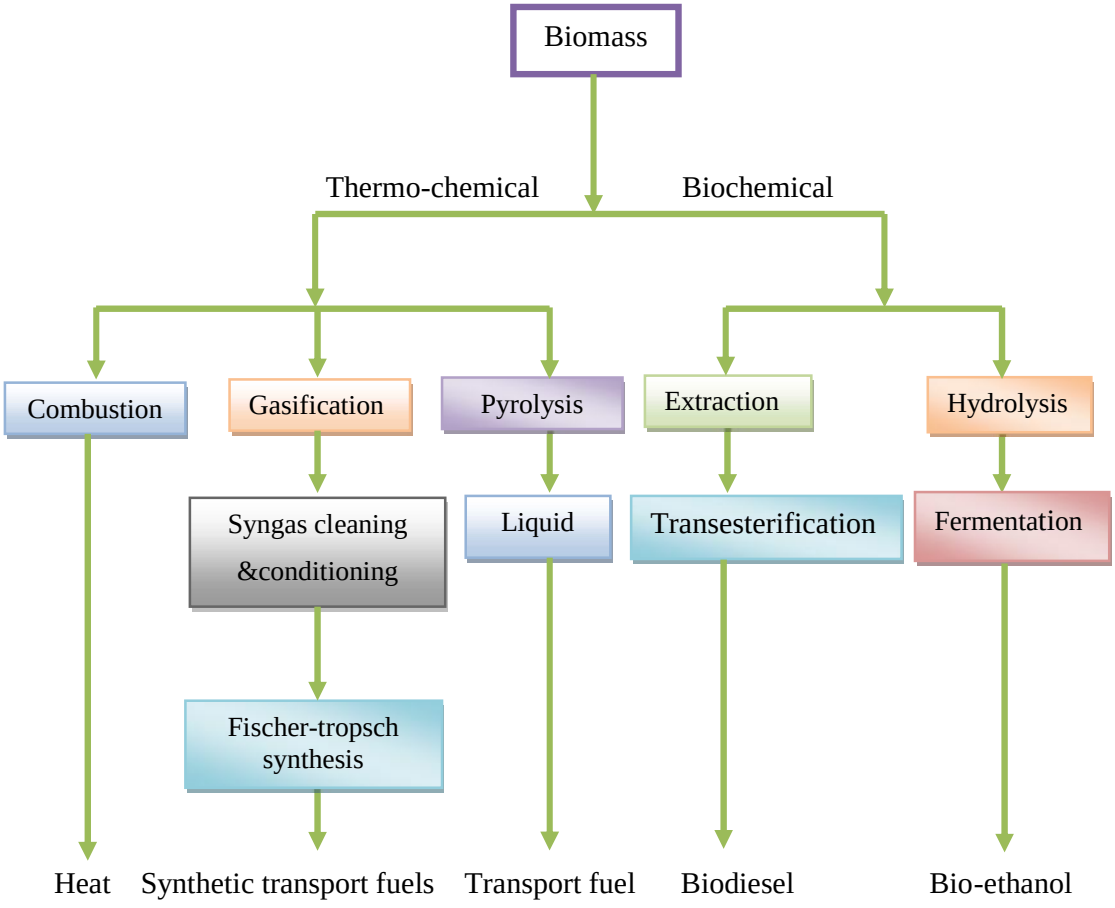

Figure 1-1: Simplified schematic illustration of the two main biofuel production pathways [8]

Biomass is a CO_2 neutral resource for its life cycle [9, 10].Biomass combines solar energy and carbon dioxide into chemical energy in the form of carbohydrates via photosynthesis. The use of biomass as a fuel is a carbon neutral process in which carbon dioxide captured during photosynthesis is released during its combustion and it also possesses zero CO_2 net emission energy [11]. Compared to biomass, burning fossil fuels takes carbon that was locked away underground (as crude oil, gas, and coal) and transfers it into the atmosphere as CO_2. However, biomass combustion, such as wood combustion, only recycles carbon that was already in the natural carbon cycle, with the net effect being that no new CO_2 was added to the atmosphere as long as the forests from which the wood came are sustainably managed [12].This is because biomass gasification is CO_2 neutral and the carbon content of biomass is absorbed by the $CO_2$of the atmosphere through photosynthetic plants for which net CO_2production is zero. The product gas (mixture of CO, CH_4 and H_2) which has a high percentage of hydrogen syngas is advantageous to all other fuels. All these reasons make biomass gasification a promising alternative for heat and power production.

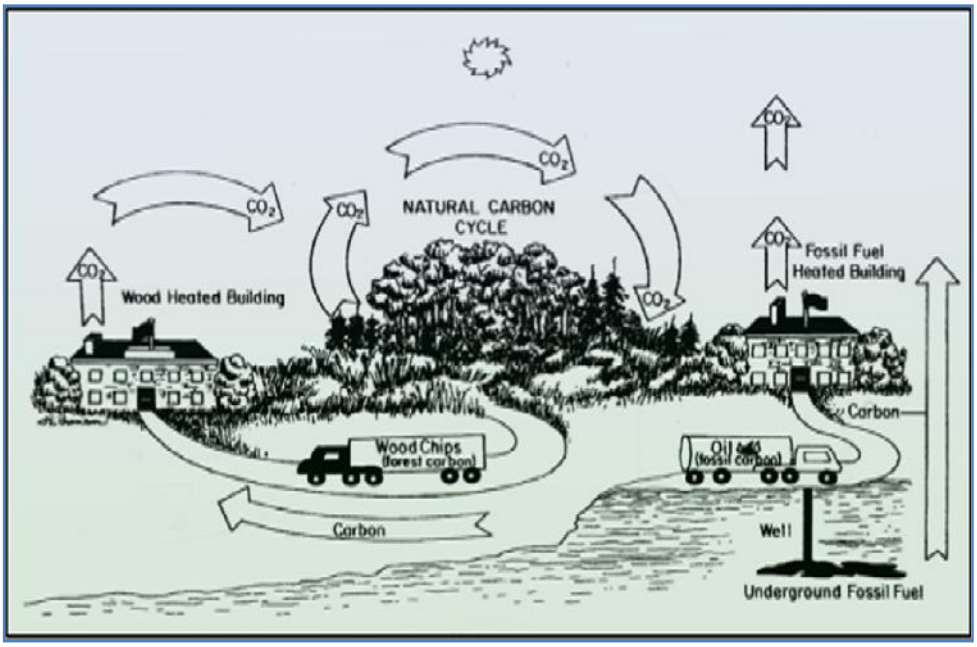

Figure 1-2: Comparison of carbon flows for fossil fuel and biomass heating systems [12].

Biomass gasification is a thermo-chemical process which converts the carbonaceous materials of biomass into a combustible syngas mainly consists of carbon monoxide, Hydrogen, carbon dioxide and methane by partial oxidation in the presence of air, steam or oxygen at high temperature (500 -700℃) within a gasifier.

Char gasification is the endothermic process where the char, solid residue from a pyrolysis process, is transformed into a gaseous mixture of CO, CO_2, CH_4, H_2 and H_2O in a reducing atmosphere usually composed of CO_2 and H_2O. Being char gasification an endothermic process, some source of heat is required. The common heat source is the combustion of the volatile matter released during pyrolysis. The addition of an oxidation agent is necessary for this combustion process. As already mentioned previously, the thermal degradation of biomass in the presence of an oxidation agent should rather be referred as devolatilization and not pyrolysis. Figure 1.3 shows schematically the char gasification process [13].

It is however common to denote as “biomass gasification” the overall process where not only the char is transformed into gas but where all drying, devolatilization, volatile matter combustion and char gasification take place. The biomass gasification process is also referred as “pyrolysis by partial oxidation”.. In drying process, evaporation of water from the solid is takes place. In the Pyrolysis process, the biomass is devolatilized by heat to char and volatiles in the absence of oxygen. Depending on the feedstock, the volatiles include H_2O, H_2, N_2, O_2, CO_2, CO, CH_4, H_2S, NH_3, C_2H_6 and very low levels of unsaturated hydrocarbons like acetylenes, olefins, aromatics and tars. After Pyrolysis, char having a higher carbon concentration than the dry feedstock is converted to gases through gasification by using gasification agents like, air, oxygen, carbon dioxide or steam [13].

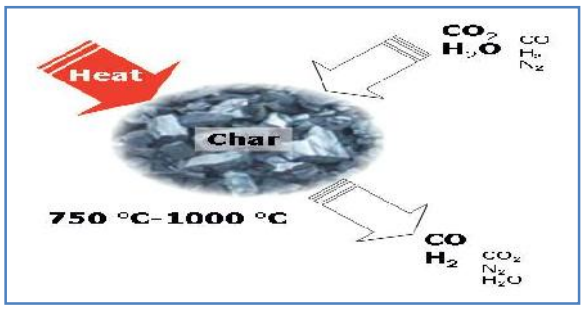

Figure 1-3: Sketch of the char gasification process [13]

The oxidant or gasifying agents can be air, pure O_2steam, CO_2 or mixtures them. Air, while a cheap and widely used gasifying agent contains a large amount of nitrogen, which lowers the heating value of the syngas produced. If pure $O_2$is used as the gasifying agent, the heating value of syngas will increase but the operating costs will also increase due to the O_2 production costs. Partial combustion of biomass with air or O_2 can provide heat for drying the

biomass, raising the biomass temperature and driving the endothermic gasification reactions, and generate water and CO_2 for further reduction reactions [14]. The heating value and H_2 content of the syngas can be increased if steam is used as the gasifying agent, in which case the heating value of the product gas is about 10–15 MJ/Nm^3[15,16], compared with 3–6 MJ/Nm^3 for air gasification of biomass [17,18]. The use of CO_2 as the gasifying agent is promising because of its presence in the syngas. CO_2 with a catalyst such as Ni/Al can transform char, tar and CH_4 into H_2 and CO, thus increasing H_2 and CO contents [19–21]. Pure steam or CO requires an indirect or external heat supply for endothermic gasification reactions [22–25]. Alternatively, a mixture of steam or CO_2 and air or O_2 can be used as the gasifying agent, and the partial combustion of biomass with air/O_2 provides the heat required for gasification [26–29].

1.1. Problem statement

The use of renewable energy sources is becoming increasingly necessary, if we are to achieve the changes required to address the impacts of global warming and energy security. Biomass is the most common form of renewable energy, widely used in the third world. Latterly much attention has been focused on identifying suitable biomass species, which can provide high-energy outputs, to replace conventional fossil fuel energy sources. The type of biomass required is largely determined by the energy conversion process and the form in which the energy is required. The conversion of biomass by gasification into a fuel suitable for use in a gas form increases greatly the potential usefulness of biomass as a renewable resource. Unlike combustion where oxidation is substantially complete in one process, gasification converts the intrinsic chemical energy of the carbon in the biomass into a combustible gas in subsequent thermo chemical conversion stages in robust proven technology that can be operated either as a simple, low technology system based on a fixed-bed gasifier, or as a more sophisticated system using fluidized-bed technology. However, Biomass gasifiers are complex facilities, which make it difficult to investigate their various operating conditions. Most of the time, syngas composition and the thermo chemical processes take place inside the gasifier stages (drying, pyrolysis, reduction and combustion stages) are unknown until the gasification work is conducted. Experimental work is often time and money intensive and a mathematical model predicting syngas composition (concentration of H_2, CO, CH_4 and CO_2) using elemental analysis of biomass would be helpful. Several models such as thermodynamic equilibrium model, kinetics model, Artificial Neural Network model and computational fluid dynamics model has been developed to determine the syngas composition and thermo-chemical process takes place in the gasifier. The numerical simulation techniques such as Aspen Plus thermodynamic equilibrium model offer an effective way of quantifying the physical and chemical process in the biomass thermo-chemical reactors under various operating conditions within a virtual environment.

1.2. Objectives

1.2.1. General objective

- ✓ The general objective of this thesis is biomass gasifier process simulation of fixed bed reactor for sustainable energy production from different biomass feedstock

1.2.2. Specific objective

- ✓ Treatment and characterization of the biomass which used as a feed stock through its proximate and ultimate analysis
- ✓ Sensitivity analysis of various process parameters, including air/fuel ratio, temperature, pressure and steam/fuel ratio on gasification by simulating suggested model.
- ✓ Analyzing the effects of operating parameters on calorific value of gasification products and gasifier performance.

1.3. Significance of the study

Gasifiers convert biomass into a gaseous mixture, small quantities of char and condensable compounds through thermo chemical conversion process. Gasification is considered one of the most efficient ways of converting the energy embedded in biomass to produce heat and electricity, or used in combined heat and power plants, and it is becoming one of the best alternatives for the reuse of waste solids using these reactors. But, due to the inherent complexity of biomass gasifiers to convert biomass to energy, modeling for simulation and performance prediction of the gasifier is still an emerging area of research. Mathematical models have been found to be effective in providing qualitative guidance on number of inter-related parameters concerning the type of fuel, the gasifier design and operating parameters on the gasifier performance. Computational modeling tools are advantageous in many situations due to their capability of allowing the user to find the optimum conditions for a given reactor without going for actual experimentation which is both time consuming and expensive. Thus, mathematical model becomes a useful tool in representing the real life situation viz. the gasification process. This representation of the gasification process in mathematical models helps in gaining an insight about the significance of the operating parameters affecting the gasifier performance and biomass types for energy production as fossil fuel replacement.

1.4. Scope

In the present work, an existing updraft gasifier with comprehensive thermodynamic equilibrium based model has been developed for predicting producer gas composition of an fixed bed rector using ASPEN PLUS and analyzing effects of process parameters on product composition and heating value of syngas.

2. LITERATURE REVIEW

2.1. Biomass Gasifier Technologies

Gasification reactor designs have been researched for more than a century, which has resulted in the availability of several designs at the small and large scales. They can be classified in several ways [30]:

- By gasification agent: Air-blown gasifiers, oxygen gasifiers and steam gasifiers.
- By heat source: Auto-thermal or direct gasifiers (heat is providedby partial combustion of biomass) and allothermal or indirectgasifiers (heat is supplied by an external source via a heatexchanger or an indirect process).
- By gasifier pressure: Atmospheric or pressurized.
- By reactor design: Several biomass gasification reactor designs have been developed and evaluated and can be generally classified into two broad categories; namely, fixed bed and fluidized bed gasifiers.

2.1.1. Fixed Bed Gasifiers

Fixed bed reactors get their name from the movement in the flow of biomass as it passes through the reactor is very slow in the main forward direction. There arethree types of fixed bed reactor: up draft or countercurrent, downdraft or co-current and cross draft. All reactor types are based on natural slowly descending fuel flow caused by gravity. The residence time of fuel in the gasifier is long and the gas velocity is low. Generally these gasifier types are used in small scale energy production <10MW. The traditional fixed bed gasifiers are suitable only sized feed stocks, which have high enough bulk density to guarantee stable fuel flow [31].

i. Up draft

In the updraft mode, biomass is fed from the top and air is supplied at the bottom via the grate of the gasifier. When the biomass is in the top section of the gasifier it is dried and moving downward to reach the devolatilization zone (pyrolysis zone, as marked in Figure 2.1). In the devolatilization zone biomass is decomposed into volatiles and considerable quantities of tars are formed. Following this, in the reduction zone, the volatiles evolve and produce the permanent gas. After that, the residuals of biomass reach to the grate where the solid char is formed and the remaining of biomass is combusted at approximately 1000 ºC. The hot gases

formed in the combustion zone (oxidation zone) start moving in the upward direction into the reduction step. Tar condenses partially on the descending biomass, also leaving the gasifier with the product gas [32]. In this gasification system, the product gas exits from the low temperature Pyrolysis and drying zone, and thus assumed to be contaminated with substantial amount of tars, which is the major problem of updraft gasifiers [33]

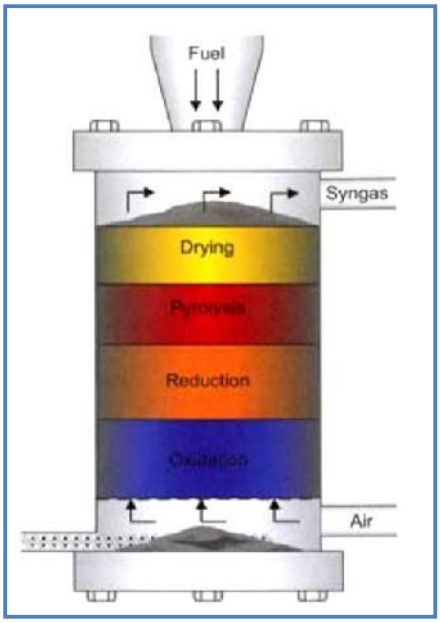

Figure 2-1: Schematic of an updraft gasifier [32]

ii. Down draft

Downdraft gasifiers are similar to updraft gasifiers, except that the locations of the zones are reversed and, as such, the Pyrolysis products are allowed to pass through the high temperature oxidation zone where they undergo further decomposition. The design of the downdraft gasifier is basically the same as the updraft, the major difference being that in the downdraft design, biomass and air move concurrently downward from the top to the bottom (as illustrated in Figure 2.2). The downdraft gasifier has four distinct zones: (1) upper - drying zone, (2) upper medium - Pyrolysis zone, (3) lower medium – oxidation zone and (4) lower - reduction zone. The synthesis gas gets out from the upper medium section, after passing the lower medium zone, where partial cracking of the formed tars occurs, thus resulting in a synthesis gas with low amount of tar. In the oxidation zone, the temperature is about 1000-1400°C and the tars produced are almost exclusively tertiary tars. Particulates and tars in the synthesis gas have low concentration, as the majority of the tar is combusted in the gasifier.

The downdraft gasifier is thus ideal when clean gas is desired [34]. Negative aspects in this design are the low thermal relatively efficiency and the difficulty to process the moisture and ash contents of the biomass.

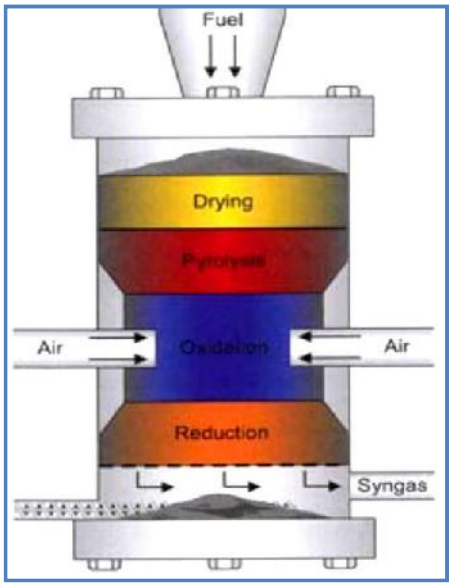

Figure 2-2: Schematic of downdraft gasifier [32]

iii. Cross-draft Gasifier

The cross-flow gasifier is designed with the biomass introduced in the top and the air fed in a lateral side. The synthesis gas exits in the opposite lateral side, more or less at the same level. Biomass moves downwards and it gets dried, devolatilized, pyrolyzed and finally gasified. Figure 4 shows a schematic of a cross-flow gasifier. A hot combustion/gasification zone is formed around the air entrance, with both pyrolysis and drying zones being formed higher up in the vessel. Due to the design of this gasifier, the residence time of the gases in the high temperature zone is small (as the gas enters and leaves from the opposite sides). As a consequence, the temperature of gasification (around 800-900 ºC) is almost the same of the leaving synthesis gas, resulting in less tar cracking, higher contents of tar and lower thermal efficiency than the other designs presented [35].

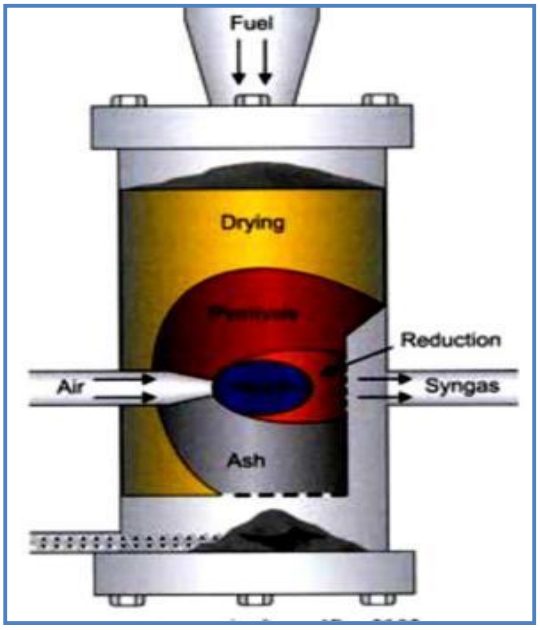

Figure: 2-3: Schematic of cross draft gasifier [32]

Table 2-1*: Advantage and disadvantage of fixed bed gasifiers [36, 37]*

	Gasifier type	Advantage	Disadvantage
1	updraft	o Small pressure drop o Good thermal efficiency o Little tendency of slag formation	o Great sensitivity to tar and moisture content of fuel o Poor reaction capacity with heavy gas load
2	Downdraft	o Flexible adaptation of gas production load o Low sensitivity to charcoal, dust and tar content of fuel	o Design tends to be tall o Not feasible for very small particle size of fuel
3	Cross-draft	o Short design height o Very fast response time to load o Flexible gas production	o Very high sensitive to slag formation o High pressure drop

2.1.2. Fluidized Bed Gasifiers

The operation of fluidized bed gasifiers is based on the fluidization of a bed of grainy porous material offering the advantage of a uniform temperature distribution and better solid–gas contact and heat transfer rates [38]. These gasifiers are equipped with cyclone separators in the top section for the removal of the particulates from the product stream.

A fluidized bed gasifier has a bed made of an inert material (such as sand, ash or char) that acts as a heat transfer medium. In this design, the bed is initially heated and the fuel is fed in relatively quickly some distance from the bottom of fluidized bed. The gasifying agent is provided in the form of a fluidizing gas fed in the bottom of the reactor when the temperature has reached the appropriate level. The bed material transfers heat to the fuel and blows the reactive agent through a distributor plate at a controlled rate [39].

Unlike fixed bed reactors, fluidized bed gasifiers have no distinct reaction zones and drying, Pyrolysis and gasification occur simultaneously during the process. Compared to other gasifiers, fluidized bed gasifiers have strong gas-to-solids contact, excellent heat transfer characteristics, better temperature control, large heat storage capacity, a good degree of turbulence and high volumetric capacity. But they operate at pressures slightly above atmospheric levels and they respond slowly to load changes. Due to their complicated and expensive control systems, fluidized bed gasifiers appear to be commercially viable at larger sizes (> 30 MW thermal output).Fluidized bed reactors are classified by their configuration and the velocity of the reactive agent as bubbling and circulating fluidized bed [40].

i. Bubbling Fluidized Bed

In bubbling fluidized bed gasifiers (Figure 2.4), fuel is fed into the reactor and gases are introduced at a flow rate that maintains pressure at a level sufficient to keep the fuel particles in suspension. The gases are injected into the reactor from the bottom through a gas distributor and pass through the reactor bed in the form of bubbles that rise and grow in size until they reach the surface of the bed, where they burst. The fluidization gas then passes through the bed creating a bubbling section where the reactions take place. The advantages of bubbling fluidized-bed gasification are [41]. Yields a uniform product gas, exhibits a nearly uniform temperature distribution throughout the reactor, able to accept a wide range of fuel

particle sizes, including fines, provides high rates of heat transfer between inert material, fuel and gas and high conversion possible with low tar and unconverted carbon.

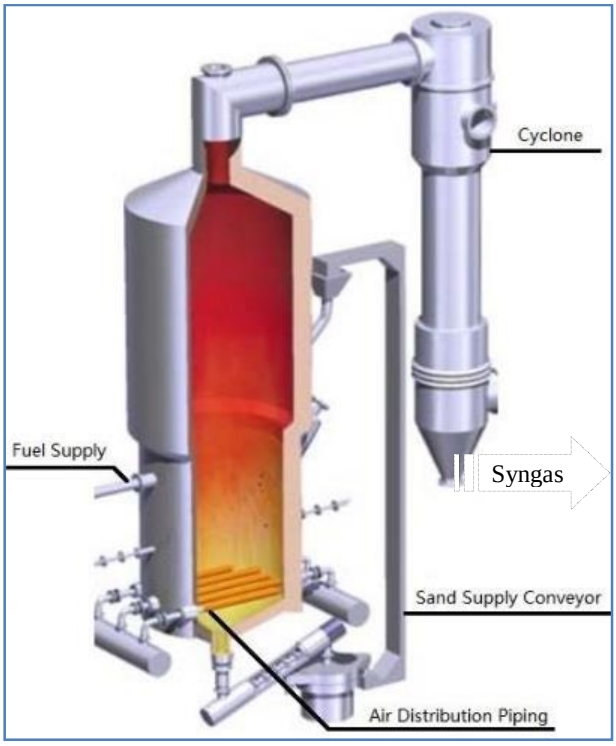

Figure 2-4: Schematic of a bubbling fluidized-bed gasifier [42]

ii. Circulating Fluidized Bed

A circulating fluidized bed gasifier (also called a fast fluidized bed gasifier) is a modified bubbling bed gasifier in which the solids leaving the reactor vessel are returned through an external collection system. The system consists of a refractory-lined reactor where the gasification takes place, a uniform cyclone separator the circulating material from the gas and of return leg for returning circulating material to the bottom part of the gasifier. The operating temperature is typically 800 to 100^0C, depending on the fuel and the application. The fuel is fed into the lower parts of the gasifier above a certain level distance from the air distribution grid. When air entering the reactor, the biofuel particles start to dry rapidly and Pyrolysis also occurs. The gaseous products of drying and Pyrolysis flow upward in the reactor, part of the char flow down to the more dense part of the fluidized bed while part of the char flow upward together with the circulating media into the uniform cyclone. Most of the solids are separated from the gas in the cyclone and returned to the bottom of the bed, where the char is combusted with air that is introduced through the grid nozzle to the fluidized bed [43].

Compared to other gasifiers, circulating fluidized bed gasifiers have a higher processing capacity, better gas-solid contact, suitable for rapid reactions and high heat transport rates possible due to high heat capacity of bed material and the ability to handle cohesive solids that might otherwise be difficult to fluidize in bubbling fluidized beds. Despite these advantages, circulating fluidized beds are still less commonly used because their height significantly increases their cost[41][17]*:*

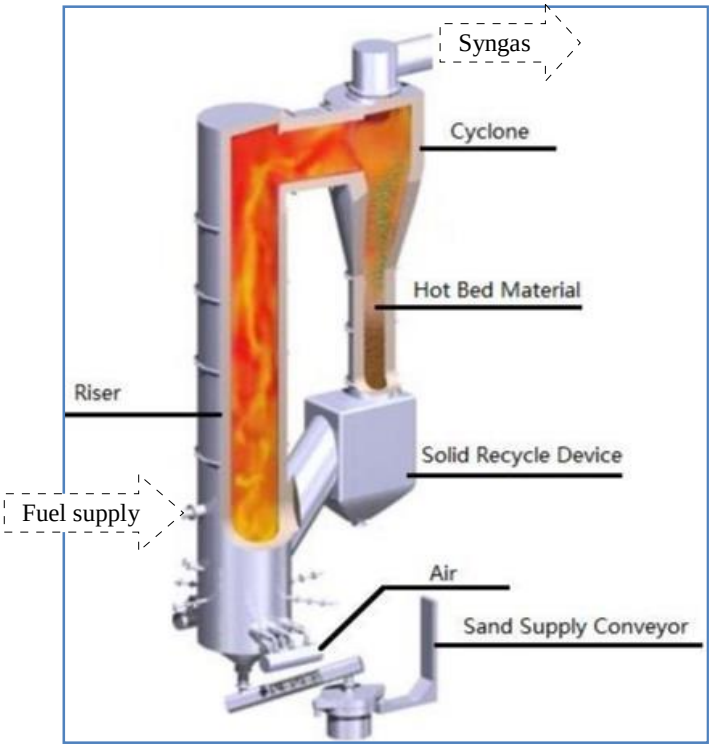

Figure 2-5: Schematic of a circulating fluidized-bed gasifier [42]

2.2. Biomass gasification principles

Gasification is partial thermal oxidation, which results in a high proportion of gaseous products (CO_2, water, carbon monoxide, hydrogen and gaseous hydrocarbons), small quantities of char(solid product), ash, and condensable compounds (tars and oils).Steam, air or oxygen is supplied to the reaction as an oxidizing agent. The gas produced can be standardized in its quality and is easier and more versatile to use than the original biomass (e.g. itcan be used to power gas engines and gas turbines or as a chemical feedstock for the production of liquid fuels). Gasification adds value to low- or negative-value feedstock by converting it into marketable fuels and products.

The chemistry of biomass gasification is quite complex. Broadly speaking, the gasification process consists of the following stages **[**44–47]:

2.2.1. Drying

In this stage, the moisture content of the biomass is reduced, typically, the moisture content of biomass ranges from5% to 35%. Drying occurs at about 100–200^{0}C with a reduction inthe moisture content of the biomass of <5%.

2.2.2. Pyrolysis

This is essentially the thermal decomposition of the biomass in the absence of oxygen or air. In this process, the volatile matter in the biomass is reduced. This results in the release of hydrocarbon gases from the biomass, due to which the biomass is reduced to solid char. The hydrocarbon gases can condense at a sufficiently low temperature to generate liquid tars.

2.2.3. Oxidation/combustion

This is a reaction between solid carbonized biomass and oxygen in the air, resulting in formation of CO_2. Hydrogen present in the biomass is also oxidized to generate water. A large amount of heat is released with the oxidation of carbon and hydrogen. The energy needed for the reduction and pyrolysis reactions is generated at this stage. If oxygen is present in sub-stoichiometric quantities, partial oxidation of carbon may occur, resulting in the generation of carbon monoxide.

2.2.4. Reduction/gasification

In the absence (or sub-stoichiometric presence) of oxygen, several reduction reactions occur in the 800–1000^{0}Ctemperature range. The gasification of biomass char entails several reactionsbetween the char and the gasifying agents which produce CO and H_2. The main reactions that take place in the gasification process are shown in Table 2.

Table 2-2: Main gasification reaction at 25^0C

No.	Reactions	Heat of reaction, [illegible]
1. Char gasification reactions		
R1 (Bourdouard)	$C + CO_2 \leftrightarrow 2CO$	+172 kJ/mol
R2 (steam)	$C + H_2O \leftrightarrow CO + H_2$	+131 kJ/mol
R3 (hydrogasification)	$C + 2H_2 \leftrightarrow CH_4$	-74.8 kJ/mol
R4	$C + 0.5\ O_2 \leftrightarrow CO$	+110 kJ/mol
2. Oxidation reactions		
R5	$C + O_2 \rightarrow CO_2$	-393 kJ/mol
R6	$CO + 0.5\ O_2 \rightarrow CO_2$	-284 kJ/mol
R7	$H_2 + 0.5\ O_2 \rightarrow H_2O$	-242 kJ/mol
R8	$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$	-803 kJ/mol
R9	$CH_4 + 0.5\ O_2 \leftrightarrow CO + 2H_2$	-206 kJ/mol
3. Shift reaction		
R10	$CO + H_2O \leftrightarrow CO_2 + H_2$	-41.2 kJ/mol
4. Methanization reactions		
R11	$CO + 3H_2 \leftrightarrow CH_4 + H_2O$	+206 kJ/mol
R12	$CO_2 + 4H_2 \leftrightarrow CH_4 + 2H_2O$	+165 kJ/mol
R13	$2CO + H_2 \leftrightarrow CH_4 + CO_2$	-247 kJ/mol
5. Steam reactions		
R4	$CH_4 + H_2O \leftrightarrow CO + 3H_2$	+206 kJ/mol

2.3. Factors affecting gasification performance

There are a few parameters that have an effect on the performance of the syngas production during gasification. The performance of biomass gasifiers could be characterized by several parameters. Here, we will review some such parameters: producer-gas composition, which directly influences the heating value of the gas, and gasification efficiency. The composition of the gas obtained from a gasifier depends on a number of parameters, such as fuel composition, gasifying medium, operating pressure, temperature, moisture content of the fuels, the types of the gasifier (mode of bringing the reactants into contact inside the gasifier i.e. gasifier design), etc., and this is very difficult to predict the exact composition of the gas from a gasifier [48]. These factors have been further highlighted in following sections.

2.3.1. Gasifier type

The gasifiers can be divided into different types such as early mention section 2.1. Due to the low and non-uniform heat and mass transfer between the solid biomass feedstock and the gasifying medium in the fixed bed gasifier, large quantities of tar and char were generated. Conversely, fluidized bed gasifiers allowed for good mixing and good gas–solid contact, which increased the reaction rate and the conversion efficiencies. Furthermore, a lower concentration of tar in the product gas was achieved with the usage of the bed material as the heat transfer medium and catalyst, and thus improving the quality of the gas. In addition, efficient heat and mass transfer properties in the fluidized bed gasifier enabled the usage of assorted types of biomass wastes with different compositions and heating values [49, 50].

2.3.2. Temperature

In the gasification of biomass, temperature is one of the most important parameters that can control the gas composition, tar concentration, reaction rate, ash build-up etc. Therefore, it needs to be highly controlled [51]. Low temperature gasification is attributed to high tar content and low CO and H_2content in the product gas [52, 53]. On the other hand, high temperature gasification leads to a desired high yield of CO and H_2, while reducing the tar content. However, two major problems limit high temperature gasification above 1000°C: (1) the ash melting, especially when high ash containing biomass is used suchas rich and wheat straw (ash content around 20%) and (2) the requirement of stringent reactor specification. Therefore, numerous studies have been conducted to investigate the gas composition, tar concentration and other requirements within the temperaturerange of 750–900°C.

According [54], when temperatures greater than 1200–1300°C was utilized, the following results were obtained: little to no methane, the formation of higher hydrocarbons or tar, and maximized production of H_2 and CO without the need for further conversion steps [55]. Rapagnà and Latif [56], who studied the steam gasification of almond shells in a continuous fluidized bedreactor, reported that the highest yield of total gas, H_2, CH_4, CO and CO_2obtained were 1.55, 0.7, 0.42, 0.12, and 0.300 m^3/kg biomass, respectively.

The study indicated that by increasing the temperature from 600 to 800°C, the concentration of char and heavy tars decreased while the overall gas yield increased, and theyields of $H_2$and CO also increased.

Similar studies of [56, 57] also performed a gasification in the fluidized bed using empty fruit bunch as the feedstock and air as the gasifying agent in the temperature range between 700 and 1000°C. When the temperature increased from 700 to 1000°C, the H_2 concentration increased from 10.27 to 38.02 volume%, the CO_2concentrationdecreased, the concentration of CO increased, and the concentration of CH_4increased from 5.84 to 14.72 vol%. By utilizing a higher temperature, the percentage of char and tar decreased, the LHV value increased, and the total gas yield increased then reached a maximum value of ~92 wt% when the temperature was 1000°C.

In other research, Lv *et al*. [58] varied the temperature in the fluidized bed reactor from 700 to900°Cand used sawdust as the feedstock material. The results indicated that higher temperatures contributed to higher gas yields and increased hydrogen production, while the concentration of CO, CH4, and CO_2 decreased. The gas yield and hydrogen composition increased from 1.43 Nm^3/kg biomass to 2.53 Nm^3/kg biomass and21 volume% to 39 volume%, respectively, with the increase in temperature. The carbon conversion efficiency also increased with the temperature increase. When the temperature was increased from700 to 750°C, the LHV value first decreased, and increased to its maximum value of 8.56 MJ/Nm^3, CO, $CH_4$when the temperature was further increased to 800°C. The LHV value started to decrease when the temperature was further increased from 800 to 900°C, and lowest LHV value was recorded at 900°C. Li et al. [59] studied the steam gasification of empty fruit bunch in a fixed bed reactor using a tri-metallic catalyst in the temperature range between 750 and 900°C, and the results demonstrated that higher temperatures contributed to increased hydrogen production and total gas yields. The maximum value of the total gas and hydrogen gas yields were 2.48m^3/kg and 1.481m^3/kg respectively was obtained at temperature of 900°C. As the temperature increased from 750 to 900°C, the concentrations of $H_2$and CO increased, while the concentrations of CO_2 and CH_4 decreased. The LHV of the product gas decreased from 11.26 to 9.13 MJ/Nm^3 when a temperature was increased from750 to 900°C. However, the gasification process of bamboo in a fluidized bed reactor using air in a study that was conducted by[60] yielded a different trend with regards to the effect oftemperature. With an increase in temperature, the concentration of H_2 and CO decreased, while the concentration of CO_2 increased. When the temperature was increased from 400°C to 600°C, the total gas yield remained constant. The maximum LHV value and the cold gas efficiency were recorded when the operating temperature was 500°C.From these studies, it can be concluded that, the higher temperatures contributed to lower concentrations of char and heavy

tars and higher concentrations of H_2gas as well as an increase in the overall gas yield. Higher temperature contributed to higher gas yield due to release of more volatiles.

The increase in H_2 production was due to the tar thermal cracking reaction, which also decrease the tar concentration [61]. According to Le Chatelier's principle, increased temperature favors the products of endothermic reactions and favors the reactants in exothermic reactions. Therefore, the endothermic reforming reactions of hydrocarbon were improved with the increasing temperature (Boudouard, water–gas reaction, and steam-methane reforming reaction. Increase in water–gas and steam-methane reforming reactions resulted in increased H_2 concentration. The increase in activity of steam-methane reforming also resulted in a decrease of CH_4concentration and an increase in $H_2$and CO [58].CO production was favored with increased temperature due to Boudouard, water–gas, and steam-methane reforming [61]. Meanwhile, the exothermic reaction of char partial combustion affects the CO composition in the producer gas. Higher temperature was not favorable for CO production, so the content of CO decreased with temperature. The increase of cold gas efficiency was contributed due to more carbon and steam conversion through endothermic reaction of Boudouard reaction and water–gas reaction [58]. Generally, higher temperature favored hydrogen production and gas yield but did not always favor gas heating value as too high a temperature lowered gas heating value [58].

For instance, an attempt has been made to produce $H_2$for charging a solid oxide fuel cell (SOFC) from sawdust in a downdraft gasifier at a temperature range of 750–1150°C under atmospheric pressure [62]. An increase in CO and H_2 content and a decrease in CO_2 and CH_4were observed when temperature was increased from 650 to 800°C in a bubbling fluidized bed gasifier. The raising of temperature from 750 to 850°C in a fluidized bed gasifier significantly reduced the tar content in the product gas, while increased the CO and H_2 concentration.

2.3.3. Pressure

Depending on the downstream application of the product gas, gasification of biomass is often conducted under atmospheric and high pressures. Some downstream applications of product gas such as the conversion of gas to methanol or to synthetic diesel using Fischer–Tropsch synthesis method need high pressure of product gas, where gasification under pressurized condition is beneficial. In addition, an increase in the gasifier pressure reduces the tar yield in the product gas. However, some investigations conducted in the fluidized bed gasifier have

shown that the concentration of tar, mainly naphthalene, increased with increasing gasifier pressure from 0.1 to 0.5 MPa, and thus the concentration of CO decreased, while CH_4 and CO_2 increased. Experiments conducted with a fluidized bed reactor with external heat source have shown that the proportion of tars increases when the pressure rises from 0.10 to 0.25 MPa, mainly due to the increase in naphthalene, so that the CO content in the syngas decreases and the CH_4 and CO_2 increase [58].

2.3.4. Gasifying agent-biomass ratio

Currently, there are many studies of biomass gasification using different agents such as air, oxygen, steam, or mixtures of these components. Several studies on the effect of gasification agent are discussed as follow. Lv *et al.* [63] made a comparison between air and steam as a gasification agent and reported that gasification using steam was more effective in the maximizing the yield of hydrogen compared to air. 0.49 $m^3$and 0.33 $m^3$of hydrogen per kgof biomass were produced by gasifying pine wood in a downdraft gasifier with steam and air, respectively. It was observed that the concentration of H_2 and CO only reached 52.19, 63.31% for biomass air gasification. The same trend was observed in the research conducted by [60]. The value of the cold gas efficiency and the total gas yield increased when air–steam was used as the medium in the gasification process instead of using just air. The $H_2$and CO gas compositions in the product gas were also higher when an air–steam mixture was usedas the gasifying medium. In contrast, Gil *et al* [64] reported that as gasification with air produced a gas yield of 1.4–2.4 m^3 dry base/ kg of dry ash free biomass that was significantly larger than what was produced in the gasification with steam, which was 0.81 – 1.0 m^3dry basis/ kg of dry ash free biomass. The steam/biomass ratio (SB) also affected the gasification output. In the research conducted by [59], the increase of the steam to biomass ratio from 0 to 1.33 resulted in an increase of the total gas and hydrogen yield to their highest values of 2.48m^3/kg and 1.481m^3/kg, respectively. However, when the steam to biomass ratio was increased from 1.33 to 2.67, the total gas and hydrogen yield started to decrease. The same trend was observed with the H_2gas composition in the product gas. The percentage of the H_2 gas composition increased significantly when the steam to biomass ratio was increased from 0 to 1.33, but it slightly decreased when the steam to biomass ratio was further increased from 1.33 to 2.67. These results were in agreement with the results obtained by [65], who studied the effects of a-cellulose gasification at 800°C and an equivalent ratio of 0.27 using steam to biomass ratio of 0, 0.5, 1, and 1.5. The H_2 gas composition increased significantly from13.50

vol. % to 18.56 vol. % when the steam to biomass ratio was increased from 0 to 1.However, it started to decrease when the steam to biomass ratio was further increased from 1 to 1.5. The steam to biomass ratio value of 1 also produced the highest value of the total gas yield. The lowest LHV value was obtained using steam to biomass ratio of 1.5. Meanwhile, for an air gasification process, the equivalence ratio (ER) may affect the quality of the syngas that is produced. In the research conducted by [57], the ER was increased from 0.15 to 0.35, and as a result, the char and tar yields were reduced from 13.65 to 2.12 wt% and 9.83 to 2.82 wt%, respectively. The product gas LHV decreased slightly from 15.38 to12.35MJ/m^3. The total gas yield increased and recorded its highest value of 86.46 wt% at an ER of 0.35; the H_2concentration increased to its maximum value of 27.42 vol% at an ER of 0.25 and started to decrease when the ER was increased from 0.25 to 0.35. For other gases, the CO_2gas increased significantly while the yields of CH_4 and CO gas decreased. In another research conducted by [61], when the ER was increased from 0.14 to 0.42, the concentration of char and tar slightly decreased. Meanwhile, the total gas yield increased to a maximum value of 0.8 wt% at an ER of 0.21, but it slightly decreased to 0.79 wt% at an ER of 0.42. Both the highest value of the LHV and H_2gas composition was recorded at an ER of0.21, and the lowest values were recorded at an ER of 0.42. With an increase in the value of the ER, the $CH_4$and CO gas yields decreased, but the CO_2 gas yield increased and reached a maximum value at ER of 0.42.

According to [65], the yield of hydrogen could be enhanced with the presence of steam. Based from these studies findings, higher steam to biomass ratio generally contributed to higher total gas yields, H_2yield, and H_2composition. Higher steam to biomass ratio enhanced the steam concentration, thus improve the steam reforming reactions, which resulted in the increase of gas yield, hydrogen yield and $H_2$and CO_2composition [59], but reduced the CO and CH_4composition [66]. This trend agrees well with Le Chatelier's principle, which stated that, the reaction will shift to the side that would counteract the change. Increasing steam to biomass ratio influences the reaction to favor a direction that reduces the concentration of steam. Thus, the high steam rate promotes the gasification and methane reforming forward to produce more $H_2$and also pushes the water gas shift reaction forward to give an upward trend of CO_2. Nonetheless, excessive steam quantity in the gasifier can cause the system to lose a large amount of energy in order to heat up the steam, which is not favorable for the production of energy [65].An excess amount of steam present in the gasifier also may lowered reaction temperature and then produced the low gas quality [59].

Based from these studies findings, higher ER generally contributed to lower char and tar yields, lower LHV, lower concentration of CO, and higher concentration of CO_2. According to [61], the ER affects the gasification behavior in two opposite trends. As the ER increases, the gas quality becomes less attractive due to an increase in the oxidation reactions that lead to an increase in theCO_2 production and decrease in the production of combustible gases. Too high of an ER can result in lower concentrations of $H_2$and CO with an increase in the CO_2 concentration in the output gas [67]. However, higher equivalence ratios also lead to exothermic oxidation reactions that, at the same time, offer more heat to the gasification process, which can optimize the product quality to some extent (the destruction of tar) [61]. But, when the ER value is too small, temperature in the gasifier is low, which is unfavorable for further reactions of biomass gasification gas [63]. Generally, ER value is an important parameter to determine the quality of gas produced from biomass gasification [63].

2.3.5. Moisture Content

Biomass contains moisture in both ways: intrinsically by its nature, and extrinsically wherein moisture is absorbed from the surrounding atmosphere. as any increase in the fuel's moisture content means that more energy is required for water evaporation and steam gasification reactions, which in turn lowers the gasifiers operating temperature. Bed temperatures remain more or less stable with moisture contents below 15% [68]. Even so, the moisture level of the biomass depends on the gasifier in which it is to be processed: in updraft type reactors it may be as high as 50% [69]. Moisture content in the biomass, during gasification, increases CO_2 concentration by the water-shift reaction which consumes CO and liberates H_2 [70]. While the equilibrium constant for water-shift reaction varies little over a wide range of temperatures, the direction tends to reverse at higher temperature. Since more heat is required for moisture evaporation than the small amount of heat gained due to the exothermic behavior of the water-shift reaction, thermal energy inside the gasifier reduces when gasifying biomass with higher moisture content [71]. Thus, the decrease in temperature further exacerbates the scenario and forms more CO_2 since the water-shift reaction is improved at lower temperature. The overall effect is the reduction in calorific value of syngas because, the small increase in H_2isnot sufficient to compensate the loss of significant amount of CO with increase in moisture content [72]. Perez et al. [73] experimented with a downdraft reactor and pine bark as the feedstock, and concluded that the optimum fuel moisture level was 10.62%.

Fluctuations in biomass moisture content lead to variations in bed temperature, and this produces changes in the composition of the syngas generated, with this being one of the causes of instability. Experiments have been conducted by Sulc et al. [74] with variations in fuel moisture levels in a pilot scale gasification unit with a novel co-current, updraft arrangement in the first stage and counter-current downdraft in the second stage. Fluctuating moisture and nitrogen contents in feedstock had less impact on the composition of the resulting fuel gas in the two-stage process. Syngas composition is linked to biomass moisture content. Thus, the molar fraction of CO increases for dry fuels, while for moister fuels the molar fraction of CO increases, reducing the calorific power of the syngas and, therefore, process efficiency, according to tests conducted in an updraft fixed bed gasifier with air [75].

Table 2-3: Summary of moisture content effects on common fixed bed gasifiers

Parameter	Gasifier type	CO	H_2	CH_4	Maximum limit (%wet base)
Moisture content	Updraft	Decrease with moisture content	Increase in some extent	Has no effect	<50%
	Downdraft	Decrease with moisture content	Increase in some extent	Decrease with moisture content	<40
	Fluidized	Decrease with moisture content	Increase in some extent	Decrease with moisture content	<10

A limiting condition called auto-thermal limit is reported as 65% moisture content in literature beyond which self-sustaining gasification is not possible due to an enthalpy deficiency for vaporization. In fact, supplemental fuel is required for most of the combustor when the moisture content is greater than 50% on a wet basis [76].Moisture content up to 30% (wet basis) can be used for downdraft gasifier. When air is used as the gasification agent, the amount of methane produced is small and stays almost constant with change in moisture content. Thus the temperature decrease inside the gasifier due to moisture also results lower mass conversion efficiency and increases tar content. Sheth *et al.* [34] reported the decrease in biomass consumption rate with increase in moisture content which is due to

the higher amount of heat necessary for drying those wood chips inside the reactor before they can be pyrolysis. However, some moisture content is always desirable since it enhances steam reforming and helps to crack tar, and at higher temperature, also enhances other reactions such as char gasification. Steam injection is widely used in industrial applications to adjust syngas composition in the gasification process but often, in the presence of higher temperature provided by some external source.

2.3.6. Biomass Type

Biomass elemental composition has a significant effect on syngas composition. The release of Pyrolysis gas is highly dependent on hydrogen, oxygen and carbon in the biomasses when these content increase, especially with an increase in hydrogen and carbon content. A higher oxygen concentration in biomass needs lower ER for gasification because of its inherent oxygen that will also be available for gasification. Another important factor is the ash content of the feedstock. Although formation of clinkers can cause problems for the gasifier operation with biomass having ash-content above 5%, successful gasification with ash-content up to 25% is reported. Higher ash content causes slugging, and consequently ash agglomeration due to fusion, the rate of which is dependent upon the ash content in biomass and ash composition [52]

2.3.7. Particle Size

Fixed bed gasifiers have lower biomass feedstock size restrictions compared to fluidized bed gasifies. Usually, feed size less than 51 mm and 6 mm is recommended for fixed bed and fluidized bed, respectively. Use of larger size feedstock has been tried and reported by several authors. In addition to the effect of temperature, some researchers have also determined that the feedstock particle size has an effect on the gasification product.

From the research conducted by [57], a biomass particle size between <0.3 and 1.0 mm was studied, and it was determined that feedstock particle sizes not in this range generally caused a blockage in the feeder. In this study, the total gas yield decreased and the char and tar yields increased with an increase in the particle size. The highest gas yield that was obtained was 74.79 wt% using the smallest particle size (<0.3 mm), and the lowest gas yield of 72.74 wt% was obtained using the largest particle size (0.5–1.0 mm).The study also determined that smaller particle sizes produced more CO and CH_4 and lower amounts of CO_2 than the larger

particles; the hydrogen yields remained almost constant (32.99 and 33.93 vol.%) for particles sizes of <0.3 mm and between 0.3 and 0.5 mm, respectively, and then they decreased to 21.57 vol.% for particle sizes between 0.5 and 1.0 mm. The highest LHV of the product gas and the optimum product gas composition were achieved when the feedstock particle size was in the range of 0.3–0.5 mm.

Similarly, Li *et al.* [59] reported that by using particle sizes between <0.15 and5 mm, the total gas yield and H_2 yield decreased. The smaller particles also produced more H_2 and CO_2 and less CH_4, CO and C_2H_4 than the larger particles. However, the LHV value increased with an increase in particle size, and a maximum value of 10.28 MJ/Nm^3 was obtained using the largest particle size.

In another research conducted by [58], the smaller particles produced more CH_4, CO and less CO_2. The smallest particles also produced the highest gas yields, the highest LHV and the highest carbon conversion efficiency, which were 1.53–2.57 Nm^3/kg biomass, 8.7 MJ/Nm^3, and 95.10%, respectively.

In the research conducted by [77], the biomass particle sizes between <0.075and 1.2 mm were used in the steam gasification of pine sawdust at a temperature of 900°C. In this study, the total gas yield, H_2 gas yield and carbon conversion efficiency decreased while the concentration of char and tar increased when the particle size increased. The highest overall gas yield and H_2 gas yield obtained were 1.62 m^3/kg biomass and 0.8 m^3/kg biomass, respectively, when the smallest particle size was used. The study also determined that the smaller particles produced more CO and less CO_2 than the larger particles.

However, a different trend was observed in the study conducted by [77] using a biomass feedstock particle size from <0.15 mm to 3 mm. At a temperature of 850°C,when particle size was increased from <0.15 mm to 0.45–0.9 mm, the dry gas yield first gradually increased from 1.72 to 1.84 Nm^3/kg, but then it slightly declined to 1.81 Nm^3/kg as the particle size increased. A maximum dry gas yield of 1.84 Nm^3/kg was obtained at a particle size of 0.45–0.9 mm. Meanwhile, the composition ofthe product gas remained nearly constant when the particle size was varied,

Saravanakumar *et al [78]* have successfully gasified long sticks with length of 68 cm and diameter of 6 cm successfully in a top-lit updraft gasifier. Bridgewater *et al* [79] reported that

the maximum particle size suggested for a conventional downdraft gasifier with throated design is one-eighth of the reactor throat diameter. The larger particles form bridges preventing the efficient flow of biomass inside a gasifier while smaller particles interfere with the air/gasifying agent passage creating high pressure drop and consequently can result in gasifier shut-down [79].

Smaller particle sizes contributed to higher total gas yields, higher H_2concentrations, and lower char and tar yields. Larger feedstock particle size increase the temperature gradient inside the particle, so that at a given time the core of the particle has lower temperature compared to the particle surface, which resulted to the increase of the char and liquids yields and decrease in gases [57]. Smaller particles can obtain faster heating rate due to larger surface area. High heating rates produce more light gases and less char and condensate. Besides that, the smaller particle size has greater contact area of biomass and steam, which resulted to higher chemical reaction rates and sufficient gasification reactions [59]. Several studies concluded that the effect of particle size was less important compared to other parameters. Although larger particle size increased reaction time needed for the completion of reactions, but at higher temperature, the effect of particle size on gasification may decreased.

2.4. Biomass Gasification Models

Due to the inherent complexity of biomass gasification processes, modeling for simulation and prediction of performance of the processes is still an emerging area of research [8]. Approaches for mathematical modeling of gasification process could be categorized into (i) thermodynamic equilibrium, (ii) kinetic (iii) computational fluid dynamics (CFD) and (iv) Artificial Neural Network (ANN) routes. In the following sections a review of recent work done by various authors using different modeling approaches are being discussed. Each modeling approach has been discussed by giving an introduction to the approach followed by key results from different studies. The studies have been arranged based on the year of publication (in ascending order), the model considerations, the feedstock used and the parameters studied.

2.4.1. Thermodynamic equilibrium model

A thermodynamic equilibrium model is used to predict the composition of the product gas based on the assumption that the reactants react in a fully mixed condition for an infinite

period of time [80]. Thermodynamic equilibrium model predicts the maximum achievable yield of a desired product from a reacting system. In other words, if the reactants are left to react for an infinite time, they will reach equilibrium yield. The yield and composition of the product at this condition is given by the equilibrium model, which concerns the reaction alone without taking into account the geometry of the gasifier. Moreover, thermodynamic equilibrium calculations, which are independent of the gasifier design, may be more suitable for process studies on the influence of the most important fuel and process parameters.

Equilibrium models have two general approaches: stoichiometric and non-stoichiometric. Stoichiometric models are based on equilibrium constants [81]. The stoichiometric approach requires a clearly defined reaction mechanism that incorporates all chemical reactions and species involved. In this modeling approach the specific chemical reactions are identified and used for the estimation of end gas composition. In the non-stoichiometric approach, no particular reaction mechanisms or species are involved in the numerical simulation. The only input needed to specify the feed is its elemental composition, which can be readily obtained from ultimate analysis data [82]. The non-stoichiometric equilibrium model [83] is based on minimizing Gibbs free energy in the system without specifying the possible reactions taking place. The stoichiometric chemical equilibrium model is based on selecting those species that are present in the largest amounts, i.e. those which have the lowest value of free energy of formation. As shown by various authors [84, 85], the two approaches (stoichiometric and non-stoichiometric) are essentially equivalent. A stoichiometric model may also use free energy data to determine the equilibrium constants of a proposed set of reactions.

Equilibrium models are based on some general assumptions that are in better agreement with some specific types of reactors for which equilibrium models have better predictive capabilities. Prins *et al*. [86] presented these assumptions:

- The reactor is implicitly considered to be zero-dimensional.
- The gasifier is often regarded as a perfectly insulated apparatus,i.e. heat losses are neglected. In practice, gasifiers have heatlosses to the environment, but this term can be incorporated in the enthalpy balance of the equilibrium model.
- Perfect mixing and uniform temperature are assumed for the gasifier although different hydrodynamics are observed in practice, depending on the design of the gasifier.

- The model assumes that gasification reaction rates are fast enough and residence time is long enough to reach the equilibrium state.No information about reaction pathways and the formation of intermediates is given in the model.
- Tars are not modeled.

Due to these assumptions, equilibrium models yield great disagreements under some circumstances. Typical pitfalls at relatively low gasification temperatures are the overestimation of H_2 and CO yields and the underestimation of CO_2 , methane, tars and char (in fact, null values for these last three components above 800℃) [87].

Baggio *et al*. [88] conducted an experimental and modeling analysis of a batch gasification pyrolysis reactor with the help of an equilibrium model. The overall agreement between model predictions and experimental results were reasonably satisfactory; in particular the residual solid yield (char) obtained experimentally at 800℃ was very close (in the range ±5% to ±15%) to the value obtained by the simulation. Considering the gas phase, the agreement was quite good also for CO and CO_2; it was fair for H_2, while it was poor for CH_4. It was commented that the computed compositions were representative only of processes where the residence time was long enough to establish thermodynamic equilibrium conditions.

Deydier *et al*. [89] developed a mathematical model for the prediction of the influence of the operating parameters of a gasification process composed of a drying and a gasifying section (travelling bed gasifier). The effect of the ratio of mass flow rate of air used for drying to the mass flow rate of incoming biomass and the ratio of the mass flow rate of air for gasification to the mass flow rate of incoming biomass was studied. An optimal value of the ratio of mass flow rate of air used for drying to the mass flow rate of incoming biomass was found to exist and this value corresponded to the complete and exact drying of the biomass. Optimal value of the ratio of the mass flow rate of air for gasification to the mass flow rate of incoming biomass was found to be associated with the exact and complete gasification of solid carbon.

Nilsson *et al.* [90] developed a model of a new three-stage gasification system and used the model to compare the performance of a three-stage system and a stand-alone fluidized bed gasifier using dried sewage sludge as fuel. The equivalence ratio (ER), steam to oxygen ratio and reactor temperature were the process parameters whose effect on the system was studied.

It was found that for a given steam to oxygen ratio, there was an optimum range of ER within which the cold gas efficiency (CGE) was maximum. As steam to oxygen ratio increased, cold gas efficiency decreased. The Carbon Conversion efficiency increased as reactor temperature increased up to a certain reactor temperature and then remained constant. The results showed that the reforming of tar in the gasifier is not significant for temperatures below 900℃. When increasing thesteam to oxygen ratio the temperature decreased, leading to lessreforming oftar and higher tar content in the gas. These results suggested that the addition of steam is not suitable to enhance tar reforming for atmospheric auto-thermal fluidized bed gasifier.

Bhattacharya *et al.* [91] developed a thermodynamic model to evaluate the yield of hydrogen from biomass through gasification in an oxygen-rich environment followed by carbon monoxide shift reaction with the injection of water. Effect of gasifier equivalence ratio, amount of water injected in the shift reactor for complete conversion of carbon monoxide and percentage of oxygen in the gasifying agent were studied using the model. The hydrogen yield was found to be marginally affected by the percentage of oxygen in the gasifying agent. However, it was commented that higher the oxygen percentage, less would be the hydrogen in the gas mixture and easier would be the purification process to obtain hydrogen. The energy consumption per unit mass of hydrogen generated was higher or the higher purity of oxygen in the gasifying agent.

Pirc *et al.* [92] developed a model for a theoretical universal biomass gasifier capable of producing different syngas compositions. It was reported that when the syngas consisted of methane, hydrogen and carbon monoxide, the highest net efficiency of the system was achieved. Producing syngas for methanol synthesis was found to be interesting area due to its use in a mobile technique. The lowest net efficiency of gasification was achieved when producing a hydrogen-rich syngas; this was because of water reduction, which was an endothermic reaction. It was sensible to use a hydrogen-rich syngas in fuel-cell systems. Using oxygen as an oxidant was found to be efficient in all cases.

Table 2-4: Equilibrium model in the study of some specific gasifier designs.

No.	Authors (years)	Type of gasifier studied	Feedstock used	Parameters studied
1	Baggio et al. (2009)	Indirectly heated batch reactor set in an external furnace	Spruce wood	Char yield

2	Deydier et al. (2011)	Travelling bed gasifier	Coal, wood and grass	Air-fuel ratio for drying and gasification
3	Nilsson et al. (2012)	Three-stage system and a stand-alone fluidized bed gasifier	Dried sewage sludge	Equivalence ratio, steam to oxygen ratio, reactor temperature
4	Bhattacharya et al.(2012)	Oxygen blown biomass gasifier followed by a water gas shift reactor for the production of hydrogen	Wood	Equivalence ratio, amount of water injected in the shift reactor for complete conversion of carbon monoxide and percentage of oxygen in the gasifying agent
5	Pirc et al. (2012)	Universal gasifier	Various biomass	Type of wood, amount of wood moisture, outlet temperature of the syngas, oxidant (oxygen or air)

An equilibrium model for studying biomass gasification with steam in a fluidized-bed gasifier was presented by Schuster *et al* [93]. The results of the equilibrium model for the gasifier were in the range of the measured results, though the CH_4content in the product gas was overestimated. It was shown that the discrepancies in the prediction of the gas composition did not significantly influence the overall efficiency. Li *et al.* [94] used a non-stoichiometric equilibrium model based ob minimization of Gibbs free energy to predict the producer-gas composition from a circulating fluidized-bed coal gasifier. Li *et al.*[95] employed the equilibrium model to predict the producer-gas compositions, product heating value and cold gas efficiency for circulating fluidized-bed gasification. They observed that real gasification processes deviate from chemical equilibrium. Therefore, in order to correct the deviations, they developed a phenomenological model to modify the equilibrium-based framework to account for key non-equilibrium factors. As they knew from a pilot-plant study the experimental carbon conversion and methane yield, it was possible to correct non-equilibrium effects by withdrawing the corresponding carbon and hydrogen from the equilibrium system.

This literature review has shown that equilibrium models are useful tools for preliminary comparison, but that they cannot give highly accurate results for all cases. As mentioned above, thermodynamic equilibrium models do not require any knowledge of the mechanisms

of transformation. Moreover, they are independent of the reactor configuration and not limited to a specified range of operating conditions. They are valuable because they predict the thermodynamic limits of the gasification reaction system. Thus, inorder to describe the behavior of gasifiers more accurately, modifications have been made to equilibrium models. According to Villanueva et al. [96], chemical equilibrium is a good approach when simulating entrained-flow gasifiers in chemical process simulators. Similar conclusions were made, by the same authors, for downdraft fixed-bed gasifiers as long as high temperature and gas residence time is achieved in the throat. In contrast, updraft fixed-bed, dual fluidized-bed and stand-alone fluidized-bed gasifiers should be modeled by revised equilibrium models or,in some extreme cases, by detailed rate-flow models. In these gasifiers, knowledge of biomass de-volatilization is crucial for a successful prediction of performance. In dual fluidized-bed gasifiers, char is converted in the combustor, while in fluidized bed gasifiers, char conversion is rather limited and its prediction seems to be a key parameter for proper prediction.

2.4.2. Kinetic model

Kinetic models provide essential information on kinetic mechanisms to describe the conversion during biomass gasification, which is crucial in designing, evaluating and improving gasifiers. These rate models are accurate and detailed but are computationally intensive [97]. A kinetic model is used to predict the gas yield and product composition that a gasifier achieves after a finite time (or in a finite volume in a flowing medium) [80]. A kinetic model can predict the profiles of gas composition and temperature inside the gasifier and overall gasifier performance for a given operating condition and gasifier configuration. The model studies the progress of reactions in the reactor, giving the product compositions at different positions along the gasifier. It takes into account the reactor's geometry as well as its hydrodynamics. Kinetic models describe the char reduction process using kinetic rate expressions obtained from experiments and permit better simulation of the experimental data where the residence time of gas and biomass is relatively short.

Nikoo and Mahinpey [98] developed a comprehensive process model based on reaction kinetics and reactor hydrodynamics for biomass gasification in an atmospheric fluidized bed gasifier using the ASPEN PLUS simulator. After satisfactory validation of the model using experimental values, the effect of reactor temperature, equivalence ratio, and steam to biomass ratio and biomass particle size were investigated using the model. Higher

temperature was found to improve the gasification process. It increased both the production of hydrogen and the carbon conversion efficiency. Carbon monoxide and methane showed decreasing trends with increasing temperature. Carbon dioxide production and carbon conversion efficiency were found to increase by increasing the ER. Although, hydrogen, carbon monoxide, and methane decreased when ER was increased, increasing steam-to biomass ratio increased hydrogen and carbon monoxide production and decreased carbon dioxide and carbon conversion efficiency. Particle average size did not show a significant influence on the composition of product gases.

Saravanakumar *et al*. [99] developed a computational model to evaluate the anticipated performance characteristics of an updraft fixed bed gasifier utilizing long stick wood as the source of fuel. The computational model indicated that all incoming air was consumed in the charcoal combustion region, and that maintaining specific air/fuel ratio could lead to a poor measure of gasifier performance. Higher combustion temperature was found to enhance the gasification time, but could waste more energy due to energy carried out by the hot exhaust gases. It was commented that air to fuel ratio could be a more useful measure when moisture was present in the lower portion of the bed to maximize/minimize specific gasification products.

Gordillo *et al*. [100] studied a numerical model of a solar downdraft gasifier, utilizing high carbon content feedstock viz. biomass char (biochar) with steam, based on the systems kinetics with char reactivity factor varying exponentially. The model was aimed to calculate the dynamic and steady state profiles while also predicting the temperature and concentration profiles of gasand solid phases, based on the mass and heat balances. The study suggested that downdraft set-up could be a great solution in order to improve the performance of the packed bed and fluidized bed gasifiers with concentrated solar radiation in the upper side of the reactor. The gas produced was found to be high quality syngas, in which the hydrogen was the principal component followed by carbon monoxide; the carbon dioxide yield was small because no combustion was conducted. The system efficiency was reported to be as high as 55% for small steam velocities. The energy conversion efficiency was found to decrease when the steam velocity was increased and when the bed was heated quickly. The model predictions were in very good agreement with the trends found experimentally and reported in the literature. Moreover, varying CRF exponentially was found to improve the representation of the heat transfer throughout the bed.

Inayat *et al*. [101] reported the results of a parametric study performed using process modeling for hydrogen enriched gas production via steam gasification in the presence of calcium oxide. The model incorporated the reaction kinetics calculations of the steam gasification with in-situ CO capture, as well as mass and energy balances calculations. Temperature and steam/biomass ratio were reported as the most important variables, as the hydrogen concentration in the product gas increased on increasing the value of both variables. Hydrogen production efficiency was found to decrease on increasing steam/biomass ratio as more energy was required for additional steam usage despite the increased hydrogen yield. Additionally, it was observed that temperature had a more significant influence on the hydrogen yield compared to the steam/biomass ratio.

Kinetic rate models contain parameters that limit their applicability to different plants. Also, with increasing complexity in the design of the gasifiers, the complexity of the model increases because the models are based on reactor hydrodynamics.

Table 2-5: Kinetic model in the study of specific gasifiers designs

No.	Authors (years)	Type of gasifier studied	Feedstock used	Parameters studied
1	Nikoo and Mahinpey (2008)	Fluidized bed gasifier	Pine wood	Reactor temperature, equivalence ratio, steam to biomass ratio and biomass particle size
2	Saravanakumar et al. (2011)	Updraft fixed bed gasifier	Long stick wood	Air-fuel ratio, gasification temperature
3	Gordillo et al. (2011)	Downdraft solar packed bed gasifier	High carbon contentfeedstock	Gas flow rate, reactor height, reactor temperature
4	Inayat et al. (2012)	Steam gasifier	Oil palm empty fruit bunch	Temperature and steam to biomass ratio

2.4.3. Computational fluid dynamics

Computational fluid dynamics can have an important role in the modeling of a fluidized-bed gasifier. A CFD-based code involves a solution of conservation of mass, momentum, species, and energy over a defined domain or region. The equations can be written for an element,

where the flux of the just-mentioned quantities moving in and out of the element is considered with suitable boundary conditions [102].

2.4.4. Artificial Neural Network Models

Neural Network Models is an alternative to the sophisticated modeling of a complex process, especially for one not well understood, is an artificial neural network (ANN). An ANN model mimics the working of the human brain and provides some human characteristics in solving models. It cannot produce an analytical solution, but it can give numerical results. This technique has been used with reasonable success to predict gas yield and composition from gasification of bagasse, cotton stem, pine sawdust, and poplar in fluidized beds; in municipal solid waste; and also in a fluidized bed [103].

Non-mechanistic, non-equilibrium modeling using neural networks for biomass gasification has also been reported [104, 105]. Artificial neural networks (ANN) have been extensively used in the fields of pattern recognition; signal processing, function approximation and process simulation. Sometimes a hybrid neural network (HNN) model is synthesized for process modeling [106].This modeling approach usually combines a partial first-principles model, which describes certain characteristics of the process being simulated and involves a multilayer feed forward neural network(MFNN) that serves as an estimator of unmeasured process parameters that are difficult to model from first principles. MFNN is a universal function approximator, which has the ability to approximate any continuous function to an arbitrary precision even without a priori knowledge of the structure of the function tobe approximated [100]. In Guo *et al*. [104], HNN model was developed to simulate biomass gasification in a steam fluidized bed gasifier. A series of gasification runs were conducted on abench-scale facility, with four types of biomass as feedstock. The data obtained from these experiments were used to train the HNN model.

2.4.5. Aspen Plus gasification models

Some authors, trying to avoid complex processes and develop the simplest possible model that incorporates the principal gasification reactions and the gross physical characteristics of the reactor, have developed models using the process simulator Aspen Plus [108]. Aspen Plus is a problem-oriented input program that is used to facilitate the calculation of physical, chemical and biological processes. It can be used to describe processes involving solids in

addition to vapor and liquid streams. Aspen Plus makes model creation and updating easier, since small sections of complex and integrated systems can be created and tested as separate modules before they are integrated. This process simulator is equipped with a large property data bank containing the various stream properties required to model the material streams in a gasification plant, with an allowance for the addition of in-house property data. In recent years, the simulation of biomass gasification has been presented using Aspen Plus in various reactors and by obtaining a series of research findings.

Mathieu and Dubuisson [109]simulated sawdust gasification process with air using aspen plus in fluidized bed reactor and analyzed the influence of air temperature, air oxygen content, and operating pressure on gasification. Gao *et al*. [110] used aspen plus to establish an interconnected fluidized bed model based on non-catalytic rice raw gasification with limestone and discussed the effects of gasification temperature and steam to biomass ration the process. Similarly, Zhang*et al*. [111] established an interconnected fluidized bed model using aspen plus to simulate the straw gasificaion: they analyzed the effects of gasification temperature, pressure, and steam to biomass ratio on gasification on the yield of methanol.

Nikoo and Mahinpey [112] constructed a biomass gasification model with aspen plus in fluidized bed and discussed the effects of gasification temperature, equivalence ratio, partial average size, and steam-to-biomass ratio on gasification result. However, the neglect of incomplete carbon conversion, there are some deviation between simulated and experimental results. The overall trend in the experimental and predicted results was identical, demonstrating that the Aspen Plus software can be used for simulation of biomass gasification. Chen *et al*. [113] developed an air gasification model in fluidized bed with aspen plus and compared peanut shell gasification results to verify the accuracy and reliability of the model. Aspen plus software can thus be applied to the simulation of agricultural waste gasification. Turn *et al*. [114] experimentally investigated with increase in temperature hydrogen yield and total gas yield increases yields which can be attributed to increased steam and carbon dioxide gasification reaction rates brought about by higher reactor temperatures. Higher hydrocarbon concentrations decreased as reactor temperature increased, the result of more favorable conditions for thermal cracking and steam reforming reactions. With decrease in Equivalence ratio the hydrogen and gas yield increases. With increasing the steam to biomass ratio the hydrogen and gas yield increases but it is least sensitive parameter compared to others. Nikoo M. B. and Nader M. developed [115] a process model is for

biomass gasification in an atmospheric fluidized bed gasifier using the ASPEN PLUS simulator. The model addresses both hydrodynamic parameters and reaction kinetic modeling. Using pine sawdust as raw material they investigated that temperature increases the production of hydrogen. Equivalence ratio is directly proportional to carbon dioxide production and carbon conversion efficiency. With increase in steam to biomass ratio hydrogen and carbon monoxide production increases and carbon dioxide and carbon conversion efficiency decreases

Other authors have worked with Aspen Plus to model the gasification process for coal and biomass. Yan and Rudolph [116] developed a model for a compartmented fluidized-bed coal gasifier process; Sudiro *et al.* [117] modeled the gasification process to obtain synthetic natural gas from pet coke. Pavi *et al* [118] describes a very simple two-step model of chemical equilibrium in the wood biomass gasification process. Robinson and Luyben [119] presented an approximate gasifier model that can be used for dynamic analysis using Aspen Dynamics. They used a high molecular-weight hydrocarbon that is present in the Aspen library as a pseudo fuel and the proposed approximate model captured the essential macro scale thermal, flow, composition and pressure dynamics. Doherty *et al.* [120, 121] developed a model for a circulating fluidized bed and studied the effect of varying the equivalence ratio, temperature, level of air preheating, biomass moisture and steam injection on the product gas composition, the gas heating value and the cold gas efficiency.

3. MATERIALS AND METHODS

3.1. Materials and Equipments used

The gasification process was simulated using three different biomass types available in Ethiopia. Rice husk and coffee husk from farm fields, saw dust from furniture industries by-product were chosen to present non-wood biomass. The agricultural residue is important because of its abundant supply or availability within the region. The analysis of fuel samples was done through ultimate and proximate properties of the biomass in laboratories of Addis Ababa University.

Oven, electronic balance, furnace, crucible, and Desiccators were used in proximate analysis experiments of the samples. Oven was used to determine the moisture content of the biomasses. Electronic balance was used to determine the weight of samples throughout moisture content, volatile matter and ash content determination of the biomass samples. Furnace was used to determine the volatile mater and ash content of samples. Crucibles were used to hold the samples inside furnace during volatile mater and ash content determination experiments. Desiccators were used to prevent the samples from absorbing moisture in the surrounding once moisture content, volatile matter and ash content experiments were done. CHNS analyzer (EA 1112 Flash CHNS/O- analyzer) was used to determine the elemental composition of biomasses.

Figure 3-1: Rice husk, coffee husk and saw dust biomass samples

3.2. Experimental Methods

Experimental phase include proximate analysis and ultimate analysis determination. All the analysis was conducted according to the Standard methods.

3.2.1. Proximate Analysis

Proximate analysis, which is a standardized procedure that gives an idea of the bulk components that make up a fuel, was done to determine the average percentage volatile matter content, percentage ash content, moisture content and percentage content of fixed carbon of the biomass. The procedure of the ASTM standard was adopted to obtain the following parameters:

i. Moisture content

Moisture was determined by using standard oven dry method. Standard method for moisture determination involves heating of 1 gm biomass sample in a hot air oven Model PHP –30at 105^{o}Cusingthe following procedure.

1. Prepare a representative portion of the samples to be tested
2. Determine the mass of the test sample and record this mass as the “wet mass”.
   - ❖ The most convenient procedure for determining the mass of the sample before and after drying is to place it in crucible where it will remain throughout the test. The mass of the container and sample are determined and the mass of the crucible subtracted.
   - ❖ If the mass of the test sample is not determined immediately after preparation, place the desiccators to prevent evaporation
   - ❖ Dry to constant mass at 105°C.The drying time required to achieve constant mass will vary depending on the type, quantity, and condition of the material. In most cases, an overnight (16 h) drying period is sufficient.
   - ❖ To reduce the drying time, break lumps of material into small fragments and spread in a thin layer over the bottom of the containers. Position the containers in the oven to allow the maximum air circulation and exhaust of the moisture laden air.
   - ❖ Constant mass has been achieved when subsequent drying periods to verify constant mass shall be of at least 1 h duration.
3. Remove the sample from the drying device and cool to room temperature.

Note: If the mass of the test sample is not determined immediately after cooling, placethe desiccators to prevent absorption of moisture from the air.

4. Determine the mass of the test sample and record this weight as the “dry mass”. Finally,

The percentage moisture content (PMC) was found by weighing of the biomass sample and oven drying it at 105°C until constant weight of the sample was obtained. The loss in weight resulted in the amount of moisture present and sample left are total solids present in the sample. The change in weight was then used to determine the samples’ PMC using the equation below:

$$PMC = \frac{W_1 - W_2}{W_3} \times 100 \quad (3.1)$$

W_1= the weight of crucible and raw samples (rice husk, coffee husk and saw dust) (g)

W_2 = the weight of crucible and oven dried samples (rice husk, coffee husk and saw dust) (g)

W_3= weight of raw samples taken (rice husk, coffee husk and saw dust) (g)

ii. Ash Content

Ash is defined as the weight of the residue remained after complete burning of 1gm of the biomass at 550°C. The percentage ash content (PAC) was determined by heating oven dried biomass samples in open silica crucible at 550°C for 6 hours in a Box type resistance furnace Model BK –5 –12GJhaving hearth dimension: 300×200×120mm, design temperature: 1200°C, voltage: 220 V and power: 5kW specification as shown in the figure 3.2.

Figure 3-2: Ash content determination experiment

Then samples were cooled in desiccators and weighed to obtain the weight of ash.

$$PAC = \frac{W_4 - W_5}{W_6} \times 100 \quad (3.3)$$

W_4= the weight of crucible and air dried sample (g)

W_5 = the weight of crucible and residue (g)

W_6= weight of air dried or raw sample taken in (g)

iii. Volatile Matter

It is termed as the weight loss due to heating of 1gm of biomass at 925^{o}C ± 5 C in furnace for 7 minutes.Thenoven dried samples were weighed using electronic balance andthe oven dried biomass sample pulverized in aopen silica crucible and placing it at 925^{o}C for7 minutes in a Box type resistance furnace Model BK –5 –12GJ as indicated in the figure 3.3.

Figure 3-3: Volatile mater content determination experiment

The PVM was then calculated using the Equation below:

$$PVM = \frac{W_2 - W_3}{W_2} \times 100 \quad (3.3)$$

W_2 = the weight of crucible and oven dried sample (g)

W_3= weight of crucible and heated sample taken (g)

iv. Fixed Carbon

The percentage fixed carbon (PFC) was computed by subtracting the sum of PVM, PAC and PMC from total of 100 % composition as shown in the Equation below:

$$\text{Fixed Carbon} = 100\ \% - (\%\ Ac + \%\ VM + \%MC) \tag{3.4}$$

3.2.2. Ultimate analysis

This analysis is important for determining the elemental composition (C, N, H, S, O etc.) of the biomass fuels and the empirical formula of biomass as $CH_xO_yN_z$ in order to establish stoichiometric equation for complete combustion of biomasses and equivalence ratio for the gasification reaction. It was carried out using CHNS analyzer(EA 1112 Flash CHNS/O-analyzer) under carrier gas flow rate of 120 ml/min, reference flow rate 100 ml/min, oxygen flow rate 250 ml/min; furnace temperature of 900°C and oven temperature of 75°C conditions at Addis Ababa University, Department of Chemistry.

3.3. Empirical Formula and Stoichiometric Air to Fuel Ratio for Air Gasification

The mathematical model of gasification process was created to determine the possibilities of non-wood biomass use for syngas production. Model was based on the thermodynamic equilibrium reactions. Model describes gasification process for the updraft gasifier where air or steam was used as gasifying agent. The oxygen and nitrogen make 100% of air in the model and steam at 120°C and 1 bar was used. Similarly, the fuel and ambient temperatures were chosen as constant in the model for all biomass types. It was decided to take the air flow rate for all biomass types to make comparison more credible. The model was based on the global gasification reaction:

$$CH_xO_yN_Z(\text{biomass}) + wH_2O + m\,(O_2 + 3.76N_2) \rightarrow nCO_2 + nH_2 + nCH_4 + nCO + nH_2O + \left(\frac{z}{2} + 3.76m\right)N_2 \tag{3.5}$$

Where: n- the numbers of moles of CO_2, H_2, CO, H_2O CH_4, N_2 in the syngas; m–the air moles; w–the moisture associated with the biomass.

The empirical formula of biomasses as $CH_xO_yN_z$was computed based on experimental results of section 3.2.1 and 3.2.2in order to establish the stoichiometric equation for complete combustion of biomasses and air to fuel ratio for the gasification reaction.Therefore,it can be computed by considering the equation of biomass as $CH_xO_yN_z$, where x, y and z is subscripts.

Assume, $C = 1$, normalized with respect to carbonand using the formula for samples;

$$x = \frac{\text{mass fraction of Hydrogen} \times \text{Molecular weight of Carbon}}{\text{mass fraction of Carbon} \times \text{Molecular weight of Hydrogen}}$$

$$\text{y} = \frac{\text{mass fraction of Oxygen} \times \text{Molecular weight of Carbon}}{\text{mass fraction of Carbon} \times \text{Molecular weight of Oxygen}}$$

$$\text{z} = \frac{\text{mass fraction of Nitrogen} \times \text{Molecular weight of Carbon}}{\text{mass fraction of Carbon} \times \text{Molecular weight of Nitrogen}}$$

The stoichiometric relation for complete oxidation of the biomass samples

$$\text{Air} - \text{to} - \text{fuelratio}\left(\frac{\text{A}}{\text{F}}\right)\text{stoichio} = \frac{\text{Volune of air}}{\text{mass of biomass samples}}$$

$$\text{Steam to Biomass ratio(SBR)} = \frac{\text{mass of steam}}{\text{mass of biomass samples}}$$

3.4. Process Simulation Model Development

A countercurrent gasification process schematic in figure 3.4 consists of continuously filled with biomass chips; while air or a mixture of air and steam is injected from the bottom was selected for the current study. Combustion takes place here, supplying heat for the globally endothermic processes of gasification, Pyrolysis and drying, for preheating the cold feed. These processes are stratified along the reactor height because biomasses particles encounter a gas at successively higher temperatures while descending toward the grate.

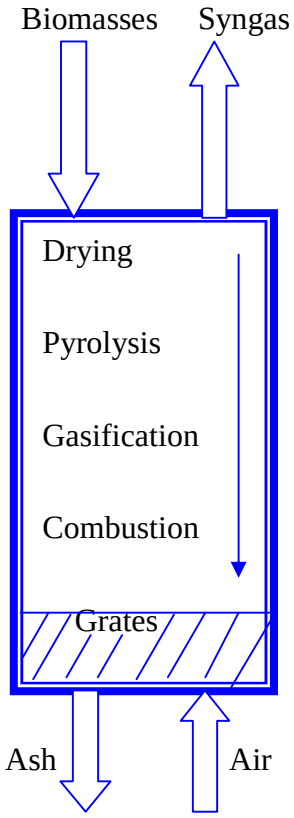

Figure 3-4: Updraft (Countercurrent) biomass gasifier.

Aspen Plus was used in this study to develop a model and simulate this gasification process for rice husk, coffee husk and saw dust feed stocks. The simulations of the biomass gasification process were based on the mass-energy balance and chemical equilibrium for the overall process. Aspen Plus is based on "blocks" related to unit operations as well as chemical reactors, through which most industrial operations can be simulated. It comprises several data bases containing physical, chemical and thermodynamic data for awide variety of chemical

compounds, as well as a selection of thermodynamic models required for the accurate simulation of any given system[18].

The main processes (drying, Pyrolysis, gasification and combustion) were simulated by RStoic, RYield and RGibbs based on the following assumptions: The model is steady state, isothermal, there is no pressure loss, no information about reaction kinetics (the detail kinetic data for each gasification reaction is not given in the model) and chemical reactions take place at an equilibrium state in the gasifier. All elements except sulphur content take part in the chemical reaction, all gases are ideal gases, including H_2, CO, CO_2, steam (H_2O), N_2 and CH_4, Char contains mainly Carbon, volatile matters composed of carbon, H_2 and O_2 and ash in solid phase and tars are assumed as non-equilibrium products to reduce the hydrodynamic complexity [79].The gasification process begins with the decomposition (Pyrolysis) region and continues with the combustion region. The relevant reactions considered in these processes are.

$$2C + O_2 \rightarrow CO \quad (\text{partial oxidation reaction}) \tag{3.6}$$

$$C + O_2 \rightarrow CO_2 (\text{complet oxidation reaction}) \tag{3.7}$$

$$C + 2H_2 \leftrightarrow CH_4 (\text{hydrogasi ication reaction}) \tag{3.8}$$

$$C + H_2O \leftrightarrow CO + H_2 (\text{water} - \text{gas reaction}) \tag{3.9}$$

$$CO + H_2O \leftrightarrow CO_2 + H_2 (\text{shift reaction}) \tag{3.10}$$

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 (\text{steam reforming reaction}) \tag{3.11}$$

$$C + CO_2 \leftrightarrow 2CO \ (Boudourd\ reaction) \tag{3.12}$$

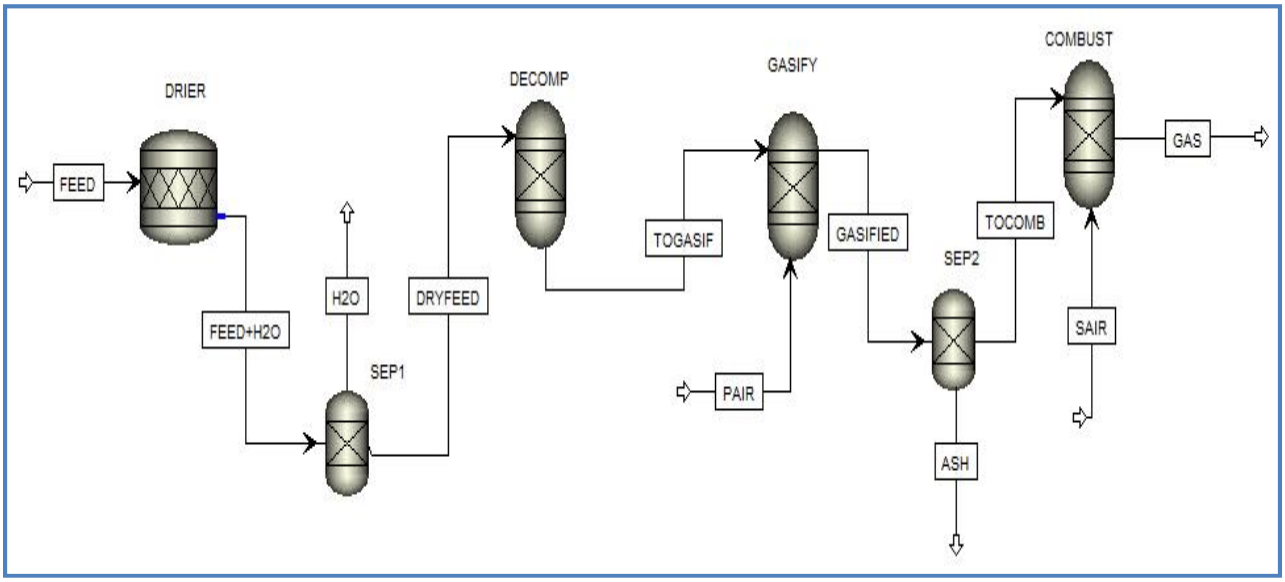

Figure 3-5: Simulation models for gasification process

3.4.1. Simulation Models Description

The purpose of drying is to reduce the moisture content of the feedstock. The Aspen Plus stoichiometric reactor, RStoic (modelID: DRIER), was used to simulate the evaporation of moisture. RStoic converts a part of feed to form water which requires the extent of reaction known as;

$$Feed \rightarrow 0.0555H_2O$$

In this step, the moisture of each feedstock is partially evaporated and then separated using a separator model, (model ID: Sep1) through split fractionation of the components. The dried feedstock was placed into the next region for decomposition after being separated from the evaporated moisture. The evaporated moisture was drained out from the process.

Pyrolysis is one of the main steps of the gasification process where each feedstock is decomposed into its constituent elements. The Aspen Plus yield reactor, Yield (model ID: Decmpose), was used to simulate the decomposition of the feed. The yield reactor converts non-conventional feed into conventional components.

In this step, feed is converted into its components including carbon, O_2, N_2, H_2, sulphur and ash by specifying the yield distribution according to the feedstock's ultimate analysis. The decomposed elements then reacted with air at Gibbs reactor are ready for gasification. The

RGibbs reactor is a rigorous reactor for multiphase chemical equilibrium based on Gibbs free energy minimization. A separator model, (model ID: Sep2) was used to separate ash from the gas mixture using split fractionation of the components.

Another RGibbs reactor (model ID: Combust) was used in the combustion section with minimum air mixing to complete the gasification process. This combustion process is also based on the principle of minimization of Gibbs free energy.

3.4.2. Simulation Modeling Sequence

In this study, the developed Aspen Plus model for a fixed bed gasifier involves the following sequential steps:

i. Define the process flow sheet
ii. Stem and sub-stream class specification
iii. Component specification
iv. Property method selection
v. Feed stream specification (specify all thermodynamic properties, flow rate, composition)
vi. Block specification (specify all thermodynamic properties, operating parameters)

i. Defining the process flow sheet

The feeds (rice husk, coffee husk and saw dust) are fed to the process unit drier, and, the water bound to the biomass is vaporized. The dried feed and the water vapor are fed into separator model. This block allows splitting the feed directly from the knowledge of the compositions, without the need to define reaction stoichiometry and reaction kinetics. The separated water is drained out of the process and the dry feed continues to the next block, where the second step of gasification starts with the decomposition of the dried feed. The decomposed biomasses mainly consisting of C, O_2,H_2, N_2, HCl, H_2S, ash and here, C will partly compose the gas phase to take part in de-volatilization and the remaining part of carbon comprises the solid phase (char) consequently is later introduced in the char gasification reactor. Then, the generated stream gets through the mixer with air steam to perform a perfect mixing in the reactor and to start the new

level of the process, which is the gasification step. In the third step, gas phase reactions including the partial combustion reaction of combustible gases (CO, H_2), the water-gas shift reaction and the steam-methane reforming reaction are takes place during the gasification of char particles and the syngas and the ash leaves stream. The ash and the gas leaving the stream separated in block separator and passes though last steps of gasification process where the remaining fixed carbon in the biomass combusted with air. The syngas is then separated from unwanted products (HCl, H_2S, and water vapor) through separator model in indicated figure 3.6.

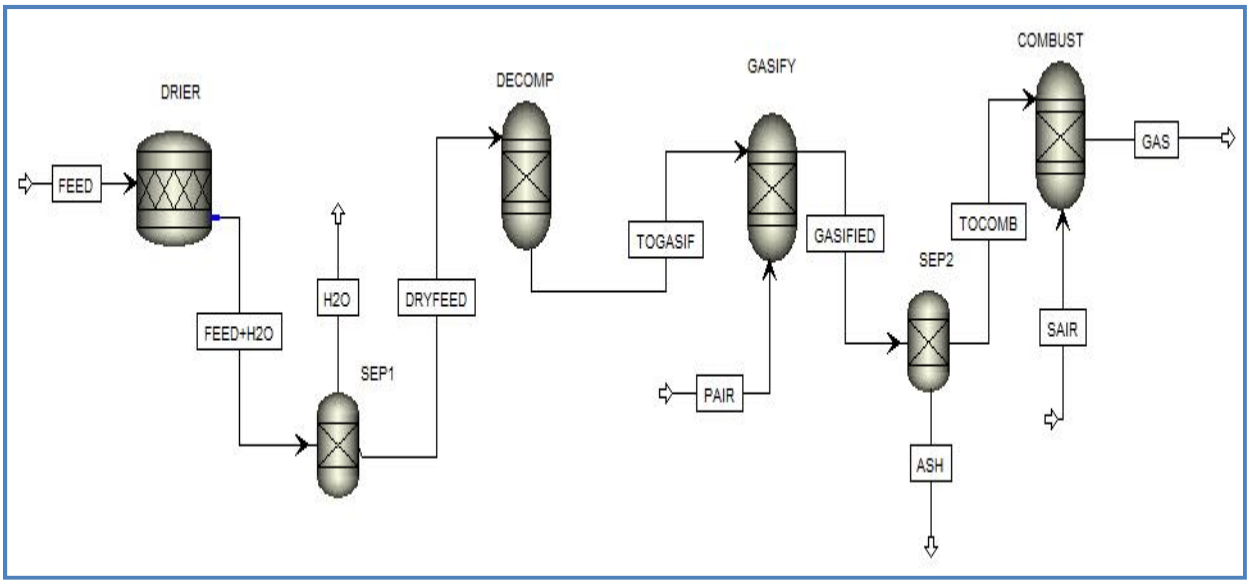

Figure 3-6: Aspen plus flow sheet of the gasification process

ii. Stream and sub stream class specification

Stream classes are used to define the structure of simulation streams wheninert solids are present. The default global stream class for most Aspen simulations is CONVEN. The CONVEN stream class has a single sub stream i.e. the MIXED sub stream. By definition, all components in the MIXED sub stream participate in phase equilibrium whenever flash calculations are performed. To introduce inert solid components to a simulation, one or more additional sub streams must be included. Aspen Plus has two other types of sub streams available (i.e. the CISOLID sub stream type and the NC sub stream type).

The CISOLID sub stream (Conventional Inert Solid) is used for homogeneoussolids that have a defined molecular weight. The NC sub stream (Nonconventional) is used for heterogeneous

solids that have no defined molecular weight. Both the CISOLID sub stream and the NC sub stream give the option of including a Particle Size Distribution (PSD) for the sub stream. Sub streams are combined in different ways to form different stream classes.

The MIXNCPSD stream class contains two sub streams: MIXED and NCPSD.The default stream class of the Solids application type, MIXCISLD, is insufficient for the simulations where there will be NC sub stream with particle size distribution for the feed biomass.

In this simulation, the stream class for global was specified as “MIXCINC”. This option was for the situation when both conventional and nonconventional solids are present without particle size distribution of the feed.

iii. Component specification

At the beginning of simulation, all the components were specified properly. Table 3.1 lists the components modeled in the simulation. By default, Aspen Plus assumes all components are of the type Conventional, indicating that they participate in phase equilibrium calculations. Because the uncertainty of exact formulas of biomass and ash, they were defined as nonconventional solid components. For these components, only enthalpy and density were calculated during the simulation. Aspen Plus includes special models for estimating both enthalpy and density for coal-derived materials. These models can be used to estimate biomass properties as well since biomass can be considered as coal-derived material.

Table 3-1: Detailed components modeled in the simulation

Formula	*Type*	*Component Name*
Biomass	Nonconventional	-
C	Solid	Carbon-Graphite
CO_2	Conventional	Carbon-Dioxide
CO	Conventional	Carbon-Monoxide
H_2	Conventional	Hydrogen
H_2O	Conventional	Water
N_2	Conventional	Nitrogen

O_2	Conventional	Oxygen
CH_4	Conventional	Methane
Cl_2	Conventional	Chlorine
S	Conventional	Sulfur
Ash	Nonconventional	-
HCl	Conventional	Hydrogen-Chloride
NH_3	Conventional	Ammonia
H_2S	Conventional	Hydrogen-Sulfide
SO_2	Conventional	Sulfur-Dioxide

iv. Property method selection

The BK10 property method was selected as the global property method for this model. BK10has been used to estimate all physical properties of the conventional components in gasification process. This method is generally used for non-polar or mildly polar mixtures, like hydrocarbons and light gases such as CO_2, hydrogen sulfide and H_2O with reasonable results at all temperatures and pressures. Biomass and ash were defined as nonconventional components, additionally; the HCOALGEN and DCOALIGT models were selected to calculate the density and enthalpy during simulation.

v. Feed stream specification

The stream 'FEED' was a non-conventional stream and its thermodynamic property was specified by its proximate and ultimate analyses, obtained from the experimental methods in section 3.2.1 and 3.2.2. The biomass lower heating value (*LHV*) was also specified with the HCOALGEN and DCOALIGT property models chosen to estimate the biomass enthalpy of formation, specific heat capacity and density based on the ultimate and proximate analyses. Finally, the stream thermodynamic condition (1 atm and 25°C) and mass flow rate of 1.5 kg/hr of each feed were used as input for the model indicated in table 3.2.

Table 3-2: Inlet stream specification of model inputs

Stream	Component	Temperature	pressure	Mass Flow (kg/hr)
Biomass	Specified as its ultimate and proximate analysis	25°C	1atm	1.5 kg/hr
Primary air	21% O_2 79% N_2 (mole fraction)	25°C	1atm	0.5kg/hr
Secondary air	21% O_2 79% N_2 (mole fraction)	25°C	1atm	0.05kg/hr

vi. Block specification

After specifying the inlet streams, all the blocks were specified according to the design operating condition. The pressure of all feed streams and unit operation blocks were set to 1 am (i.e. no pressure drop in the system). Table 3.3 gives a brief description of the unit operation blocks presented in the flow sheet.

Table 3-3: Description of Aspen Plus simulation blocks

Module Name	Scheme	Block ID and Operating parameters	Description
RStoic		Drier Temp =150°C	Stoichiometric rector can be used to simulate a reaction with the unknown or unimportant reaction kinetic and known stoichiometry by specifying the extent of reaction or the fractional component of the key component. In this simulation, DRIER was used for modeling the drying of BIOMASS.

Sep		Sep_1	Separates inlet stream components into multiple outlet streams, based on splitfractions. SEP1 separates the dried biomass from a portion of water.
RYield		Decompose Temp =500°C	Nonstoichiometric rector based on known yield distribution. In this simulation DECOMPOS was used for the conversion of the non-conventional stream BIOMASS into conventional components (C, H_2, O_2, N_2, S)
RGibbs		Gasify Temp =700°C	Rigorous reaction and/or multiphase equilibrium reactor is used when reaction Stoichiometric is unknown, but the reactor t emperature and pressure are known.
Sep		Sep_2	A separator model, Sep_2 was used to separate ash from the gas mixture using split fraction of the components
RGibbs		Combust Temp=900°C	Rigorous reaction and/or multiphase equilibrium reactor is used when reaction Stoichiometric is unknown, but the reactor t emperature and pressure are known.

4. RESULTS AND DISCUSSION

4.1. Biomass characterization

Proper understanding of the physical and thermo chemical properties of biomasses is necessary for the design of thermo chemical conversion systems. This study provides information on moisture content, proximate analysis, and ultimate analysis of three biomass varieties (Rice husk, coffee husk and saw dust) for preliminary simulation model inputs and how they are represented empirically from their elemental composition to establish stoichiometric relation between the feed and gasifying agent.

4.1.1. Proximate Analysis

The result of the proximate analysis of the biomass is presented in table 4.1 below. During heating process the biomass decomposes into volatile gases and solid char. Volatile matter refers to the part of the biomass that is released when the biomass is heated up to 400 to 500°C. Biomass typically has high volatile matter content up to 80 percent. As it can be seen from the table 4.1, Volatile content of 54.6%, 58.9 and 72.8 wt. %was recorded for the rice husk, coffee husk and saw dust respectively. However, saw dust has high volatile mater content and this is high volatile content signifies easy ignition of the feed. The high volatile matter content indicates that during combustion and gasification; most of the feed will volatilize and burn as gas in combustion chambers.

Ash, which is the inorganic matter left out after complete combustion of the biomass was found to be 19.8%, 3.7 and 2.61wt% was obtained for the rice husk, coffee husk and saw dust respectively. This is the percentage of impurity that will not burn during and after combustion. The low ash content indicates that it is suitable for thermal utilization. Higher ash content in a fuel usually leads to lower calorific value. The fixed carbon of a fuel is the percentage of carbon available for char combustion. In each samples, it was found to be 13.26, 28.28 and 17.6 % for rice husk, coffee husk and saw dust correspondingly. The low fixed content makes it tend to prolong cooking time by its low heat release.

Table 4-1: Proximate properties of biomass samples

Properties	Rice Husk	Coffee Husk	Saw Dust
Moisture content	8.76	9.65	6.84
Volatile matter	63.23	58.37	72.94
Fixed Carbon	13.59	28.28	17.61
Ash	14.42	3.70	2.61

4.1.2. Ultimate Analysis

Table 4.2 below, presents the ultimate analysis of the rice husk, coffee husk and saw dust. Ultimate Analysis involves the estimation of important chemical elements that makes up the biomass, namely carbon, hydrogen, oxygen, nitrogen and Sulphur.

Elemental analysis of the biomasses from CHNS (EA 1112 Flash CHNS/O- analyzer) under carrier gas flow rate of 120 ml/min, reference flow rate 100 ml/min, oxygen flow rate 250 ml/min; furnace temperature of 900 ℃ and oven temperature of 75 ℃ conditions were done in duplicated run at Addis Ababa University and the average values were taken as represented in the table 4.2. The amount of carbon and hydrogen content in the biomass is very satisfactory as they contribute immensely to the combustibility of any substance in which they are found. The low sulphur and nitrogen contents in the biomasses are welcomed development as there will be minimal release of sulphur and nitrogen oxides into the atmosphere and that is an indication that the burning of biomasses examined in this work will not pollute the environment.

Table 4-2: Elemental properties of biomass samples

Properties	Rice Husk	Coffee Husk	Saw Dust
Carbon	35.36	43.39	44.29
Hydrogen	4.83	6.36	5.40
Oxygen	59.51	48.79	40.27
Nitrogen	0.21	1.41	0.57
Sulfur	0.09	0.05	-

As it clearly seen in the table 4.3, the general Rice husk, coffee husk and saw dust chemicals formula was $CH_{1.64}O_{1.27}N_{0.005,}$ $CH_{1.76}O_{0.84}N_{0.028}$ and $CH_{1.50}O_{0.7}N_{0.01.}$ The stoichiometric reaction for complete oxidation of these biomasses were; 2.71 m^3/kg, 4.35m^3/kg and 4.83 m^3/kg. The stoichiometric relation estimates that 2.71 m^3 (3.14 kg) of air is required for complete oxidation of 1 kg of rice husk to CO_2 and H_2O. Similarly, 4.35 m^3 (5 kg) and 4.83 m^3 (6 kg) of air is required for complete oxidation of 1 kg of coffee husk and saw dust to CO_2 and H_2O. The detail computation of empirical formula of biomasses and stoichiometric reaction to oxidized is presented in the Appendix table A1.

Table 4-3: Empirical formulas of biomasses and stoichiometric air to fuel ratio of the feeds

Biomass	Empirical Formula	Stoichiometric (A/F ratio) [m^3/kg]
Rice Husk	$CH_{1.64}O_{1.27}N_{0.005}$	2.71
Coffee Husk	$CH_{1.76}O_{0.84}N_{0.028}$	4.35
Saw Dust	$CH_{1.50}O_{0.7}N_{0.01}$	4.83

As it seen from the table, coffee husk and saw dust required higher stechoimetric air as compared with rice husk. This is due to the fact that coffee husk and saw dust has lower inherent oxygen content as indicated in the elemental properties of biomass samples. Rice husk requires lower air lower stechoimetric air as it has higher inherent oxygen composition as obtained from the elemental analysis.

4.2. Sensitivity Analysis

Sensitivity analysis is advantageous to obtain best operating condition for a process which will be beneficial to improve the performance and efficiency of the system. Feed stock flow rate, gasifying agent flow rate, equivalence ratio, reactor pressure and reactor temperature are some of the important operating parameters which influence the gasification process greatly. Change in any of the parameter has considerable effect on the end-gas composition and hence on the performance of the gasifier. Also, different feedstock has inherent heterogeneity in terms of their composition and thermo-chemical properties. The following variables were chosen in sensitivity analysis to perform the model analysis.

4.2.1. Air/Fuel Ratio

Variation of air equivalence ratio will change the amount of air introduced into the reactor. As air to fuel ratio is increased, the amount of oxygen supplied to the gasifier increased due to which conversion of carbon present in the fuel increases. But excess amount of oxygen oxidizes the fuel completely and the production of syngas declines. Therefore, three different reaction conditions can be identified: complete combustion to CO_2, complete gasification to CO and partial combustion (gasification) to CO_2 and CO. This ratio has a strong effect on syngas production. Air to fuel ratio was varied from 0.0 to 1.0 in this simulation (i.e. mass flow rate of fuel is set to constant 1.5 kg/hr.) and only the mass flow rate of air is varied from 0 – 6 kg/hr.)While gasification temperature 700°C, pressure 1bar parameters remain unchanged. The simulation result is presented in figure 4.4

As it can be seen in figure 4.1, CO_2 mole fraction increased from 0.431 to 0.185 with the increase of air flow rate from 0 kg/h to 4 kg/hr (ER 0 to 1). While, CO, H_2 and CH_4 show an inverse significant decrease from 0.516 to 0.187, 0.33 to 0.233 and 0.879 to almost zero respectively for rice husk. This indicates the increase of air to fuel ratio means more oxygen is placed in carbon combustion toward CO_2production. This excess amount of oxygen oxidizes the fuel completely to CO_2 and declines syngas production because an excess air addition prevailed with complete combustion of carbon to CO_2 over partial oxidation of carbon to CO.

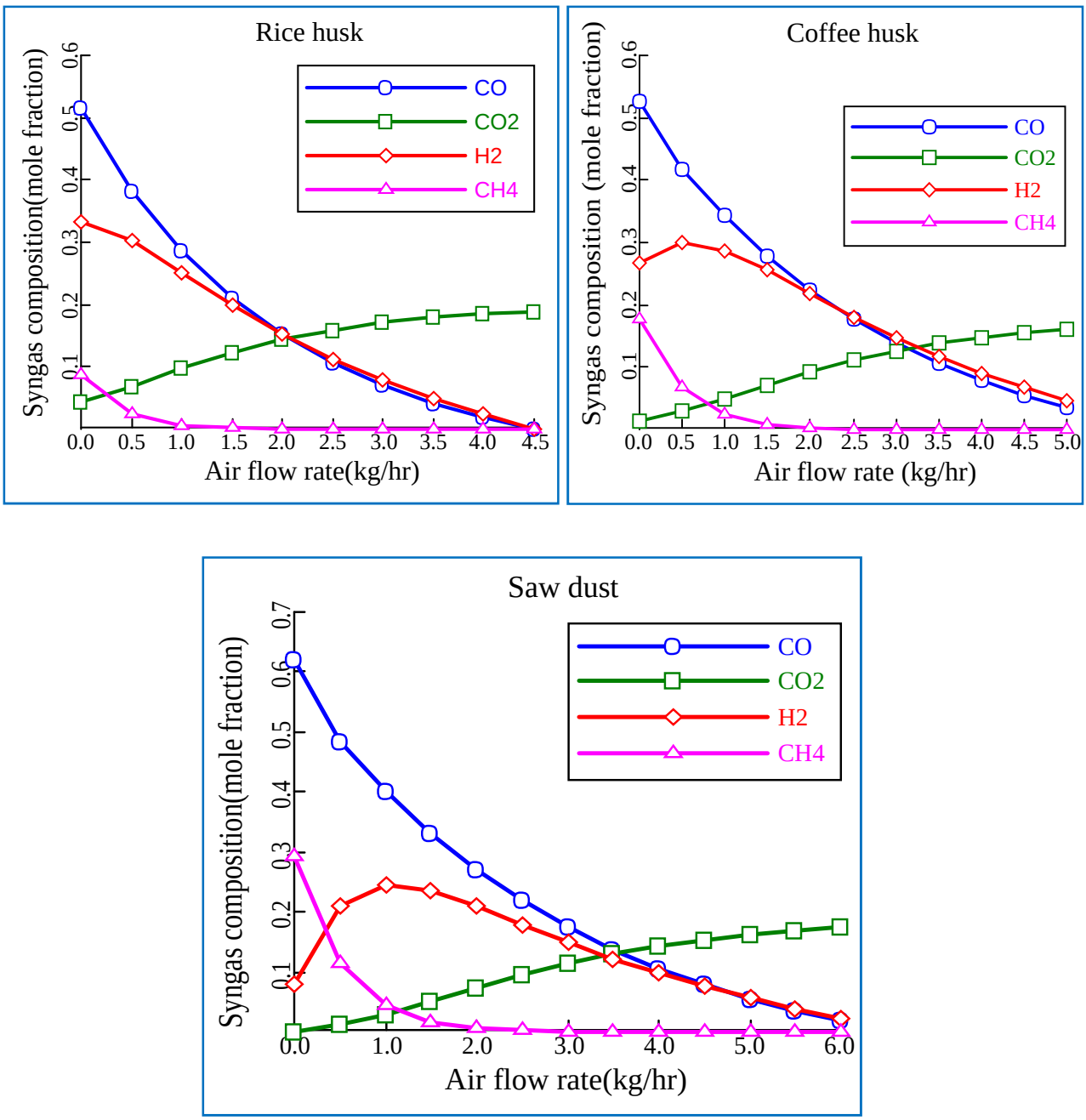

Figure 4-1: Effects of air/ fuel ratio on syngas composition of rice husk, coffee husk and saw dust

Within the same parametric study, coffee husk CO_2 mole fraction increased from 0.144 to 0.162 with an increase of air flow rate from 0 kg/h to 5 kg/hr (ER 0 to 1). Whereas, CO, H_2 and CH_4 show an inverse significant decrease from 0.5259 to 0.36, 0.268 to 0.458 and 0.176 to almost zero respectively.

Similarly, saw dust CO_2concentration significantly increased from 0.001 to 0.174 with the increase of air flow rate from 0 kg/h to 6 kg/hr (ER 0 to 1). At the same time, the concentration of CO, H_2 and CH_4 showed an inverse significant decrease from 0.62 to 0.018, 0.08 to 0.02 and 0.29 to almost zero percent respectively.

As it can be seen in figure 4.1, the contents of combustible components of producer gas (CO, $H_2$and CH_4) reaches maximum values of 0.516,0.526 and 0.62for rice husk, coffee husk and saw dust air gasification at air flow rate of 0 kg/hr (ER of 0.0). When air flow rate is above 0.0, the Pyrolysis and gasification reaction has gone to completion with continued combustion of the combustible components (CO, H_2 and CH_4) to generate CO_2 and H_2O as presented in Appendix table I in accordance with the following reaction schemes.

$$C + O_2 \rightarrow CO_2 \quad \text{Carbon combustion,}$$

$$H_2 + \frac{1}{2}O_2 \rightarrow H_2O \quad \text{Hydrogen combustion}$$

$$CH_4 + 2O_2 \leftrightarrow CO_2 + 2H_2O \quad \text{Methane combustion}$$

On the other hand, an increase in amount of O_2, promotes the complete reaction of fixed carbon and O_2, which leads to a decrease in CO concentration and an increase in CO_2 concentration. The increase in ER causes the intense combustion of combustible components releasing heat. When the ER reaches very high level, excessive complete combustion generates CO_2 and H_2O at the expense of CO and H_2 [122]

4.2.2. Gasification Temperature

Gasification can occur in any of the zones but is prevalent in the gasification zone due to the lack of oxidizer. This process can be carried out at a range of temperatures from 400 to 1500K. Therefore, the influence of temperature on syngas composition of rice husk, coffee husk and saw dust has been shown in figure 4.2. All gas components are plotted on a mole fraction basis. The gas components H_2S, $NH_3$and HCl are omitted due to their very low content and the fact that they would be removed by syngas separator mode (cleaning equipment). The gasification temperature was varied from 500°C to 1500°C and its effect on syngas production at air flow rate of 0.5 kg/hr and 1 bar operating conditions were presented.

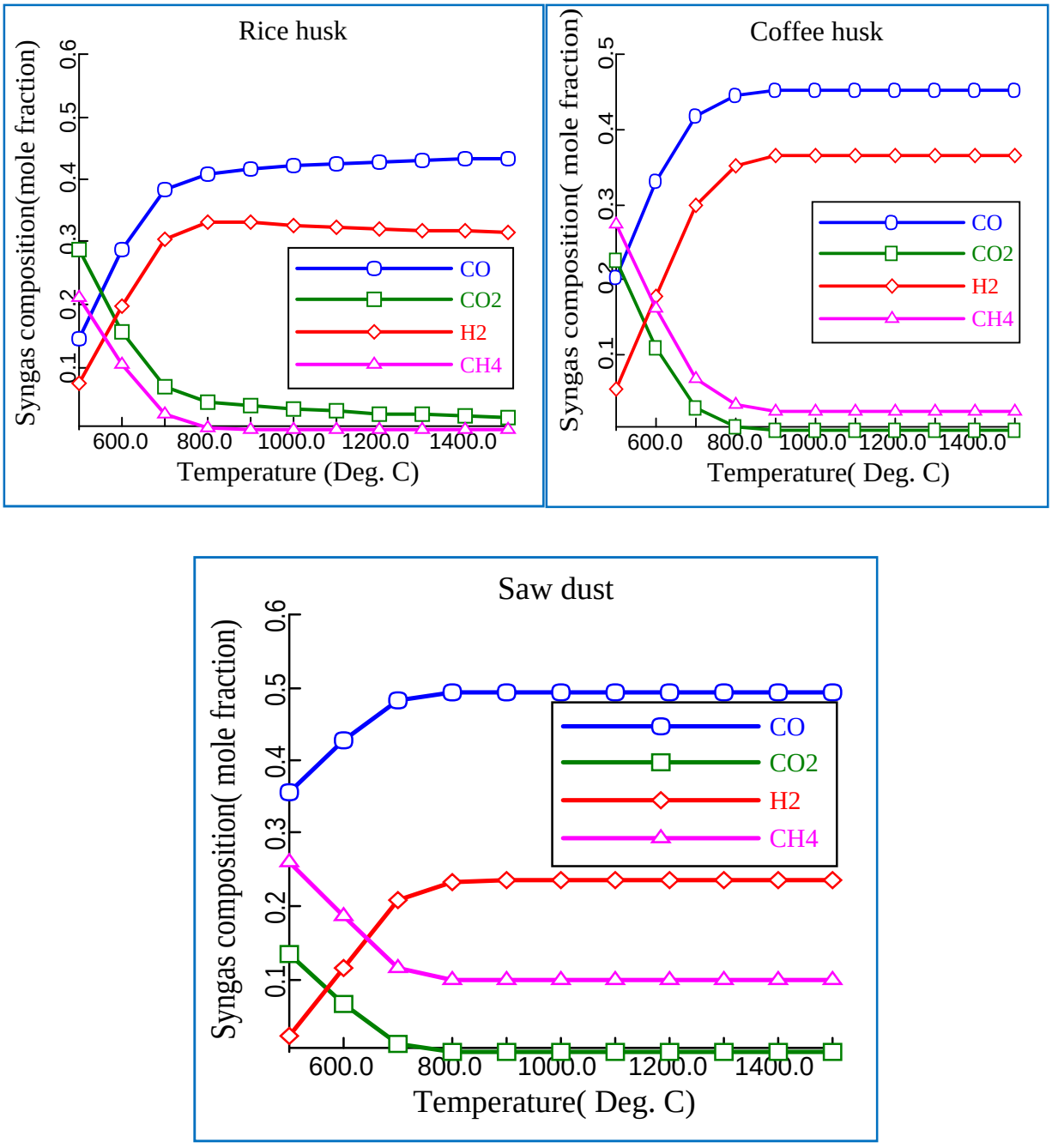

Figure 4-2: Effects of gasification temperature on syngas composition of rice husk, coffee husk and saw dust

It can be seen from Figures 4.2 that, for each feedstock, CO and H_2concentration increases with increase in gasification temperature while CO_2 and CH_4 concentration follow an opposite trend. Based on the developed model, it is observed that a gasifier temperature of 900°C provides an ideal condition for rice, coffee husks and saw dust.

$$C + CO_2 \leftrightarrow CO \quad \text{(Bourdourd reaction)}, \qquad \Delta H = +172.4\ \text{kJ/mol}$$

$$C + 2H_2 \leftrightarrow CH_4 \text{(Hydrogasificatio reaction)}, \qquad \Delta H = -74.6\ \text{kJ/mol}$$

$$C + H_2O \leftrightarrow CO + H_2 \text{(Water} - \text{gas reaction)}, \qquad \Delta H = +131.3\ \text{kJ/mol}$$

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 \text{(Steam reforming)}, \qquad \Delta H = +206\ \text{kJ/mol}$$

As it can be seen from the figure 4.2, at low temperature (500°C) the carbon present in the biomass is not utilized completely. At this temperature both un-burnt carbon and methane are present in the syngas but is converted into carbon monoxide in accordance with Boudouard reaction and CH_4 is converted into H_2 by the reverse hydrogasification reaction.

The equilibrium chemical reaction will shift in the opposite direction at those elevated temperature to offset the change [123], Therefore, the equilibrium shifts the chemical reaction when the temperature is increased which causes increase in H_2 concentration and a decrease in CH_4 concentration. This result is also supported by [124], which state that the higher temperature contributes more favorable conditions for thermal cracking and steam reforming.

The water gas reaction implies that a high temperature increases the production of both CO and H_2. The increase of H_2 concentration could be explained by the endothermic reactions in water gas, and steam reforming, and CO concentration would increase because endothermic reactions in Bourdouard, water gas and steam reforming are more dominant than the exothermic reaction in carbon combustion because the equilibrium chemical reaction will shift in the opposite direction at those elevated temperature to offset the change. Higher gasification temperature favors endothermic products in endothermic reactions and favored the reactants in exothermic reaction reactions towards the positive direction.

A possible reason to explain the result is that reforming reaction of CH_4, Boudouard reaction, water gas reaction and reverse reaction of hydro gasification were accelerated when the gasification temperature increases. Therefore, CH_4 and CO_2 gradually converted to CO and H_2 which leads to faster increase of H_2 and CO contents, while faster decline in CH_4 and CO_2 mole fraction.

Rise in temperature leads to an increasing in concentration of CO and decreasing concentration of CO_2 in the Bourdourd reaction until the temperature reaches 1000°C, at which point the reaction only generates CO [122];[125]. At this temperature the simulated result also reveals that concentration of CO_2 almost zero and the Bourdourd reaction generates only CO. In the heating process, the chemical equilibrium constant of hydrogasification reaction becomes reduced, consequently, CH_4 concentration almost reduced to zero [122]. The simulated result shows that the concentration of CH_4 becomes zero when the gasification temperature reaches to 1000°C and this result was in agreement with the result reported by [122].

4.2.3. Gasification pressure

Pressure is very important factor in the gasification process. Figure 4.3 shows the effects of the change in gasification pressure from 1 to 10 bars on the result of gasification when 1.5 kg/hreach feeds entered at 700°C with 0.5 kg/hr air.

As it can be seen in figure 4.3, for each feedstock, H_2 and CO mole fraction were declined, while the contents of CO_2 and CH_4 were raised at higher gasification pressure. The constant reduction in H_2 concentration and the constant rise in CH_4 concentration content may be from the following reactions:

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 (\text{Steam reforming}), \Delta H = +206\ \text{kJ/mol}$$

$$C + 2H_2 \leftrightarrow CH_4 (\text{hydrogasificati}), \Delta H = -74.6\text{kJ/mol}$$

$$C + CO_2 \leftrightarrow 2CO \quad (\text{Bourdourd reaction}), \Delta H = +172.4\ \text{kJ/mol}$$

Le Chatelier's principle, states that if a system in equilibrium is disturbed by changes in determining factors, such as temperature, pressure and concentration of components, the system will tend to shift its equilibrium position so as to counteract the effect of the disturbance. Pressure is caused by gas molecules hitting the sides of their container. The more molecules you have in the container; the higher pressure will be. The system can reduce the pressure by reacting in such a way as to produce fewer molecules. In steam reforming reaction of methane, there are 4 moles (1mol CO + 3mole CH_4) on the right side of equation and fewer molecules, 2 mole (CH_4 +

H_2O) on the left side. Similarly, fewer molecules are on the right side of hydrogasification reaction.

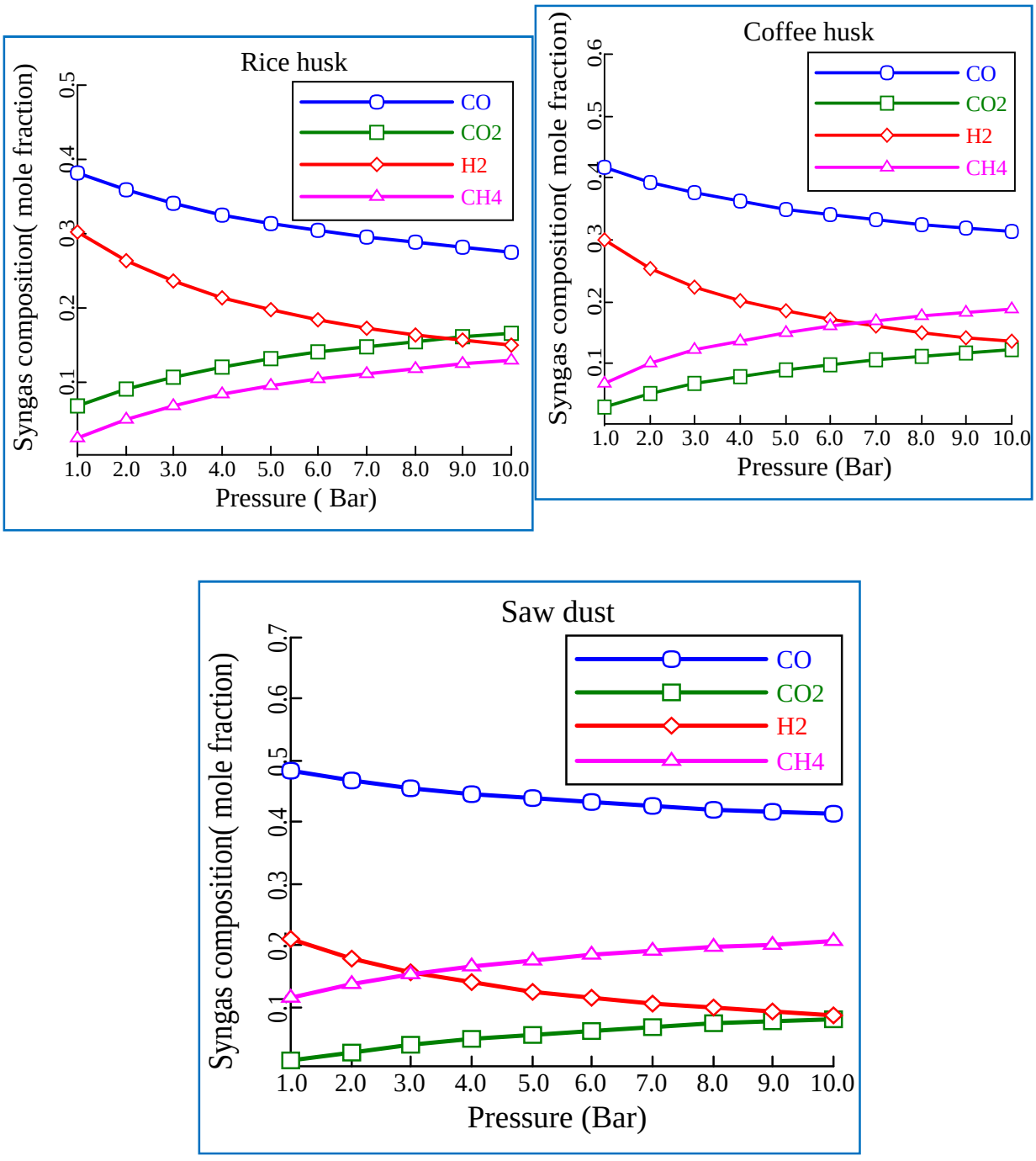

Figure 4-3: Effects of gasification pressure on syngas composition of rice husk, coffee husk and saw dust

According to this principle, pressurization causes the balance of steam reforming reaction of CH_4 to move in direction to reduce volume which gives rise to a slowdown of positive reaction and accelerates the reverse reaction.

Similarly, Increase the gasification pressure makes the balance of hydro gasification reaction move in the right direction of reaction, and H_2 is constantly consumed to form CH_4 in all cases of the feed stocks. Consequently, some H_2 and CO are consumed to generate CH_4. The Bourdourd reaction is kept to generate CO_2 during pressurization, the positive direction of Bourdourd reaction generates CO_2 which results increasing CO_2 concentration and increases CO consumption. The results are close to the gasification experiment conducted by [126, 47] on a fluidized bed reactor using oak and momiki biomass.

4.2.4. Effects of steam injection

The model Performance was also examined with respect to the addition of the steam as a gasifying agent instead of air. The steam-to-biomass ratio (SBR) was increased from 0 to 1.0 by altering the steam flow rate to the gasifier at fixed fuel flow rate of 1.5 kg/hrat a temperature of 700°C and pressure of 1 bar and the simulation result is presented in figure 4.4

As can be seen in figure 4.4, CO_2 mole fraction increased from 0.0451 to 0.1543 with the increase of steam flow rate from 0 kg/h to 1.5 kg/hr (SBR of 0.0 to 1.0) while CO and CH_4 show an inverse decrease from 0.516 to 0.085 and 0.088 to 0.0007 respectively. The H_2 concentration increases from 0.33 and reaches maximum 0.446 at steam flow rate of 0.4, beyond this flow rate concentration slightly decreases from 0.446 to 0.357 was observed for rice husk.

As is clearly seen in the figure 4.4, coffee husk CO_2 mole fraction increased from 0.014 to 0.14 with an increase of steam flow rate from 0 kg/h to 1.5 kg/hr (SBR 0.0 to 1.0), Whereas, CO and CH_4 show an inverse decrease from 0.526 to 0.139, and 0.177 to 0.0034 respectively. The H_2 concentration increases from 0.268 and reaches maximum 0.48 at steam flow rate of 0.6, further than this flow rate concentration slightly decreases from 0.48 to 0.4396 was seen for coffee husk.

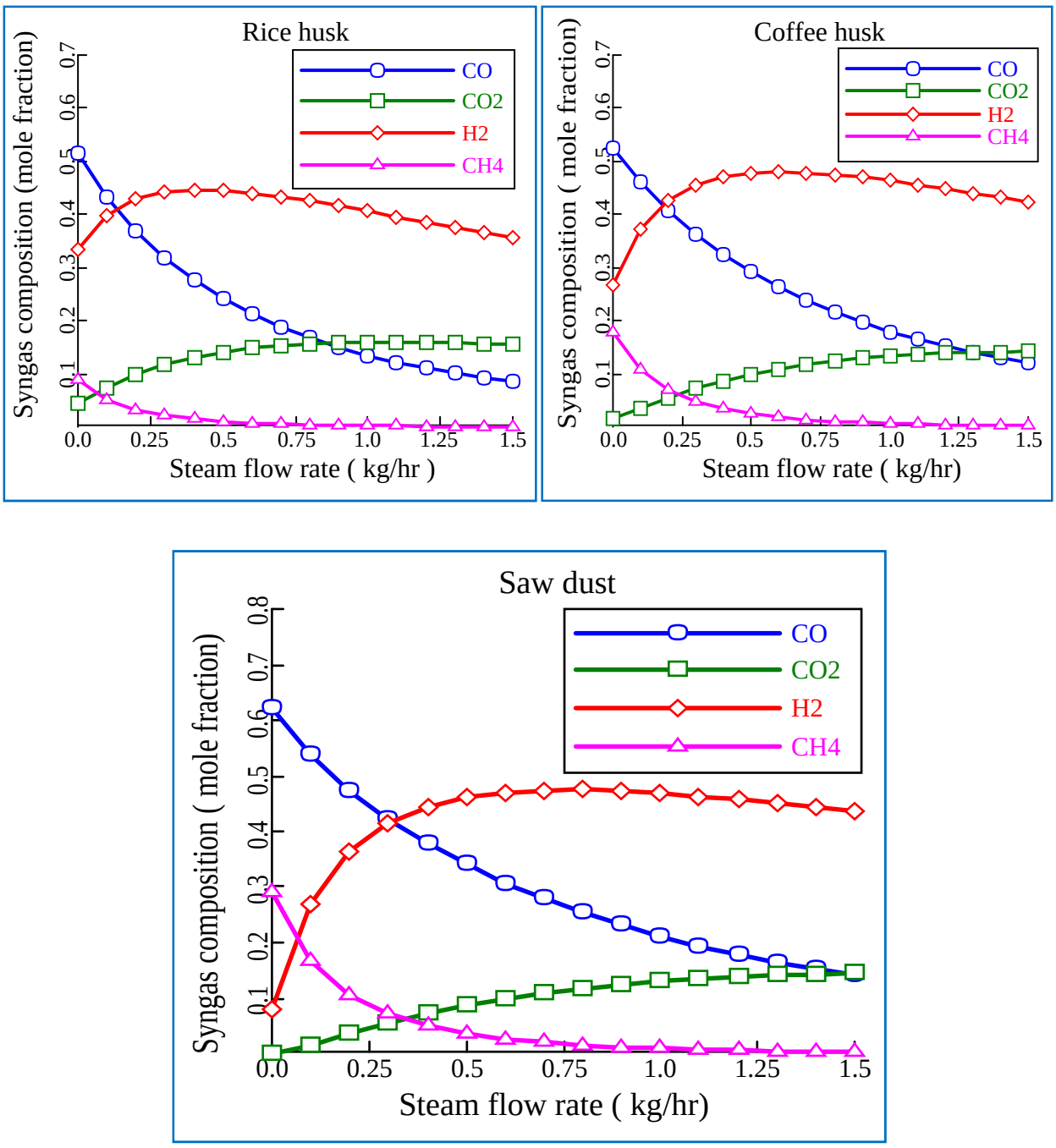

Figure 4-4: Effects of steam to fuel ratio on syngas composition of rice husk, coffee husk and, saw dust

In figure 4.4, similarly, saw dust CO_2mole fraction increased from 0.001 to 0.144 with the increase of air flow rate from 0 kg/h to 1.5 kg/hr (SBR 0.0 to 1.0). Simultaneously, the concentration of CO and CH_4 decrease from 0.62 to 0.14 and 0.292 to 0.0034 respectively. The H_2mole fraction increases from 0.08 and reaches maximum 0.475 at steam flow rate of 0.8 kg/s,

away from this flow rate concentration slightly decreases from 0.475 to 0.43 for saw dust. Similar results were reported by [126].

In all feed stocks, the H_2 and CO_2 production increase within steam flow rate and decline in CH_4 and CO production in accordance with the reaction;

$$CH_4 + 2H_2O \leftrightarrow CO + 4H_2 (Steam\ reforming), \Delta H = +165\ \text{kJ/mol}$$

$$CO + H_2O \leftrightarrow CO_2 + H_2 (CO - \text{shift reaction}), \qquad \Delta H = -41.1\ \text{kJ/mol}$$

This is due to the fact that the addition of steam would have two major effects on the kinetics of gasification: (i) accelerating the rates of reforming reactions and (ii) displacing the equilibrium of the CO shift reaction. The increase of the steam will enhance CH_4 reforming reactions to produce a significant amount of H_2 and CO. On the other hand, the increase of steam will shift the equilibrium of the CO shift reaction to the right side, to increase the CO_2 and H_2 concentrations results a decrease in the CO concentration.

4.3. LHV of syngas

In this paper, the lower heating value of the syngas was reported as an indicator of the gasifier efficiency. The LHV can be calculated from the following equation [127]:

$$LV_g = Y_{CO} \times LVH_{CO} + Y_{H2} \times LVH_{H2} + Y_{CH4} \times LVH_{CH4}$$

Where *Y* is the mole fraction of each gas species, the lower heating values of the gas species are LHV_{CO} = 10.11 MJ/kg, LHV_{H2}=119.5 MJ/kg andLHV_{CH4} = 49.915 MJ/kg [128].

From the above equation, we can see the LHV is dependent on the concentration of combustible gases and the LHV value was calculated based on the individual heating value of components including H_2, CO, and CH_4.

The LHV of syngas of rice husk, coffee husk and saw dust was declined from 49.41 to 2.98 MJ/kg, 49.16 to 0.98 MJ/kg, and 30.46 to 2.69 MJ/kg with raise in air flow rate from 0kg/hr to air required for complete combustion of feeds. This is due to the increase in air flow favors the

combustion reactions of H_2, CO and CH_4 and CO, H_2 and CH_4 oxidation reactions are dominant as seen in figure 4.1a-c, resulting in a significant decrease in the calorific value of the product gas as indicated in figure 4.5a.

Although elevated temperature reduces the methane concentration, the H_2 and CO content of product gas was in higher percentage, as seen in figure 4.2. The LHV of syngas for rice husk, coffee husk and saw dust was in the range of 20.87 – 42 MJ/kg, 22.16 – 49.57 MJ/kg and 19.37 – 37.87MJ/kg respectively at gasification temperature ranges of 500 – 1500^0C. The calculated LHV of the product gas of feeds enhanced within the temperature as indicated in figure 4.5b. it is clear to see from the figure 4.5b syngas generated from coffee husk has registered maximum value in all ranges of temperature because coffee husk generates H_2 having higher calorific value among the feeds at all ranges of temperatures.

Pressurized gasification system favors the production and CH_4 and CO_2, and decline in H_2 and CO production in accordance with Le Chatelier's principle as discussed in figure 4.3, as a result the LHV of rice husk, coffee husk and saw dust syngas was declined from 41.36 to 27.05 MJ/kg, 43.5 to 28.76 MJ/kg, and 35.64 to 24.73 MJ/kg with pressure.

An increase in the steam flow rate also results in production of CO_2 and H_2. In the CO- shift reaction and steam reforming reaction, CO and CH_4 are consumed towards the formation of CO_2 and H_2. Consequently, CO_2 and $H_2$the dominate syngas product composition based on chemical equilibrium as seen in figure 4.5. It was observed that using steam as gasifying agent results increase in mole fraction of hydrogen which has significant heating value, as seen in figure 4.5, results anincrease in LHV of the syngas seen in figure 4.5c.It can be seen from the figure 4.5d,at higher steam flow rate the heating value of syngas slightly declined for each case of feeds. The maximum LHV of syngas was obtained 57.1 MJ/kg at steam flow rate of 0.3kg/hr, 61.31 MJ/kg at 0.5kg/hr and 60.68MJ/kg at 0.7kg/hr for rice husk, coffee husk and saw dust correspondingly.

It can be concluded from syngas composition obtained in this simulation result indicates that the concentration of combustible gas decreased with the increase of air flow rate, pressure, and steam to biomass ratio. As a result, LHV of the syngas was decreased as the ER ratio, steam to biomass ratio, and the LHV of syngas increased with temperature that trend can be found in

figure 4.5

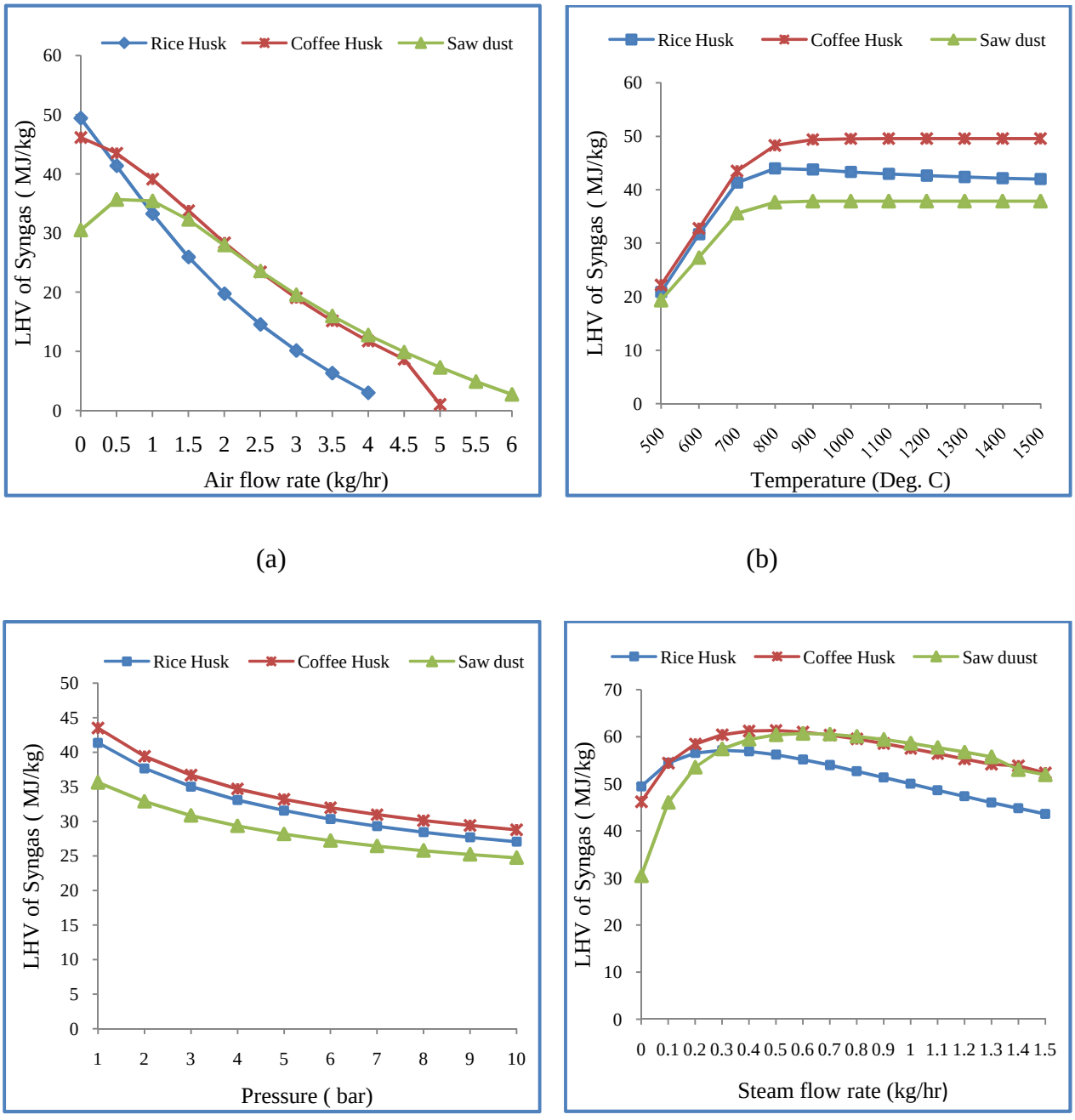

Figure 4-5: Effects of (a) air to fuel ratio, (b) temperature, (c) pressure and (d) steam to biomass ratio on LHV of syngases

5. CONCLUSION AND RECOMMENDATION

5.1. Conclusion

The comparison of biomass for the use in gasification process was done from different biomass based wastes. The thermo-chemical equilibrium model was used in the study. Produced syngas composition and the LHV of the syngas were main criteria for gasification process efficiency in comparison of biomass types. Gasification process was simulated with dry basis from rice husk, coffee husk and saw dust by-product. All these herbaceous agricultural and processing industry by-products are available in Ethiopia.

The ultimate analysis result shows that the lower oxygen content is found in coffee husk and saw dust along with the more air must be supplied in the simulation of gasification process to acquire the required equivalent ratio for the gasification process. In contrast, the higher is oxygen content is found in rice husk and lower amount of air is supplied for rice husk gasification.

The hydrogen amount in the fuel has effect on the H_2 content in the syngas. From the ultimate analysis result it can be concluded that coffee husk has higher hydrogen percentage and registered the higher heating value of produced syngas in all gasification operating parameters.

Simulation results show that the developed model can be successfully used in gasification process on the same basis for all biomass samples. The results obtained from these comprehensive simulations within this work were a promising outcome and can be used as pre-condition for fixed bed reactor design. Thus, this Simulation model provides the powerful theoretical basis for next steps on fixed bed gasifier modeling, optimizing the operation and scale-up/down designs. Gasification efficiency was about 80% for major samples including wood, but produced syngas calorific value was around 5.5 MJ Nm^{-3}. The consumption of wood fuels for energy production is continuously increasing and non-wood material can be used to satisfy the demand for biomass. In such a way the greater amount of the fossil fuel can substitute with biomass. It is important that one non-wood biomass type can be substituted by another without altering the produced syngas quality.

The dependence of syngas composition on different parameters of a fixed bed biomass gasifier performed thermodynamic equilibrium model by assuming all chemical reaction reaches in equilibrium using Aspen plus 2006. The temperature and air equivalence ratio (ER), Gasification

pressure and moisture content of the biomass are analyzed in order to evaluate their effects on the composition of the produced gas and on the LHV of gasification gas. The simulation results for the product gas composition versus temperature and air equivalence ratio were compared with literature review and the results obtained from the sensitivity analyses were in good agreement with literature review. Therefore, the developed model was capable of predicting accurately gasifier performance over a wide range of operating conditions.

Based on the result obtained, the following specific conclusion can be drawn:

1-The syngas composition was strongly influenced by the air to fuel ratio. Higher air to fuel ratio results complete combustion of carbon present in the feed resulting higher percentage of CO_2 in the product gas as it results lower calorific value of product gas. Lower air to fuel ratio results higher percentage of combustible gases (CO, H_2 and CH_4) in the product gas with a high heating value was obtained.

2-High temperature contributes more favorable conditions for endothermic and steam reforming reaction as a result carbon and steam were converted to CO and H_2. Higher temperature generates only CO and H_2 which has significant heating value thus syngas calorific value was enhance.

3- Pressurized conditions were not advantageous for formation of H_2rich producer gas and decline the calorific value of syngas as increasing gasification pressure, increase the contents of CO_2 and CH_4 whereas the reduction in content of H_2 and CO.

4- Higher steam to biomass ratio increase the CO_2 and H_2 concentrations by displacing the equilibrium of the CO shift reaction and CH_4 reforming reactions to produce a significant amount of H_2 and CO results however in a raise in lower heating value of syngas.

5.2. Recommendation

Prediction of gasifier performance is generally carried out using two techniques, namely, kinetic modeling and thermodynamic equilibrium modeling. Thermodynamic equilibrium modeling was adopted in the current study because it is not governed by any of the gasifier design parameters and the only input needed is the elemental composition of the feed, which is known from its ultimate analysis. Therefore, Thermodynamic equilibrium modeling was deployed to compute and ascertain the influence of the equivalence ratio, moisture content and reaction temperature on biomass gasification. The gasification process is assumed to be steady state. In real life, gasification process is involved with unsteady state processes. For future work we recommend,

- Further study is needed on combined thermodynamic equilibrium modeling and reaction kinetics modeling to complete the full design on these processes.
- Model validation of the simulation result with experimental data from gasification of biomasses in a pilot-scale fixed bed gasifier

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APENDICES

APPENDIX A: EMPIRICAL FORMULA OF BIOMASSES

The empirical formula of samples can be computed by considering the equation of biomass as $C_aH_bO_cN_d$, where a, b, c and d are subscripts.

Assume, $a = 1$, normalized with respect to carbonand using the formula for Rice Husk;

$$b = \frac{mass\ fraction\ of\ Hydrogen \times Molecular\ weight\ of\ Carbon}{mass\ fraction\ of\ Carbo \times Molecular\ weight\ of\ Hydrogen} = \frac{0.0483 \times 12}{0.3536 \times 1}$$

$$= 1.64$$

$$c = \frac{mass\ fraction\ of\ Oxygen \times Molecular\ weight\ of\ Carbon}{mass\ fraction\ of\ Carbon \times Molecular\ weight\ of\ Oxygen} = \frac{0.5951 \times 12}{0.3536 \times 16}$$

$$= 1.27$$

$$d = \frac{mass\ fraction\ of\ Nitrogen \times Molecular\ weight\ of\ Carbon}{mass\ fraction\ of\ Carbon \times Molecular\ weight\ of\ Nitrogen} = \frac{0.0021 \times 12}{0.3536 \times 14}$$

$$= 0.005$$

Similarly, the empirical formula of coffee husk and saw dust was computed based on their elemental composition from table 4.2 and subscripts a, b, c, and d were calculated from the above formula.

Table A1: Empirical Formula of biomass samples

Biomass	*C*	*H*	*O*	*N*	*Empirical Formula*
Rice Husk	1	1.64	1.27	0.005	$CH_{1.64}O_{1.27}N_{0.005}$
Coffee Husk	1	1.76	0.84	0.028	$CH_{1.76}O_{0.84}N_{0.028}$
Saw Dust	1	1.50	0.70	0.01	$CH_{1.50}O_{0.7}N_{0.01}$

APENDIX B: STOICHIOMETRIC AIR -TO – FUEL OF BIOMASSES

The general Rice husk, coffee husk and saw dust chemicals formula used was $CH_{1.64}O_{1.27}N_{0.005,}CH_{1.76}O_{0.84}N_{0.028}$ and $CH_{1.50}O_{0.7}N_{0.01.}$ The stoichiometric relation for complete oxidation of a hydrocarbon fuel, C_nH_m, is given by:

$$C_nH_m + (n + \frac{m}{4})(O_2 + 3.78N_2) \rightarrow nCO_2 + \frac{m}{2}H_2O + 3.78(n + \frac{m}{4})N_2$$

For Rice husk($CH_{1.64}O_{1.27}N_{0.005}$), stoichiometric relation for complete oxidation,

$CH_{1.64}O_{1.27}N_{0.005} + \alpha (O_2 + 3.78N_2) \rightarrow xCO_2 + y H_2O + z N_2$

$C: \quad 1 = x \quad \rightarrow \quad x = 1$

$H: \quad 1.64 = 2y \quad \rightarrow \quad y = 0.82$

$O: \quad 1.27 + 2\alpha = 2x + y \quad \rightarrow \quad \alpha = \frac{(2 * 1 + 0.82 - 1.27)}{2} = 0.775$

$N: \quad 0.005 + 3.78 * 2\alpha = 2z \quad \rightarrow \quad z = 2.932$

Therefore, stoichiometric relation for complete oxidation of rice husk, becomes

$CH_{1.64}O_{1.27}N_{0.005} + 0.275(O_2 + 3.78N_2) \rightarrow CO_2 + 0.82\ H_2O + 2.932N_2$

Mass of fuel (RH) $= 1 \times 12 + 1.64 \times 1 + 1.27 \times 16 + 0.005 \times 14 = 34.03$ gram

Mass of Air $= 0.775(2 \times 16 + 3.78 \times 2 \times 14) = 106.83$ gram

$$\text{volume of air} = \frac{\text{mass of air}}{\text{Density of air}} = \frac{0.10683 \text{ kg}}{1.1588 \text{ kg/m}^3} = 0.092 \text{ m}^3_{\text{air}}$$

$$\text{Airtofuelratio} \left(\frac{A}{F}\right) \text{stoichio} = \frac{\text{Volune of air}}{\text{massofrice husk}} = \frac{0.092 \text{ m}^3}{0.03403 \text{ kg}} = 2.71 \text{ m}^3_{\text{air}}/\text{kg}_{\text{rice husk}}$$

For coffee husk($CH_{1.76}O_{0.84}N_{0.028}$), stoichiometric relation for complete oxidation,

$CH_{1.76}O_{0.84}N_{0.028} + \alpha (O_2 + 3.78N_2) \rightarrow xCO2 + yH2O + zN2$

$C: \quad 1 = x \quad \rightarrow \quad x = 1$

$H: \quad 1.76 = 2y \quad \rightarrow \quad y = 0.8$

$O: \quad 0.84 + 2\alpha = 2x + y \quad \rightarrow \quad \alpha = \frac{(2 * 1 + 0.88 - 0.84)}{2} = 1.02$

$N: \quad 0.028 + 3.78 * 2\alpha = 2z \quad \rightarrow \quad z = 3.8696$

$CH_{1.76}O_{0.84}N_{0.028} + 1.02 (O_2 + 3.78N_2) \rightarrow CO_2 + 0.88\ H_2O + 3.8696\ N_2$

Mass of fuel $=1 \times 12 + 1.76 \times 1 + 0.84 \times 16 + 0.028 \times 14 = 27.6$ gram

Mass of Air $=1.02(2 \times 16 + 3.78 \times 2 \times 14) = 140.6$ gram

$$\text{volume of air} = \frac{\text{mass of air}}{\text{Density of air}} = \frac{0.1406 \text{ kg}}{1.1588 \text{ kg/m}^3} = 0.12 \text{ m}^3{}_{\text{air}}$$

$$\text{Airtofuelratio} \left(\frac{\text{A}}{\text{F}}\right) \text{stoichio} = \frac{\text{Volune of air}}{\text{massofsaw dust}} = \frac{0.12 \text{ m}^3}{0.0276 \text{ kg}} = 4.35 \text{ m}^3{}_{\text{air}}/\text{kg}_{\text{coffee husk}}$$

For saw dust ($CH_{1.50}O_{0.7}N_{0.01}$), stoichiometric relation for complete oxidation,

$CH_{1.50}O_{0.7}N_{0.01} + \alpha\ O_2 + 3.78N_2) \rightarrow xCO_2 + \ y\ H_2O + z\ N_2$

C: $1 = x \quad \rightarrow \quad x = 1$

H: $1.5 = 2y \quad \rightarrow \quad y = 0.75$

O: $0.7 + 2\alpha = 2x + y \quad \rightarrow \quad \alpha = \frac{(2 * 1 + 0.75 - 0.70)}{2} = 1.025$

N: $0.01 + 3.78 * 2\alpha = 2z \quad \rightarrow \quad z = 3.8795$

$CH_{1.5}O_{0.7}\ N_{0.01} + 1.025(O_2 + 3.78N_2) \rightarrow CO_2 + 0.75\ H_2O + 3.8795\ N_2$

Gram of fuel $=1 \times 12 + 1.5 \times 1 + 0.7 \times 16 + 0.01 \times 14 = 24.84$ gram

Gram of Air $=1.025(2 \times 16 + 3.78 \times 2 \times 14) = 141.28$ gram

$$\text{Airtofuelratio} \left(\frac{\text{A}}{\text{F}}\right) \text{stoichio} = \frac{\text{Volune of air}}{\text{massofsaw dust}} = \frac{0.12 \text{ m}^3}{0.02484 \text{ kg}} = 4.83 \text{ m}^3{}_{\text{air}}/\text{kg}_{\text{saw dust}}$$

APENDIX C: SYNGAS COMPOSION AND LHV OFRICE HUSK

Table C1: Syngas composition at different air flow rate

Rice husk = 1.5 kg, Temperature = 7000C and pressure = 1 bar.

Air (kg/hr)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
0	0.51628868	0.04314634	0.33310324	0.08799569	49.41782
0.5	0.3824121	0.06902362	0.30335999	0.02502922	41.36704
1	0.286275	0.09807432	0.2511726	0.00676748	33.24716
1.5	0.21124641	0.12384445	0.19834117	0.00181972	25.9283
2	0.15269245	0.14415523	0.15202507	0.00047986	19.73467
2.5	0.10686534	0.15945576	0.11256096	0.00011649	14.53726
3	0.07070171	0.17073445	0.07874074	2.3304e-05	10.12548
3.5	0.04190177	0.1788837	0.04933589	3.0179e-06	6.319416
4	0.01878121	0.18460232	0.02337516	1.3259e-09	2.98321

Table C2: Syngas composition at different gasification temperature

Rice husk = 1.5 kg, air flow rate = 0.5 kg/hrandpressure = 1 bar.

Temperature (^0C)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV(MJ/kg)*
500	0.14635158	0.28787991	0.07421470	0.21088573	20.87463
600	0.28866438	0.15536205	0.19656114	0.10514594	31.65581
700	0.3824121	0.06902362	0.30335999	0.02502922	41.36704
800	0.40930873	0.04418438	0.33237853	0.00269682	43.99196
900	0.41665503	0.03705634	0.33109892	0.00027689	43.79252
1000	0.42123712	0.03249328	0.32711691	3.9128e-05	43.35113
1100	0.42483148	0.02890020	0.32360467	7.4109e-06	42.96617
1200	0.42772926	0.02600196	0.32072261	1.7774e-06	42.65078
1300	0.43008462	0.02364608	0.31837144	5.1578e-07	42.39357
1400	0.43201387	0.02171639	0.31644375	1.749e-07	42.1827
1500	0.43360734	0.02012244	0.31485114	6.7468e-08	42.00848

Table C3: Syngas composition at different gasification pressure

Rice husk = 1.5 kg, air flow rate = 0.5 kg/hr and Temperature = 700^0C

Pressure (bar)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV(MJ/kg)*
1	0.3824121	0.06902362	0.30335999	0.02502922	41.36704
2	0.35828693	0.09080675	0.26370283	0.05050074	37.65551
3	0.33997466	0.10741336	0.23556498	0.06905067	35.03382
4	0.32552383	0.12056192	0.21454772	0.08321481	33.08317
5	0.31368122	0.13136593	0.1980996	0.09451261	31.56182
6	0.30369632	0.14049505	0.18477152	0.1038225	30.33287
7	0.29509475	0.14837395	0.17368394	0.1116852	29.31341
8	0.28756	0.15528665	0.16426976	0.11845358	28.45008
9	0.28087124	0.16143177	0.15614498	0.12436918	27.70682
10	0.27486862	0.16695336	0.14903902	0.12960392	27.05826

Table C4: Syngas composition at different steam flow rate

Rice husk = 1.5 kg, Temperature = 700^0C and pressure =1bar

Steam flow rate (kg/hr)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
0	0.51628868	0.04314634	0.33310324	0.08799569	49.41782
0.1	0.4336024	0.07379931	0.39725676	0.05161095	54.43206
0.2	0.36919857	0.09814136	0.42796714	0.03265615	56.5047
0.3	0.31761437	0.11676854	0.44196701	0.02166404	57.1075
0.4	0.27558101	0.13073024	0.44648522	0.01486730	56.88321
0.5	0.24091	0.14098366	0.44525956	0.01047788	56.16712
0.6	0.21203525	0.14831973	0.44044672	0.00754955	55.1539
0.7	0.18778931	0.15337341	0.43337945	0.00554442	53.96414
0.8	0.16727927	0.15664843	0.42491783	0.00414102	52.67557
0.9	0.14981062	0.15854189	0.41563227	0.00313986	51.33937
1	0.13483677	0.15936569	0.40590725	0.00241344	49.98958
1.1	0.12192355	0.15936434	0.39600406	0.00187826	48.64889
1.2	0.11072352	0.15872932	0.38610052	0.00147844	47.33222
1.3	0.10095666	0.15761056	0.37631693	0.00117594	46.04924
1.4	0.09239589	0.15612553	0.36673353	0.00094435	44.80592
1.5	0.08485593	0.15436633	0.35740246	0.00076513	43.60568

APENDIX D: SYNGAS COMPOSION AND LHV OF COFFEE HUSK

Table D1: Syngas composition at different air flow rate

Coffee husk = 1.5 kg, Temperature = 7000C and pressure = 1 bar

Airflow rate (kg/hr)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
0	0.52597772	0.01444551	0.26806323	0.17665994	46.16864
0.5	0.41831356	0.02924877	0.29983733	0.06904424	43.50585
1	0.34297597	0.04875927	0.28770317	0.02563429	39.12747
1.5	0.27906083	0.07102454	0.25508174	0.00915828	33.76068
2	0.22388234	0.09233472	0.21697726	0.00328075	28.35598
2.5	0.17724091	0.11075048	0.18036949	0.00118463	23.40519
3	0.13824223	0.1259591	0.14722766	0.00042219	19.01241
3.5	0.10567813	0.13827014	0.11764857	0.00014351	15.13457
4	0.07840940	0.14812612	0.09121278	4.4331e-05	11.69486
4.5	0.05547963	0.15594379	0.06742839	1.1521e-05	8.619167
5	0.03611695	0.16207642	0.04585575	2.1698e-06	0.986797

Table D2: Syngas composition at different gasification temperature

Coffee husk = 1.5 kg, air flow rate = 0.5 kg/hrandpressure = 1 bar

Temperature (^0C)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
500	0.20306461	0.22561414	0.05395169	0.27376452	22.16517
600	0.3301644	0.10882814	0.17854686	0.16202082	32.76158
700	0.41831356	0.02924877	0.29983733	0.06904424	43.50605
800	0.44607726	0.00476448	0.35239761	0.03339020	48.28803
900	0.45096177	0.00056080	0.3642895	0.02594591	49.38691
1000	0.45153943	7.3346e-05	0.36599601	0.02493379	49.54616
1100	0.45161187	1.2507e-05	0.36625285	0.0247887	49.57034
1200	0.45162303	2.7142e-06	0.36630287	0.0247622	49.57511
1300	0.45162489	7.1917e-07	0.36631562	0.02475613	49.57635
1400	0.45162506	2.2471e-07	0.36631988	0.02475445	49.57677
1500	0.45162484	8.054e-08	0.3663218	0.0247539	49.57697

Table D3: Syngas composition at different gasification pressure

Coffee husk = 1.5 kg, air flow rate = 0.5 kg/hr and Temperature = 700^0C

Pressure(bar)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV(MJ/kg)*
1	0.41831356	0.02924877	0.29983733	0.06904424	43.50605
2	0.39332463	0.05137553	0.25455199	0.1002334	39.39862
3	0.3753294	0.06740759	0.22434212	0.12162773	36.67451
4	0.36138412	0.07988725	0.2022984	0.1376003	34.69657
5	0.35006342	0.09005358	0.18527299	0.15018081	33.17554
6	0.34057637	0.09859745	0.17160037	0.1604595	31.95881
7	0.33243969	0.10594265	0.16030303	0.16908469	30.95704
8	0.32533698	0.11236754	0.1507631	0.17647104	30.1139
9	0.31905004	0.1180646	0.14256764	0.18289865	29.39181
10	0.31342222	0.12317237	0.13542849	0.18856492	28.76462

Table D4: Syngas composition at different steam flow rate

Coffee husk = 1.5 kg, Temperature = 700^0C and pressure =1bar

Steam flow rate (kg/hr)	*CO*	*CO2*	*H2*	*CH4*	*LHV (MJ/kg)*
0	0.52597772	0.01444551	0.26806323	0.17665994	46.16917
0.1	0.46054107	0.03484867	0.37103757	0.1075595	54.36389
0.2	0.40728715	0.05456504	0.42487772	0.07044996	58.40707
0.3	0.36250296	0.07197693	0.45443205	0.04839925	60.38538
0.4	0.32423652	0.08680599	0.47035662	0.03439582	61.20251
0.5	0.2912395	0.09918054	0.47796131	0.02508091	61.31272
0.6	0.26261238	0.10935549	0.48020118	0.01866927	60.97093
0.7	0.23766045	0.11761348	0.4788603	0.01413752	60.33223
0.8	0.21582554	0.12422702	0.47508241	0.01086511	59.49668
0.9	0.19664897	0.1294439	0.46963381	0.00845932	58.53161
1	0.17974887	0.13348264	0.46304472	0.00666310	57.48369
1.1	0.16480492	0.13653245	0.45569063	0.00530366	56.38594
1.2	0.15154707	0.13875527	0.44784185	0.00426221	55.26199
1.3	0.13974691	0.14028854	0.43969516	0.00345552	54.1289
1.4	0.12921072	0.14124836	0.43139479	0.00282436	6.629066
1.5	0.1197738	0.14173237	0.42304674	0.00232592	6.393452

APENDIX E: SYNGAS COMPOSION AND LHV OF SAW GUST

Table E1: Syngas composition at different air flow rate

Saw dust = 1.5 kg, Temperature = 700^0C and pressure = 1 bar

Airflow rate (kg/hr)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
0	0.62146699	0.00109740	0.08038500	0.29193437	30.46094
0.5	0.48249082	0.01142654	0.20950667	0.11479361	35.64395
1	0.39956908	0.02785267	0.24409173	0.04384109	35.39693
1.5	0.33163784	0.04996736	0.23531886	0.01564646	32.25446
2	0.27120204	0.07381750	0.20844855	0.00555763	27.92886
2.5	0.21876365	0.09572236	0.17776893	0.00202823	23.55633
3	0.17422782	0.11444577	0.14846182	0.00075048	19.54009
3.5	0.13664521	0.13000652	0.12176012	0.00027334	15.94546
4	0.10490525	0.14278574	0.09766528	9.4378e-05	12.7363
4.5	0.07801032	0.1532088	0.07586450	2.9348e-05	9.855957
5	0.05513228	0.16165619	0.05600701	7.5717e-06	7.250603
5.5	0.03560078	0.16844718	0.03778058	1.1882e-06	4.874763
6	0.01887418	0.1738456	0.02092452	3.6944e-09	2.691298

Table E2: Syngas composition at different air flow rate

Saw dust = 1.5 kg, Temperature = 700^0C and pressure = 1 bar

Temperature (^0C)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
500	0.3570519	0.1354016	0.02322458	0.26131781	19.32036
600	0.42654903	0.06666921	0.11565091	0.18520351	27.30027
700	0.48249082	0.01142654	0.20950667	0.11479361	35.59631
800	0.49321335	0.00085574	0.23251517	0.09894192	37.66957
900	0.49400459	7.4148e-05	0.23454671	0.09761699	37.85476
1000	0.49406887	9.2387e-06	0.2347455	0.09749434	37.87309
1100	0.49407565	1.5642e-06	0.2347744	0.09747813	37.87581
1200	0.49407628	3.3901e-07	0.23478075	0.09747519	37.87643
1300	0.49407611	8.9801e-08	0.2347829	0.09747448	37.87665
1400	0.4940759	2.8056e-08	0.23478389	0.09747426	37.87676
1500	0.49407562	1.0054e-08	0.23478477	0.09747414	37.87685

Table E3: Syngas composition at different gasification pressure

Saw dust = 1.5 kg, air flow rate = 0.5 kg/hrandTemperature = 700^0C

Pressure(bar)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
1	0.48249082	0.01142654	0.20950667	0.11479361	35.64395
2	0.46741007	0.02630219	0.17875007	0.13633401	32.89126
3	0.45539558	0.03815953	0.15589172	0.15272244	30.85625
4	0.44584164	0.04759254	0.13871632	0.1652841	29.33422
5	0.43804171	0.05529633	0.12533269	0.17523963	28.15294
6	0.43151921	0.06174017	0.11457117	0.18336264	27.20646
7	0.42595765	0.06723587	0.10569855	0.19014639	26.42857
8	0.42114038	0.07199697	0.09823531	0.19591803	25.7761
9	0.41691394	0.07617478	0.09185471	0.20090323	25.21972
10	0.41316623	0.07987988	0.08632599	0.2052631	24.73877

Table E4: Syngas composition at different steam flow rate

Saw dust = 1.5 kg, Temperature = 700^0C and pressure =1bar

Steam flow rate (kg/hr)	*CO*	*CO_2*	*H_2*	*CH_4*	*LHV (MJ/kg)*
0	0.62146699	0.00109740	0.08038500	0.29193437	30.46094
0.1	0.53821302	0.01630690	0.2700993	0.16635864	46.02199
0.2	0.47438098	0.03623525	0.36349959	0.10534006	53.49224
0.3	0.42208896	0.05533277	0.41483329	0.07112765	57.39023
0.4	0.37792079	0.07227122	0.44443505	0.05010985	59.432
0.5	0.34000481	0.08682802	0.46142513	0.03639054	60.39419
0.6	0.30713574	0.09911993	0.47050088	0.02704597	60.68
0.7	0.278452	0.10936953	0.47434355	0.02047760	60.52134
0.8	0.25329547	0.11782316	0.47460539	0.01574654	60.06215
0.9	0.23114192	0.1247195	0.47236452	0.01227102	59.39691
1	0.21156264	0.13027742	0.46835521	0.00967542	58.59029
1.1	0.19420115	0.13469208	0.46309379	0.00770935	57.68789
1.2	0.17875785	0.1381345	0.45695199	0.00620148	56.72255
1.3	0.16497916	0.14075294	0.45020219	0.00503210	55.71828
1.4	0.1526494	0.14267496	0.44304644	0.00411606	54.69279
1.5	0.14158446	0.1440097	0.4356359	0.00339184	53.65921