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Wayne Doherty
Dublin Institute of Technology, wayne.doherty@dit.ie

Anthony Reynolds
Dublin Institute of Technology, anthony.reynolds@dit.ie

David Kennedy
Dublin Institute of Technology, david.kennedy@dit.ie

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Wayne G. Doherty* Anthony Reynolds† David Kennedy‡

*Dublin Institute of Technology, wayne.doherty@dit.ie

†Dublin Institute of Technology, anthony.reynolds@dit.ie

‡Dublin Institute of Technology, david.kennedy@dit.ie

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The effect of air preheating in a biomass CFB gasifier using ASPEN Plus simulation

Wayne Doherty [*], Anthony Reynolds, David Kennedy

Department of Mechanical Engineering,

Dublin Institute of Technology, Bolton Street, Dublin 1, Ireland

Abstract

In the context of climate change, increasing efficiency and energy security, biomass gasification is likely to play an important role in energy production. Atmospheric circulating fluidised bed (CFB) technology was selected for the current study. The primary objective of this research is to develop a computer simulation model of a CFB biomass gasifier that can accurately predict gasifier performance under various operating conditions. An original model was developed using ASPEN Plus (**A**dvanced **S**ystem for **P**rocess **EN**gineering Plus). The model is based on Gibbs free energy minimisation. The restricted equilibrium method was used to calibrate the model against experimental data. This was achieved by specifying the temperature approach for the gasification reactions. The model predicts syn-gas composition, process conversion efficiency and syn-gas heating values in good agreement with experimental data from a laboratory rig. Operating parameters were varied over a wide range. Parameters such as equivalence ratio (ER), temperature, air preheating, biomass moisture and steam injection were found to influence product gas (syn-gas) composition, syn-gas heating value, and process conversion efficiency. The results indicate an ER and temperature range over which hydrogen (H_2) and carbon monoxide (CO) production is maximised, which in turn ensures a high heating value and cold gas efficiency (CGE). Gas heating value was found to decrease with ER. Air preheating increases H_2 and CO production, which increases gas heating value and gasifier CGE. Air preheating is more effective at low ERs. A critical air temperature exists after which additional preheating has little influence. Steam

[*] Corresponding author. Tel.: +353 1 4022976; fax: +353 1 4023991.
E-mail address: wayne.doherty@dit.ie (W. Doherty).

has better reactivity than fuel bound moisture. Increasing moisture degrades gasifier performance therefore the input fuel should be pre-dried. Steam injection should be employed if a H_2 rich syn-gas is desired.

Keywords: Biomass gasification; Modelling; Circulating fluidised bed; Gibbs free energy minimisation; Equivalence ratio; Steam injection

1. Introduction

Biomass is of major interest as a renewable energy source in the context of both climate change mitigation and energy security. Energy security is of utmost importance and is vital for any country's continued economic growth. Currently Ireland imports 90% of its energy [1]. According to a recent study, oil and gas prices are set to double by 2050 [2]. Also global energy demand is set to more than double by the middle of the century, fuelled by the rapid increase in the energy demand of developing countries. Another study, World Energy Outlook [3] predicts similar trends. This study predicts a 53% increase in global energy demand by 2030, 70% of which will come from China, India and other developing countries and a 55% increase in carbon dioxide (CO_2) emissions. Biomass gasification coupled with other renewable energy options could cut dependency on imported energy and would help to ensure energy security.

Gasification is a process for converting carbonaceous materials to a combustible or synthetic gas [4]. It can be considered an upgrading process that takes in a solid which is difficult to handle, strip it of undesirable constituents and convert it into a gaseous product that can be handled with maximum convenience and minimum cost and can readily be purified to a clean fuel or feedstock for synthesis of other chemicals [5]. Gasification occurs when oxygen (O_2) or air and steam or water is reacted at high temperatures with available carbon in biomass or other carbonaceous material within a gasifier. The syn-gas produced can be combusted in an engine or gas turbine to generate electricity and heat or more recently syn-gas is considered a candidate fuel for fuel cell applications. Air gasification produces a poor quality gas with regard to the heating value, around 4-7 MJ m^{-3} higher heating value (HHV), while O_2 and steam blown processes result in a syn-gas with a heating value in the range of 10-18 MJ m^{-3} (HHV) [6]. However, gasification with pure O_2 is not practical for biomass gasification due to prohibitively high

costs for O_2 production using current commercial technology (cryogenic air separation). This has encouraged research of novel gasification processes such as the dual indirectly heated fluidised bed [6-9]. The modelling of air gasification is conducted in this research work.

Stoichiometric combustion occurs when all the carbon in the fuel is converted to CO_2 and there is no excess O_2 left over. The basis of gasification is to supply less oxidant than would be required for stoichiometric combustion of a solid fuel. The resulting chemical reactions produce a mixture of carbon monoxide (CO) and hydrogen (H_2), both of which are combustible. The energy value of this gaseous fuel is typically 75% of the chemical heating value of the original solid fuel. In addition, the syn-gas temperature will be substantially higher than the original solid fuel due to the gasification process. The process of biomass gasification is represented by the reactions given in Table 1.

Table 1
Gasification reactions

Reaction	Heat of reaction[a]	Reaction name	Reaction number
Heterogeneous reactions:			
$C + 0.5O_2 = CO$	(-111 MJ kmol^{-1})	Char partial combustion	(R1)
$C + CO_2 \leftrightarrow 2CO$	(+172 MJ kmol^{-1})	Boudouard	(R2)
$C + H_2O \leftrightarrow CO + H_2$	(+131 MJ kmol^{-1})	Water-gas	(R3)
$C + 2H_2 \leftrightarrow CH_4$	(-75 MJ kmol^{-1})	Methanation	(R4)
Homogeneous reactions:			
$CO + 0.5O_2 = CO_2$	(-283 MJ kmol^{-1})	CO partial combustion	(R5)
$H_2 + 0.5O_2 = H_2O$	(-242 MJ kmol^{-1})	H_2 partial combustion	(R6)
$CO + H_2O \leftrightarrow CO_2 + H_2$	(-41 MJ kmol^{-1})	CO shift	(R7)
$CH_4 + H_2O \leftrightarrow CO + 3H_2$	(+206 MJ kmol^{-1})	Steam-methane reforming	(R8)
Hydrogen sulphide (H_2S) and ammonia (NH_3) formation reactions:			
$H_2 + S = H_2S$	nr[b]	H_2S formation	(R9)
$0.5N_2 + 1.5H_2 \leftrightarrow NH_3$	nr	NH_3 formation	(R10)

[a][10] Negative sign indicates an exothermic reaction and a positive sign indicates an endothermic reaction.
[b]nr = Not reported.

The gasification process can be split into three linked processes; pyrolysis (also called devolatilisation, thermal decomposition or carbonisation), gasification, and combustion. Gasification and combustion may be combined, for example Di Blasi et al. [11] described gasification as two stages: solid pyrolysis and char conversion (gasification and combustion). Partial combustion is necessary because it supplies the heat required by the endothermic gasification reactions. Pyrolysis occurs in a temperature range of 350-800 °C and results in the production of char, CO, H_2, methane (CH_4), CO_2, H_2O, tars and hydrocarbons. Tars are extremely undesirable because they represent a loss of efficiency and degrade downstream plant equipment. If temperatures are high enough some tars will be cracked to form H_2, CO, CO_2 and others. The products of pyrolysis are then used in the gasification and combustion reactions.

To date there has been a lack of research and reporting on biomass gasification, which makes it difficult to design plants. The primary objective of this research was to develop a computer simulation model of a CFB biomass gasifier that can accurately predict gasifier performance under various operating conditions. In this paper an original model of a biomass CFB gasifier developed using the commercial chemical process simulator ASPEN Plus is presented. The model presented is based on Gibbs free energy minimisation. The approach assumes that only a limited number of chemical reactions (R1 to R10) are required with respect to the prediction of syn-gas composition, gas heating value and process efficiency. The influence of operating conditions on gasifier performance was investigated and the results and conclusions from these investigations are presented.

2. Methodology

2.1 Gasifier types and selection

The differences between classifications are in the movement of the fuel through the vessel, the operating pressures and temperatures and the size and condition of the entering fuel [12]. The primary configurations are moving/fixed bed, fluidised bed, and entrained flow. Atmospheric CFB was selected for the current study because of its near commercial status. This technology is proven for biomass gasification, has potential for scale-up (low MW to over 100 MW) and high fuel flexibility. The Värnamo biomass integrated gasification combined cycle (BIGCC) demonstration plant, which operated in Sweden from 1996 to 2000, used pressurised CFB gasification technology. Fig. 1 shows a schematic of a typical CFB gasifier.

2.2 Process simulation software

ASPEN Plus was selected for modelling the gasifier. This simulation package has been used for modelling coal and biomass power generation systems in many research projects [13-23]. It is a steady state chemical process simulator, which was developed at Massachusetts Institute of Technology (MIT) for the US DOE, to evaluate synthetic fuel technologies. It uses unit operation blocks, which are models of specific process operations (reactors, heaters, pumps etc.). The user places these blocks on a

flowsheet, specifying material and energy streams. An extensive built in physical properties database is used for the simulation calculations. The program uses a sequential modular (SM) approach, i.e. solves the process scheme module by module, calculating the outlet stream properties using the inlet stream properties for each block. ASPEN Plus has the capability to incorporate Fortran code, written by the user, into the model. This feature is utilised for the definition of non-conventional fuels, e.g. biomass, municipal solid waste (MSW), specific coals and for ensuring the system operates within user defined limits and constraints. User models can be created in Excel or written using Fortran and can be fully integrated into the ASPEN Plus flowsheet.

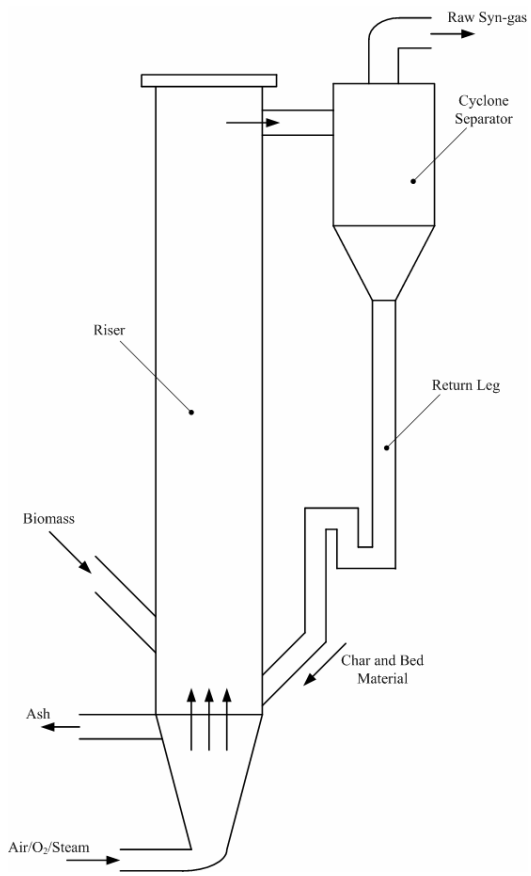

Fig. 1. Schematic diagram of a CFB biomass gasifier.

2.3 Uncoupling the gasification process

To model a CFB gasifier using ASPEN Plus, the overall process must be broken down into a number of sub-processes. For example a model may include the following zones: drying and pyrolysis,

partial oxidation, and gasification. The modeller may choose to model each of these zones separately or combine them in one unit. Fig. 2 shows the overall gasification process broken down or uncoupled into its sub-processes.

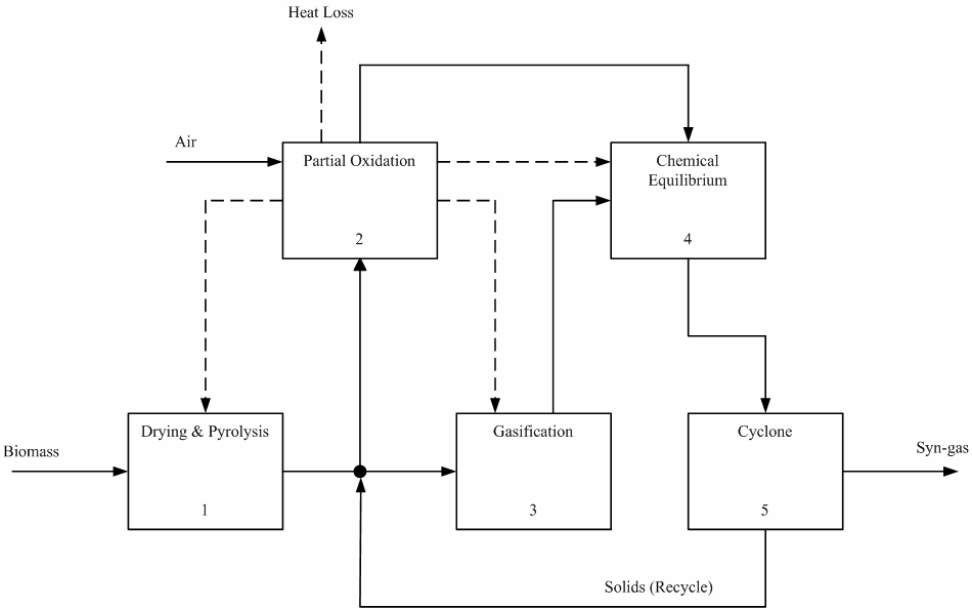

Fig. 2. Uncoupled CFB gasification process.

The drying and pyrolysis zone simulates the first stage of gasification and produces char, H_2, CO, CH_4, CO_2, H_2O, other hydrocarbons, and tars. These products are then either burnt or gasified. The partial oxidation zone simulates the burning of char as well as some H_2 and CO, which generates the heat required for all the sub-processes. This heat is represented by broken lines in Fig. 2. A percentage of the heat generated is lost from the system and products other than heat from this zone include CO, CO_2, and H_2O. The third zone, the gasification zone, simulates the gasification reactions, reactions such as the Boudouard, the water-gas and the methanation. The products of both the partial oxidation and the gasification zone are fed into an additional zone. This zone sets the final syn-gas composition, which is composed mainly of H_2, CO, CO_2 and some CH_4. In this zone the chemical equilibrium of the gasification reactions is restricted in order to give a realistic syn-gas composition. The final zone, box 5, represents the CFB cyclone separator, which separates out and recycles the solids entrained in the gas.

3. Modelling

3.1 ASPEN Plus flowsheet

Fig. 3 displays the CFB biomass gasifier ASPEN Plus flowsheet. Table 2 presents a brief description of the unit operation blocks shown in Fig. 3. It gives the ASPEN Plus name, that is the name given to each unit operation block by the software developers, the block ID, which is the name given to each block by the user and a short description.

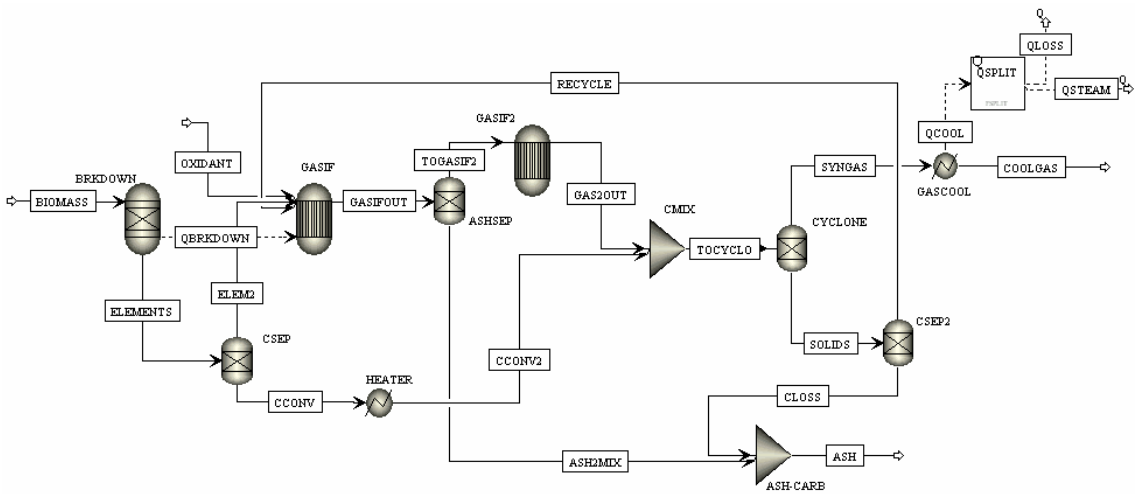

Fig. 3. ASPEN Plus flowsheet of biomass CFB gasifier.

Table 2
Description of ASPEN Plus flowsheet unit operation blocks presented in Fig. 3

ASPEN Plus name	Block ID	Description
RYIELD	BRKDOWN	Yield reactor – converts the non-conventional stream 'BIOMASS' into conventional components
SEP2	CSEP	Separator – extracts a portion of the carbon contained in the fuel so that it remains un-reacted
	ASHSEP	Separator – separates the inert ash from the gas to allow removal from the system
	CYCLONE	Separator – simulates the CFB cyclone by separating out a specified percentage of the solid carbon
	CSEP2	Separator – extracts a portion of the carbon to simulate carbon loss in the ash, with the rest recycled
RGIBBS	GASIF	Gibbs free energy reactor – simulates drying and pyrolysis, partial oxidation, and gasification
	GASIF2	Gibbs free energy reactor – restricts chemical equilibrium of the specified reactions to set the syn-gas composition
HEATER	HEATER	Heater – increases the temperature of the un-reacted carbon to the reactor temperature
	GASCOOL	Cooler – simulates syn-gas cooling to a typical gas cleanup temperature
MIXER	CMIX	Mixer – mixes the un-reacted carbon separated in block 'CSEP' with the syn-gas
	ASH-CARB	Mixer – mixes the carbon lost with the ash before leaving the system
FSPLIT	QSPLIT	Splitter – splits the heat available from syn-gas cooling in 'GASCOOL' into two heat streams with one of them representing the heat lost from the gasifier

3.2 Model description

The main model assumptions are: steady state conditions, zero-dimensional model, isothermal (uniform bed temperature), drying and pyrolysis are instantaneous in a CFB [24, 25], char is 100% carbon (graphite), all of the sulphur reacts to form H_2S [6], only NH_3 formed no nitrogen oxides considered [6, 12, 23], cyclone separation efficiency is 85% [26], 2% carbon loss in ash [27], and heat loss from the gasifier is equal to 3% of the total heat input [16, 28, 29].

From Fig. 3, the stream 'BIOMASS' was specified as a non-conventional stream and the ultimate and proximate analyses were inputted. The stream thermodynamic condition and mass flow rate were also entered. The block 'BRKDOWN' yields are set by a calculator block, which in turn determines the mass flow of each component in the block outlet stream 'ELEMENTS'. The enthalpy of this stream will not equal the enthalpy of the feed stream 'BIOMASS', as the enthalpies of the individual constituents that make up a fuel do not equal the enthalpy of the fuel because chemical bonds etc. are not taken into consideration. The energy value of any heat stream leaving a block is equal to the block heat duty, which is equal to the difference in enthalpy between the inlet and outlet streams. Thus, the heat stream 'QBRKDOWN' was inserted to add back the enthalpy loss to the system.

The function of the next block is to simulate carbon conversion by separating out a specified portion of the carbon from the fuel. Reported carbon conversion for CFB gasifiers in the literature ranged from 90 to 99% [29-32]. Before this carbon can be mixed with the gas downstream it must be brought up to the gasifier temperature, which is accomplished using the block entitled 'HEATER'. The un-reacted carbon represents solids contained in the product gas that must be removed by the CFB gasifier cyclone or other solids removal steps downstream. In reality there would also be fly ash and bed material entrained in the gas but these components cannot be modelled in ASPEN Plus. Thus, in this model the solid carbon that remains in the syn-gas represents all solids.

The streams 'ELEM2', 'OXIDANT', and 'RECYCLE' enter the block 'GASIF', where pyrolysis, partial oxidation, and gasification reactions occur. The mass flow of air entering the reactor is set using a design specification, which varies the oxidant mass flow rate so that a specific gasifier temperature is achieved. Alternatively, the air mass flow is set by a calculator block that calculates the air flow using a user specified equivalence ratio (ER). ER is defined as the ratio of the actual oxidant mass flow rate to the stoichiometric oxidant mass flow rate. Most existing fluidised bed biomass gasifiers

operate in the ER range 0.2-0.45 [33]. All the sulphur in the system reacts with H_2 to form H_2S. Due to the low contents of sulphur in the fuel, inaccuracies of this simplification are negligible [6]. The simplification that only NH_3 is formed and nitrogen oxides are omitted was adopted in this work. This assumption has been applied by others such as Pickett [12]; Schuster et al. [6]; and Zhu [23]. Char, which is a product of pyrolysis, is assumed to be 100% carbon (graphite). Demirbaş [34] reported the elemental analysis of various wood chars and the carbon content ranged from 90.5 to 92.1 wt. %, therefore the assumption is valid.

Ash removal is simulated in the model using the unit operation block 'ASHSEP'. The material stream 'TOGASIF2' is fed to the unit operation block 'GASIF2', which is an 'RGIBBS' reactor. 'RGIBBS' reactors allow restricted equilibrium specifications for systems that do not reach complete equilibrium. Specifying the temperature approach for each reaction results in restricted equilibrium, which means that the syn-gas composition can be adjusted to match data reported in the literature. This approach is well established, it was introduced by Gumz [35]. Reactions (R7), (R8), and (R10) were specified in the 'RGIBBS' reactor.

The next block mixes the un-reacted carbon that was separated upstream with the gas from 'GASIF2' and its product stream is fed to a separator that simulates the operation of the CFB gasifier cyclone. The block 'CYCLONE' was specified so that it removes 85% of the solid carbon from the gas stream [26]. Zhang and Basu [26] investigated the operation of a CFB cyclone by experiment and developed a model that predicted the cyclone collection efficiency for different particle diameters and gas velocities and showed that the efficiency ranged from 70 to 100%. The average value of 85% was used in this work. The bottom outlet stream from 'CYCLONE' with the stream name 'SOLIDS' is composed of solid carbon only and is sent to a separator block 'CSEP2'. The top outlet stream, which is called 'SYNGAS', is composed of all the gases from 'GASIF2' and a small amount of solid carbon (15% of the un-reacted carbon). This material stream represents the final output, i.e. the product gas from the gasifier.

'CSEP2' splits the 'SOLIDS' stream into a recycle stream 'RECYCLE', that is sent back through the gasifier, and another stream named 'CLOSS', which represents the carbon lost from the system in the ash. The recycle was added because in a real CFB gasifier, inerts (bed material and fly ash) and un-reacted char are collected in the cyclone and re-injected into the reaction zone of the gasifier via the return leg (see Fig. 1). The 'CSEP2' split fraction is set by a calculator block using the specification

that the ash exiting the gasifier contains 2% carbon [27]. The stream 'CLOSS' is then mixed with the ash in the block 'ASH-CARB'.

The stream 'SYNGAS' is fed to a cooler entitled 'GASCOOL' that cools the gas to the required gas cleanup temperature of 375 °C [36]. The energy that would be lost through cooling could be recovered by generating steam or by supplying heat for air preheating. The heat stream 'QCOOL' represents the energy that could be recovered during gas cooling. This stream is fed to 'QSPLIT', which is used to split the heat stream 'QCOOL' into two heat streams 'QLOSS' and 'QSTEAM'. Two calculator blocks are used, one calculates and sets the amount of heat loss 'QLOSS' and the other calculates and sets the amount of heat available for steam generation or air preheating 'QSTEAM'. The heat loss from the gasifier is assumed to be 3% of the total heat input [16, 28, 29].

3.3 Model validation

The model was validated against the experiments of Li et al. [27], which were conducted on a pilot scale air-blown biomass CFB gasifier. The fuel used for model validation is hemlock wood. The ultimate and proximate analyses for the wood are given in Table 3.

Table 3
Ultimate and proximate analyses of hemlock (wood)

Ultimate Analysis (dry basis)[a]		
Carbon	wt. %	51.8
Hydrogen	wt. %	6.2
Oxygen	wt. %	40.6
Nitrogen	wt. %	0.6
Sulphur	wt. %	0.38
Chlorine	wt. %	0.0
Ash	wt. %	0.4
	wt. %	100
Proximate Analysis (dry basis)		
Volatile Matter[b]	wt. %	84.8
Fixed Carbon[c]	wt. %	14.8
Ash[a]	wt. %	0.4
	wt. %	100
Moisture[a]	wt. %	11.7
HHV (dry basis)[a]	MJ kg^{-1}	20.3

[a][27].
[b][37] Material: wood, fir, western hemlock and ID-number: 242.
[c]Determined by difference.

Li et al. [27] reports results for six experimental runs using hemlock wood as input fuel. The input data for three of these runs (run 4, 6 and 7) were entered into the model and the predictions were found to be in good agreement with the reported results. For example for run number 4 the model

predicts the following syn-gas composition: 3.23% H_2, 72.82% N_2, 8.9% CO, 15.04% CO_2 and 0.01% CH_4 and for the same input data Li et al. [27] reports 3.0% H_2, 68.4% N_2, 9.6% CO, 17.1% CO_2 and 1.9% CH_4. Experimental run number 7 [27], was chosen for a detailed comparison and analysis. The input data for run number 7 are as follows: input fuel stream mass flow - 33.626 kg h^{-1}, gasification temperature - 991 K, and gasification pressure - 1.05 bar.

Table 4 compares the experimental results (run 7) as reported by Li et al. [27] to the model predictions using the input data presented above. The model predictions are in good agreement with the experimental data. For example H_2, CO and CO_2 are predicted within 2.5% and N_2 is under-predicted by 6.8%. However the CH_4 is over-predicted, which causes an error in the calculation of the gas heating value and ultimately the CGE. The under or over-prediction of methane is quite a common problem for modellers; the product gas of fluidised bed gasifiers generally contains tar, which is not considered in equilibrium models, and much more hydrocarbons (especially methane) than predicted [38]. Also, the model done by Giltrap et al. [39] which is a steady state model of a biomass downdraft gasifier, over-predicts CH_4 by a substantial amount. The low operating temperature (991 K) results in high CH_4 content and the CH_4 content decreases rapidly with temperature (at ~870 °C the model predicts virtually zero CH_4). This is further discussed in section 4.

Table 4
Experimental results versus model predictions

	Experimental (run # 7)	**Model**
Gas composition[a]		
H_2	5.5	5.53
N_2	59.5	55.42
CO	16.6	16.79
CH_4	3.4	7.65
CO_2	15.0	14.62
HHV[b]	4.82	5.87
CGE[c]	71.4	62.61

[a]Volume %, dry basis.
[b]MJ m^{-3}, dry basis at 0 °C and 1 atmosphere.
[c]%.

4. Sensitivity analysis and discussion

The model described was used to perform sensitivity analyses. The effects of varying ER, temperature, level of air preheating, biomass moisture and steam injection on product gas composition, gas heating value, and CGE were investigated. During the sensitivity analyses the model input data was

kept the same as for model validation (run 7 input data), which was presented in section 3.3, with one parameter being varied at any given time. The CGE (η_{CGE}) is a means of indicating the performance of a gasifier and is defined as: $\eta_{CGE} = \dot{m}_{gas} \cdot HHV_{gas} / \dot{m}_{fuel} \cdot HHV_{fuel}$. Where $\dot{m}_{gas}$ and $\dot{m}_{fuel}$ are the mass flow rate (in kg s^{-1}) of the gas and fuel respectively and HHV_{gas} and HHV_{fuel} are the higher heating value (in kJ kg^{-1}) of the gas and fuel respectively.

4.1 Effect of ER and gasification temperature (T_g)

The influence of ER on product gas composition is illustrated in Fig. 4. T_g depends on the amount of air fed to the gasifier, i.e. it is controlled by the ER. As a result, varying ER or T_g will have the same effect on product gas composition, heating value, and CGE. For this reason only ER is plotted against product gas composition, heating value and CGE. The corresponding temperatures for each ER are also given. In Fig. 4 H_2, H_2O, CO, CO_2, and CH_4 are plotted, with H_2S and NH_3 omitted, first of all because of their very low content and second because they would be removed by downstream air pollution control equipment in a gasification plant. All gas components are plotted on a dry basis except for water vapour. The nitrogen (N_2) content is not displayed; its value may be calculated by summing the other components and subtracting this from 100%. The N_2 content varied between 53 and 61% over the ER and T_g range. The most interesting point from Fig. 4 is that both H_2 and CO reach a maximum at an ER of 0.35 or at a temperature of 874 °C. After this peak their contents decrease steadily. H_2O increases over the whole range but experiences a small decrease close to the point of maximum H_2 and CO. CO_2 decreases rapidly up to an ER of 0.35 and then increases slowly. CH_4 decreases and eventually reaches zero between an ER of 0.4 and 0.45 or a temperature of 1046 and 1195 °C. These trends may be explained as follows:

- The Boudouard reaction (R2) is endothermic; therefore as the temperature rises, so does the amount of CO_2 reacted with char to produce CO. For ERs up to 0.35 sufficient char is available for the Boudouard reaction but for ERs greater than this there is insufficient char, so CO decreases and CO_2 increases.
- The water-gas reaction (R3) is endothermic, which means for increasing ER and temperature CO and H_2 production are increased and more char and H_2O are consumed.

- The methanation reaction (R4) is exothermic, which means as ER and T_g increase the production of CH_4 decreases, which in turn leaves more H_2 in the gas.
- The CO is reacted with available O_2 (R5) producing CO_2.
- H_2 is reacted with available O_2 (R6) producing H_2O. This reaction produces more water than is used up by the reactions (R3) and (R8) because the H_2O content increases over the whole ER and temperature range. The slight drop in H_2O occurs at an ER of 0.34 and a T_g of 837 °C. One possible explanation would be that at a sufficiently high temperature reaction (R8) begins to consume more water than is produced by reaction (R6); however, this trend is short-lived because the other reactant required for this reaction, CH_4 is decreasing rapidly.
- The CO shift reaction (R7) being exothermic, produces less CO_2 and H_2 at higher temperatures, which means less CO and H_2O are consumed.
- The CH_4 is reduced by the steam-methane reforming reaction (R8). This reaction is endothermic meaning the forward reaction is favoured as temperature increases. Hence, CH_4 and H_2O decrease while H_2 and CO increase.

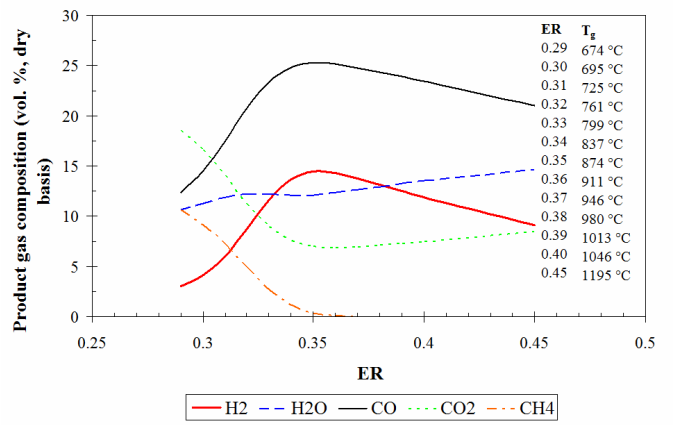

Fig. 4. Effect of ER on product gas composition.

Commercial fluidised bed biomass gasifiers operate in the temperature range 800-1000 °C, which for this model corresponds to an ER range of 0.33-0.38. ER values higher than 0.3 have to be used to get tar contents below 2 g m^{-3} [40]. Taking these points into consideration the authors recommend operation at ER = 0.34 to 0.35 or T_g = 837 to 874 °C.

The influence of ER on product gas heating value and CGE is shown in Fig. 5. The gas HHV on a mass basis and the CGE on a HHV basis are plotted against ER. It is evident that the heating value

decreases with increasing ER. The HHV is high for low ERs due to the high CH_4 content. The CGE increases between an ER of 0.29 and 0.34, it reaches a maximum value of 66.1% at an ER of 0.34 and then decreases steadily. It is worth noting that the point of maximum CGE corresponds to the point of maximum H_2 and CO content in Fig. 4. The CGE for an ER of 0.31 as reported by Li et al. [27] is indicated on the plot. It is indicated for comparison with the model prediction at the same ER value.

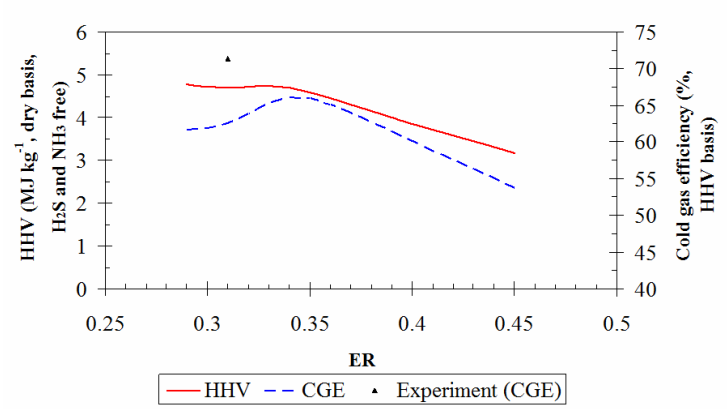

Fig. 5. Effect of ER on gas HHV on a mass basis and CGE on a HHV basis (▲: indicates CGE as reported by Li et al. [27]).

4.2 Effect of air preheating

Air preheating is a means of increasing the conversion efficiency of the gasification process. The sensible heat in the air causes a rise in the gasification temperature, which in turn influences the product gas composition, causing an increase in the production of combustible gases, H_2 and CO. This change in syn-gas composition affects the gas HHV and hence the gasifier CGE. Air preheating offers an alternative and more economical approach than oxygen blown systems [41]. The overall efficiency of the process on a thermal basis would be increased if the heat required for air preheating was recovered from the gas cooling section of the plant. Use of high temperature air as an oxidant achieves downsizing of the plant [42]. Downsizing is achieved because a smaller volume of air is needed to bring the gasifier to the required operating temperature, which in turn reduces the size of the reactor and gas cleanup system needed.

The influence of air preheating on the reactor or gasification temperature T_g was investigated over the complete ER range. It was found that T_g increased almost linearly with air temperature (T_a) for

all ERs. It was discovered that a limit on the level of air preheating exists for each ER. This level is limited by the effectiveness of the heat exchange equipment used to transfer the heat from the syn-gas to the air but it is also limited by the operating temperature constraint of fluidised beds. Fluidised bed biomass gasifiers should not be operated over 1000 ℃, so as to ensure that the ash melting temperature is not reached, which would cause agglomeration and de-fluidisation. It was found that at high ERs air preheating is limited to a low level, e.g. at an ER of 0.37 a T_a no more than 114 ℃ would be recommended because the corresponding T_g is 987 ℃ whereas for an ER of 0.29 the air could in theory be heated to 825 ℃ as the T_g stays below the limit at a value of 978 ℃.

The influence of T_a on product gas composition is shown in Fig. 6. The product gas composition for an ER value of 0.29 is plotted against T_a. The gas composition changes reflect the change in the gasifier temperature. The rising temperature promotes the products of the endothermic reactions, (R2), (R3) and (R8), and simultaneously the reactants of the exothermic reactions (R4) and (R7). A detailed discussion on the effects of gasifier temperature was given in section 4.1. The major conclusion drawn from this sensitivity analysis is that T_a has a greater influence on the product gas composition for low ERs. For an ER of 0.29 CO and H_2 content increases 17.3 and 15.8 percentage points respectively over the T_a range whereas for an ER of 0.34 CO and H_2 content increases by only 2.7 and 1.8 percentage points respectively over the same temperature range. It was also found that T_a has a significant influence on composition only up to a certain level, after which additional preheating has little effect. For an ER of 0.29 this T_a is high at a value of ~560 °C but for an ER of 0.34 it is significantly lower at ~200 °C. This finding agrees with published work [43, 44]. Lucas et al. [43] reported that the H_2 rises with increasing air preheat temperature but exhibits no rise between 700 and 830 ℃. Yang et al. [44] refers to a critical T_a above which air preheating is no longer efficient if the purpose is to maximise the yield of gaseous products. This critical T_a for CO and H_2 was reported as 530 ℃. The results of this work indicate a critical temperature of ~560 ℃ for an ER of 0.29.

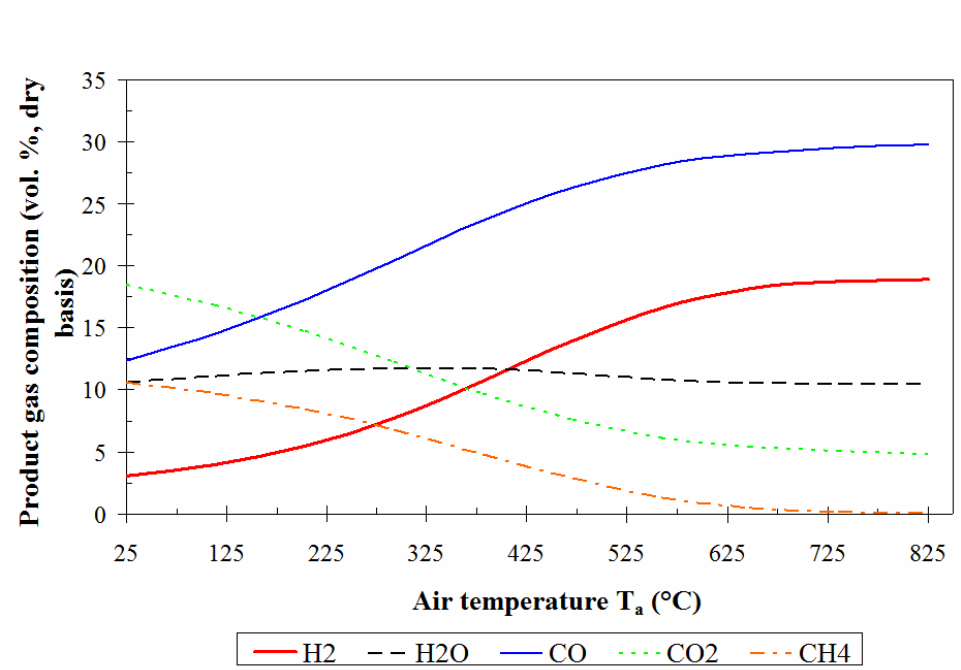

Fig. 6. Effect of T_a on product gas composition for ER = 0.29.

The influence of air preheating on product gas heating value and CGE is shown in Fig. 7. The gas HHV on a mass basis and the CGE on a HHV basis, plot (a) and (b) of Fig. 7 respectively, were plotted against T_a for the complete range of ER. As expected, the gas heating value increases with T_a. The increase is in line with the gas composition change for each ER, i.e. the increase is greater for low ERs than for high ERs due to the greater change in gas composition for low ERs. The CGE trends are in agreement with the changes in gas composition and HHV. T_a has a significant influence on CGE at low ER values. Its influence ceases for ERs greater than 0.35. As already seen for gas composition, T_a has a significant effect on CGE only up to a certain level, after which additional preheating has little influence. For an ER of 0.29 this T_a is high at a value of ~650 °C but for an ER of 0.33 it is significantly lower at ~290 °C. For an ER of 0.34, which is a point of interest because it was the point of maximum CGE for gasification without air preheating, the CGE increases from 66.1 to 67.2% for a T_a of ~110 °C and then increases by a lesser degree to 67.7% for a T_a of ~200 °C.

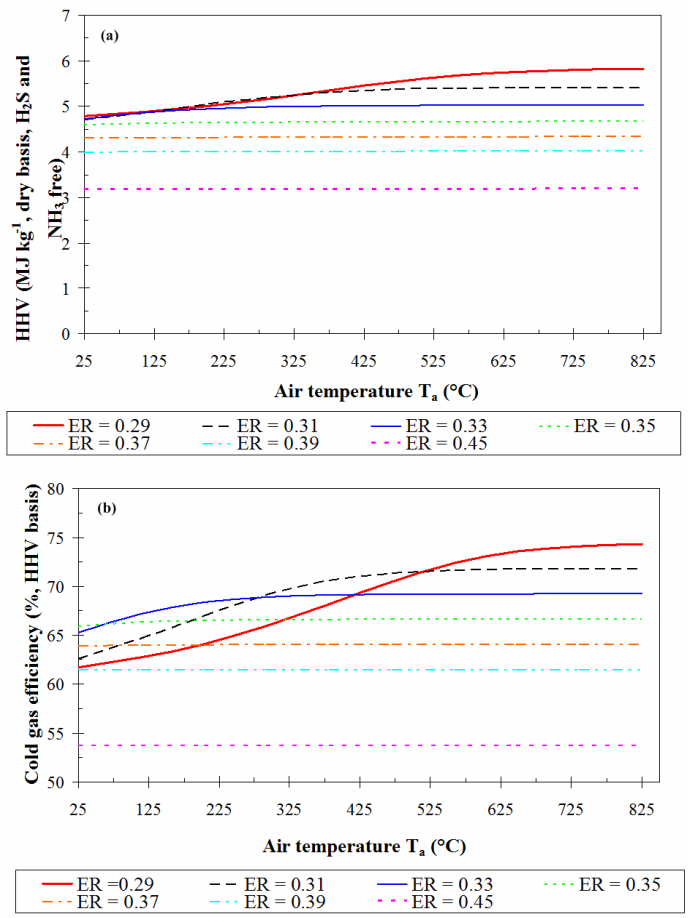

Fig. 7. Effect of T_a on (a) gas HHV on a mass basis and (b) CGE on a HHV basis for complete ER range.

4.3 Effect of biomass moisture and injected steam

The effect of fuel bound moisture on gasifier performance for ER = 0.34 is shown in Fig. 8 (a). The moisture level was varied over a realistic range for woody biomass (5 – 30%) and as expected increasing the moisture content degrades gasifier performance. Both syn-gas HHV and gasifier CGE reach their maximum level, 5.138 MJ kg^{-1} and 73.81% respectively, at the lowest moisture content (5%). The gas heating value and CGE decrease over the entire moisture range, for comparison HHV = 3.338 MJ kg^{-1} and CGE = 44.24% at a moisture content of 30%. These trends are a direct result of changes in the syn-gas composition with moisture. The rising H_2O content is the main cause for the decline in syn-gas HHV. CO and CH_4 are shifted and reformed respectively with the additional H_2O decreasing their contents and producing CO_2. There is little change in H_2 content at this ER however at lower ERs H_2 content was found to increase with moisture level. Similar performance trends were seen for other ERs but maximum gasifier performance was predicted for an ER range of 0.34 to 0.35.

The influence of steam injection on gasifier performance for an ER of 0.34 is illustrated in Fig. 8 (b). The steam injection rate was varied from 0 to 10.5 kg h^{-1} [27]. The syn-gas HHV decreases only slightly from 4.69 to 4.62 MJ kg^{-1} and gasifier CGE increases from 66.1 to 66.5%. This small increase in CGE is due to increased syn-gas mass flow rate. As was seen for increasing moisture level, steam injection causes a rise in H_2O content, which results in a lower syn-gas HHV. CO and CH_4 are shifted and reformed respectively with the additional H_2O decreasing their contents and producing CO_2. The most important effect of steam injection is the rise in H_2 content, in this case H_2 increases by 3% (13.7 to 16.7%) over the range of steam injection. The gasifier temperature will decrease with increasing steam injection due to the highly endothermic reforming and water-gas reactions (R3 and R8) unless heat is supplied from an external source. A decrease in temperature is undesirable as this would degrade gasifier performance and could lead to high tar yield. Similar performance trends were seen for other ERs but maximum increase in H_2 content was predicted for an ER range of 0.34 to 0.35.

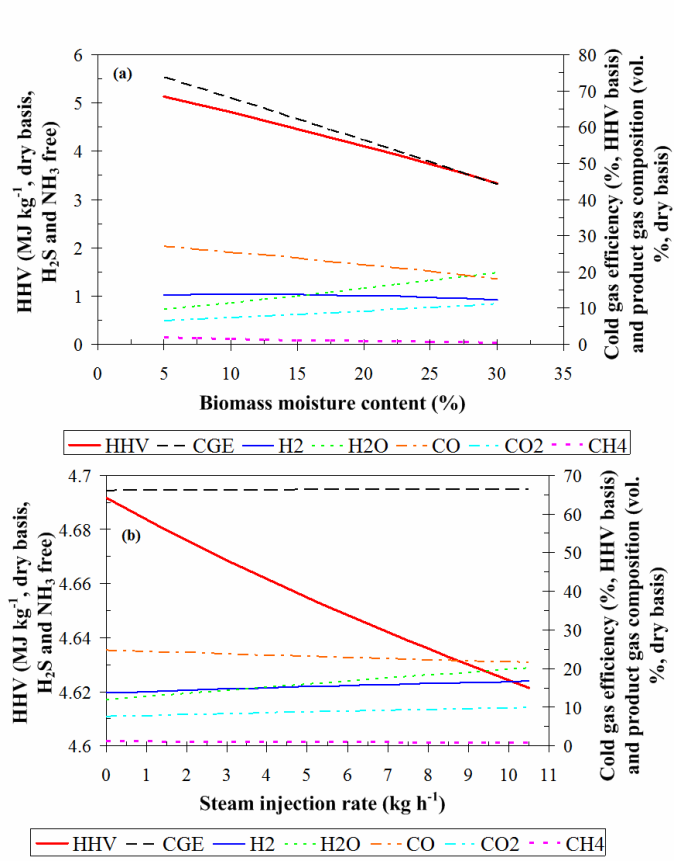

Fig. 8. Effect of (a) biomass moisture content and (b) steam injection rate on product gas composition, gas HHV on a mass basis and CGE on a HHV basis for ER = 0.34.

The main conclusion from the analysis above is that injected steam, although chemically equivalent, has better reactivity than fuel bound moisture. A possible reason for this is that the time it takes to bring the moisture up to the gasifier temperature is greater than the residence time and therefore the moisture does not participate in the chemical reactions. This finding agrees with other work [27]. The results indicate that the input fuel should be pre-dried to ensure low moisture content and if a H_2 rich syn-gas is required steam injection should be employed. Air preheating should be considered when using high moisture fuels and steam injection because air preheating causes an increase in the gasifier temperature (see section 4.2), which would in turn offset both fuel pre-drying and the temperature decrease as a result of steam injection.

5. Conclusions

A computer simulation model of a circulating fluidised bed biomass gasifier was developed using ASPEN Plus. The model was calibrated against experimental data. The restricted equilibrium method was used. In addition the results obtained from the sensitivity analyses are in good agreement with published work. Therefore, it may be said that the model is capable of predicting accurately gasifier performance over a wide range of operating conditions. The effects of varying ER, temperature, level of air preheating, biomass moisture and steam injection on product gas composition, gas heating value, and CGE were investigated, the results of which revealed the following:

- Without air preheating, the optimum operating conditions were found to be: ER = 0.34 to 0.35 and gasifier temperature = 837 to 874 °C.
- The syn-gas heating value decreases with increasing ER.
- Without air preheating, the CGE reaches a maximum value of 66.1% at an ER of 0.34.
- Air preheating increases the production of combustible gases, H_2 and CO, which increases the product gas heating value and the gasifier CGE.
- Air preheating is more effective at low ERs. It should not be used for ERs greater than 0.35.
- A critical T_a exists after which additional preheating has little influence. This temperature is high for low ERs and low for high ERs.
- For ER = 0.34, which is the point of maximum CGE for gasification without air preheating, the CGE increases 1.6 percentage points for a low level of air preheating (T_a = 200 °C).

- If economical, the gasifier could be operated using very high temperature air, which would mean it could operate using a low ER. A lower ER means reduced volume and hence, the size and cost of the gasifier, gas cleanup system and other plant components would be reduced.
- Steam has better reactivity than fuel bound moisture.
- Increasing moisture degrades gasifier performance therefore the input fuel should be pre-dried.
- Steam injection should be employed if a H_2 rich syn-gas is desired.

In a future study, the biomass gasifier model presented here will be integrated with a high temperature fuel cell stack model and balance of plant models all developed in ASPEN Plus.

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