Integrating advanced biomass gasifiers into the New Zealand wood industry

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Abstract

Biomass gasification offers an appealing cogeneration option for the energy intensive wood industry. The appeal of biomass gasification stems from the fact that gasification transforms a solid fuel, often waste, into a gaseous fuel which retains 75-88% of the heating value of the original (Higman and Burgt, 2003). A gaseous fuel offers easier handling and the ability to be utilized in either a gas engine or a gas turbine. Conventional biomass cogeneration plants utilize steam turbines and manage an electrical efficiency of 15-28%, while integration of a gasifier with a gas turbine or engine allow efficiencies of 25-40% (Franco and Giannini, 2005). This paper presents a model for assessing the feasibility of gasification based cogeneration plants. Firstly a chemical equilibrium model for a fast internally circulating fluid bed biomass gasifier is presented allowing estimation of the product gas composition. Secondly basic process flowsheets for two heat and power applications are considered; these are integration with a gas engine and integration with a gas turbine combined cycle.

Introduction

This work is undertaken to provide a modeling tool to evaluate the economic feasibility of an advanced gasification technology for woody biomass energy plant as part of a larger programme to develop biomass gasification systems for electricity production. There are a number of biomass energy plants constructed in the world but most of these plants are limited to demonstration scale. One of the issues is high capital costs and operation costs. However, these costs can be significantly reduced by optimizing the scale, location and the level of integration of the system. For example at small scales simple system may be more feasible whereas at large scales integrated systems may be preferred. The decision on these options will need both technology evaluation and economic analysis. The approach taken has been to model the gasifier using chemical equilibrium so that a product gas composition and heating value can be estimated. The gasification technology used in this work is fast internal circulating fluidized bed (FICFB) as reported by Brown et al. (2006). In conjunction process flow-sheets and costing models have been created for two possible energy plant concepts.

FICFB gasification

The FICFB gasifier produces a high hydrogen gas yield due to the use of steam as the gasifying agent. The endothermic nature of the gasification reactions combined with the use of steam as a gasifying agent requires that there is heat transfer to the gasification reactor in order for the gasification to take place. This is achieved through a twin bed system. The bubbling fluid bed (BFB) gasification reactor is combined with a circulating fluid bed (CFB) combustor. The CFB heats an inert heat carrying medium (sand) which flows from the CFB to the BFB providing the heat of reaction. A diagram of the system is shown below.

The BFB reactor is screw-fed woody biomass accompanied by a nitrogen purge gas. The nitrogen purge gas is used to ensure positive gas flow into the gasifier.

Figure 1: Diagram of FICFB gasifier



The biomass is fed in above the fluid bed. Drying and devolatilzation of the biomass occur immediately upon the biomass entering the reactor. The heterogeneous chargasification reactions have longer reaction rates (Kinoshita and Wang, 1993, Fiaschi and Michelini, 2001) and will occur throughout the BFB. The BFB has a sand bed fluidized with steam. During gasifying the bed will also contain significant amounts of char. The sand and char bed material flow from the BFB through a chute fluidized with either air or steam into the CFB. Inside the CFB, the char and any additional fuel in the form of LPG is combusted. The CFB is a sand bed fluidized with air. Air rates are maintained to provide excess air conditions. The CFB air velocity is significantly greater than the steam velocity in the BFB and hence the sand is entrained up and out of the CFB. The sand entrained out of the CFB is separated from the flue gases by a cyclone and fed back through a siphon into the BFB. The hot sand settles at the bottom of the siphon preventing flow of the BFB product gas out through the siphon. The sand is then fluidized with either air or steam up and over into the BFB. The sand, having passed through the combustion reactor, is hotter than the BFB bed and cools providing the heat for the gasification reactions. The product gas from the BFB flows out of the top of the BFB and through a cyclone, to separate particulates, before being burnt in an afterburner. When the FICFB is integrated into a process the afterburner would

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be replaced with either a boiler system, chemical reactor, gas engine or a gas turbine. Also, re-circulated product gas would be used to replace LPG as the fuel for the CFB.

Chemical equilibrium model for biomass gasification

The various flows around the BFB are simplified and modeled as shown in Figure 2. A steady-state equilibrium model has been developed to predict the composition of the product gas from a FICFB gasifier so that preliminary feasibility studies can be undertaken for the integration of a FICFB gasifier into different heat and power applications. The variables used in the gasification model are defined as follows.

Dependant variables

The variables determined by the model are the molar fractions and flows of the product gas species. These parameters define the quality and heating values of the product gas.

 y_{CH4} is the mol fraction of methane y_{H2} is the mol fraction of hydrogen y_{CO2} is the mol fraction of carbon dioxide y_{co} is the mol fraction of carbon monoxide y_{H2O} is the mol fraction of steam N_{gas} is the molar flow of product gas N_{CH4} is the molar flow of CH4 out of the BFB $N_{_{CO}}$ is the molar flow of CO out of the BFB $N_{\scriptscriptstyle CO2}\,$ is the molar flow of CO2 out of the BFB $N_{\rm \scriptscriptstyle H2}\,$ is the molar flow of H2 out of the BFB N_{H2O} is the molar flow of H2O out of the BFB $k_{(CO+3H, \bigcirc CH_4+H, O)}$ is the equilibrium constant at T_{BFB} for the steam methane reforming reaction $k_{(CO+H,O} \Leftrightarrow CO_{2}+H_{2})$ is the equilibrium constant at T_{BFB} for the water gas shift reaction

Independent variables

The independent variables are the inputs required for the model. These values can be determined from chemical analysis of the feed biomass or are determined by the operator.

 N_{char} is the molar flow of char

 N_{Steam} is the molar flow of steam into the BFB

 $N_{moisture}$ is the molar flow of moisture into the BFB

 $N_{_{\rm IVood}}$ is the molar flow of carbon in the wood into the BFB

H/C is the hydrogen to carbon ratio in the wood

O/C is the oxygen to carbon ratio of the wood

 N_{Purge} is the molar flow of nitrogen purge gas

 $T_{BFB}^{-m_s}$ is the temperature of the BFB and defines the equilibrium constants

Modeling Approach

The objective of the modeling is to determine the composition and concentration of the gasification product gas by considering the chemical reactions and mass balances in the gasification process. The model of the FICFB gasifier is based around equation 3, which represents the governing reaction for the BFB gasifier. Equations for three of the unknown variables are found using elemental balances for C,



H and O (Equations 4-6), which can be rearranged to show the dependence of the molar flow of CO, CO_2 and CH_4 on the molar flow of H_2 and H_2O . The mol fractions of H_2 and H_2O are found through chemical equilibrium (Equations 7-8) using the reactions 1-2. The equilibrium constants for equations 7 and 8 are derived using Gibbs energies from the TRC (1994). The molar flow of product gas can be found by Equation 9.

$$CO_2 + H_2 < => CO + H_2O$$
 (1)
 $CH_1 + H_2O < => CO + 3H_2$ (2)

Governing Reaction

$$N_{Wood} CH_{H/C} O_{O/C} + (N_{Steam} + N_{moisture}) H_2 O + N_{purge} N_2 = N_{gas} [y_{CH_{\downarrow}} y_{CO_{\downarrow}} + y_{CO_{\downarrow}} + y_{H_{\downarrow}} + y_{H_{\downarrow}O} + y_N] + N_{char} C$$
(3)

Carbon Balance $N_{wood} = N_{CH_4} + N_{CO_2} + N_{CO} + N_{char}$ (4)

Hydrogen balance

$$\frac{H}{C}N_{wood} + 2N_{steam} + 2N_{moisture} = 4N_{CH_{i}} + 2N_{H_{i}} + 2N_{HO}$$
(5)

Oxygen Balance

$$\frac{O}{C} N_{wood} + N_{steam} + N_{moisture} = 2N_{CO_2} + N_{CO} + N_{H_2O}$$
(6)

Steam
$$y_{HO} = \frac{y_{CO_2} y_{H_2}}{y_{CO} k_{[CO+HO <=> CO_2+H]}}$$
 (7)

Hydrogen

$$y_{\rm H_2} = \sqrt[3]{\frac{y_{\rm CH_4} y_{\rm H_2O}}{y_{\rm CO} k_{[CO+3H_2 \Leftrightarrow CH_4 + H_2O]}} \left(\frac{P_{reac}}{P_0}\right)^{1+1-3-1}}$$
(8)

Product Gas Yield

$$N_{gas} = \frac{N_{CH} + N_{CO} + N_{CO} + N_{purge}}{1 - y_{H_2O} - y_{H_2}}$$
(9)

Equations 4-9 provide a system of linear and non-linear equations which Microsoft Excel Solver is used to solve.

The model presented above was then further adapted to include the possibility of solid carbon as a product. At low temperatures and low H_2O to biomass ratios it is possible



for incomplete carbon conversion. Hence some carbon will remain in solid form and not be gasified. To enable the model to deal with solid carbon, the reaction set was adapted to include 3 reactions (Equations 10-12) in situations where carbon was present as a product.

$$C + H_2 O <=> CO + \dot{H}_2 \tag{10}$$

$$C + H_2 <=> CH_4$$
 (11)
 $C + CO_2 <=> 2CO$ (12)

$$C + CO_2 < -> 2CO$$
 (

Results

Figures 3 and 4 present the model results from gasifying one kmol of wood, modeled as $CH_{1.43}O_{0.62}$, with varying H_2O to biomass ratios and temperatures. Chemical equilibrium is dependant only on the elemental abundances in the reactor, hence moisture content of the wood has not been specified in reporting these results but is included in the H_2O to biomass ratio. In order to clearly illustrate the trends evident from thermodynamic modeling char circulation and nitrogen flow have been set to zero.

Figure 3: Lower heating value of product gas $(M\mathcal{J}/kmol \text{ of } C \text{ in system})$.





Figure 4: Lower heating value of product gas (MJ/kmol of gas).

Two major trends are evident in Figures 3 and 4. At low H_2O to biomass ratios not all of the carbon in the system is gasified. Increasing the H_2O to biomass ratio results in greater carbon conversion, hence greater gas yield and increased heating value of the product gas per kmol of carbon in the system. This is shown in Figure 3. The plateau in Figure 3 represents conditions where complete carbon

conversion is attained. Once the H_2O to biomass exceeds the minimum value required for complete carbon conversion dilution of the product gas occurs, reducing the heating value per kmol of gas. This is shown in Figure 4.

Increasing temperature increases the heating value at H_2O to biomass ratios below complete carbon conversion due to shifting the endothermic char gasification reactions (Equations 10-12) to the right. However above the carbon conversion boundary increased temperature decreases the heating values at H_2O to biomass ratios above complete carbon conversion due to the water-gas shift (Equation 1) promoting the formation of H_2 at the expense of CO.

It should be noted that this model assumes that only CH₄, CO, CO₂, H₂O, H₂ and solid carbon exist in greater than trace quantities at equilibrium conditions. For standard gasification operating conditions (temperatures greater than 650°C and the presence of some oxidant) this is a valid assumption. The results of Li (2001) who used 44 different species for calculating equilibrium was used to validate this assumption. Chemical equilibrium has been used widely in the literature (Schuster et al., 2001, Zainal et al., 2001, Li et al., 2001, Kinoshita et al., 1991) as a practical method for estimating product gas composition. However, equilibrium based models perform better at temperatures generally higher than typical gasification conditions. Equilibrium calculations have been shown to underestimate the methane yield (Li et al., 2001). A comparison of chemical equilibrium results with experimental results published in the literature is shown in Table 1.

Table 1 shows that steam gasification results in the literature do not approach equilibrium closely. Actual results will lie between devolatilisation compositions and equilibrium. Better mixing, higher temperature and longer residence times will drive the composition towards equilibrium. The primary discrepancy between results and equilibrium is the methane yield. Semi-empirical methods of adjusting equilibrium to allow for greater methane yield have been shown to work well (Li et al., 2001). These methods will be used once a bank of experimental data is available from the University of Canterbury gasifier.

For the purpose of assessing the feasibility of gasification-based cogeneration plants, the model is adequate as it predicts the product gas heating value, shown in Table 1, with reasonable accuracy. Chemical equilibrium calculations are also important in indicating the effects of altering different operating parameters.

Gasification Energy plant

The following is a discussion on how a FICFB gasifier may be integrated into a gasification-based energy plant.

Gasifier and gas cleaning

FICFB gasifiers require inflows of wood feed, superheated steam, air and re-circulated product gas, as shown in Figures 5 and 6. For modeling purposes an acceptable moisture content of the wood feed of 25 wt(daf)% has been used. This is based on reports by Schuster et al. (2001) from an operational FICFB plant. To dry the wood

	Equilibrium	(Hofbauer et al., 1997)	Equilibrium	(Herguido et al., 1992)
Wood Chemical Formula	CH _{1.47} O _{0.8}	CH _{1.47} O _{0.8}	CH _{1.52} O _{0.91}	CH _{1.52} O _{0.91}
Hydrogen	53.6%	31.5%	50.0%	59%
Carbon Monoxide	33.7%	22.7	36.5%	11%
Carbon Dioxide	12.3%	27.5%	12.9%	21%
Methane	0.4%	11.2%	0.6%	3%
C _x H _y	0%	4.4%	0%	2.5%
Lower Heating Value (MJ/Nm ³)	10.2	13.0	10.2	10.8

Table 1: Dry gas comparison with experiment at 750°C and biomass to steam ratio of 2 kg/kg.

feed a rotary drum dryer using either warm flue gas or heated air has been used. Re-circulated product gas is required for combustion in the CFB to provide the heat for the endothermic gasification reactions, where char circulation is insufficient. Char circulation is assumed to be 15% based upon values published from the work at Gussing (Schuster et al., 2001).

At Gussing, Austria, which is a successful demonstration biomass FICFB gasifier-gas engine cogeneration plant, the product gas is cooled to 170?C before entering the gas cleaning stage. The gas is then cleaned by a filter and an esterified rape-seed oil scrubber (Hofbauer et al., 2002). This cools the gas to 40?C, which enables it to be used in either a gas engine or a gas turbine. Details on the gas cleaning are being undertaken in other part of the larger programme.

Gas engine plant

Figure 5: Gasification-Gas Engine Plant



The process shown in Figure 5 uses a turbocharged, intercooled, spark ignition engine. They can either be operated at stoichiometric air for maximum power or at lean burn conditions which minimize NO_x emissions. For modeling purposes it is assumed that the engine is operated in lean-burn conditions, with an air to fuel ratio of 1.6 times the stoichiometric air-fuel ratio (Major, 1995). Gas engine are often de-rated, when run on fuels of lower heating value than natural gas, due to the lower thermal energy input per volume of the cylinders. However, while the FICFB gas has a lower heating value it also requires less air to combust.

On a basis of volume of lean burn air/fuel mixture the FICFB product gas has similar heating value as natural gas. Another obstacle to maintaining the efficiency of a gas engine on FICFB product gas is its knock tendency due to its low methane number (high hydrogen content).

Gas turbine combined cycle plant

Figure 6: Gasification-Combined Cycle Plant



A typical gas-fired natural gas turbine combined cycle (GTCC) unit consists of a single fuel gas turbine, unfired multi-pressure heat recovery steam generator with no bypass stack, multi-pressure condensing steam turbine, electric generators, step-up transformer, and water cooled heat rejection. A benefit of gas turbine combined cycles is when integrated into a process they can provide high-grade heat at the expense of steam power generation. Typical exhaust temperatures of gas turbines are 500-550°C (Traverso et al., 2004). However it is generally considered that gas turbines have stricter gas cleaning requirements than gas engines (Scharpf and Carrington, 2005). Adaptation of gas turbines to product gas may require modification to the combustion chamber in order to be suitable for burning lower calorific value fuel. Standard gas turbines are designed for natural gas, which has a HHV of around 39 MJ/Nm³ (Baines, 1993) compared to 10-13 MJ/Nm³ for FICFB producer gas. Rodriques et al (2003) suggests that this could add between 3 and 20%

to the capital cost of a gas turbine. These modifications are not novel, GE has developed combustion chambers specifically for lower heating value fuels from gasification and has gained 340,000 hours experience in operating these turbines (Jones and Shilling, 2003).

Heat flow from exhaust gas stream

In order to increase the energy efficiency of the process, the thermal energy possessed by the exhaust gas stream can be used in the wood processing plant. Medium density fibreboard (MDF) has been taken as an example for such an application. An energy demand model for the MDF production has been established by Li and Pang (2006) which provided estimates of the energy demands from a MDF plant, shown in Table 2.

Table 2:	Energy	Demand	of a	MDF	Plant.
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MDF Panel Output	120,000m³/yr		
Electricity	4.8 MW		
Thermal oil at 280°C	2.6 MW		
9 bar saturated steam	2.5 tonnes/hr		
4 bar saturated steam	4.6 tonnes/hr		
Flue gas at 380°C	74 tonnes/hr		

Modeling results for the integrated system

HYSYS, a heat and material balance software package, has been used to develop a computer simulation of the processes shown in Figure 5 and 6. From the HYSYS model, mass flows, energy flows and major cost driving parameters can be extracted for each equipment item. This allows estimates of plant efficiencies and capital costs. Both processes are designed so that sufficient heat is generated to meet the heat requirements presented in Table 2. The fuel is a mixture of waste residues from the MDF process and externally sourced biomass. The costs and availability of this fuel is derived from work by Li and Pang (2006) and biomass availability modeling undertaken as part of the larger programme.

Results for a system using gas engine

The model was run for the simulation of a biomass energy system using gas engines for electricity generation and the results are given in Figure 7 and Figure 9. The model shows a large reduction in electrical efficiency at low MW_{el} scales due to the need to meet the heat demands of the MDF plant. This results in the need to have a boiler in parallel with the engine decreasing electrical efficiency and increasing capital cost on a per kW_{el} basis. The 'Model - No heat plant' line shows the efficiencies of a pure gasifier-gas engine process without the need for a boiler in parallel. In Figure 7 and Figure 9 efficiencies and costs for existing plant as reported by Dornburg and Faiij (2001), Brammer and Bridgewater (2002)and Li and Pang (2005) are shown as well as efficiency and cost correlations reported by Bridgewater (1995). The largest gas engine used in the model was a three MWel engine based on the Jenbacher JMS 620. Therefore electrical outputs above this scale are obtained by using a number of engines in parallel.

Figure 7: Gasifier-Gas Engine Efficiency



The reduction in electrical efficiency from the 38-43% that gas engines are capable of on natural gas (Jenbacher, 2003) is due primarily to the gasification step. After gasification the heating value of the product gas is typically 75-88% of the heating value of the original (Higman and Burgt, 2003). Results from the energy transformations from chemical equilibrium are shown in Figure 8. This illustrates the conversion of the chemical energy in the wood into the various post-gasifier forms. It should be noted that this represents results from chemical equilibrium studies. Further energy flows can be developed for the University of Canterbury gasifier once the data is available.

Figure 8: Gasifier Energy Flows



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From Figure 9 a FICFB gasifier gas engine plant will have a capital cost in the vicinity of $33,000/kW_{el}$ to $4,000/kW_{el}$. The major expense being the gas engine, which contributes half the capital cost. These estimates are similar to those presented in the literature, as shown in Figure 9. In Figure 9 two capital costs lines are shown. The more expensive line is assuming that the reliability of the gas engine and quality of the gas cleaning is insufficient to be relied upon to meet the MDF plant heat demands at all times. This assumption leads to the process including a boiler capable of meeting the full heat demand of the MDF plant. The cheaper line represents the process, where the boiler (if needed) is not sized to meet the full heat demand of the plant.





Results for a gas turbine combined cycle

The model was also run for the simulation of a biomass energy system using gas turbine combined cycle for electricity generation applied to the same MDF plant as discussed in the above section. The simulation results are given in Figure 10 and Figure 11. Figure 10 and Figure 11 show comparisons of the modeled efficiencies and costs with the efficiencies and costs of nine BIGCC_{atm} plants reported in Dornburg and Faaij (2001) and Li and Pang (2005), a literature relationship for the efficiency and cost of









BIGCC_{atm} plants based on in-house data (Bridgwater, 1995) and natural gas combined cycles efficiencies and costs (Gas Turbine World, 2005).

The nine plants in Figure 10 show considerable variation in efficiency but compare well with the modeled results. Higher efficiencies, especially at small scale, can be expected in a pure BIGCC_{atm} plant compared to the modeled process, as the modeled process integrates the BIGCC_{atm} plant with an MDF plant. This results in heat being utilized in the process rather than generating steam for the electricity generation. The effect of this is reduced with scale as the heat demands of the process become smaller relative to the total energy input. Natural gas combined cycle efficiencies are shown to illustrate the effect of using a solid fuel and gasification compared to using a gaseous fuel. The major decrease in efficiency is due to the conversion efficiency (chemical efficiency) of the gasifier, which is typically ~80%. Hence, one would expect BIGCC efficiencies to be $\sim 80\%$ of natural gas combined cycle efficiencies. The efficiency of BIGCC plants is greater than gasifier-gas engine plants at scales above $10MW_{el}$ but the capital costs are greater than the costs of gasifier gas engine plant. BIGCC plants vary between \$4,500 and \$8,500 over a scale of seven to $55 MW_{el}$. Due to the wood fuel costs being low, the benefit of increased efficiency associated with BIGCC systems compared to gas engine systems is unlikely to compensate for the significant increase in capital costs.

Conclusions

Gasification power generation represents the next generation of higher efficiency technology for biomass electricity generation. Further development and deployment are required internationally to reduce the capital costs of these technologies, which are currently greater than conventional technologies. BIGCC technology has a very large capital cost, while gasifier gas engine plants show a moderate capital cost. The gasifier-gas engine plant efficiency and costs show this to be an appealing process for further research and development. Continued work by the BIGAS consortium in characterizing the FICFB product gas, developing gas cleaning procedures and adaptation of the prime movers to use with the product gas is planned. Once this work has been completed more detailed discussions about the integration of gasification energy plants into wood processing plants can be made.

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