

Biomass steam gasification – an extensive parametric modeling study

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Abstract

A model for steam gasification of biomass was developed by applying thermodynamic equilibrium calculations. With this model, the simulation of a decentralized combined heat and power station based on a dual fluidized-bed steam gasifier was carried out. Fuel composition (ultimate analysis and moisture content) and the operating parameters, temperature and amount of gasification agent, were varied over a wide range. Their influences on amount, composition, and heating value of product gas and process efficiencies were evaluated. It was shown that the accuracy of an equilibrium model for the gas composition is sufficient for thermodynamic considerations. Net electric efficiency of about 20% can be expected with a rather simple process. Sensitivity analysis showed that gasification temperature and fuel oxygen content were the most significant parameters determining the chemical efficiency of the gasification. © 2001 Elsevier Science Ltd. All rights reserved.

Keywords: Biomass; Gasification; Fluidized bed; Parametric study; Chemical equilibrium

1. Introduction

During recent years, the interest on biomass utilization for power generation has increased since it has significant environmental benefits. It is a renewable energy resource that provides low SO₂ and CO₂ emissions, being a positive contribution to limit the greenhouse effect. Currently 14% of the total world energy consumption is covered by biomass (Bain et al., 1998). However, there is even a much bigger potential to produce, for instance, at least 50% of Europe's total energy requirement on the basis of biomass fuels (Bridgewater, 1995). The technologies for the primary thermal conversion of biomass for electricity production are combustion, gasification, and pyrolysis. An overview of existing technologies is given by Bain et al. (1998). Gasification combined with a gas engine or gas turbine has the advantage of having a higher electric efficiency and lower electricity production costs than direct combustion (Bridgewater, 1995). Van Swaaij et al. (1994) give an overview of principles and practice of biomass gasification.

Fluidized-bed gasifiers provide excellent mixing and gas/solid contact, causing high reaction rates and con-

version efficiencies. Further, there is the possibility of addition of catalysts to the bedmaterial to influence product gas composition and reduce its tar content (US Department of Energy, 1992). Since the gasification reactions are endothermic, the process must be supplied with energy. This can be done by partial combustion of the biomass within the gasifier using an hypostoichiometric amount of air as gasification agent. Air gasification produces a poor quality gas with regard to the heating value (around 4–7 MJ/m³, higher heating value (HHV)). Gasification with pure oxygen, producing a higher quality gas (around 10–18 MJ/m³, HHV) requires, however, additional costs for oxygen production. A gas of similar quality can be produced by using a dual fluidized-bed system. Thereby, the gasifier is divided into two zones. The gasification zone is fluidized with steam (fluidization with recirculated product gas is also possible) supplied with the process energy from oxidation of part of the char in a second zone. Such a system applying an internally circulating fluidized-bed has been developed for instance by Hofbauer et al. (1995).

The design and operation of a gasifier requires understanding of the influence of fuel- and operating parameters on the performance of the plant. For this, numerous models for fluidized-bed gasifiers have been developed. Some of these are summarized by Gururajan

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Notation		π	pressure ratio (–)
a, b, c	coefficients from combustion calculations (–)	ω	linear sensitivity coefficient (–)
h^f	heat of formation (kJ/kg)	<i>Subscripts</i>	
HHV	higher heating value (kJ/kg)	F	fuel (biomass)
k_{800}	thermodynamic equilibrium constant at 800°C (–)	C	compressor
LHV	lower heating value (kJ/kg)	chem	chemical (efficiency)
\dot{m}	mass flow (kg/s)	C_{pg}	product gas compressor
P	power (kW)	DH	district heating
<i>Greeks</i>		el	electric
λ	air to fuel ratio (–)	GT	gas turbine
η	efficiency (–)	ov	overall
		pg,GT	product gas entering gas turbine
		rel	relative

et al. (1992). The models can be divided into thermodynamic equilibrium and kinetic rate models.

Rate models always contain parameters which make them hardly applicable to different plants. Thus thermodynamic equilibrium calculations, which are independent of the gasifier design, may be more convenient for process studies on the influence of the most important fuel and process parameters. However, it is known that thermodynamic equilibrium may not be achieved mainly because of the relatively low operation temperatures (product gas outlet temperatures between 750–1000°C, Bridgewater, 1995), (Buekens and Schouters, 1984). Nevertheless, models based on thermodynamic equilibrium have been used widely (Buekens and Schouters, 1984; Kinoshita et al., 1991; Watkinson et al., 1991; Denn et al., 1979; Kosky and Floess, 1980; Kovacic et al., 1989; Cousins, 1978; Shesh and Sunavala, 1990; Shand and Bridgewater, 1984; Kilpinen et al., 1991). Watkinson et al. (1991) demonstrated that discrepancies in comparison of equilibrium gas composition with experimental data may reflect unknown temperature gradients from the gasification zone to the outlet zone. Hence, there might be differences in the temperature, which determines the equilibrium composition of the product gas, and the specified gasification temperature. Adapting this value, reasonable correlation to data of semi-commercial and commercial units can be obtained. Kilpinen et al. (1991) showed that solid carbon and CH₄ content were underpredicted to some extent by an equilibrium approach. This was mainly caused by the slow kinetics of the heterogeneous gasification and the decomposition of CH₄. Reducing the input amounts of carbon and hydrogen by the measured values of CH₄ and solid carbon, fine agreement to the measurements was obtained, even for a laboratory-scale unit. Norman et al. (1997) conclude that for industrial-scale plants, only HCN and NH₃ may be kinetically controlled but not the other species.

In this work, an extensive parametric study of the influence of operating conditions and fuel parameters on the heating value of the product gas and the overall performance of a gasification plant to produce electricity and energy for district heating was performed. (Though thermodynamic equilibrium models were used in a significant number of publications, to our knowledge no such study on a power plant based on biomass steam gasification has been performed yet.) As the plant was designed for operating in a decentralized combined heat and power station (thermal power of 10 MW), its design was kept compact and simple. The concept of the gasifier itself is based on the fast internally circulating fluidized-bed (FICFB, Hofbauer et al., 1995) process.

Special attention throughout the present investigations was paid to the influence of fuel composition, since biomass and derived fuels, which may be used in such an unit for electricity production, differ widely. Since the exact composition of the product gas is of minor importance for thermodynamic considerations and the presented analysis should be independent of the detailed plant design, an equilibrium model was incorporated into a commercial, equation-oriented simulation tool (IPSEpro™) developed to simulate energy processes (Perz, 1990). Due to the object-oriented structure of IPSEpro™, modules describing the gasification process can be integrated into an existing power plant model library.

2. Process scheme and process parameters

2.1. Process scheme

The flowsheet of the process is depicted in Fig. 1. The gasifier (GAS) is split – according to FICFB process – into a gasification zone (GZ) and a combustion zone (CZ). The energy necessary for the gasification process is

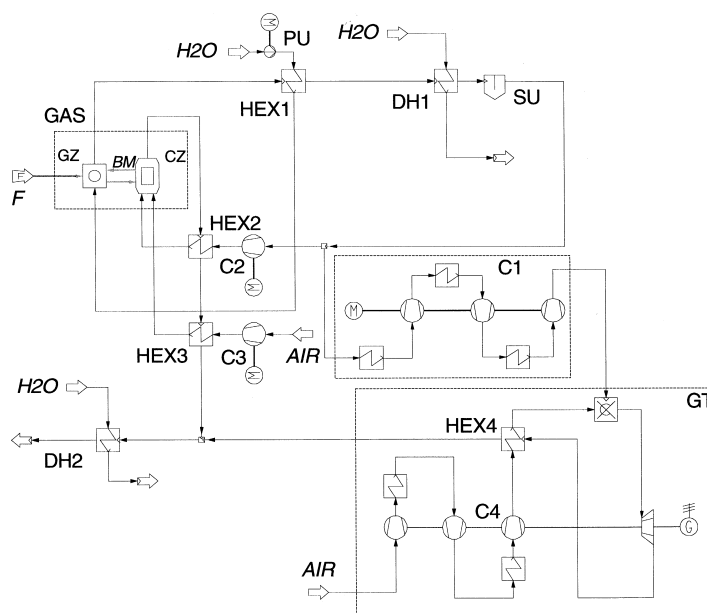


Fig. 1. Flowsheet of the process: GAS – Gasification unit; GZ – gasification zone; CZ – combustion zone; BM – bedmaterial; HEX – heat exchanger; DH – district heating; SU – separation unit; C – compressor; PU – pump; GT – gas turbine; AIR – air of combustion; H₂O – water; F – fuel (biomass).

supplied by combusting a part of the char and a part of the product gas, where heat and char are transported by the bedmaterial (BM) (quartz sand) from the gasification zone to the combustion zone. According to experience obtained in a 10 kW_{th} pilot power plant, the char leaving the gasification part was assumed to be 15 wt% of the carbon content in the biomass. As this value depends on different parameters such as temperature, residence time of biomass in gasification zone, amount of circulating bedmaterial or nature of biomass, the effect of a deviation from this value was examined and the results are presented below. The gasification zone was fluidized with steam at 400°C and the product gas after the gasifier utilized to evaporate and heat up fluidization steam (HEX 1).

After heat exchanger HEX 1, the product gas was cooled down to 100°C by transferring heat to a district heating system (DH 1) and cleaned up in a gas cleaning unit. Compressed in three stages with intercooling, it was fed into a gas turbine (GT) with recuperative air preheating (HEX 4). Combustion chamber, turbine, recuperator and air compression (three-stage with intercooling) are modeled separately. Air to fuel ratio in the gas turbine was chosen to maintain a required turbine inlet temperature of 1150°C and the turbine inlet pressure set to 12 bar absolute. In the recuperator (carried out as a countercurrent heat exchanger), a minimum temperature difference of 100°C at the hot side was assumed. Compression ratios in each step of compressing were equal, i.e. $\pi = 2.34$ for the combustion air and $\pi = 2.44$ for the product gas into gas turbine, respectively.

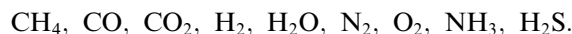
The exhaust gas from the combustion zone of the gasification unit was used for preheating recirculated product gas (up to 400°C) and afterwards for preheating combustion air (up to 300°C). Finally, it was mixed with exhaust gas from the gas turbine and the incorporated energy transferred to a district heating system (DH 2).

2.2. Model description

In the gasification zone, temperatures of product gas, outgoing bedmaterial and ash, as well as in the combustion zone, temperatures of exhaust gas and outgoing bedmaterial were assumed to be equal. Heat loss was specified as percentage of heat input into the gasifier; 2.5% in the gasification zone and the same value for the combustion zone.

The model for combustion chamber handles mass and heat balances assuming complete combustion (air to fuel ratio $\lambda = 1.1$).

For the gasification zone, an overall mass balance and one for each element are implemented. Fuel was described by its ultimate analysis; for product gas, the following components were considered:



The product gas composition was calculated considering thermodynamic equilibrium of the main components CH₄, CO, CO₂, H₂ and H₂O and the presence of solid carbon. Sulfur from biomass was converted into hydrogen sulfide and nitrogen from biomass was converted into NH₃. Due to low contents of sulfur and nitrogen in the fuel, inaccuracies of this simplification are negligible

anyway. The formation of tar was neglected in thermodynamic calculations because of its low concentrations, but it has to be considered in plant operation.

For enthalpy balance, the heat of formation of biomass is required, which is derived as shown in Eq. (1)

$$h_{\text{biomass}}^f = bh_{\text{CO}_2}^f + ch_{\text{H}_2\text{O}}^f - (\text{LHV} + ah_{\text{O}_2}^f). \quad (1)$$

Lower heat value (LHV) is calculated by the equation of Boie (Netz, 1982) which is an empirical calculation based on ultimate analysis. The equation of Boie gives suitable results for coal, lignite and biomass. Alternatively, measured values of LHV can be used in the gasifier.

The overall heat balance gives the energy demand of the gasification reaction that has to be provided by the circulating bedmaterial.

Equilibrium determination is based on the fact that Gibbs free energy (ΔG) at equilibrium state is zero. The equilibrium is calculated considering the components CH_4 , CO , CO_2 , H_2 , H_2O and solid carbon C. System analyses show that simultaneous equilibrium is described by three partial mass balances (for C, H and O) and three equations for the chemical equilibrium of three independent reactions.

In the case of excess of gasification agent, no solid carbon remains in the equilibrium state. This means that one equation describing the system becomes $C = 0$, and subsequently, only two reactions need to be considered for the equilibrium equations. In the case that solid carbon remains, the equations of the reactions R1–R3 in Table 1 are used, otherwise R4 and R5. Table 1 shows additionally, the thermodynamic equilibrium constant k_{800} at 800°C .

As described above, a certain part of the carbon content in the fuel does not participate in the equilibrium reaction. This amount of solid carbon plus the amount resulting from equilibrium calculation gives the total quantity of solid carbon leaving the gasification zone together with bedmaterial.

Table 2 gives a summary of all required settings.

Table 1
Reaction equations

		k_{800}
R1: $\text{C(s)} + \text{CO}_2 \rightleftharpoons 2\text{CO}$	Boudouard equilibrium	0.04723
R2: $\text{C(s)} + 2\text{H}_2 \rightleftharpoons \text{CH}_4$	Hydrogenating gasification	6.90803
R3: $\text{C(s)} + \text{H}_2\text{O} \rightleftharpoons \text{CO} + \text{H}_2$	Heterogenous water gas shift reaction	7.49508
R4: $\text{CH}_4 + \text{H}_2\text{O} \rightleftharpoons \text{CO} + 3\text{H}_2$	Methane decomposition	158.706
R5: $\text{CO} + \text{H}_2\text{O} \rightleftharpoons \text{CO}_2 + \text{H}_2$	Water gas shift reaction	1.08498

3. Results and discussion

3.1. Standard conditions

For the base case, beech (chips) serves as fuel for the simulation (standard water content, wet weight basis is assumed to be 25 wt%). Proximate and ultimate analyses are taken from Reisinger et al. (1996) (Table 3).

Product composition resulting from the described model and the given parameters is shown in Table 4. In order to achieve a thermal power of 10 MW, 0.81 kg/s biomass have to be fed into the gasifier. This yields 1.28 m_s^3/s product gas, from which 0.47 m_s^3 have to be recirculated into the combustion zone of the gasifier. 0.81 m_s^3 (63%) can be utilized for electric power generation in the gas turbine. Thus, a considerable amount of the produced gas has to be used to maintain the gasification process.

The values for thermal and electric power are summarized in Table 5. By cooling down the product gas and the exhaust streams, a sum of 4.43 MW thermal power can be gained for the use in district heating.

These results allow rating of the process with the efficiencies shown in Table 6, where $\dot{m}_{\text{pg,GT}}$, LHV_{pg} stand for mass flow and LHV of product gas fed into gas turbine and \dot{m}_{F} , LHV_{F} for mass flow and LHV of biomass into gasifier, $P_{\text{el,GT}}$ for net electric power produced by turbine, $\sum P_{\text{el,C}}$ for electric power consumed by all compressors, and $P_{\text{el,C,pg}}$ consumed by product gas compression.

3.2. Effect of gasification temperature

Fig. 2 shows product composition and LHV versus gasification temperature.

It can be seen that LHV of humid product gas is nearly constant (between 8300 and 8500 KJ/m_s^3); LHV for dry product gas ranges between 9900 and 10 800 KJ/m_s^3 . Above approximately 800°C , no methane is produced; some measurements given in the literature (Herguido et al., 1992; de Souza-Santos, 1999; Jennen et al., 1999; Wang and Kinoshita, 1993; Fercher et al., 1998) differ from that considerably and show methane contents up to 10 vol% (see Table 7). This indicates that equilibrium state is not reached in these references; devolatilization of biomass gives high contents in methane and higher hydrocarbons (e.g., C_xH_y : 24.8 vol% water-free basis; Zanzi et al., 1994), which do not react completely to equilibrium concentrations of CO , CO_2 and H_2 . Deviations of the other product gas components between the model and measured values are within the deviations reported from different plants.

The product of $\text{LHV} \times \text{gas yield}$ is 17 500 kJ/kg dry biomass for the equilibrium model and $15\,900 \pm 2000$ kJ/kg is given for experimental measurements by Fercher et al. (1998) and by Fleck (1997).

Table 2
Summary of all required settings

Thermal power input of the gasifier ($\dot{m}_{\text{biomass}} \times \text{LHV}_{\text{biomass}}$)		10 MW
Temperatures	Ambient (fuel, combustion air and water for steam production)	20°C
	Gasifier/gasification zone – product gas and bedmaterial out	800°C
	Gasifier – steam in	400°C
	Gasifier/combustion zone – exhaust gas and bedmaterial out	870°C
	Gasifier/combustion zone – combustion air in	300°C
	Gasifier/combustion zone – product gas in	400°C
	District heatings 1 and 2 – water in	60°C
	District heatings 1 and 2 – water out	90°C
	District heating 1 – product gas out	100°C
	District heating 2 – exhaust gas out	90°C
	Product gas compression, stages 1, 2 and 3 in	10°C above dew point
	Gas turbine–air compressor, stages 2 and 3 in	50°C
	Turbine inlet temperature	1150°C
Minimum temperature difference	Recuperator, hot side (gas turbine)	100°C
Heat loss	Gasifier	5% of thermal power input
Cooling by heat loss	Gas separation unit	2°C
Amount fluidizing agent	Gasifier/gasification zone	500 kg/h
	per mass fuel	0.17 kg/kg
Pressures	Ambient (fuel, combustion air and water for steam production)	1 bar
	Gasifier/gasification zone – product gas out	0.98 bar
	Exhaust gas into atmosphere	1.1 bar
	Turbine inlet pressure	12 bar
Pressure drops	Gasifier/gasification zone	0.15 bar
	Gasifier/combustion zone	0.15 bar
	Gas turbine/combustion chamber	0.5 bar
	Gas separation unit	0.05 bar
	All heat exchangers (hot and cold side)	0.02 bar
Efficiencies	Turbine η_s	0.8
	Compressors ^a , pump η_s	0.75
	Motors, generator η_{el}	0.98
	Turbine, compressors, pump, motors, generator η_m	0.98
Air to fuel ratio λ	Gasifier/combustion zone	1.1
Solid carbon in bedmaterial (non participating in equilibrium reaction)		15% of carbon content in biomass

^a Remark to the internal efficiency η_s of the three-stage compressors: the efficiencies of each stage are set resulting in an overall efficiency of 0.75.

Table 3
Data of fuel (beech chips)

Proximate analysis		Ultimate analysis (maf)	
LHV	12 430 kJ/kg	C	48.26 wt%
HHV	13 994 kJ/kg	H	5.82 wt%
Volatiles (mf)	80 wt%	O	45.67 wt%
Water content	25 wt%	N	0.22 wt%
Ash content (mf)	0.61 wt%	S	0.03 wt%

3.3. Effect of amount of fluidizing agent

An increase in fluidizing agent results in almost linear decreases of cycle efficiencies: efficiencies fall from 19.4% (η_{el}) and 64.6% (η_{ges}) at 0.034 kg steam/kg fuel (500 kg/h) to 14.9% (η_{el}) and 55.4% (η_{ges}) at 0.68 kg/kg (2000 kg/h).

Table 4
Composition (vol%) and lower heating value of raw and dry product gas

	Raw	Dry
CH ₄	0.09%	0.11%
CO	25.84%	31.17%
CO ₂	10.30%	12.42%
H ₂	46.57%	56.18%
H ₂ O	17.11%	
Others	0.09%	0.11%
LHV	11 781 kJ/kg	14 629 kJ/kg
	8316 kJ/m _s ³	10 032 kJ/m _s ³

The ratio of product gas used in the gas turbine to produced product gas falls from 64.4% to 56.8%. These decreases in efficiency are mainly due the fact that the enthalpy of vaporization of fluidization steam is not recovered in the considered process.

Table 5
Thermal and electric power on the basis of 10 MW gasifier

District heating 1 (DH1)	891.2 kW
District heating 2 (DH2)	3536.4 kW
Compression of product gas into gas turbine (C1)	429.5 kW
Compression of combustion air into gasifier/ combustion zone (C2)	60.9 kW
Compression of product gas into gasifier/ combustion zone (C3)	37.5 kW
Compression of water for steam production (PU)	0.003 kW
Gas turbine (GT), net power	2366.7 kW

Table 6
Efficiencies of the process

Chemical efficiency	$\eta_{\text{chem}} = \frac{\dot{m}_{\text{PG,GT}} \times \text{LHV}_{\text{PG}}}{\dot{m}_{\text{F}} \times \text{LHV}_{\text{F}}}$	66.9%
Gas turbine efficiency (considering compression of product gas)	$\eta_{\text{GT}} = \frac{P_{\text{el,GT}} - P_{\text{el,C,PG}}}{\dot{m}_{\text{PG,GT}} \times \text{LHV}_{\text{PG}}}$	28.9%
Overall net electric efficiency	$\eta_{\text{el}} = \frac{P_{\text{el,GT}} - \sum P_{\text{el,C}}}{\dot{m}_{\text{F}} \times \text{LHV}_{\text{F}}}$	18.4%
Overall efficiency	$\eta_{\text{ov}} = \frac{P_{\text{el,GT}} + P_{\text{DH1}} + P_{\text{DH2}} - \sum P_{\text{el,C}}}{\dot{m}_{\text{F}} \times \text{LHV}_{\text{F}}}$	62.7%

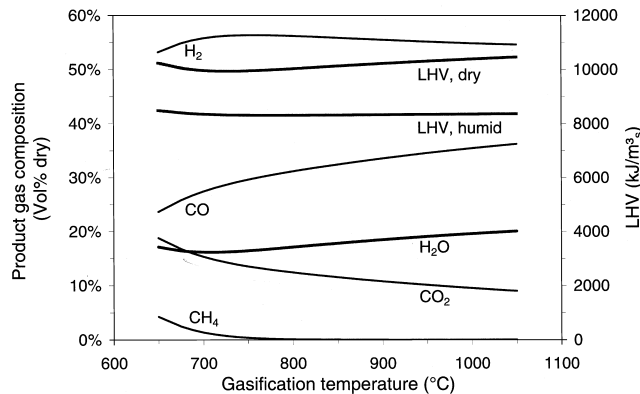


Fig. 2. Product gas composition and lower heating value versus gasification temperature.

Table 7
Product gas composition of equilibrium model compared to measurements given in the literature

	Results from equilibrium model (%)	Fercher et al. (1998) (%)	Herguido et al. (1992) (%)
H ₂	56.2	34.2	52
CO	31.2	27.2	23
CO ₂	12.5	22.7	18
CH ₄	0.1	11.1	7
Remaining C _x H _y	–	4.8	–

3.4. Effect of water content

The effect of biomass water content on product gas composition, on LHV (of humid product gas) and on chemical efficiency at constant amount of fluidizing agent is shown in Fig. 3.

Water in biomass shifts the equilibrium of product gas towards H₂ and CO₂; CO and CH₄ decrease. LHV of humid product gas declines mainly because of the rising H₂O content in the product gas. Chemical efficiency goes down due to low LHV firstly, and secondly, because biomass moisture has to be evaporated in the gasifier resulting in a higher energy demand, so that more product gas has to be recirculated into the combustion zone of the gasifier. At a biomass water content of about 66%, the whole product gas has to be used to maintain the gasification process leading to a chemical efficiency of zero, which is, of course, only a theoretical extrapolation.

3.5. Effect of amount of char leaving the gasification zone

As mentioned above, a certain amount of biomass leaves the gasification zone without taking part in the chemical equilibrium. According to measurements, this value is assumed to be 15 wt% of the carbon content in the biomass. For lower values, more product gas has to be recirculated to provide the energy for the gasification reactions. If no biomass is transported into the combustion part, a chemical efficiency of 65.6% is obtained leading to an overall electric efficiency of 18.1%. In this case, 47.8% of produced gas is recirculated into the combustion zone of the gasifier. If the amount of carbon carried into the combustion part can be increased, the demand on recirculated product gas decreases. With 42.6 wt% of the carbon leaving the gasification zone as char, no product gas has to be recirculated. This leads to an increase in chemical efficiency to 70.9% and in overall electric efficiency to 19.0%. This increase is due to the

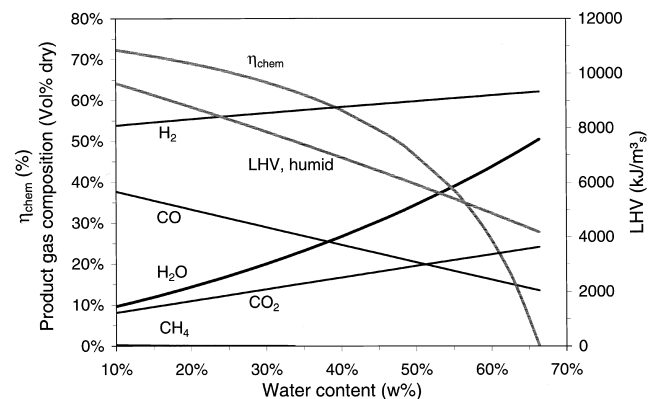


Fig. 3. Product gas composition, lower heating value and chemical efficiency versus biomass water content (wet weight basis).

fact, that the product recirculation is thermodynamically disadvantageous.

3.6. Effect of fuel composition

The influence of different fuel compositions on LHV of humid product gas (Fig. 3) and on chemical efficiency (Figs. 4 and 5) was examined by varying oxygen content (in wt%, moisture- and ash-free – maf) on one hand and the ratio of carbon content (in wt%, maf) to hydrogen

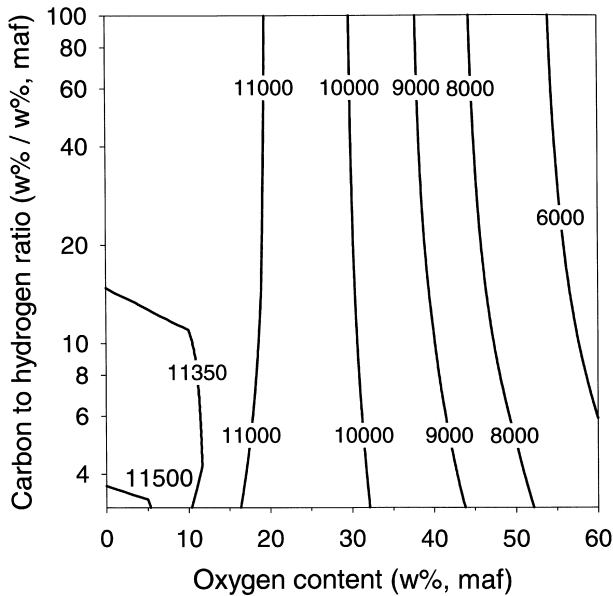


Fig. 4. Lower heating value (humid product gas) of product gas versus carbon to hydrogen ratio and oxygen content.

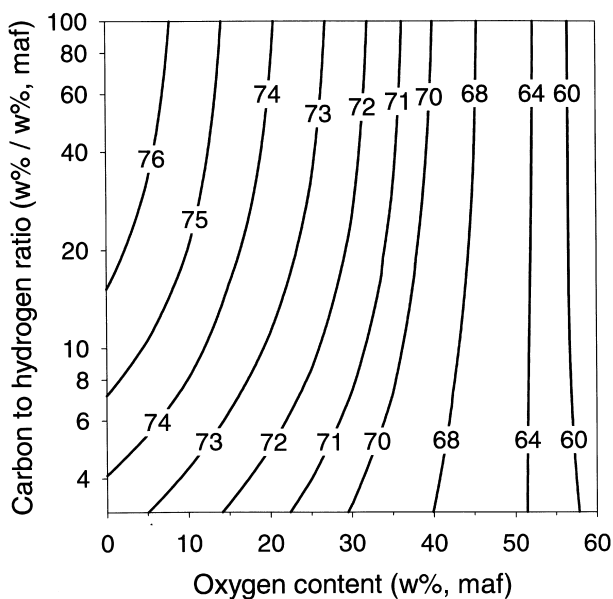


Fig. 5. Chemical efficiency versus carbon to hydrogen ratio and oxygen content.

content (in wt%, maf) on the other hand (C/H ratio). This variation was carried out over a wide range of fuel compositions: oxygen contents from 0 to 60 wt%, and C/H ratio from 3 (corresponds to pure methane) to 100 (approximate composition of coke) giving different fuels with a strongly varying demand in gasification agent. In order to get comparable results, the ratio of carbon content in biomass to water into gasifier considering water content of biomass and fluidization steam (C/H₂O ratio) was kept constant in this variation (0.856 kg/kg; this value is taken from calculation with parameters shown in Tables 2 and 3). One has to consider that the gasifier thereby has not a constant geometry.

LHV for humid product gas declines naturally with increasing oxygen content, but seems to be almost independent of C/H ratio at low oxygen contents in fuel. A more detailed analysis of the results shows that due to the constant C/H₂O ratio, water content in product gas varies to a comparatively small extent and that the sum of H₂ and CO content is nearly constant (93.5–95 vol%). As CH₄ content is low and has negligible contribution to LHV and as LHV of H₂ (10 783 kJ/m³) and CO (12 633 kJ/m³) are quite similar, the resulting LHV of humid product gas is nearly constant for a given oxygen content (between 11 000 and 11 500 KJ/m³ for oxygen contents up to 20 wt%).

Higher C/H ratios result in higher chemical efficiencies. This can be explained by the fact that at higher C/H ratios, less product gas has to be recirculated into the gasifier. Instead of this, more carbon of the biomass is transported directly into the combustion zone, which is thermodynamically beneficial.

3.7. Linear sensitivity coefficient

In order to compare the influence of the various parameters on chemical efficiency η_{chem} (refer to Table 6) of the gasifier, a sensitivity analysis was carried out by calculating at the base case, the relative linear sensitivity coefficient $\omega_{i,\text{rel}}$ of η_{chem} towards the parameter x_i by the finite difference method (Rabitz et al., 1983).

$$\omega_{i,\text{rel}} = \frac{\frac{\Delta \eta_{\text{chem}}}{\eta_{\text{chem}}}}{\frac{\Delta x_i}{x_i}} \quad (2)$$

In Fig. 6, the values of $\omega_{i,\text{rel}}$ for the examined parameters are depicted. Gasification temperature has the strongest influence on chemical efficiency ($\omega = -1.10$), since a higher amount of product gas has to be recirculated to maintain the increased gasification temperature. Nevertheless, in plant operation, a minimum temperature has to be kept as tar emissions increase at lower temperatures (e.g., Fercher et al., 1998). The influences of amount of carbon that is transported into the combustion zone and the ratio of carbon to hydrogen (wt%/wt%) in biomass are quite low.

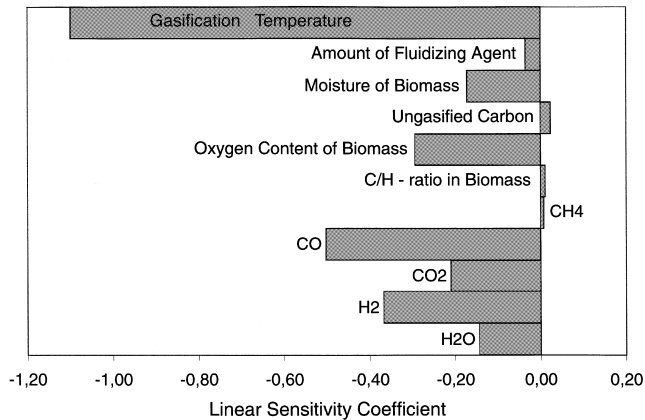


Fig. 6. Linear sensitivity coefficient of chemical efficiency towards the examined parameters.

The linear sensitivity coefficient of product gas components is estimated in such a way that one component content is varied while the value of the other component is fixed. The remaining remaining components were calculated to fulfill the mass balances. Among all possible variations, the ones giving maximum derivation for each component are taken.

The maximum value of $\omega_{i,rel}$ becomes -0.5 for increasing the component CO, keeping the content of CO₂ constant. Due to the carbon balance, the amount of CH₄ in product gas decreases in that case and the total lower heating value increases. Since the inaccuracies are more significant for the prediction of CH₄, which does not significantly influence chemical efficiency, this analysis also confirms the application of an equilibrium model for thermodynamic considerations.

4. Conclusions

A simulation model was developed for a gasification unit according to a dual fluidized-bed gasification process, that handles mass and heat balances and calculates thermodynamic equilibrium as well. This model was integrated into a process scheme with district heating and electric power generation by using a gas turbine. The effects of varying biomass moisture, amount of fluidizing agent, gasification temperature and biomass composition were shown. Among these parameters, gasification temperature had the strongest influence on chemical efficiency (linear sensitivity coefficient $\omega = -1.10$, refer to Eq. 2). As the tar content in the product gas increases with decreasing temperature, the choice of the operating temperature is restricted. Thus for plant operation, the oxygen content of biomass, which has also considerable influence on the gasifier's chemical efficiency ($\omega = -0.3$), is of greater interest.

The model offered the opportunity to evaluate different gasification processes as well as variations of the fuel.

The results of the equilibrium model for the gasifier (LHV \times gas yield) were in the range of measured results, though CH₄ content in product gas was overestimated. It was shown that the discrepancies in the prediction of the gas composition did not significantly influence the overall efficiency. So an overview over the wide range of application of steam biomass gasification and the expected LHV of product gas, the chemical efficiency of the gasifier η_{chem} and total process efficiencies could be given.

In order to keep investment costs low and to achieve simple operation, the design was chosen to be compact and use only a few units, and the plant should be combined with district heating to improve economics. A net electric efficiency of about 20% can be expected with this process.

References

- Bain, R.L., Overend, R.P., Craig, K.R., 1998. Biomass-fired power generation. *Fuel Process. Technol.* 54, 1–16.
- Bridgewater, A.V., 1995. The technical and economic feasibility of biomass gasification for power generation. *Fuel* 74 (5), 631–653.
- Buekens, A.G., Schouters, J.G., 1984. Mathematical modelling in gasification. In: Bridgewater, A.V. (Ed.), *Thermochemical Processing of Biomass*. Butterworths, London, UK, pp. 177–199.
- Cousins, W.J., 1978. A theoretical study of wood gasification processes. *N. Z. J. Sci.* 21, 175–183.
- de Souza-Santos, M.L., 1999. A feasibility study of an alternative power generation system based on biomass gasification/gas turbine concept. *Fuel* 78, 529–538.
- Denn, M.M., Yu, W.-C., Wei, J., 1979. Parameter sensitivity and kinetics-free modeling of moving bed coal gasifier. *Ind. Eng. Chem. Fundam.* 18, 286–288.
- Fercher, E., Hofbauer, H., Fleck, T., Rauch, T., Veronik, G., 1998. Two years experience with the FICFB-gasification process. In: *Proceedings of the 10th European conference and technology exhibition, Würzburg, June 1998*.
- Fleck, T., 1997. *Gaserzeugung aus Biomasse in schnell intern zirkulierender Wirbelschicht Integration des FICB-Prozesses*. Ph.D. thesis, Vienna University of Technology, Vienna, Austria.
- Gururajan, V.S., Agarwal, P.K., Agnew, J.B., 1992. Mathematical modelling of fluidized bed coal gasifiers. *Trans. Inst. Chem. Eng.* 70, 211–238.
- Herguido, J., Corella, J., Gonzalez-Saiz, J., 1992. Steam gasification of lignocellulosic residues in a fluidized bed at a small pilot scale effect of the type of feedstock. *Ind. Eng. Chem. Res.* 31, 1274–1282.
- Hofbauer, H., Stoiber, H., Veronik, G., 1995. Gasification of organic material in a novel fluidization bed system. In: *Proceedings of the first SCEJ Symposium on fluidization, Tokyo*, pp. 291–299.
- Jennen, T., Hiller, R., Köneke, D., Weinspach, P.-M., 1999. Modellierung der Vergasung von Holz in einer zirkulierenden Wirbelschicht. *Chem.-Ing.-Tech.* 71, 459–464.
- Kilpinen, P., Hupa, M., Leppälähti, J., 1991. Nitrogen chemistry at gasification – a thermodynamic analysis. Report 91-14, Combustion Chemistry Research Group, Abo Academi University, Finland.
- Kinoshita, C.M., Wang, Y., Takahashi, P.K., 1991. Chemical equilibrium computations for gasification of biomass to produce methanol. *Energy Sources* 13, 361–368.
- Kosky, P.G., Floess, J.K., 1980. Global model of coal gasifiers. *Ind. Eng. Chem. Proc. Des. Dev.* 19, 586–592.

- Kovacik, G., Oguztoreli, M., Chambers, A., Ozum, B., 1989. Equilibrium calculations in coal gasification. *Int. J. Hydrogen Energy* 15, 125–131.
- Netz, H., 1982. *Verbrennung und Gasgewinnung bei Festbrennstoffen*, first edition. Technischer Verlag Resch KG, Gräfelfing.
- Norman, J., Pourkashanian, M., Williams, A., 1997. Modelling the formation and emission of environmentally unfriendly coal species in some gasification processes. *Fuel* 76 (13), 1201–1216.
- Perz, E., (1990). A computer method for thermal power cycle calculation. ASME-Paper IGTI GT-351.
- Rabitz, H., Kramer, M., Dacol, D., 1983. Sensitivity analysis in chemical kinetics. *Ann. Rev. Phys. Chem* 34, 419–461.
- Reisinger, K., Herger, M., Hofbauer, H., Haslinger, C., 1996. BIOBIB – a database for biofuels. In: *Proceedings of THERMIE- conference: renewable energy databases*, Harwell, UK, 1996.
- Shand, R.N., Bridgwater, A.V., 1984. Fuel gas from biomass status and new modelling approaches. In: Bridgwater, A.V. (Ed.), *Thermochemical Processing of Biomass*. Butterworths, London, UK, pp. 229–255.
- Shesh, K.K., Sunavala, P.D., 1990. Thermodynamics of pressurized air–steam gasification of biomass. *Indian J. Technol.* 28, 133–138.
- US Department of Energy, 1992. *Electricity from biomass – a development strategy*. Report No. DOE/CH10093-152.
- Van Swaaij, W.P.M., van den Aarsen, F.G., Bridgwater, A.V., Heesink, A.B.M., 1994. A review of biomass gasification. Report to the Commission of the European Communities, DG XII.
- Wang, Y., Kinoshita, C.M., 1993. Kinetic model of biomass gasification. *Sol. Energy* 51 (1), 19–25.
- Watkinson, A.P., Lucas, J.P., Lim, C.J., 1991. A prediction of performance of commercial coal gasifiers. *Fuel* 70, 519–527.
- Zanzi, R., Sjöström, K., Björnbom, E., 1994. Rapid pyrolysis of wood with application to gasification. In: Bridgwater, A.V. (Ed.), *Advances in Thermochemical Biomass Conversion*. Blackie, London, UK, pp. 977–985.