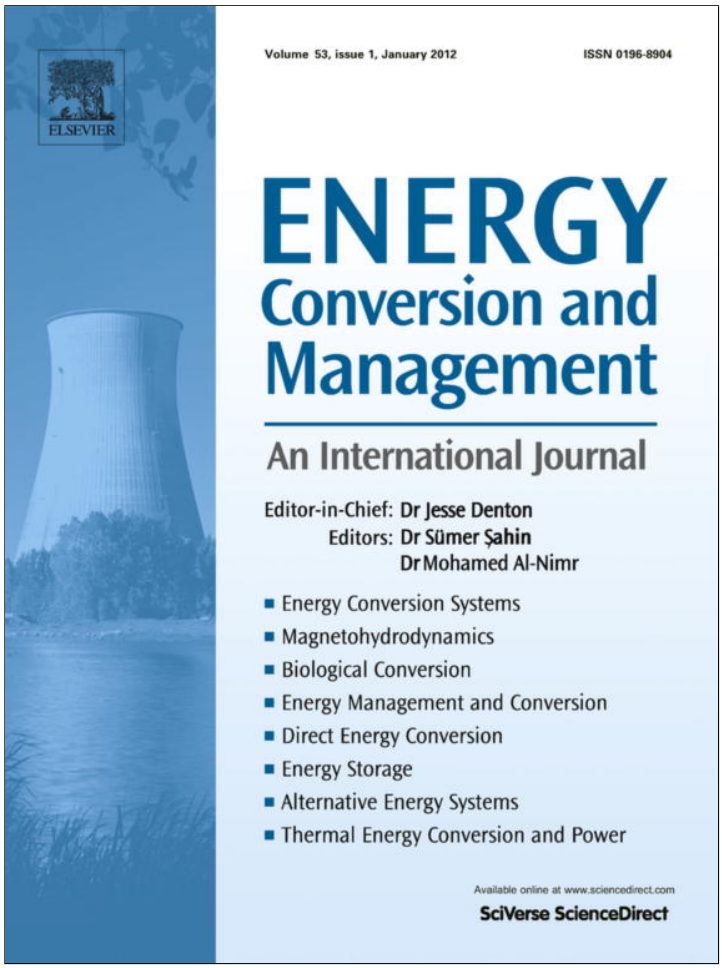

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Performance of entrained flow and fluidised bed biomass gasifiers on different scales

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ABSTRACT

This biomass gasification process study compares the energetic and economic efficiencies of a dual fluidised bed and an oxygen-blown entrained flow gasifier from 10 MW_{th} to 500 MW_{th}. While fluidised bed gasification became the most applied technology for biomass in small and medium scale facilities, entrained flow gasification technology is still used exclusively for industrial scale coal gasification. Therefore, it is analysed whether and for which capacity the entrained flow technology is an energetically and economically efficient option for the thermo-chemical conversion of biomass. Special attention is given to the pre-conditioning methods for biomass to enable the application in an entrained flow gasifier. Process chains are selected for the two gasifier types and subsequently transformed to simulation models.

The simulation results show that the performance of both gasifier types is similar for the production of a pressurised product gas (2.5 MPa). The cold gas efficiency of the fluidised bed is 76–79% and about 0.5–2 percentage points higher than for the entrained flow reactor. The net efficiencies of both technologies are similar and between 64% and 71% depending on scale. The auxiliary power consumption of the entrained flow reactor is caused mainly by the air separation unit, the oxygen compression, and the fuel pulverisation, whereas the fluidised bed requires additional power mainly for gas compression.

The costs for the product gas are determined as between €4.2 cent/kWh (500 MW_{th}) and €7.4 cent/kWh (10 MW_{th}) in the economic analysis of both technologies.

The study indicates that the entrained flow reactor is competitive technology for biomass gasification also on a smaller scale.

1. Introduction

Today, it is increasingly realised and accepted that the well-being of our society is closely bound to the future energy supply. The increasing demand for safe, secure, sustainable but still affordable energy will make this issue a challenge over the next decades. With the intention of the European Union to supply 20% of its overall energy demand from renewable sources by 2020 [1], biomass is a very promising resource regarding the seasonal and weather-limited fluctuations of wind and solar power.

Amongst other technologies, biomass gasification is increasingly being considered for future power generation from renewable energies. In contrast to wind or solar power, biomass applications can deliver reliable energy on demand because biomass can be stored to balance seasonal fluctuations. Furthermore, the material utilisation of biomass is feasible by gasification. Biomass can be converted to fuels (SNG, FT or methanol) or synthetic products (plastics, ammonia). However, today biomass is used rather in small- and medium-scale applications. Regarding biomass gasification, the preferred technologies are fluidised and fixed bed gasifiers [2], developed and already commercially operated by several companies [3]. In contrast, biomass entrained flow gasifiers have only been applied in the research stage. The technology is however widely used for industrial-scale coal gasification (IGCC and chemical synthesis applications with several 100 MW_{th}). Entrained flow gasification is used there because of the higher availability, the higher throughput and the better product gas quality. Furthermore, the co-gasification of biomass in large scale integrated gasification combined cycles (IGCC) has already been tested [4,5].

This leads to the question of whether and for which size entrained flow gasification could be an alternative for the full scale biomass application. Technical, energetic and economic issues of biomass entrained flow gasification are discussed in this study.

Regarding the scale this paper analyses gasification systems with a thermal input of 10, 50, 100 and 500 MW_{th} respectively, which is still relatively small compared to coal-fired state-of-the-art facilities. In general, scale effects in biomass systems are significant.

The economy of scale is used in almost all technologies. Larger system sizes reduce the specific investment and operation costs and usually improve the process efficiency due to lower specific heat losses and higher component efficiencies. Regarding biomass

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Nomenclature

C	investment costs
$\dot{m}$	mass flow rate
m_N^3	cubic metre at standard conditions (273.15 K, 0.1 MPa)
p	pressure
P	power
q	heat
Q	thermal fuel input
T	temperature

Greek letters

η	efficiency
λ	air stoichiometry

Subscripts

aux	auxiliary
CGE	cold gas efficiency
cold	lower temperature at the outer reactor surface
EF	entrained flow reactor
FB	fluidised bed reactor
loss	losses
Syngas	synthesis gas

Acronyms

ASU	air separation unit
CGE	cold gas efficiency
DME	dimethyl ether
FT	Fischer–Tropsch synthesis
HTC	hydrothermal carbonisation
IGCC	integrated gasification combined cycle
LHV	lower heating value
MeOH	methanol
PSA	pressure swing adsorption
REA	restricted equilibrium approach
SNG	synthetic natural gas

systems the specific transport effort is increased if larger plants are considered. The increased costs and energy demand for transportation can hamper the realisation of large scale biomass applications. Therefore, the optimum size of a biomass plant is not only set by the positive effects of the economy of scale but also by biomass specific issues (transportation, sustainability, social impacts, etc.).

Due to the inherent limitations of biomass-based systems caused by low energy density, challenging storability and regional distribution the selected scales seem to be appropriate. However, compared to combustion-based systems, gasification technologies show a lower dependency on biomass transportation costs, distribution density and fuel cost [6]. Gasification systems have the potential to achieve a higher efficiency and therefore consume less biomass for a given power output.

Different process simulations of fluidised bed biomass gasification are available in the literature (e.g. [7–10]) and a life cycle assessment of an integrated biomass gasification combined cycle is available [11].

The cold gas efficiency (CGE) is simulated to be between 66% and 81% depending on the simulation parameters. For instance, Pröll and Hofbauer [8] present a detailed simulation of the DBCFB plant in Güssing, Austria and report a net efficiency of gas generation (clean product gas that is fed to the gas engine) based on the LHV of 71.5% for a thermal fuel input of 7.4 MW_{th}.

Only a few process simulations are known analysing entrained flow gasification of biomass or co-gasification (e.g. [12–14]). The CGE is 77–82% depending on the entrained flow gasifier type.

Only a few studies directly compare entrained flow and fluidised bed gasification technologies. These studies focus on an overall process evaluation including downstream units. Meijden et al. [15] compare process efficiencies from biomass to SNG using three different gasification technologies (entrained flow, circulating fluidised bed, and allothermal fluidised bed). The CGE of the atmospheric allothermal gasifier (81.1%) is slightly higher compared to the entrained flow gasifier (77.4%). This is due to the assumption of higher carbon conversion in the allothermal gasifier and identical heat losses for both technologies. Both assumptions are questionable as the main advantages of entrained flow gasifiers are the high fuel conversion and small heat losses due to their compact design.

The production of FT fuel based on fluidised bed and entrained flow biomass gasification was evaluated in a recent techno-economic analysis [16]. Although the study is focused on an economic evaluation on a large scale (389 MW_{th}), technical aspects of entrained flow gasification (pulverisation, feeding, reactivity, ash behaviour) are not considered. The biomass to fuel efficiency (biomass to FT on a LHV basis) for the entrained flow gasifier (50%) is significantly higher compared to the fluidised bed gasifier (39%).

Marechal et al. use a thermo-economic model to analyse the production of SNG [17] and liquid fuels [18] from lignocellulosic biomass. Entrained flow and fluidised bed gasification are evaluated for liquid fuel (FT, DME, MeOH) generation and the gasification technology is identified to be the most critical choice defining the performance of the overall system. For the production of liquid fuel, the best configuration includes indirectly heated circulating fluidised bed gasification.

As the focus of these studies is not on the gasifier, the direct comparison of the gasification processes and the consideration of technical issues are not discussed. No study is known to the authors that directly compares fluidised bed and entrained flow gasification technologies and assesses the influence of scale and of specific technical issues (e.g. fuel reactivity, slagging requirements, and operation temperature) on these processes. This study accounts for the influence of conversion reactivity in different gasifier technologies and considers the slagging requirements of large scale entrained flow gasifiers.

Both gasification technologies for biomass are simulated and both technologies are compared directly. An allothermal fluidised bed and an oxygen blown entrained flow reactor are modelled. We selected these technologies because we consider these prior art and expect a wider application of both technologies in the future. In order to enable a wide range of utilisations (e.g. gas turbine, chemical synthesis) in a large scale, this study aims at a high quality, almost nitrogen free product gas at a pressure of 2.5 MPa.

2. Gasification technologies

A classification of gasification technologies can be made by the type of the reactor (fixed bed, fluidised bed and entrained flow), the energy supply (allothermal or autothermal), the gasification agent (air, oxygen, steam or carbon dioxide), as well as by the working pressure in the reactor (pressurised or atmospheric). The characteristic feature is the reactor type with most influence on the product gas composition and efficiency.

In the following sections, fluidised bed and entrained flow gasification are introduced briefly, as these technologies are most appropriate for large scale industrial applications.

The challenges and as yet unsolved obstacles are presented and discussed regarding prior art technology from the literature as well as from reported data published on pilot and commercial plants. This extended literature overview is required to evaluate the simulation parameters in Section 3.

2.1. Fluidised bed gasification

2.1.1. Application of fluidised bed gasification

Fluidised bed gasification is a well-known technology in small- to medium-scale (500 kW to 50 MW thermal biomass input) biomass applications. However, fluidised beds have found only limited application in hard coal gasification because of their temperature limitation due to the bed material agglomeration and the resulting low carbon conversion rate of coal. Furthermore, lignite is a possible feedstock for fluidised bed gasification and gasifiers on a larger scale were installed in Germany [19] and are discussed in Australia [20].

Especially the elements Ca, K and Na in the fuel – respectively in the ash – influence the agglomeration behaviour. K reduces and Ca increases the softening temperature of the ash. Therefore a high K content can cause deposit formation and bed sintering. Na and K have a high affinity to Cl and SO_4 and are therefore found in the ash particles mainly as sulphates, carbonates or chlorides. Sulphates are formed especially at fuel rich (air ratio <1) conditions. In excess of alkalis, agglomerations with silicate are likely [21].

If biomass is co-fired with sulphur containing fuels (coal), Ca and K reduce primarily the sulphur emissions, and elementary chlorine is available [22]. This probably leads to increased high temperature corrosion.

Current research and development activities favour a fluidised bed reactor for biomass gasification. There are however some different fluidised bed concepts which differ significantly in the process engineering as well as in important parameters such as product gas composition. A review of prior art technologies was published recently [23]. In larger scale biomass gasification facilities circulating beds are used. Gasifiers up to a thermal input of 60 MW_{th} are realised [3].

A well known facility is the so called Güssing gasifier in Austria [24]. This plant was chosen as the reference plant for the simulations. The plant concept is realised so far in Güssing, Oberwart, Villach and Ulm with a thermal input of approximately 10 MW_{th} [25].

The gasifier consists of two fluidised beds with circulating bed material for combustion and gasification respectively. Besides the Güssing plant, there are more projects with already operating plants like the Milena gasifier in the Netherlands, the Rentech-SilvaGas gasification plant in the US, the Foster-Wheeler gasifier in Lathi or the Heatpipe Reformer in Germany [3,26]. The technologies and technical specifications of these gasifiers vary, but they are all examples for the wider application of fluidised bed reactors for biomass gasification.

All of the different technologies and facilities have some inherent challenges to solve, depending on the design. But in general, the main issues arise repeatedly.

2.1.2. Important system and operation parameters

An important issue allowing high cold gas efficiency is the complete carbon conversion within the process. In a stationary fluidised bed there is a wide range of residence times for individual particles [27]. Partly reacted particles are removed from the hot zone which leads to a decrease of carbon conversion that is on average only 90% in circulating fluidised bed (CFB) reactors [28]. The best of existing fluidised bed processes have a carbon conversion of 97% [27]. However, it is expected that the overall carbon conversion can be further increased by a combination with a fluidised bed combustor.

Another main barrier for biomass gasification is the formation of organic impurities (tars) in the gasifier. The tars may form coke in the filters, blocking them, or condense in cold spots (below about 300 °C) causing operational interruptions. Together with their carcinogenic character and their destructive power for engines and turbines, the tar problem has to be considered closely [29].

In comparison to other gasification designs, fluidised bed gasifiers are known to have relatively high tar contents of 2–10 g/m^3 [30].

The literature shows different approaches of how to separate or convert the tar downstream of the gasifier or catalytically remove the tar in situ, both resulting in additional costs. At first glance, additional equipment downstream of the gasifier seems more expensive and is less favoured [29,31]. Catalysts suitable for the tar conversion are dolomite, iron-based or nickel and other supported catalysts and carbon-supported catalysts [32,33].

In order to minimise the reactor size or maximise the capacity for a given size, the fluidised bed can be pressurised. Operation pressures up to nearly 2.0 MPa (e.g. Värnamo) have been proved in one-stage circulating fluidised bed reactors [34]. The pressurised gasification enables and favours elaborate downstream processes.

However, the big challenge to solve for pressurised operation is the feeding system and thus the input of biomass into the gasifier. It is known from literature and own experimental observations that the feeding of biomass under pressure causes serious problems and a need for elaborate and costly process design. Furthermore, the inert gas requirement for purging increases, which may lead to a dilution of the product gas [35]. Today only few fluidised bed gasifiers are operated under moderate pressure which can lead to the conclusion that pressurised operation still creates more problems than the advantage of a pressurised product gas can compensate [3]. However, a pressurisation up to about 0.5 MPa seems to be feasible without major difficulties in the middle future and can enhance the potential for downstream processes.

The combustion fluidised bed is operated with air as the reaction medium. The operation with pure oxygen is not considered. Due to the two stage process the two fluidised beds can be operated in different gas atmospheres and a dilution of the product gas with nitrogen does not occur. The operation with air avoids the installation of an air separation unit that would have a significant energy consumption. Furthermore, the operation with air leads to a higher gas flow rate which facilitates the fluidisation and improves the temperature distribution in the fluidised bed.

2.2. Entrained flow gasification

2.2.1. Application of entrained flow gasifiers

In general, entrained flow gasification is a well-researched and developed technology in coal gasification. According to the US Department of Energy database on gasification [36], in the years from 2005 to 2011, 94% of all registered gasification projects worldwide that are in operation or in the planning stage are based on entrained flow gasification. These are 82 entrained flow gasification projects out of 87 total gasification projects. As almost all gasifiers are designed for coal and pet coke as a feedstock, the adaption of entrained flow gasification to biomass is still under development.

The most challenging issues to solve for entrained flow gasification of biomass in contrast to fluidised bed gasification are fuel pre-treatment, oxygen supply, and ash behaviour.

2.2.2. Important system and operation parameters

In entrained flow gasification the residence time of fuel particles in the hot reaction zone is short, typically below 10 s. In order to achieve high conversion the solid feedstock is grinded to a small particle size; in the case of coal below 200 μm. The fuel particles are either pneumatically transported to the pressurised reactor or pumped as a slurry. Due to the thermal energy required for the heat up and evaporation of the slurry, dry-fed gasification systems have the potential for higher efficiency [27]. For dry-fed coal-fired entrained flow reactors the pulverised coal is fluidised in a non-bubbling mode using an inert gas (N_2 or CO_2). Alternatively, piston feeder systems in combination with screws are a very attractive solution due to the lower energy consumption for the compression of inert gas and only little dilution of the synthesis gas [12]. However, there is no experience of such feedings systems on a larger scale.

Svoboda et al. [37] recently compared advantages and disadvantages of different fuel pre-treatment strategies for entrained flow gasification. They reviewed drying, torrefaction, flash pyrolysis and dissolution of wood in organic solvents. Furthermore, the hydrothermal carbonisation (HTC) is a method to disintegrate the biomass structure and improve the grindability.

For the first generation of biomass entrained flow gasifiers, torrefaction and HTC are expected to be the best fuel pre-treatment technologies. The intermediate products are storable and smooth dry feeding can be achieved. The wet process HTC is especially suitable for biomass with high moisture content. Torrefaction is suitable for woody biomass as it disintegrates the wood structure at low temperature. As larger scale biomass processes are likely to be based on wood, torrefaction seems to be the best pre-treatment technology for the entrained flow gasification of biomass. As the low temperature heat demand for the torrefaction can be taken from the gasification process, a large scale torrefaction process next to the gasifier offers the potential for a high overall efficiency. Simultaneously, due to the feasibility of storage and transport of torrefaction and HTC products, distributed small scale processes are possible, and the fuel supply chain for the gasification process is more flexible. The commercial application of both technologies cannot be considered as state of the art. However, the torrefaction process seems to be less complex and a faster commercial implementation should be possible.

Torrefaction occurs at relatively low temperatures (200–300 °C), atmospheric pressure and in the absence of oxygen [37–39]. The fuel properties are improved: the carbon content and the LHV increase, whereas the total mass decreases. For torrefied wood mass yield is typically 70–90% [37] and energy yields of the solid product are between 83% and 97% (LHV_{daf}) [38].

After the torrefaction, the fuel is ground to a particle size suitable for entrained flow gasification. By torrefaction, fuel fibres are shortened and the particles become more spherical, which improves the fluidisation behaviour in the dense flow feeding system. Fluidisation experiments using willow torrefied at 270 °C proved that such a powder with a size range of approx. 30–400 μm can be fluidised smoothly [38].

The reactivity of biomass is generally higher than the reactivity of coal. Therefore, it is expected that at given reaction conditions, biomass particles can be larger compared to coal particles to achieve complete conversion. There are some experimental studies [40–43] on the influence of biomass particle size at reaction conditions similar to entrained flow gasification. In general, higher conversions are achieved for a very small particle size, but sufficient conversions are achieved with a particle size of 0.5 mm [42].

The experimental studies and requirements for a pneumatic feeding system indicate that biomass particles of approximately <0.5 mm can be accepted as fuel in entrained flow gasification systems. The exact allowable particle size depends on biomass reactivity, reactions conditions and particle fluidisation properties.

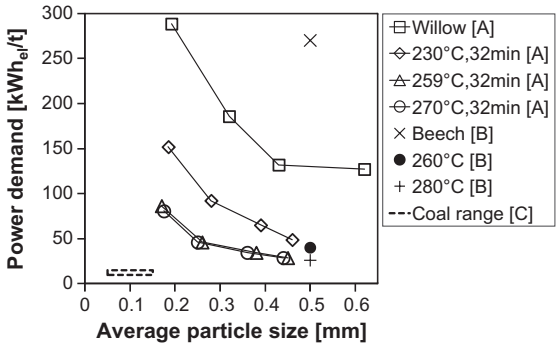

Fig. 1. Specific power demand for the pulverisation of solid fuels. (A) Bergmann et al. [38] (torrefaction at different temperature and residence time); (B) Govin et al. [44] (torrefaction at different temperature and 20 min residence time); (C) suppliers information by Loesche, Mr. Thomas Leppak (average values for bituminous coal and lignite).

The power demand for pulverisation to the desired particle size depends heavily on biomass species and pre-treatment process. Fig. 1 summarises different data for the energy demand of fuel pulverisation [38,44]. The energy requirements reported for the grinding of lignocellulosic biomass (20–150 kWh/t) are much higher than those reported for coal (7–36 kWh/t) [45].

After torrefaction, the energy demand decreases by a factor of 4–7. Torrefied wood can be pulverised down to 250–400 μm with an electrical energy demand of 25–45 kWh/t (see Fig. 1). This particle size is expected to be suitable for fluidisation, pneumatic dense flow feeding, and complete conversion in an entrained flow gasifier.

Entrained flow coal gasifiers are operated with pure oxygen that is produced in an air separation unit (ASU). The operation with air is possible but not applied due to the dilution of the synthesis gas with nitrogen. The additional electricity demand for the ASU is expected to be disadvantageous for entrained flow gasification compared to fluidised bed gasification.

Oxygen can be produced by cryogenic air separation (generally used for applications >1000 m^3/h) or pressure swing adsorption (PSA) at a smaller scale, besides electrolysis and high temperature air separation by ceramic ion transfer as niche technologies [17].

There are standardised cryogenic air separation units available for medium oxygen demands. The energy consumption is dependent on the ASU scale. According to manufacturers' data [46] the power demand for a 1000 m^3/h cryogenic air separation plant is 0.35 kWh per m^3_N oxygen. On a larger scale (12,000 m^3/h) the power demand is reduced to 0.29 kWh/m^3_N. In both processes the purity of the gaseous oxygen is 99.5% and oxygen is supplied at a slight overpressure of 0.05 MPa.

In entrained flow gasification, two types of operation modes are distinguished: slagging and non-slagging. In a slagging gasifier the operation temperature is above the ash melting temperature and molten ash flows down the gasifier wall. The slag layer partially solidifies and prevents the wall material from further corrosion by the slag. For a stable slag flow at the gasifier wall, the slag mass flow should be at least 6% of the fuel flow [12]. In a non-slagging reactor the wall is kept free from slag. The gasifier is operated below the ash melting temperature and is suitable for low-ash fuels.

The stable long-term operation of a non-slagging reactor seems to be very difficult [12] and depends heavily on the combination of specific fuel properties and the reactor design of the gasifier. In contrast, the stable operation in entrained flow coal gasification is achieved at a reaction temperature above the ash melting temperature. Hence, all known coal IGCC plants are operated in a

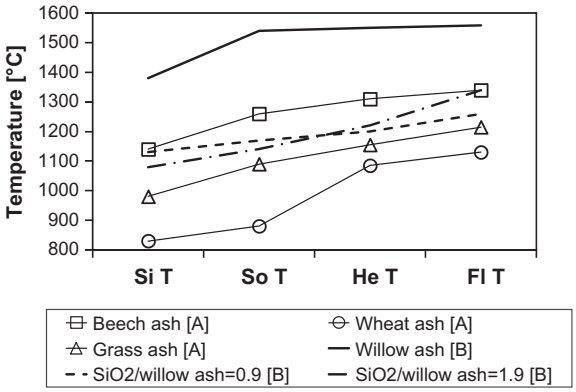

Fig. 2. Sintering temperature (Si *T*), softening temperature (So *T*), hemi-spherical temperature (He *T*), and flow temperature (Fl *T*) of different biomass fuel ashes. Measurements according to DIN 51730. Silica/ash mixtures in [kg/kg dry]. (A) Obernberger et al. [47]; (B) Drift et al. [12].

slagging mode. However, if the biomass feed contains a low amount of ash with a high softening temperature compared to coal, a dry operation mode could be preferred. Characteristic ash softening and melting temperatures for willow and straw ash are summarised in Fig. 2 [12,47]. Soft biomass like straw and grass has a low melting temperature and is therefore preferable for slagging gasifiers. As woody biomass (willow/beech) can have a high melting temperature, the operation of a non-slagging gasifier might be feasible. By blending high melting ash with flux material or using low melting biomass ash, slagging operation is possible in the temperature range of 1200 °C to 1400 °C. Slag recycling might be needed to achieve a completely covered gasifier wall. Below 1200 °C some fuels might be suitable for conversion in a non-slagging gasifier.

3. Process simulations

The process models are developed using the ASPEN PLUS V7.1 process simulator using default convergence settings and the available material databases.

For the evaluation of size effects both base cases are run for a thermal input of 10 MW_{th}, 50 MW_{th}, and 100 MW_{th}, respectively. The entrained flow gasifier is further scaled up to 500 MW_{th}. The up-scaling of a fluidised bed reactor to 500 MW would be a very difficult task and is not believed to happen in the middle future. For throughputs >100 MW_{th} an installation of several reactors in parallel is suggested. Furthermore, a large scale biomass facility (500 MW_{th}) seems not to be possible without a pre-treatment technology to increase energy density and to reduce the biomass transportation effort. As a pre-treatment technology is not considered in the fluidised bed process, the fluidised bed gasifier is simulated only up to 100 MW_{th}. The feeding system of the fluidised bed gasifier does not require a biomass pre-treatment process. The implementation of such a process would not be advantageous, but the efficiency of the fluidised bed gasifier would be significantly decreased.

The product gas stream (synthesis gas) is set to 200 °C and 2.5 MPa to enable further use in a combined-cycle or chemical synthesis plant. The temperature of 200 °C is selected to enable a direct feeding to a gas turbine as the fuel gas inlet temperature is usually below 250 °C. The pressure of 2.5 MPa is above the usual combustion pressure of gas turbines and a direct feeding including pressure losses should be feasible. Furthermore, the pressure is in the range of pressures used in chemical syntheses. The gasifiers run at operating pressures that are available (2.8 MPa, entrained flow) or thought to be commercially available in the near future (0.6 MPa, fluidised bed). Power that is required within the processes (e.g. compressors, pumps, pulverisation and oxygen production) is produced on site using the product gas from the gasifiers. The efficiency of power generation is dependent on the plant size.

3.1. Simulation of fluidised bed gasification

The allothermal fluidised bed gasifier is designed with circulating bed material in two separated beds. The combustion bed provides the heat supply for the bed material. With this configuration a dilution of the product gas with nitrogen by air is prevented, which guarantees a high caloric synthesis gas. However, the second fluidised bed increases the complexity of the system. The simulation is modelled on the basis of a reference plant located in Güssing, Austria where data are available in the literature [8,24].

Besides the gasifier itself, the simulation concept contains a flue gas cooling section, a subsequent product gas conditioning unit and a process steam production system. A pre-conditioning of biomass is not considered as wood chips are the feed material for both gasifier types and wood chips can be directly fed to a fluidised bed reactor. Therefore, efficiency gains are not expected by using a pre-conditioning process for fluidised bed gasification.

Fig. 3 shows the process configuration. The central gasifier is modelled as an equilibrium reactor (RGibbs reactor) where the product gas composition is adapted via a restricted equilibrium approach (REA). Biomass feed specified by moisture content first enters a decomposer block (RYield reactor). In parallel to the main gasifier, an external methanation reactor (RYield reactor) is used for additional methane generation. Thus, the gas composition modelled can be adjusted to real gas data from the reference gasifier. In addition, the tar problem is also solved by an external tar reactor which produces tar or naphthalene in the same amount literature suggests from the carbon feedstock. 90% of the carbon is converted in the gasifier which leaves 10% for char combustion with preheated air and some additional biomass in the second fluidised bed (RGibbs reactor). The assumed carbon conversion is in agreement with the range of carbon conversions measured and simulated in the gasification zone of a two stage fluidised bed gasifier [48,49]. The gas outlet temperature of the gasifier is 850 °C. For heat generation the mixture (unconverted char and biomass) is burned under excess air conditions ($\lambda = 1.2$). Both reactors are operated at a pressure of 0.6 MPa. The thermal input is controlled by the biomass feed rate. Heat losses for both fluidised beds are considered.

The flue gas exits the fluidised bed combustor at a temperature of 1000 °C. Stepped cooling in heat exchangers provides the required process heat for air preheating, steam production and synthesis gas preheating. The temperature of the flue gas never falls below 180 °C.

The gas clean-up of the product gas is carried out in a typical cold gas cleaning section. Raw synthesis gas is cooled down to

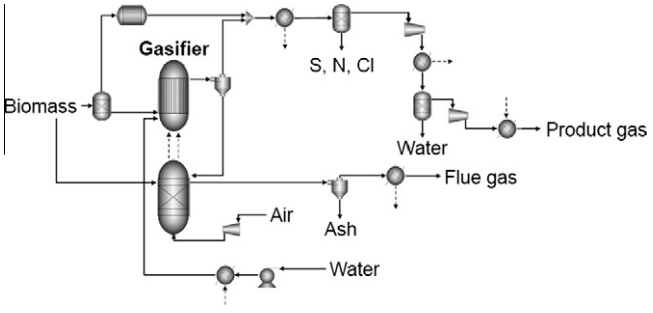

Fig. 3. Process simulation scheme of fluidised bed gasification (simplified).

30 °C for the separation of tar and subsequent cleaning of miscellaneous contamination. After the removal of particular matter the gas temperature is decreased in a heat recovery steam generator. Depending on the detailed process configuration a direct injection of steam or water may also be required. The detailed technical specification of the gas cleaning unit strongly depends on the specific fuel and is beyond the scope of this work. The cleaning steps are modelled as simple separators. Water is condensed and separated. The clean gas is then compressed in a two-stage process with intercoolers, remaining water is removed, and the gas is finally slightly preheated with heat from the flue gas cooling unit to the required condition of 200 °C at a pressure of 2.5 MPa. The final heat-up to 200 °C is chosen to enable product gas conditions that are identical to the entrained flow gasifier. The heat-up does not decrease the efficiency of the process since the utilisation of process heat is not implemented and excess heat is available at the gasifier site.

3.2. Simulation of entrained flow gasification

The process simulation of the entrained flow gasification plant contains a torrefaction process, a fuel pulverisation system, and a steam generation unit, as well as the gasifier itself and the subsequent product gas conditioning unit. Fig. 4 shows a schematic overview of the process simulation.

To model torrefaction the wet biomass is heated to 260 °C with subsequent reduction of the water content to 3 wt.%. Heat is provided from the product gas cooling section. The modelling of the reaction mechanisms occurs in a decomposition reactor (RYield reactor). Reactions are not modelled in detail but biomass is converted to reference state, whereas the energy required for breaking the molecular bonds is fed to the gasifier. The energetic loss of biomass in the torrefaction is set to 5% based on the LHV which is in accordance with the literature [38]. The torrefaction process is considered as a requirement for entrained flow gasification and an energy penalty is accounted for.

Torrefied biomass is fed to the pulverisation unit. To reach a particle size of <0.5 mm, the power demand required for the grinding of the torrefied residue is defined as 36 kWh/t.

Due to the drying of biomass in the torrefaction process, an absolute dry fuel is fed to the gasifier. The oxygen content of the dry biomass is not sufficient to convert all carbon to the gas phase (CO and CO_2). Furthermore, the flame temperature of a dry fuel is very high which could have a negative impact on the burner. Therefore, the addition of steam to the gasifier is required to homogenise the temperature distribution in the reactor and to provide enough gasification agent. The required steam is internally generated by a stepped use of process heat and final steam parameters are a temperature of 450 °C at a pressure of 3.5 MPa.

The gasifier is modelled as an equilibrium reactor (RGibbs). The product gas quality is adjusted by the restricted equilibrium approach which leads to concentrations close to that reported for the reference entrained flow gasifier [50]. The gasifier outlet temperature is set to 1350 °C to operate the reactor in a slagging mode. It is not expected that the moderate temperatures in the torrefaction process changes the ash properties.

The supply of oxygen for the gasification is modelled by a black-box-model. The power demand for oxygen production is adjusted to the gasifier scale by a linear interpolation of the manufacturer's data [46]. Slag is removed from the gasifier by a particle separator and the loss of energy for melting ash is considered by an enthalpy stream. Heat losses through the reactor walls are considered depending on reactor size.

The hot product gas from the gasifier is cooled down in steps to the required end temperature of 200 °C. After the removal of particular matter cold gas is used to quench the temperature to 800 °C. At this temperature the installation of a relatively inexpensive heat recovery steam generator is possible. The gas is cleaned of contaminants such as sulphur, nitrogen and chloride components in a gas cleaning unit modelled by separators. This simple simulation is used to approach a hot gas cleaning unit operating around 450 °C. The specific technical specification of the gas cleaning unit is strongly influenced by the trace element concentration of the fuel and is beyond the scope of this work. After the gas cleaning unit a second heat recovery steam generator is installed that cools the gas to a temperature of 200 °C. A warm gas recirculator is used to recycle gas to the gas quench. Alternatively to the hot gas cleaning section also a conventional cold gas cleaning unit is possible. This would slightly increase the auxiliary power demand because of the possible power demand of scrubbers.

By cooling down the product gas, heat is integrated in the steam generation unit and the fuel pre-treatment section.

In both gasification processes (fluidised bed and entrained flow) the heat production is much higher than the heat requirement. Furthermore, the utilisation of heat for external purposes (e.g. power generation) is not considered. Therefore, an optimisation of the heat management is not considered.

3.3. Process parameter definition and input specification

Some general parameters, compositions and starting points have to be defined prior to the actual process simulation. Default state for all educts is ambient condition (15 °C and 0.1 MPa). Biomass composition is taken from the ECN database [51] and average values for untreated woody biomass are selected. Table 1 shows the elementary composition of biomass used for the simulation.

Biomass consists of 82.1 wt.% volatiles and 16.8 wt.% bonded carbon with a higher heating value of 19.745 MJ/kg. The water content is 20 wt.%.

Power generation for the auxiliary demand is considered by different power machines depending on scale. For the 10 MW_{th} gasifiers a gas engine is assumed with an electrical efficiency of 40%. At 500 MW_{th}, power is produced by a gas turbine combined cycle with an efficiency of 55%. The electrical efficiencies for the different scales are shown in Table 2.

Literature data on heat losses are only available for larger scale entrained flow reactors. For a thermal input of ca. 500 MW_{th} heat

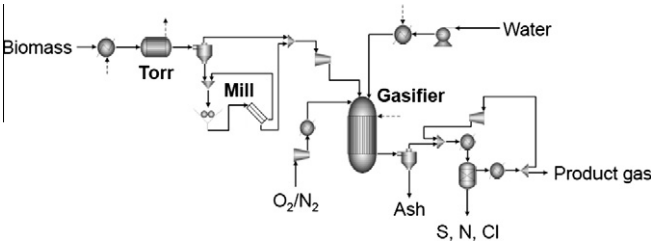

Fig. 4. Process simulation scheme of entrained flow gasification (simplified).

Table 1
Composition of biomass (untreated wood) and proximate analysis [51].

Ultimate analysis	wt.% (dry)	Proximate analysis	wt.% (dry)
C	49.5	Volatile matter	82.1
H	6.0	Fixed carbon	16.8
O	43.0	Ash content	1.1
N	0.31		
S	0.05		
Cl	0.04		
Ash	1.1		

Table 2
The key process parameters for the simulation of entrained flow and fluidised bed gasifiers.

	Entrained flow gasifier				Fluidised bed gasifier		
	10 MW	50 MW	100 MW	500 MW	10 MW	50 MW	100 MW
Gasification temperature (°C)	1350				850		
Gasifier pressure (MPa)	2.8				0.6		
Torrefaction temperature (°C)	260				n.a.		
Air pre-heater (°C)	n.a.				650		
Steam temperature (°C)	450				450		
Carbon conversion gasification	99%				90%		
Carbon conversion combustion	n.a.				99%		
Grinding energy ($kWh_{el}/t_{Biomass}$)	36				n.a.		
ASU energy demand (kWh_{el}/m^3_N)	0.35	0.34	0.32	0.29	n.a.		
Heat loss ($MW/MW_{Biomass}$)	3%	2%	1.5%	1%	5.2%	3.4%	2.6%
Pressure drop gasifier (MPa)	0.2				0.05		
Pressure drop heat exchanger (MPa/MPa_{gas})	2%				2%		
Compressor eff. isentropic	0.75	0.8	0.85	0.88	0.75	0.8	0.85
Compressor eff. mechanical	0.95				0.95		
Power machine efficiency	0.4	0.45	0.5	0.55	0.4	0.45	0.5
Ash melting enthalpy (kJ/kg)	1000				n.a.		
Raw gas tar content (g/m^3_N)	n.a.				1.5		

losses through the reactor wall range from 0.3% to 1.4% of the thermal fuel input [52] depending on the entrained flow gasifier type (membrane wall, refractory lined). Therefore, a generic gasifier with a heat loss of 1.0% is assumed for the 500 MW_{th} entrained flow technology. Since data for smaller scale systems are not available, heat losses are assumed to be 1.5% (100 MW_{th}), 2.0% (50 MW_{th}) and 3.0% (10 MW_{th}), respectively.

The heat losses of fluidised bed gasifiers are estimated from a heat transfer calculation. It is assumed that heat losses are directly proportional to the outer surface area of the reactor vessel and the temperature gradient across the reactor wall. The surface area of the pressure vessel is described as a function of reactor volume and reaction pressure. When a cylindrical shape of the reactor is assumed, heat losses can be calculated as follows

$$q_{loss,FB} = q_{loss,EF} \cdot \left(\frac{p_{EF}}{p_{FB}}\right)^{2/3} \cdot \frac{T_{FB} - T_{cold}}{T_{EF} - T_{cold}} \quad (1)$$

where $q_{loss,FB}$ and $q_{loss,EF}$ are the heat losses of the fluidised bed and entrained flow reactor at the same scale, p_{EF} and p_{FB} are the operating pressures of the reactors, T_{FB} and T_{EF} are the operating temperatures and T_{cold} is the temperature of the outer reactor surface. T_{cold} is set to 50 °C.

Other heat losses (from auxiliary process units, piping, etc.) are not considered. The elevated temperature of the product gas after the heat recovery can be considered as heat loss. However, this is considered in the calculation of process efficiency.

Although both gasifier types are modelled by an equilibrium approach, biomass reactivity in both technologies is considered. The small particle size of torrefied biomass in the entrained flow reactor enables an almost complete conversion in the short residence and at high temperature. The lower operation temperature and larger particle size in the fluidised bed reactor result in a slightly reduced fuel conversion.

Table 2 summarises heat losses and gives an overview of other important process parameters for the entrained flow gasifier at 10, 50, 100 and 500 MW_{th}, and the fluidised bed gasifier at 10, 50, and 100 MW_{th}.

4. Simulation results

4.1. Gasifier performance

The process performance is evaluated using the cold gas efficiency η_{CGE} and the net efficiency η_{net} defined as follows

$$\eta_{CGE} = \frac{\dot{m}_{Syngas} \cdot \mathrm{LHV}_{Syngas}}{\dot{m}_{Biomass} \cdot \mathrm{LHV}_{Biomass}} \quad (2)$$

$$\eta_{net} = \frac{\dot{m}_{Syngas} \cdot \mathrm{LHV}_{Syngas} - \frac{P_{aux}}{\eta_{aux}}}{\dot{m}_{Biomass} \cdot \mathrm{LHV}_{Biomass}} \quad (3)$$

where P_{aux} is the auxiliary power consumption of each process (including all process units and consumers) and η_{aux} is the electrical efficiency of the on-site power machine. As the absolute heat release in each process configuration (fluidised bed gasifier and entrained flow gasifier) is higher and at a higher temperature level than the heat requirements, the consumption of additional power for heat generation is not required. In each process several different heat integration configurations are feasible that eliminate the need for an external heat source or an electrical heater.

The cold gas and net efficiencies of both gasification technologies at different scales are summarised in Fig. 5.

Despite the technical differences and the different process parameters, the performances of entrained flow and fluidised bed gasifiers are similar. The CGE of a fluidised bed gasifier is higher because of its lower operating temperature and is 0.4–1.9 percentage points above the entrained flow gasifier. The main reason for the increased difference on a larger scale is the significant reduction of heat losses of the fluidised bed at a larger system size. Regarding the net efficiency, the entrained flow gasifier has a slightly better

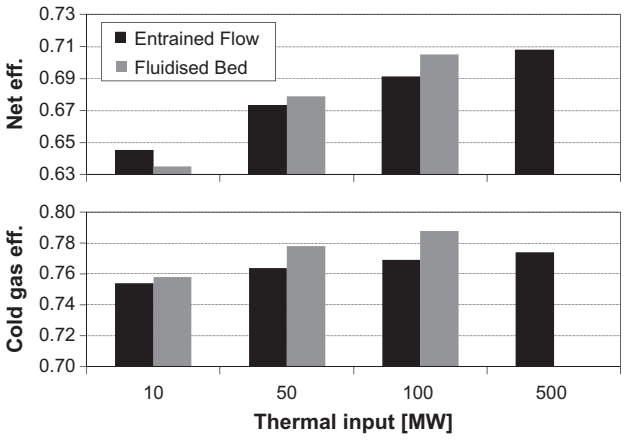

Fig. 5. Calculated cold gas and net efficiencies for entrained flow and fluidised bed gasification on different scale.

performance at a thermal input of 10 MW_{th}. However, on a larger scale the fluidised bed reactor is slightly more efficient. The effect is caused by the difference of power consumption of auxiliary systems.

The slightly lower performance of entrained flow gasifiers can be balanced by their ability to operate on a larger scale. If the entrained flow process is operated at 100 MW_{th} or 500 MW_{th}, its efficiency (CGE and net) is higher than a fluidised bed that is operated at 10 MW_{th}. The downscaling of the entrained flow technology still offers an efficiency in the range of fluidised bed gasifiers.

The total auxiliary power demand of both gasifier types is similar. The power consumer of each process for a thermal biomass input of 10 MW_{th} and 50 MW_{th} are shown in Fig. 6. The product gas compression and the air blower for the fluidised bed combustor operated at 0.6 MPa are the main power consumers of the fluidised bed gasifier. The entrained flow process requires its electrical load mainly for the ASU (oxygen supply), the oxygen compression, and the fuel pulverisation (grinding). The total power demand of the fluidised bed gasifier is slightly higher due to the large requirements for air compression (combustion fluidised bed) and product gas compression. If a product gas pressure of 2.5 MPa is needed, the production and compression of oxygen that is required for the entrained flow gasifier is less costly in terms of energy consumption. The power consumption for grinding can be compensated which makes the entrained flow gasifier advantageous if a high pressure product gas is required. If a low pressure gas or an atmospheric pressure gas is produced, the fluidised bed gasifier is expected to be advantageous. The power demand for air compression would be significantly reduced and product gas compression would no longer be necessary. An entrained flow gasifier operated at atmospheric pressure would still require an air separation unit, but could reduce its demand for oxygen compression and slightly for biomass pulverisation. At lower pressure a higher volumetric gas concentration in the dense flow feeding system could be accepted which would reduce the particle size requirements and the energy demand for grinding. However, the power demand of an oxygen blown entrained flow is expected to be higher than the power demand of a fluidised bed gasifier if both gasifiers are operated at low pressure or at atmospheric pressure.

Due to the different operating conditions the product gas compositions of the two technologies are different. The concentrations of the main gas components (H_2, CO, CO_2, CH_4, and N_2) are shown in Table 3. The H_2 concentration of the fluidised bed is higher due to the higher steam content in the gasification zone. The lower gasification temperature of the fluidised bed reactor results in a higher CH_4 content, whereas methane is not expected to be produced during entrained flow gasification.

4.2. Sensitivity analysis

In order to evaluate the influence of input parameters on the simulation results, the sensitivity of the simulation parameters is analysed. The analysis is based on a process scale of 50 MW_{th} because this is a medium process size for both gasification technologies considered here. The conclusions from a sensitivity analysis of the gasifiers on other scales are identical and the sensitivity analyses on other scales are not shown here.

Parameters that significantly influence the performance of the fluidised bed reactor are the carbon conversion, the tar content of the product gas, and the air-to-fuel stoichiometry (lambda) in the combustion bed. The influence of these parameters on the net efficiency is shown in Fig. 7. The carbon conversion is a very important process parameter. Only a high carbon conversion in the fluidised bed combustion chamber enables a high net efficiency. The stoichiometry of the combustion process (lambda) is also a very important process parameter. An increase of the air-to-fuel ratio will increase the power demand for air compression, and will also raise the heat losses due to a higher mass flow rate of the flue gas. Both effects significantly reduce the net efficiency. The CGE that is not shown here is less affected because an additional specific power consumption that is caused downstream of the gasifier does not influence the CGE. The tar content in the gasifier product gas has only a minor influence on the process performance. Due to the condensation and removal of tars in the process configuration, the net efficiency and CGE decrease. Even if the tar content is as high as 8 g/m^3, the net efficiency is still 66.0%. The influence of other parameters is also evaluated, but significance in the range of carbon conversion and combustion stoichiometry is not found.

In case of the entrained flow reactor, very important parameters are the gasification temperature, the energy demand for the ASU, the efficiency of the biomass pre-treatment process, and the energy demand for fuel pulverisation. Their quantitative impact on the process performance is shown in Fig. 8. The most important process parameters are the gasification temperature and the pre-

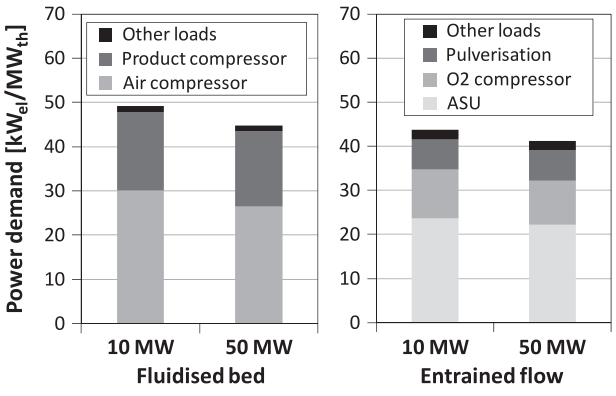

Fig. 6. Auxiliary power demand (specific power demand in kW_{el} per MW of thermal biomass input) for the fluidised bed and the entrained flow gasifier on small and medium scale (10 and 50 MW_{th}). PG: product gas; ASU: air separation unit.

Table 3
Product gas composition (mol%) of the fluidised bed and the entrained flow gasifier for a thermal input of 10 MW.

mol%	H_2	CO	CO_2	CH_4	N_2
Fluidised bed	43.2	21.3	20.5	10.5	4.6
Entrained flow	33.7	45.1	15.4	0.2	5.6

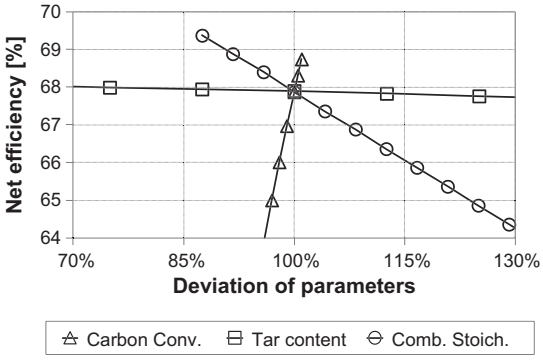

Fig. 7. Influence of carbon conversion, tar content and stoichiometry of the fluidised bed combustor on the net efficiency of the fluidised bed gasifier (50 MW_{th}).

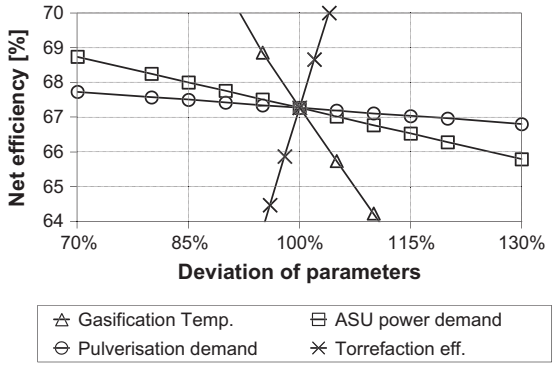

Fig. 8. Influence of gasification temperature, torrefaction efficiency, ASU specific power demand, and power demand for fuel pulverisation on the net efficiency of the entrained flow gasifier (50 MW_{th}).

treatment efficiency. A higher reaction temperature requires an increased oxygen demand that causes an additional power demand for the ASU and the oxygen compression. Furthermore, the calorific value of the product gas is reduced. If the gasification temperature is changed by 100 °C, this results in a variation of the net efficiency of the process of 2.5 percentage points. The torrefaction efficiency in the simulation is set to 95% on a LHV basis which is in the range reported in literature [38]. If this efficiency is varied by 1 percentage point, the net efficiency of the total gasification process is changed by 0.73 percentage points. This significant impact shows that losses upstream of the gasifier result in the largest decrease of process efficiency.

A variation of the specific electricity demand for the pulverisation of the torrefied biomass has a smaller impact on the process performance. An increase by 30% (from 36 kWh/t to 47 kWh/t) reduces the net efficiency only by 0.5 percentage points. The sensitivity analysis shows that a trade-off in the biomass pre-treatment has to be found. If the pre-treatment severity increases, the specific power demand for pulverisation is reduced. However, more intensive operating conditions during torrefaction are also expected to reduce the energy efficiency of the pre-treatment process. Therefore, torrefaction conditions have to be found that cause only small energy losses of the biomass, but significantly improve the pulverisation properties. If the torrefaction efficiency was improved by 1 percentage point in the 50 MW_{th} process, the power demand for pulverisation would be allowed to be increased by 17 kWh/t in order to achieve a constant net process efficiency. The exclusion of a pre-treatment process (100% energy efficiency) for the 50 MW_{th} gasifier would allow a power demand of 117 kWh/t that can hardly be achieved for untreated woody biomass (see Fig. 1). The application of torrefaction is beneficial for the performance of the whole gasification process, but both energy yield and pulverisation properties have to be considered for the optimisation of torrefaction process parameters. If the influence of torrefaction severity on energy yield and pulverisation properties is known for a specific fuel, the process parameters of the torrefaction unit can be chosen to optimise the total gasifier efficiency.

The power demand for the ASU also has an influence, however the impact is smaller. Even, if the power demand is increased by 30% (0.44 kWh/m^3), the net efficiency is still 65.8%.

4.3. Comparison of the gasification technologies

The gasification efficiencies of entrained flow and fluidised bed reactors are similar. Both gasifier types enable a CGE in the range 75–79% depending on the scale of the processes. Due to the higher operation temperature (higher oxygen consumption) the CGE of the entrained flow reactor is slightly lower.

The net efficiency is 63–71%, but there is no large difference of the auxiliary power consumption of both technologies. The entrained flow reactor requires electrical power mainly for inlet gas pressurisation, fuel pulverisation, and oxygen production; whereas the fluidised bed gasifier needs auxiliary power mainly for air compression to 0.6 MPa in the combustion bed and product gas compression to 2.5 MPa. A lower outlet gas pressure will of course favour the fluidised bed gasifier. However, a product gas at high pressure is required for a gas turbine application or a chemical synthesis on a larger scale as considered here. If the entrained flow technology is applied on a scale of 500 MW, a similar net efficiency as for a 100 MW fluidised bed gasifier would be expected. The net efficiency in Fig. 5 suggests that fluidised bed gasifiers are more suitable for scaling up as their performance significantly increases. This is mainly due to the reduction in heat losses on a larger scale. The simulation results suggest the potential of a scale up which is in accordance with commissioned plants [3]. The net efficiency of the entrained flow gasifier is less dependent on scale (see Fig. 5). It is improved from 64.5% at 10 MW_{th} to 70.8% at 500 MW_{th}. This range is smaller than for the fluidised bed gasifier because input parameters (e.g. heat losses, energy demand for grinding and oxygen supply) are less dependent on scale. The high performance of large scale systems can be transferred to a smaller scale with fewer losses compared with the fluidised bed gasifier. Therefore, the simulation results suggest the high potential of the entrained flow technology also on a smaller scale.

A direct comparison with literature data is not possible as in all references the gasification process is integrated with downstream process units and/or electricity production is included in the performance evaluation. Cold gas efficiencies reported in the literature are 66–81% (fluidised bed gasification; [7–10,15]) and 70–82% (entrained flow gasifier; [12–15]), whereas the process simulations of entrained flow reactors are typically run at a larger scale. The results from this study compare both gasification technologies at different scales and the process efficiencies are within the expected range. The detailed simulation enables the quantification of the scale effect and other important operating parameters.

The process evaluation shows that a general decision about the preferable gasification technology cannot be made based solely on the CGE and the net efficiency. The additional power consumption of the entrained flow gasifier for fuel pulverisation and oxygen supply is comparable to the power demand for downstream gas compression of a fluidised bed gasifier. Due to the required compressed air mass flow for the combustion bed and the product gas compression, the auxiliary power demand of a fluidised bed gasifier is significant if a pressurised (2.5 MPa) gas is produced. In comparable simulations that are not shown here, the operation of a 10 MW_{th} fluidised bed gasifier at atmospheric and high pressure, respectively, is considered. The simulations show only a small effect of the gasifier pressure on the efficiency if a final product gas pressure of 2.5 MPa is approached. The pressurised gasifier is slightly favoured; however, operation at atmospheric pressure is also possible without a major decrease of performance. The net efficiency is only reduced by 0.4 percentage points for an atmospheric gasifier compared to the gasifier operated at 0.6 MPa. But the atmospheric operation is expected to result in higher investment costs and higher heat losses due to the larger equipment design size. Therefore, the pressurised fluidised bed gasifier is considered in this study.

The notable energy demand for air compression gives rise to the question if the operation of the combustion fluidised bed with oxygen could be an alternative. This would require the installation of an air separation unit which would also consume power. The combustion with oxygen would make the operation of the fluidised bed

more difficult because the gas flow rate would be reduced and the combustion reaction would be more intensive (hot spots are likely). Moreover, the supply of oxygen would enable the operation of a one stage fluidised bed gasifier. However, these concepts are beyond the scope of this paper but could be considered in future studies.

5. Economic evaluation and results

The economic evaluation of the syngas production in fluidised bed and entrained flow gasifiers using the full costing method is based on the German guideline VDI 2067 [53]. The different types of costs are divided in costs based on capital, the consumption costs, the operating costs and the other costs. Taxes are not considered.

As information about investment costs is hardly reported in the literature, the estimation of investment costs for the gasifiers is difficult. The entrained flow reactor consists of only one pressure vessel that is usually small in volume due to the short fuel residence time. The fuel pre-treatment and feeding system is costly due to the additional torrefaction process and the particle size requirements. Furthermore, an air separation unit is needed. Specific investment costs for large scale biomass-to-liquid plants (ca. 400 MW_{th}) are reported in literature. The specific investment of the total plant is 1568 $/$kW_{th}$ [54]. The costs of the entrained flow gasifier (including biomass pre-processing, air separation unit and syngas cleaning) are reported to be 381 $/$kW_{th}$ [16]. The specific costs of a smaller scale (ca. 25 MW_{th}) biomass-to-power plant based on an entrained flow gasifier are estimated to be 1378 €/kW_{th} [55]. The specific costs of the gasifier (including auxiliary system: biomass pre-treatment, ASU, etc.) are not given in detail, but are assumed to have the highest share.

A two stage fluidised bed gasifier may be as complex as an entrained flow reactor, and the reactor size and volume may be larger due to the higher fuel residence time and the additional bed material. Furthermore, technical equipment is needed for product gas cleaning and gas compression units. The investment of a biomass-to-power plant (25 MW_{th}) based on fluidised bed gasification is reported to be 1641 €/kW_{th}, but can be decreased to 1063 €/kW_{th} on a larger scale (70 MW_{th}) [6]. These costs are slightly higher compared to the entrained flow costs given by Vogel et al. [55]. However, Vogel et al. directly compare biomass-to-power routes based on an entrained flow reactor and a two stage fluidised bed gasifier. The costs of the entrained flow gasifier plant are about 20% higher compared to the fluidised bed gasifier plant [55]. The higher costs may be explained by the number of facilities installed. The cost data for entrained flow gasifiers are based on either unique or expensive prototype plants, like the Choren gasifier [56], whereas a larger number of fluidised bed gasifiers is already commercially available.

Therefore, a clear difference in cost of the two gasifier types is not found in literature and the construction and realisation of an entrained flow gasifier is assumed to be as complex as a pressurised fluidised bed gasifier. For both technologies, a larger number of process units would bring down the costs.

As a similar complexity of both technologies is assumed, the same investment for both fluidised bed gasifier and entrained flow gasifier is expected in the middle future. The investment costs of the gasifiers include all auxiliary systems. For the fluidised bed gasifier these are for example the compression stations (air and product gas), the biomass feeding system, and the gas cleaning section. The investment costs of the entrained flow gasifier comprise for example the air separation unit, the biomass pre-treatment (torrefaction and grinding), and the gas cleaning section.

It is assumed that the specific investment costs decrease for a gasifier with a larger throughput due to the economy of scale. The investment of each gasifier is calculated using

$$C_2 = \left(\frac{Q_2}{Q_1}\right)^{0.65} \cdot C_1 \quad (4)$$

where $C_{1,2}$ denote the investment cost and $Q_{1,2}$ the thermal fuel input on each scale. The cost estimate is based on specific investment costs for a 10 MW_{th} gasifier. These costs are assumed to be 1100 €/kW_{th} which are slightly lower compared to biomass-to-power plant costs reported in literature [6,55] on slightly larger scales. The total investment costs of the 10 MW_{th} gasifiers are 11 million €. The cost estimate (economy of scale) results in specific costs of the 500 MW_{th} gasifier of 280 €/kW_{th} which is in good agreement with gasifier costs (including biomass pre-treatment, ASU, product gas cleaning) given in literature for large scale installations (381 $/$kW_{th}$ [16]).

Further economic parameters are the interest rate of 0.06 and the utilisation period of 20 years. The annual capital costs are calculated with an annuity method.

The investment comprises the gasifier including all necessary auxiliary components, a gas cleaning unit for particle, H_2S and HCl removal, and in the case of the fluidised bed gasifier a compressor for the compression of the product gas to 2.5 MPa.

For the annual cost of repairs and servicing, 1.5% of the investment and 1% for insurance and other costs are assumed, respectively. The operational staff for the 10 MW_{th} plant is assumed to be 8 persons (three shifts), for the 50 MW_{th} plant 16, for the 100 MW_{th} plant 24 and for the 500 MW plant 30 persons – each at €40,000/year. The fuel costs for wood chips with a fuel water content of 20 wt.% are assumed to be €98/t (€2.5 cent/kWh) based on the German fuel market in the years 2011 and 2012. Full operating hours are assumed to be 7000 h/year, which correlates with practical experience, e.g. from Güssing [57]. Table 4 shows all costs of the fluidised bed and entrained flow gasifiers. The product gas output of each gasifier and the resulting gas production costs are given in Table 5. The production costs are €4.2 ct/kWh to €7.4 ct/kWh. Due to similar efficiencies and investment costs of both gasification technologies the production costs are only marginally different for gasifiers of the same scale. If different investment costs of the gasification technologies are assumed, the production costs are expected to differ. The influence of a variation of input parameters on the product gas costs is examined in a sensitivity analysis.

For the three most important parameters – biomass costs, operating hours and investment costs – the effect of a parameter variation is shown for the 10 MW_{th} and 100 MW_{th} plant. The parameters are varied in the range ±20% and the result (exemplary for the entrained flow gasifier) is shown in Fig. 9. The fuel costs and the annual full operating hours are the main factors which are influencing the costs of the product gas. Especially the fuel costs are of importance because these costs have been growing for years due to the high demand of biomass in the energy sector. The influence of the investment costs is lower for bigger plants due to lower

Table 4
Investment and operating costs of the fluidised bed (10 MW, 50 MW, 100 MW) and the entrained flow gasifier (all scales) (m: million; k: thousand).

	10 MW	50 MW	100 MW	500 MW
Total investment (€)	11.0 m	31.3 m	49.1 m	139.9 m
Annual invest costs (€/a)	959 k	2.73 m	4.28 m	12.2 m
Annual fuel costs (€/a)	1.7 m	8.7 m	17.3 m	86.7 m
Repairs and servicing (€/a)	165 k	470 k	737 k	2.1 m
Personnel costs (€/a)	320 k	640 k	960 k	1.2 m
Insurance and others (€/a)	110 k	313 k	491 k	1.4 m
Total annual costs (€/a)	3.29 m	12.82 m	23.80 m	103.6 m

Table 5
Product gas output and production costs of entrained flow and fluidised bed gasifiers.

	Entrained flow gasifier				Fluidised bed gasifier		
	10 MW	50 MW	100 MW	500 MW	10 MW	50 MW	100 MW
Product gas output (MW)	6.45	33.7	69.1	354.0	6.4	34.0	70.5
Gas production costs (€ ct/kWh)	7.3	5.4	4.9	4.2	7.4	5.4	4.8

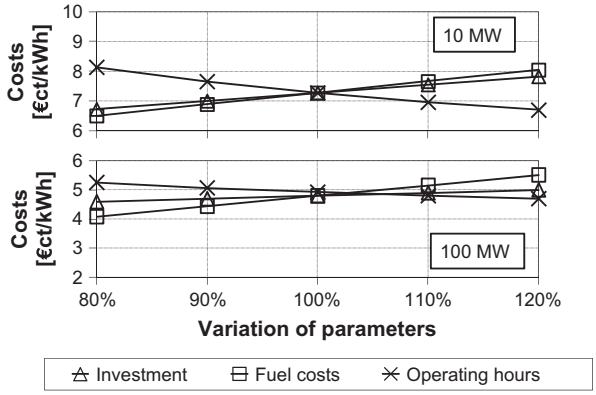

Fig. 9. Influence of investment costs, fuel costs, and annual full load hours on the product gas production costs of entrained flow gasification for a thermal input of 10 MW_{th} and 100 MW_{th}.

specific costs. The influence of the investment costs is almost comparable to the influence of the fuel costs if the economics of the 10 MW_{th} gasifier are considered. On all scales, the operating hours and the fuel costs are very essential for the product gas costs. The influence of annual full load hours is more significant for a small gasifier size. As the sensitivity analysis of fluidised bed gasification leads to similar conclusions, it is not shown in detail.

The cost analysis is based on several simplifications (e.g. static calculation, no detailed analysis for investment costs) and does not claim to predict precisely the future product gas costs of a biomass gasifier. However, the cost analysis gives the magnitude of the costs, the influence of scale on the costs and identifies crucial parameters (in the sensitivity analysis) that have to be focused on in a more detailed economic study.

6. Conclusions

The gasifier performance for the production of a pressurised product gas (2.5 MPa) from woody biomass is analysed in different scale. Dual stage fluidised bed (gasifier pressure 0.6 MPa, compression of the product gas to 2.5 MPa) and entrained flow reactors (gasifier pressure 2.8 MPa) are compared directly. The CGE of the fluidised bed is 0.4–1.9 percentage points higher than for the entrained flow reactor. However, if the auxiliary power demand is considered, the net efficiency of both gasifier technologies is similar (64–71%). Product gas compression and air compression for the combustion bed are the main power consumers of the fluidised bed gasifier. The entrained flow reactor needs electricity mainly for the ASU, the oxygen compression, and the fuel pulverisation. The biomass pre-treatment technology (torrefaction) of the entrained flow gasifier is identified as a crucial process unit for the overall plant performance. Energy losses in this unit significantly reduce the overall net efficiency, but the disintegration of the biomass structure reduces the power demand for pulverisation. Therefore, a trade-off has to be found in the severity of operating conditions of the torrefaction process.

The determination of investment cots of both gasifier types is a difficult task as biomass gasifiers are not commercially available in the larger scale. The conclusions from the economic evaluation are that the plant scale is an important parameter for the production costs of gas. Due to the specific lower investment costs, a larger gasifier throughput is favourable. However, logistics and transportation issues that will increase at a larger scale are not considered here.

If a pressurised product gas is required on a larger scale for a gas turbine application or a chemical synthesis, both gasification technologies show a similar performance. However, based on the given boundary conditions, the entrained flow reactor might be advantageous due to its simpler operation and higher reliability compared to the fluidised bed gasification process. The potential of the entrained flow technology for the conversion of biomass is shown and entrained flow reactors should be considered in future biomass applications.

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