DIPLOMARBEIT

Co-gasification of sewage sludge and wood in a dual fluidized bed steam gasifier

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Contents

1 Abstract

Power generation from biomass is an attractive technology for a sustainable energy supply of future generations. Integrated gasification combined cycle (IGCC) power plants have been developed for more efficient usage of solid fuels. In most cases the fuel is coal but in the same way biomass can be used for combined heat and power production. Research on gasification of wood was successfully done in the past two decades and pilot plants in industrial size were built. As the feedstock for these systems is actually limited to wood, seasonal fluctuations of the prize and availability affect the plant business. Therefore research activities on increasing fuel flexibility are underway.

In this thesis the usability of dried sewage sludge for co-gasification with wood was investigated. The required experiments have been carried out at the dual fluidized bed steam gasification pilot plant of the University of Canterbury in Christchurch, New Zealand. In these experiments bed material was graywacke sand and an olivine/calcite mixture, the operation conditions were 750 °C gasification temperature at atmospheric pressure and a steam to fuel ratio of 0.89. As feedstock wood pellets, mainly from radiata pine, and granulated dried sewage sludge form a municipal sewage treatment plant in Auckland, New Zealand were used. The ratio of sewage sludge in the feed was raised from 10 wt% to 100 wt%. From each run the average producer gas composition and the cold gas efficiency have been determined.

It was found that gasification of sewage sludge can be done in this type of reactor at regular gasification conditions. Increasing sewage sludge ratios in the feed caused higher nitrogen contents and also moderately higher hydrogen contents. On the other hand the percentages for CO and CO_2 dropped while the CH_4, C_2H_4 and C_2H_6 fractions in the producer gas remained nearly constant. Although the producer gas composition changed, the lower heating value of the achieved gas was nearly equal for all mixtures with just a slight tendency down. Higher sewage sludge percentages in the feed caused a decreasing producer gas yield and an also decreasing cold gas efficiency. The reason for this behavior is the high ash content of the sludge.

Based on the observations during this experimental series it seems possible to use sewage sludge for gas generation in fluidized bed gasifiers. After appropriate gas cleaning the produced gas should be suitable for usage in gas engines or turbines for example in a combined heat and power plant.

2 Kurzfassung

Verwendung von Biomasse zur Energiegewinnung ist eine attraktive Technologie für eine nachhaltige Stromversorgung zukünftiger Generationen. Kombi-Prozesse mit integrierter Vergasung (Integrated gasification combined cycle - IGCC) wurden für eine effiziente Nutzung von festen Brennstoffen entwickelt. Meist wird dabei Kohle als Brennstoff verwendet, aber auch Biomasse kann auf diese Weise für die kombinierte Erzeugung von elektrischer Energie und Wärme (Kraft-Wärme-Kopplung) genutzt werden. Erfolgreiche Forschung in den letzten beiden Jahrzehnten ermöglichten den Bau von Pilotanlagen in industriellem Maßstab. Da der Brennstoff bei diesen Anlagen zur Zeit aber noch auf Holz beschränkt ist, wird der Anlagenbetrieb sehr von den saisonalen Schwankungen des Preises und der Verfügbarkeit beeinflusst. Aus diesem Grund werden derzeit Möglichkeiten zur Erhöhung der Brennstoffflexibilität untersucht.

In dieser Diplomarbeit wird die Verwendbarkeit von getrocknetem Klärschlamm zur Co-Vergasung mit Holz untersucht. Die erforderlichen Versuche wurden am Zweibett-Wirbelschicht-Dampfvergaser der Universität von Canterbury in Christchurch, Neuseeland durchgeführt. In diesen Versuchsläufen wurde Graywacke Sand und eine Olivin/Calcit Mischung als Bettmaterial verwendet, die Versuchsbedingungen waren 750 °C Vergasungstemperatur bei Umgebungsdruck und ein Dampf-Brennstoffverhältnis von 0,89. Als Feed wurden Holzpellets, hauptsächlich aus Monterey-Kiefer, und granulierter, getrockneter Klärschlamm von einer kommunalen Kläranlage in Auckland, Neuseeland verwendet. Der Klärschlammanteil im Feed wurde schrittweise von 10 bis 100 m% gesteigert und von jedem Versuchs-

lauf die durchschnittliche Gaszusammensetzung und der Kaltgaswirkungsgrad bestimmt.

Es konnte gezeigt werden, dass Klärschlamm mit diesem Reaktor bei den oben genannten Prozessbedingungen vergast werden kann. Mit steigendem Anteil von Klärschlamm im Feed stieg der Anteil von Stickstoff im Produktgas stark, der Gehalt an Wasserstoff nahm moderat zu. Dagegen fielen die Werte für CO und CO_2 kontinuierlich, die Anteile von CH_4, C_2H_4 und C_2H_6 blieben etwa gleich. Obwohl sich die Produktgaszusammensetzung änderte, blieb der errechnete Heizwert des erhaltenen Gases nahezu konstant und fiel nur leicht. Höhere Anteile von Klärschlamm im Brennstoff bewirkten aber eine reduzierte Produktgasausbeute und dadurch einen geringeren Wirkungsgrad, dies lässt sich aufgrund des hohen Aschegehalts des Klärschlamms erklären.

Aus den Ergebnissen der vorliegenden Versuche lässt sich ableiten, dass Klärschlamm grundsätzlich zur Gaserzeugung in Wirbelschichtreaktoren und nach entsprechender Aufbereitung das erzeugte Produktgas z.B. für die Verwendung in Gasmotoren oder Turbinen zur Strom- und Wärmeerzeugung geeignet ist.

3 Introduction

Increasing energy demand all over the world, problems with the security of energy supplies, global warming and the greenhouse effect brought the energy industry into the focus of billions of people. The traditional energy industry is based on fossil fuels like coal, petroleum and natural gas. As these fuels and also nuclear fuels are limited, energy generation from renewable sources and waste-to-energy concepts are necessary for a new, sustainable energy industry. Additionally to the limited resources the main problem with traditional energy generation is the production of CO_2, which is next to water vapor one of the most important greenhouse gases. Therefore research and development of alternative energy generation technologies are among the most important topics of actual studies at the universities and research centers.

Biomass gasification is one high potential technology in this research field. As the flexibility of the process allows generation of burnable gases from many different carbonaceous substances, it is universally applicable and could be widely used in combined heat and power (CHP) plants. At the Vienna University of Technology a dual fluidized bed system steam gasification of wood was developed and is successfully used in the $8\,\mathrm{MW}_{\mathrm{fuel\ input}}$ pilot plants Güssing and Oberwart in Austria [1, 2, 3]. Actual further projects for increasing fuel flexibility and refining of the produced gas, to produce synthetic natural gas (SNG) via steam reforming or synthetic liquid fuels via Fischer-Tropsch synthesis are carried out.

Mainly coal or woody biomass is used in gasification applications, but also straw, organic waste from household or industry and sewage sludge would be imaginable as energy source, as they mainly contain hydrocarbons and have a lower heating value between 10 and 18 MJ/kg [4]. Therefore the usability of dried sewage sludge for co-gasification with wood pellets should be investigated in this thesis. Firstly a literature research should help to define the state of the art in co-combustion and co-gasification of difficult fuels. Secondly tests at a dual fluidized bed steam gasification pilot plant with different ratios of dried sewage sludge and wood pellets should prove the usability. Knowledge about the general behavior of sewage sludge during the gasification process and possibly optimized mixing rates should be gained. Finally the results should be compared with results from previous experiments at Vienna University of Technology.

4 Theory of gasification and fluidized bed technology

4.1 Gasification [4, 5, 6]

4.1.1 Fundamentals

Gasification is the thermo-chemical conversion of solid or liquid carbonaceous fuels mainly into H_2, CO and CO_2 using a gasifing agent like air, oxygen or steam. Fuels can be coal, petroleum and biomass, their derivates like coke, plastics or other organic materials. Additional to the main components H_2, CO and CO_2 the produced gas contains CH_4, N_2 and SO_2. Gasification is a mainly endothermic process, therefore heath supply from partial combustion or external heating is necessary.

Auto-thermal gasification The thermal energy for the gasification is produced inside the reactor through partial combustion. As gasification agent air or oxygen with an equivalence ratio λ between 0.3 and 0.4 is commonly used.

Allo-thermal gasification For allo-thermal gasification the thermal energy is provided from an external source, supplied through a heat exchanger or a circulating bed-material.

The gasification process for a typical particle is characterized through four stages: heating and drying, pyrolysis, oxidation and reduction. In the first two stages the particle has a temperature of 100 - 500 °C and does not react with the gasification agent, only thermal energy is used to heat up the particle and for decomposition of macromolecules to gas, coke and oil. In the other two stages the temperature is between 500 and 1000 °C and the conversion is dominated through chemical reactions of the pyrolysis products with the gasification agent.

4.1.2 Main gasification reactions

The following reactions describe different ways for conversion of solid carbon to gaseous components as well as conversion of some gaseous components.

$$\text{Oxidation} \qquad C + O_2 \rightleftharpoons CO_2 \qquad \Delta H = -406.1 \frac{MJ}{kmol} \tag{4.1}$$

$$\text{Partial Oxidation} \qquad 2C + O_2 \rightleftharpoons 2CO \qquad \Delta H = -246.2 \frac{MJ}{kmol} \tag{4.2}$$

$$\text{Boudouard} \qquad C + CO_2 \rightleftharpoons 2CO \qquad \Delta H = +159.9 \frac{MJ}{kmol} \tag{4.3}$$

$$\text{Water-gas} \qquad C + H_2O \rightleftharpoons CO + H_2 \qquad \Delta H = +118.1 \frac{MJ}{kmol} \tag{4.4}$$

$$\text{Water-gas shift} \qquad CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad \Delta H = -41.8 \frac{MJ}{kmol} \tag{4.5}$$

$$\text{Methanation} \qquad C + 2H_2 \rightleftharpoons CH_4 \qquad \Delta H = -88.3 \frac{MJ}{kmol} \tag{4.6}$$

$$\text{Steam reforming} \qquad CH_4 + H_2O \rightleftharpoons CO + 3H_2 \qquad \Delta H = +206.4 \frac{MJ}{kmol} \tag{4.7}$$

4.1.3 Influencing parameters

Temperature At lower temperatures the exothermic oxidation reactions are dominating the process and the product gas contains mainly CO_2, CO and in case of steam gasification H_2O. Increasing temperature promotes endothermic reactions, above 800 °C the CO_2 and H_2O content is clearly reduced and the CO and H_2 content is rising, the Boudouard reaction becomes more dominant (fig. 4.1).

Pressure Higher pressure leads to a higher CH_4 and CO_2 and lower H_2 and CO content. This is caused through the pressure dependence of the equilibrium of the reactions (4.3), (4.4), (4.6) and (4.7).

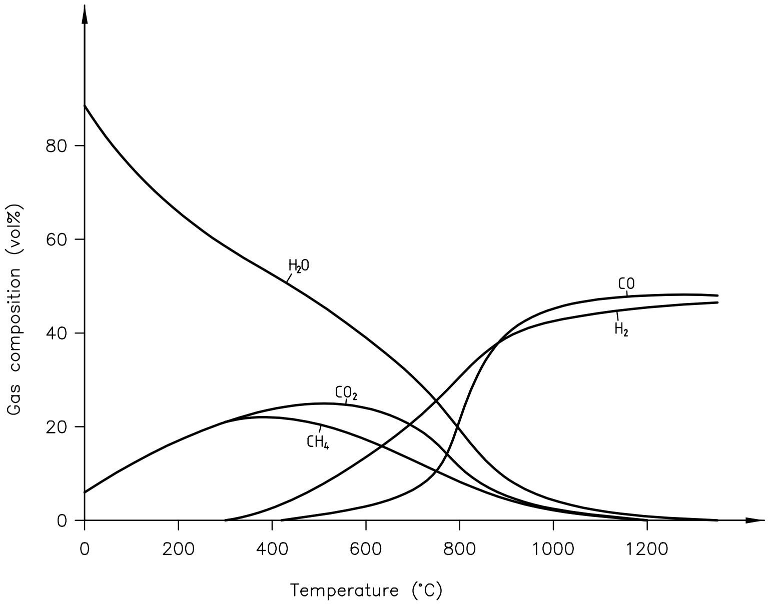


Figure 4.1: Producer gas composition vs. gasification temperature (coal) [6]

Gasifying agent The product gas composition depends essentially on the selected gasification agent. *Air* is easily available and cheap, but as major disadvantage it contains 79% of nitrogen. This high nitrogen content causes a nitrogen content in the producer gas between 40 and 60% which lowers the caloric value of the producer gas significantly to 3 - 6.5 MJ/Nm^3. *Oxygen* as gasification agent is more expensive than air, but the achieved caloric value of the producer gas is with 12 - 16 MJ/Nm^3 about the double of using air. *Steam* is an alternative to air and oxygen and can be used in allo-thermal gasifiers. The produced gas is hydrogen rich and has, like using oxygen, a quite high caloric value.

Reactor type The producer gas composition also depends heavily on the type (fixed bed, fluidized bed or entrained bed gasifier, see chapter 4.1.4) and on the size of the reactor. In contrast to the gas composition the caloric value is nearly independant and equal for all types.

4.1.4 Gasification reactors

Fixed bed gasifier In a fixed bed gasifier usually lumpy fuels are used which are fed on top of the gasifier, moving down during the gasification process. The subprocesses of gasification proceed in separated areas while the gasification agent flows upwards through the whole gasifier and all gasification zones (*updraft gasifier*) or downwards through the oxidation and reduction zone (*downdraft gasifier*). The main advantage of the updraft gasifier compared to the downdraft gasifier is the high gasification efficiency, caused through the lower producer gas outlet temperature since the hot gas flows through the heating/drying and pyrolysis zone. However the gas-flow through the pyrolysis zone causes a remarkable tar load in the producer gas in contrast to a downdraft gasifier. A possibility to compensate the disadvantage of the higher gas outlet temperature is to use an external heat exchanger for preheating the fuel or the gasifying agent or a combination of both systems like the twin fire gasifier.

Fixed bed gasifiers are used in rather small plants with a thermal power output below 10 MW, updraft gasifiers just for heat generation while downdraft gasifiers are used in CHP-plants with an electrical power output up to 500 kW [4].

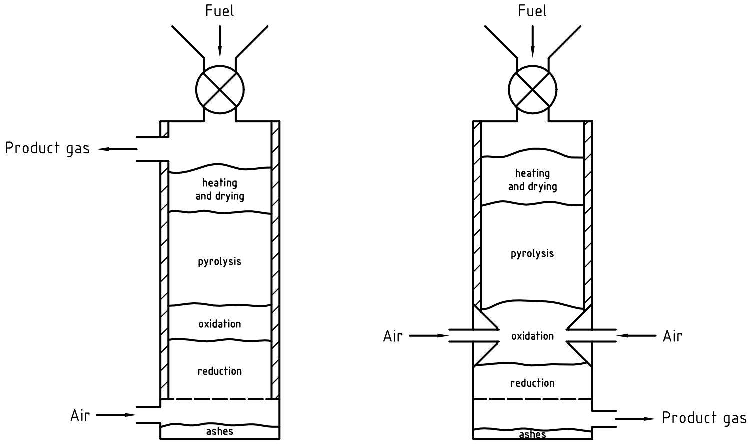


Figure 4.2: Fixed bed gasifiers. left: updraft gasifier, right: downdraft gasifier [4]

Fluidized bed gasifier A fluidized bed gasifier can be used for granulated, pelletized or chopped fuel which is inserted into the bed of inert or catalytic sand. Caused through the fast turbulence there a no separated gasification zones for heating/drying, pyrolysis, oxidation and reduction and each subreaction runs parallel in the whole reactor. The turbulence causes also an excellent heat exchange between the fuel and the bed-material which is essential in allo-thermal gasifiers. The gasification temperature is nearly equal in the whole bed and well controllable. The producer gas leaves the reactor with nearly reaction temperature, therefore heat recovery systems for preheating the gasifying agent or the fuel have to be installed. Usually the tar load of the producer gas from a fluidized bed gasifier is clearly below an updraft gasifier but about a downdraft fixed bed gasifier. As a consequence of the fluidized bed small particles could be discharged out of the reactor with the producer gas.

Fluidized bed gasifiers can be built as bubbling fluidized bed, circulating fluidized bed or a combination of both like the dual fluidized bed gasifier, described in chapter 4.2.3. These are build for a thermal power output up to 100 MW or as IGCC-plants (Integrated gasification combined cycle) for combined thermal and electrical power generation with an electrical output up to 12 MW [4].

Entrained bed gasifier The fuel is blown finely milled in cocurrent flow with the gasifying agent into the reactor pipe and gasifies at temperatures between 1200 and 2000 °C within a few seconds. When using biomass the fuel is often prepared as milled coke from pyrolytic decomposition as it is very difficult to reach this high temperature with biomass-fibres because of the low heating value and the high water content. At these high temperatures the ash is molten and is removed liquid, which is an advantage for biomass fuels with a low ash melting point.

Entrained bed gasifiers are technologically complex and too expensive for small scale reactors. For these reasons they became not as important as fixed bed or fluidized bed gasifiers and are only built for high power output of 130 MW and more [4].

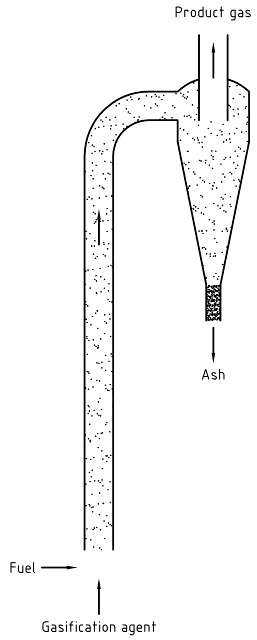


Figure 4.3: Entrained bed gasifier [4]

4.2 Fluidized bed technology [7, 8, 9]

4.2.1 Fundamentals

A fluidized bed is formed by granulated solids when a fast enough upwards flowing gas or liquid lifts the particles and the mixture achieves a fluid-like state. In 1921 F. Winkler observed this behavior the first time when he tried to gasify coal with a steam/oxygen mixture, streaming from the bottom of crucible through the particles [10]. The fluidized particles behave like a liquid, the surface remains horizontal when the reactor is tilted and when fluidized with high gas streams bubbles raise up like in a boiling liquid.

Depending on the fluidization velocity different states of fluidization are recognized (fig. 4.4). At low velocities the particles remain in a packed state, the *fixed bed*. If the velocity increases the bed expands and the particles begin moving until they fully suspend and the *minimum fluidization* state is reached. Further increasing flow rate, above 5 to 6 times of the minimum fluidization velocity, leads to *bubbling fluidization* and in tall reactors, if the bubble diameter exceeds the reactor's internal diameter, to *slugging*. Even higher fluidization velocities can cause a turbulent state and lead finally to a *fast fluidized bed*, an *entrained bed* or *pneumatic transport*.

4.2.2 Characteristics of fluidized beds

Particle size and bulk density The mean parameter to describe the particle size is the diameter. As the most particles of technical interest are not ideal spherical an *equivalent diameter* has to be used. In fluidized bed applications the equivalent diameter is normally defined as the diameter of that sphere which has the same surface/volume ratio like the real particle.

$$d_{sv} = \frac{6 \cdot V}{S} \tag{4.8}$$

The *particle density* is defined as the particle mass divided by the volume of the particle, including small pores.

$$\rho_p = \frac{m_P}{V_P} \tag{4.9}$$

By contrast the *bulk density* is defined as the mass of all particles divided by the whole volume they occupy.

$$\rho_b = \frac{m}{V_B} \tag{4.10}$$

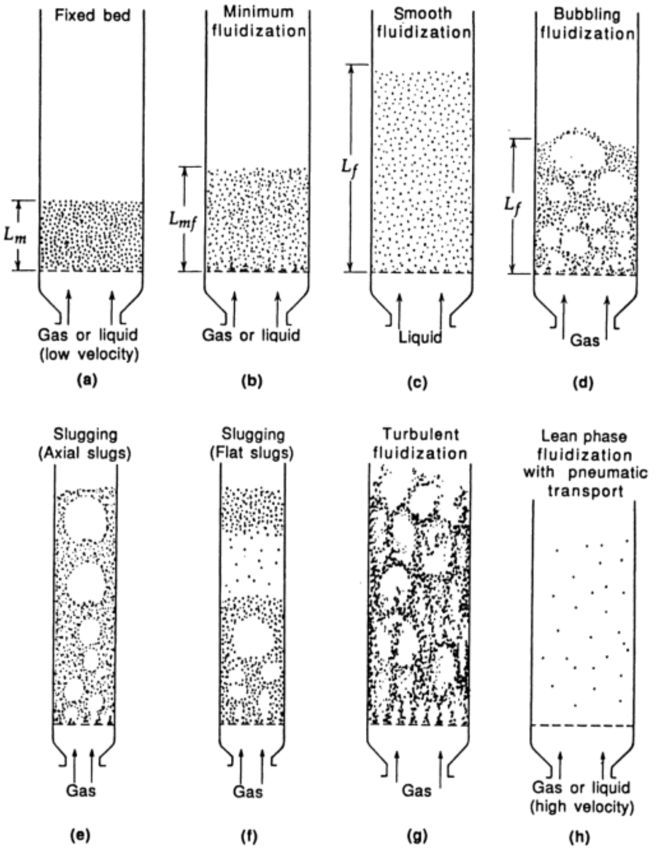


Figure 4.4: States of fluidization [9]

Porosity is the ratio between the void volume and total volume of the bulk and can be calculated as

$$\varepsilon = 1 - \frac{\rho_b}{\rho_p}. \tag{4.11}$$

Minimum fluidization velocity In a fluidized bed the resisting force of the fluid stream is equal to the weight of the particles minus the buoyant force. Therefore the *pressure drop* is determined as

$$\Delta p = (1 - \varepsilon) \cdot (\rho_p - \rho_g) \cdot g \cdot H \tag{4.12}$$

which is the fundamental fluidization condition. The pressure drop in a fixed bed is calculated with the Ergun equation.

$$\frac{\Delta p}{H} = 150 \cdot \frac{(1-\varepsilon)^2}{\varepsilon^3} \cdot \frac{\mu \cdot U}{d_{sv}^2} + 1.75 \cdot \frac{1-\varepsilon}{\varepsilon^3} \cdot \frac{\rho_g \cdot U^2}{d_{sv}} \tag{4.13}$$

The *Archimedes number* is a dimensionless number used to determine the motion of particles in fluids due to density differences.

$$Ar = \frac{\rho_g \cdot d_{sv}^3 \cdot (\rho_p - \rho_g) \cdot g}{\mu^2} \tag{4.14}$$

Together with the Archimedes number the two equations for the pressure drop are used to calculate the minimum fluidization velocity, because the pressure drop in the fluidized bed is equal to the fixed bed at the point of minimum fluidization (fig. 4.6). As ε_m was found nearly constant over the range of Reynolds numbers from 0.001 up to 4000 in extensive experiments the equation simplifies to

$$U_m = \frac{\mu}{\rho_g \cdot d_{sv}}(\sqrt{33.7^2 + 0.0408 \cdot Ar} - 33.7). \tag{4.15}$$

Transport velocity Transition form a captive state to a transport stage of fluidized beds is mark through the terminal velocity of the particles. For a single particle in an upwards flowing fluid the gravitational force is opposed to the buoyant force, the drag force and the acceleration force (fig. 4.5).

$$m_p \cdot g = m_p \cdot \frac{\rho_g}{\rho_p} \cdot g + C_D \cdot \frac{\pi(U - U_p)^2 \rho_g \cdot d_p^2}{8} + m_p \cdot \frac{dU_p}{dt} \tag{4.16}$$

The terminal velocity for a stationary hovering particle is reached when the acceleration term is zero.

$$U_t = \sqrt{\frac{4}{3} \cdot \frac{\rho_p - \rho_g}{\rho_g} \cdot \frac{d_p \cdot g}{C_D}} \tag{4.17}$$

If the bed is fluidized above the terminal velocity single particles are entrained. Further increasing flow rate leads to a critical velocity, the transport velocity, where the time to empty the reactor suddenly drops and the regime changes to

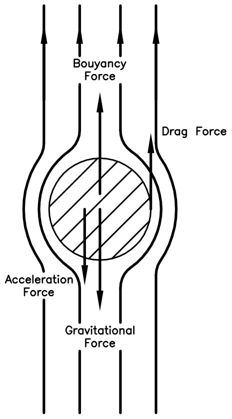


Figure 4.5: Force balance on a moving particle

fast fluidization. Perales developed an empirical equation, using the Archimedes number, valid for Reynolds numbers between 20 and 50 000.

$$U_{tr} = 1.45 \cdot \frac{\mu}{\rho_g \cdot d_p} \cdot Ar^{0.484} \tag{4.18}$$

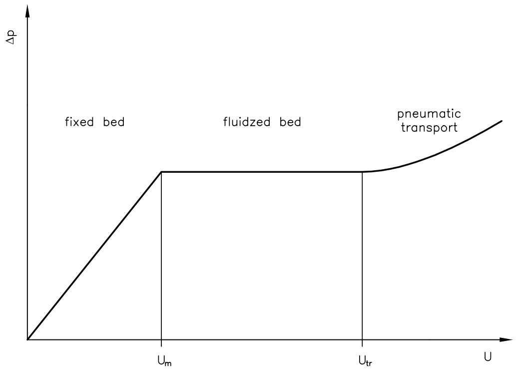


Figure 4.6: Pressure drop vs. fluidization velocity [8]

4.2.3 Fluidized bed reactors

Bubbling fluidized bed reactor In a bubbling fluidized bed the flow velocity is between the minimum fluidization velocity and the transport velocity and a clearly visible expanded bed is formed. Increasing flow rate causes growing bubbles and maybe leads to turbulent fluidization. Above the dense phase of the bubbling bed is the freeboard. Due to eruption of bubbles on the surface particles are thrown in free space above the bed and travel upwards (fig. 4.7). As their kinetic energy is not high enough to entrain from the reactor they fall back into the bed zone. Above the *transport disengaging height* (TDH) are only fine particles, whose fluid drag is greater than their weight and which are carried out of the reactor.

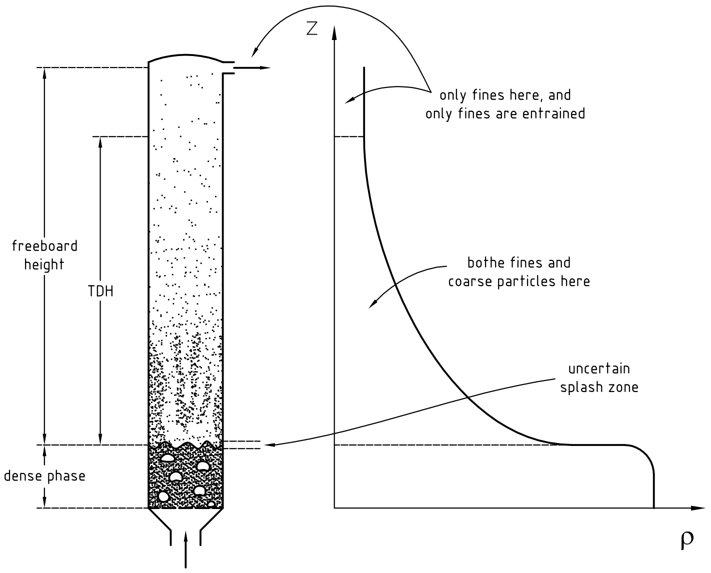


Figure 4.7: Freeboard and transport disengaging height [9]

Fast fluidized bed reactor In a fast fluidized bed the gas flow rate is above the minimum transport velocity and the bed material is carried out of the reactor. The particles will be separeted from the gas stream and return to the bottom of the riser. At moderate fluidization velocities the lower section in a fast fluidized bed is, similar to the dense phase and the freeboard in a bubbling fluidized bed, denser than the higher section, which causes different pressure gradients in the two sections (fig. 4.8). At higher fluidization velocities these two zones unify and the pressure gradient becomes constant.

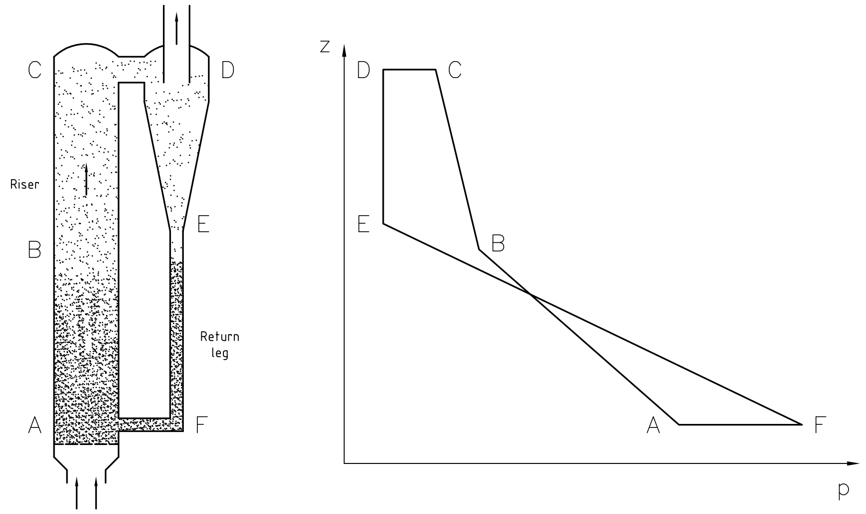
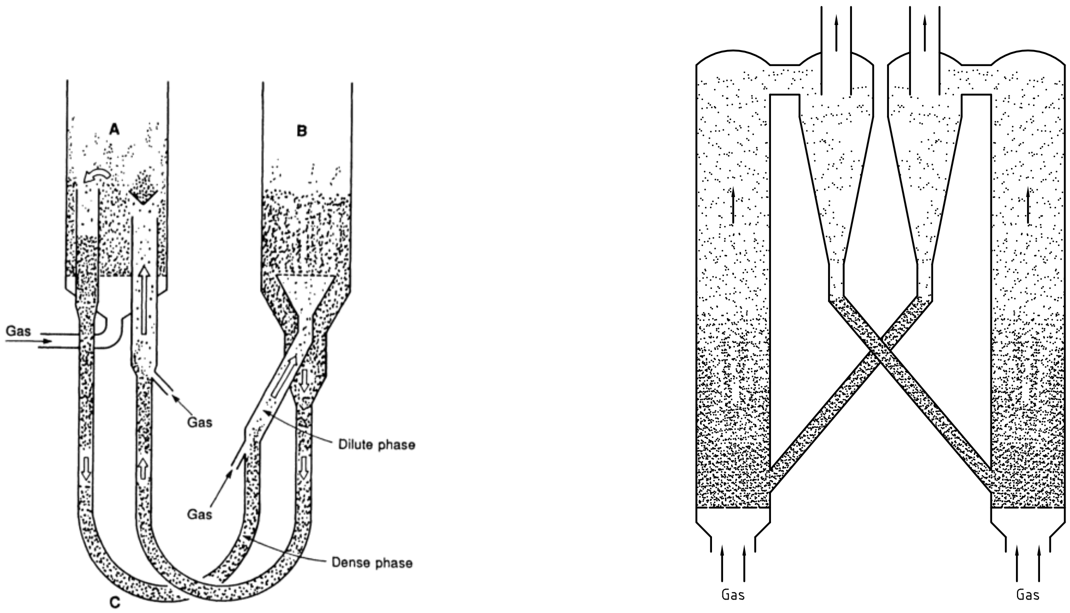


Figure 4.8: Pressure drop in a fast fluidized bed [7]

Dual fluidized bed reactor Dual fluidized bed reactors are usually used for allothermal processes like fluid catalytic cracking or gasification. The circulating bed material can be used for heat transfer and also as a catalyst, it is regenerated or reheated in one part of the reactor while the main reaction runs in the other part of the reactor. This concept is suitable for all possible combinations, as two bubbling beds, a bubbling and a fast bed or two fast beds. For capable operation of such a solid flow system, proper fluidization of the solid phase all over the reactor and in any operating condition must be maintained, as any settling of particles can cause a complete blockage of the circulating flow, and will kill the process.



(a) two fixed beds [9]

(b) two fast beds, adapted from [9]

Figure 4.9: Dual fluidized bed systems

5 State of the art of co-firing

Co-firing is the usage of different fuels at the same time for combustion or gasification. For example biomass is co-fired in existing coal plants or fuels, which could not be burned alone because of the low energy content could be burned together with natural gas to reach a good performance. Part of this thesis is a literature review to determine the state of the art of co-combusting and co-gasifying biomass with other potential fuels.

5.1 Co-combustion of biomass

5.1.1 Co-combustion of agricultural residues with coal in a bubbling bed combustor

Gahni et al., 2009 [11]

Agricultural residues like rice husk and palm kernel shell are among the main sources of biomass for energy utilization in Malaysia. These two sources provide an energy potential of about 245 petajoules every year. Therefore their usability for co-firing in coal-fired fluidized bed boilers was investigated in a study of the "University Putra Malaysia". This study brought up that co-firing of rice husk and palm kernel shell is possible with minimal modifications in existing boilers.

Experimental setup and test procedure For the tests an laboratory scale atmospheric fluidized bed combustor was used. A scheme of this combustor is shown in fig. 5.1. It has a combustion camber with 0.15 m diameter and 2.3 m height, is build for a bed hight up to 0.3 m and use of 850 µm silica sand. Fluidization air is distributed through a nozzle plate and provides also oxygen for the combustion. The premixed fuel is fed with a screw feeder and pneumatically transported on top of the bed. A cyclone separates flue gas from ashes and other bed carryover.

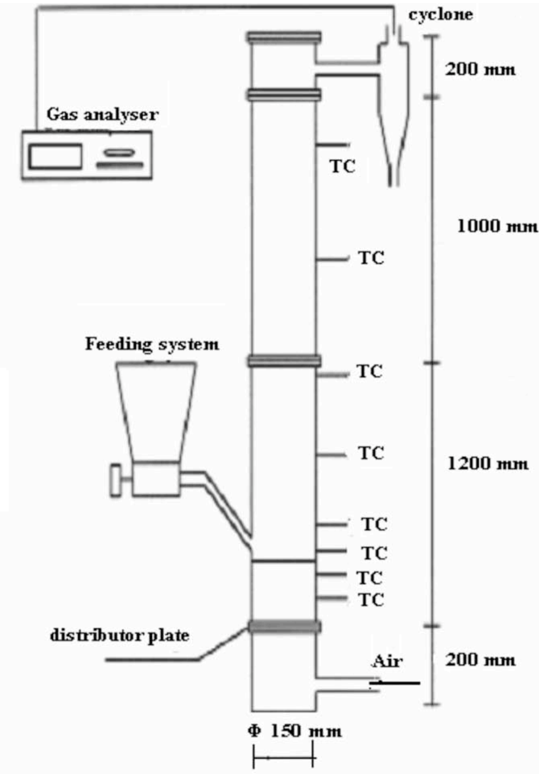


Figure 5.1: Scheme of the fluidized bed combustor (University Putra Malaysia)

Tests with the pure components have been undertaken as well as tests with mixtures of 50 wt% palm kernel shell and also 50 wt% of rice husk, both mixed with coal. The temperature profile of the reactor was recorded, also the flue gas and the ash composition was analysed and the combustion efficiency was calculated.

Results The combustion efficiencies range between 67-88% for burning rice husk and 80-92% for burning palm kernel shell. Coal combustion causes a higher bed temperature and a lower freeboard temperature compared to the co-firing runs and those with pure biomass. This behavior is explained by the devolatilisation process of the fuel. The CO-content of the flue gas was between 200 and 900 ppm and not significantly depending on the feed, but tending to higher rates for the biomass co-firing runs compared to the 100% coal runs. The analysis of the collected ash shows a low amount of unburned carbon, less than 5% of the total carbon input. For the pure biomass lower carbon content was measured than for the mixtures and the pure coal.

5.1.2 Co-combustion of cynara with two coals in a bubbling bed combustor

Aho et al., 2009 [12]

Cynara is a herbaceous plant with an oil content of 25 wt% (dry). The oil yield per hectare of the seeds of cynara is lower than that of sunflowers, but if the other parts of the plant would also be used for energy production, it could be a competitive plant for producing biodiesel. Like many other herbaceous plant, like cereal straws, its ash content is very high, between 5 wt% in the stems and 11 wt% in the leaves. Because of the usage of KCl fertilisers, the biomass contains up to 2 wt% chlorine which could cause corrosion problems in superheaters. There are two possibilities to reduce chlorine in the flue gas, first the usage of chlorine free fertilisers and second the use of substances which could capture the chlorine during combustion. Sulphur and aluminosilicates, which are normally contents of coals, have this ability. So co-firing of cynara with coal could reduce the operational problems, caused through the chlorine content, and could also reduce CO_2 emissions compared to coal firing alone.

Experimental setup and test procedure The experimental test rig (fig. 5.2) is a 4 m high bubbling fluidized bed reactor with an internal diameter of 0.16 m and a 0.55 m high bed of silica sand. For the tests seven different mixtures of cynara pellets with South African bituminous coal or spanish sub-bituminous coal were used. The mixtures have been burned at a bed temperature of 860 °C and reduced oxygen concentration, simulating flue gas recirculation.

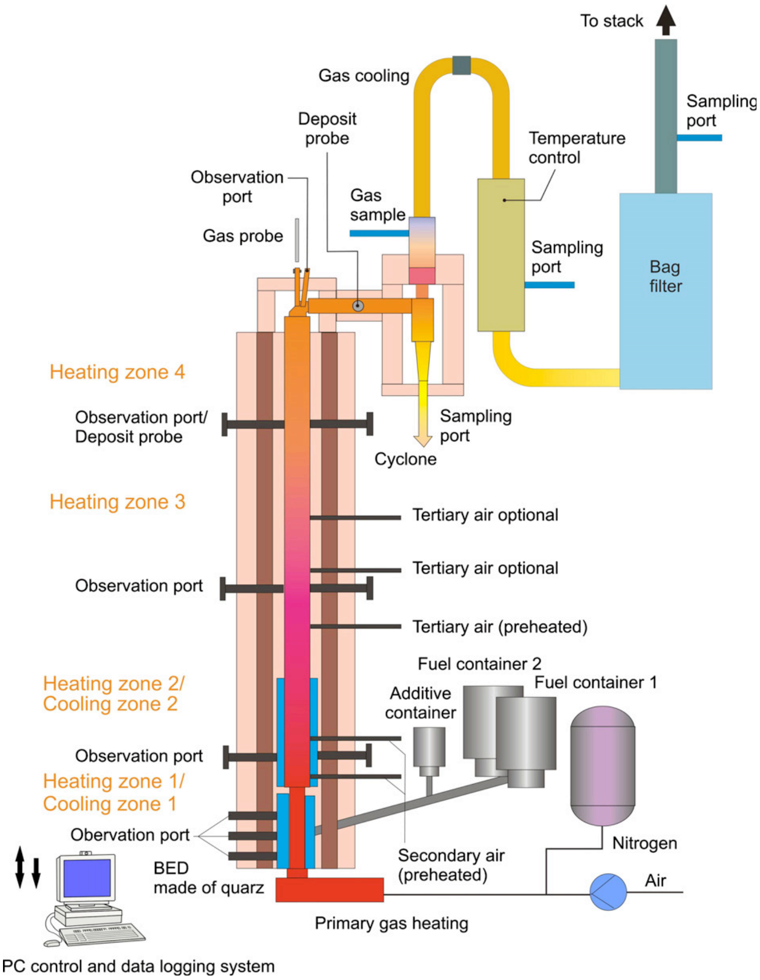


Figure 5.2: Scheme of the test rig (Technical Research Centre of Finland)

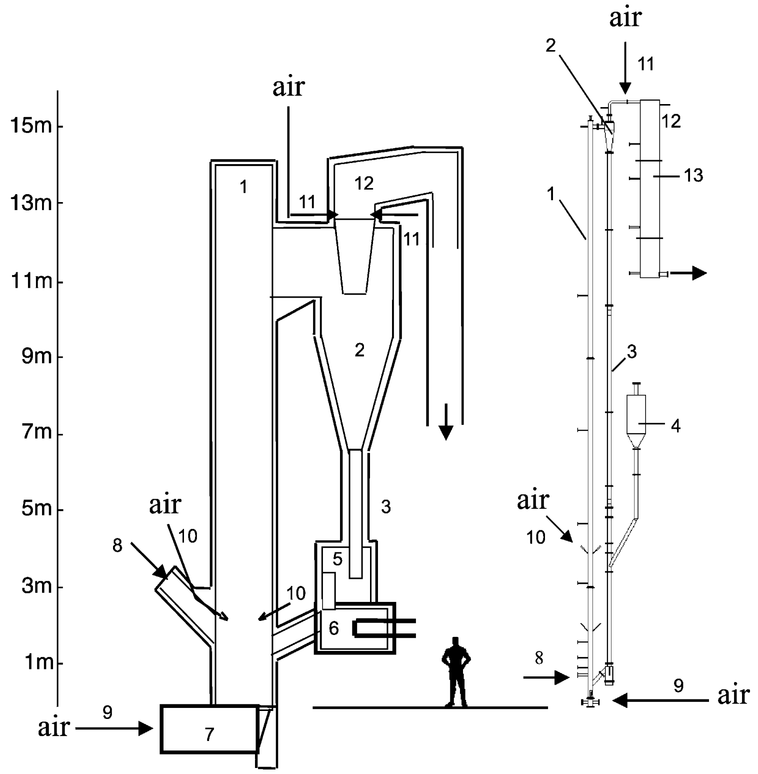
Results Cynara with a high chlorine content like the sample in this project can not be recommended for single firing in fluidized bed boilers. Co-firing of up to 10% of cynara on energy basis with coal is safe, and, using South African coal, SO_2 emissions are so low that limestone addition is not necessary. HCl emissions are below 400 mg/Nm^3, but nevertheless flue gas cleaning should be intended, specially if Spanish coal should be used which causes much higher SO_2 emissions because of its much higher sulphur content (8 wt% compared to 0.6 of South African coal). Earlier studies have shown that refuse-derived fuel with chlorine content lower than 0.6 wt% could be burnt safely up to 60% with South African coal. Cynara with chlorine contents in this range could be possibly burnt in as high proportion. It should be tried to reduce Cl and ash content by use of other fertilisers and harvesting technologies.

5.1.3 Co-combustion of sewage sludge with coal or biomass in a circulating fluidized bed

Leckner et al., 2004 [13]

Sewage sludge combustion is usually proceeded in special plants, as co-combustion in common power plants is difficult due to emissions of harmful gases and heavy metals. Aim of this feasibility investigation was to find ways of co-combustion with coal or biomass in existing fluidized bed combustors considering the legal conditions for gaseous emissions.

Experimental setup and test procedure The experiments have been undertaken in two circulated fluidized bed combustors, a lab scale unit and a pilot scale 12 MW$_{th}$ CFB boiler (fig. 5.3). The combustion chamber of the lab scale unit has a diameter of 0.1 m, equivalent 0.008 m^2, and is 15 m high, while the combustion chamber of the pilot scale plant has square cross section of 2.25 m^2 and is 13.6 m high. The feedstock is inserted with a screw feeder into the lab scale combustor or through a fuel chute at the pilot plant. Furthermore there are no other significant differences between these two systems and if adequate similarity rules are observed the measured results are practically identical. Each plant is equipped with secondary air inlets into the combustion chamber and between the cyclone and the afterburner chamber. Tested fuels have been polish coal, wood pellets and dried (19 wt% water content) respectively wet (>70 wt% water content) sewage sludge in different ratios. It was possible to run the pilot plant combustor with an energy fraction up to 50% of dried sludge or 10% wet sludge, while the lab scale unit could burn even higher concentrations due to electrical heating support.



(1) combustion chamber, (2) cyclone, (3) particle return line, (4) bed material hopper, (5) particle seal, (6) heat exchanger, (7) windbox, (8) fuel feed, (9) primary air supply, (10) secondary air addition into combustion chamber, (11) secondary air addition after cyclone, (12) after-burner chamber, and (13) probe for flue gas extraction

Figure 5.3: Combustion pilot plant (Chalmers Technical University) and lab scale combustor (Technical University Hamburg-Harburg)

Results The experiments had shown that sewage sludge could be burned togehter with the regular fuel without technical problems. If the moisture content exeeds 10 wt% storage problems occure because of the increase of odors and biological activities during longer storage periods. An energy fraction of dried sewage sludge up to 50% showed only slight differences in combustion performance and could be fired through the conventional feeding system. A continuous ash removing system designed for the ash content of around 40 wt% (daf) and the higher stickiness of the ash is neccesary. Combustion of up to 25% of sludge does not exceed EU emission limits for CO, NO_x and SO_2, but high limestone addition, specially during co-combustion with wood, was needed. Therefore in a wood fired boiler installation of special gas cleaning equipment is necessary, as in a coal fired boiler existing equipment can be used. Injection of hydrated lime into the flue gas prior to the filter would also be a suitable method for sulfur capture and would also reduce HCl emissions.

5.2 Co-gasification of biomass

5.2.1 Co-gasification of different biomass wastes in an entrained-bed gasifier

Lapuerta et al., 2008 [14]

The use of different biomass wastes from forestry, agriculture and industry as fuel for an air gasifier was investigated. These fuels are pruning wastes from pine, olive and wine, further sawdust and marc of grape. Additionally some tests for co-firing with coal and coke have been carried out. The selected biomass wastes are available in different times of the year and need operation flexibility. Therefore a main subject of this study was to evaluate if these fuels could be burned under similar conditions in the same gasifier.

Experimental setup and test procedure For this tests an electrical heated entrained-bed gasifier was used, which is 1.2 m high and has an internal diameter of 75 mm (fig. 5.4). The fuel samples have been milled to particles with a diameter less than 800 µm and inserted in the reactor tube through an injector together with the fluidization air. The gasification temperature was 1250 °C and the average residence time of the particles was between 0.6 and 1.7 s, depending on the biomass/air ratio.

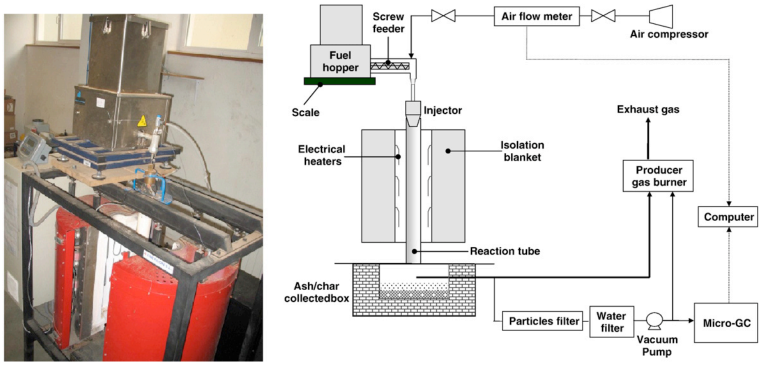


Figure 5.4: Gasification test rig (University of Castilla-La Mancha)

Results The tests had not shown significant differences between the tested fuels, which allows flexible usage of the fuel. The agricultural pruning wastes turned out to be a bit more efficient than forestry and industrial wastes, however sawdust gasification brought the highest H_2 concentration. The co-gasification experiments indicated more efficient usage of coal, combined with lower emissions and improvement on the producer gas quality. As expected the biomass/air ratio is the main parameter influencing the producer gas composition. An increasing biomass/air ratio causes increasing concentration of CO, H_2 and CH_4 and a decreasing CO_2 concentration. Higher biomass/air ratios cause also a reduced total amount of produced gas while the gasification efficiency goes up. A similar effect could be observed with rising temperature but the differences have been in a smaller range.

5.2.2 Co-gasification of biomass and HDPE in a fixed bed downdraft gasifier

García-Bacaicoa et al., 2008 [15]

The permanently increasing volume of plastic residues could be used for energy generation as the lower caloric value of high density polyethylene (HDPE) is 44 MJ/kg, which is quite high and should not get lost. As fixed bed gasifiers are not suitable for use with powdery or liquid residues, co-firing HDPE with wood is an option to consider the characteristics of this thermoplastic fuel. Downdraft gasification has the advantage of low tar generation and is suitable for small scale power generation.

Experimental setup and test procedure The gasification pilot plant (fig. 5.5) consists of a 2 m high cylindrical reactor with an internal diameter of 0.44 m with rotating grate, gas cleaning and cooling equipment and is connected to a modified diesel engine coupled to a 25 kVA alternator. The process runs autothermic using air with an equivalence ratio between 0.3 and 0.4 as gasification agent. Wood was supplied as approximately 40 mm big chips, while the HDPE chips had just a size around 5 mm. Mixtures with 10 and 15 wt% of polyethylene have been gasified during this study.

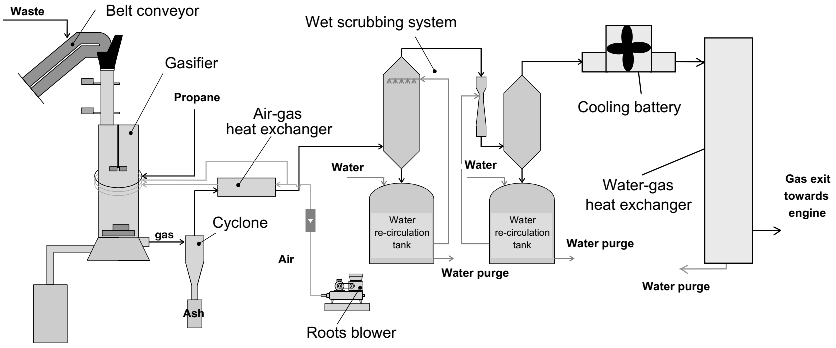


Figure 5.5: Gasification plant scheme (University of Zaragoza)

Results The mixtures of biomass with HDPE could be gasified without problems caused through agglomeration. HDPE is much more reactive than wood, this causes higher temperatures in the oxidation zone and in the drying and reduction zones, further a higher consumption of oxygen. As a consequence the biomass conversion and the energetic yield increases. Another result of using HDPE is the appearance of higher hydrocarbons like C_2H_6 and C_2H_4 and a higher percentage of CO and CH_4. The lower heating value of the producer gas increases from around 3.5 MJ/Nm^3 to 6.1 MJ/Nm^3 for the 10 wt% mixture and to about 6.7 MJ/Nm^3 for the 15 wt% mixture. The high gas production caused operational problems with the installed gas cleaning system and limited the HDPE content to 17 wt%.

5.2.3 Co-gasification of Columbian coal and biomass in a bubbling fluidized bed

Vélez et al., 2009 [16]

Columbia has not only large coal reserves, also a relevant amount of residual biomass, particularly sawdust, rice and coffee husk. Therefore co-gasification of coal with biomass would be an obvious option for efficient usage of fossil fuels and to reduce greenhouse gases. Biomass/coal mixtures with up to 15w% biomass have been investigated for this study.

Experimental setup and test procedure The tests have been carried out in a bubbling fluidized bed gasifier which has an internal diameter of 0.22 m and is 4 m high, shown in fig. 5.6. As gasification agent air/steam mixtures with different ratios were used. Coal and biomass were milled to a particle size between 1.2 and 1.6 mm and provided premixed, fed with a screw feeder. A second feeder for gravity feeding from the top of the reactor is also installed to this gasifier. An overflow pipe maintains a constant bed height and evacuates permanently ashes and bed material. Unburned solids in the producer gas are separated in two cyclones, before the gas is burned in an afterburner, a partial stream is led to the gas analysis unit.

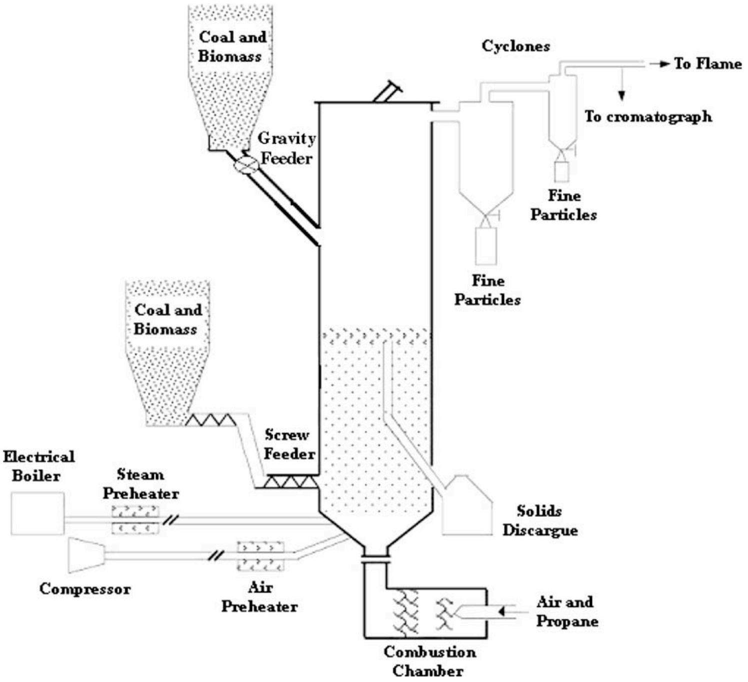


Figure 5.6: Scheme of the gasifier (Universidad Nacional de Colombia)

Results As a result of temperature variation the tests brought up that biomass with high volatile matter (e.g. coffee husk) needs a lower gasification temperature to produce a good combustible gas than biomass with less content of volatile matter like rice husk. The produced gas was, compared to pure coal, rich in hydrogen, and contains also CO, CH_4 and higher hydrocarbons (tars). Cold gas efficiency reached up to 57% using coffee husk or sawdust, and around 45% using rice husk. Higher percentages of biomass, up to 15%, caused decreasing efficiency values, except when using sawdust where the efficiency is constant. In general the reduction of efficiency for a biomass percentage of 6% was minor and should not overshadow the reached reduced fossil CO_2 emissions compared to usage of pure coal.

5.2.4 Co-gasification of coal and olive oil wastes in a bubbling fluidized bed

André et al., 2005 [17]

In Southern Europe olive oil industry produces large volumes of semi-solid (foot cake) or solid waste (bagasse), depending on the pressing process. Disposal of these wastes cause considerable environmental problems due effluent discharges, which consist of oil and water phases. Until now waste water treatment, like filtration, reverse osmosis, chemical methods or thermal concentration, is inefficient or too expensive. On the other hand large amounts of low grade coal are available in Spain, which are difficult to gasify due the ash and sulphur content. Using a mixture of both provides the possibility to combine the advantages of both fuels and to reduce their particular disadvantages.

Experimental setup and test procedure The experimental setup (fig. 5.7) consists of a fluidized bed reactor with an internal diameter of 70 mm and 500 mm total hight, and a cyclone to remove discharged particles. As bed material silica sand or dolomite with an average particle size of 350 µm was used, the gasifying agent was an air/steam mixture. The feedstock, containing up to 70 wt% of bagasse, was fed from the top of the reactor and gasified at temperatures between 730 and 900 °C. The producer gas was led through a quenching system to condensate tars and other liquids before it was analyzed by a gas chromatograph.

Results The amount of biomass has a significant influence to the gas composition, a higher percentage of bagasse causes decreasing H_2 and increasing CO content while the variation of CO_2 concentration is quite small. Higher biomass ratio led to greater gas yields and a slightly higher caloric value of the producer gas, but it caused also problems in gasification stabilisation and promoted higher amounts of tars and heavier hydrocarbons. Usage of dolomite as catalytic bed material brought a reduction in tar and hydrocarbon content, the same effect was observed

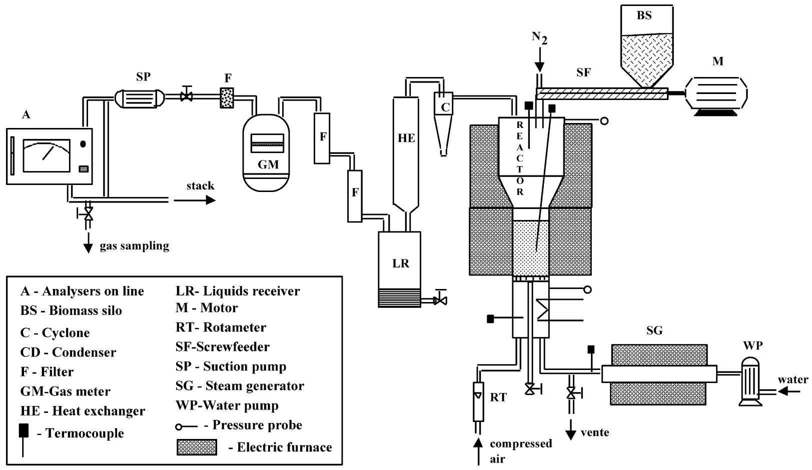


Figure 5.7: Schematic diagram of the gasification system (National Institute of Engineering, Technology and Innovation, Portugal)

at higher gasification temperatures and air flow rates, respectively higher air/steam ratio. Pure steam as gasification agent showed a higher hydrogen and hydrocarbon concentration, pure air gasification led to a great increase of CO_2, caused through favoured oxidation reactions. These oxidation reactions provide heat supply for the gasification reactions compared to the steam gasification process where additional heat supply is necessary.

5.2.5 Co-gasification of plastic waste with coal and biomass in a bubbling fluidized bed

Aznar et al., 2006 [18]

Waste reduction and recycling is one of the most important targets of todays environmental politics. Due to the high amount of produced goods plastic waste is very important in this process. Mechanical recycling of plastic waste is only feasible for products formed by one type or special mixtures of plastics. Instead of mechanical methods several chemical processes, like liquefaction, cracking or gasification, are used to break down the polymers to light hydrocarbons or hydrogen for usage as burning gas or raw material. It is imaginable that technologies developed for coal or biomass gasification will be used for plastic waste and plastic waste is used for co-gasification with one of the established fuels.

Experimental setup and test procedure The gasification reactor, shown in fig. 5.8, consists of a 1 m high bed zone with an internal diameter of 92 mm and an also 1 m high freeboard with 154 mm internal diameter. As bed material a mixture of silica sand and dolomite as tar cracking catalyst was used. Plastic waste, consisting of 50 wt% polyethylene and 50 wt% polypropylene, premixed with Spanish coal or pine sawdust was used as feedstock and introduced into the bed zone of the reactor through a screw feeder. Air was used as gasification agent, equivalence ratio has been varied between 0.30 and 0.46 as well as gasification temperature between 750 and 880 °C. The gas sampling point is located after two cyclones for particle removal, analyses of composition and tar content of the producer gas have been undertaken.

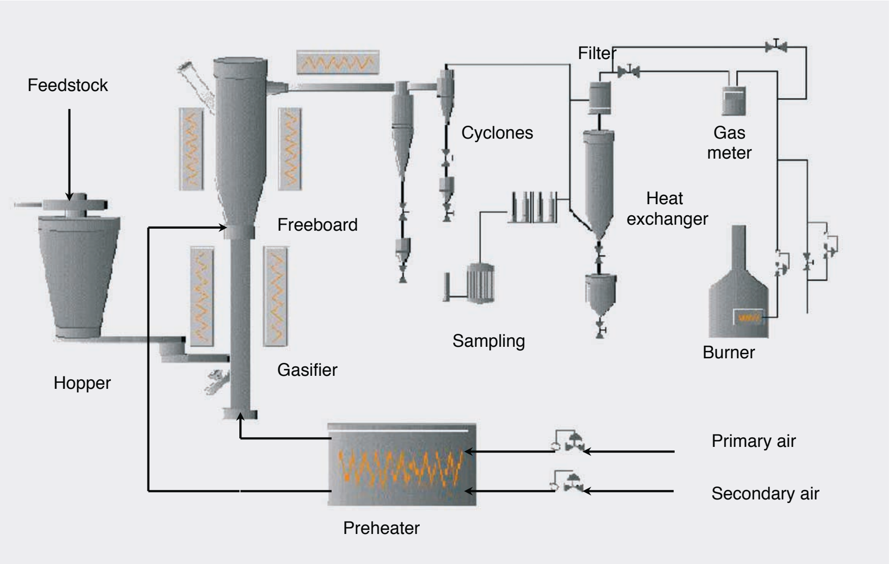


Figure 5.8: Scheme of the reactor (University of Zaragoza)

Results Increasing plastic waste concentration in the feedstock causes higher concentrations of methane and light hydrocarbons and accordingly a higher heating value and fewer tars, but also a lower gas yield. The effect of temperature variation showed increasing heating value and rising gas yield at higher temperatures, combined with a decrease in tar content. Rising equivalence ratio brought a reduced char yield and due to dilution with nitrogen reduced concentrations of H_2, CO, CO_2, CH_4, C_2H_n and a reduced heating value while the tar content and the total gas yield was nearly constant. Optimal working conditions where located at a bed temperature of 850 °C with an equivalent ratio of 0.36, independent of the feedstock mixture. The main problem of the process was the tar content in the producer gas, which could be lowered by introduction of secondary air (10% of total air volume) into the freeboard by 50%, but the effected reduction of the heating value was up to 20%.

5.2.6 High pressure co-gasification of coal and petroleum coke with biomass wastes in a fixed bed downdraft gasifier

Fermoso et al., 2009 [19]

Gasification of coal, petroleum coke and wood are well established technologies and widely used. Co-gasification of these fuels could provide several improvements in producer gas quality like reduction of sulfur compounds generated from coal or reduced tar content, a common problem at wood fired plants. Another advantage is avoiding problems due seasonal differences of the feedstock, especially if biomass wastes are used. In the current studies coal and petcoke were co-fired with three different types of biomass, almond shells, olive stones and eucalyptus.

Experimental setup and test procedure An electrical heated fixed bed downdraft gasifier(fig. 5.9) was used to gasify the premixed, ground and sieved (particle size between 75 and 150 µm) feedstock. The reaction tube has an internal diameter of 13 mm and is 305 mm high. As gasification agent mixtures of nitrogen, oxygen and steam where used, the experiments where carried out at gasification temperatures from 850 up to 1000 °C and a pressure between 0.5 and 2 MPa. The formed tars and excess steam were separated from the producer gas in a gas cooler before the gas composition was analysed in an online micro-GC.

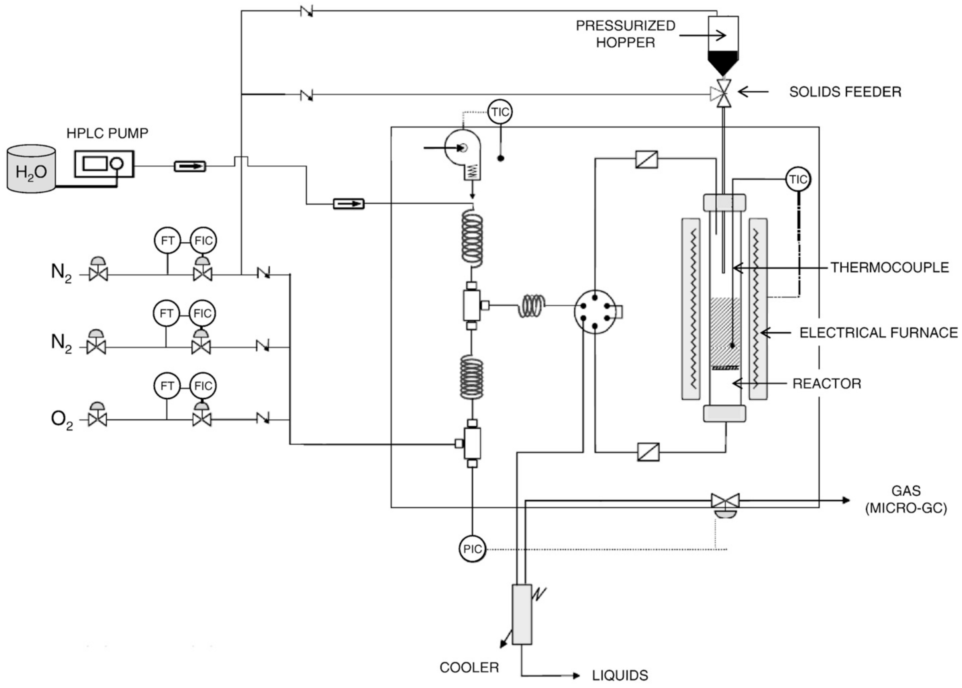


Figure 5.9: Experimental device (Instituto Nacional del Carbón)

Results The effect of variation of operational variables on gas production, cold gas efficiency and high heating value was analysed as well as the effect of the different fuel blends. The results showed a great dependency on gasification temperature and oxygen concentration, while the gasification pressure had practically no influence. Higher gasification temperatures caused an increase in H_2 and CO concentration, while rising O_2 concentration in the gasifying agent led to increasing CO_2 concentration, due to coal combustion. Increasing steam concentration results in higher H_2 and CO_2 concentrations, the CO value remains nearly constant. By addition of biomass (up to 10%) to coal higher H_2 and CO production of was observed and carbon conversion and cold gas efficiency improve considerably. Ternary blends of coal (45%), petroleum coke (45%) and biomass (10%) brought only slight variations in CO and CO_2 concentration while H_2 production was almost constant.

5.3 Critical comments and conclusions about the reviewed literature

The presented co-firing systems for combustion or gasification can be related to two groups. Most of them have been designed for coal, only two systems where designed for biomass fuels. In some plants it was possible to run co-fuel percentages up to 100%, on the other hand some could only use 10 to 17% of other fuels, but mainly co-firing ratios between 40 and 70% were realized (table. 5.1). Additional to these facilities the gasification pilot plant at the Vienna University of Technology has to be mentioned, where also lots of different fuels with ratios up to 100% have been tested [20, 21]. Most of the pilot plants and lab scale units in this review are equipped with electrical heat support which reflects in high fuel flexibility.

In industrial coal-fired power plants co-firing is often used to add a green fingerprint without any loss in efficiency and only minor changes in plant settings, therefore only low percentages of other fuels are used. But as some of the reviewed literature shows, co-firing of even higher percentages of biomass or refuse derived fuels is possible and should be further investigated. Nevertheless, the reactor has to be designed for flexible usage of fuels and appropriate gas cleaning systems are necessary. Therefore existing facilities are often unsuitable and have to be redesigned, but to reach real reductions of fossil CO_2 emissions this investment is essential. Otherwise most of the investigated fuels get burned in waste incineration plants or get composted, in some countries still they get land-filled, and the energy content remains unused.

Researcher	Main fuel	Additional fuels (max. ratio)
Combustion plants		
Gahni et al.	coal	rice husk, palm kernel shell (100 en%)
Aho et al.	coal	cynara (50 en%)
Leckner et al.	coal, wood	sewage sludge (50 en%)
Gasification plants		
Lapuerta et al.	coal	grapevine pruning waste, pine pruning waste, marc of grape, sawdust (100 en%)
García-Bacaicoa et al.	wood	HDPE (17 wt%)
Vélez et al.	coal	sawdust, rice husk, coffee husk (15 wt%)
André et al.	coal	olive oil waste (70 wt%)
Aznar et al.	coal	plastic waste, wood (40 wt%)
Fermoso et al.	coal	petroleum coke (60 wt%), wood (10 wt%)
Hofbauer et al.	wood	lignite, coal, trefoil pellets, straw pellets, sewage sludge (100 en%)

Table 5.1: Reviewed plants

6 Experiments on co-gasification of dried sewage sludge with wood

6.1 Description of the 100 kW pilot-plant in Christchurch, New Zealand

The experimental work was carried out in a "Dual fluidized bed (DFB) steam gasifier", located at the Department of Chemical and Process Engineering of the University of Canterbury, Christchurch, New Zealand, designed in co-operation with the Vienna University of Technology. A schematic diagram of the installation is shown in fig. 6.1. The reactor is an allo-thermal reactor with a circulating bed material and therefore separated into two reaction zones, the gasification column and the combustion column. The gasification section has an internal diameter of 207 mm, a height of 2 m and is built as a bubbling fluidized bed (BFB), while the gasification section has an internal diameter 107 mm, is 3.7 m high and works as a circulating fluidized bed (CFB). As bed material graywacke sand or a mixture of olivine and calcite with a particle size between 200 and 1000 µm was used. Possible operation conditions for the gasifier are gasification temperatures between 700 and 900 °C and atmospheric pressure.

The feedstock is transported premixed with a screw feeder from a main hopper into an intermediate hopper and than with a second screw feeder directly into the bed. To avoid pyrolysis in the second feeding screw, the feeder is water cooled.

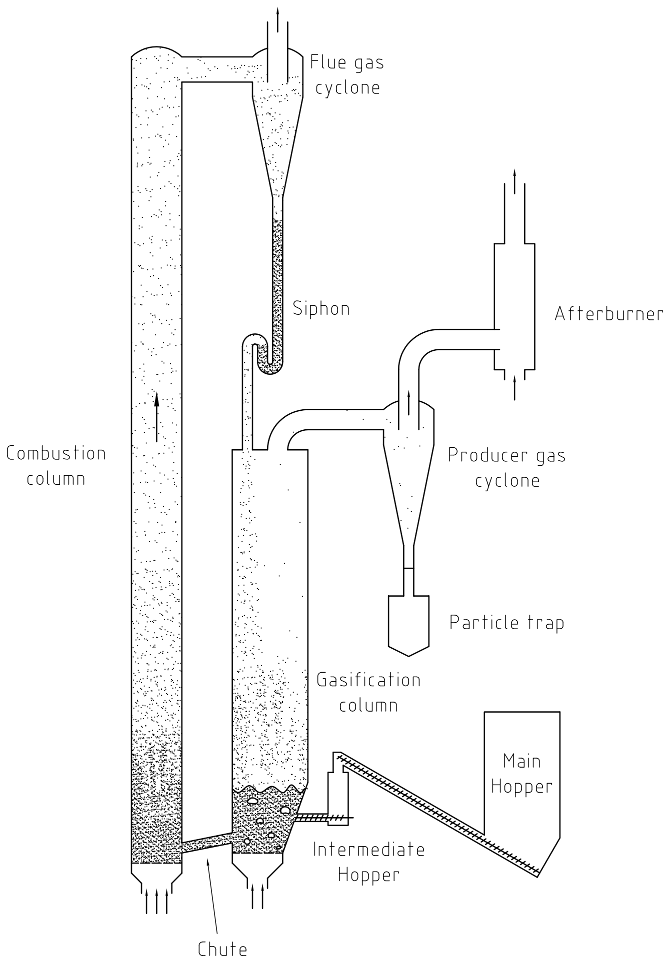


Figure 6.1: DFB Gasifier

The whole feeding system is sealed and nitrogen at a slightly positive pressure is fed into the hoppers to prevent backflow of producer gas from the gasification column. For longtime runs the main hopper is equipped with an airlock, which allows refilling the hopper without a pressure loss (fig. 6.2). The feeding system is designed for wood pellets and allows also the usage of granulated fuels, like wood or plastic chips, with diameters up to 10 mm.

The gasification column is made of steel and is refractory lined, it has a cone shaped base where a chute allows the bed material and char to circulated to the

Figure 6.2: Main hopper with airlock

combustion column. Steam at 400 °C, 10 kPa, is used as fluidization medium and oxidizing agent in the gasification column and introduced through a nozzle plate into the reactor. The chute is also fluidized with steam to inhibit an air leakage from the combustion column. The producer gas rises upwards through the freeboard and leaves the reactor at the top. There it is separated in a cyclone from small entrained bed material and char particles, then led to the sampling point and afterwards it is combusted in the afterburner.

Like the gasification column the lower part of the combustion column is also made of steel and refractory lined, while the upper part is made of a high temperature resistant stainless steel tube which is insulated with Kaowool blanket (fig. 6.3). In this column the bed material is reheated through combustion of char and additional LPG. Fluidization air and LPG is introduced to the bottom section through a nozzle plate and fluidizes the bed material coming from the chute at low gas velocities, behaving like a bubbling bed. Primary air is added 250 mm above the nozzle plate and secondary air another 200 mm above the primary air inlets.

In this section the gas velocity increases above the transport velocity and the hot particles are entrained to the top of the reactor. The operation temperature of the gasification column, and furthermore of the gasification column, is controlled by the LPG gas flow, the amount of char and the circulation rate of the bed material.

At the top of the combustion column the entrained bed material is led to a cyclone, where it is separated from the flue gas and than it gets through the siphon back into the gasification column. The siphon is fluidized with steam and prevents gas exchange between the two columns while it allows the bed material to circulate. The flue gas passes a sampling point, is led to a heat exchanger to preheat the combustion air and leaves the plant together with the flue gas from the afterburner.

Figure 6.3: Gasifier without insulation [22]

General startup and gasification procedure During the startup procedure of the gasifier both columns are fluidized with air and a second LPG burner in the gasification column is activated, additional to the burner in the combustion column. The bed material circulates between the two columns and provides a smooth temperature gradient and an equal temperature distribution. As soon as the bed material reaches the planed gasification temperature, fluidization of BFB, chute and siphon is changed to steam and the burner in the gasification column is shut off. Then the feeding system is launched and the gasification reactions start. As soon as the temperatures remain constant at the planned values steady-state is reached and the measurements can be taken.

Measuring equipment Gas samples for analyses are taken from the producer gas and the flue gas simultaneously every half hour and analysed in a dual channel Agilent 3000A Micro Gas Chromatograph (Micro-GC). The producer gas is analysed for He, H_2, N_2, CO, CO_2, CH_4, C_2H_4 and C_2H_6, the flue gas for He, O_2, N_2 and CO_2. Tars in the producer gas are collected with SPE columns and externally analysed in a gas chromatography-mass spectrometry (GC-MS), moisture content is measured by adsorption on silica gel.

6.2 Properties of the tested fuels

For the current experiments different mixtures of wood pellets and dried sewage sludge were prepared. The wood pellets are made of sawdust and waste wood, mainly from radiata pine, manufactured from a local factory in Christchurch. The sewage sludge (DSS) comes from a sewage treatment plant in Auckland where it has been dried to a water content less than 8 wt% and granulated. Table 6.1 shows the proximate and ultimate analysis for both fuels and their main physical parameters, the full analysis of both fuels are attached in appendix A and B.

Analysis	Wood pellets	Sewage sludge
proximate (wt%)		
moisture	8.0	81.7
volatile matter	77.4	n.a.
fixed carbon	14.2	n.a.
ash	0.4	5.4
ultimate (wt% dry basis)		
C	51.3	34
H	5.8	5
N	< 0.2	5
S	0.01	1.2
Cl	-	0.04
O	42.4	25
LHV (MJ/kg dry basis)	18.6	12.1
ash softening temperature (°C)	1180	1180

Table 6.1: Fuel characterization

6.3 Experimental procedure

The Experiments were carried out at a constant gasification temperature of 750 °C with a total feed mass flow of 18 kg/h and a steam mass flow of 16 kg/h, equivalent to a steam fuel ratio (sfr) of 0.89. It was chosen to set the feed mass flow constant to reach equal fluidization conditions and a constant sfr. For the main series the ratio between wood pellets and DSS was varied between 0 and 100 wt% and graywacke sand was used as bed material. Additionally two runs with 0 wt% and 20 wt% of DSS with a mixture of 75 wt% olivine and 25 wt% calcite as bed material were undertaken at a gasification temperature of 700 °C.

In a pre-test pellets from sawdust and granulated sewage sludge in the planed ratios were pressed to evaluate the risk of bridging in the hoppers or blockage of the augers due to the stickiness of the sewage sludge. It was found that an enduring bondage appeared only at very high pressure (more than 10 MPa), so the experiments could be carried out without changes to the regular feeding system. Fig. 6.4 shows some of these pressed pellets before they where crushed.

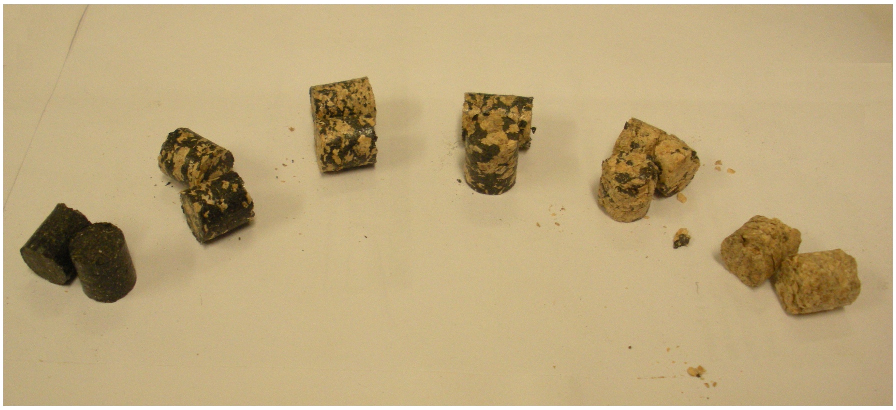


Figure 6.4: Pressed pellets from sawdust and sewage sludge

6.4 Results and discussion

The effect of the ratio DSS to wood pellets on producer gas composition and gasification performance was studied. All gas composition values are calculated for a dry, ash free basis, dilution with nitrogen upcoming from the hopper inertization was corrected. Producer gas yield was determined from the helium concentration which was used as tracer, the lower heating value was calculated with the heating values of the measured components.

Fig. 6.5 illustrates the producer gas composition for the seven runs using graywacke sand as bed material. It is clearly visible that the nitrogen content increases significantly with higher DSS ratios. Also the other components were effected from the amount of DSS, CO and CO_2 drop while H_2 rises slightly, CH_4, C_2H_4 and C_2H_6 remain nearly constant with just a marginal tendency down.

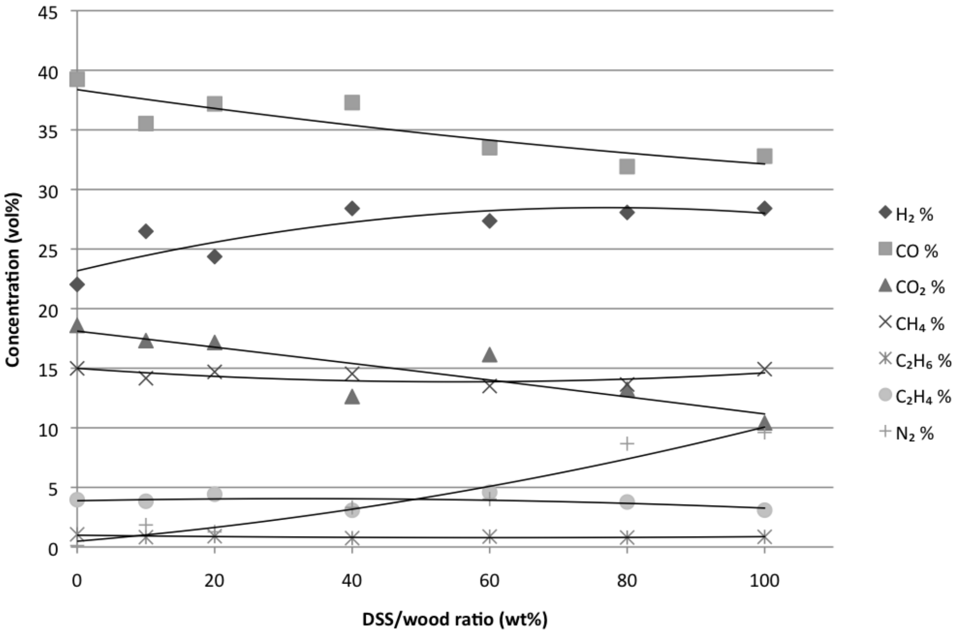


Figure 6.5: Effect of DSS ratio on producer gas composition

The increasing nitrogen concentration is caused through the nitrogen content of the sewage sludge which is about 5 wt% dry weight and about twenty times higher than the content in wood pellets. In the same way the reduced CO/CO_2 content can be explained. The carbon concentration in DSS is 34 wt% dry weight compared to 51 wt% in wood pellets. As the hydrogen concentration in these two fuels is only slightly different, the hydrogen concentration in the product gas is approximately constant. Another clearly visible trend is that the $(CO+CO_2)/H_2$ ratio in the produced gas is decreasing with lower C/H ratios of the feed as they appear due to increasing DSS percentages (fig. 6.6).

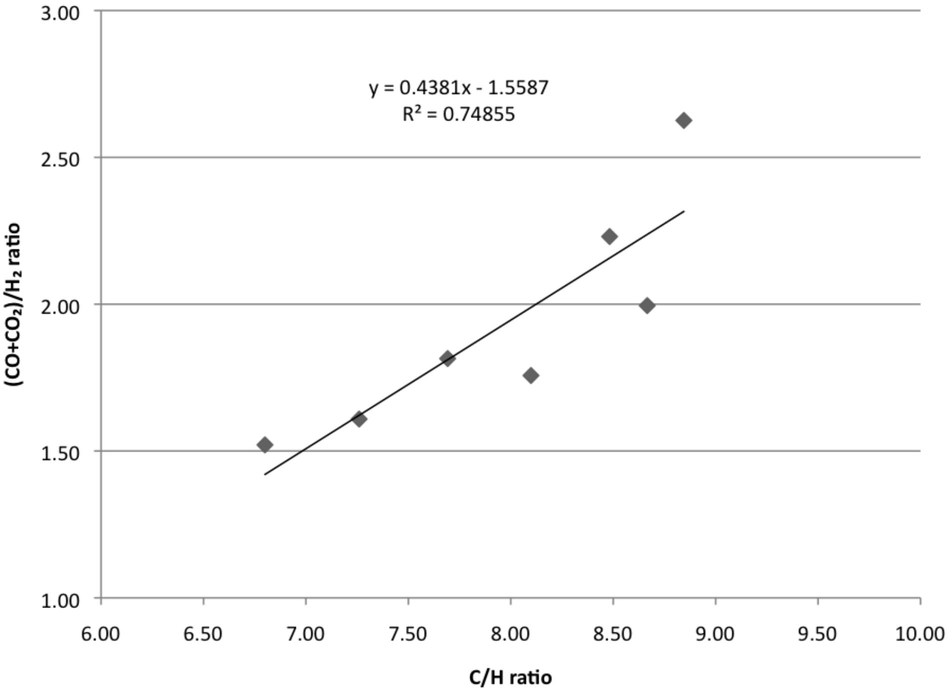


Figure 6.6: Effect of C/H ratio (feed) on the $(CO+CO_2)/H_2$ ratio (product gas)

Gasification of sewage sludge is more energy consumptive due to the high nitrogen and ash content, proportional to the sewage sludge percentage more supplementary LPG was needed to reach the intended gasification temperature. Therefore the cold gas efficiency (CGE, defintion in appendix C) decreases with higher DSS ratios, although the heating value of the produced gas is nearly equal (fig. 6.7).

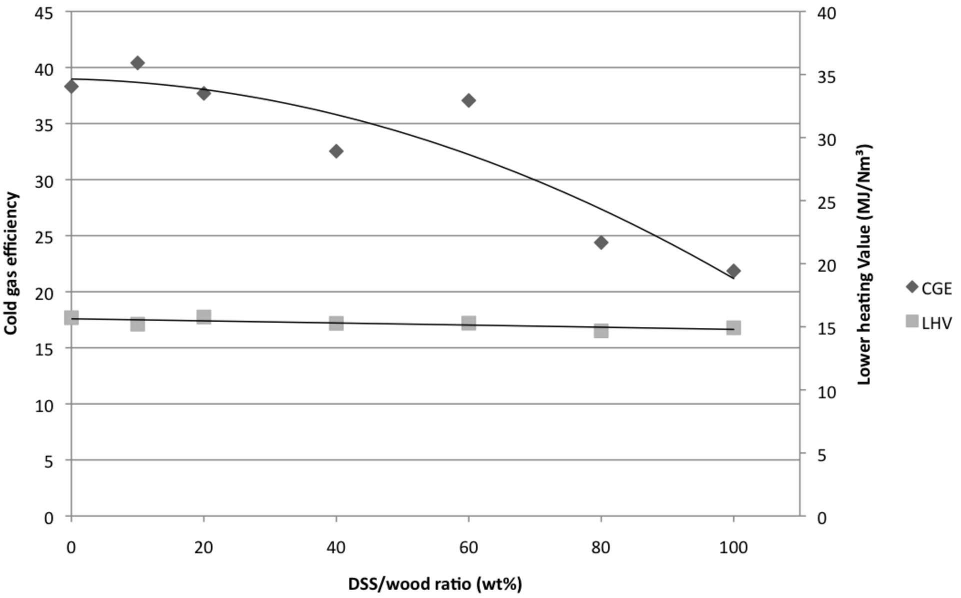


Figure 6.7: Effect of DSS ratio on CGE and LHV

The slight decrease of the LHV is caused through the increasing amount of nitrogen in the producer gas, but as the concentration of methane (which has the biggest share of the total heating value) remains constant, the effect is only little. The lower heating value is calculated with eq. (6.1), based on the heating values of the single components (table 6.2), regarding to DIN 51857 [23].

$$\begin{aligned} LHV_{prod.gas} &= LHV_{H_2} \cdot C_{H_2} + LHV_{CO} \cdot C_{CO} + LHV_{CH_4} \cdot C_{CH_4} + \\ &+ LHV_{C_2H_4} \cdot C_{C_2H_4} + LHV_{C_2H_6} \cdot C_{C_2H_6} \end{aligned} \tag{6.1}$$

	H_2	CO	CO_2	CH_4	C_2H_4	C_2H_6	N_2
LHV (MJ/Nm^3)	10.79	12.63	0	35.81	59.03	63.74	0
HHV (MJ/Nm^3)	12.75	12.63	0	39.74	62.96	69.63	0

Table 6.2: Heats of combustion of typical producer gas components [23]

The high ash content also caused some problems in the product gas cleaning system, as the particle trap is not designed for such high ash loads. At DSS ratios above 40 wt% the ash fills up the container within 2-3 hours and the ash starts flowing into the afterburner where it covers the walls and the pilot burner. A continuous ash removing system has to be installed to guarantee the stabilisation of the gasification process and to prevent unexpected shutdowns before long time runs can be carried out. Except the problems with the high ash build up, gasification and co-gasification of dried sewage sludge performed well. Detailed figures of the results are shown in table 6.3, the complete data sheets of these runs can be found in appendix D.

DSS	Producer gas (vol%)							LHV	CGE
wt%	H_2	CO	CO_2	CH_4	C_2H_6	C_2H_4	N_2	MJ/Nm^3	-
0	22.0	39.2	18.6	15.0	1.1	4.0	0.1	15.7	38.3
10	26.5	35.5	17.3	14.2	0.8	3.8	1.8	15.2	40.4
20	24.4	37.2	17.2	14.7	0.9	4.4	1.3	15.8	37.7
40	28.4	37.3	12.6	14.5	0.7	3.1	3.3	15.3	32.5
60	27.4	33.5	16.1	13.5	0.9	4.6	4.0	15.3	37.1
80	28.1	31.9	13.2	13.6	0.8	3.8	8.7	14.7	24.4
100	28.4	32.8	10.4	14.9	0.9	3.1	9.6	14.9	21.9

Table 6.3: Fuel ratio, producer gas composition, LHV and CGE

Additional runs with olivine and calcite as bed material Olivine and calcite are catalytically active bed materials used for tar elimination in the producer gas. Another observed effect of using olivine as bed material is a higher hydrogen content of the producer gas. In preliminary tests with wood pellets a mixture of 75 wt% of olivine and 25 wt% showed the best performance at this reactor. As expected the producer gas showed a higher hydrogen content while the lower heating value was less effected. Using 20 wt% of DSS in the feedstock caused a unexpected serious drop of the cold gas efficiency, due to the need of extreme high LPG supply and a significant decrease of the product gas yield. Also H_2 and CO_2 content in the produced gas dropped, while CO and CH_4 raised several percent (fig. 6.8). The recognized effects have to be invesigated in future studies, as there was currently no obvious explanation found. Results from these two runs are summarized in table 6.4, the complete data sheets from the olivine/calcite runs are attached in appendix E.

DSS	Producer gas (vol%)							LHV	CGE
wt%	H_2	CO	CO_2	CH_4	C_2H_6	C_2H_4	N_2	MJ/Nm^3	-
0	30.5	28.0	19.5	11.4	0.7	3.0	7.0	13.1	54.3
20	28.2	38.3	10.6	14.5	0.7	2.1	5.6	14.8	26.0

Table 6.4: Fuel ratio, producer gas composition, LHV and CGE (olivine/calcite)

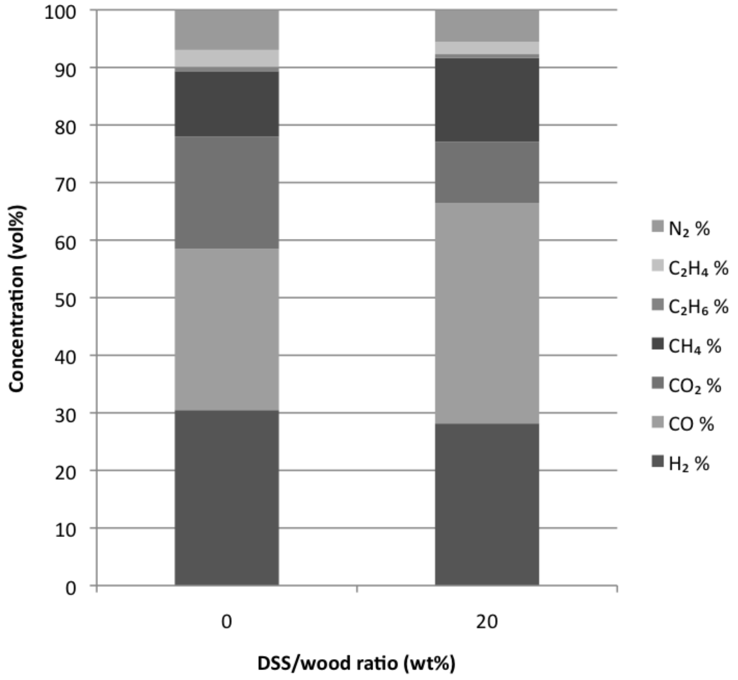


Figure 6.8: Effect of DSS ratio on producer gas composition (olivine/calcite)

Summary Both bed materials allowed stable gasification conditions. The performance of the fuel mixtures in long time runs is a topic for future studies as well as the unexpected differences in the gasification efficiency and the producer gas quality when using olivine. Another aspect which was not possible to investigate in this project is formation of polychlorinated biphenyls (PCBs) and polychlorinated dibenzo-p-dioxine/-furans (PCDD/Fs) due to the presence of chlorine in the sewage sludge.

6.5 Comparison with experiments at the Vienna Univerisity of Technology

6.5.1 Description of the 100 kW pilot plant I

At the Vienna University of Technology first comparable tests with 100% dried sewage sludge and 100% wood pellets were carried out in 1998, using silica sand and olivine as bed material. For these tests the former pilot plant I (fig. 6.9) was used. The pilot plant was built as a fast internally circulating fluidized bed (FICFB) gasifier with a bubbling bed in the gasification zone and a fast bed in the combustion zone. The gasification process is designed as allo-thermal steam gasifier and the circulating bed material is used as heat carrier.

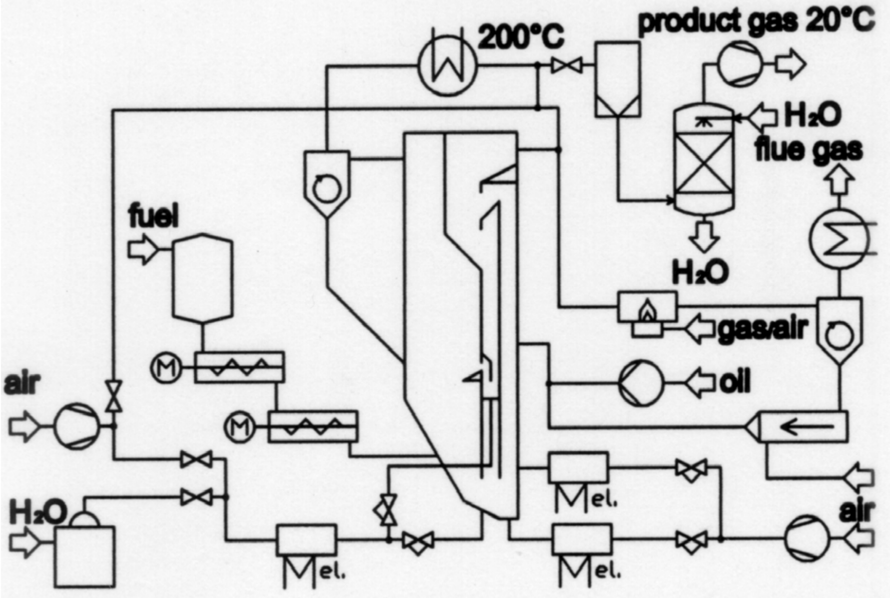


Figure 6.9: Flow sheet of the pilot plant I [20]

A FICFB reactor allows a very compact design and due to heat exchange between the different rector parts a very uniform temperature distribution. The casing is made of steel and outside insulated with glass wool. The required fluidization and combustion air is provided from two blowers while the steam is produced in an electrical heated boiler.

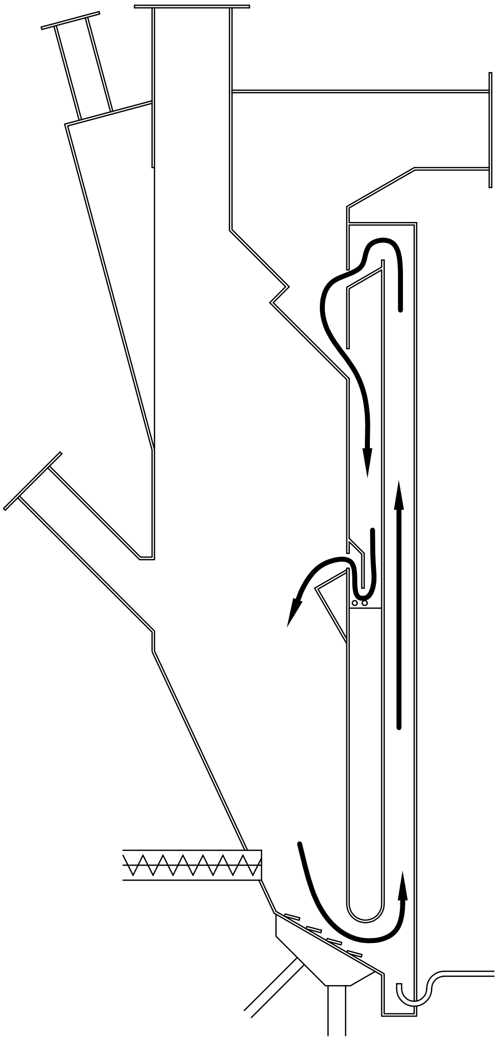


Figure 6.10: FICFB reactor. Bed material circulation flow

Usual bed materials for this reactor are silica sand as inert bed material or olivine as catalytically active bed material. The hot bed material gets continuously through the siphon into the steam fluidized gasification reactor. At the bottom of the gasification zone cold bed material flows over the downwards angled distributer plate to the combustion zone. There the remaining char and also additional oil is burned in the air fluidized riser to reheat the bed material. At the top of the riser the hot bed material is separated from the flue gas and returns to the steam fluidized siphon (fig. 6.10).

The feedstock is prepared in a hopper and introduced with two screw feeders into the bed zone of the gasification zone. The first feeder controls the mass flow of the biomass while the second feeder runs at constant high speed to push the fuel into the bed where it is gasified at 700 to 900 °C. The produced gas leaves the reactor at the top and is led through a cyclone where is separated form small char and bed material particles. The particles return to the gasification zone while the clean producer gas is cooled down to 200 °C in an air/air heat exchanger. Afterwards the gas is led together with the flue gas to the afterburner where additional natural gas guarantees the total combustion of all components. A second cyclone collects entrained particles from the flue gas, which are returned through an injector to the riser. A part of the product gas stream can be separated directly after the heat exchanger and after further cooling and cleaning used for analysis or other applications.

For the heating up procedure the whole reactor is also fluidized with air, the biomass is also burned in the gasification zone until the intended gasification temperature is reached. Then fluidization of siphon and gasification zone is changed to steam, the process reaches steady state and is running as described above. During the gasification process the reactor temperature is controlled through the oil feed to the combustion zone or the circulation rate of the bed material.

6.5.2 Description of the 100 kW pilot plant III

Other comparable tests have been carried out in 2007 at the actual pilot plant III and this time olivine was us as bed material. As the previous plant II, the pilot plant III is built as a dual fluidized bed gasifier with separated gasification and combustion zones. The reactor principle is again an allo-thermal gasifier with circulating bed material. It was built as further development of the reactor technology using most of the peripheral equipment of the pilot plant II.

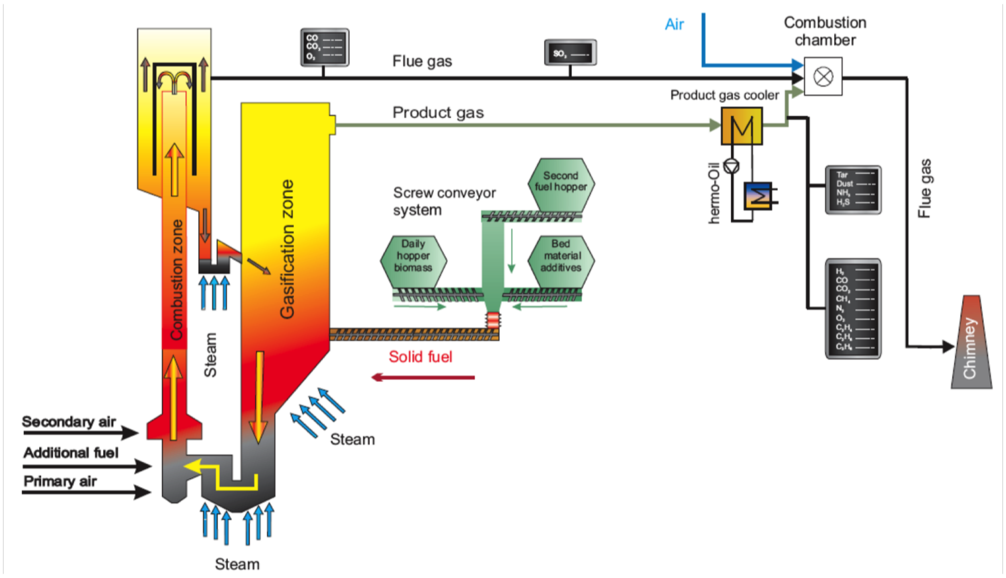


Figure 6.11: Flow sheet of the pilot plant III

The gasification column has a square cross section with a side length of 285 mm and is 3000 mm high. It is made of made of a high temperature resistant steel and outside insulated with glass wool. The gasification section is fluidized at moderated velocity to generate a bubbling bed. Fluidization steam is introduced to the the reactor in the bottom section, which is cone shaped and opens into the bottom siphon, also fluidized with steam and allowing bed material circulation to the combustion column. This siphon allows to minimize nitrogen cross flow from the combustion column to practically zero and therefore the nitrogen content of the produced gas is reduced to the nitrogen generated from the fuel. Typical gasification conditions are 700 to 900 °C at ambient pressure.

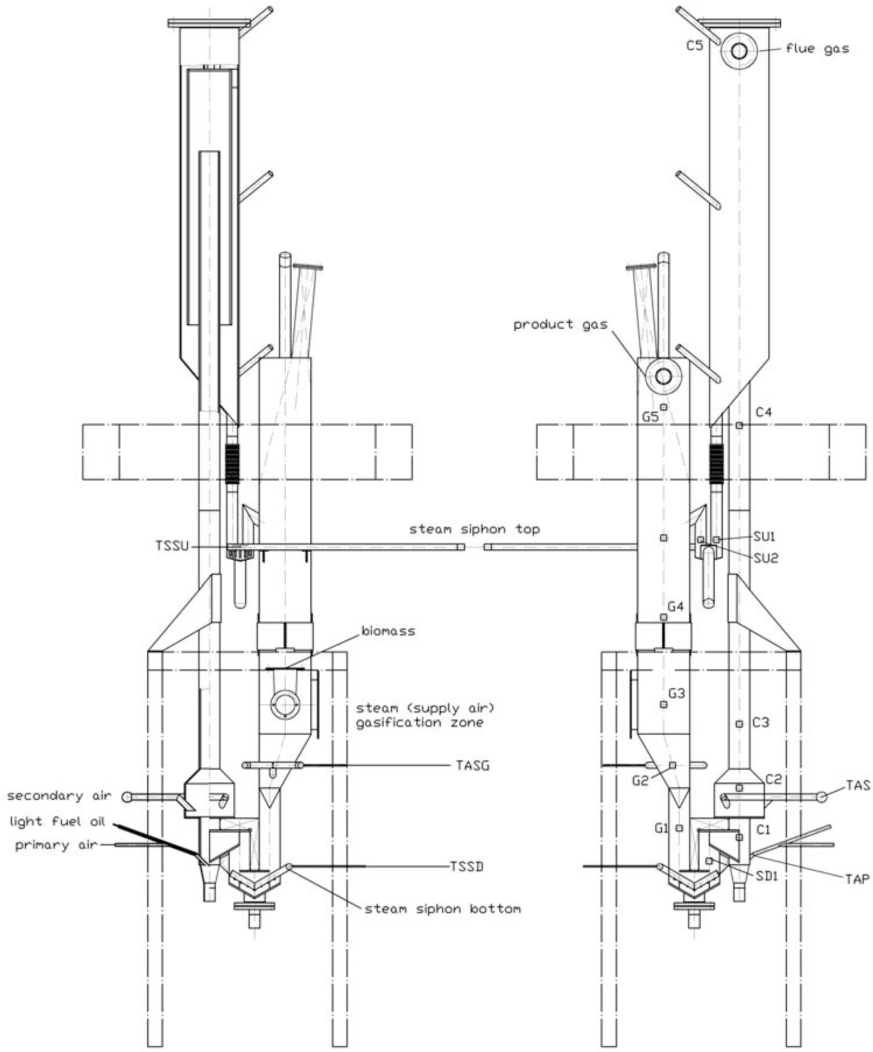


Figure 6.12: Dual fluidized bed reactor

The combustion column has a round cross section with a diameter of 105 mm and is 4800 mm high. It contains a fast fluidized bed, using air as fluidization and combustion agent to burn additional fuel (light fuel oil) and residual char from the gasification column. The bottom siphon is connected to the combustion column between the primary and secondary air inlets. Together with the primary air the additional fuel is fed into the column. The secondary air accelerates the particles to the terminal velocity and carries them out of the reactor. In the top section the reheated bed material is separated from the flue gas stream and returned through the top siphon to the gasification column.

For the feedstock three hoppers in different sizes are available, which allow co-firing of different fuels at any desired ratio. Usually the main hopper is used for wood pellets and additional fuels are stored in the medium hopper. The small hopper could be used for bed material additives or also for additional fuel. Each hopper is equipped with a screw feeder with variable speed and which lead to a central mixing chamber. From the mixing chamber a screw feeder with constant high speed feeds the mixture into the bed zone of the gasifier.

The produced gas leaves the reactor at the top of the gasification column where it is cooled down to 150 to 300 °C. Afterwards it is introduced together with the flue gas and air to the combustion chamber where a natural gas pilot burner guarantees complete combustion. The combustion chamber is built as the entrance duct of a cyclone which cleans the flue gas from ash and bed material dust, entrained from both columns.

For gas analysis part streams of the producer gas and the flue gas are separated and led to the measurement system. The producer gas composition is analysed in an online gas analyser for H_2, CO, CO_2, CH_4 and O_2 and in an online GC with 20 min cycle time for N_2, C_2H_4 and C3/C4 components and also CO, CO_2, CH_4 and O_2. Furthermore samples for analytical determination of the tar, NH_3 and SO_2 load are taken. The flue gas is analysed in a second online gas analyser for O_2, CO and CO_2.

6.5.3 Discussion of the results from pilot plant I

In this experiments wood chips and granulated dried sewage sludge where used. The composition of the fuels is shown in table 6.5. Olivine and silica sand was used as bed material during this series, the gasification temperature was varied between 770 to 850 °C. Detailed results of this runs are described in [20], therefore only a short overview is given here.

Analysis	Wood	Dried sewage sludge
proximate (wt%)		
moisture	14.3	13.0
ash	1.3	40.1
ultimate (wt% dry basis)		
C	50.6	28.3
H	6.02	4.07
N	0.35	2.96
S	0.03	1.25
Cl	0.02	0.07
O (by difference)	41.65	23.3
HHV (MJ/kg dry basis)	20.3	12.1
LHV (MJ/kg dry basis)	19.2	10.6
ash softening temperature (°C)	1210	1120

Table 6.5: Fuel characterization

Fig. 6.13 shows the producer gas composition for the runs with silica sand as bed material. Compared to wood the CO content in case of gasifying sewage sludge is significantly smaller while CO_2 increases in the same ratio. The differences between the concentrations of the other components are rather small, basically H_2

is higher but it is slightly decreasing with higher temperatures. The reduced CH_4 percentage is nearly counterbalanced by C_2H_4. Caused through an air crossflow from the combustion zone the nitrogen content is about 6 vol% when using wood and increases to 8 vol% when firing sewage sludge, due to the higher nitrogen content of the feed.

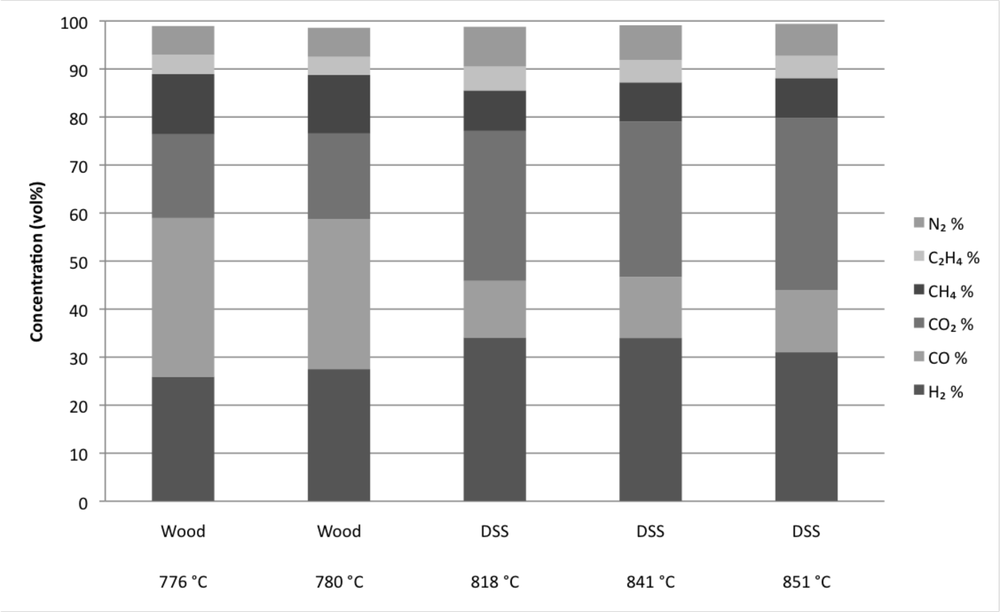


Figure 6.13: Effect of the used fuel on producer gas composition (silica sand)

The results from the second series, using olivine as bed material are displayed in fig. 6.14. Similar to the first series the CO concentration when using sewage sludge is approximately one third of the concentration when firing wood. On the other hand the CO_2 and H_2 contents are higher, with further increasing H_2 at higher temperatures. The CH_4 content is reduced, while C_2H_4 tends to slightly higher concentrations. Compared to the runs using silica sand bed material the concentrations of H_2 and CO_2 are at higher levels while the CO content is lower, the influence on CH_4 and C_2H_4 is marginal.

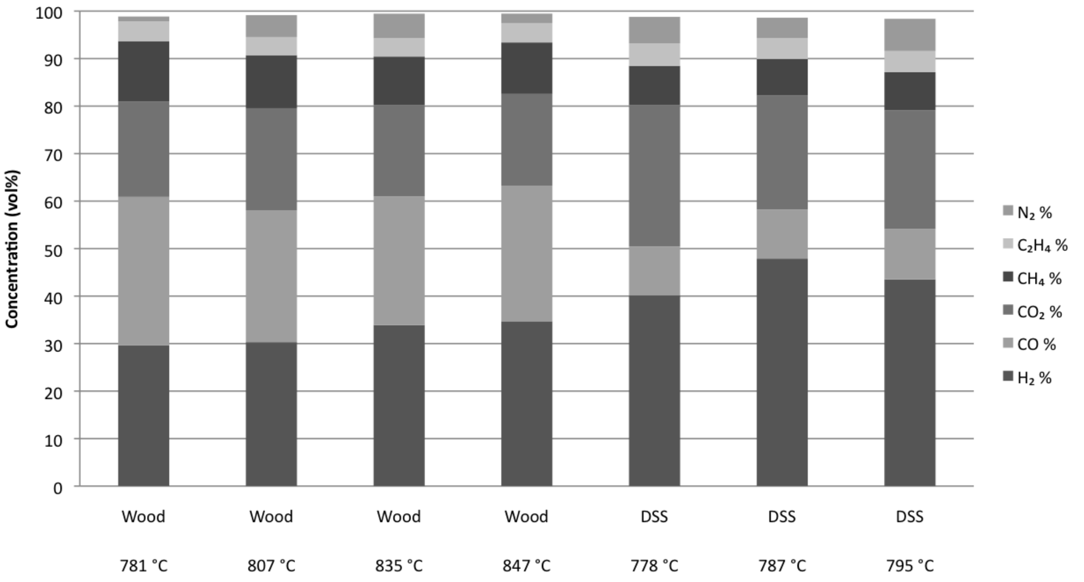


Figure 6.14: Effect of the used fuel on producer gas composition (olivine)

As the plant was designed for wood gasification, those runs using wood as feed could be carried out without troubles, but even the runs gasifying sewage sludge performed well. The high ash concentration led to adherences at different parts of the reactor and to high fly ash load of the producer gas, which could only be separated in the flue gas cyclone as the producer gas particle separation system was overloaded. As the stickiness of the ash was low, the adherences could be removed during regular maintenance without any great effort. The lower heating value of the generated gas from sewage sludge is lower than from wood. For both fuels the heating values where higher if olivine was used as bed material than if silica sand was used. All figures of these results are summarized in table 6.6.

DSS	Bed temp.	Producer gas (vol%)						LHV
wt%	°C	H_2	CO	CO_2	CH_4	C_2H_4	N_2	MJ/Nm^3
runs with silica sand bed material								
0	776	25.93	33.05	17.41	12.54	4.04	5.98	14.72
0	780	27.49	31.20	17.87	12.20	3.76	6.06	14.71
100	818	34.09	11.78	31.24	8.42	4.97	8.29	12.18
100	841	33.97	12.71	32.34	8.19	4.65	7.25	11.74
100	851	31.01	12.94	35.90	8.26	4.61	6.68	11.18
runs with olivine bed material								
0	781	29.67	31.22	20.03	12.75	4.17	1.01	15.13
0	807	30.30	27.77	21.48	11.16	3.83	4.61	13.74
0	835	33.90	27.10	19.25	10.17	3.89	5.14	13.46
0	847	34.70	28.55	19.29	10.85	4.10	1.97	14.11
100	778	40.13	10.33	29.80	8.21	4.75	5.56	13.07
100	787	47.88	10.32	24.01	7.69	4.39	4.32	13.01
100	795	43.52	10.65	24.94	8.03	4.46	6.78	12.96

Table 6.6: Fuel ratio, producer gas composition and LHV

6.5.4 Discussion of the results from pilot plant III [21]

This series of runs contained three runs with different DSS/wood ratios and were carried out at a gasification temperature of 820 °C. As mentioned in 6.5.2 olivine was used as bed material. Table 6.7 shows the composition of the used fuels, while the figures for wood are nearly identical with table 6.5, the dried sewage sludge has a little different compostion and a 15% higher heating value.

Analysis	Wood pellets	Dried sewage sludge
proximate (wt%)		
moisture	6.1	10.95
volatile matter	77.7	49.3
ash	0.47	36.97
ultimate (wt% dry basis)		
C	49.6	26.4
H	6.3	3.3
N	0.3	3.5
S	0.02	0.9
Cl	0.01	0.04
O (by difference)	43.3	18.0
HHV (MJ/kg dry basis)	20.1	13.2
LHV (MJ/kg dry basis)	18.5	12.4
ash softening temperature (°C)	1390	1140

Table 6.7: Fuel characterization

The producer gas composition of the different runs is shown in fig. 6.15. Similar to the results from pilot plant I the CO content is decreasing significantly when gasifying sewage sludge, while CO_2 rises in a similar ratio. The H_2 fraction is nearly constant, just as the CH_4 percentage, which is tending slightly downwards. Due to the high nitrogen content of the sludge a NH_3 amount of about 35 000 ppm was found in the producer gas. The further analysed tar content was 3.5 g/Nm^3 dry basis for sewage sludge, compared to 2.5 g/Nm^3 dry basis for wood pellets.

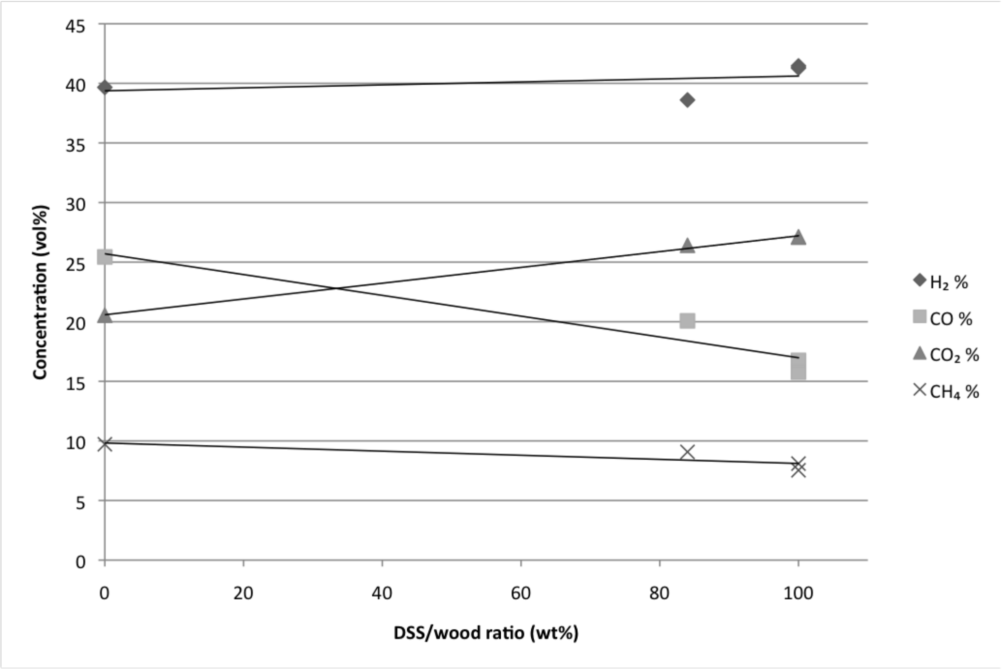


Figure 6.15: Effect of DSS ratio on producer gas composition

The gasification process performed well, only the high ash content of the sewage sludge caused a slightly different behavior of the system. Normally ash particles are entrained from the reactor with the producer gas and separated in the producer gas cyclone, but the sewage sludge produced coarse ash particles which enriched in the bed zone and led to increasing bed pressure. This was no matter in this series, as the runs finished before it could affect problems, but it would be more critical during long time runs or continuous processes. In such cases the ash has to be removed from the reactor during the run, otherwise fluidization and bed material circulation would collapse. A summary of all results from these runs is shown in table 6.8.

DSS	Producer gas (vol%)				LHV
wt%	H_2	CO	CO_2	CH_4	MJ/Nm^3
0	39.7	25.4	20.5	9.7	11.0
84	38.6	20.1	26.4	9.1	9.9
100	41.5	16.8	27.1	7.5	9.3
100	41.3	15.8	27.1	8.1	9.3

Table 6.8: Fuel ratio, producer gas composition and LHV

6.6 Cumulative experiences from gasification of sewage sludge and wood

All three experimental series showed a good performance when gasifying sewage sludge. The differences in the producer gas compositions are mainly plant specific and occur also when gasifying wood pellets. As mentioned in 6.4, olivine and calcite are catalytically active bed materials used for tar elimination from the producer gas. Based on the good experiences with olivine, this bed material is now the standard bed material used at the Vienna University of Technology as well as at the demonstration plants in Güssing and Oberwart. But olivine also enriches the hydrogen concentration in the producer gas, which explains the significant difference between the producer gas compositions between the last runs in Vienna and in Christchurch (fig. 6.16).

Regarding to the high ash content of the sludge every series had more or less serious problems with the ash load of the producer gas and the residues remaining

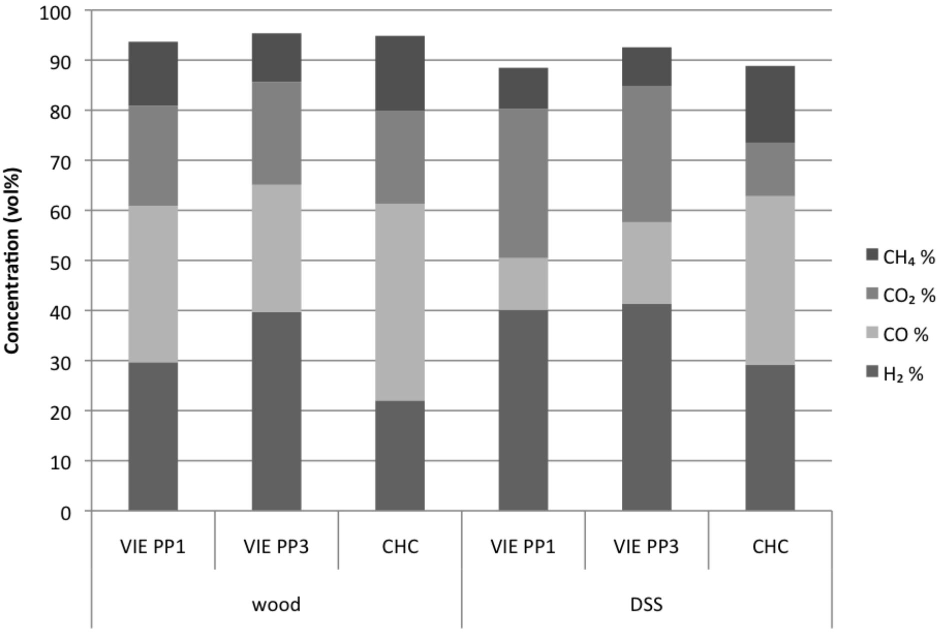


Figure 6.16: Producer gas composition on the different pilot plants

in the bed zone. No one of these gasification reactors is equipped for bed material exchange during a hot run. Therefore long time runs where not possible, because the pressure drop in the bubbling bed of the gasifier increases and bed material circulation would break down. The actual pilot plant III in Vienna allows addition of bed material during a run, but it would be necessary to remove contaminated bed material too.

Depending on the capacity of the gas cleaning system, the producer and flue gas stream was more or less loaded with ash when it reached the afterburner. This caused some problems as the flame monitoring system reported failures as the ash covered the photocell. The regularly occurring problem of tar covered inspection glasses, has been solved in Christchurch by usage of ball valves for protection from the gas streams. This solution will be also applied to the next pilot plant in Vienna, as industrial available protection solutions for inspection glasses are normally not gas tight.

As the improvements from pilot plant I to pilot plant III have been already described in other publications [24, 25, 26], just the differences between the pilot plan III in Vienna and the pilot plant in Christchurch are discussed here. Both plants have been designed for 100 kW fuel power input, following the same principles and appear similarly. Compared to the pilot plant I, which has one casing for both reactors, they have separate casings for the gasification and the combustion reactor. While the casing of the pilot plant III is made of high temperature resistant stainless steel and insulated, the casing of the pilot plant in Christchurch is refractory lined. The bottom connection between the gasification column and the combustion column is realized with a steam fluidized chute in Christchurch, opposite to an also steam fluidized siphon in Vienna, which allows an even stricter separation between the two gas streams, with nearly zero air leakage. The bed material is separated form the flue gas stream in the pilot plant III using a baffle plate, whereas Christchurch's pilot plant uses a cyclone.

The biggest difference between the two plants is the amount of the bed material, as usually 100 kg are used in Vienna while in Christchurch the standard amount is 15 kg. This difference is mainly caused through the specific design. The pilot plant III has the same main dimensions like the former pilot plant II but instead of a chute, like the pilot plant in Christchurch, now the bottom siphon is installed. Furthermore a diameter expansion in the lower part of the combustion column is built to reduce the gas velocity and to extend the gas residence time. These two changes made it necessary to increase the amount of bed material by 25 kg up to 100 kg. The cross sectional area of the BFB in Christchurch is 33650 mm^2 compared to 81225 mm^2 in Vienna, also the hight of the CFB is just 3700 mm instead of 4800 mm (the diameter of the CFB is nearly equal). Considering these figures a bed material amount of approx. 25 kg would have been expected for the gasifier in Christchurch.

A minor difference can be found between the feeding systems. The in-bed auger of the pilot plants in Vienna runs with constant speed and pushes the feed as fast as possible into the bed zone. As the other augers run at lower speeds, depending on the feed rate, the in-bed auger is not filled completely with biomass and hot gases from the reactor could pass the auger towards the hopper system and cause pyrolysis in the feeder. However the in-bed auger of the pilot plant in Christchurch runs with a lower, the feed rate according, speed out of the intermediate hopper (fig. 6.17). This hopper is filled from the main auger to a constant level and therefore the feeder screw is always filled with biomass which reduces leakage of producer gas. Combined with a cooling system for the feeder pipe, pyrolysis in the feeding system is effectively inhibited.

Figure 6.17: Intermediate hopper and in-bed auger in Christchurch/NZ

7 Conclusion

The co-gasification of dried sewage sludge and wood pellets was investigated in a dual fluidized bed steam gasification pilot plant. Runs with different ratios of DSS were carried out, beginning with an initial run with pure wood and a first run with 10 wt% of DSS, then in 20% steps from 20 up to 100 wt%. The experimental results showed that gasification of sewage sludge in existing gasification facilities is possible and that the produced gas has a heating value comparable to the produced gas from wood. The gas composition was found different, with increasing DSS/wood ratio the contents of H_2 and N_2 are rising, while the concentrations of CO and CO_2 drops. The amounts of CH_4, C_2H_4 and C_2H_6 are nearly constant, just a slight tendency down is visible.

The energy balance of the total gasification system brought cold gas efficiencies 20 and 40% which is not too bad for a small scale pilot plant. Due to the small size of the plant heat losses through the reactor casing have a big influence on the results, as well as the heat of the produced gas and also from the flue gas could not be recovered effectively. A comparable industrial power plant could reach cold gas efficiencies of more than 70%, the CHP plant in Güssing reaches 80% and above when firing wood [2]. The main reason why the efficiency decreases due to the usage of DSS is that DSS transports lots of inert material, ash, into the reactor which has to be heated up to the reaction temperature but it does not generate any benefit. Therefore the producer gas yield per kg feed is significantly lower for DSS than for wood. Another reason is the high nitrogen content of the sewage sludge (compared to wood), which leads to the already mentioned high nitrogen content of the producer gas and lowers the heating value. All things considered the producer gas seems to be suitable for usage in a gas engine or a turbine after appropriate gas cleaning.

The high ash content makes enhanced gas cleaning systems necessary, which could bear the high fly ash load of the producer gas. On the other hand bottom ash also has to be removed from the circulating bed in an appropriate way, for example batch-wise or continuous bed material exchange, to avoid accumulation of ash particles in the reactor.

From a total point of view gasification of sewage sludge seams to be suitable for industrial application, as long as the plant is optimized for this fuel. Regarding to the data from the pilot plant 20 wt% of DSS in the feed do not affect the efficiency significantly, up to 60 wt% of DSS the drop of the efficiency seams to be acceptable. A main reason for using sewage sludge for gasification could be the price compared to wood, but the higher investment and operation costs of the necessary gas cleaning equipment have to be considered. At the moment sewage sludge has to be treated in waste incineration plants as usage of sewage sludge in agriculture is limited due to the heavy metal content. According to the landfill directive of the European Union and to Austrian law landfill of untreated sewage sludge is not permitted as the total organic carbon is above the limit of 5 wt%.

Under these conditions co-gasification of sewage sludge could be competitive to other bio-energy solutions as well as to regular waste incineration. Therefore further experiments with sewage sludge should be carried out at the pilot plant as well as at a suitable industrial plant. Additional investigations should be done to gain more experience and for optimization of the process.

Abbreviations

af	ash free
BFB	Bubbling fluidized bed
CFB	Circulating fluidized bed
CGE	Cold gas efficiency
CHC	University of Canterbury - Pilot plant
CHP	Combined heat and power
daf	dry and ash free
DFB	Dual fluidized bed
DSS	Dried sewage sludge
FICFB	Fast interanlly circulating fluidized bed
GC	Gas chromatography
GC-MS	Gas chromatography-mass spectrometry
HDPE	High density polyethylene
HHV	Higher heating vlaue
IGCC	Integrated gasification combined cycle
LHV	Lower heating value
LPG	Liquified petroleum gas
n.a.	not analyzed
PCB	polychlorinated biphenyl
PCDD	polychlorinated dibenzo-p-dioxine
PCDF	polychlorinated dibenzo-p-furan
sfr	Steam fuel ratio

SNG	Synthetic natural gas
SPE	Solid Phase Extraction
TDH	Transport disengaging height
VIE PP1	Vienna University of Technology - Pilot plant I
VIE PP3	Vienna University of Technology - Pilot plant III

Symbols

Ar	Archimedes Number
C_D	Drag coefficient
C_x	Concentration of component x
d_{sv}	Equivalent diameter
H	Height
m	Mass
m_p	Mass of the Particle
S	Surface
U	Velocity
V	Volume
Δp	Pressure difference
ε	Porosity
λ	Equivalence ratio
μ	Dynamic viscosity
ρ_b	Bulkdensity
ρ_p	Density of the Particle

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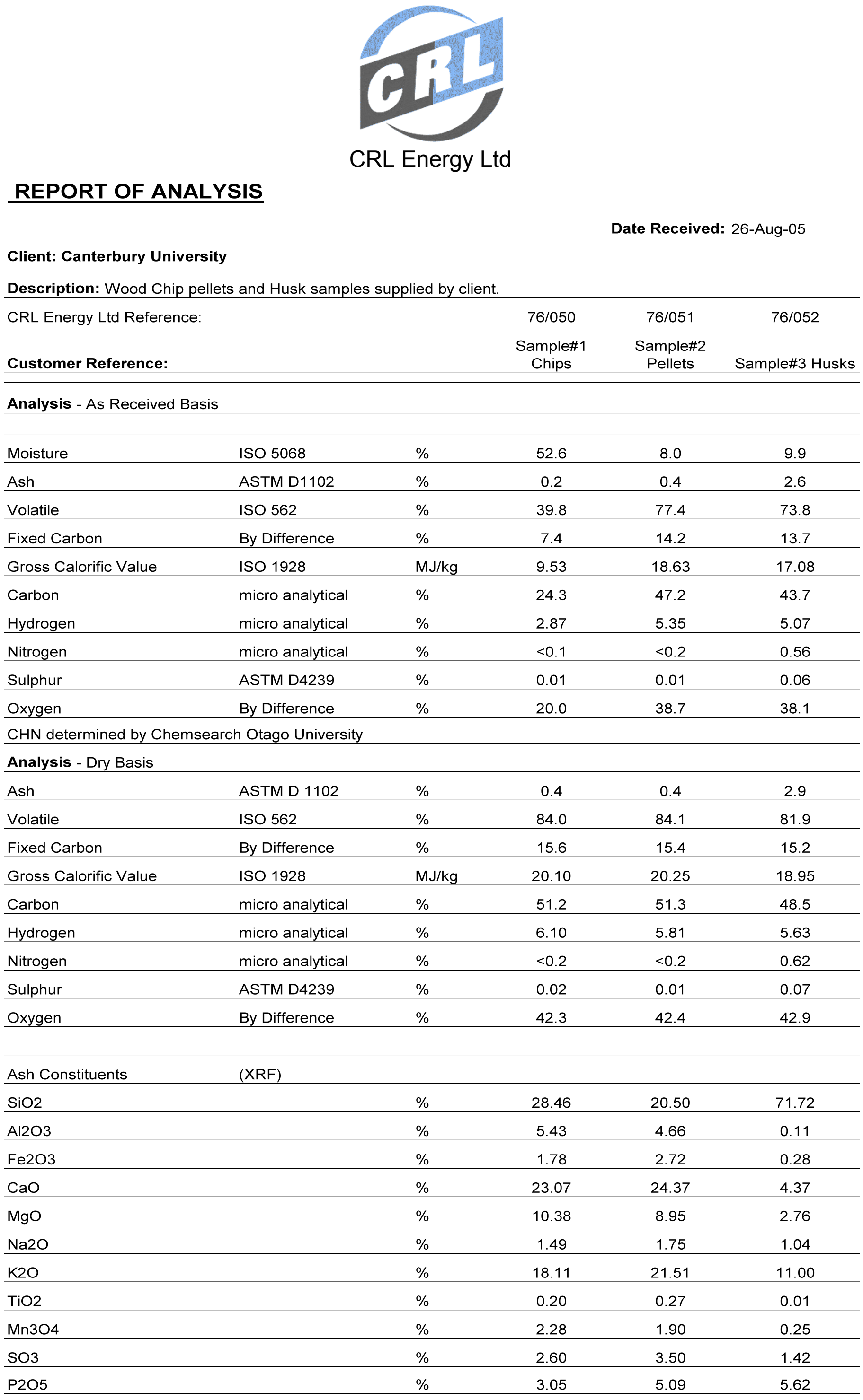
List of Figures

List of Tables

Appendix A. Detailed ultimate analysis of wood pellets - Sample#2



CRL Energy Ltd

REPORT OF ANALYSIS

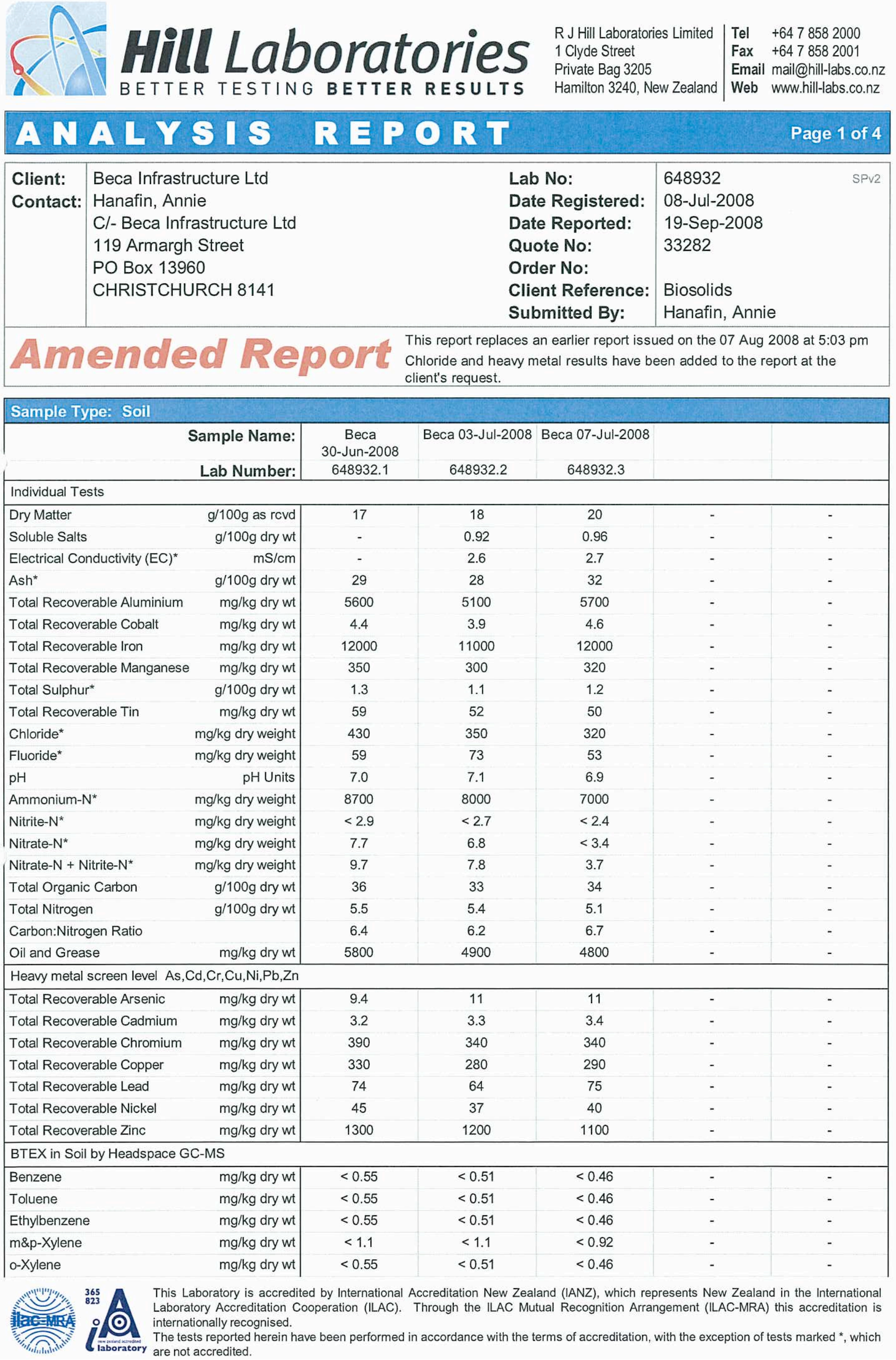
Date Received: 26-Aug-05

Client: Canterbury University

Description: Wood Chip pellets and Husk samples supplied by client.

CRL Energy Ltd Reference:			76/050	76/051	76/052
Customer Reference:			Sample#1 Chips	Sample#2 Pellets	Sample#3 Husks
Analysis - As Received Basis					
Moisture	ISO 5068	%	52.6	8.0	9.9
Ash	ASTM D1102	%	0.2	0.4	2.6
Volatile	ISO 562	%	39.8	77.4	73.8
Fixed Carbon	By Difference	%	7.4	14.2	13.7
Gross Calorific Value	ISO 1928	MJ/kg	9.53	18.63	17.08
Carbon	micro analytical	%	24.3	47.2	43.7
Hydrogen	micro analytical	%	2.87	5.35	5.07
Nitrogen	micro analytical	%	<0.1	<0.2	0.56
Sulphur	ASTM D4239	%	0.01	0.01	0.06
Oxygen	By Difference	%	20.0	38.7	38.1
CHN determined by Chemsearch Otago University					
Analysis - Dry Basis					
Ash	ASTM D 1102	%	0.4	0.4	2.9
Volatile	ISO 562	%	84.0	84.1	81.9
Fixed Carbon	By Difference	%	15.6	15.4	15.2
Gross Calorific Value	ISO 1928	MJ/kg	20.10	20.25	18.95
Carbon	micro analytical	%	51.2	51.3	48.5
Hydrogen	micro analytical	%	6.10	5.81	5.63
Nitrogen	micro analytical	%	<0.2	<0.2	0.62
Sulphur	ASTM D4239	%	0.02	0.01	0.07
Oxygen	By Difference	%	42.3	42.4	42.9
Ash Constituents	(XRF)				
SiO2		%	28.46	20.50	71.72
Al2O3		%	5.43	4.66	0.11
Fe2O3		%	1.78	2.72	0.28
CaO		%	23.07	24.37	4.37
MgO		%	10.38	8.95	2.76
Na2O		%	1.49	1.75	1.04
K2O		%	18.11	21.51	11.00
TiO2		%	0.20	0.27	0.01
Mn3O4		%	2.28	1.90	0.25
SO3		%	2.60	3.50	1.42
P2O5		%	3.05	5.09	5.62

Appendix B. Detailed ultimate analysis of sewage sludge



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ANALYSIS REPORT **Page 1 of 4**

Client:	Beca Infrastructure Ltd	**Lab No:**	648932 SPv2
Contact:	Hanafin, Annie	**Date Registered:**	08-Jul-2008
	C/- Beca Infrastructure Ltd	**Date Reported:**	19-Sep-2008
	119 Armargh Street	**Quote No:**	33282
	PO Box 13960	**Order No:**	
	CHRISTCHURCH 8141	**Client Reference:**	Biosolids
		Submitted By:	Hanafin, Annie

Amended Report This report replaces an earlier report issued on the 07 Aug 2008 at 5:03 pm Chloride and heavy metal results have been added to the report at the client's request.

Sample Type: Soil

Sample Name:		Beca 30-Jun-2008	Beca 03-Jul-2008	Beca 07-Jul-2008		
Lab Number:		648932.1	648932.2	648932.3		
Individual Tests						
Dry Matter	g/100g as rcvd	17	18	20	-	-
Soluble Salts	g/100g dry wt	-	0.92	0.96	-	-
Electrical Conductivity (EC)*	mS/cm	-	2.6	2.7	-	-
Ash*	g/100g dry wt	29	28	32	-	-
Total Recoverable Aluminium	mg/kg dry wt	5600	5100	5700	-	-
Total Recoverable Cobalt	mg/kg dry wt	4.4	3.9	4.6	-	-
Total Recoverable Iron	mg/kg dry wt	12000	11000	12000	-	-
Total Recoverable Manganese	mg/kg dry wt	350	300	320	-	-
Total Sulphur*	g/100g dry wt	1.3	1.1	1.2	-	-
Total Recoverable Tin	mg/kg dry wt	59	52	50	-	-
Chloride*	mg/kg dry weight	430	350	320	-	-
Fluoride*	mg/kg dry weight	59	73	53	-	-
pH	pH Units	7.0	7.1	6.9	-	-
Ammonium-N*	mg/kg dry weight	8700	8000	7000	-	-
Nitrite-N*	mg/kg dry weight	< 2.9	< 2.7	< 2.4	-	-
Nitrate-N*	mg/kg dry weight	7.7	6.8	< 3.4	-	-
Nitrate-N + Nitrite-N*	mg/kg dry weight	9.7	7.8	3.7	-	-
Total Organic Carbon	g/100g dry wt	36	33	34	-	-
Total Nitrogen	g/100g dry wt	5.5	5.4	5.1	-	-
Carbon:Nitrogen Ratio		6.4	6.2	6.7	-	-
Oil and Grease	mg/kg dry wt	5800	4900	4800	-	-
Heavy metal screen level As,Cd,Cr,Cu,Ni,Pb,Zn						
Total Recoverable Arsenic	mg/kg dry wt	9.4	11	11	-	-
Total Recoverable Cadmium	mg/kg dry wt	3.2	3.3	3.4	-	-
Total Recoverable Chromium	mg/kg dry wt	390	340	340	-	-
Total Recoverable Copper	mg/kg dry wt	330	280	290	-	-
Total Recoverable Lead	mg/kg dry wt	74	64	75	-	-
Total Recoverable Nickel	mg/kg dry wt	45	37	40	-	-
Total Recoverable Zinc	mg/kg dry wt	1300	1200	1100	-	-
BTEX in Soil by Headspace GC-MS						
Benzene	mg/kg dry wt	< 0.55	< 0.51	< 0.46	-	-
Toluene	mg/kg dry wt	< 0.55	< 0.51	< 0.46	-	-
Ethylbenzene	mg/kg dry wt	< 0.55	< 0.51	< 0.46	-	-
m&p-Xylene	mg/kg dry wt	< 1.1	< 1.1	< 0.92	-	-
o-Xylene	mg/kg dry wt	< 0.55	< 0.51	< 0.46	-	-

This Laboratory is accredited by International Accreditation New Zealand (IANZ), which represents New Zealand in the International Laboratory Accreditation Cooperation (ILAC). Through the ILAC Mutual Recognition Arrangement (ILAC-MRA) this accreditation is internationally recognised.
The tests reported herein have been performed in accordance with the terms of accreditation, with the exception of tests marked *, which are not accredited.

Sample Type: Soil

Sample Name:		Beca 30-Jun-2008	Beca 03-Jul-2008	Beca 07-Jul-2008		
Lab Number:		648932.1	648932.2	648932.3		
Polycyclic Aromatic Hydrocarbons Trace in Soil						
Acenaphthene	mg/kg dry wt	0.050	0.049	0.055	-	-
Acenaphthylene	mg/kg dry wt	0.037	0.036	0.048	-	-
Anthracene	mg/kg dry wt	0.060	0.065	0.067	-	-
Benzo[a]anthracene	mg/kg dry wt	0.10	0.10	0.10	-	-
Benzo[a]pyrene (BAP)	mg/kg dry wt	0.085	0.086	0.097	-	-
Benzo[b]fluoranthene + Benzo[j] fluoranthene	mg/kg dry wt	0.20	0.20	0.23	-	-
Benzo[g,h,i]perylene	mg/kg dry wt	0.086	0.099	0.10	-	-
Benzo[k]fluoranthene	mg/kg dry wt	0.074	0.064	0.061	-	-
Chrysene	mg/kg dry wt	0.11	0.10	0.11	-	-
Dibenzo[a,h]anthracene	mg/kg dry wt	0.033	0.033	0.023	-	-
Fluoranthene	mg/kg dry wt	0.21	0.17	0.19	-	-
Fluorene	mg/kg dry wt	0.11	0.092	0.12	-	-
Indeno(1,2,3-c,d)pyrene	mg/kg dry wt	0.054	0.068	0.077	-	-
Naphthalene	mg/kg dry wt	0.50	0.55	0.61	-	-
Phenanthrene	mg/kg dry wt	0.29	0.36	0.30	-	-
Pyrene	mg/kg dry wt	0.58	0.61	0.58	-	-
Poychlorinated Biphenyls Trace in Soil						
PCB-101	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-105	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-110	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-114	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-118	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-121	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-123	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-126	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-128	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-138	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-141	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-149	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-151	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-153	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-156	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-157	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-159	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-167	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-169	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-170	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-180	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-189	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-194	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-206	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-209	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-28 + PCB-31	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-44	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-49	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-52	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-60	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-77	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-81	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
PCB-86	mg/kg dry wt	< 0.00099	< 0.0010	< 0.0010	-	-
Total PCB (Sum of 33 congeners)	mg/kg dry wt	< 0.02	< 0.02	< 0.02	-	-

Analyst's Comments

"We were unable to test for Soluble Salts on sample 648932.1 using our standard method because the sample absorbed all

Analyst's Comments

the water added, leaving no free liquid to measure EC on."

SUMMARY OF METHODS

The following table(s) gives a brief description of the methods used to conduct the analyses for this job. The detection limits given below are those attainable in a relatively clean matrix. Detection limits may be higher for individual samples should insufficient sample be available, or if the matrix requires that dilutions be performed during analysis.

Sample Type: Soil

Test	Method Description	Default Detection Limit	Samples
Environmental Solids Sample Preparation*	Air dried at 35°C and sieved, <2mm fraction.	-	1-3
Heavy metal screen level As,Cd,Cr,Cu,Ni,Pb,Zn	Dried sample, <2mm fraction. Nitric/Hydrochloric acid digestion, ICP-MS, screen level.	-	1-3
Total Organic Carbon and Total Nitrogen	Catalytic Combustion (900°C, O^2), separation, Thermal Conductivity Detector [Elementar Analyser]	-	1-3
BTEX in Soil by Headspace GC-MS	Solvent extraction, Headspace GC-MS analysis US EPA 8260B	-	1-3
Polycyclic Aromatic Hydrocarbons Trace in Soil	Sonication extraction, SPE cleanup, GC-MS SIM analysis US EPA 8270C	-	1-3
Poychlorinated Biphenyls Trace in Soil	Sonication extraction, SPE cleanup, GPC cleanup (if required), GC-MS analysis	-	1-3
Dry Matter (Env)	Dried at 103°C (removes 3-5% more water than air dry), gravimetry.	0.10 g/100g as rcvd	1-3
esFIAextn*	2M potassium chloride extraction for FIA determination. Analyst, 109, 549, (1984).	-	1-3
esICextn*	Potassium phosphate extraction for Ion Chromatography. In House.	-	1-3
Total Recoverable digestion	Nitric / hydrochloric acid digestion. US EPA 200.2.	-	1-3
Soluble Salts	1:5 soil:water extraction followed by potentiometric determination of conductivity. SS=EC*0.35 Calculated from EC measurement.	0.050 g/100g dry wt	2-3
Conductivity from soluble salts*	1:5 soil:water extraction, potentiometric conductivity determination (Soluble salts/0.35) .	0.20 mS/cm	2-3
Ash*	Ignition in muffle furnace 550°C, 6hr, gravimetric. APHA 2540 G 21st ed. 2005.	0.040 g/100g dry wt	1-3
Total Recoverable Aluminium	Dried sample, sieved as specified (if required). Nitric/Hydrochloric acid digestion, ICP-MS, screen level. US EPA 200.2.	10 mg/kg dry wt	1-3
Total Recoverable Cobalt	Dried sample, sieved as specified (if required). Nitric/Hydrochloric acid digestion, ICP-MS, screen level. US EPA 200.2.	0.40 mg/kg dry wt	1-3
Total Recoverable Iron	Dried sample, sieved as specified (if required). Nitric/Hydrochloric acid digestion, ICP-MS, screen level. US EPA 200.2.	40 mg/kg dry wt	1-3
Total Recoverable Manganese	Dried sample, sieved as specified (if required). Nitric/Hydrochloric acid digestion, ICP-MS, screen level. US EPA 200.2.	1.0 mg/kg dry wt	1-3
Total Sulphur (Sub)*	LECO SC32 Sulphur Determinator, high temperature furnace, infra-red detector. Subcontracted to SGS, Waihi. ASTM 4239.	0.0050 g/100g dry wt	1-3
Total Recoverable Tin	Dried sample, sieved as specified (if required). Nitric/Hydrochloric acid digestion, ICP-MS, screen level. US EPA 200.2.	1.0 mg/kg dry wt	1-3
Chloride*	Ion Chromatography determination of es potassium phosphate extraction. APHA 4110 B 21st ed. 2005.	3.0 mg/kg dry weight	1-3
Total Fluoride in solids*	Alkaline fusion of sample. Ion selective electrode determination. Methods of Soil Analysis 2nd Edition, Pt2, 26-4.3.3.	10 mg/kg dry weight	1-3
pH	1:2 (v/v) soil : water slurry followed by potentiometric determination of pH.	0.1 pH Units	1-3
Ammonium-N*	2M potassium chloride extraction. Phenol/hypochlorite colorimetry. Discrete Analyser. APHA 4500-NH_3 G 21st ed. 2005.	5.0 mg/kg dry weight	1-3
Nitrite-N*	FIA determination of es 2M potassium chloride extraction. APHA 4500-NO_3^- I (Proposed) 21st ed. 2005.	1.0 mg/kg dry weight	1-3
Nitrate-N*	Calculation: (Nitrate-N + Nitrite-N) - Nitrite-N.	1.5 mg/kg dry weight	1-3
Nitrate-N + Nitrite-N*	Automated cadmium reduction, FIA determination of es 2M potassium chloride extraction. APHA 4500-NO_3^- I (Proposed) 21st ed. 2005.	1.0 mg/kg dry weight	1-3

Sample Type: Soil

Test	Method Description	Default Detection Limit	Samples
Oil and Grease	Chemical drying, Soxhlet extraction, gravimetric determination of extracted Oil & Grease.	100 mg/kg dry wt	1-3

These samples were collected by yourselves (or your agent) and analysed as received at the laboratory.

Samples are held at the laboratory after reporting for a length of time depending on the preservation used and the stability of the analytes being tested. Once the storage period is completed the samples are discarded unless otherwise advised by the client.

This report must not be reproduced, except in full, without the written consent of the signatory.

Carole Rodger-Carroll

Carole Rodgers-Carroll BA, NZCS
Client Services Manager - Environmental Division

Appendix C. Definition of the cold gas efficiency

The cold gas efficiency is defined as the ratio of the energy content of the producer gas and the total energy input of the gasification process. The energy content of the producer gas stream has to be calculated for standard temperature, therefore it is equal to the heating value.

$$CGE = \frac{\text{heating value of the cold producer gas stream}}{\text{total energy input}}$$

The total energy input contains not only the heating value of the feed, but also the enthalpy of the used steam and preheated air, the additional fuel in the combustion area (LPG) and electrical heat support.

Appendix D. Data sheets graywacke runs CHC

COG-I-090120

Operating conditions

Gasification temperature	750 °C	Comissioning run	
feed mass flow	18.1 kg/h wood		
bed material	13 kg Graywacke sand		
Chute steam flow	4 kg/h	CFB distributer air flow	42.9 Nm³/h
BFB steam flow	6 kg/h	CFB Primary air flow	17.4 Nm³/h
Siphon steam flow	3 kg/h		

Startup 07:30 Stable 12:45 Shutdown 13:50

Producer gas samples

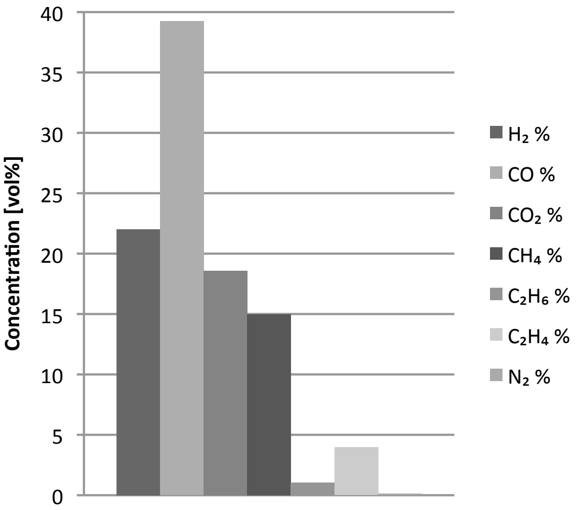
Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1253 PG1	2.41	20.4	5.88	14.2	41.3	12.5	2.55	0.75
1327 PG2	1.24	21.4	1.58	14.6	38.2	18.1	3.86	1.03

Flue gas samples

Sample	He %	O_2 %	N_2 %	CO_2 %
1254 FG1	0.28	7.03	82.8	9.87
1328 FG2	0.29	3.22	82.4	14.10

Producer gas compositon

H_2 %	22.03
CO %	39.24
CO_2 %	18.59
CH_4 %	14.99
C_2H_6 %	1.06
C_2H_4 %	3.97
N_2 %	0.12



Gasification efficiency

Lower CV (MJ/Nm³)	15.72
Gas Gen Rate(Nm³/h)	10.35
Combustion E gas (kW)	45.18
Total Energy input (kW)	118
Cold Gas Efficiency	38.31

For comparison with the sewage sludge series, the 1328 PG1 sample was used, as the feed rate at 12:53 was just 11 kg/h

COG-10B-090319

Operating conditions

Gasification temperature	750 °C		
feed mass flow	16.3 kg/h wood, 1.8 kg/h DSS		
bed material	13 kg Greywacke sand		
Chute steam flow	4 kg/h	CFB distributer air flow	16.2 Nm3/h
BFB steam flow	6 kg/h	CFB Primary air flow	39.4 Nm3/h
Siphon steam flow	4.5 kg/h		

Startup	07:30	Stable	14:30	Shutdown	16:05

Producer gas samples

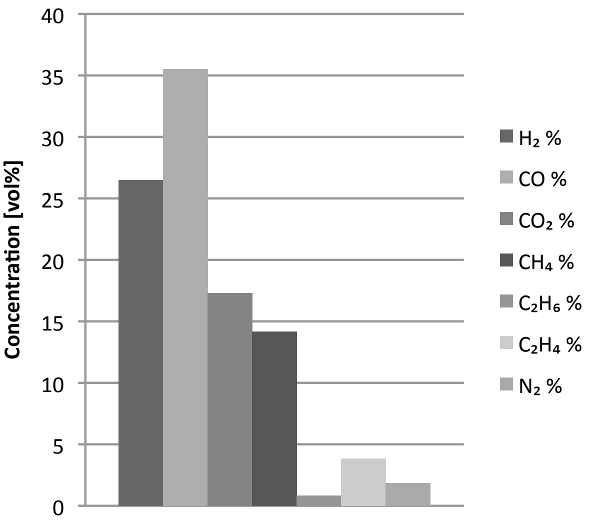
Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1448 PG1	1.24	24.8	6.91	13.9	35.9	13.8	2.83	0.69
1513 PG2	1.18	26.2	6.08	13.4	33.6	15.3	3.48	0.79
1539 PG3	1.13	26.3	5.69	13.3	32.5	16.7	3.70	0.75
1600 PG4	0.96	22.7	6.48	12.9	32.1	19.5	4.47	0.88

Flue gas samples

Sample	He %	O_2 %	N_2 %	CO_2 %
1449 FG1	0.31	1.78	83.5	14.4
1514 FG2	0.30	3.98	82.3	13.4
1540 FG3	0.29	4.91	81.9	13.0
1601 FG4	0.30	5.27	81.8	12.6

Producer gas compositon

H_2 %	26.49
CO %	35.53
CO_2 %	17.32
CH_4 %	14.16
C_2H_6 %	0.82
C_2H_4 %	3.84
N_2 %	1.85



Gasification efficiency

Lower CV (MJ/Nm^3)	15.27
Gas Gen Rate(Nm^3/h)	9.12
Combustion E gas (kW)	38.72
Total Energy input (kW)	119
Cold Gas Efficiency	32.53

COG-20B-090402

Operating conditions

Gasification temperature	750 °C		
feed mass flow	14.5 kg/h wood, 3.6 kg/h DSS		
bed material	13 kg Graywacke sand		
Chute steam flow	4.5 kg/h	CFB distributer air flow	17.9 Nm3/h
BFB steam flow	6.5 kg/h	CFB Primary air flow	45.6 Nm3/h
Siphon steam flow	5 kg/h		

Startup	07:30	Stable	13:00	Shutdown	15:40

Producer gas samples

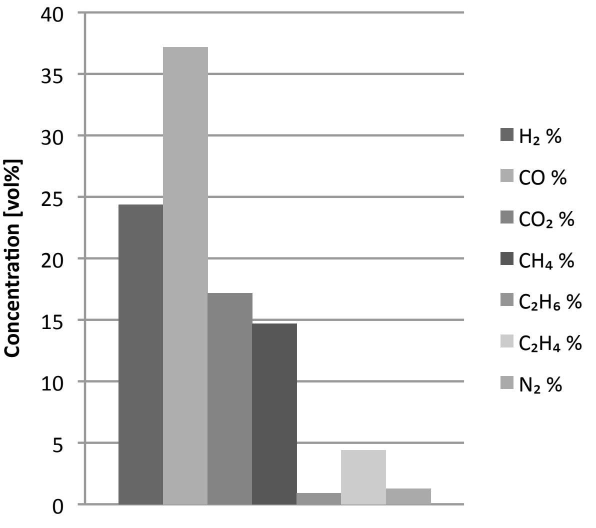
Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1310 PG1	1.32	25.7	5.2	15.0	37.3	11.9	2.97	0.75
1335 PG2	1.30	24.3	5.9	13.9	34.7	15.2	3.93	0.84
1407 PG3	1.31	23.2	5.7	13.8	34.9	16.1	4.30	0.85
1438 PG4	1.31	21.7	5.8	13.4	34.1	18.1	4.75	0.88
1508 PG5	1.26	21.0	6.4	13.4	34.2	18.4	4.48	0.90
1537 PG6	1.36	21.7	6.1	13.5	34.7	17.2	4.54	0.89

Flue gas samples

Sample	He %	O_2 %	N_2 %	CO_2 %
1311 FG1	0.25	5.31	81.8	12.6
1336 FG2	0.25	5.84	81.8	12.1
1408 FG3	0.25	6.70	81.4	11.7
1439 FG4	0.24	8.09	81.4	10.3
1509 FG5	0.24	7.25	81.3	11.2
1538 FG6	0.24	7.10	81.5	11.2

Producer gas compositon

H_2 %	24.4
CO %	37.2
CO_2 %	17.2
CH_4 %	14.7
C_2H_6 %	0.9
C_2H_4 %	4.4
N_2 %	1.3



Gasification efficiency

Lower CV (MJ/Nm^3)	15.77
Gas Gen Rate(Nm^3/h)	9.46
Combustion E gas (kW)	41.43
Total Energy input (kW)	110
Cold Gas Efficiency	37.69

COG-40B-090428

Operating conditions

Gasification temperature	750 °C		
feed mass flow	10.9 kg/h wood, 7.3 kg/h DSS		
bed material	13 kg Greywacke sand		
Chute steam flow	4.5 kg/h	CFB distributer air flow	17.2 Nm3/h
BFB steam flow	6.5 kg/h	CFB Primary air flow	49.3 Nm3/h
Siphon steam flow	5 kg/h		

Startup 07:00 Stable 13:00 Parameter change 14:20

Producer gas samples

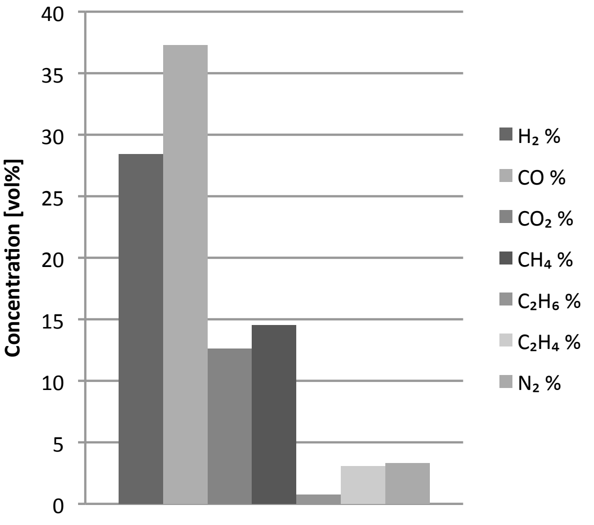
Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1324 PG1	1.49	28.6	7.79	14.4	38.7	7.2	1.30	0.48
1352 PG2	1.36	25.6	6.46	14.0	35.3	13.2	3.25	0.79
1416 PG3	1.30	27.1	4.78	13.3	32.7	15.7	4.27	0.86

Flue gas samples

Sample	He %	O_2 %	N_2 %	CO_2 %
1325 FG1	0.25	6.53	82.1	11.1
1353 FG2	0.25	5.69	82.0	12.0
1417 FG3	0.26	4.80	82.2	12.7

Producer gas compositon

H_2 %	28.41
CO %	37.29
CO_2 %	12.63
CH_4 %	14.54
C_2H_6 %	0.74
C_2H_4 %	3.08
N_2 %	3.32



Gasification efficiency

Lower CV (MJ/Nm^3)	15.27
Gas Gen Rate(Nm^3/h)	9.12
Combustion E gas (kW)	38.72
Total Energy input (kW)	119
Cold Gas Efficiency	32.53

COG-60B-090428

Operating conditions

Gasification temperature	750 °C		
feed mass flow	10.9 kg/h wood, 7.3 kg/h DSS		
bed material	13 kg Greywacke sand		
Chute steam flow	4.5 kg/h	CFB distributer air flow	17.2 Nm3/h
BFB steam flow	6.5 kg/h	CFB Primary air flow	49.3 Nm3/h
Siphon steam flow	5 kg/h		

Parameter change 14:20 Stable 14:40 Shutdown 15:30

Producer gas samples

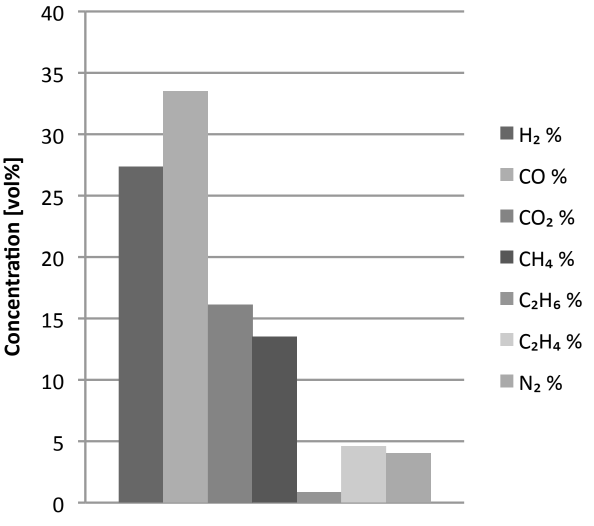
Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1447 PG4	1.33	26.6	5.48	13.1	32.5	15.7	4.46	0.85
1528 PG5	1.37	25.9	5.39	14.6	31.3	16.3	4.23	0.91

Flue gas samples

Sample	He %	O_2 %	N_2 %	CO_2 %
1448 FG4	0.260	5.6	82.1	12.1
1529 FG5	unable to sample due to high ash load of the flue gas			

Producer gas compositon

H_2 %	27.36
CO %	33.51
CO_2 %	16.14
CH_4 %	13.50
C_2H_6 %	0.88
C_2H_4 %	4.59
N_2 %	4.02



Gasification efficiency

Lower CV (MJ/Nm^3)	15.29
Gas Gen Rate(Nm^3/h)	9.62
Combustion E gas (kW)	40.85
Total Energy input (kW)	110
Cold Gas Efficiency	37.06

COG-80B-090505

Operating conditions

Gasification temperature	750 °C		
feed mass flow	10.9 kg/h wood, 7.3 kg/h DSS		
bed material	13 kg Greywacke sand		
Chute steam flow	4.5 kg/h	CFB distributer air flow	17.4 Nm3/h
BFB steam flow	6.5 kg/h	CFB Primary air flow	50.4 Nm3/h
Siphon steam flow	4.5 kg/h		

Startup	07:30	Stable	13:30	Parameter change	14:30

Producer gas samples

Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1406 PG1	1.86	20.9	14.88	11.4	26.5	18.0	5.45	0.96
1428 PG2	2.57	28.2	20.75	12.4	29.3	5.2	1.15	0.44

Flue gas samples

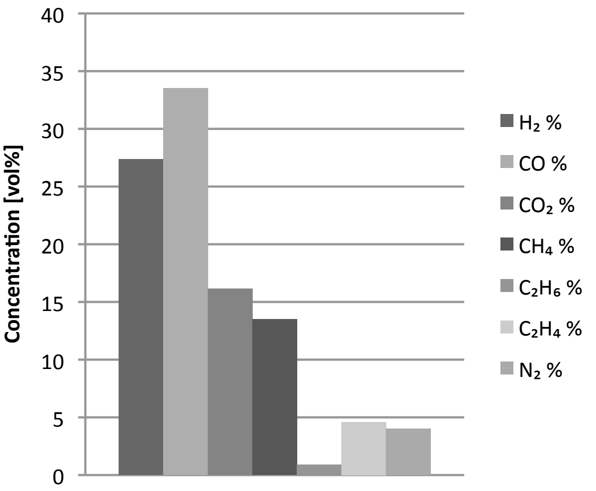
Sample	He %	O_2 %	N_2 %	CO_2 %

Producer gas compositon

H_2 %	27.36
CO %	33.51
CO_2 %	16.14
CH_4 %	13.50
C_2H_6 %	0.88
C_2H_4 %	4.59
N_2 %	4.02

Gasification efficiency

Lower CV (MJ/Nm^3)	15.29
Gas Gen Rate(Nm^3/h)	9.62
Combustion E gas (kW)	40.85
Total Energy input (kW)	110
Cold Gas Efficiency	37.06



COG-100B-090505

Operating conditions

Gasification temperature	750 °C		
feed mass flow	10.9 kg/h wood, 7.3 kg/h DSS		
bed material	13 kg Greywacke sand		
Chute steam flow	4.5 kg/h	CFB distributer air flow	17.4 Nm3/h
BFB steam flow	6.5 kg/h	CFB Primary air flow	50.4 Nm3/h
Siphon steam flow	4.5 kg/h		

Parameter change	14:30	Stable	14:40	Shutdown	15:10

Producer gas samples

Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1448 PG1	2.60	25.9	17.72	13.3	30.1	7.6	2.08	0.67
1408 PG2	2.51	23.9	18.76	12.9	27.3	10.6	3.33	0.82

Flue gas samples

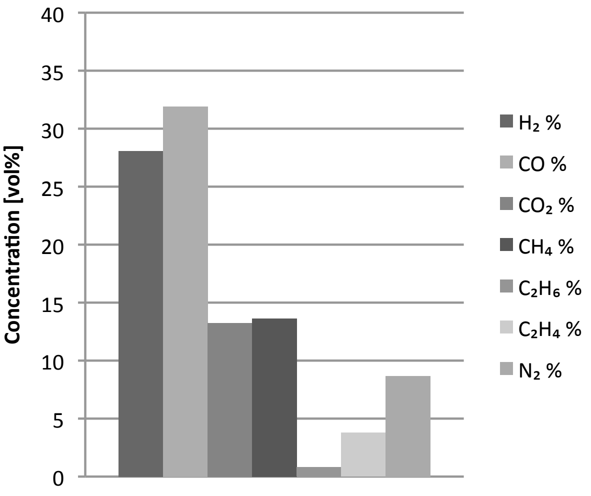
Sample	He %	O_2 %	N_2 %	CO_2 %

Producer gas compositon

H_2 %	28.07
CO %	31.92
CO_2 %	13.24
CH_4 %	13.64
C_2H_6 %	0.80
C_2H_4 %	3.78
N_2 %	8.67

Gasification efficiency

Lower CV (MJ/Nm^3)	14.68
Gas Gen Rate(Nm^3/h)	6.17
Combustion E gas (kW)	26.50
Total Energy input (kW)	109
Cold Gas Efficiency	24.39



Appendix E. Data sheets olivine/calcite runs CHC

COG-I-OC-090302

Operating conditions

Gasification temperature	700 °C		
feed mass flow	18.1 kg/h wood		
bed material	9 kg Olivine, 3 kg Calcite		
Chute steam flow	4 kg/h	CFB distributer air flow	18.3 Nm3/h
BFB steam flow	6 kg/h	CFB Primary air flow	50.1 Nm3/h
Siphon steam flow	4 kg/h		

Startup 07:45 Stable 13:20 Shutdown 16:20

Producer gas samples

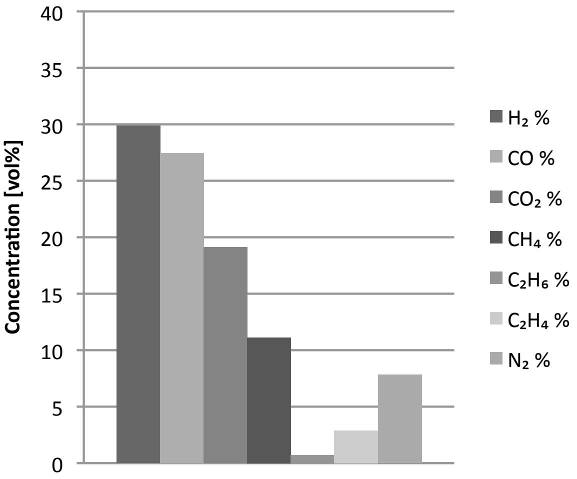
Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1337 PG1	1.18	32.8	9.1	11.4	27.7	14.7	2.44	0.62
1410 PG2	0.99	31.1	7.9	11.0	27.0	18.4	2.91	0.68
1444 PG3	0.85	30.1	9.3	10.3	26.3	19.5	2.89	0.68
1510 PG4	0.81	28.2	4.4	11.8	29.8	21.0	3.19	0.84
1536 PG5	0.95	27.2	8.5	11.1	26.4	22.0	3.07	0.80
1559 PG6	1.13	27.1	5.5	11.6	28.3	22.4	3.08	0.88

Flue gas samples

Sample	He %	O_2 %	N_2 %	CO_2 %
1340 FG1	0.25	5.12	81.9	12.8
1411 FG2	0.24	7.02	81.9	10.8
1445 FG3	0.24	6.49	80.9	12.4
1511 FG4	0.24	11.39	79.8	8.6
1537 FG5	0.24	12.69	79.8	7.3
1600 FG6	0.23	9.95	80.0	9.8

Producer gas compositon

H_2 %	29.9
CO %	27.5
CO_2 %	19.1
CH_4 %	11.1
C_2H_6 %	0.7
C_2H_4 %	2.9
N_2 %	7.8



Gasification efficiency

Lower CV (MJ/Nm^3)	13.06
Gas Gen Rate(Nm^3/h)	14.98
Combustion E gas (kW)	54.45
Total Energy input (kW)	100
Cold Gas Efficiency	54.28

COG-20B-OC-090421

Operating conditions

Gasification temperature	700 °C		
feed mass flow	18.2 kg/h wood		
bed material	9 kg Olivine, 3 kg Calcite		
Chute steam flow	4.5 kg/h	CFB distributer air flow	17.7 Nm3/h
BFB steam flow	7 kg/h	CFB Primary air flow	55.6 Nm3/h
Siphon steam flow	6.5 kg/h		

Startup	07:30	Stable	12:00	Shutdown	15:15

Producer gas samples

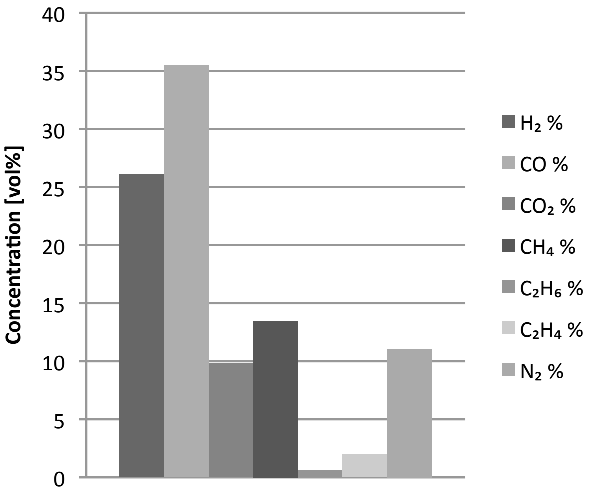
Sample	He %	H_2 %	N_2 %	CH_4 %	CO %	CO_2 %	C_2H_4 %	C_2H_6 %
1227 PG1b	1.56	29.5	11.2	14.2	35.9	6.3	0.95	0.45
1258 PG2	1.46	25.8	11.7	13.2	35.2	10.1	1.97	0.65
1326 PG3	1.45	23.0	10.2	13.0	35.4	13.2	2.95	0.77

Flue gas samples

Sample	He %	O_2 %	N_2 %	CO_2 %
1214 FG1	0.24	6.29	81.9	11.5
1259 FG2	0.25	6.63	82.3	10.9
1327 FG3	0.24	6.11	82.1	11.6

Producer gas compositon

H_2 %	26.1
CO %	35.5
CO_2 %	9.8
CH_4 %	13.5
C_2H_6 %	0.6
C_2H_4 %	2.0
N_2 %	11.0



Gasification efficiency

Lower CV (MJ/Nm^3)	14.76
Gas Gen Rate(Nm^3/h)	8.18
Combustion E gas (kW)	33.52
Total Energy input (kW)	129
Cold Gas Efficiency	25.99

Appendix F. Positions of thermocouples and pressure sensors of the gasification pilot plant in Christchurch