



# TECHNISCHE UNIVERSITÄT WIEN

## DIPLOMARBEIT

### **Influence of fluid dynamics on the gasification in a dual fluidized bed**

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.....  
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## **Abstract**

Our increasing demand of energy requires the construction of new power plants. Although new deposits of oil and natural gas are found, our supply of fossil fuels is limited.

The working group of Professor Hofbauer "Future Energy Technology" at the Institute of Chemical Engineering at the Vienna University of Technology is conducting research with a 100 kW dual fluidized bed gasifier. Alternative fuels and catalytic bed materials are not the only materials that are investigated to make this type of energy production more interesting for the industry. With the cooperation between the Vienna University of Technology and the industrial power plant in Güssing the results of the research can easily be verified on an industrial scale.

This cooperation with others is important to the workgroup. There is a permanent correspondence between universities worldwide including New Zealand, Germany, United States of America, Sweden, Portugal and France, exchanges are also supported.

This thesis was written during one of these exchanges. It covers the fluid dynamics of the dual fluidized bed at the University of Canterbury in Christchurch, New Zealand and at the Vienna University of Technology in Austria. During three months in New Zealand data were collected for this thesis and compared with those in Austria. For this purpose bed materials were used that have the same composition or at least almost the same catalytic influence on the gasification. In Christchurch greywacke sand and in Vienna silica sand was used. There are certain parameters that are important to describe a fluidized bed. For this thesis they were calculated and compared. The result of this calculation shows that the way the two gasifiers are operated is very similar.

The two gasifiers differ in the structure, geometry, bed material and gasification temperature. These differences have an influence on operating factors, like the Archimedes number or the superficial gas velocity.

These parameters have an influence on the product gas. The composition of the product gas is significantly different but there are reasons which explain these differences.

All the differences in geometry and operation of these pilot plants make it difficult to compare the collected data. In this thesis some explanations are given.

## Kurzfassung

Durch den steigenden Verbrauch an Energie und den damit verbundenen Verbrauch an fossilen Rohstoffen, stellt sich die Frage nach Alternativen zu Erdöl und Erdgas. Trotz neuer Funde wird der Vorrat an Erdöl und Erdgas zur Neige gehen.

Damit kommende Generationen nicht auf Erdölprodukte verzichten müssen, wird an der technischen Universität Wien an der Weiterentwicklung der Verwendung von Biomasse als Alternative gearbeitet. In der Arbeitsgruppe für „Zukunftsfähige Energietechnik“ von Professor Hofbauer wird unter anderem auch an einem 100 kW Zweibettwirbelschichtvergaser geforscht. Diese Technologie ist mittlerweile in ganz Europa und auch über dessen Grenzen hinweg verbreitet, zum Beispiel: Neuseeland, Deutschland, Vereinigte Staaten von Amerika, Schweden, Portugal, Frankreich, nicht nur um Strom und Wärme sondern auch ein für andere Zwecke verwendbares Synthesegas zu erzeugen. Es ist ein erklärtes Ziel der Arbeitsgruppe, diese Technologie zu verbessern und somit für industrielle Anwendungen interessant zu machen. Es gibt bereits einige industrielle Großanlagen wie zum Beispiel in Güssing in Österreich.

Um dieses Ziel zu erreichen, werden an der technischen Universität Wien laufend Versuche gemacht. Es wird auch auf den Kontakt zu anderen Universitäten und den Austausch von Ergebnissen großer Wert gelegt.

Im Zuge dieses Austausches wurde diese Diplomarbeit verfasst, um die fluiddynamischen Eigenschaften der Zweibettwirbelschichten in Christchurch, Neuseeland, und Wien, Österreich zu untersuchen. In einem 3 monatigen Aufenthalt in Christchurch wurden Daten für diese Diplomarbeit gesammelt. Um eine Vergleichsbasis zu schaffen, wurden ähnliche Bettmaterialien verwendet, deren katalytische Wirkung auf Grund ihrer Zusammensetzungen ähnlich ist. In Christchurch wurde der dort übliche Grauwacke Sand verwendet und in Wien Quarzsand. Für einen Vergleich beider Anlagen wurden im Rahmen dieser Diplomarbeit Berechnungen für signifikante Größen zur Beschreibung der Wirbelschicht gemacht. Die Ergebnisse dieser Berechnungen haben ergeben, dass die

Anlagen in Christchurch und Wien im Grunde auf die gleiche Art und Weise betreiben werden.

Unterschiede finden sich jedoch in der Geometrie und in der Betriebsweise durch einen Unterschied in der Korngröße des Bettmaterials und der Vergasungstemperatur. Diese Unterschiede haben natürlich einen Einfluss auf andere Betriebsparameter und erschweren den Vergleich. Wichtige Kennzahlen für die Wirbelschichtvergasung wie z.B. die Archimedes Zahl oder die Leerrohrgeschwindigkeit sind davon nicht ausgeschlossen.

Diese Parameter beeinflussen das Produktgas. Es gibt plausible Begründungen für die signifikanten Unterschiede in der Zusammensetzung des Produktgases.

# Table of Contents

<b>1</b>	<b>INTRODUCTION.....</b>	<b>1</b>
<b>2</b>	<b>BACKGROUND THEORY .....</b>	<b>4</b>
2.1	GASIFICATION .....	4
2.1.1	<i>Basic principles .....</i>	4
2.1.2	<i>Gasification reactions .....</i>	4
2.1.3	<i>Thermo-chemical conversion of solid fuels .....</i>	5
2.1.4	<i>Influencing parameters.....</i>	6
2.1.5	<i>Gasification reactor .....</i>	7
2.2	FLUIDIZATION.....	10
2.2.1	<i>Basic Principles .....</i>	10
2.2.2	<i>Characteristics of fluidized beds .....</i>	11
2.3	DUAL FLUIDIZED BED GASIFIER .....	15
<b>3</b>	<b>STATE OF THE ART OF DUAL FLUIDIZED BED GASIFICATION .....</b>	<b>17</b>
3.1	EXPERIMENTAL DEVICES .....	17
3.1.1	<i>Battelle Columbus.....</i>	17
3.1.2	<i>MILENA Gasifier.....</i>	18
3.1.3	<i>Heatpipe-Reformer.....</i>	20
3.1.4	<i>“New Chalmers” research-gasifier.....</i>	21
3.2	INDUSTRIAL SCALE.....	24
3.2.1	<i>Güssing .....</i>	24
3.2.2	<i>Oberwart .....</i>	25
3.2.3	<i>Villach .....</i>	27
3.2.4	<i>Klagenfurt.....</i>	27
3.2.5	<i>Sweden .....</i>	27
3.2.6	<i>Germany.....</i>	28
<b>4</b>	<b>EXPERIMENTAL .....</b>	<b>29</b>
4.1	GASIFICATION IN AUSTRIA .....	29
4.1.1	<i>Mechanical setup.....</i>	29
4.1.2	<i>Procedure.....</i>	32
4.1.3	<i>Measurement .....</i>	32
4.2	GASIFICATION IN NEW ZEALAND .....	34
4.2.1	<i>Mechanical setup.....</i>	35
4.2.2	<i>Procedure.....</i>	37
4.2.3	<i>Measurement .....</i>	37
<b>5</b>	<b>CALCULATION OF FLUID DYNAMICS.....</b>	<b>38</b>
5.1	GENERAL DATA OF THE PILOT PLANTS .....	38
5.2	COMPARISON OF RESULTS AND OPERATION PARAMETERS.....	39

6	EXPERIMENTAL RESULTS .....	43
7	CONCLUSION .....	47
	NOMENCLATURE.....	49
	ABBREVIATIONS.....	50
	REFERENCES.....	51
	TABLE OF FIGURES .....	53
	TABLE OF TABLES .....	54
	APPENDIX .....	A

## 1 Introduction

It is well known that the demand of energy is increasing. Therefore we need more energy sources but fossil fuels are limited which means we have to look for renewable energy. In Austria only 25% of the whole energy is supplied by renewable energy. Figure 1.1 shows the development of the energy used in the years from 2000 to 2007. In these years the amount of renewable energy has increased, but also the totally amount of energy consumed, so the percentage of fossil and renewable energy did not really change.

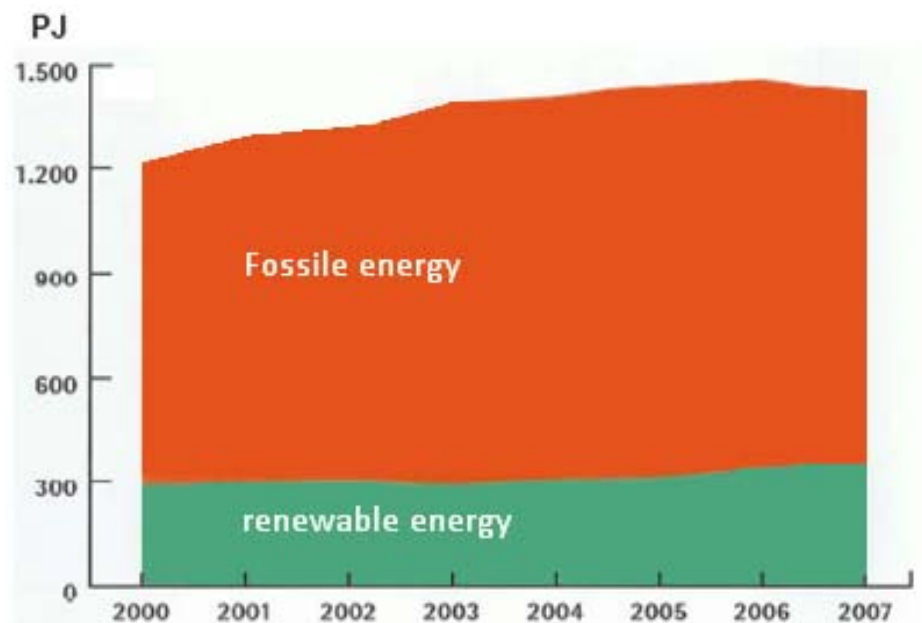


Figure 1.1: Temporal growth of energy input in Austria, [1]

There are several sources for renewable energy like hydroelectric power, photovoltaic or biomass. It is not reasonable to say that there is only one true renewable energy resource. For example, it does not really make sense to use wind energy in a place where there is hardly any wind the whole year. Another point of view is that there are times when there is no wind and no sun and we still need energy. The renewable energy source that can help in this situation could be biomass. The amount of different energy sources used is shown in Figure 1.2. About two thirds of the whole renewable energy produced in Austria is produced by biomass.

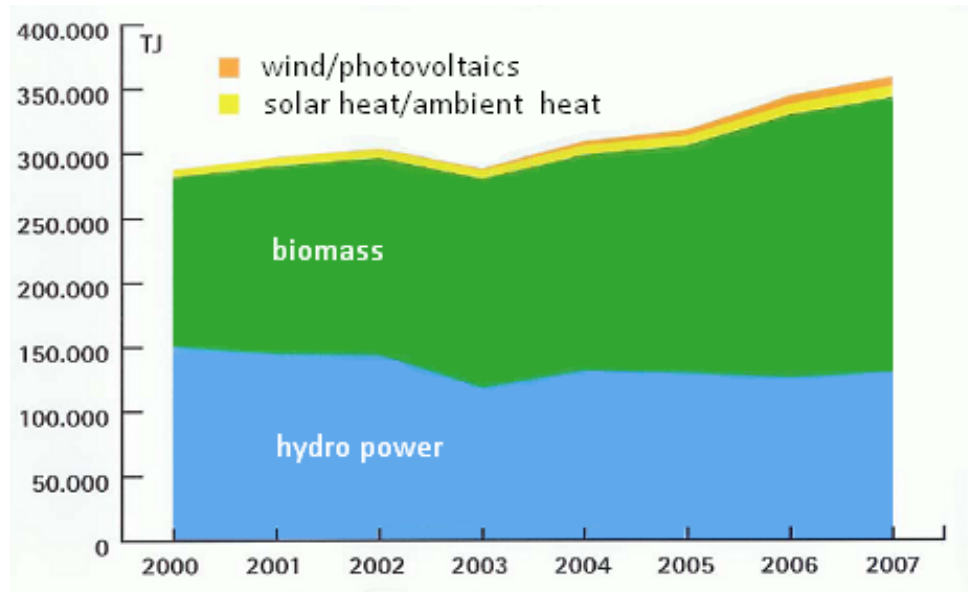


Figure 1.2: Temporal growth of renewable energy in Austria, [2]

The working group of Professor Hermann Hofbauer for “Future Energy Technology” does research on biomass as fuel for gasification. To make this technology competitive this working group searches for other useful feed into the gasifier than wood chips and also tests the influence of bed material. In cooperation with the University of Canterbury in Christchurch, New Zealand already some projects have already been successfully accomplished.

To increase the efficiency of dual fluidized bed gasifiers experiments are made all over the world. At the Vienna University of Technology different research projects are carried out. Besides experiments on gasifying waste wood and reed, the influence of the bed material and its catalytic effects are reviewed.

The task for this thesis was to investigate the influence of the fluid dynamics on the gasification in a dual fluidized bed. Therefore experiments were planned and made at the University of Canterbury in Christchurch and at the Vienna University of Technology.

One request for these tests was that these experiments need to be comparable. That means to define certain standards, for example, using the same bed material or at least ones that were as similar as possible. Also the construction of the dual fluidized bed needs to be similar. This was one reason why cooperation with the University of Canterbury in Christchurch, New Zealand was planned. With the help of the Institute of Chemical Engineering at the Vienna University of Technology a gasifier was built in Christchurch in 2006. The construction is similar to the power plant in Güssing, Austria, but it was designed to have the same nominal fuel power load as the gasifier at the University in Vienna.

To be able to compare these two pilot plants it is important to compensate the differences in construction, for example, the cross section area of the columns and to calculate some important parameters for fluidized bed gasification.

For this task important parameters were calculated. But not only characteristics, which are important for the operation of a dual fluidized bed gasifier, were calculated, but also parameters that might have an influence on the product gas. Another focus of interest was the height of bed material in the columns and the mass of bed material in the system.

## 2 Background Theory

### 2.1 Gasification

#### 2.1.1 Basic principles

Gasification is a thermo-chemical conversion of fluids or solids into a gas. The preferred components in such a synthetic gas are  $H_2$ ,  $CO$ ,  $CH_4$  [3]. But there are also components that cannot be avoided like  $CO_2$  or tar. Air, oxygen or steam is used to convert the fuels into gas. As fuel coal, biomass, plastics or other organic materials can be used. Gasification is an endothermic process so the energy for the gasification has to be provided. Mainly there are two ways to provide this energy: auto-thermally and allo-thermally

**Auto-thermal gasification** means that the energy for the gasification is produced by the fuel itself. That means that a part of the fuel is combusted. As gasification agent oxygen or air is used with an equivalence ratio between 0.3 and 0.4

**Allo-thermal gasification** uses an external fuel, which is combusted, to produce the heat that is necessary for the gasification. This heat is transferred by a heat exchanger or a circulating bed material.

Gasification is a process that takes place in a temperature range between  $500\text{ }^{\circ}\text{C}$  and  $1000\text{ }^{\circ}\text{C}$ . To reach this temperature a particle of the fuel has to pass through several other chemical reactions. First the particle is heated up, and then it is dried. Drying starts at about  $100\text{ }^{\circ}\text{C}$ . The next treatment is pyrolysis which is in a temperature ratio from  $150\text{ }^{\circ}\text{C}$  and  $500\text{ }^{\circ}\text{C}$ . [2]

#### 2.1.2 Gasification reactions

There are many different gasification agents. As a result of this fact there are also different equations, which describe the gasification reactions.

The main reactions that occur are summarized chart:

Table 2.1: Important gasification reactions [2]

Oxidation	$C + O_2 \rightleftharpoons CO_2$	$\Delta H = - 393.5 \left[ \frac{MJ}{kmol} \right]$	(R 2.1)
Partial Oxidation	$2CO + O_2 \rightleftharpoons 2CO_2$	$\Delta H = - 221.0 \left[ \frac{MJ}{kmol} \right]$	(R 2.2)
Boudouard	$C + CO_2 \rightleftharpoons 2CO$	$\Delta H = + 159.9 \left[ \frac{MJ}{kmol} \right]$	(R 2.3)
Water-gas	$C + H_2O \rightleftharpoons CO + H_2$	$\Delta H = + 118.1 \left[ \frac{MJ}{kmol} \right]$	(R 2.4)
Water-gas shift	$CO + H_2O \rightleftharpoons CO_2 + H_2$	$\Delta H = - 40.9 \left[ \frac{MJ}{kmol} \right]$	(R 2.5)
Methanation	$C + 2H_2 \rightleftharpoons CH_4$	$\Delta H = - 87.5 \left[ \frac{MJ}{kmol} \right]$	(R 2.6)
Steam reforming	$CH_4 + H_2O \rightleftharpoons CO + 3H_2$	$\Delta H = + 203.0 \left[ \frac{MJ}{kmol} \right]$	(R 2.7)
Methanation-Reaction	$CH_4 + CO_2 \rightleftharpoons 2CO + 2H_2$	$\Delta H = - 393.5 \left[ \frac{MJ}{kmol} \right]$	(R 2.8)

### 2.1.3 Thermo-chemical conversion of solid fuels

Gasification is one of five states (heating up, drying, pyrolysis, gasification and oxidation) biomass has to pass through for thermo chemical treatment.

Heating up and drying are the first states in this treatment. From 20 °C to about 200 °C the fuel is heated up and the moisture captured in the porose structure of the biomass evaporates. This is an endothermic process. While the water is evaporated the temperature of the biomass does not change a lot. Though the biomass is heated up there is no change in the structure. If the heating rate is high the moisture expands very fast and the structure can be blown up.

Pyrolysis starts at about 200 °C. For pyrolysis there must not be any oxygen. In this state the structure of the biomass is irreversibly destroyed. The volatiles (e. g.  $CO_2$ ,  $CH_4$ ,  $C_xH_y$ ,  $CO$ ) are released from the fuel. Three products are produced by pyrolysis a gas, a liquid and a solid fraction. It depends on the temperature which fraction is preferably produced. At about 500 °C almost the whole biomass is pyrolyzed.

The thermo chemical conversion between 500 °C and 850 °C is called gasification. Shown in Table 2.1 heat is required for this reaction as the overall gasification is an endothermic process. Thus, the heat requirement is supplied by partial oxidation e.g. of the carbon and hydrogen in the fuel for auto-thermal gasification (e.g. Reaction (R 2.2)) or by oxidation of the char that is produced at the allo-thermal gasification in a separate part of the gasifier (e.g. Reaction (R 2.1)). If even more energy is required additional fuel supplies this

energy. Fugacious carbon is gasified. The products of the pyrolysis react under higher temperature with the oxygen put into the system. The influencing parameters are discussed in the next chapter. [2]

From about 430 °C oxidation takes place. That means that, if there is more oxygen in the system as required for stoichiometric reaction, all the carbon will be oxidized. Instead of producing a gas that still is energy-rich like in the gasification, a gas is produced that contains mostly CO<sub>2</sub>. For whole combustion there are certain parameters that have to be taken care of: time, temperature and turbulence. If the ratio is out of place there won't be full combustion.

#### 2.1.4 Influencing parameters

Gasification is a process that is influenced by many parameters. Some important parameters should be mentioned and discussed.

**Gasification agent:** It is obvious that the gasification agent is essential influencing the composition of the product gas. For example, if the gasification agent is air the percentage of H<sub>2</sub> is lower than if steam is taken as gasification agent. One reason is that there is a lot of Nitrogen in the air that dilutes the product gas and moreover at such high temperatures the steam dissociates. As a result the calorific value of the product gas increases. Gasification with steam as gasification agent is an allo-thermal gasification therefore one has to make sure that it can transfer enough energy to the gasification area.

Paying attention to the principle of le Chatilier that some other influencing parameters are obvious

**Pressure:** Looking at the equations (R 2.3), (R 2.5), (R 2.6) and (R 2.7) we see that the concentration of CH<sub>4</sub> and CO<sub>2</sub> will increase if the pressure is increased. As a result of this fact the concentration of H<sub>2</sub> and CO will decrease. [4]

**Temperature:** At a low temperature level more combustion reaction will proceed. Is steam used as gasification agent mainly CO<sub>2</sub>, CH<sub>4</sub> will be produced and H<sub>2</sub>O passes through the gasification column. By increasing the gasification temperature, the composition of the product gas changes. At a temperature of about 800° C the concentration of CO<sub>2</sub> an H<sub>2</sub>O is significantly decreasing while the concentration of CO and H<sub>2</sub> is rising. Figure 2.1 shows the influence of the temperature on the components in the product gas.[4]

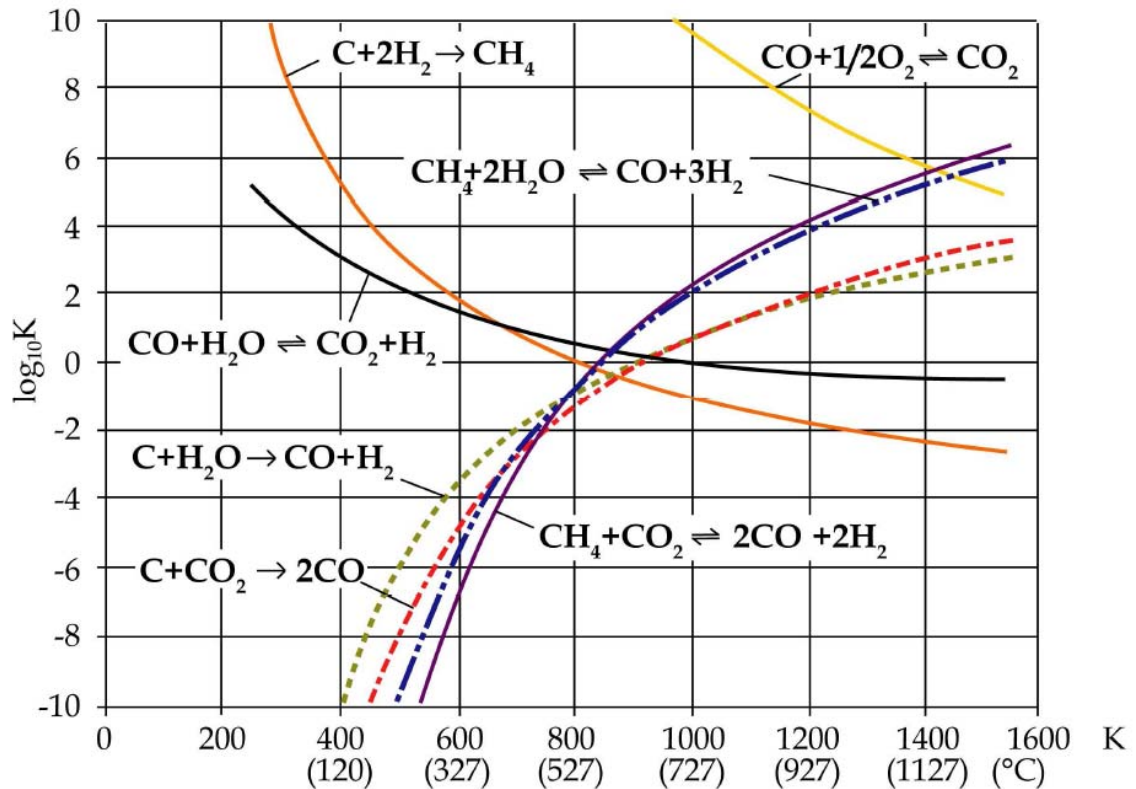


Figure 2.1: The temperature influence on the product gas, [5]

**Types of gasifier:** There is an influence of the reactor on the composition of the product gas. It is not only the size that influences the product gas but also the type of the reactor.

### 2.1.5 Gasification reactor

There are different types of gasifiers. One of the main differentiating factors is fluid dynamical conditions. Mainly there are three different gasification reactors. There are fixed bed gasifiers, fluidized bed gasifiers and entrained flow reactors.

**Fixed bed gasifier:** In a fixed bed gasifier the feedstock moves from the top of the reactor downwards under gravity in a stacked state. The gasification agent, most of the time air, is passing by the biomass. As a result different zones are formed, heating-, drying-, pyrolysis-, gasification and oxidation zone. There are different types of fixed bed gasifier. The name depends on the flow of the gasification agent compared to the biomass. Figure 2.2 shows the different types of fixed bed gasifier.

- Updraft gasifier:** In an updraft gasifier the biomass is fed into the gasifier from the top and moves downwards producing different types of products. The part of the fuel that cannot be converted into any product leaves as ash the reactor at the bottom. The air is put in at the bottom of the gasifier and moves upwards. The product gas leaves the updraft gasifier at the top. The gas is heated up in the oxidation zone. After the gasification and pyrolysis zone the gas dries the feed. So fuels with high water content can be fed into this gasifier type. The product gas produced in this type of gasifier has a high content of tar and because the hot product gas flows through the other zones and is cooled down. At lower temperature, between 200 °C and 500 °C, pyrolysis takes place and those substances produced there are entrained by the gas flow.
- Downdraft gasifier:** Compared to the updraft gasifier the air is fed into the middle of the downdraft gasifier and product gas leaves the gasifier at the bottom like the ash. The tar level in the product gas is very low, but the content of small particles like ash and char is higher than at the updraft gasifier because the product gas leaves the gasifier without being filtered by the feedstock of the biomass. Another disadvantage is that the loss of heat with the product gas is very high. So the efficiency is much lower than of the updraft gasifier.

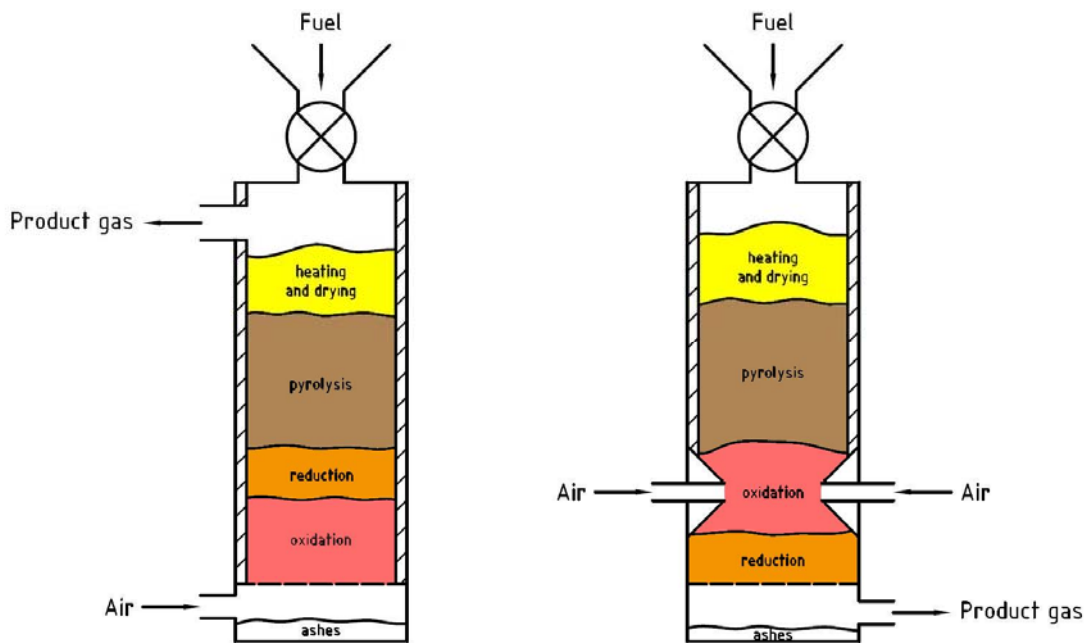


Figure 2.2: a) Updraft and b) downdraft gasifier, [6]

**Fluidized bed gasifier:** The main difference between the fixed bed gasifier and the fluidized bed gasifier is the velocity of the gasification agent. In a fluidized bed gasifier this velocity is high enough to move the particles in the gasifier. This circumstance leads to a temperature which is almost the same over the whole column. The heat transfer is done by bed material. For stabilized conditions this bed material has to be in a certain range of size. In this fluidized bed material the fuel is fed and gasified. To control the composition of product gas the bed material also can be catalytic.

Fluidized bed gasifier can be divided in to 3 different types as shown in Figure 2.3. The fluidization velocity of the gasification agent is different. If it is between the minimum fluidization velocity and the floating velocity, the particles of the bed material will be raised and form bubbles but they will not be entrained. If the velocity is increased more particles will be entrained.

The entrained particles can be transported back into the column. In this case an internally circulating fluidized bed is formed. The bed material could also be transported to another column and than a dual fluidized bed, which is detailed in chapter 2.3, is formed where the particles move from one column to another.

The gasification temperature is typically between 700 and 900° C. As a result of the small and moved particles the heat transfer between the bed material and the biomass is very good. This good heat transfer leads to a decreasing dwell time.

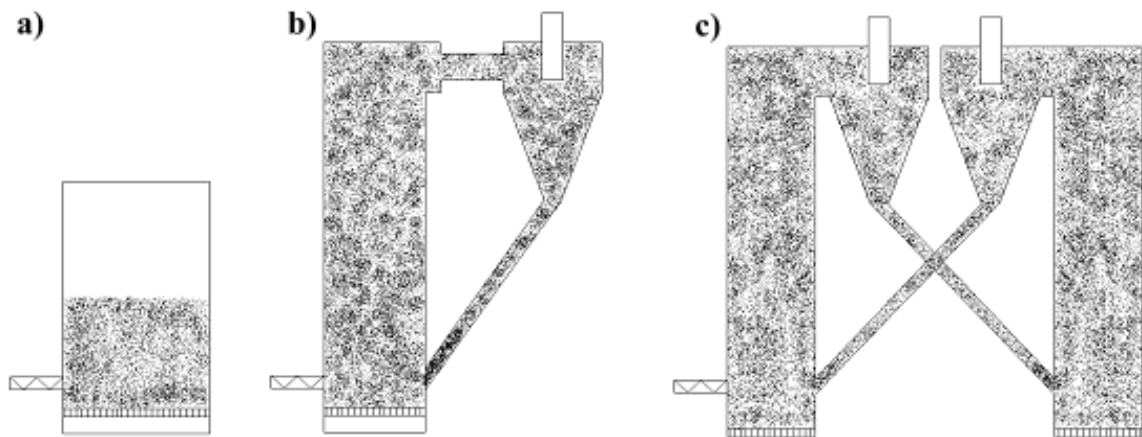


Figure 2.3: a) bubbling fluidized bed gasifier, b) circulating fluidized bed gasifier and c) dual fluidized bed gasifier, [6]

**Entrained bed gasifier:** An entrained bed gasifier works at high temperature and high flow velocity. As a result of the high speed the fuel has to be very small, so that in this short time the particles can be gasified. Biomass often is pre-treated because of the high water content and the low heating value. Entrained bed gasifier usually work at a

temperature range between 1200° and 2000 ° C. At this high temperature the ash melts and has to be removed liquid. For biomass with a low melting point this is an advantage. Depending on the ash composition the melting temperature is between 700 °C and 1500 °C.

Entrained bed gasifiers, shown in Figure 2.4, are not as important as fixed bed gasifiers, or fluidized bed gasifiers because they are very complex and expensive for small scale reactors.

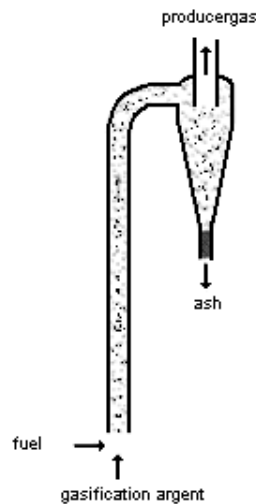


Figure 2.4: Entrained bed gasifier

## 2.2 Fluidization

### 2.2.1 Basic Principles

Fluidization means to change the behaviour of packed solid particles. When the force created by the gas that is feed into the bottom of the column is equal to the gravitation, the solid particles start to behave like a liquid. The first person, who discovered this, was F. Winkler in 1921, when he tried to gasify coal with an oxygen/steam mixture. [4] As a result of the fluidization the surface stays horizontal even if the column is tilted and if the gas stream is higher than the stream that is necessary to fluidize the particles bubbles raising like in a boiling liquid.

There are three different states of fluidization. The kind of state is related to the gas velocity. If the gas flows slow the particles stay in a packed state, the so called fixed bed. At the minimum fluidization velocity the particles start to expand so a fluidized bed is formed. If the velocity is further increased bubbles are created. This is the bubbling bed.

When the velocity is increased the particles will be entrained. That is the last state the entrained bed.

### 2.2.2 Characteristics of fluidized beds

**Particle size and bulk density:** The best way to describe the size of a particle is its diameter. The particles used are no spheres and so the diameters used for the calculations have a form factor that approximates the real size to a sphere. This calculated diameter is now called equivalent diameter. Its use makes the calculations closer to the reality. In fluidized beds the equivalent diameter is usually defined as the diameter of a sphere with the same surface to volume ratio as the real particle.

$$D_{sv} = \frac{6 \cdot V}{S} \quad \text{Equation 2.1}$$

Another important characteristic to describe a particle is its density. The particle density is defined as the mass of the particle divided by its volume.

$$\rho_p = \frac{m_p}{V_p} \quad \text{Equation 2.2}$$

The bulk density is the mass of all particles divided by the volume the particles need.

$$\rho_b = \frac{m}{V_b} \quad \text{Equation 2.3}$$

Now we can define another very important characteristic for fluidized beds, its porosity. The porosity is the ratio of the space between the particles and the whole volume. This can be calculated as:

$$\varepsilon = \frac{1 - \rho_b}{\rho_p} \quad \text{Equation 2.4}$$

Figure 2.5 shows the change of the porosity with the increase of the fluidization rate. It also shows the formation of bubbles when the volume flow is higher than the one needed for the minimum fluidization.

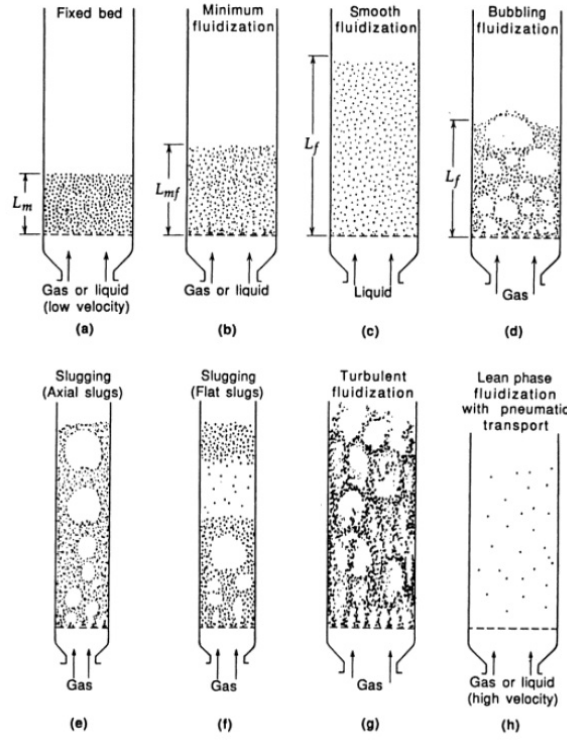


Figure 2.5: Different states of fluidization, [6]

**Pressure drop in a fluidized bed and minimum fluidization velocity:** Referring on Figure 2.5 a pressure drop occurs with the flow of a gas through a pipe. This pressure drop can be measured easily. With this pressure drop other parameters, like the mass of bed material, can be calculated.

The pressure drop in a fluidized bed column can be calculated by the following equation:

$$\Delta p = (1 - \varepsilon) \cdot (\rho_p - \rho_g) \cdot g \cdot H \quad \text{Equation 2.5}$$

The pressure drop stays approximately constant in the fluidized bed state. In that case the pressure drop is related to the mass of bed material and the area of the column. This makes the calculation of the pressure drop easier and can be calculated in this form:

$$\Delta p = \frac{M \cdot g}{A} \quad \text{Equation 2.6}$$

For the calculation of the minimum fluidization velocity  $U_{mf}$  we need another dimensionless number, the Archimedes number. The Archimedes number is independent of any process factors. The Archimedes number is defined as:

$$Ar = \frac{\rho_g \cdot d_{sv}^3 \cdot (\rho_p - \rho_g) \cdot g}{\mu^2} \quad \text{Equation 2.7}$$

With the Archimedes number we now can calculate the minimum fluidization velocity  $U_{mf}$ .

$$U_{mf} = \frac{\mu}{\rho_g \cdot d_{sv}} \cdot (\sqrt{33.7^2 + 0.0408 \cdot Ar} - 33.7) \quad \text{Equation 2.8}$$

Experiments have shown that this equation has a very large validity area

**Terminal velocity:** For a complete examination of a fluidized bed it is not enough to know when it begins to be a fluidized bed but we also need to know when it ends. Therefore we calculate the terminal velocity. This is the flow velocity where a particle is floating.

To calculate this velocity we need a force balance, shown in Figure 2.6. All forces that affect a particle must neutralize each other.

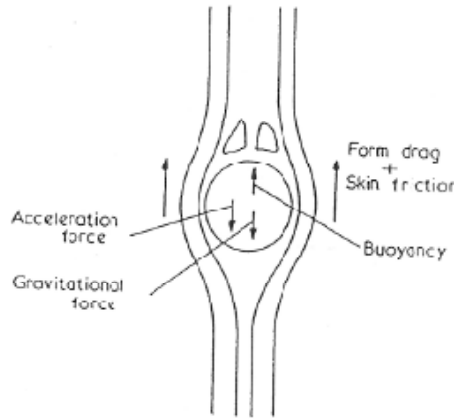


Figure 2.6: Force balance for a single particle, [7]

If we equalize all forces the equation to calculate the terminal velocity  $U_t$  is:

$$U_t = \sqrt{\frac{4}{3} \cdot \frac{\rho_p - \rho_g}{\rho_g} \cdot \frac{g \cdot d_{sv}}{C_D}} \quad \text{Equation 2.9}$$

As we see in this equation we need the drag coefficient  $C_D$ . The drag coefficient is related to the type of the stream. It is not the same, if the stream is laminar or turbulent. For small Reynolds numbers ( $Re < 0.2$ ) it can be defined as:

$$C_D = \frac{24}{Re} \quad \text{Equation 2.10}$$

Measurement showed that for high Reynolds numbers ( $Re > 1000$ ) the drag coefficient is

$$C_D = 0.43 \quad \text{Equation 2.11}$$

The calculations in this thesis showed that the Reynolds number is between 0.2 and 1000. Therefore the equation for the drag coefficient is:

$$C_D = \frac{24}{Re} + \frac{4}{\sqrt{Re}} + 0.4 \quad \text{Equation 2.12}$$

In Figure 2.7 the dependency of the drag coefficient and the Reynolds number is shown.

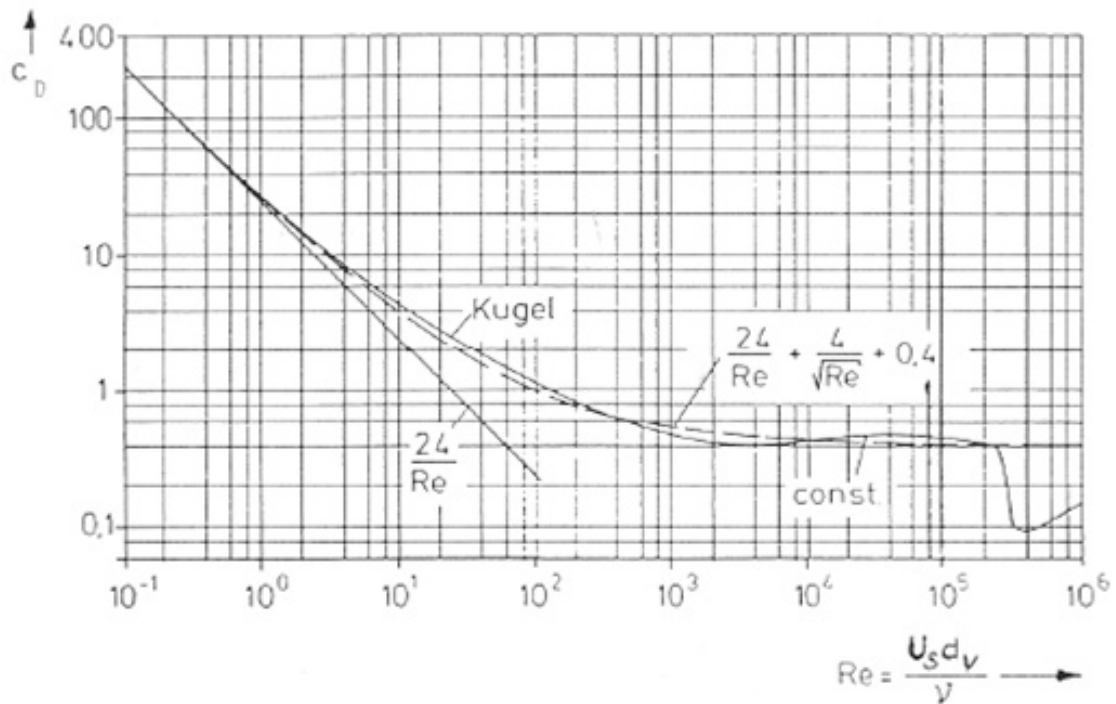


Figure 2.7: Dependency of the drag coefficient and the Reynolds number, [7]

The drag coefficient is related to the Reynolds number and the Reynolds number is calculated with the terminal velocity. As a result of this relation the true value is calculated by iteration.

To characterize the quality of a fluidized bed certain parameters are of great importance. Bi and Grace created a regime map [8] which uses the Archimedes number and the velocity. With this diagram the state of the fluidized bed can be detected easily. There is another regime map created by Reh in 1968 [9] that gives a little more information. But in this thesis only the regime map according to Bi and Grace is used because in this regime map all important values are integrated but it is not that loaded with too much information.

## **2.3 Dual fluidized bed gasifier**

As mentioned before different types of gasifiers have been invented and are used for research. This thesis concentrates on the dual fluidized bed gasifier. That means that there are two fluidized beds connected to each other. As a consequence of this connection the bed material circulates between the two columns. The bed material that is cooled down because of the endothermic gasification reaction in the gasification column is transferred to the combustion column. There it is heated up by burning the char and additional fuel and transferred back into the gasification column. These two columns are operated as one bubbling bed and one fast fluidized bed. In Vienna the combustion chamber is the fast fluidized bed and the particles are entrained and separated from the flue gas at the top of this column. Via a siphon the bed material goes back to the endothermic gasification reactor. One advantage of two separated columns is that two different fluidization agents can be used and with a good loop seal the flue gas will not dilute the product gas. That means that steam can be used for the gasification and also the char of the fuel can be burned with air. So the energy that is needed for the gasification can be provided by the fuel itself. If there is not enough char to heat up the bed material, additional fuel can be used to supply the needed energy. Another advantage is that the conditions for gasification and combustion can be set separately. That means that for example the fluidization and the temperature can be adjusted separately. But the setting of two different temperatures is not only related to the amount of combustible substances like char or additional fuel but also, and this is an even more influencing parameter, to the circulation rate. With a high circulation rate the temperatures will be close together in both columns. Because the hot bed material is transported to the gasification column but there it has not got much time to cool down because it is transported back to the combustion column very fast. With a low circulation

rate the bed material has got a lot of time to absorb the heat in the combustor and also much time to cool down. So the temperatures differ a lot. This effect can be used for the sorption enhanced reforming process for transportation of  $\text{CO}_2$  from the gasification to the combustion column. Cycled carbonation and calcination of  $\text{CaO}/\text{CaCO}_3$  at a gasification temperature below  $690^\circ\text{C}$  and a temperature in the combustor higher than  $820^\circ\text{C}$  allows selective removal of  $\text{CO}_2$  from the product gas. In any case the temperature has an influence on the product gas composition and the tar content. Figure 2.8 shows a schema of the dual fluidized bed gasification.

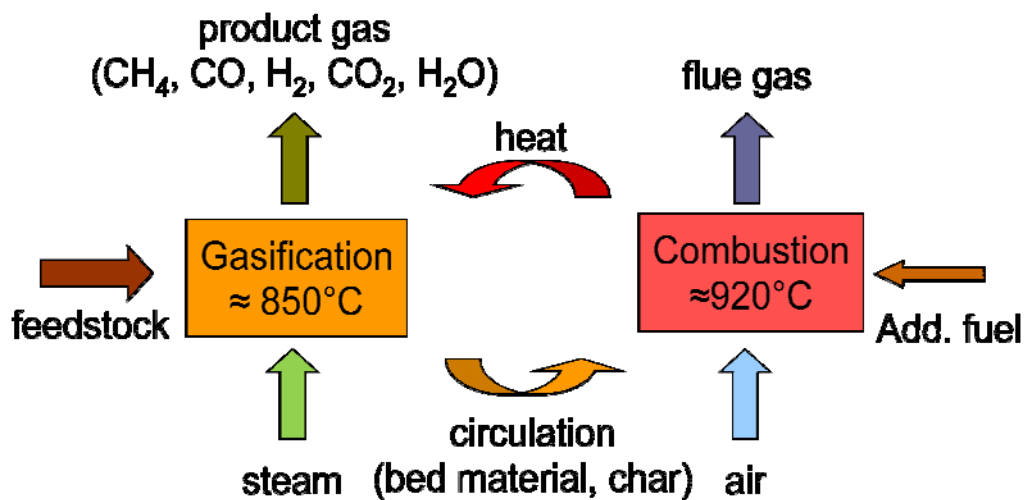


Figure 2.8: Principal of a dual fluidized bed gasifier

### 3 State of the art of dual fluidized bed gasification

The gasification technology was invented in the late nineteen hundreds. In the 1930's Hitler-Germany used this technology because of the embargo. After the Second World War and the use of oil for industrial purposes, this technology was forgotten. In the 1980's a lot of different types of gasifier were invented. The focus in this thesis is on dual fluidized bed gasifiers. So the research with regard to power plant was done for this type of gasifier.

#### 3.1 Experimental devices

##### 3.1.1 Battelle Columbus

This technology was invented by the FERCO – Company and partners. Though this technology is one of the oldest this kind of gasifier is still produced and used in the United States of America. In this concept two separated circulating fluidized bed are connected crosswise, shown in Figure 3.1. In the one column, the combustor, the char of gasifier is burned. The bed material is separated from the flue gas with a cyclone separator and is transported via a chute to the gasification column. There the hot bed material cools down and this energy is used to gasify the feed. On the top of this column the product gas is cleaned from the bed material with another cyclone separator. Again the bed material is filled back into the combustor where it is reheated again. As gasification agent steam is used and for the combustion air. [10] This is one of the oldest DFBs and so there are a lot of data for the product gas composition one possible composition is shown in Table 3.1

Table 3.1: Product gas composition of the Battelle Columbus [10]

H <sub>2</sub>	[vol%]	17.5
CO	[vol%]	50.0
CO <sub>2</sub>	[vol%]	9.4
CH <sub>4</sub>	[vol%]	15.5
C <sub>2</sub> H <sub>4</sub>	[vol%]	6.0
C <sub>2</sub> H <sub>6</sub>	[vol%]	1.1

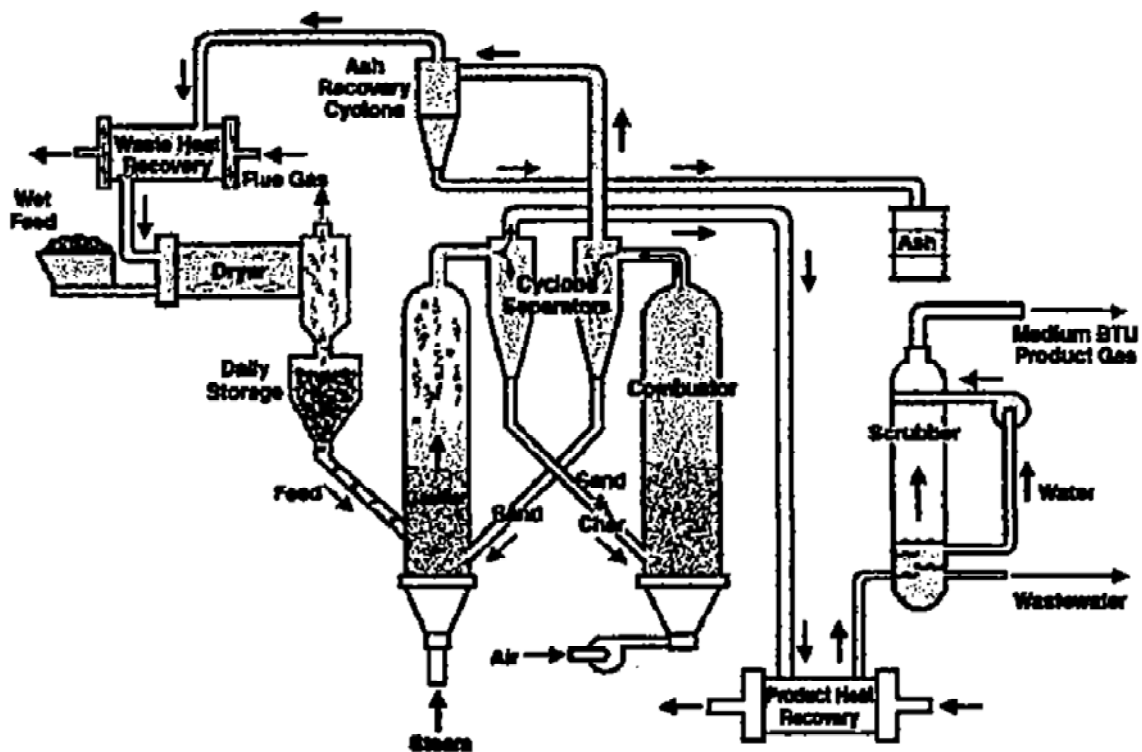


Figure 3.1: Schematic drawing of the FERCO Biomass Gasification Process, [10]

### 3.1.2 MILENA Gasifier

Another type of gasifier is the MILENA gasifier. This type was firstly built in the Netherlands by ECN. There is a big column in which a second column is integrated. So the heat transfer is not only done by the bed material but also directly by the wall that separates gasification and combustion section. The only thing is, if the wall is insulated the heat transfer might not have a big influence. But insulation increases the weight of the tubes. So the construction of this gasifier with insulation is difficult. The gasification is done in the riser and for combustion a bubbling bed is used. The advantage of this gasifier is that it is more compact and better suited for operations at elevated pressure [11]. Figure 3.2 shows a schematic drawing of the MILENA gasifier.

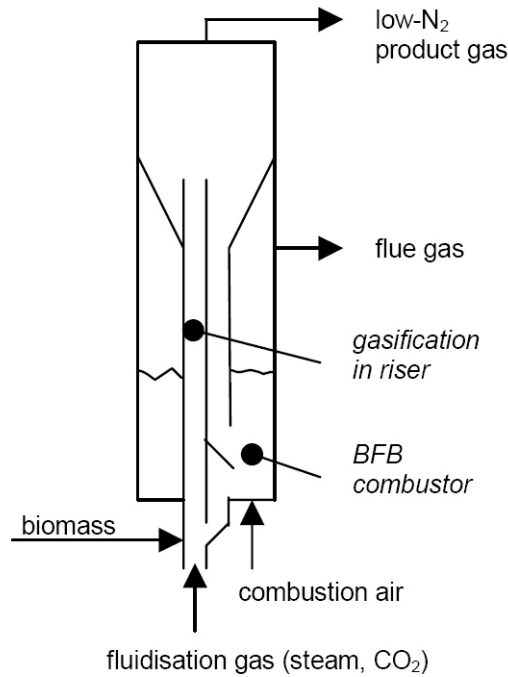


Figure 3.2: Drawing of the MILENA gasifier, [11]

*“[...] MILENA is the name of an allothermal gasification technology developed by ENC. It is a compact and easy design, which can produce medium calorific (essentially  $N_2$  free) gas already high in methane without the use of pure oxygen. The product gas contains high amounts of hydrocarbons, which makes the MILENA concept very suitable for a biomass-to-SNG plant. [...]”* [11] To compare different concepts of gasification it is important to look at the composition of the product gas, shown in Table 3.2 and also some operation parameters, shown in Table 3.3

Table 3.2: Typical gas composition of the MILENA gasifier

		Product gas	Flue gas
CO	[vol% dry]	44	
H <sub>2</sub>	[vol% dry]	18	
CO <sub>2</sub>	[vol% dry]	11	15
N <sub>2</sub> +Ar	[vol% dry]	4	71
CH <sub>4</sub>	[vol% dry]	15	
C <sub>2-5</sub>	[vol% dry]	6	
C <sub>6-7</sub>	[vol% dry]	1	
tar (C <sub>8+</sub> )	[vol% dry]	45	
H <sub>2</sub> O	[vol% dry]	25	2
O <sub>2</sub>	[vol% dry]		4

Table 3.3: Data of the MILENA gasifier

fuel capacity [kg/h]	5
bed material	silica sand
gasification agent	steam or CO <sub>2</sub>
feedstock	wood
gasifier type	riser
combustor type	bubbling bed

### 3.1.3 Heatpipe-Reformer

To provide the energy that is needed for allo-thermal gasification with steam Dr.-Ing. Jürgen Karl and Prof. Dr. Dietmar Hein invented the heatpipe-reformer at the University of Munich. Figure 3.3 shows the setup of this heatpipe-reformer. Here the fluidized beds are located on the top of each other. The lower fluidized bed is used as a combustion chamber. On the top there is the gasification area which is fluidized with steam. Between these two chambers heat is transferred by the heatpipes. The composition of the infill of the heatpipes is not published. The companies which build this reformer state that it could be sodium or potassium which is vaporized in the combustion chamber [12]. The gas rises and is condensed in the gasification area. The char of the gasification is transported via a spill and a siphon into the combustion chamber. [13]

The composition of the product gas is analyzed and the results of these measurement is shown in Table 3.4

Table 3.4: Product gas composition of the heatpipe-reformer

H <sub>2</sub>	[vol%]	~43
CO	[vol%]	~16
CO <sub>2</sub>	[vol%]	~22
CH <sub>4</sub>	[vol%]	~7

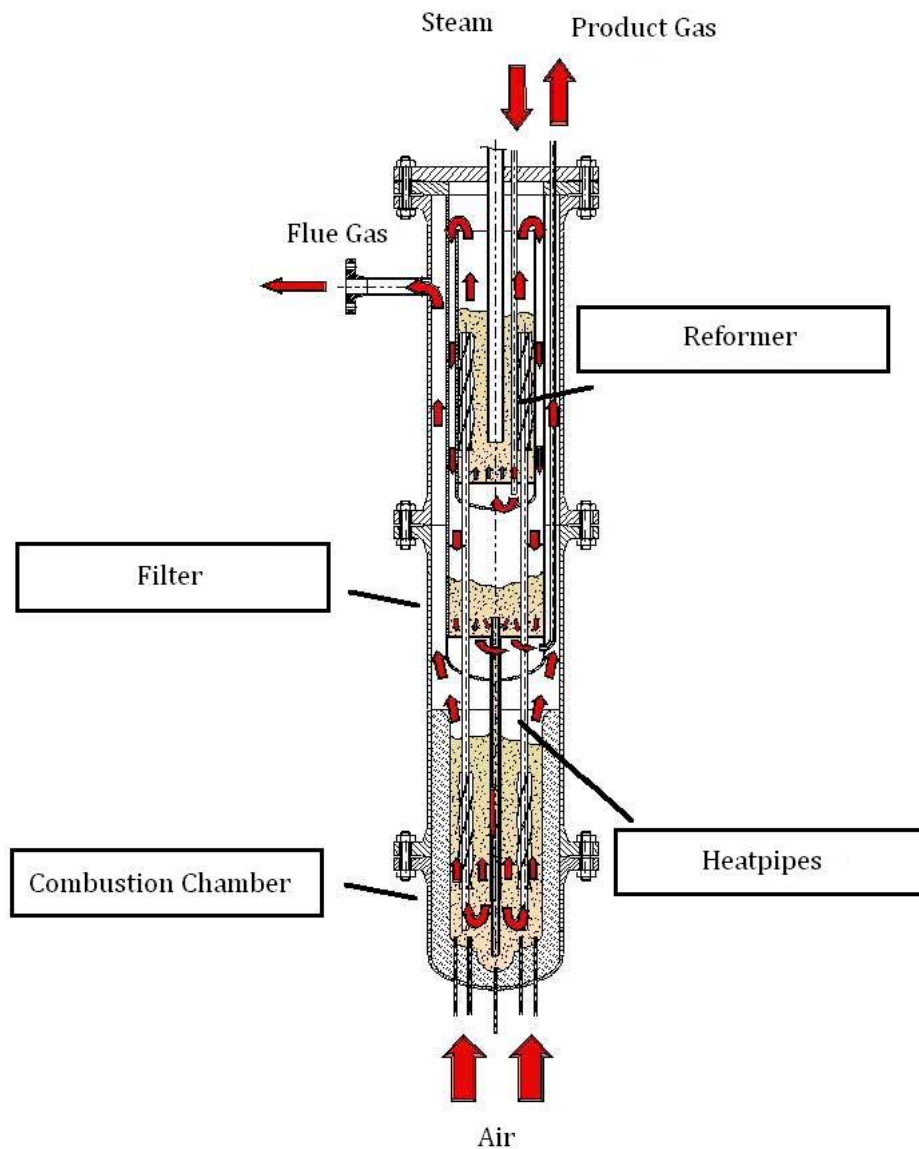


Figure 3.3: Schematic drawing of the eatpipe-Reformer, [14]

#### 3.1.4 “New Chalmers” research-gasifier

The new Chalmers gasifier is an expansion of the existing CFB-boiler. Therefore in 2007 a gasifier was integrated into the power plant, shown in Figure 3.4. As gasification agent steam is used but also other fluids for gasification have been tested. The boiler has a capacity of 12 MW<sub>th</sub>. The heat transferred by the bed material provides enough for 6 MW biomass, but to keep the temperature of the boiler at a certain level only 2 MW biomass is fed into the gasifier. In the way that gasifier is integrated the CFB-boiler still can be used without the new gasifier. In winter 2008/2009 this power plant was operated for more

than 4000 hours. In this time the gasifier was used for 250 hours with fuel and 1800 hours without fuel. [9]

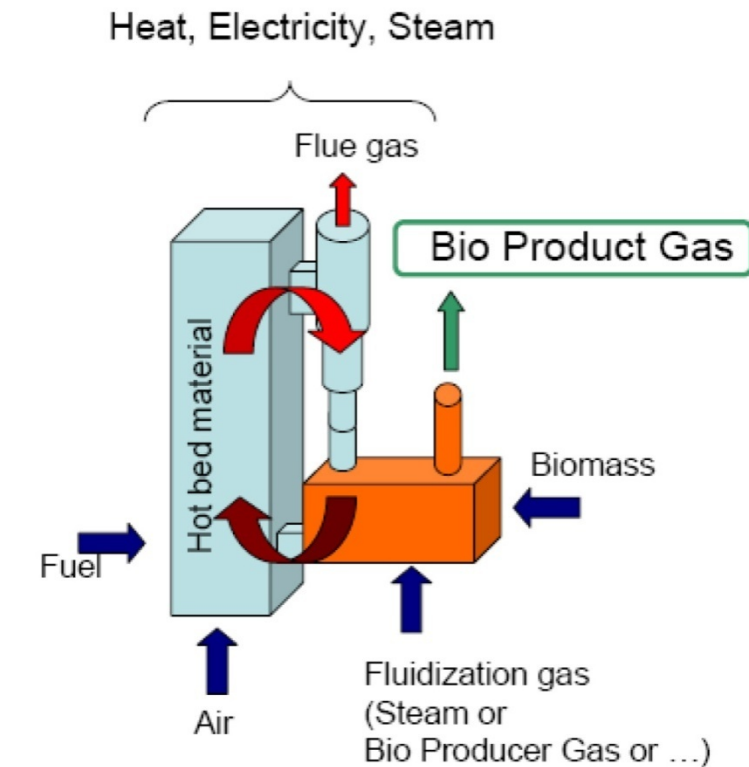


Figure 3.4: Schematic drawing of the new Chalmers research-gasifier, [9]

During the operation no negative effect on the boiler was noticed.

*"[...] The main scope of the Chalmers gasifier is to investigate the fundamentals of the gasification process in a large scale facility. For that reason, the gasifier is designed so that gasification takes place under well defined conditions. The goal is to derive results that are valid independent of reaction design and operation. [...]" [9]*

To sample the gas 12 ports are installed into the gasifier so that samples can be taken from different areas of the gasification column. The results of this sample are listed in Table 3.5.

Table 3.5: Product gas composition of the “New Chalmers” research gasifier,

H <sub>2</sub>	[mol%]	25.1
CO	[mol%]	33.1
CO <sub>2</sub>	[mol%]	14.8
CH <sub>4</sub>	[mol%]	11.8
C <sub>2</sub> H <sub>4</sub>	[mol%]	4.2
C <sub>2</sub> H <sub>6</sub>	[mol%]	0.4
N <sub>2</sub>	[mol%]	9.3
tars (gravimetric)	[g/nm <sup>3</sup> ]	7.8

There are a lot of different concepts which are quite similar to those which are already explained. For the sufficiency of this thesis these missing concepts are mentioned here with the literature where more about them can be found. J. Corella et al. give a clear view over gasification plants all over the world. [15]

There some DFBs in Japan one of them in Yokohama. It feeds 3-5 kg/h. In this pilot plant different concepts for DBFs have been tested. There are two papers published by Xu et al. which are easily available [16] and [17].

In the People’s Republic of China there are some places where DFBs are operated. There is one in Beijing with an capacity of 100 kg/h [18]. At the Zhejiang University in Hangzhou there is a 1 MW DFB gasification plant. It is running with coal but the results there could be useful for biomass gasification too. [8]



Table 3.6: Data of the gasifier in Güssing

fuel capacity [kg/h]	2000
bed material	olivine
gasification agent	steam
feedstock	wood pellets
gasifier type	bubbling bed
combustor type	riser

There is an intensive cooperation between the University of Technology of Vienna and power plant in Güssing. So a lot of things invented or discovered at the University are tested on a bigger scale in the power plant in Güssing.

### 3.2.2 Oberwart

After the first power plant in Güssing another gasifier for industrial use was built in Oberwart. Figure 3.6 shows the flowsheet of the power plant with the feeding system, the gasifier itself and the energy production section. With the experiences of Güssing some new inventions were put into effect, so this plant is more efficient than that one in Güssing. With an ORC-Process the heat is used for electricity production in two gas engines. The hospital and industrial companies use the heat that is produced in the power plant in Oberwart. If there is more need for district heating, the production of electricity can be reduced and so more heat can be exported. Table 3.7 gives an over view of characteristic data of the power plant in Oberwart [20].

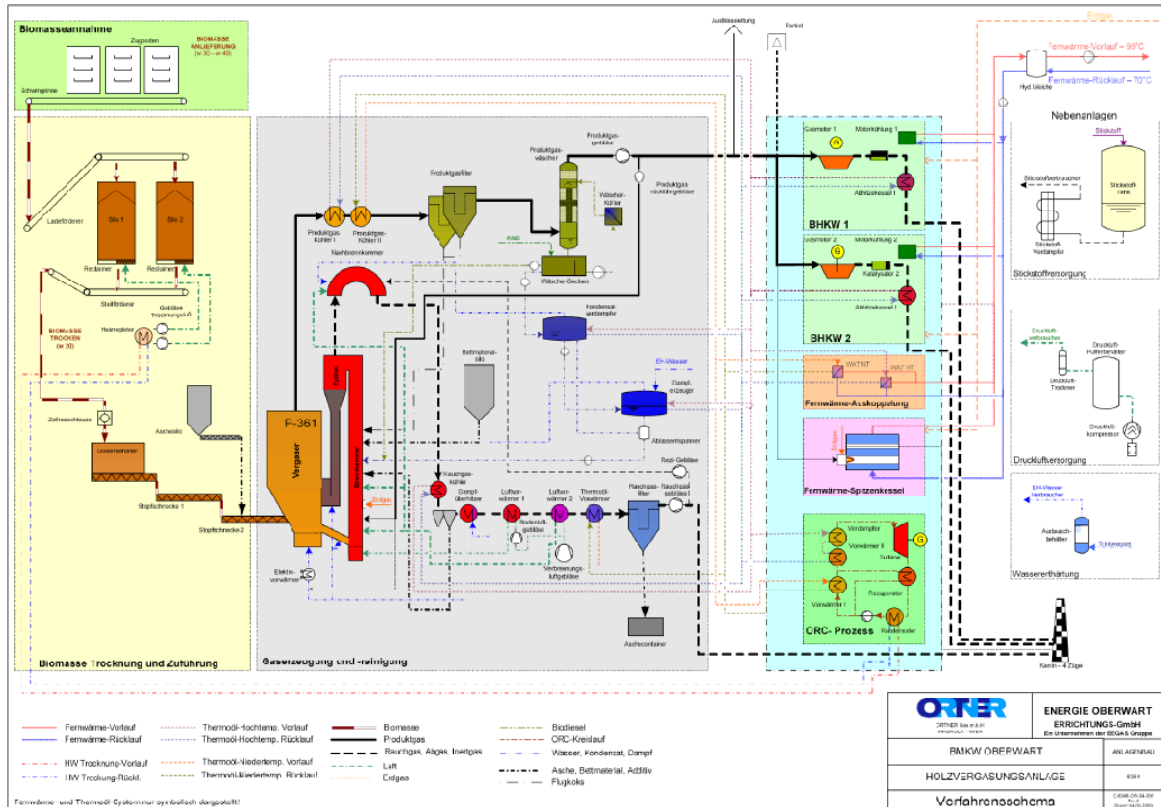


Figure 3.6: Flowsheet of the power plant in Oberwart, [20]

Table 3.7: Data of the power plant in Oberwart. [21]

fuel capacity [kg/h]	3000
bed material	olivine
gasification agent	steam
feedstock	wood chips
gasifier type	bubbling bed
combustor type	riser
$KW_{th}$	4100
$KW_{el}$	2400

There are power plants planned or right now even under construction. Table 3.8 gives an overview of power plants that use the DFB concept.

Table 3.8: Power plant in Europe

Location	Electricity production	Fuel / electr. MW, MW <sub>el</sub>	Start up	Status
Güssing, AT	gas engine	8.0 / 2.0	2002	operational
Oberwart, AT	gas engine / ORC	8.5 / 2.8	2008	operational
Villach, AT	gas engine	15 / 3.7	2010	in commissioning
Klagenfurt, AT	gas engine	25 / 5.5		detailed engineering
Ulm, DE	gas engine / ORC	15 / 5.3	2011	under construction
Geislingen, DE	gas engine / ORC	10 / 3.3		detailed engineering
Göteborg, SW	BioSNG	30 / 20 (SNG)		detailed engineering

### 3.2.3 Villach

In the past years a new gasification power plant was built in Villach in the province of Carinthia in Austria. The power plant is about to be commissioned. This gasifier is also a dual bed gasifier. The annual demand of wood, also shown in Table 3.9, is 33.000 t the expected capacity is 3900 kW<sub>el</sub> and 6700 kW<sub>th</sub>.

The gas produced in this power plant shall be used in two 20-cylinder-gas engines. [21]

Table 3.9: Expected performance of the planned gasifier in Villach.

MW <sub>th</sub>	6.7
MW <sub>el</sub>	3.9

### 3.2.4 Klagenfurt

In the capital city of Carinthia Klagenfurt another power plant is planned. The detailed engineering of this plant is happening now.

### 3.2.5 Sweden

The government in Sweden proclaimed a goal to increase the use of renewable sources. According to the speech of Möller in conference in 2010 of "Gasification – Feedstock, Pretreatment and Bed Material" [22] Sweden wants to gain up to 50% of the energy from renewable sources until 2020. Until 2030 the transport sector in Sweden will be independent of fossil fuels. Therefore new gasification plants are planned. Ingemar

Gunnarsson spoke about the GoBiGas project in Sweden. The goal of the GoBiGas project is to produce a gas out of biomass that has the quality of natural gas so that it could be fed into the gas distribution system in Sweden. To achieve this goal there are two steps. The first is to build a large scale demonstration plant in Göteborg. With the experiences of power plant Güssing and its synthesis gas production the demonstration plant in Göteborg will be highly efficient. After this phase the capacity in Sweden will be increased so that the goal of independency of the transport sector in Sweden can be achieved. More information is given by script of Gunnarssons speech [23].

### **3.2.6 Germany**

In Germany there are two power plants planned. One of them is in Neu-Ulm [24]. This power plant is about to be built and expected to start working in 2011. There is a great public interest in this power plant and a lot of information for the public has been published. To involve the people into this project the company Stadtwerke Ulm installed a webcam. Those who are interested in the progress of work can trace it online. [25]

Another power plant shall be built in Geislingen. But it is still in planning phase.

## 4 Experimental

As mentioned before experiments have been made for this thesis to compare the fluid dynamics and the product gas composition in pilot plants in Christchurch, New Zealand and Vienna, Austria. In this chapter these two pilot plants are described.

### 4.1 Gasification in Austria

This part describes the pilot plant currently in use at the Vienna University of Technology. This is also the place where the first experiments for this thesis took place.

#### 4.1.1 Mechanical setup

The gasification plant at the Vienna University of Technology is now the third generation of gasification plants. The first one was built in 1995 and the second in the year 1999. The gasifier, which is used for research now, was built in 2004 with the capacity of 100 kW. This plant is built as a dual fluidized bed gasifier in which one column is a fast fluidized bed and the other is a bubbling bed, its flowsheet is shown in Figure 4.1. The fast fluidized bed is used as combustor to heat up the bed material and transport it via the upper siphon into the bubbling bed column. The fluidization agent in the riser is air to combust the char from the gasification and if it is necessary additional fuel too.

The bubbling bed is used for the gasification and the fluidization agent is steam. The ash and not gasified carbon is transported via the siphon at the bottom to the riser.

To ensure that there are no toxic substances emitted into the surrounding air the product gas and the flue gas is burned together in an afterburner.

The size and construction is very important for the description of a gasifier and this is one of the most obvious parameters to compare the power plants at the Vienna University of Technology Austria and at the University of Canterbury in Christchurch New Zealand.

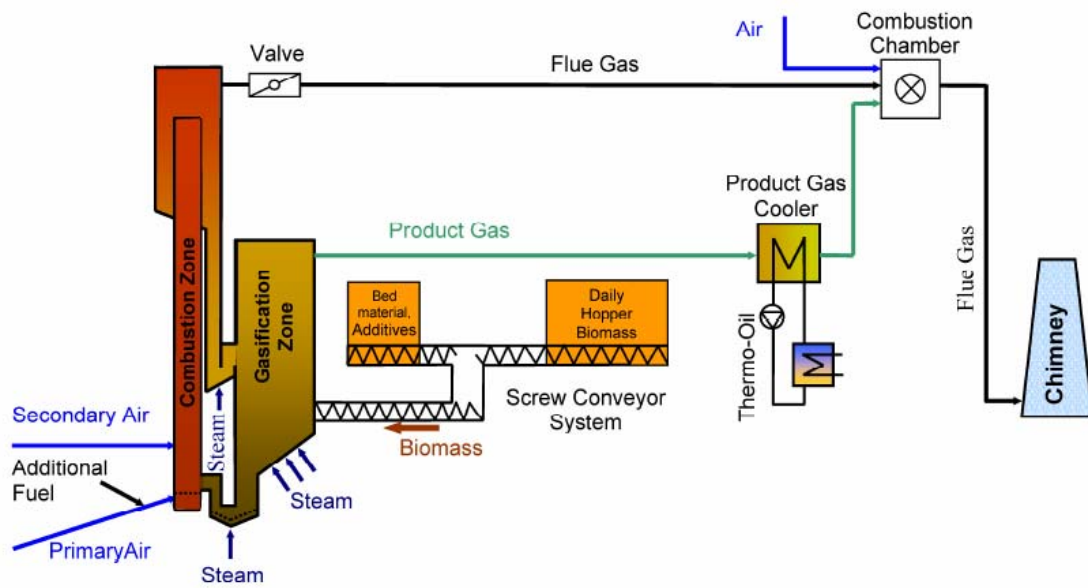


Figure 4.1: Flowsheet of the power plant at the Vienna University of Technology

The combustion column has a circular cross section area with the inner diameter of 98 mm. The column is 4800 mm high. Air is used as to fluidize the bed and burn char and additional fuel (light fuel oil). The air is preheated to a temperature of 250 °C and put into the column at high velocity so that the combustion column is operated as a fast fluidized bed and the bed material particles are entrained. The velocity is about 10.2 m/s. Primary air and fuel are fed into the riser together. Between the primary and secondary air inlets the siphon down opens out into the riser. With a separation plate the reheated bed material is separated and transported into the upper siphon and from there into the gasification area.

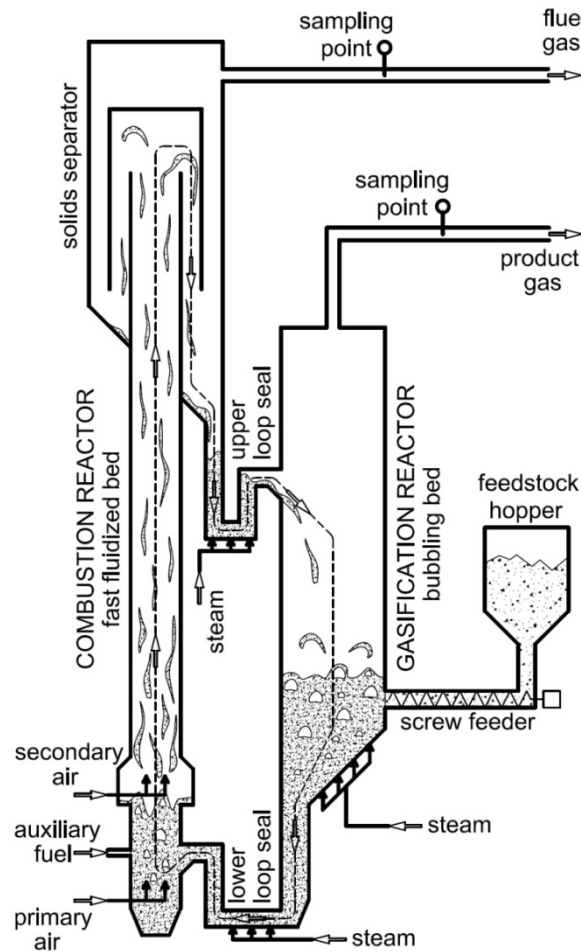


Figure 4.2: Design drawing of the gasifier at the University in Vienna, [18]

The gasification column is 269 mm in square wide and 3000 mm high. This column is operated as a bubbling bed. The fluidization agent, steam, is put into the column at the bottom. There is the siphon which is also fluidized with steam. That prohibits the flow of air from the combustion area to the gasification area. The nitrogen in the air would dilute the product gas. The only nitrogen in the product gas is from the fuel as well as from purge gas streams (e.g. hopper).

Both columns are made of high temperature resistant steel because the usual gasification temperature is between 650 °C and 1000 °C. Figure 4.2 shows the construction of the columns. To minimize the loss of energy the columns are insulated with glass wool.

The fuel is provided in 3 hoppers with different sizes. That allows to feed different fuels and mixtures of fuels into the gasifier. Most of the time the biggest hopper is filled with wood pellets. These pellets are used to heat up the power plant and also for gasification. Each hopper has an auger. It is possible to set the speed of each auger separately. These augers end in a mixing chamber where another auger feeds its content into the gasifier.

To prevent product gas to diffuse through the hoppers, nitrogen is attached to them with a volume flow of 1.6 l/h.

Runs are done with different bed materials but most of the time olivine or silica sand is used.

For the analysis the product gas, that leaves the gasification column, needs to be cooled down to 150 °C to 300 °C. This happens in a heat exchanger. The flue gas is cleaned from entrained bed material and ash with a cyclone separator.

Flue gas and product gas are separately analyzed. The product gas composition is analyzed with a 5 component online gas analyzer for H<sub>2</sub>, CO, CO<sub>2</sub>, CH<sub>4</sub>, and O<sub>2</sub> and an online gas chromatograph which analyzes N<sub>2</sub>, CO, CO<sub>2</sub>, CH<sub>4</sub>, C<sub>2</sub>H<sub>4</sub>, C<sub>2</sub>H<sub>6</sub> and C<sub>3</sub>H<sub>8</sub>. This GC runs in 20 minutes cycle. For identification of tar, SO<sub>2</sub> and NH<sub>3</sub> samples are taken. The flue gas is assayed for O<sub>2</sub>, CO<sub>2</sub> and CO

#### **4.1.2 Procedure**

This part shall explain how a trial takes place in Vienna.

The heating up of the columns starts with hot air. At about 250 °C the first of three charges of bed material is filled into the gasifier. Then wood pellets are fed into the gasifier and they start to burn. At about 500 °C the second charge of bed material is filled in. After activating the light fuel oil the third and last charge of bed material is filled into the system. When a stationary state is reached the fluidization agent for the gasification section is changed from air to steam. After reaching a stationary state after changing from wood pellets to the fuel that shall be researched, the analytic staff of the institute starts the measurements.

#### **4.1.3 Measurement**

For the measurement of the product gas composition 6 wash bottles are required. The first two bottles are empty. Then three bottles are filled with rapeseed methyl ester. The last one is empty to trap any liquid that might be entrained from the other bottles. The wash bottles are situated in a water bath with 4 °C. Then the product gas is analyzed in the gas analyzer.

Figure 4.3 shows the setup of the gas sampling line for the analysis of tar, H<sub>2</sub>S, NH<sub>3</sub> and moisture. The equipment to sample tar, H<sub>2</sub>S, NH<sub>3</sub> and moisture is the same but temperature and the solvent are different. The samples from the product gas are taken

isokinetically. Particles are separated from the gas with a cyclone and a thimble stuffed with quartz wool. The sampling line is heated so that condensation can be avoided. The gas is pumped through 6 washing bottles which are situated in a cool bath. To normalize the values the temperature and the mass flow is recorded.

For the tar measurement the washing bottles are filled with toluene and the temperature of the cool bath is  $-8^{\circ}\text{C}$ . This is different to the norm but it makes it possible to measure the water content in the gas with the same solvent. The solid fraction that is in the cyclone and the thimble are weighed and tars are extracted with isopropanol under reflux. To get the total amount of particles in the gas stream these particles are dried again. The solid fraction is now burned to get the amount of ash and char.

The water is separated from the toluene and then a sample of the toluene phase and the isopropanol phase is analyzed in a GC/MS. In a rotary evaporator the solvent is removed and by storing in a drying oven. The weight of these residues is the amount of gravimetric tar.

For the GC/MS-detectable tar tetrahydronaphthalene is added to the samples which are analyzed by a PerkinElmer Autosystem XL GC with PerkinElmer Turbomass mass spectrometer.

Ammonia is sampled with  $\text{H}_2\text{SO}_4$  with a molar concentration of 0.05 at a temperature of  $-2^{\circ}\text{C}$ . With ion chromatography the amount of ammonium in the solution is analyzed and the concentration in the gas phase can be calculated.

For the measurement of Hydrogen sulphide the bottles are filled with 35% KOH at a temperature of  $-2^{\circ}\text{C}$ . The analysis of the fluid is based on the standard ISO 6326-3.

In the flue gas  $\text{SO}_2$  is measured from the wet, dust-free flue gas.  $\text{N}_2$ ,  $\text{CO}_2$ ,  $\text{CO}$ ,  $\text{O}_2$  is analyzed with a gas analyzer.

Further information is given by Aigner et al. 2011 [26]

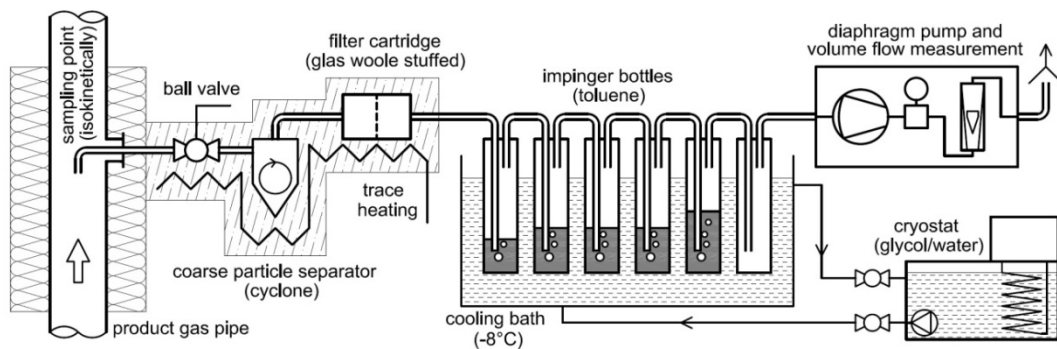


Figure 4.3: Analysis equipment

## 4.2 Gasification in New Zealand



Figure 4.4: Photo of the gasifier in Christchurch, New Zealand, [4]

Tests have been taken at the dual fluidized bed gasifier of the University of Canterbury in Christchurch, New Zealand. Figure 4.4 shows a photo of this gasifier and the combustor. This gasifier is slightly different to the gasifier at the Vienna Technical University. This part of the thesis will explain the gasifier in New Zealand and the handling of it.

#### 4.2.1 Mechanical setup

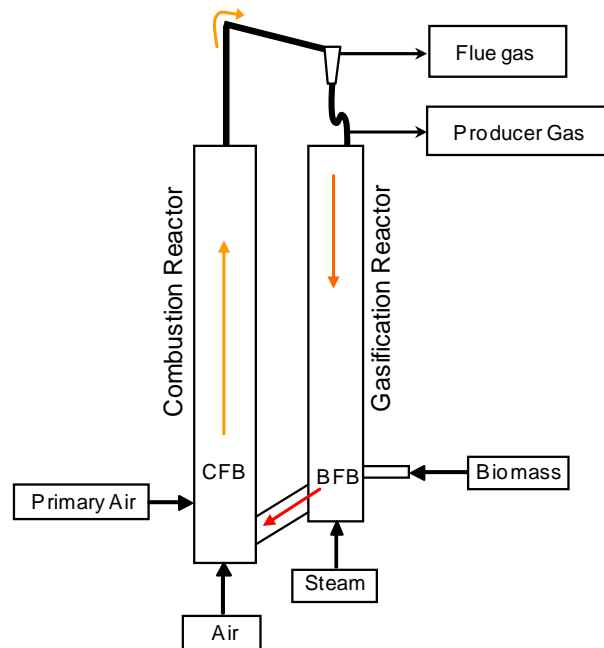


Figure 4.5: Schematic drawing of the pilot plant in Christchurch, [27]

There are a lot of differences in the structure of the gasifier in Christchurch and in Vienna.

Figure 4.5 shows the flowsheet of the pilot plant. The riser has an inner diameter of 107 mm and is 3700 mm high. Air is used as fluidization agent. In the riser the char is burned together with the additional fuel (natural gas). The riser is operated as a fast fluidized bed.

The 800 °C hot bed material is separated from the flue gas stream with a cyclone separator and the bed material moves via a siphon to the gasification area.

The gasification column has an inner diameter of 200 mm and is 2000 mm high. This column is operated like the gasifier in Vienna as a bubbling bed. Steam is used as gasification agent. In contrast to Vienna the bed material leaves the gasification column via a chute and not via a siphon. This chute and the upper siphon are fluidized with steam.

Both columns are made of steel with a refractory of cement. Figure 4.6 shows a photo of the inside of the riser. The usual temperatures are between 650 °C and 850 °C.



Figure 4.6: A view into the riser at the University of Canterbury, Christchurch

The fuel is provided in one hopper. Usually this hopper is filled with wood pellets. This hopper is rather small but it can be refilled with a sewer port while running the gasifier. Figure 4.7 shows a photo of the sewer port. To prevent any gas to leave the hopper nitrogen is attached to it with 5 l/min. There is a long auger that transports the bed material from the hopper into a mixing chamber. From there it is fed into the gasification column with another auger.



Figure 4.7: Mechanism to refill the hopper.

Usually greywacke is used as bed material. The composition of greywacke is similar to silica sand. For details see Appendix.

#### **4.2.2 Procedure**

The way the gasifier is heated up is different to the procedure in Vienna. The whole bed material is in the columns right from the start. The LPG is stitched on in the gasifier and the riser. The air used to burn the LPG is electrically preheated. When the system reaches the requested temperature the LPG in the gasifier is switched off and the fluidization agent is switched to steam. Unfortunately there is no software that records the mass flows, temperatures and pressures in the columns automatically so these things have to be recorded manually. But this software is about to be established.

#### **4.2.3 Measurement**

The moisture is measured gravimetrically. One cubic meter is extracted from the product gas through a U-tube filled with silica gel. This tube is weighted before and after the extraction. The difference in weight is a proportion for the moisture in the product gas. For this kind of measurement the room temperature has to be measured because water may condense when the gas cools down. To compensate this failure this procedure is necessary. To analyze the product gas and the flue gas samples are taken and injected into a GC. Tar is measured gravimetrically. Therefore a Bakerbond amino normal phase SPE column is used [28]. For measurement of  $\text{NH}_3$  and  $\text{H}_2\text{S}$  tubes from the company DRÄGER are used. These tubes are operated as specified in the manual. Product gas is pushed through the tube and the change of colour in the tube proceeds the higher is the content of ammonia and hydrogen sulfide.

## 5 Calculation of fluid dynamics

The main goal of this work was to calculate the fluid dynamical conditions in the gasifier in Vienna and Christchurch. With the equations, mentioned in section “theoretical background”, calculations were made. This chapter shall compare the fluid dynamic of the gasifier in Vienna and Christchurch.

### 5.1 General data of the pilot plants

General operational and geometrical data of the reactor system in Vienna and Christchurch are tabled in Table 5.1 and Table 5.2.

Both systems are based on the dual fluidised bed reactor system and designed for a nominal fuel power input of 100 kW<sub>th</sub>. However, the pilot plants differ marginally in geometry and operable ranges.

The range for the particle size is the same though normally a smaller fraction of bed material is used in Christchurch. The reactor height and cross section are slightly bigger in Vienna but the gas residence time in Christchurch is longer. A comparison of both sieve analysis is shown in the Appendix.

The operable temperature ranges are different, due greywacke material, which has a low agglomeration point. This temperature lies under 900 °C

The superficial gas velocity differs marginally due to: slightly different fluidization feed and different cross section area. Also the ratio  $U/U_{mf}$  is different, as in Christchurch and Vienna University of Technology different particle size ranges are applied.

Table 5.1: Typical operating parameters in Vienna

nominal fuel power load	[kW]	100	
feedstock particle size	[mm]	up to 40 (limited by feeding system)	
applicable bed material particle size	[μm]	200 – 600	
		gasifier	riser
cross section area	[m <sup>2</sup> ]	0.072	0.008
operable temperature range	[°C]	650 – 870	750 – 920
fluidisation agent		steam	air
fluidisation regime		bubbling fluidised bed	fast fluidisation
steam/fuel ratio	[-]	0.5–1.2	n/a
U (superficial gas velocity)	[m/s]	~0.4	~9
ratio $U/U_{mf}$	[-]	3 – 4	80 – 85
ratio $U/U_t$	[-]	$U \ll U_t$	2 – 3

Table 5.2: Typical operating parameters in Christchurch

nominal fuel power load	[kW]	100	
feedstock particle size	[mm]	~25	
applicable bed material particle size	[ $\mu\text{m}$ ]	180 – 700	
		gasifier	riser
cross section area	[m <sup>2</sup> ]	0.031	0.009
operable temperature range	[°C]	700 – 780	800 – 850
fluidisation agent		steam	air
fluidisation regime		bubbling fluidised bed	fast fluidisation
steam/fuel ratio	[-]	0.72	n/a
U (superficial gas velocity)	[m/s]	~0.2	8 – 9
ratio $U/U_{mf}$	[-]	3 – 6	100-250
ratio $U/U_t$	[-]	$U \ll U_t$	3 – 4

## 5.2 Comparison of results and operation parameters

For a better understanding of the two gasifiers Table 5.3 shows some typical data. There is a big difference in the Archimedes number that is shown in Table 5.4. Reasons for this can be found in the different size of the particles and its density. A very interesting thing is that in Christchurch only 18 kg of bed material is used. These are only about one fifth of the bed material used in Vienna. This is interesting because there is just a little difference between the area of the gasification columns and an even smaller difference between the combustion columns.

There is also a significant difference in the pressure drop in the columns. The explanation for this is the mass difference. What is remarkable is that the pressure drop in the riser in Christchurch is higher than in Vienna. One reason for that is on the one hand that there is more bed material in the riser in Christchurch, measured after an experiment, than in Vienna. It is possible to measure this because the bases can be removed. On the other hand there are parameters which are different e.g. the particle size and the cross section area.

The size of the particles has an influence on the whole system e.g. the terminal velocity or the Archimedes number. In the calculation it was noticed that the temperature difference does not have a big influence on the calculation. The calculated height of the bed material shows that the biomass is not fed into the bubbling bed. But the calculation is only done for the minimum fluidization velocity because the porosity is hard to calculate for other conditions.

Table 5.3: Comparison of mechanical parameter

		Vienna		Christchurch	
		Gasifier	Riser	Gasifier	Riser
height	[m]	3	4.8	2	3.7
cross section area	[m <sup>2</sup> ]	0.072	0.008	0.031	0.009
temperature	[ °C]	850	864	750	850
pressure drop	[Pa]	7500	3600	3800	4200
particle size	[μm]	450		250	
mass bed material	[kg]	100		18	
height of bed	[m]	0.527	0.253	0.242	0.268

Important parameters for fluidized beds are the ratio of the superficial velocity  $U$ , the ratio between the superficial velocity and the minimum fluidization velocity  $U/U_{mf}$  and  $U/U_t$  the ratio between the superficial velocity and the terminal velocity. There is not much difference in these parameters for the gasification column between Christchurch and Vienna. But for the riser there this ratio in Christchurch is about twice the ratio in Vienna. Also important for the gasification is the steam to carbon ratio.

Not only for the calculation but also for the whole gasification process the fluidization of the connection between the gasification column and the riser is important. The reason for this is the function this connection has. One function is to transport the bed material from one column to the other. Another function is to prevent gas to flow from one to the other column. This is even more important for the air not to come into the gasification column because the nitrogen in the air would dilute the product gas and degrade the quality of the product gas. Therefore both Christchurch and Vienna use steam these connections.

The primary air and secondary air are an indication for the combustion in the riser. Again the value in Christchurch is higher than in Vienna. It is the same reason as for the pressure drop, the bigger amount of bed material.

Table 5.4: Comparison of process parameter

		Vienna		Christchurch	
Archimedes number	[-]	279.287	357.488	59.1982	62.691
steam to carbon ratio	[-] <sub>kg</sub>	1.672		1.455	
ratio U/U <sub>mf</sub>	[-]	3.315	143.358	11.259	376.398
ratio U/U <sub>t</sub>	[-]	0.060	2.726	0.160	5.392
superficial gas velocity	[m/s]	0.260	10.193	0.293	8.470
steam fluidisation gasifier	[kg/h]	9.800		6.000	
steam fluidisation for upper	[kg/h]	3.000		2.000	
steam fluidisation for lower siphon	[kg/h]	6.500		2.000	
primary air	[Nm <sup>3</sup> /h]		5.038		17.700
secondary air	[Nm <sup>3</sup> /h]		50.190		44.700

Another way to characterize fluidized beds is to use the regime map of Bi and Grace. [8] The calculated ratios are plotted in this diagram.

Figure 5.1 shows the fluidized beds in the regime map according to Bi and Grace. Both gasification columns, Christchurch and Vienna, are over the minimum fluidization velocity. That means that they are operated as a bubbling bed. They could be fluidized with an even higher velocity but that would just dilute the product gas.

The regime map of Bi and Grace shows the significant entrained velocity to make sure the particles are entrained. But an increase of the velocity is not only positive. Here are some reasons to choose the velocity as low as possible:

- The erosion of the columns is increasing with higher velocity.
- For combustion a certain time is required.
- If the velocity is too high bed material could be transported out of the column.
- For a high velocity a high volume flow is required.

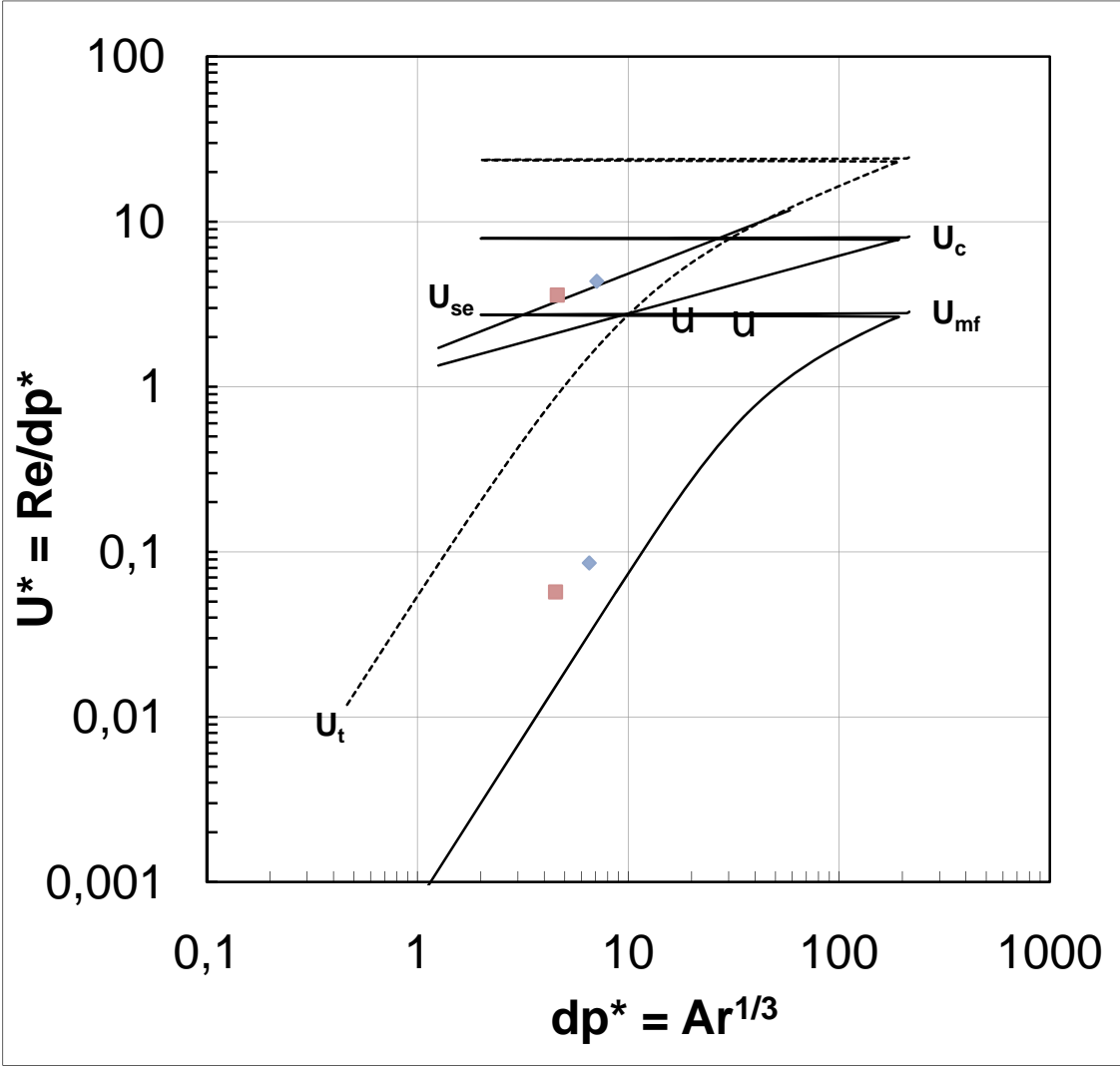


Figure 5.1: The dual fluidized bed in the regime map according to Bi and Grace [8]

## 6 Experimental results

For this thesis experiments have been carried out in Christchurch and Vienna. Both pilot plants were operated with a similar bed material. In Christchurch the bed material was greywacke sand and in Vienna silica sand.

Table 6.1 shows the composition of the product gas in the trials done for this thesis. For a better understanding and to get a better look the composition of the product gas is also shown in Figure 6.1 and Figure 6.2. The reason for two diagrams is that in Christchurch the content of nitrogen is much higher than in Vienna. For a better comparison the nitrogen was eliminated from the composition. The result of this calculation is shown in Figure 6.2.

The concentration of tar in the product gas differs quite a bit. In Christchurch it is about twice as high as in Vienna. The reason for that seems to be the lower gasification temperature in Christchurch [18]. Unfortunately dust and char was not measured in Christchurch so they cannot be compared.

The moisture in the product gas in Christchurch is slightly higher than in Vienna. The moisture samples have a certain variation and also the variation in the results of the measurement has to be considered. So this might be the main reason for the differences in the moisture.

Table 6.1: Comparison of the composition of the product gas produced in Vienna and Christchurch

Component	Unit	Vienna	Christchurch
H <sub>2</sub>	[vol.% <sub>db</sub> ]	34.69	25.5
CO	[vol.% <sub>db</sub> ]	34.19	26.3
CO <sub>2</sub>	[vol.% <sub>db</sub> ]	13.63	20.0
CH <sub>4</sub>	[vol.% <sub>db</sub> ]	11.07	9.9
C <sub>2</sub> H <sub>4</sub>	[vol.% <sub>db</sub> ]	3.01	3.3
C <sub>2</sub> H <sub>6</sub>	[vol.% <sub>db</sub> ]	0.17	0.6
N <sub>2</sub>	[vol.% <sub>db</sub> ]	2.98	14.5
tar	[g/Nm <sup>3</sup> <sub>db</sub> ]	8.22	16
tar GC/MS	[g/Nm <sup>3</sup> <sub>db</sub> ]	10.82	n/a
dust	[g/Nm <sup>3</sup> <sub>db</sub> ]	2.76	n/a
char	[g/Nm <sup>3</sup> <sub>db</sub> ]	16.82	n/a
H <sub>2</sub> O	[vol.%]	32	33.8
fuel		wood pellets	wood pellets
bed material		silica sand	greywacke

The composition of the product gas in Christchurch and Vienna seems very different at a first glance. The main reason for that is the nitrogen content in the gas. There are some reasons for that. One reason is the high pressure of nitrogen that is applied on the hopper but even with these 5 litres per minute it not enough to explain this high content. Another reason might be a leakage from the combustion column to the gasification column. In this case also other components are leaking to the gasification area e.g. oxygen or carbon monoxide. That could lead to a reaction in the gasification area and hydrogen and carbon monoxide are oxidized. This could explain the higher content of carbon dioxide in the product gas in Christchurch. Also the bed material could have some catalytic effect on the gasification and could support the reaction of  $H_2$  and CO. Generally catalytic effects of the bed material could be the absorption of  $CO_2$  or the assistance of the production of CO or the increase of dissociation of  $H_2O$ . There is a higher fraction of  $Fe_2O_3$  in greywacke than in silica sand and there is no CaO and  $K_2O$  fraction in the silica sand. The composition of the bed materials is shown in the Appendix. It can be assumed that these parts of the greywacke are supporting the CO-shift reaction. This catalytic effects are described by Christoph Pfeifer et al. [29].

Other components like  $CH_4$ ,  $C_2H_4$  or  $C_2H_6$  are at the same level in both power plants.

In chapter 5 it was mentioned that the wood pellets are not fed into the bubbling bed but the result of the gasification shows that the wood pellets have a very good contact with the bed material.

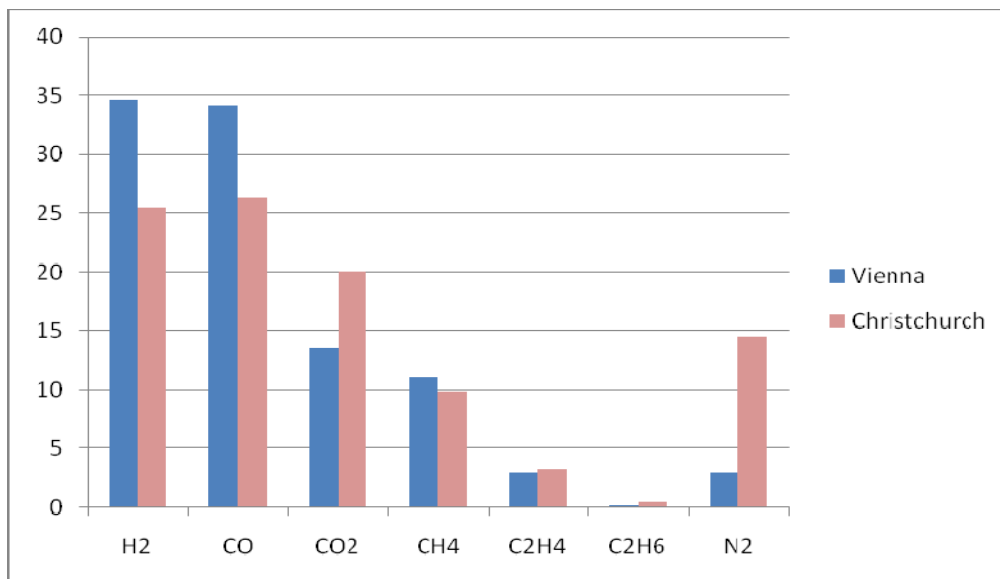


Figure 6.1: Composition of the product gas as measured

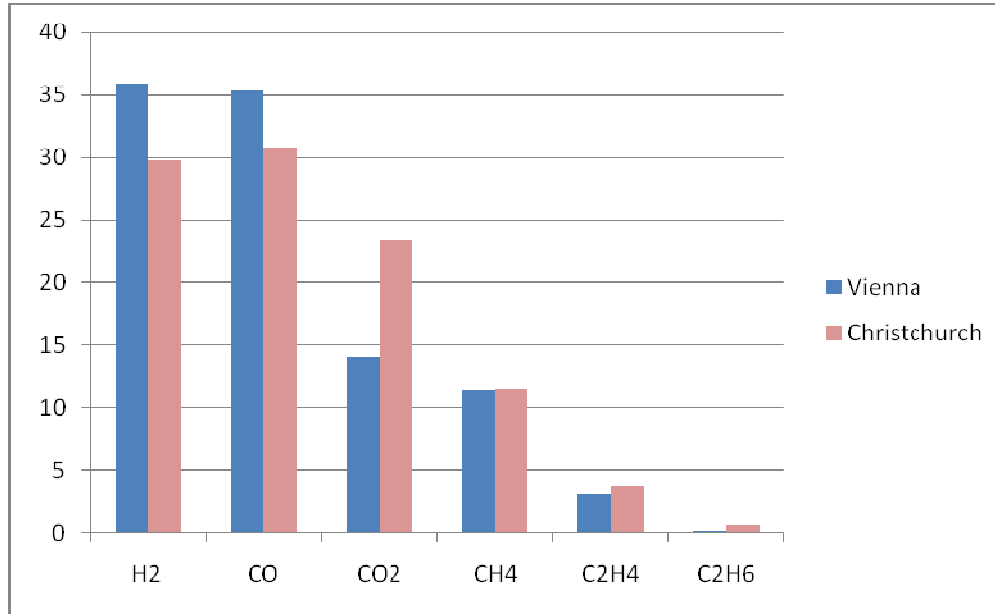


Figure 6.2: Composition of the product gas calculated on zero nitrogen

The composition of the flue gas is shown in Table 6.2 and Figure 6.3. As shown in this Table and Figure the composition is quite the same.

Carbon dioxide is next to nitrogen the component with the highest percentage in the flue gas. There is still oxygen in the flue gas and the content of carbon monoxide is almost zero. The small amount of CO in the flue gas is a result of the char particles that are circulating in the system because the residential time is not enough to burn them. But they keep on oxidizing so near the separator still some CO is produced.

Table 6.2: Comparison of the composition of the flue gas produced in Vienna and Christchurch

Component	Unit	Vienna	Christchurch
CO	[vol.% <sub>db</sub> ]	0.05	0.0
CO <sub>2</sub>	[vol.% <sub>db</sub> ]	13.72	15.2
O <sub>2</sub>	[vol.% <sub>db</sub> ]	4.25	4.8
N <sub>2</sub>	[vol.% <sub>db</sub> ]	81.99	80

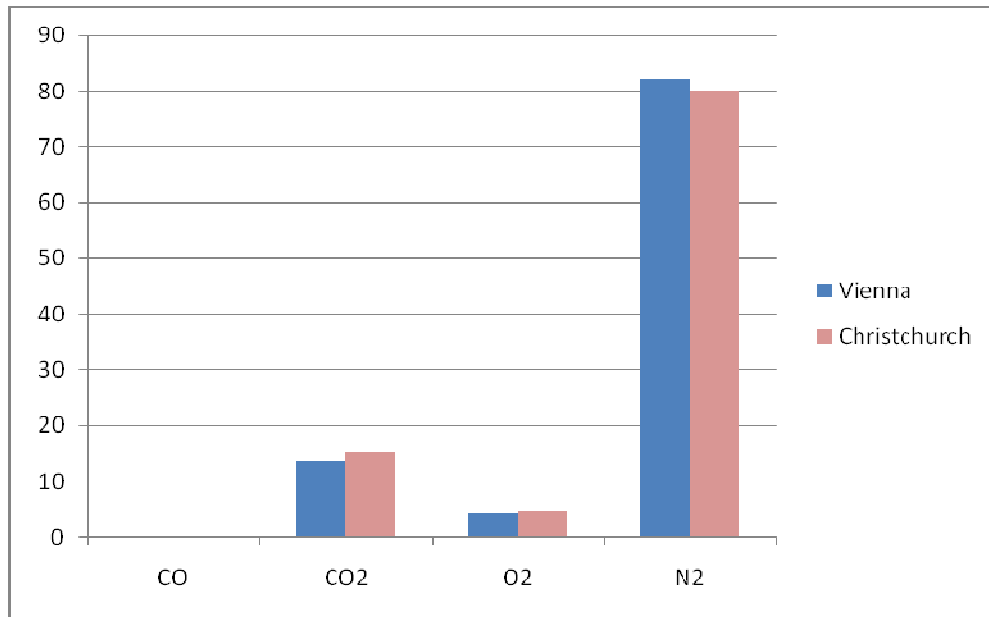


Figure 6.3: Composition of the flue gas

## 7 Conclusion

In this thesis calculations and collected data from experiments in Christchurch and Vienna are compared. To simplify the demonstration of these data graphics and tables were made. The results are summarized in this chapter.

For this thesis experiments at two dual fluidized bed gasifiers have been made. A DFB is a mechanical set up of two columns which are connected. In this work the experiments took place in Austria and New Zealand where these DFBs are operated with a bubbling bed for gasification and a fast fluidized bed as the combustion area

The calculation in this thesis shows that there are differences in operation and in geometry between the two gasifiers Christchurch and Vienna. Some of these differences are rather small and can be neglected. Others are big and therefore cannot be neglected.

One main discovery during the work for this thesis was that there is a leakage between the combustion column and the gasification column. Only with a leakage the high content of nitrogen in the product gas can be explained. With the calculation to zero nitrogen the composition of the product in Christchurch and Vienna is similar and if one can assume that not only nitrogen but also oxygen is passing through the leakage, which certainly happens, the difference in carbon monoxide, carbon dioxide, hydrogen and the moisture can be explained. But this leakage is not the only reason for the differences. Also the different temperature is an explanation because in New Zealand the gasification temperature is not as high as in Austria. One reason for this lower temperature is that the bed material in New Zealand has a melting point lower than 900 °C. That means that in the combustion column the bed material cannot be heated up higher than this temperature and so the gasification column cannot be supplied with bed material of the same energy content as in Austria. The different bed material is also a reason for the different product gas composition because the bed material in New Zealand is more catalytically active than that one in Austria. The  $F_2O_3$ ,  $CaO$  and  $K_2O$  in the bed material support the CO-shift reaction.

Another finding was that the content of tar in the product gas is higher in Christchurch than in Vienna. This is interesting because there is an influence of the temperature and also the particle size. Higher temperature and smaller particles lead to lower tar content. It seems that the influence of the temperature is greater than the influence of the particle size.

One main issue that should be checked in this thesis was the difference in the mass of bed material in the two power plants. The mass differs in a fifth. But it seems that the

influence of mass of bed material is not big on the gasification reaction as long as the fuel is well worked into the bed material and has a good contact with it so that the heat can be well transferred.

The combustion column is operated in the same way and as the composition of the flue gas shows the combustion of char and additional fuel is well done.

With the impact of the earthquake at the 4<sup>th</sup> of September in Christchurch, New Zealand the collection of the data has been negatively affected. However in this thesis influences of the fluid dynamics on the dual fluidized bed gasification were discovered. The influence of the fluid dynamics is an interesting research topic which offers possibilities for a lot of scientific acknowledgement. Therefore more research seems to be of great profit.

## Nomenclature

$A$	Area
$Ar$	Archimedes Number
$C_D$	Drag Coefficient
$d_{sv}$	Equivalent Diameter
$M$	Mass of Bed Material
$Re$	Reynolds Number
$S$	Surface
$U$	Superficial Gas Velocity
$U_{mf}$	Minimum Fluidization Velocity
$U_t$	Terminal Velocity
$V$	Volume
$V_b$	Bulk Volume
$V_p$	Particle Volume
$\Delta p$	Pressure drop
$\varepsilon$	Porosity
$\mu$	Viscosity
$\rho_b$	Bulk Density
$\rho_g$	Gas Density
$\rho_p$	Particle Density

## Abbreviations

BFB	Bubbling Fluidized Bed
CFB	Circulating Fluidized Bed
db	Dry Basis
DFB	Dual Fluidized Bed
GC	Gas Chromatograph
GC/MS	Gas Chromatography-Mass Spectrometry
LPG	Liquid Petroleum Gas
n/a	Not Analysed
ORC	Organic Rankine Process

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## Table of figures

Figure 1.1: Temporal growth of energy input in Austria, [1] .....	1
Figure 1.2: Temporal growth of renewable energy in Austria, [2].....	2
Figure 2.1: The temperature influence on the product gas, [5].....	7
Figure 2.2: a) Updraft and b) downdraft gasifier, [6] .....	8
Figure 2.3: a) bubbling fluidized bed gasifier, b) circulating fluidized bed gasifier and.....	9
Figure 2.4: Entrained bed gasifier .....	10
Figure 2.5: Different states of fluidization, [6] .....	12
Figure 2.6: Force balance for a single particle, [7] .....	13
Figure 2.7: Dependency of the drag coefficient and the Reynolds number, [7].....	14
Figure 2.8: Principal of a dual fluidized bed gasifier.....	16
Figure 3.1: Schematic drawing of the FERCO Biomass Gasification Process, [10] .....	18
Figure 3.2: Drawing of the MILENA gasifier, [11] .....	19
Figure 3.3: Schematic drawing of the eatpipe-Reformer, [14] .....	21
Figure 3.4: Schematic drawing of the new Chalmers research-gasifier, [9] .....	22
Figure 3.5: flowsheet of the power plant in Güssing, [19].....	24
Figure 3.6: Flowsheet of the power plant in Oberwart, [20] .....	26
Figure 4.1: Flowsheet of the power plant at the Vienna University of Technology .....	30
Figure 4.2: Design drawing of the gasifier at the University in Vienna, [18] .....	31
Figure 4.3: Analysis equipment .....	33
Figure 4.4: Photo of the gasifier in Christchurch, New Zealand, [4] .....	34
Figure 4.5: Schematic drawing of the pilot plant in Christchurch, [27] .....	35
Figure 4.6: A view into the riser at the University of Canterbury, Christchurch.....	36
Figure 4.7: Mechanism to refill the hopper.....	36
Figure 5.1: The dual fluidized bed in the regime map according to Bi and Grace [8].....	42
Figure 6.1: Composition of the product gas as measured .....	44
Figure 6.2: Composition of the product gas calculated on zero nitrogen.....	45
Figure 6.3: Composition of the flue gas.....	46

## Table of tables

Table 2.1: Important gasification reactions [2] .....	5
Table 3.1: Product gas composition of the Battelle Columbus [10].....	17
Table 3.2: Typical gas composition of the MILENA gasifier.....	19
Table 3.3: Data of the MILENA gasifier.....	20
Table 3.4: Product gas composition of the heatpipe-reformer .....	20
Table 3.5: Product gas composition of the “New Chalmers” research gasifier, .....	23
Table 3.6: Data of the gasifier in Güssing .....	25
Table 3.7: Data of the power plant in Oberwart. [21].....	26
Table 3.8: Power plant in Europe .....	27
Table 3.9: Expected performance of the planned gasifier in Villach. ....	27
Table 5.1: Typical operating parameters in Vienna.....	38
Table 5.2: Typical operating parameters in Christchurch .....	39
Table 5.3: Comparison of mechanical parameter .....	40
Table 5.4: Comparison of process parameter .....	41
Table 6.1: Comparison of the composition of the product gas produced in Vienna and Christchurch.....	43
Table 6.2: Comparison of the composition of the flue gas produced in Vienna and Christchurch.....	45

## APPENDIX

**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	6.0	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)**

SiO <sub>2</sub>	%	28.46	20.50	71.72
Al <sub>2</sub> O <sub>3</sub>	%	5.43	4.66	0.11
Fe <sub>2</sub> O <sub>3</sub>	%	1.78	2.72	0.28
CaO	%	23.07	24.37	4.37
MgO	%	10.38	8.96	2.76
Na <sub>2</sub> O	%	1.49	1.75	1.04
K <sub>2</sub> O	%	18.11	21.51	11.00
TiO <sub>2</sub>	%	0.20	0.27	0.01
Mn <sub>3</sub> O <sub>4</sub>	%	2.28	1.90	0.25
SO <sub>3</sub>	%	2.60	3.50	1.42
P <sub>2</sub> O <sub>5</sub>	%	3.05	5.09	5.62

The content of the wood pellets at the Vienna University of Technology as excerpt from the report of analysis.

### 3 Analysenergebnisse

Probe-Nr. 09023/1		Bezeichnung: Pellets	
		TS	Roh
Wassergehalt *	[Masse-%]	-	6,11
Aschegehalt 550°C	[Masse-%]	0,29	0,27
<i>Kohlenstoffgehalt</i>	<i>[Masse-%]</i>	<i>50,23</i>	<i>47,16</i>
<i>Wasserstoffgehalt</i>	<i>[Masse-%]</i>	<i>6,04</i>	<i>5,67</i>
<i>Stickstoffgehalt</i>	<i>[Masse-%]</i>	<i>0,05</i>	<i>0,05</i>
Schwefelgehalt	[Masse-%]	0,005	0,005
Chlorgehalt	[Masse-%]	0,003	0,003
Gehalt an flüchtigen Bestandteilen *	[Masse-%]	86,45	81,17
Tiegelkoks *	[Masse-%]	13,55	12,72
Brennwert $H_o$ *	[kJ/kg]	20080	18853
Heizwert $H_u$ *	[kJ/kg]	18753	17458

kursiv dargestellte Werte sind Ergebnisse externer Analysen

\*...Messmethoden im akkreditierten Umfang

Probe-Nr. 09023/2		Bezeichnung: Rinde Pellets	
		TS	Rch
Wassergehalt *	[Masse-%]	-	11,91
Aschegehalt 550°C	[Masse-%]	5,68	5,00
Kohlenstoffgehalt	[Masse-%]	52,00	45,81
Wasserstoffgehalt	[Masse-%]	5,51	4,85
Stickstoffgehalt	[Masse-%]	0,56	0,49
Schwefelgehalt	[Masse-%]	0,045	0,040
Chlorgehalt	[Masse-%]	0,14	0,12
Gehalt an flüchtigen Bestandteilen *	[Masse-%]	72,96	64,27
Tiegelkoks *	[Masse-%]	27,04	23,82
Brennwert H <sub>o</sub> *	[kJ/kg]	20076	17685
Heizwert H <sub>u</sub> *	[kJ/kg]	18866	16329
Ascheschmelzverhalten *			
Erweichungstemperatur A	°C	1150	
Spärischtemperatur B	°C	1190	
Halbkugeltemperatur C	°C	1200	
Fließtemperatur D	°C	1210	

n.a. ... nicht aufgetreten

kursiv dargestellte Werte sind Ergebnisse externer Analysen

\*...Messmethoden im akkreditierten Umfang

Die Analysenergebnisse sind Mittelwerte aus mindestens zwei Einzelbestimmungen.

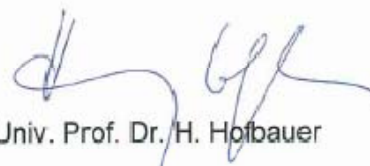
Die Ergebnisse beziehen sich ausschließlich auf die Prüfgegenstände zum Zeitpunkt der Prüfung.

Der Sachbearbeiter



Dipl. Ing. H. Kohl

Der Leiter



Univ. Prof. Dr. H. Hofbauer

The silica sand applied was delivered by Quarzwerke Österreich GmbH (Austria). The raw material is washed and is purged from lime and organic compounds.



## Datenblatt

### Quarzsand ME 0,3 - 0,8

Quarzsand ME 0,3 -0,8 ist ein aufbereiteter natürlicher Rohstoff. Der Quarzsand wird attritiert, gewaschen, hydroklassiert, gesiebt und ist von Kalk und organischen Verunreinigungen befreit. Durch laufende Kontrollen garantieren wir Ihnen eine hohe und gleichmäßige Qualität.

Lieferform : feucht oder trocken (feuergetrocknet)  
lose, abgesackt in PE Säcken, foliert auf Palette

#### Korngrößenverteilung

Körnung in mm	Siebrückstand in %		Siebdurchgang in %
	Richtwert	Toleranz	Richtwert
1,000	1	max.5	99
0,900	1		98
0,710	7	min.90	91
0,500	30		61
0,355	46		15
0,300	11		4
0,250	3	max.5	1
<	1		

#### Chemische Analyse (Gew.-%)

Element	Richtwert	Toleranz
SiO <sub>2</sub>	97,00	96,0-98,0
Al <sub>2</sub> O <sub>3</sub>	1,40	max.2,0
Fe <sub>2</sub> O <sub>3</sub>	0,17	max.0,25

#### Physikalische Eigenschaften

Schüttgewicht	1,5 t/m <sup>3</sup>
Dichte	2,65 t/m <sup>3</sup>
Härte	7 Mohs

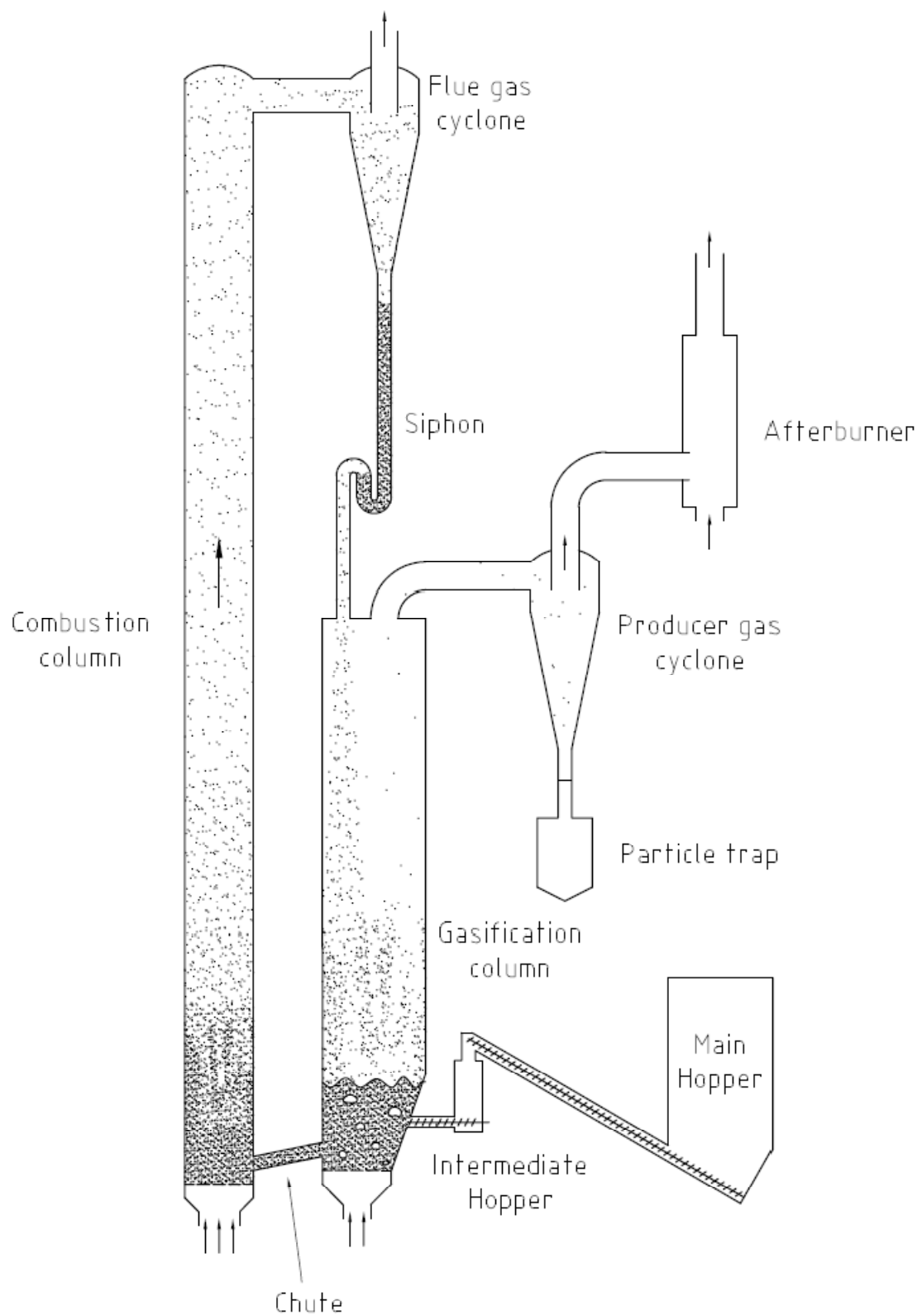
Quarzsand ist ein aufbereiteter natürlicher Rohstoff. Alle Daten sind Richtwerte mit vorkommens- und produktionsbedingter Toleranz. Sie dienen nur zur Beschreibung und stellen keine zugesicherten Eigenschaften dar. Größere und feinere Anteile sind in Spuren möglich. Dem Benutzer obliegt es, die Tauglichkeit für seinen Verwendungszweck zu prüfen. Wir geben auf Wunsch gerne Auskunft über Toleranzbreiten und anwendungstechnische Erfahrungen. Verkäufe erfolgen gemäß unseren Verkaufs- und Lieferbedingungen.

QUARZWERKE Österreich GmbH  
Wachbergstr. 1  
A-3390 Melk  
www.quarzwerke.at

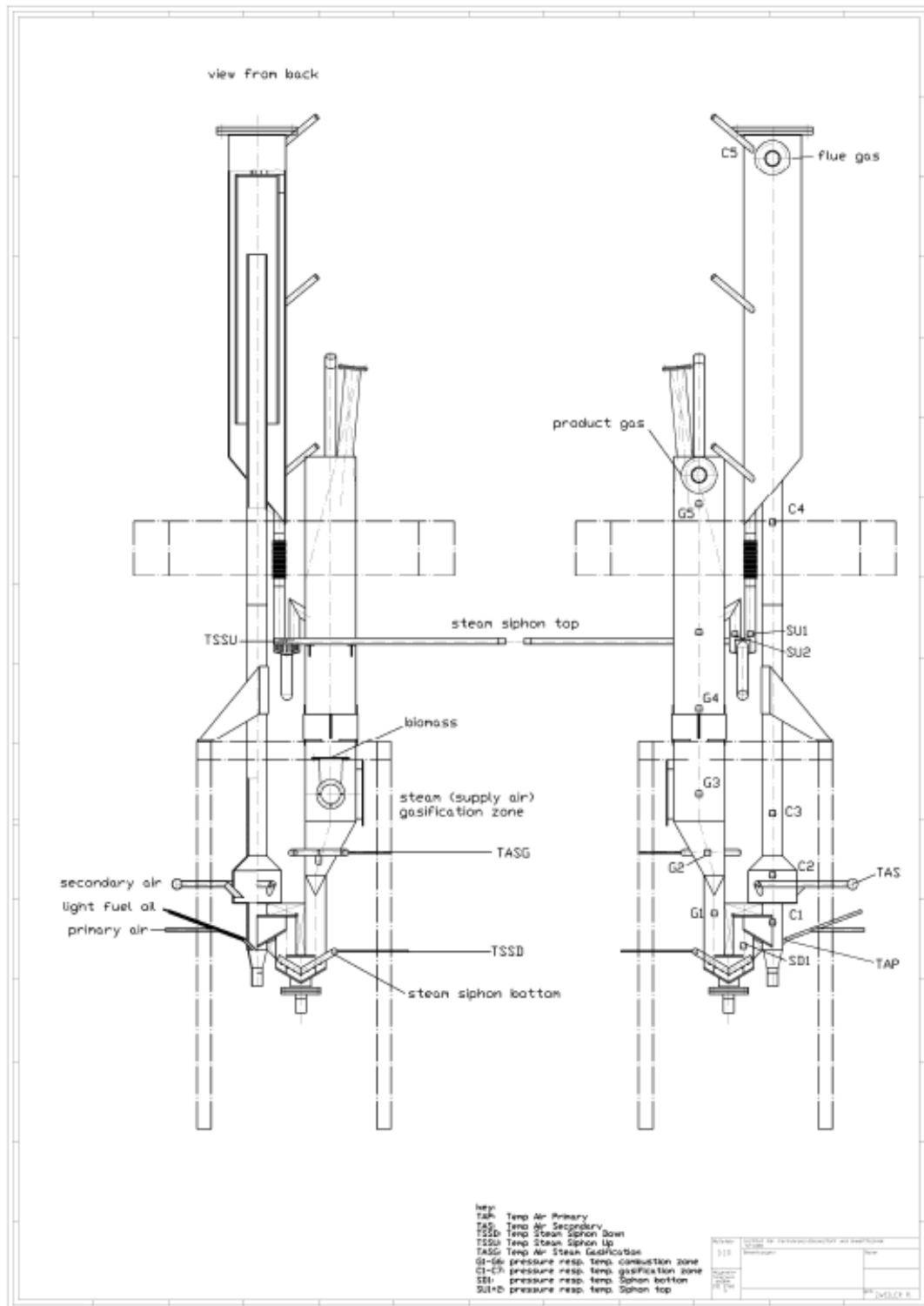
Telefon (02752) 50040-0  
Telefax (02752) 50040-30

The composition of greywacke sand at the University of Canterbury is shown in the table below:

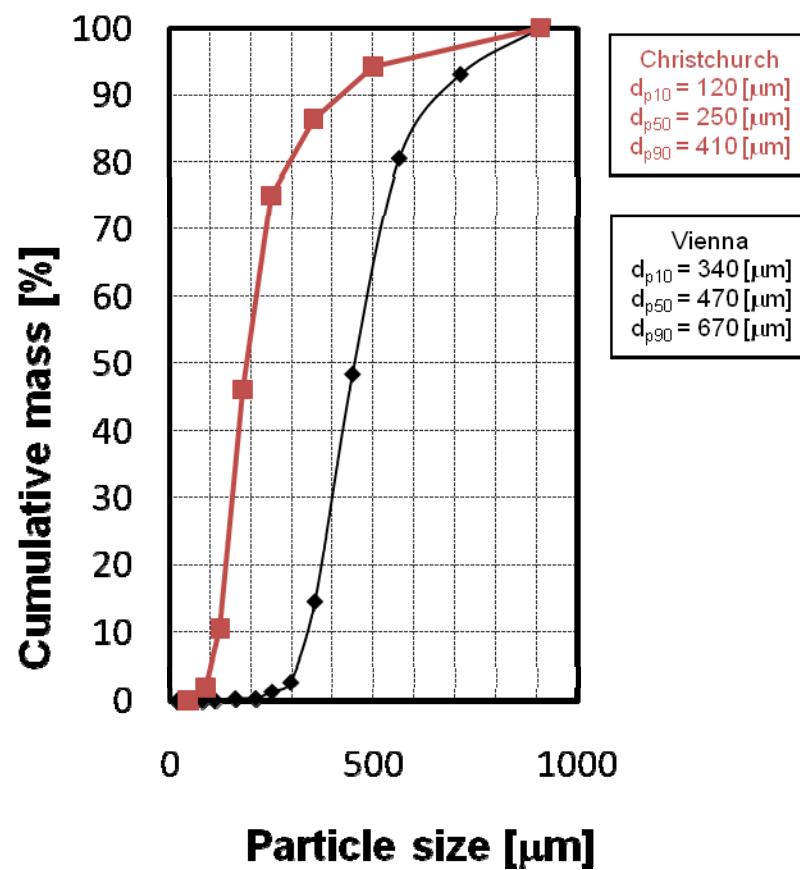
SiO <sub>2</sub>	71.32
Al <sub>2</sub> O <sub>3</sub>	14.42
Na <sub>2</sub> O	4.06
Fe <sub>2</sub> O <sub>3</sub>	3.59
K <sub>2</sub> O	2.51
CaO	1.89
MgO	1.39
P <sub>2</sub> O <sub>5</sub>	0.15
MnO	0.05
TiO <sub>2</sub>	0.01
LOI	1.53



Drawing of the gasifier in Christchurch with the measurement points, [4]



Drawing of the gasifier in Vienna with the measurement points, [4]



Sieve analysis of greywacke sand and silica sand