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**RENEW**

**Renewable fuels for advanced powertrains**

**Integrated Project**

**Sustainable energy systems**

***Deliverable 2.2.1: Report on experimental results and operating conditions regarding the variation of biomass***

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### Task 2.2.1

Optimising of process properties one step gasification fluidised bed reactor

## 1 Introduction

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For the production of biomass to liquid (BTL) fuels there are various technical possibilities. In order to carry out a technical, economic and energetic comparison of various BTL concepts, different technical processes are included in the integrated project "Renew". In principle, BTL fuels can be produced in decentralized plants, centralized plants or in a combination of both systems.

The approach used by CUTECH of a central gasification unit with a circulating fluidized bed (CFB) followed by Fischer-Tropsch synthesis (FTS) has been set up in the technical centre in a size capable of scale-up, with a thermal power of 400 kW. As will be shown below, a thermal power above 400 kW is also possible, depending on the material used.

The aim of the first test phase with the gasification reactor was the setting up of constant operating conditions with the measurement of the gas components important for the following FTS - hydrogen ( $H_2$ ) and carbon monoxide (CO) - depending on the essential operating parameters. Of particular interest is, among other things, the methane concentration, which, in the gasification with the aim of maximum hydrogen yield, is an unwanted accompanying component, because the hydrogen bonded in this form is not available for FTS without further treatment (reforming). In addition, for comparison, evaluation and plausibility testing, mass and energy balance statements were drawn up for the different operating conditions.

The following sections are divided into the description of the plant technology and the measurement techniques used. This also includes a stationary laboratory fluidized bed reactor, in which initial pre-investigations into fuel characteristics (e.g. ash melting behaviour, and release of volatile products) were carried out.

In the section 'Test procedure' the parameter variations are described and the necessity for cold tests in the plant is discussed.

Further to this, the mass and energy balance statements for the gasifier are drawn up. In section 5 the results of the cold and hot tests carried out so far are presented.

The report ends with a summary and a description of future perspectives.

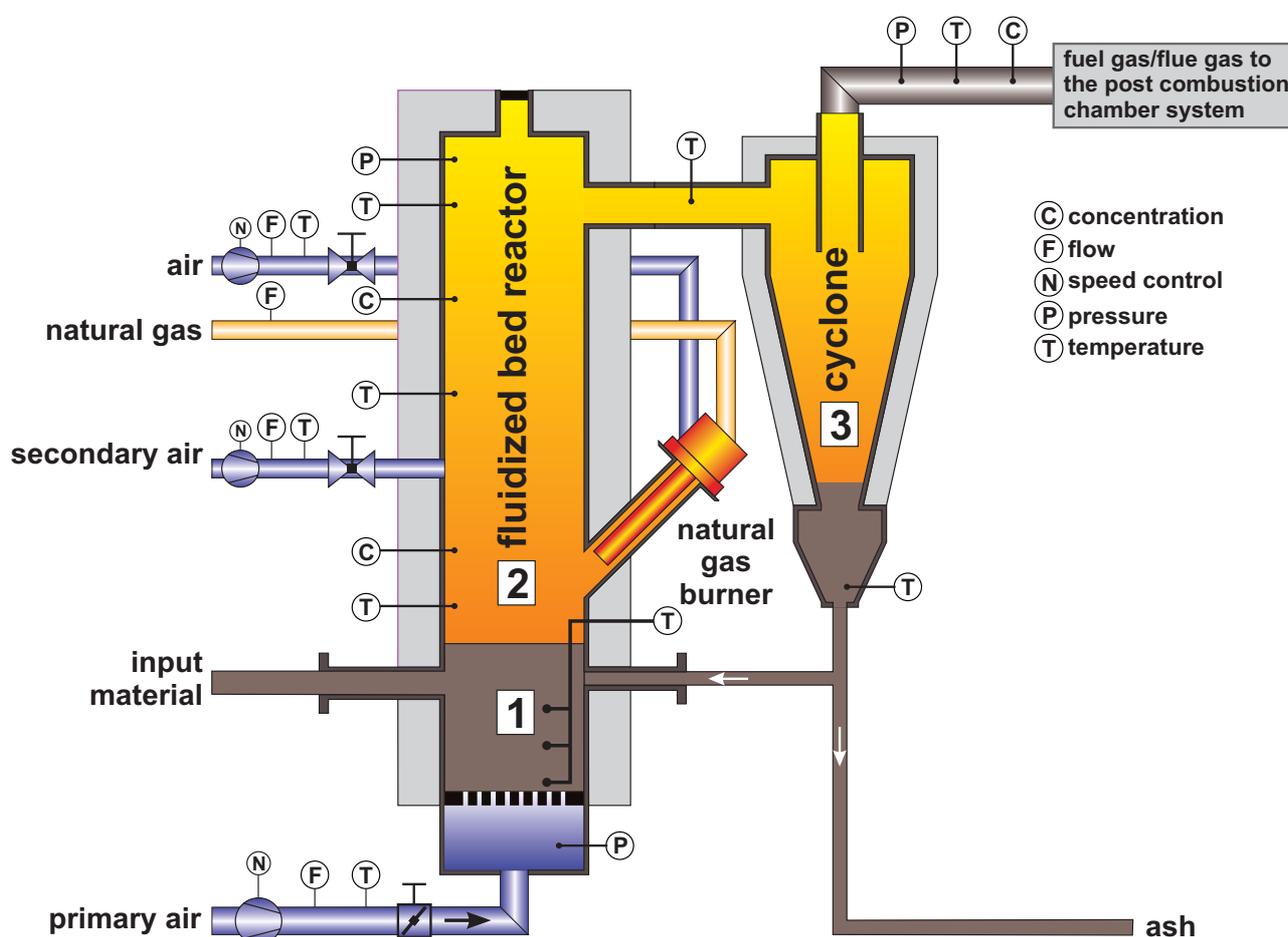
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## 2 Plant technology

In this section the test plants available at CUTEC and relevant for the project are described.

### 2.1 Laboratory fluidized bed reactor

Since 1998 CUTEC has had at its disposal a laboratory fluidized bed reactor which can be used in stationary or circulating modes (thermal power approx. 50 kW), in which various investigations, especially with biomass (Olive kernels, peach stones, straw, sugar beet shreds) for gasification and combustion have been carried out. **figure 2-1** shows a process-technology scheme of this plant:

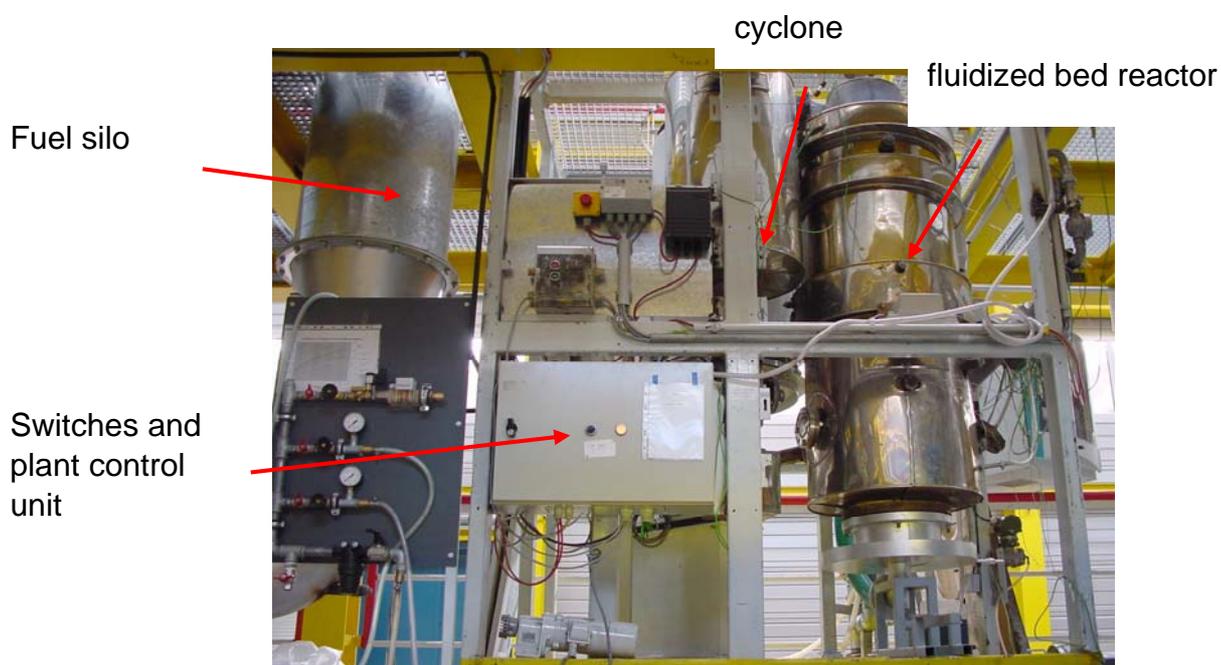


**figure 2-1:** Schematic diagram of the laboratory fluidized bed reactor

The fluidized bed reactor consists of the following main components:

- ◆ Fluidized bed (1 in **figure 2-1**),

- ◆ Free-board (2 in **figure 2-1**) and
- ◆ Hot gas cyclone (3 in **figure 2-1**)



**figure 2-2:** Side view of the test plant laboratory fluidized bed reactor

A revolution-controlled conveyor worm transports the input material, broken down to a maximum grain size of 5 mm, from the charging bunker to the sieve of the fluidized bed reactor. A maximum mass flow of about 20 kg/h (depending on the input material) can be achieved. The maximum fluidized bed height in steady-state mode amounts to about 300mm.

The heating of the fluidized bed, and a controlled supplementary firing when using low calorific input material, is achieved with a controllable natural gas burner situated at about the same height as the input supply pipe connection. Depending on the chosen parameters, load ratio and primary air ratio, secondary air can be supplied via four radially connected inlets above the fluidized bed in order to optimise burnout. The gas generated in the fluidized bed is fed to a hot gas cyclone where the dust and ash are removed. A cellular wheel sluice transports the dust and ash to a bunker and/or with the help of a conveyor screw optionally back to the fluidized bed (steady state or circulation operations mode).

When operating the fluidized bed reactor substoichiometrically, the combustion gas is fed through the cyclone to the post-combustion chamber where an independent and staged combustion takes place.

In the case of hyperstoichiometrical operation of the fluidized bed, the flue gas is fed through the combustion chamber in order to improve burnout (long throughput times at high temperatures).

Finally, the flue gas is released into the atmosphere via the flue gas purification plant.

The measuring and control scheme presented in the processing flow chart in **figure 2-1** gives an overview of the parameters measured during operation. Here, those values significant for the operation and balancing of the plant are collected and saved for further processing in a computer system.

### **Variation of operating parameters**

The operating conditions in the pilot plant can be varied via the following parameters:

- ◆ composition of input material,
- ◆ throughput time,
- ◆ air ratio,
- ◆ oxygen enrichment of the reaction gas,
- ◆ simulation of flue gas recycling by supply with nitrogen,
- ◆ variation of the load ratio,
- ◆ variation of the air distribution,
- ◆ ash feedback (optional).

The use of this fluidized bed as a CFB gasifier in the "Renew" project is restricted for the following reasons:

- ◆ A scale-up of the results to the MW level of a central gasification unit, by a factor of 100, is not possible from a technical perspective, even though the CFB technology in general represents the current state of the art.
  - ◆ The turnaround time of the solid material and the gas is insufficient to allow the sought after main reactions to run almost to completion; thus there would be a low yield, and a lot of tar and by-products would be produced.
  - ◆ A steam / oxygen gasification is not possible at this time; enriched air can only be gasified with air or with oxygen.
  - ◆ The plant has a high heat loss (partly > 50 %).
  - ◆ It has old sensor technology.
-

- ◆ The flue gas purification is specified for combustion products using a grate firing with a thermal power of approx. 1 MW. It is not suitable for partial capacity operation with 50 kW power and is also not set up for the cleaning of a synthesis gas.

Due to these technical limitations CUTEC used a considerable amount of its own resources in order to build a plant to produce synthesis gases from various biomasses. The completion of this plant was delayed because Basic Engineering had first planned to build an allothermal stationary gasifier.

In connection with the heat transfer into the biomass, the required heat can be added either indirectly via a heat carrier, as used in the plants in Güssing [3] or FZK [4] or alternatively using heat pipes or radiator tubes [5]. These possibilities are already in use at the above named research institutions, or are protected by patents, so that at CUTEC the tried and tested technology of a direct heat transfer through part combustion of the biomass direct in the CFB was seen as the aim.

The gasifier was set up for a thermal power of 400 kW. This is a power level which allows a scale-up to industrial dimensions and, as previously, allows technical flexibility in the plant's modification with acceptable technical and financial costs.

This Technology Centre plant is described in the following section.

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## 2.2 Circulating fluidized bed on a pilot scale

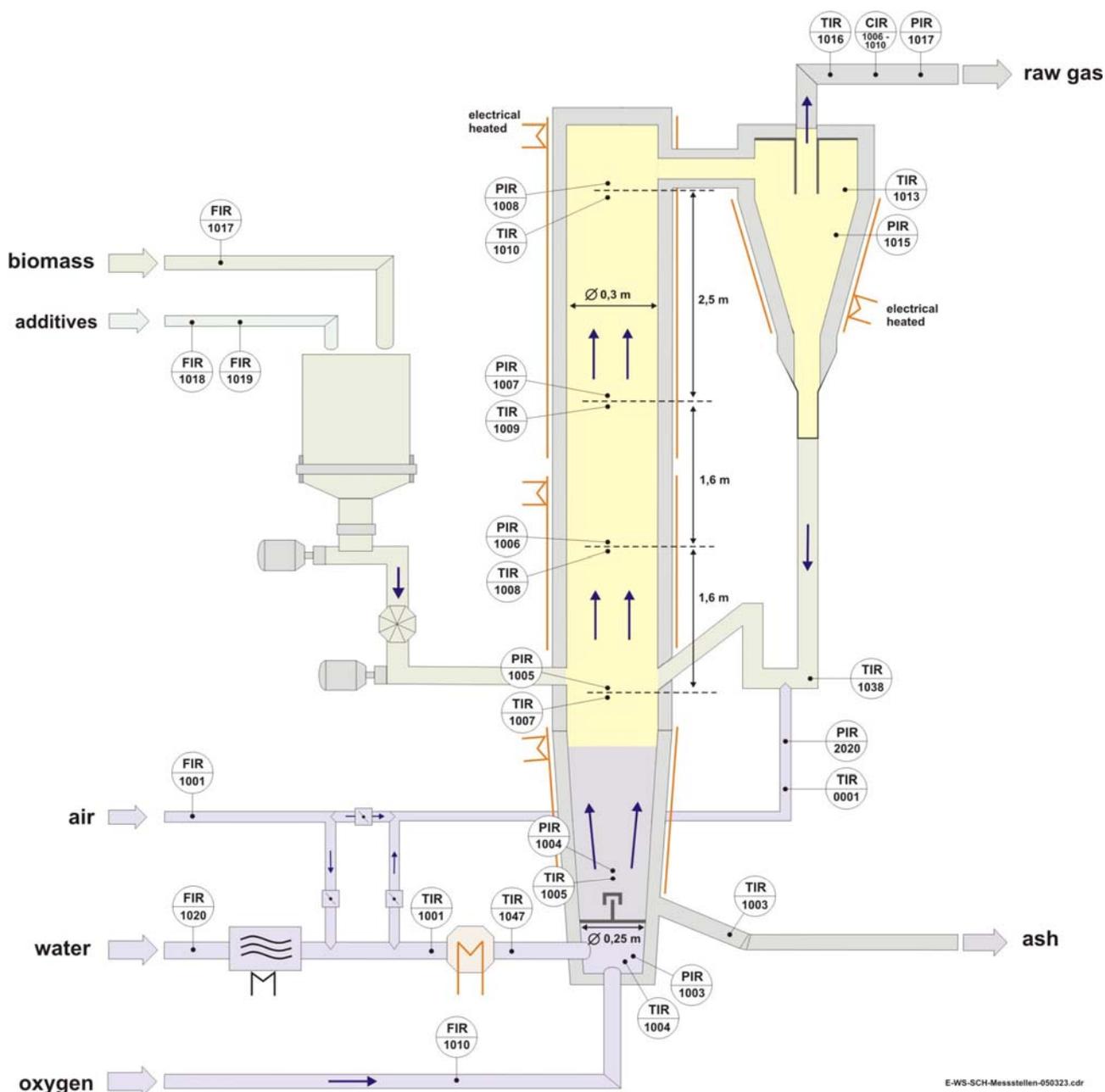
The applied gasification technology at CUTEC can be classified and explained as follows:

Reactor type:	CFBR ( <b>C</b> irculating <b>F</b> luidized <b>B</b> ed <b>R</b> eactor)
Bed material:	Sand, catalytic active material
Feed:	Biomass (wood, straw, miscanthus...)
Gasification agent:	Steam/oxygen or air
Heat transfer:	autothermic, via partial oxidation
Pressure:	atmospheric
Temperature:	950°C max.
Product gas application:	Fischer-Tropsch synthesis
Scale:	Pilot (400kW)

A typical attribute of any gasification process is its endothermic behaviour, which means that in all cases a gasification reactor has to be supplied with heat. At CUTEC this is done by internal partial oxidation, also known as the *autothermic* operation mode. As a gasification agent either air, oxygen enriched air or a mixture of steam and oxygen can be used.

The process scheme of the synthesis gas production is shown in **figure 2-3**.

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**figure 2-3:** Synthesis gas production

The autothermic operation-mode gives the advantage that the heat can be directly provided in the fluidized bed, without internals or heat transfer surfaces vulnerable to fouling. The compromise which has to be tolerated is the loss in LCV (**L**ower **C**alorific **V**alue) caused by inert gas components, produced as a result of the partial oxidation. To reduce the loss in LCV, at CUTEC the gasification agent can be preheated to 500°C before it enters the reactor. Thereby the requirement of gasification heat can be partially met and, in addition, oxygen can be saved.

A 100kW steam generator, designed as a package unit supplies the CFBR with low-pressure steam. Oxygen is stored outside the pilot plant station in a 6m<sup>3</sup> cryogenic tank with adapted evaporator unit. For pressurizing the air feed a roots blower is used. The biomass is stored in an 8m<sup>3</sup> bunker and fed by screw conveyors and cell locks into the

gasifier. Subsequently, the feed together with the gasification agent is physically and chemically converted into a hydrogen and carbon monoxide rich synthesis gas. The operation of a CFBR has several advantages:

- reliable technology, with no moving parts
- safeness, availability and stability
- wide range of particle sizes can be used
- low tar concentration in the product gas
- constant gas composition because of homogenous fluidized bed

To compensate for heat losses and to speed up the starting procedure, the reactor hull is equipped with an electrical heating system. For safety reasons, the biomass feed is started at a minimum temperature of 700°C.

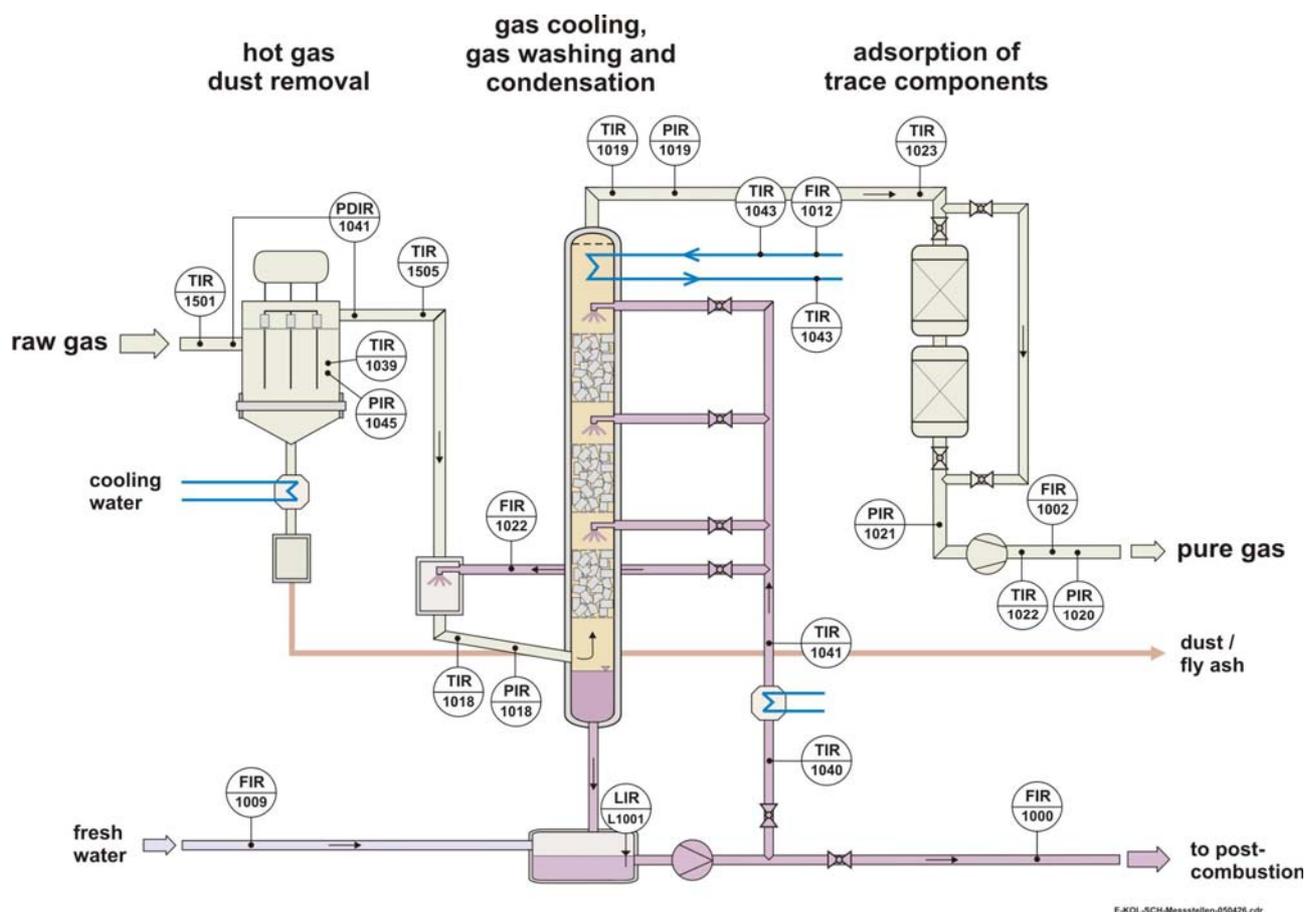
The gasification agent steam - in addition to its roll as a reaction partner - supports the fluidising of the reactor content and the circulation of the bed materials. After the exit of the two-phase flow of synthesis gas / bed material out of the freeboard of the fluidized bed, the separation of the solid material in the cyclone takes place. The bed material is fed back to the reactor via a siphon. The raw synthesis gas, pre-cleaned of dust, leaves the cyclone via a dip tube.

A part of the raw synthesis gas (up to 1/3) is then fed to the gas purification unit (process stage 2). The larger part of the raw synthesis gas receives no further processing, it is burnt in a swirled combustion chamber. The waste gases are then cooled using an exhaust gas cooler to cloth filter temperature. After passing through the flow reactor and cloth filter, the purified combustion gas passes via the extractor through an active carbon filter and is then released into the atmosphere. The swirled combustion chamber, waste gas purification stages and extractor are parts of the plant which have been operated successfully for many years by CUTEC and are available in our Technical Centre.

### Synthesis gas purification

The concept of the gas purification consists of 3 process stages. **Figure 2-4** shows the process scheme for this.

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**Figure 2-4:** Synthesis gas purification

Firstly, the gas is fed at reaction temperature into a hot gas filter equipped with a ceramic filter, where it is cleaned of dust. As an area of research for the future, it is planned to develop catalytically active filter candles in cooperation with partners. In the following gas scrubber column, which in the simplest case is operated with water, the temperature reduction leads to condensation of the steam component and simultaneously to purification of the gases through absorption of unwanted components such as hydrogen chloride (HCl) and ammonia (NH<sub>3</sub>).

Before the transfer of the gas to the Fischer-Tropsch synthesis it passes through an activated char coal filter. Here, any sulphur and tar compounds which may still be present are bound absorptively. An extractor serves to create the pressure gradient necessary for the gas transport over the three purification stages.

## 2.3 Measurement technology

In the following, only the measurement technology for the synthesis gas production (CFB gasifier) will be discussed in more detail at first; the synthesis gas purification will be described in detail in a future report. After that, the sample taking system for collecting the condensate and measuring special gas components will be described.

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### 2.3.1 Measurement technology - pilot plant CFBR

The measurement technology includes the collection and processing of data for the characterisation of the material streams and the drawing up of mass and energy balances. The measurement points are briefly described in the following and are shown in **figure 2-3**.

In order to restrict the number of measurement points in the reactor, combined measuring points are used for pressure and temperature measurement, in which the thermocouple element is contained in the inner tube of a double tube and the static pressure is measured via an annular gap. At regular time intervals the annular gap is flushed with nitrogen to avoid steam condensation and clogging with bed material. In addition, there is the possibility of removing gas samples via the measurement point.

In order to measure the temperature and pressure development in the reactor and cyclone over the height of the reactor, the following measuring points are installed:

- TIR 1004/ PIR 1003: Temperature (gasification agent mix temperature)/ pressure underneath the nozzle floor
- TIR 1005/ PIR 1004: Temperature / pressure above the nozzle floor
- TIR 1007/ PIR 1005: Temperature / pressure
- TIR 1008/ PIR 1006: Temperature / pressure above the material input point
- TIR 1009/ PIR 1007: Temperature / pressure in the middle of the reactor
- TIR 1010/ PIR 1008: Temperature / pressure at the top of the reactor
- TIR 1013/ PIR 1015: Temperature / pressure in the cyclone
- TIR 1038: Temperature in the siphon
- TIR 1003: Temperature of the ash
- TIR 1016/ PIR 1017: Temperature / pressure at the cyclone exit

In order to draw up balances, knowledge of the volume or the mass flow rate of at least the input material is necessary. This data is collected via the following measuring points:

- FIR 1017: mass flow rate of the input material (screw characteristic curve)
  - FIR 1018/ 1019: mass flow rate of the additive (screw characteristic curve)
  - FIR 1001: mass flow rate of the air (hot-wire anemometer)
  - FIR 1020: mass flow rate of the steam (circulation frequency measurement)
  - FIR 1010: mass flow rate of the oxygen (hot-wire anemometer)
-

In addition to this the temperatures before and after the super heater are measured:

- TIR 1001: Temperature before the super heater
- TIR 1047: Temperature after the super heater
- CIR 1006-1010: Further to this, in order to measure the composition of the raw gas at the cyclone exit a heated extraction probe is used. Here the standard permanent gas components  $H_2$ ,  $CH_4$ ,  $O_2$ ,  $CO$ ,  $CO_2$  and the humidity are measured online.

In order to determine the mass and enthalpy flow rates various measurement processes are used for the gaseous media. For measurement of the volume flow rates, hot-wire anemometers are used for air and oxygen, for steam a vortex meter is utilized. The temperature measurement is carried out using appropriate thermo-elements. Using the operating density and the operating volume flow rate the individual mass flow rates can then be calculated.

The mass flow rates of the solid materials (biomass and additive) are determined using screw characteristic curves. The bed ash removed is weighed gravimetrically using scales and determined over the period of the investigation.

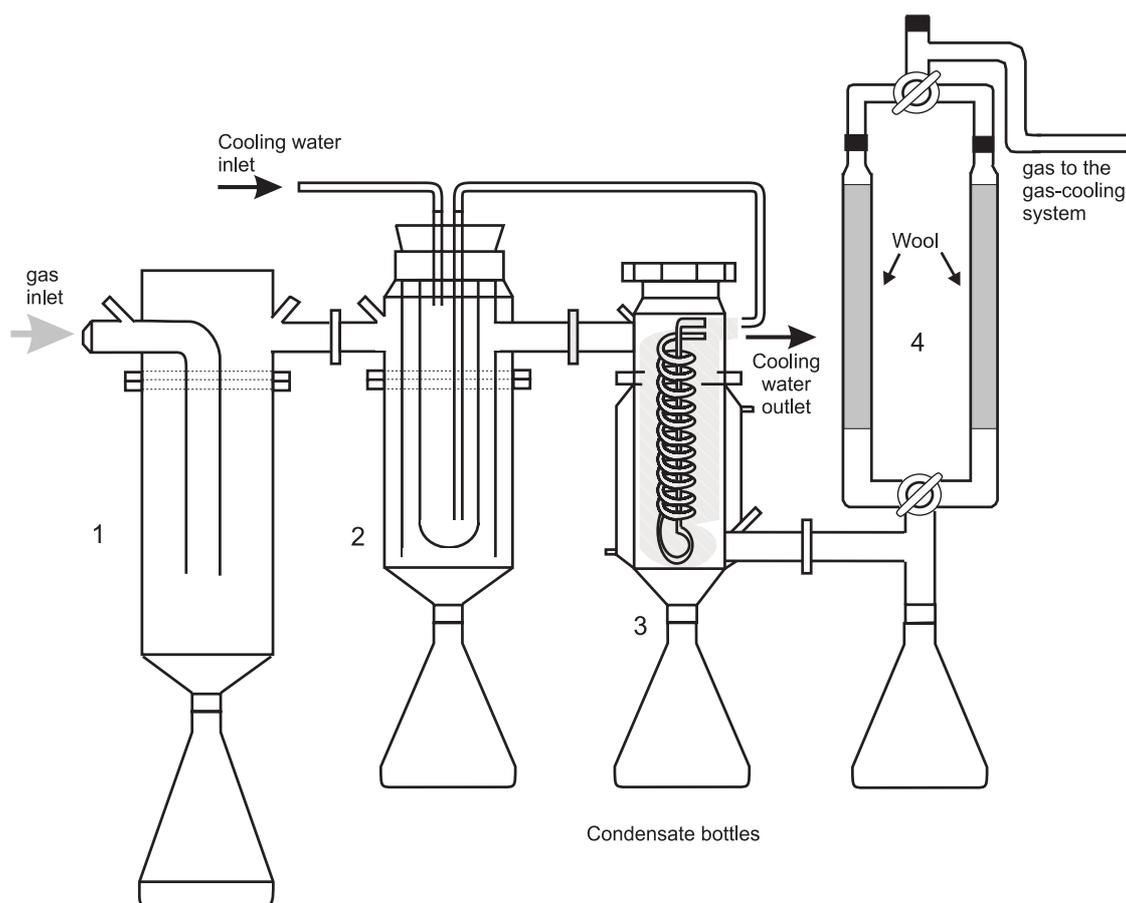
The circulation rate of the bed material can be calculated using the pressure difference  $\Delta PIR$  (1007-1008), (see. section on mass balance).

### **2.3.2 Sample taking system for condensable gas components**

The sampling point for the intermittent sampling of the composition of the raw synthesis gas in relation to components condensable at room temperature is found at the measuring point CIR 1006 - 1010.

**Figure 2-5** shows the sampling system consisting of the four main components:

- ◆ pre-separator (1),
  - ◆ finger-type cooler (2),
  - ◆ spiral cooler (3),
  - ◆ measuring gas filter (4).
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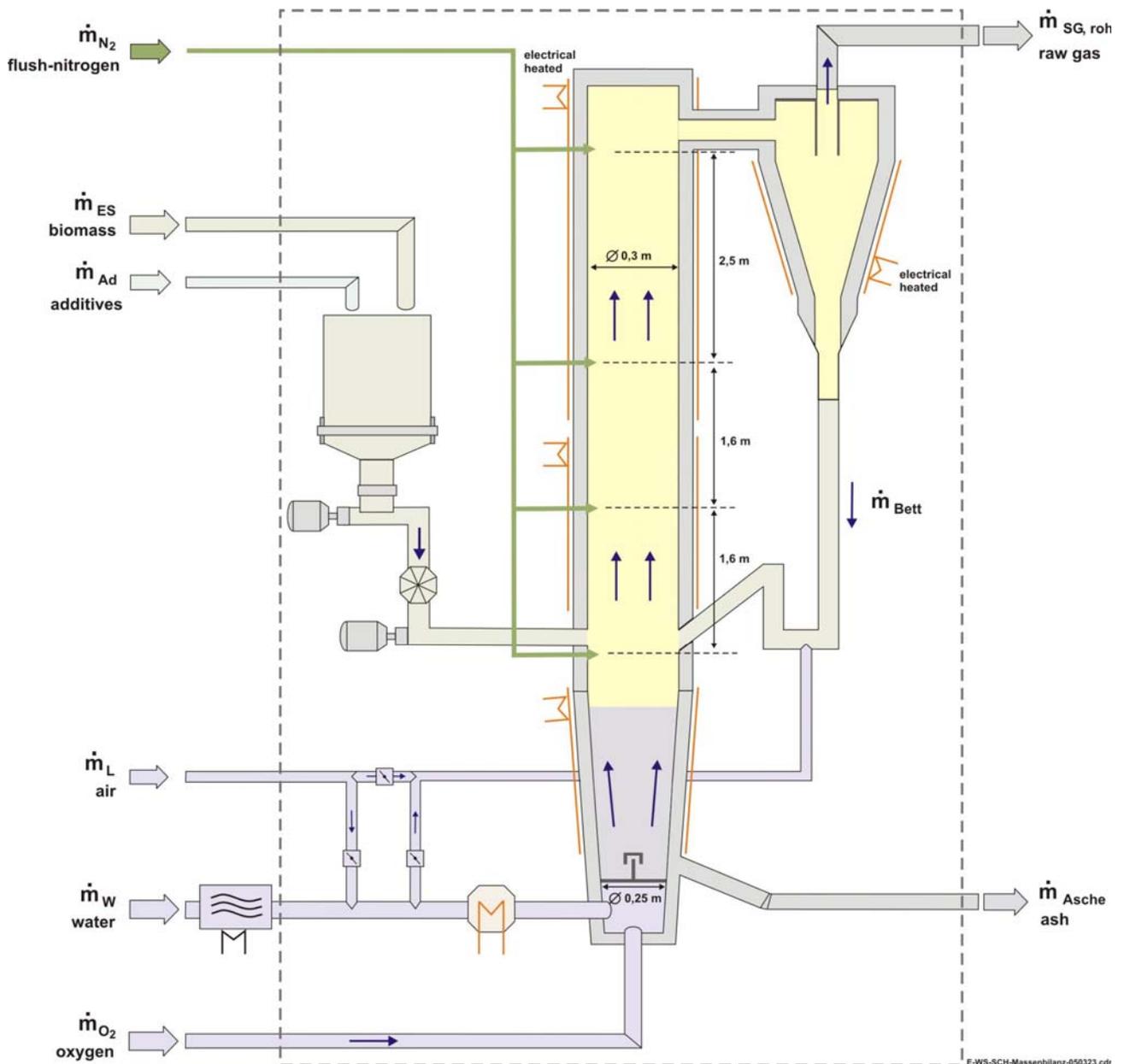


**Figure 2-5:** Intermittent gas sampling system

The measuring gas cooling system comprises three main components, namely, a pre-separator (1), a so-called finger-type cooler (2) and behind it a spiral cooler (3). The extracted hot pyrolytic gas flows through a pre-separator, thus separating the fly ash and char to prohibit slugging of the smaller pipes, then flows through the finger-type cooler, in a counter-current arrangement with the cooling water, causing rapid cooling of the gas. For the heavily volatile components in the pyrolytic gas the cooler's smooth surface acts as a good and quick run-off into the condensate collecting bottle underneath. In the case of a significant content of heavily volatile components in the gas (e.g. 30 to 40 vol.-%), even relatively small drops in temperature cause these fractions to precipitate. The pre-cooled pyrolytic gas then reaches a spiral cooler (3), also operated in a counter-current arrangement, the large surface of which causes further cooling. Via two parallel connected filters (4) filled with quartz wool the pyrolytic gas current flows to a second gas cleaning system with the possibility to take gas samples. The amount of gas volume can be measured by a gas flow meter after the second gas cleaning system.

### 3 Mass, material and energy balance

The drawing up of balances serves, firstly, to check the plausibility of the measured values (e.g. to check measured concentrations using values from the balance of materials) and secondly to determine values which are difficult or impossible to ascertain, e.g. the raw synthesis gas mass flow rate.



**figure 3-1:** Schematic depiction of the mass flows of the complete plant including the fluidized bed and the cyclone

From the mass balance around the fluidized bed, including the cyclone, the mass flow rate of the synthesis gas of the fluidized bed according to **figure 3-1** is determined using the following formula:

$$\dot{m}_{SG,roh} = \dot{m}_{ES,WS} + \dot{m}_{Ad,WS} + \dot{m}_{Luft,WS} + \dot{m}_{Wasser,WS} + \dot{m}_{O_2,WS} + \dot{m}_{N_2,WS} - \dot{m}_{Asche,WS} \quad (3-1)$$

From the ash component of the input material  $\xi_{Asche,ES}$  and the input material mass flow rate the ash mass flow rate can be calculated using formula 3-2, this is necessary because the quantities remaining in the cyclone are not regularly emptied and the quantities of ash leaving the reactor with the raw gas can only be measured using the part flow over the hot gas filter.

$$\dot{m}_{Asche,WS} = \xi_{Asche,ES} \cdot \dot{m}_{ES,WS} \quad (3-2)$$

Because the plant is operated with a slight over-pressure, the possibility of air entering through leakages can be eliminated.

The synthesis gas mass flow is essentially composed of four component flow rates:

$$\dot{m}_{SG,roh} = \dot{m}_{PG,SG,WS,tr.} + \dot{m}_{Dampf,SG,WS} + \dot{m}_{Teer,SG,WS} + \dot{m}_{Flugkoks,SG,WS} \quad (3-3)$$

- ◆ the dry permanent gas component  $\dot{m}_{PG,SG,WS,tr.}$ ,
- ◆ a steam component  $\dot{m}_{Dampf,SG,WS}$ ,
- ◆ a tar component  $\dot{m}_{Teer,SG,WS}$  and
- ◆ a fly ash component  $\dot{m}_{Flugkoks,SG,WS}$ .

These components are, similarly to the tar and fly ash composition, first unknown and must be estimated; the estimates are modified so that the error in the material and energy balance is minimal. In further investigation programmes these estimates will be checked by measurements.

### Balance of materials

The balance of materials is next drawn up for the essential elements C, H, and O, whereby the moisture from the fuel and the synthesis gas is converted to H and O. Using targeting, the raw gas moisture, for example, which can be composed of parts of the fluidisation agents steam, water from the decomposition of the organic matter and fuel moisture, can be calculated.

As an example the balance of materials for C is drawn up as follows:

$$\Psi_{C,ES} \cdot \dot{m}_{ES,WS} = \Psi_{C,Asche} \cdot \dot{m}_{Asche,WS} + \Psi_{C,SG} \cdot \dot{m}_{SG,WS} + \Psi_{C,Flugkoks} \cdot \dot{m}_{Flugkoks,SG} \quad (3-4)$$

$$+ \Psi_{C,Teer} \cdot \dot{m}_{Teer,SG}$$

$\Psi_{C,ES}$ ,  $\Psi_{C,Asche}$ ,  $\Psi_{C,Flugkoks}$ ,  $\Psi_{C,Teer}$  are proportions by mass of C from individual fractions from analyses or estimates.

Determination of  $\Psi_{C,SG}$  from the gas composition:

$$\Psi_{C,SG} = \frac{12}{16} \cdot \Psi_{CH_4} + \frac{12}{28} \cdot \Psi_{CO} + \frac{12}{44} \cdot \Psi_{CO_2} + 2 \cdot \frac{12}{30} \cdot \Psi_{C_2H_6} \quad (3-5)$$

$$+ 2 \cdot \frac{12}{28} \cdot \Psi_{C_2H_4} + 3 \cdot \frac{12}{44} \cdot \Psi_{C_3H_8} + 3 \cdot \frac{12}{42} \cdot \Psi_{C_3H_6}$$

$$+ 4 \cdot \frac{12}{58} \cdot \Psi_{C_4H_{10}} + 5 \cdot \frac{12}{72} \cdot \Psi_{C_5H_{12}} + 6 \cdot \frac{12}{86} \cdot \Psi_{C_6H_{14}}$$

- Trace components, such as HCN, have not been considered here.
- The proportion of C in the fly ash is considered as 1.
- Because currently only data from the continuous gas sampling is available, the nitrogen content of the gas is calculated using 1 minus the sum of the measured components.
- The measurement of further permanent gas components and the analysis of the tar are being prepared.

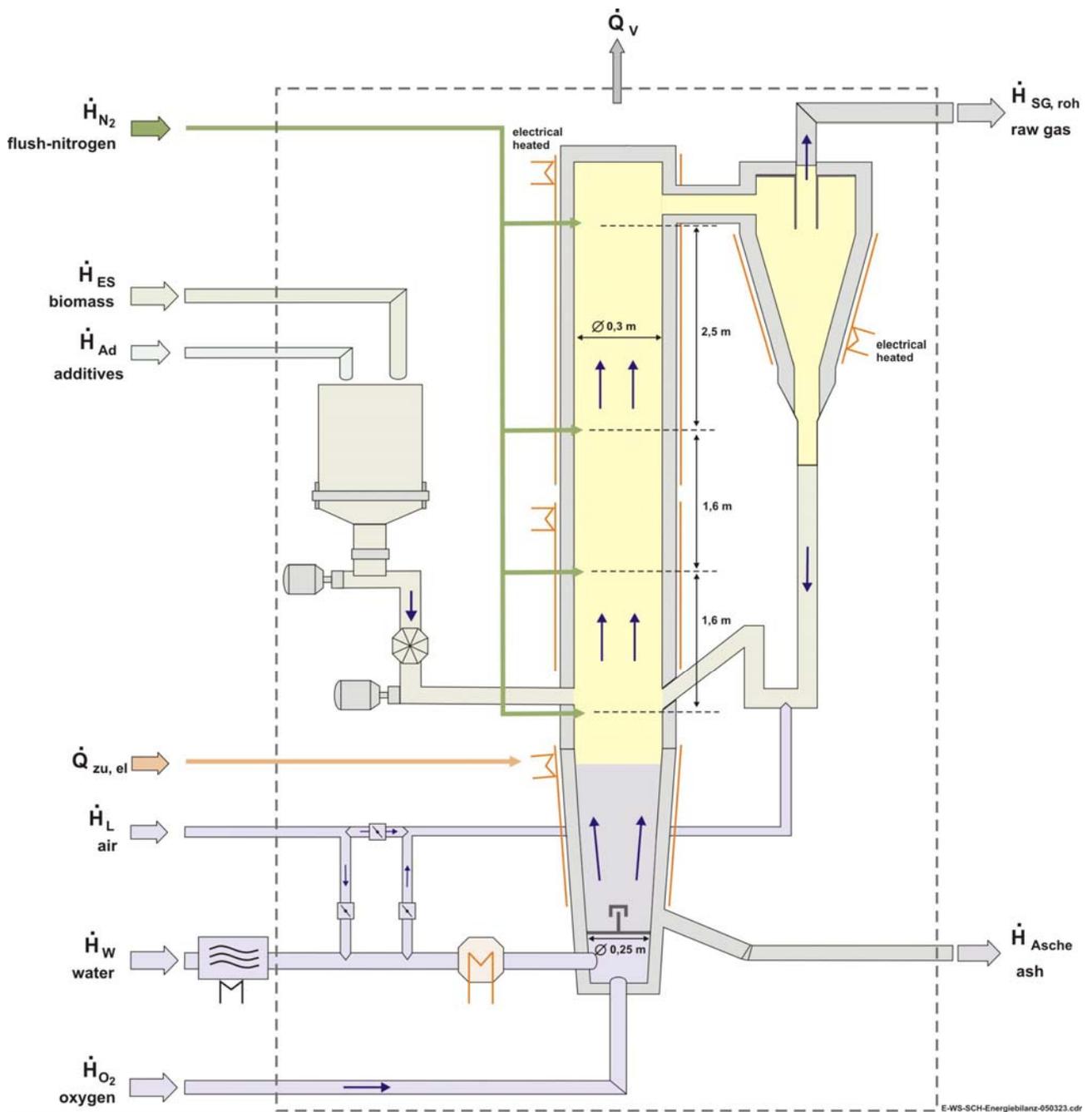
For the control of the moisture measurement in the raw synthesis gas, the moisture of the waste gas with the relevant oxygen content can be measured after the combustion chamber system. Using a combustion calculation, which takes into account the moisture from the combustion chamber head operated natural gas burner, the moisture content of the raw gas can be determined. A detailed discussion of the individual formulas of the combustion calculation will not be entered into here.

Because the moisture could not be determined in the first investigation, it was calculated using the O and H balances. The total water quantity put into the process through the fuel moisture and steam can be smaller than the quantity of steam exiting in the synthesis gas. On the one hand, water is used by the water vapour shift reaction, on the other, a part process of the pyrolysis produces decomposition water.

### Energy balance

Together with the balance of materials, the energy balance serves to determine the unknown enthalpy flow rate of the fly ash, the tar content and the steam quantity.

The aim is to minimize the error between input and output levels.



**Figure 3-2:** Energy balance in and around the fluidized bed

The equation for the energy balance in the stationary state is:

$$0 = \dot{H}_{ES,WS} + \dot{H}_{Ad,WS} + \dot{H}_{N_2,WS} + \dot{H}_{O_2,WS} + \dot{H}_{Luft,WS} + \dot{H}_{Wasser,WS} + \dot{Q}_{zu,el.} - \dot{H}_{Asche,WS} - \dot{H}_{SG,WS} - \dot{Q}_V \quad (3-6)$$

Enthalpy flow, input material:

$$\dot{H}_{ES,WS} = \dot{m}_{ES,WS} \cdot (c_{p_{ES}} \cdot (\vartheta_{ES} - \vartheta_0) + h_{u_{ES}}) [kW] \quad (3-7)$$

Enthalpy flow, flushing nitrogen:

$$\dot{H}_{N_2,WS} = \dot{m}_{N_2,WS} \cdot c_{p_{N_2}} \cdot (\vartheta_{ES} - \vartheta_0) [kW] \quad (3-8)$$

Enthalpy flow, oxygen:

$$\dot{H}_{O_2,WS} = \dot{m}_{O_2,WS} \cdot c_{p_{O_2}} \cdot (\vartheta_{O_2} - \vartheta_0) [kW] \quad (3-9)$$

Enthalpy flow, air:

$$\dot{H}_{Luft,WS} = \dot{m}_{Luft,WS} \cdot c_{p_{Luft}} \cdot (\vartheta_{Luft} - \vartheta_0) [kW] \quad (3-10)$$

Enthalpy flow, steam:

$$\dot{H}_{Dampf,WS} = \dot{m}_{Wasser,WS} \cdot (c_{p_w} \cdot (100 - \vartheta_0) + \Delta h_V + c_{p_w} \cdot (\vartheta_{Dampf} - 100)) [kW] \quad (3-11)$$

Enthalpy flow, ash:

$$\dot{H}_{Asche,WS} = \dot{m}_{Asche,WS} \cdot (c_{p_{Asche}} \cdot (\vartheta_{Asche} - \vartheta_0)) [kW] \quad (3-12)$$

Enthalpy flow, raw synthesis gas:

$$\dot{H}_{SG,WS,roh} = \dot{H}_{PG,SG,WS,tr.} + \dot{H}_{Dampf,SG,WS} + \dot{H}_{Teer,SG,WS} + \dot{H}_{Flugkoks,SG,WS} [kW] \quad (3-13)$$

$$\dot{H}_{PG,SG,WS,tr.} = \dot{m}_{PG,SG,WS,tr.} \cdot (c_{p_{PG}} \cdot (\vartheta_{SG,WS} - \vartheta_0) + h_{u_{PG,SG,tr.}}) [kW] \quad (3-14)$$

$$\dot{H}_{Dampf,SG,WS} = \dot{m}_{Dampf,SG,WS} \cdot (c_{p_w} \cdot (100 - \vartheta_0) + \Delta h_V + c_{p_{Dampf,SG}} \cdot (\vartheta_{SG,WS} - \vartheta_0)) [kW] \quad (3-15)$$

$$\dot{H}_{Teer,SG,WS} = \dot{m}_{Teer,SG,WS} \cdot (c_{p_{Teer,SG}} \cdot (\vartheta_{SG,WS} - \vartheta_0) + h_{u_{Teer,SG}}) [kW] \quad (3-16)$$

$$\dot{H}_{Flugkoks,SG,WS} = \dot{m}_{Flugkoks,SG,WS} \cdot (c_{p_{Flugkoks,SG}} \cdot (\vartheta_{SG,WS} - \vartheta_0) + h_{u_{Flugkoks,SG}}) [kW] \quad (3-17)$$

The heating values of the input material and the dry permanent gas component were determined experimentally. The fuel ash has no more heating value. Residual carbon content is taken into account for the calculations here as fly ash with 100 % carbon. The cyclone ash and the filter ash from the hot gas filter could easily have carbon contents of up to 80 Ma.-% and are therefore comparable with pyrolysis cokes with appropriate heating values.

As already mentioned in section 2.2, to estimate the heat loss of the gasifier and to collect possible leakage material, the plant was surrounded with a sheet metal jacket with a gap of a few cm between it and the isolated reactor, so that the heat loss  $\dot{Q}_{V,1}$  could be

calculated using a sucking fan intake with a defined volume flow and temperature measurement at the inlet and outlet. Furthermore, the heat loss via the screw cooling water  $\dot{Q}_{V,2}$  is taken into account.

$$\dot{Q}_V = \dot{Q}_{V,1} + \dot{Q}_{V,2} + \dot{Q}_{V,3} \quad [kW] \quad (3-18)$$

$$\dot{Q}_{V,1} = \Delta \dot{H}_{A-E} = \dot{m}_{Kuehlung,WS} \cdot c_{p\ Luft} \cdot (\vartheta_{Luft,A} - \vartheta_{Luft,E}) \quad [kW] \quad (3-19)$$

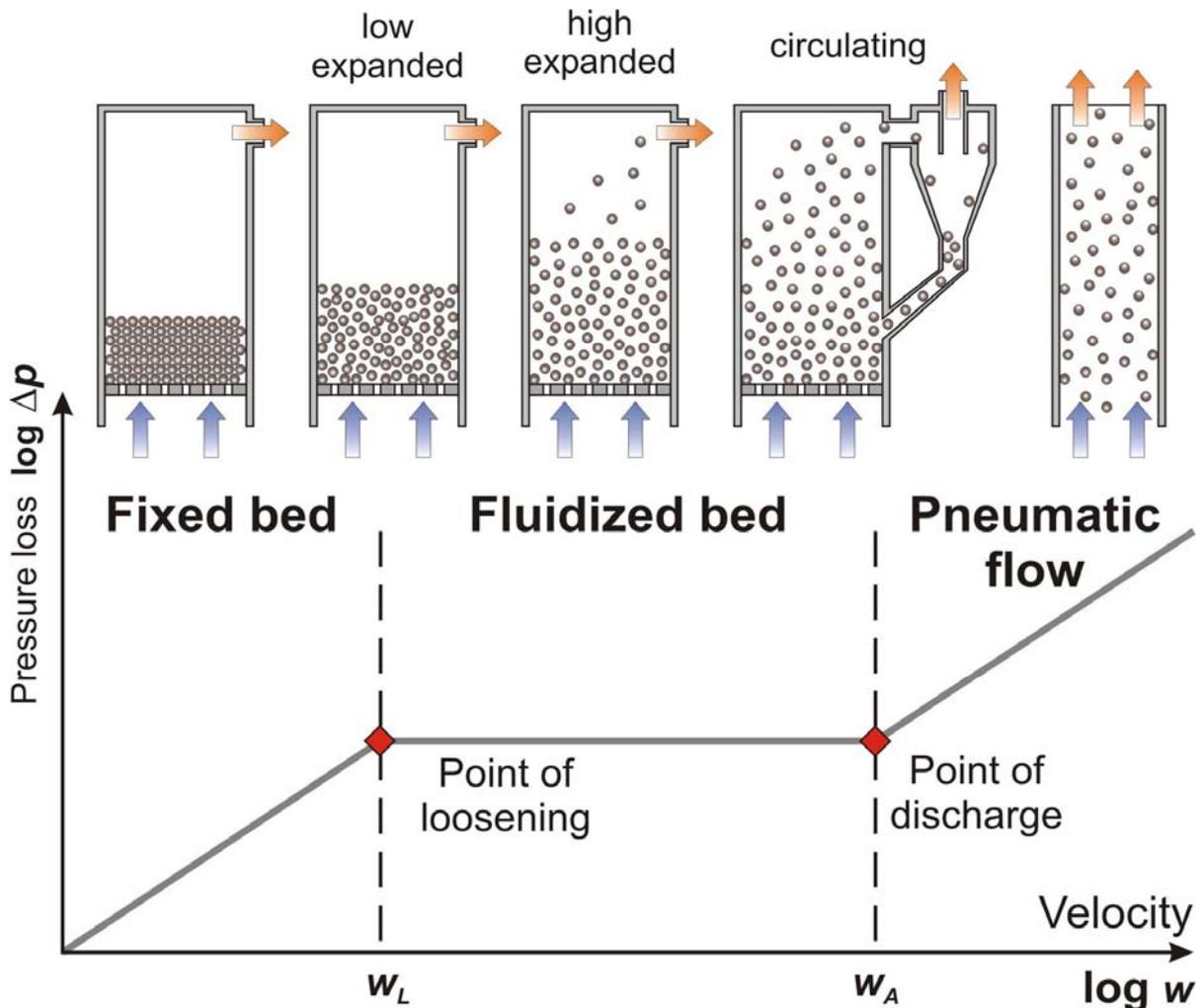
$$\dot{Q}_{V,2} = \Delta \dot{H}_{A-E} = \dot{m}_{Kuehlwasser} \cdot c_{pW} \cdot (\vartheta_{W,A} - \vartheta_{W,E}) \quad [kW] \quad (3-20)$$

In addition, the radiation and convection losses from the cooling jacket to the environment  $\dot{Q}_{V,3}$  were calculated.

### Calculation of the circulation rate of the bed material for the circulating drive mode

The circulation rate is a measure of the homogeneity of the circulating fluidized bed, e.g. with reference to an isothermal temperature profile in the reactor. The greater proportion of the heat which is transferred to heat the biomass is stored in the bed material. Conversely, the reaction enthalpy from the exothermic reactions due to partial combustion is transferred to the bed material.

The circulation rate can be calculated using the pressure loss which occurs in the fluidized bed, in dependence on the intake velocity, according to **Figure 3-3**.



**Figure 3-3:** Pressure loss in a fluidized bed depending on the fluid velocity [1]; figure from [2]

The total pressure loss of the fluidized bed reactor consists of the inflow floor pressure loss and the fluidized bed pressure loss  $\Delta p_{WS}$  and must be increased by the compressor for the inflow of the bed material.

The pressure loss of the fluidized bed  $\Delta p_{WS}$  can be calculated if the bed porosity  $\epsilon_{WS}$  is known, depending on the bed height  $h_{WS}$  using:

$$\Delta p_{WS} = h_{WS} \cdot (1 - \varepsilon_{WS}) \cdot (\rho_S - \rho_F) \cdot g \quad (3-21)$$

Since the bed porosity  $\varepsilon_{WS}$  is generally not known, the reverse process is used here: By measuring the pressure loss  $\Delta p_{WS}$  of the fluidized bed at the reactor head between the measuring points P 1007 and P 1008 the bed circulation rate  $\dot{m}_{Umlauf}$  can be calculated from the difference between the measurements  $\Delta h_{WS}$  using the following assumptions:

- 100 % of the solid material in the cell which is bounded by the measuring points P 1007 and P 1008 leaves the reactor.
- All particles in the cell move with the same speed as the gas.
- The input material content of the cell is not considered.
- The fluid density  $\rho_F$  is neglected.

The circulation rate can be calculated by considering the synthesis gas mass flow rate calculated from the mass balance:

$$\dot{m}_{Umlauf} = \frac{\dot{m}_{SG}}{\rho_{SG,Betrieb} \cdot (v_{SG})} \cdot (1 - \varepsilon_{WS}) \cdot \rho_S \cdot g = \frac{\dot{m}_{SG}}{\rho_{SG,Betrieb} \cdot (v_{SG})} \cdot \frac{\Delta p_{WS}}{\Delta h \cdot g} \left[ \frac{kg}{s} \right] \quad (3-22)$$

Depending on operating conditions and the bed material used, the circulation rates can be up to 18 t/h.

Using the loss of pressure of a fluidized bed, it is possible to assess whether a stable fluidisation is taking place. High pressure loss fluctuations during operation indicate, for example, a pounding fluidized bed. This form of the fluidized bed is usually undesirable. The progress of the pressure loss over the complete reactor height over time can show a loss of bed material, especially if the operating conditions are unfavourable for the cyclone.

## 4 Test procedure

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The emphasis for R&D in the area of synthesis gas production lies in the investigation of the raw synthesis gas composition and the content of trace components, dependent on the reaction temperature, turnaround time, composition of the gasification agent and type of fuel. In addition, an efficient catalytic bed additive (dolomite,  $\text{Al}_2\text{O}_3$  etc.) for the minimization of tar development inside the gasifier should be evaluated. The testing of additives (lime etc.) in order to shift the ash softening point of critical biomasses (straw) is also planned.

The R&D emphasis of the gas purification lies with the investigation of various wash liquids for their gas purification potential and the use of various adsorbers for the binding of trace components. This field will be investigated in the next phase of "Renew".

The target result of all variations is the production of a hydrogen-rich synthesis gas with minimal methane and tar content. To achieve this, high reactor temperatures ( $> 900\text{ }^\circ\text{C}$ ), a homogenous temperature field and sufficient turnaround time are necessary. The high temperatures are needed for kinetic reasons for the water gas reaction, sufficient turnaround time is needed for the breakdown of tars.

The following parameter variations were carried out:

- Variation of the input material:
  - ◆ Mixed wood chippings,
  - ◆ Mixed wood pellets,
  - ◆ Chipboard.
- Variation of the gasification agent:
  - ◆ Air,
  - ◆ Steam and oxygen mix in various proportions.
- With and without additional electrical heating
- Variation of the bed material (quartz sands of various particle sizes)

In planning are investigations with sugar beet chips, sunflower seed husks, wheat straw pellets and unprocessed softwood.

The reactor temperatures and the gas composition are set up as a result of these variations.

The minimum temperature in the investigations up to now, which can be regulated via the quantity of oxidation agent and the electrical heating, dependent on the input material mass flow rate, lay above  $900^\circ\text{C}$ .

---

## 4.1 Cold tests

For the cold tests the respective stock bunker is filled with the material to be investigated. Once a constant screw rotation speed is set up and there is an appropriate fill level in the screw, the material quantity per time unit is weighed.

A characteristic curve can be determined by plotting the mass flow rate via the screw speed. These characteristic curves have so far been plotted for mixed wood chippings, mixed wood pellets, straw pellets, chipboard and quartz sand.

## 4.2 Hot tests

Before loading with biomass the reactor must be preheated to a minimum of 600 to 700°C, in order to bring up the ignition energy for the biomass. For this the bed is warmed to the required temperature, in circulating mode and usually overnight, using pre-warmed air and with electrical heating. After this the biomass is introduced and, via a combustion procedure in the CFB, is warmed to approx. 900 °C. After reaching the target temperature the biomass flow rate is increased, i.e. the air volume is reduced, in order to achieve the gasification process. After this, and depending on the aim of the particular test, the composition of the gasification agent is changed from air to a steam/oxygen mixture and a stationary condition is strived for by observing the reactor temperatures and gas concentrations.

Once a stationary condition is reached, the measurements of the trace components etc. can begin.

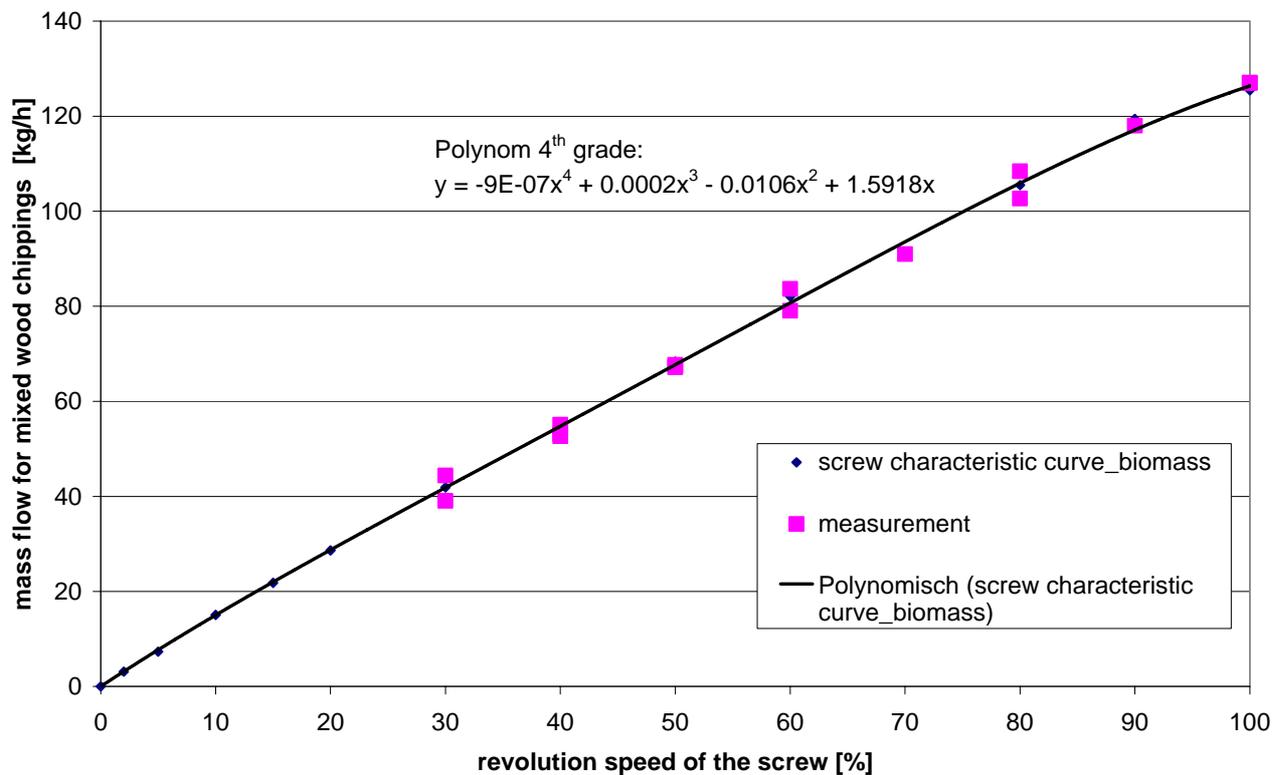
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## 5 Results

In the following two sections the first results of the investigations are presented.

### 5.1 Cold tests

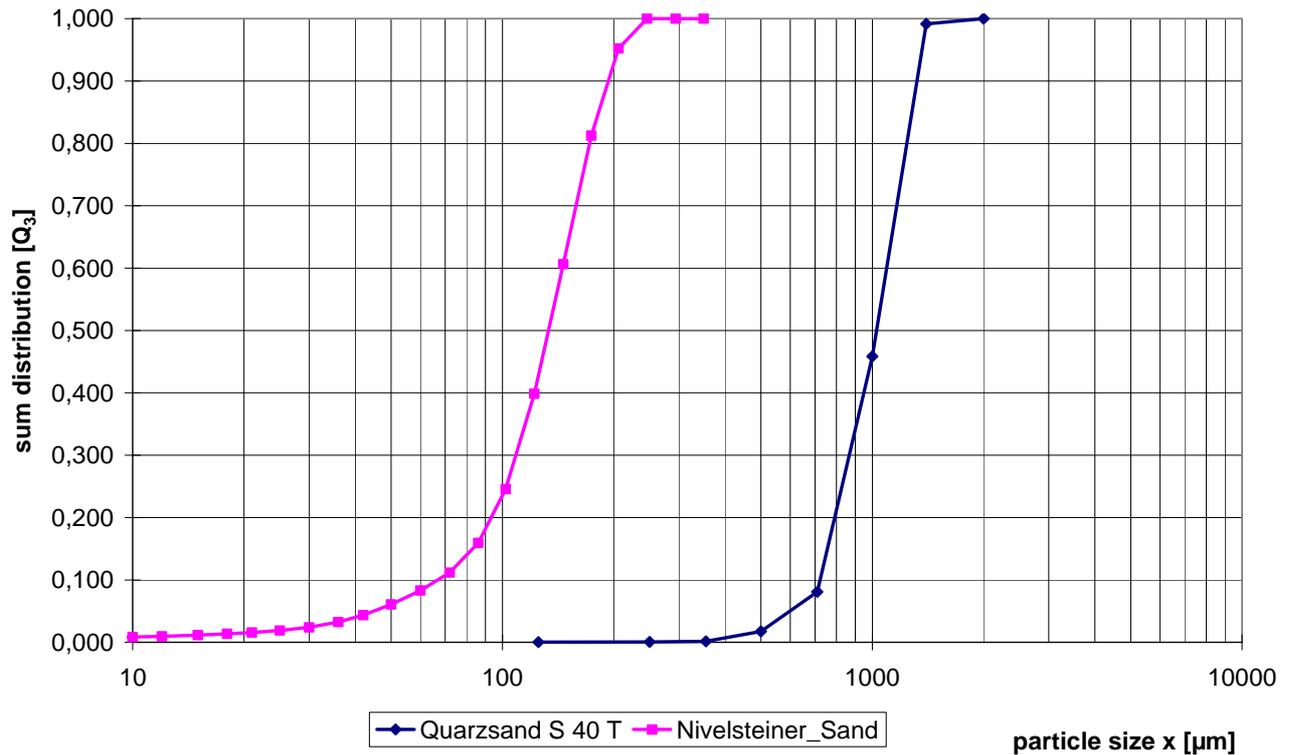
As an example of a screw characteristic curve, the process development for mixed wood chippings is depicted in **Figure 5-1**.



**Figure 5-1:** Screw characteristic curve for mixed wood chippings

Similar curves are available for other materials in the measurement collection program, so that the respective mass flow rates can be shown online.

In addition, for the characterisation of the quartz sands used, the respective value distributions were determined (see **Figure 5-2**).



**Figure 5-2:** Value distribution of the bed materials used

The median particle size of a sand determines, together with the particle size of the input material, the minimum fluidisation speed.

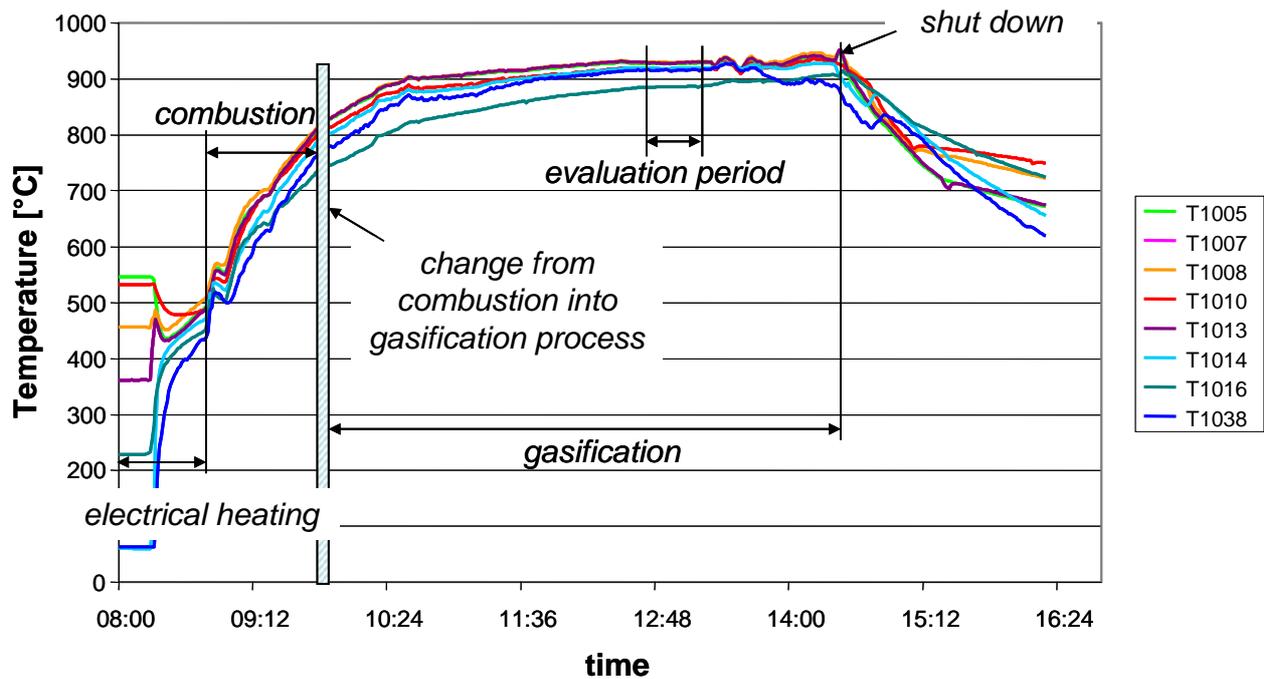
For the quartz sand S40T the median is approx. 1000  $\mu\text{m}$ , and for the Nivelsteiner sand N45 approx. 150  $\mu\text{m}$ .

## 5.2 Hot tests

For every investigation an evaluation period is defined in which the essential operating parameters (e.g. temperatures in the reactor) fluctuate only insignificantly. Short periods of non-stationary test time caused, for example, by fluctuations in the fuel input, are removed from the data for the evaluation.

The determination of a stationary operating point, for which the mass and energy balances are plotted, is decided on the basis of the progress of temperature variations during the test period.

As an example, **Figure 5-3** shows the investigation and evaluation time periods for a selected test with the reactor temperatures in the fluidized bed.



**Figure 5-3:** Temperature variation

Because stationary conditions can be reliably established in a fluidized bed the evaluation period can be short, as here for example, with 10 minutes. The balances are then plotted for this time period.

It should be noted that all the investigations described in this report serve to determine

- the main effects of using various biomasses,
- the main effects of the particle morphology of the standard biomass wood,
- technical imperfections of the apparatus,
- process technology parameters which can be improved (e.g. particle diameter of the sand, minimum temperature for the tar breakdown, material removal, precipitation behaviour of the cyclone)

In further phases it is planned to use the findings in terms of changes in the apparatus gained from these day tests in order to be able to perform multi-day tests as a follow-up. The aim must be to provide the "Chemical Process Technology" department with a purified synthesis gas for the FT synthesis.

The tests are shortened as follows:

**V1\_steam\_ox\_no\_electr\_chippings:**

Test for the steam/ oxygen gasification of mixed wood chippings without electrical heating.

**V2\_steam\_ox\_with\_electr\_chippings:**

Test for the steam/ oxygen gasification of mixed wood chippings with electrical heating.

**V3\_steam\_ox\_no\_electr\_time\_chippings:**

Test for the steam/ oxygen gasification of mixed wood chippings without electrical heating with increased residence time for the gas phase.

**V4\_air\_no\_electr\_pellets:**

Test for the airn gasification of mixed wood pellets without electrical heating.

**V5\_steam\_ox\_no\_electr\_pellets:**

Test for the steam/ oxygen gasification of mixed wood pellets without electrical heating.

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## 5.2.1 V1\_steam\_ox\_no\_electr\_chippings

Firstly, an overview of the main test parameters is shown in the test protocol in **Table 5-4**. Next, the test is described in brief and mass, material and energy balances are presented and discussed.

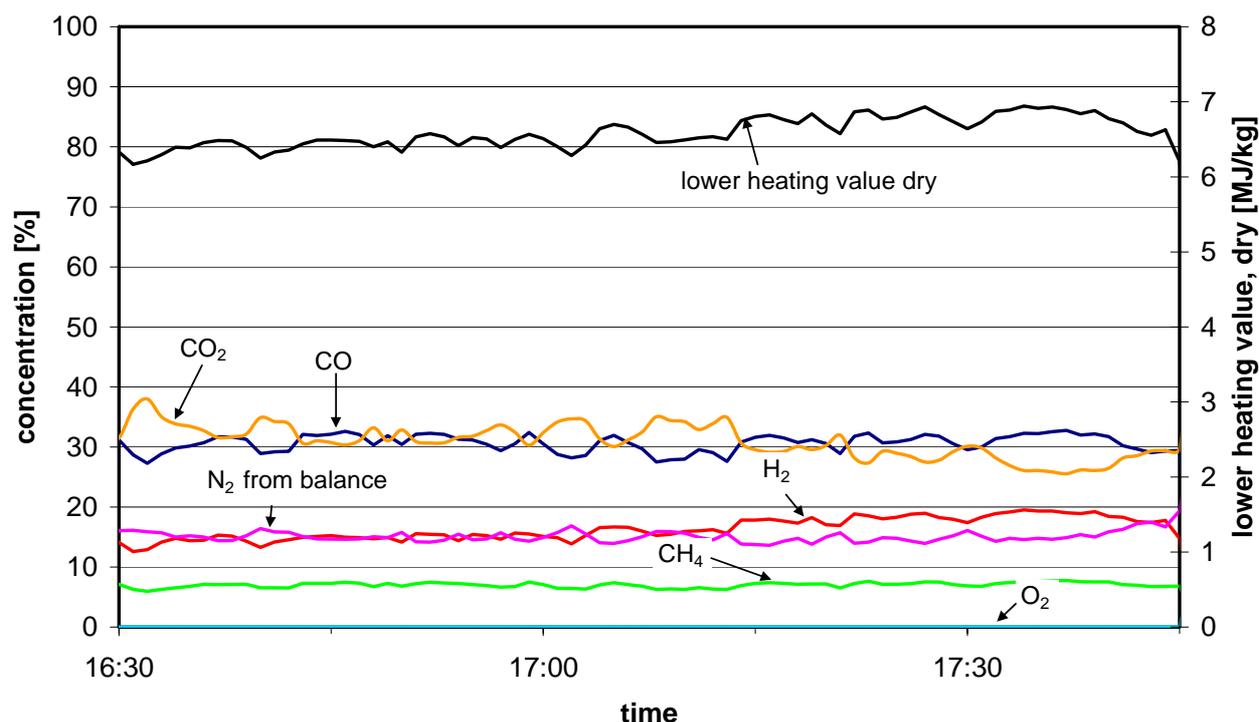
Date:	07.02.2005	Start	Stop	Duration [min]	
Total period		08:28	20:59		
Evaluation period		16:45	16:55	10,00	
<b>Type of input material:</b>	mixed wood chippings	<b>Composition of the reaction gas:</b>			oxygen/ steam
<b>total thermal power:</b>	[kW]	<b>529</b>			
<b>CFB-gasifier with cyclone</b>	mass flow		mass flow		mass flow
	[ kg/h ]		[ kg/h ]		[ kg/h ]
Input material	93,66	Ash	0,26		
Nitrogen	6,73				
Purge air	11,20			<b>Permanent gas, dry</b>	139,48
Steam	80,45	<b>Synthesis gas</b>	234,13	<b>Tar</b>	1,87
Air	0,00			<b>Steam</b>	91,78
Oxygen	42,35			<b>Fly char</b>	1,00
<b>Operational parameters</b>					
Bed circulation rate	[t/h]	6,81		Residence time of the gas in the CFB incl. cyclone [s]	
Average temperature of the reaction gas	[°C]	384,8		2,21	
Pressure above atmosphere	[ mbar ]	105,56		emptying tube speed [m/s]	
<b>Electrical heating</b>		Sum		3,86	
	[ kW ]	0,00			
<b>Distribution of temperatures</b>					
Reactor above the nozzles	[ °C ]	933	Ash	[ °C ]	14
Reactor lower third	[ °C ]	936			
Reactor middle	[ °C ]	943			
Reactor middle	[ °C ]	944			
Reacktor top	[ °C ]	933			
cyclone	[ °C ]	928			
Siphon	[ °C ]	935			
cyclone outlet	[ °C ]	904			
<b>Composition of the raw synthesis gas</b>					
compound	dry gas	water	fly char	tar	sum
mass-% (partly estimated)	59,57	39,20	0,43	0,80	100,00
<b>Gas concentration of the permanent gas from the continuously gas analysis</b>					
Choice	O <sub>2</sub> [Vol.-% dr.]	CO <sub>2</sub> [Vol.-% dr.]	CO [Vol.-% dr.]	H <sub>2</sub> [Vol.-% dr.]	CH <sub>4</sub> [Vol.-% dr.]
of compounds:	0,00	31,35	31,67	15,01	7,21

**Table 5-4:** Test protocol, V1\_steam\_ox\_no\_electr\_chippings for mixed wood chippings and steam/oxygen gasification

The investigation was carried out using mixed wood chippings and a steam/oxygen mixture with a total thermal output of approx. 530 kW. The individual mass flow rates are shown in the protocol. The purging air was introduced in the area of the input screw.

Further operating parameters are the bed circulation rate as a measurement of the mixing, at approx. 7 t/h, and the emptying tube speed, at approx. 4 m/s. The average gas throughput time was around 2 seconds. The reactor temperature can be seen as homogenous over the length of the reactor, at approx. 930 °C. As a bed material, the fine Nivelsteiner sand N45 was used. The methane content is noticeably high, at approx. 7 Vol.-% and the hydrogen content is correspondingly low with approx. 15 Vol.-%.

**Figure 5-5** shows the progress of concentration during the test period.



**Figure 5-5:** Gas concentrations, V1\_steam\_ox\_no\_electr\_chippings for mixed wood chippings and steam/oxygen gasification

One can see that the concentrations fluctuate relatively strongly. The reason for this is the batch input characteristic of the chippings from the input screw into the reactor. The material develops a characteristic which has a direct effect on the gas concentration. Furthermore, the relatively light chippings are quickly transported from the lower area of the fluidized bed into the upper part of the reactor, which, as will be shown later, in comparison to mixed wood pellets is more like a flow reactor gasification than a fluidized bed gasification. Thus the throughput and reaction times are additionally shortened for the chippings. This also explains the low hydrogen concentrations, although temperatures above 900 °C are present everywhere in the reactor.

The following **Table 5-6** shows an overview of the mass and energy balances.

The mass flow rates of the permanent gas, tar, steam and fly ash are iteratively determined together with the balance of materials. The following assumptions are made:

- Steam mass proportion of the raw synthesis gas: 40 Ma.-%
- Tar proportion in the steam: 2 Ma.-% (to calculate material values decane is used as a pseudo-component for tar)
- Ethane content in the permanent gas: 1.5 Vol.-% (substitutes higher hydrocarbon compounds in the permanent gas)
- Fly char quantity: 1 kg/h (shown in red)

Mass balance		Energy balance				
Input	m [kg/h]	$c_p$ [kJ/(kg K)]	$\vartheta$ [°C]	$h_u$ [kJ/kg]	H [kW]	$Q_{zu,elektr.}$ [kW]
Input material	93,66	1,00	20	17.114	445,8	0,00
Nitrogen	6,73	1,00	20	0	0,0	
Steam	80,45	1,94	385	0	79,2	
Oxygen	42,35	0,96	385	0	4,1	
Air	11,20	1,03	20	0	0,0	
Sum input	234,39					529,1
<b>Output</b>						$Q_v$ [kW]
Ash	0,26	1,00	14	0	0,0	40,8
Permanentgas, raw, dry	139,48	1,42	904	7.232	329,9	
Tar (estimated)	1,87	3,34	904	44.038	24,5	
Steam (estimated)	91,78	2,09	904	0	122,6	
Fly char (estimated)	1,00	2	904	33.913	9,9	
Sum Output	234,39					527,7
Failure Input/Output [%]	0,00					0,26

**Table 5-6:** Mass and energy balances, V1\_steam\_ox\_no\_electr\_chippings for mixed wood chippings and steam/oxygen gasification

The raw synthesis gas mass flow rate is calculated from all the input mass flow rates in accordance with equation 3-3. Thus the balance error for the masses is zero. The red highlighting for the fly char shows that one can vary the value in the relevant calculation table.

Using these assumptions the error in the energy balance is small enough to be ignored. The following **Table 5-7** shows an overview of the balance of materials.

Material	Input		Steam	Oxygen	Air	Nitrogen	Ash		Permanent-gas,dry	condensates (tar and water)		Fly char	
Distribution of mass raw [Ma-%]	39,96		34,32	18,07	4,78	2,87	0,11		59,51	39,96		0,43	
kg/h	93,66		80,45	42,35	11,20	6,73	0,26		139,48	93,65		1,00	
Composition	raw	dry	raw	raw	raw	raw	raw	dry	raw = dry	raw	dry	raw	raw
kg/h		86,43						0,26	139,48		1,87		1,00
Water [Ma.-%]	7,7		100,0	0,0	0,0	0,0	0		0,00	98			0
Ash [Ma.-%]	0,3	0,3	0,0	0,0	0,0	0,0	100,00	100,0	0,00	0,00	0,00	0,00	0,00
Volatiles [Ma.-%]	0,0	0,0							0,00				
Char [Ma.-%]	0,0	0,0							0,00				
C <sub>fix</sub> [Ma.-%]	0,0	0,0							0,00				
C [Ma.-%]	44,8	48,5	0,0	0,0	0,0	0,0	0,00	0,0	31,07	1,80	90,00	100,00	100,00
H [Ma.-%]	5,6	6,0	0,0	0,0	0,0	0,0	0	0,0	2,40	0,20	10,00	0,00	0,00
N [Ma.-%]	0,1	0,1	0,0	0,0	77,0	100,0	0	0,0	13,14	0,00	0,00	0,00	0,00
O <sub>diff</sub> [Ma.-%]	41,6	45,0	0,0	100,0	23,0	0,0	0	0,0	53,39	0,00	0,00	0,00	0,00
Sum	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,00	100,00	100,00	100,00	100,00
lower heating valuet h <sub>v</sub> [kJ/kg]	17.114		0	0	0	0	0	0	7.232	881	44.038	33.913	33.913
Water is calculated as H and O													
Ash- balance [kg/h]	0,26		0,0	0,0	0,0	0,0	0,26	0,26	0	0,00	0,000	0,00	0,000
C- balance [kg/h]	41,96		0,00	0,00	0,00	0,00	0,00	0,00	43,34	1,69	1,686	1,00	1,000
H- balance [kg/h]	6,03		8,94	0,00	0,00	0,00	0,00	0,00	3,35	10,39	0,187	0,00	0,000
O- balance [kg/h]	45,35		71,51	42,35	2,58	0,00	0,00	0,00	74,47	81,58	0,000	0,00	0,000
N- balance [kg/h]	0,05		0,00	0,00	8,62	6,73			18,33				
Sum [kg/h]	93,65		80,45	42,35	11,20	6,73	0,26	0,26	139,48	93,65	1,87	1,00	1,00
Element	Sum Input	Sum Output	Failure (regard to Input) [%]										
raw													
Ash- balance [kg/h]	0,26	0,26	0,0										
C- balance [kg/h]	41,96	46,02	9,7										
H- balance [kg/h]	14,97	13,73	8,3										
O- balance [kg/h]	161,79	156,05	3,5										
N- balance [kg/h]	15,40	18,33	19,0										

**Table 5-7:** Balance of materials, V1\_steam\_ox\_no\_electr\_chippings for mixed wood chippings and steam/oxygen gasification

The ash balance amounts to zero, because the ash quantities were calculated according to equation 3-2 for the input and output. No ash was removed during the investigation, because these quantities are small enough to be ignored. The figures marked in red show a simplification: Ash is considered here to be completely composed of inert material, where as the fly ash (fly char) is composed of 100 % carbon. In the test operation a mixed fraction of ash and coke is removed.

The H and O balance can be determined using the steam proportion. Thus the steam quantity increases by around 3 kg/h in relation to the sum of the occurring steam quantities. In principle, an increase in steam can be explained by water produced as a product of pyrolysis, and a reduction by the consumption of water through the water gas shift reaction.

In a comparison, the error in the N balance is the largest: The N content of the synthesis gas is calculated from the difference to 100 Vol.-% of the sum of the components. As a rule the gasification gas consists, in addition to the standard permanent gas components, as will be shown below, of further hydrocarbon compounds, which have not yet been determined in the first investigations. Thus the N content is correspondingly reduced and the balances come out more even. Generally, the error increases with a decrease in the absolute mass flow rate. At under 20 %, however, the error is still within an acceptable size range.

## 5.2.2 V2\_steam\_ox\_with\_electr\_chippings

In this investigation additional electric heating was used.

**Table 5-8** shows the important operating data.

Date:	07.02.2005	Start	Stop	Duration [min]	
Total period		08:28	20:59		
Evaluation period		17:30	17:40	10,00	
Type of input material:	mixed wood chippings	Composition of the reaction gas:			oxygen/ steam
total thermal power:	[kW]	561			
CFB-gasifier with cyclone	mass flow		mass flow		mass flow
	[ kg/h ]		[ kg/h ]		[ kg/h ]
Input material	93,66	Ash	0,26		
Nitrogen	6,73				
Purge air	11,20			Permanent gas, dry	135,16
Steam	82,00	Synthesis gas	226,93	Tar	1,82
Air	0,00			Steam	88,96
Oxygen	33,59			Fly char	1,00
<b>Operational parameters</b>					
Bed circulation rate	[t/h]	7,94		Residence time of the gas in the CFB incl. cyclone [s]	
Average temperature of the reaction gas	[ °C ]	400,9		2,16	
Pressure above atmosphere	[ mbar ]	99,97		emptying tube speed [m/s]	
<b>Electrical heating</b>					
	[ kW ]	30,00		3,94	
<b>Distribution of temperatures</b>					
Reactor above the nozzles	[ °C ]	955	Ash	[ °C ]	11
Reactor lower third	[ °C ]	957			
Reactor middle	[ °C ]	962			
Reactor middle	[ °C ]	963			
Reactor top	[ °C ]	949			
cyclone	[ °C ]	946			
Siphon	[ °C ]	958			
cyclone outlet	[ °C ]	926			
<b>Composition of the raw synthesis gas</b>					
compound	dry gas	water	fly char	tar	sum
mass-% (partly estimated)	59,56	39,20	0,44	0,80	100,00
<b>Gas concentration of the permanent gas from the continuously gas analysis</b>					
Choice	O <sub>2</sub> [Vol.-% dr.]	CO <sub>2</sub> [Vol.-% dr.]	CO [Vol.-% dr.]	H <sub>2</sub> [Vol.-% dr.]	CH <sub>4</sub> [Vol.-% dr.]
of compounds:	0,00	27,02	31,66	18,87	7,41

**Table 5-8:** Test protocol, V2\_steam\_ox\_with\_electr\_chippings for mixed wood chippings and steam/oxygen gasification with additional electrical heating

The investigation was carried out using mixed wood chippings and a steam/oxygen mixture with additional electric heating, with a total thermal power of approx. 560 kW. The individual mass flow rates are shown in the protocol. The bed circulation rate as an indicator of the mixing rate was approx. 8 t/h and thus lies in the size range of the last test. The superficial velocity and the gas throughput time remain nearly constant in comparison

to the previous test. The reactor temperature can be considered homogenous over the reactor length, at approx. 960 °C.

In comparison to the last test, the following main changes occurred under otherwise nearly constant test conditions:

- The reactor was supplied with approx. 30 kW of electrical power.
- The oxygen quantity was reduced.
- The gasification agent mixture temperature and the reactor temperature increased.
- The hydrogen content increased.

On the one hand, through the electric heating the losses from the reactor to the environment were reduced, on the other hand the radiation losses of the plant via the jacket extraction system increased, as a comparison of the heat losses from the energy balances shows:

Mass balance		Energy balance				
Input	m [kg/h]	$c_p$ [kJ/(kg K)]	$\vartheta$ [°C]	$h_u$ [kJ/kg]	H [kW]	$Q_{zu,elektr.}$ [kW]
Input material	93,66	1,00	20	17.114	445,8	30,0
Nitrogen	6,73	1,00	20	0	0,0	
Steam	82,00	1,95	401	0	81,5	
Oxygen	33,59	0,96	401	0	3,4	
Air	11,20	1,04	20	0	0,0	
Sum input	227,19				560,7	
<b>Output</b>						<b><math>Q_v</math> [kW]</b>
Ash	0,26	1,00	11	0	0,0	43,3
Permanentgas, raw, dry	135,16	1,48	926	8.110	356,1	
Tar (estimated)	1,82	3,35	926	44.038	23,8	
Steam (estimated)	88,96	2,10	926	0	120,2	
Fly char (estimated)	1,00	2	926	33.913	9,9	
Sum Output	227,19				553,3	
Failure Input/Output [%]	0,00				1,33	

**Table 5-9:** Mass and energy balances, V2\_steam\_ox\_with\_electr\_chippings for mixed wood chippings and steam/oxygen gasification with electric heating

With electric heating, approx. 43 kW are lost, without electric heating the plant loses approx. 40 kW (Table 5-8). Since the temperature level in the reactor should be held nearly constant, as a compensation for the external addition of heat the oxygen level was reduced by approx. 10 kg/h to approx. 33 kg/h, with the steam quantity remaining constant. Because of the reduced throughput due to the gasification agent heater the gasification agent mixture temperature increased from 385 °C to 401 °C. Together with the additionally added heat this resulted in a 30 °C increase in reactor temperature to 960 °C.

This temperature increase led together with less combustion of produced hydrogen with oxygen to an increase in the hydrogen component of approx. 4 Vol.-% (dry) to about 19 Vol.-% dry base (or 25 % related to the last test).

In order to minimize the difference between the input and output enthalpy flows, the following assumptions were made:

- Steam mass proportion of the raw synthesis gas: 40 Ma.-%
- Tar proportion in the steam: 2 Ma.-% (to calculate material values decane is used as a pseudo-component for tar)
- Ethane content in the permanent gas: 1.5 Vol.-% (substitutes higher hydrocarbon compounds in the permanent gas)
- Fly char quantity: 1 kg/h (shown in red)

These assumptions are the same as those from the previous test, in order to facilitate a better comparison of the tests in relation to the effect of the additional electric heating on the operating parameters.

On the basis of the mass balance and the assumptions, the following balance of materials results:

Material	Input		Steam	Oxygen	Air	Nitrogen	Ash		Permanent-gas,dry	condensates (tar and water)		Fly char	
Distribution of mass raw [Ma.-%]	41,23		36,09	14,79	4,93	2,96	0,11		59,49	39,95		0,44	
kg/h	93,66		82,00	33,59	11,20	6,73	0,26		135,16	90,77		1,00	
Composition	raw	dry	raw	raw	raw	raw	raw	dry	raw = dry	raw	dry	raw	raw
kg/h		86,43						0,26	135,16		1,82		1,00
Water [Ma.-%]	7,7		100,0	0,0	0,0	0,0	0		0,00	98			0
Ash [Ma.-%]	0,3	0,3	0,0	0,0	0,0	0,0	100,00	100,0	0,00	0,00	0,00	0,00	0,00
Volatiles [Ma.-%]	0,0	0,0							0,00				
Char [Ma.-%]	0,0	0,0							0,00				
C <sub>fix</sub> [Ma.-%]	0,0	0,0							0,00				
C [Ma.-%]	44,8	48,5	0,0	0,0	0,0	0,0	0,00	0,0	31,22	1,80	90,00	100,00	100,00
H [Ma.-%]	5,6	6,0	0,0	0,0	0,0	0,0	0	0,0	2,88	0,20	10,00	0,00	0,00
N [Ma.-%]	0,1	0,1	0,0	0,0	77,0	100,0	0	0,0	14,27	0,00	0,00	0,00	0,00
O <sub>diff</sub> [Ma.-%]	41,6	45,0	0,0	100,0	23,0	0,0	0	0,0	51,63	0,00	0,00	0,00	0,00
Sum	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,00	100,00	100,00	100,00	100,00
lower heating value h <sub>v</sub> [kJ/kg]	17.114		0	0	0	0	0	0	8.110	881	44.038	33.913	33.913
Water is calculated as H and O													
Ash- balance [kg/h]	0,26		0,0	0,0	0,0	0,0	0,26	0,26	0	0,00	0,000	0,00	0,000
C- balance [kg/h]	41,96		0,00	0,00	0,00	0,00	0,00	0,00	42,20	1,63	1,634	1,00	1,000
H- balance [kg/h]	6,03		9,11	0,00	0,00	0,00	0,00	0,00	3,89	10,07	0,182	0,00	0,000
O- balance [kg/h]	45,35		72,89	33,59	2,58	0,00	0,00	0,00	69,79	79,07	0,000	0,00	0,000
N- balance [kg/h]	0,05		0,00	0,00	8,62	6,73			19,29				
Sum [kg/h]	93,65		82,00	33,59	11,20	6,73	0,26	0,26	135,16	90,77	1,82	1,00	1,00
Element	Sum Input	Sum Output	Failure (regard to Input) [%]										
raw													
Ash- balance [kg/h]	0,26	0,26	0,0										
C- balance [kg/h]	41,96	44,83	6,8										
H- balance [kg/h]	15,14	13,95	7,9										
O- balance [kg/h]	154,41	148,86	3,6										
N- balance [kg/h]	15,40	19,29	25,2										

**Table 5-10:** Balance of materials, V2\_steam\_ox\_with\_electr\_chippings for mixed wood chippings and steam/oxygen gasification with electric heating

Here all the considered components show similar errors as in the previous case without electric heating.

### 5.2.3 V3\_steam\_ox\_no\_electr\_time\_chippings

In the next investigation with mixed wood chippings the steam mass flow was approximately halved in order to increase the gas throughput time and to find out to what extent this measure, together with an increase in the wood mass flow effects the hydrogen concentration.

Through the constant oxygen mass flow the energy availability remains the same. Through the reduction of the steam, less steam must be heated up to the reactor temperature, at the same time, however, more input material must be heated and decomposed.

**Table 5-11** shows the essential operation data.

Date:	02.03.2005	Start	Stop	Duration [min]	
Total period		08:00	20:59		
Evaluation period		14:50	17:50	180,00	
Type of input material:	mixed wood chippings	Composition of the reaction gas:			oxygen/ steam
total thermal power:	[kW]	550			
CFB-gasifier with cyclone	mass flow		mass flow		mass flow
	[ kg/h ]		[ kg/h ]		[ kg/h ]
Input material	106,32	Ash	0,29		
Nitrogen	6,65				
Purge air	0,00			Permanent gas, dry	140,88
Steam	43,22	Synthesis gas	197,05	Tar	1,10
Air	0,00			Steam	54,07
Oxygen	41,15			Fly char	1,00
<b>Operational parameters</b>					
Bed circulation rate	[t/h]	n.b.		Residence time of the gas in the CFB incl. cyclone [s]	
Average temperature of the reaction gas	[ °C ]	322,5		2,61	
Pressure above atmosphere	[ mbar ]	81,44		emptying tube speed [m/s]	
Electrical heating		Sum		3,26	
	[ kW ]	0,00			
<b>Distribution of temperatures</b>					
Reactor above the nozzles	[ °C ]	813	Ash	[ °C ]	14
Reactor lower third	[ °C ]	820			
Reactor middle	[ °C ]	906			
Reactor middle	[ °C ]	929			
Reacktor top	[ °C ]	898			
cyclone	[ °C ]	811			
Siphon	[ °C ]	813			
cyclone outlet	[ °C ]	866			
<b>Composition of the raw synthesis gas</b>					
compound	dry gas	water	fly char	tar	sum
mass-% (partly estimated)	71,49	27,44	0,51	0,56	100,00
<b>Gas concentration of the permanent gas from the continuously gas analysis</b>					
Choice	O <sub>2</sub> [Vol.-% dr.]	CO <sub>2</sub> [Vol.-% dr.]	CO [Vol.-% dr.]	H <sub>2</sub> [Vol.-% dr.]	CH <sub>4</sub> [Vol.-% dr.]
of compounds:	0,01	32,43	36,36	15,85	9,84

**Table 5-11:** Test protocol, V3\_steam\_ox\_no\_electr\_time\_chippings for mixed wood chippings and steam/oxygen gasification

The investigation was carried out using mixed wood chippings and a steam/oxygen mixture with a total thermal power of approx. 550 kW. The individual mass flow rates are shown in the protocol. The bed circulation rate could not be determined due to large pressure fluctuations and negative pressure differences. It can therefore be considered as zero. The superficial velocity fell due to the reduction in steam to around 3 m/s with a simultaneous increase in the gas throughput time to around 2.6 seconds.

The reactor temperatures are no longer homogenous because of the nonexistent or very low bed circulation: In the upper and lower areas of the reactor they are around 820 °C, whereas in the area of the fuel inlet in the middle, due to exothermic reactions, temperatures of up to 930 °C are reached. This temperature level is around the temperature of the V1\_steam\_ox\_no\_electr\_chippings. The much lower gasification agent mixture temperature of 322 °C possibly contributes to the lower temperature in the lower third of the gasifier.

The higher methane content (almost 10 Vol.-%) and the increased CO content (+ 5 Vol.-%) to 36 Vol.-%. are noticeable.

If one assumes the same energy availability as in the V1\_steam\_ox\_no\_electr\_chippings, then with the increased biomass mass flow there is possibly no longer enough energy available for the decomposition. With the inhomogeneous temperature distribution in the reactor, intermediate products from the pyrolysis which are not completely split cannot be broken down further despite the higher throughput time. This shows the importance of the relationship between homogeneous temperature distribution in the reactor on a high level (at least 900 °C) and the throughput time, especially the solid material throughput time. The solid material throughput time can be estimated with the help of suitable combustion models. With very fine materials, which, depending on the operating conditions, are quickly transported, the solid material throughput time can be considered approximately the same as the gas throughput time.

In order to minimize the errors between input and output mass and enthalpy flow rates, the following assumptions were made for drawing up the mass, material and energy balances:

- Steam mass proportion of the raw synthesis gas: 28 Ma.-%; this lower proportion takes into account the reduction of the steam mass flow
- Tar proportion in the steam: 2 Ma.-% (to calculate material values decane is used as a pseudo-component for tar)
- Ethane content in the permanent gas: 3 Vol.-% (substitutes higher hydrocarbon compounds in the permanent gas; here there is a higher proportion due to lower decomposition rates in the process)
- Fly char quantity: 1 kg/h (shown in red)

On the basis of these assumptions, the following balance results:

---

Mass balance		Energy balance				
Input	m [kg/h]	$c_p$ [kJ/(kg K)]	$\phi$ [°C]	$h_u$ [kJ/kg]	H [kW]	$Q_{zu,elektr.}$ [kW]
Input material	106,32	1,00	20	17.114	506,0	0,00
Nitrogen	6,65	1,00	20	0	0,0	
Steam	43,22	1,93	322	0	41,0	
Oxygen	41,15	0,95	322	0	3,3	
Air	0,00	1,03	20	0	0,0	
Sum input	197,34				<b>550,4</b>	
<b>Output</b>						<b><math>Q_v</math> [kW]</b>
Ash	0,29	1,00	14	0	0,0	39,1
Permanentgas, raw, dry	140,88	1,44	866	9.381	415,8	
Tar (estimated)	1,10	3,31	866	44.038	14,4	
Steam (estimated)	54,07	2,08	866	0	70,8	
Fly char (estimated)	1,00	2	866	33.913	9,9	
Sum Output	197,34				<b>550,0</b>	
Failure Input/Output [%]	0,00				0,06	

**Table 5-12:** Mass and energy balances, V3\_steam\_ox\_no\_electr\_time\_chippings for mixed wood chippings and steam/oxygen gasification

The energy balance is almost complete. On the basis of the mass balance and the assumptions the following balance of materials results:

Material	Input		Steam	Oxygen	Air	Nitrogen	Ash		Permanent-gas,dry	condensates (tar and water)		Fly char	
Distribution of mass raw [Ma-%]	53,87		21,90	20,85	0,00	3,37	0,15		71,39	27,96		0,51	
kg/h	106,32		43,22	41,15	0,00	6,65	0,29		140,88	55,17		1,00	
<b>Composition</b>	<b>raw</b>	<b>dry</b>	<b>raw</b>	<b>raw</b>	<b>raw</b>	<b>raw</b>	<b>raw</b>	<b>dry</b>	<b>raw = dry</b>	<b>raw</b>	<b>dry</b>	<b>raw</b>	<b>raw</b>
kg/h		98,11						0,29	140,88		1,10		1,00
Water [Ma.-%]	7,7		100,0	0,0	0,0	0,0	0		0,00	98		0	
Ash [Ma.-%]	0,3	0,3	0,0	0,0	0,0	0,0	100,00	100,0	0,00	0,00	0,00	0,00	0,00
Volatiles [Ma.-%]	0,0	0,0							0,00				0,00
Char [Ma.-%]	0,0	0,0							0,00				
$C_{fix}$ [Ma.-%]	0,0	0,0							0,00				
C [Ma.-%]	44,8	48,5	0,0	0,0	0,0	0,0	0,00	0,0	36,34	1,80	90,00	100,00	100,00
H [Ma.-%]	5,6	6,0	0,0	0,0	0,0	0,0	0	0,0	3,19	0,20	10,00	0,00	0,00
N [Ma.-%]	0,1	0,1	0,0	0,0	77,0	100,0	0	0,0	2,53	0,00	0,00	0,00	0,00
$O_{air}$ [Ma.-%]	41,6	45,0	0,0	100,0	23,0	0,0	0	0,0	57,95	0,00	0,00	0,00	0,00
Sum	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,00	100,00	100,00	100,00	100,00
lower heating valuet $h_u$ [kJ/kg]	17.114		0	0	0	0	0	0	9.381	881	44.038	33.913	33.913
Water is calculated as H and O													
Ash- balance [kg/h]	0,29		0,0	0,0	0,0	0,0	0,29	0,29	0	0,00	0,000	0,00	0,000
C- balance [kg/h]	47,63		0,00	0,00	0,00	0,00	0,00	0,00	51,19	0,99	0,993	1,00	1,000
H- balance [kg/h]	6,85		4,80	0,00	0,00	0,00	0,00	0,00	4,49	6,12	0,110	0,00	0,000
O- balance [kg/h]	51,48		38,42	41,15	0,00	0,00	0,00	0,00	81,64	48,06	0,000	0,00	0,000
N- balance [kg/h]	0,05		0,00	0,00	0,00	6,65			3,56				
Sum [kg/h]	106,30		43,22	41,15	0,00	6,65	0,29	0,29	140,88	55,17	1,10	1,00	1,00
Element	Sum Input	Sum Output	Failure (regard to Input) [%]										
raw													
Ash- balance [kg/h]	0,29	0,29	0,0										
C- balance [kg/h]	47,63	53,18	11,7										
H- balance [kg/h]	11,65	10,61	8,9										
O- balance [kg/h]	131,05	129,70	1,0										
N- balance [kg/h]	6,71	3,56	47,0										

**Table 5-13:** Balance of materials, V3\_steam\_ox\_no\_electr\_time\_chippings for mixed wood chippings and steam/oxygen gasification

Only the nitrogen balance has a larger error. To improve the balance the purging air quantity has been set to zero. How much of this purging air actually gets into the reactor cannot be determined. A further adjustment can only be performed via the purging nitrogen quantity.

## 5.2.4 V4\_air\_no\_electr\_pellets

With the suspicion that the chippings have too low a solid material throughput time in the reactor, the input material was changed to mixed wood pellets from the same sawmill. With this, all the chemical properties remain essentially the same; mechanical qualities are changed, such as, for example, the bulk density, the particle diameter and in connection with the reactor, the throughput time behaviour.

**Table 5-14** shows the main operating data for a test with pellets and air gasification.

Date:	9.3.05	Start	Stop	Duration [min]	
Total period		08:00	19:59		
Evaluation period		12:45	13:11	26,00	
Type of input material:	mixed wood pellets	Composition of the reaction gas:			air
total thermal power:	[kW]	462			
CFB-gasifier with cyclone	mass flow		mass flow		mass flow
	[ kg/h ]		[ kg/h ]		[ kg/h ]
Input material	92,01	Ash	0,25		
Nitrogen	6,72				
Purge air	10,00			Permanent gas, dry	273,37
Steam	0,00	Synthesis gas	288,81	Tar	0,43
Air	180,34			Steam	14,01
Oxygen	0,00			Fly char	1,00
<b>Operational parameters</b>					
Bed circulation rate	[t/h]	4,69		Residence time of the gas in the CFB incl. cyclone [s]	
Average temperature of the reaction gas	[ °C ]	459,7		2,09	
Pressure above atmosphere	[ mbar ]	153,55		emptying tube speed [m/s]	
Electrical heating		Sum		4,07	
	[ kW ]	0,00			
<b>Distribution of temperatures</b>					
Reactor above the nozzles	[ °C ]	928	Ash	[ °C ]	12
Reactor lower third	[ °C ]	929			
Reactor middle	[ °C ]	930			
Reactor middle	[ °C ]	930			
Reactor top	[ °C ]	917			
cyclone	[ °C ]	862			
Siphon	[ °C ]	919			
cyclone outlet	[ °C ]	887			
<b>Composition of the raw synthesis gas</b>					
compound	dry gas	water	fly char	tar	sum
mass-% (partly estimated)	94,65	4,85	0,35	0,15	100,00
<b>Gas concentration of the permanent gas from the continuously gas analysis</b>					
Choice	O <sub>2</sub> [Vol.-% dr.]	CO <sub>2</sub> [Vol.-% dr.]	CO [Vol.-% dr.]	H <sub>2</sub> [Vol.-% dr.]	CH <sub>4</sub> [Vol.-% dr.]
of compounds:	0,00	14,99	14,62	9,11	3,53
(dry and N <sub>2</sub> -free):	0,00	34,66	33,80	21,07	8,15

**Table 5-14:** Test protocol, V4\_air\_no\_electr\_pellets for mixed wood pellets and air gasification

The investigation was carried out using mixed wood pellets and air with a total thermal power of approx. 460 kW. The individual mass flow rates are shown in the protocol. In comparison with the tests with chippings, here the content of the fluidized bed (bed material) was mixed: Around 70 kg Nivelsteiner sand and sand S40 T was used. The bed circulation rate as an indication of the mixing lies here, with approx. 4.7 t/h under the rate with only Nivelsteiner Sand, as was used in the tests with chippings, although the superficial velocities of approx. 4 m/s, are equally high. The very fine Nivelsteiner sand (**Figure 5-2**) is removed from the reactor via the cyclone during the test period, depending on the gas velocity. Due to the fineness of the sand, in the tests with chippings high circulation rates were achieved at the beginning of the test. The bed material is reduced over time through losses via the cyclone. By mixing with courser S40T, the latter remains in the reactor and only the finer Nivelsteiner sand is gradually transported out. Therefore the bed circulation rate is also lower in the test being considered. The average gas throughput time is around 2 seconds. The reactor temperature can be considered homogeneous over the reactor length, at approx. 930 °C, and lies in the same range as the temperatures with wood chippings (V1\_steam\_ox\_no\_electr\_chippings).

In relation to the hydrogen yield one can easily see, after calculating out the nitrogen fraction, that changing the fuel from the lighter chippings to the heavier pellets has a positive effect on the hydrogen concentration, which is approx. 21 Vol.-%. In contrast, the methane concentration, at around 8 Vol.-%, is in a similar range as with the chippings and steam/oxygen gasification. Obviously the change of gasification medium plays in this case a less important role. As already mentioned, the reason for the increase in hydrogen is assumed to be a longer solid material throughput time with the pellets in the hot part of the reactor, because the gas throughput time is, at 2 seconds, fairly low.

In order to minimize the errors between input and output mass and enthalpy flow rates, the following assumptions were made for drawing up the mass, material and energy balances:

- Steam mass proportion of the raw synthesis gas: 5 Ma.-%
- Tar proportion in the gas: 3 Ma.-% (to calculate material values decane is used as a pseudo-component for tar; assumption of a slightly higher tar content than for gasification with air)
- Ethane content in the permanent gas: 1 Vol.-% (substitutes higher hydrocarbon compounds in the permanent gas)
- Fly char quantity: 1 kg/h (shown in red)

On the basis of these assumptions, the following balance results:

Mass balance		Energy balance				
Input	m [kg/h]	$c_p$ [kJ/(kg K)]	$\vartheta$ [°C]	$h_u$ [kJ/kg]	H [kW]	$Q_{zu,elektr.}$ [kW]
Input material	92,01	1,00	20	17.114	437,9	0,00
Nitrogen	6,72	1,00	20	0	0,0	
Steam	0,00	1,97	460	0	0,0	
Oxygen	0,00	0,97	460	0	0,0	
Air	190,34	1,04	460	0	24,2	
Sum input	289,06					<b>462,2</b>
<b>Output</b>						<b><math>Q_v</math> [kW]</b>
Ash	0,25	1,00	12	0	0,0	48,6
Permanentgas, raw, dry	273,37	1,28	887	3.823	376,7	
Tar (estimated)	0,43	3,32	887	44.038	5,7	
Steam (estimated)	14,01	2,08	887	0	18,5	
Fly char (estimated)	1,00	2	887	33.913	9,9	
Sum Output	289,06					<b>459,3</b>
Failure Input/Output [%]	0,00					0,62

**Table 5-15:** Mass and energy balances, V4\_air\_no\_electr\_pellets for mixed wood pellets and air gasification

The difference in the heating value of the dry permanent gas between air and steam/oxygen gasification can be easily seen: Whilst with the steam/oxygen gasification the heating value, through condensing out of the inert water, lies at 9.4 MJ/kg, here the dilution with nitrogen means that only 3.8 MJ/kg is reached.

The energy balance is almost complete. On the basis of the mass balance and the assumptions the following balance of materials results:

Material	Input		Steam	Oxygen	Air	Nitrogen	Ash		Permanent-gas,dry	condensates (tar and water)		Fly char	
Distribution of mass raw [Ma.-%]	31,83		0,00	0,00	65,85	2,32	0,09		94,57	5,00		0,35	
kg/h	92,01		0,00	0,00	190,34	6,72	0,25		273,37	14,44		1,00	
Composition	raw	dry	raw	raw	raw	raw	raw	dry	raw = dry	raw	dry	raw	raw
kg/h		84,91						0,25	273,37		0,43		1,00
Water [Ma.-%]	7,7		100,0	0,0	0,0	0,0	0		0,00	97		0	
Ash [Ma.-%]	0,3	0,3	0,0	0,0	0,0	0,0	100,00	100,0	0,00	0,00	0,00	0,00	0,00
Volatiles [Ma.-%]	0,0	0,0							0,00				
Char [Ma.-%]	0,0	0,0							0,00				
C <sub>fix</sub> [Ma.-%]	0,0	0,0							0,00				
C [Ma.-%]	44,8	48,5	0,0	0,0	0,0	0,0	0,00	0,0	15,26	2,70	90,00	100,00	100,00
H [Ma.-%]	5,6	6,0	0,0	0,0	0,0	0,0	0	0,0	1,39	0,30	10,00	0,00	0,00
N [Ma.-%]	0,1	0,1	0,0	0,0	77,0	100,0	0	0,0	57,52	0,00	0,00	0,00	0,00
O <sub>diff</sub> [Ma.-%]	41,6	45,0	0,0	100,0	23,0	0,0	0	0,0	25,83	0,00	0,00	0,00	0,00
Sum	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,00	100,00	100,00	100,00	100,00
lower heating valuet h <sub>v</sub> [kJ/kg]	17.114		0	0	0	0	0	0	3.824	1.321	44.038	33.913	33.913
Water is calculated as H and O													
Ash- balance [kg/h]	0,25		0,0	0,0	0,0	0,0	0,25	0,25	0	0,00	0,000	0,00	0,000
C- balancez [kg/h]	41,22		0,00	0,00	0,00	0,00	0,00	0,00	41,72	0,39	0,390	1,00	1,000
H- balancez [kg/h]	5,93		0,00	0,00	0,00	0,00	0,00	0,00	3,79	1,60	0,043	0,00	0,000
O- balance [kg/h]	44,55		0,00	0,00	43,78	0,00	0,00	0,00	70,61	12,45	0,000	0,00	0,000
N- balance [kg/h]	0,05		0,00	0,00	146,56	6,72			157,24				
Sum [kg/h]	92,00		0,00	0,00	190,34	6,72	0,25	0,25	273,37	14,44	0,43	1,00	1,00
Element	Sum Input	Sum Output	Failure (regard to Input) [%]										
raw													
Ash- balance [kg/h]	0,25	0,25	0,0										
C- balance [kg/h]	41,22	43,11	4,6										
H- balance [kg/h]	5,93	5,39	9,0										
O- balance [kg/h]	88,33	83,07	6,0										
N- balance [kg/h]	153,32	157,24	2,6										

**Table 5-16:** Balance of materials, V4\_air\_no\_electr\_pellets for mixed wood pellets and air gasification

In comparison to the previous balances, the nitrogen balance can be well completed here, because the large absolute nitrogen quantities here greatly reduce the error between input and output.

The remaining errors due to the assumptions are of an acceptable size.

## 5.2.5 V5\_steam\_ox\_no\_electr\_pellets

In this investigation the main change was that the gasification agent was changed from air to steam/oxygen. In addition, in this test a tar and steam determination in the raw gas was carried out, and the higher hydrocarbon concentration in the permanent gas was also determined.

**Table 5-17** shows the main operating data for a test with pellets and steam/oxygen gasification.

Date:	30.03.2005	Start	Stop	Duration [min]	
Total period		08:00	19:59		
Evaluation period		16:15	16:26	11,00	
Type of input material:	mixed wood pellets	Composition of the reaction gas:			steam/ oxygen
total thermal power:	[kW]	509			
CFB-gasifier with cyclone	mass flow		mass flow		mass flow
	[ kg/h ]		[ kg/h ]		[ kg/h ]
Input material	96,73	Ash	0,27		
Nitrogen	0,00				
Purge air	0,20			Permanent gas, dry	112,13
Steam	46,96	Synthesis gas	183,03	Tar	2,04
Air	0,00			Steam	68,06
Oxygen	39,61			Fly char	1,00
<b>Operational parameters</b>					
Bed circulation rate	[t/h]	n.b.		Residence time of the gas in the CFB incl. cyclone [s]	
Average temperature of the reaction gas	[°C]	343,2		2,69	
Pressure above atmosphere	[ mbar ]	88,20		emptying tube speed [m/s]	
<b>Electrical heating</b>					
	[ kW ]	0,00		3,17	
<b>Distribution of temperatures</b>					
Reactor above the nozzles	[ °C ]	939	Ash	[ °C ]	17
Reactor lower third	[ °C ]	938			
Reactor middle	[ °C ]	943			
Reactor middle	[ °C ]	946			
Reacktor top	[ °C ]	901			
cyclone	[ °C ]	910			
Siphon	[ °C ]	804			
cyclone outlet	[ °C ]	881			
<b>Composition of the raw synthesis gas</b>					
compound	dry gas	water	fly char	tar	sum
mass-% (partly estimated)	61,26	37,19	0,55	1,11	100,11
<b>Gas concentration of the permanent gas from the continuously gas analysis</b>					
Choice	O <sub>2</sub> [Vol.-% dr.]	CO <sub>2</sub> [Vol.-% dr.]	CO [Vol.-% dr.]	H <sub>2</sub> [Vol.-% dr.]	CH <sub>4</sub> [Vol.-% dr.]
of compounds:	0,02	33,69	29,91	26,04	8,87

**Table 5-17:** Test protocol, V5\_steam\_ox\_no\_electr\_pellets, for mixed wood pellets and hydrogen/oxygen gasification

The figure highlighted in red is the purging air quantity, which has been adapted here to complete the nitrogen balance (see **Table 5-21**).

The test was carried out using mixed wood pellets and steam/oxygen with a total thermal power of approx. 510 kW. The individual mass flow rates are shown in the protocol. In comparison to the test with air as a gasification agent, the gasification agent flow rate is reduced here in order to increase the gas throughput time to 2.7 seconds. As with the test V3\_steam\_ox\_no\_electr\_time\_chippings the bed circulation rate could not be determined because of the low, sometimes negative pressure differences occurring at the reactor head. The temperature distribution in the reactor is homogeneous to the reactor head. Because of the very low bed circulation rate mentioned above, the hot sand did not

completely reach the reactor head and the cyclone. Therefore the temperatures here, in comparison to the reactor middle, are up to 60 °C lower (cyclone exit, reactor middle).

In relation to the hydrogen yield, here one can clearly see an increase in the hydrogen concentration compared with air gasification of over 5 Vol.-% to over 26 Vol.-%, with a slightly increased methane content. With the same input material, and despite the lower bed circulation rate, this can be traced back to the change of gasification medium.

**Table 5-18** shows the gas concentrations, which were obtained intermittently using a gas probe bag:

compound		vol.-% dry
hydrogen	H <sub>2</sub>	26,0414
oxygen	O <sub>2</sub>	0,0200
nitrogen	N <sub>2</sub>	0,1700
methane	CH <sub>4</sub>	8,8735
carbonmonoxide	CO	29,9070
steam	H <sub>2</sub> O (g)	0,0000
carbondioxide	CO <sub>2</sub>	33,6900
ethane	C <sub>2</sub> H <sub>6</sub>	0,0356
ethene	C <sub>2</sub> H <sub>4</sub>	1,3393
ethine	C <sub>2</sub> H <sub>2</sub>	0,5811
propane	C <sub>3</sub> H <sub>8</sub>	0,0021
propene	C <sub>3</sub> H <sub>6</sub>	0,0103
butane	C <sub>4</sub> H <sub>10</sub>	0,0023
pentane	C <sub>5</sub> H <sub>12</sub>	0,0027
hexane	C <sub>6</sub> H <sub>14</sub>	0,0045
sum		100,6798

**Table 5-18:** Dry synthesis gas concentrations, V5\_steam\_ox\_no\_electr\_pellets, for mixed wood pellets and steam/oxygen gasification

It is noticeable that particularly the unsaturated hydrocarbons ethene and ethyne have higher concentrations in comparison to ethane. This can be attributed to incomplete decomposition reactions in the pyrolysis. Further higher aliphatic hydrocarbon compounds do not need to be determined in future; they play only a minor role.

Within an hour of sample taking with the apparatus shown in **Figure 2-5** a total of

- 5 g tar collected in the apparatus on the cooling surfaces and
- 167 g watery condensate was collected in the conical flasks.

If a few suspended solids are ignored, the condensate consists of 100 % water. As a proportion of the raw gas from the reactor, the steam percentage is approx. 37 Ma.-%, the tar approx. 1 Ma.-%.

The deposits in the cooling trap can be seen in **Figure 5-19**.



**Figure 5-19:** Deposits in the cooling trap after day V5\_steam\_ox\_no\_electr\_pellets

The exact composition of the yellowish deposits is still being investigated.

In order to minimize the difference between input and output mass and enthalpy flow rates, only the fly ash mass flow is unknown in order to complete the mass, material and energy balances:

- Tar is treated as naphthalene, and for the calculation of material values decane is used as a pseudo-component for tar
- Fly char quantity: 1 kg/h (shown in red)

On the basis of these assumptions, the following balance results:

Mass balance		Energy balance				
Input	m [kg/h]	$c_p$ [kJ/(kg K)]	$\phi$ [°C]	$h_u$ [kJ/kg]	H [kW]	$Q_{zu,elektr.}$ [kW]
Input material	96,73	1,00	20	17.114	460,4	0,00
Nitrogen	0,00	1,00	20	0	0,0	
Steam	46,96	1,93	343	0	45,1	
Oxygen	39,61	0,96	343	0	3,4	
Air	0,20	1,03	20	0	0,0	
Sum input	183,49					<b>508,9</b>
<b>Output</b>						<b><math>Q_v</math> [kW]</b>
Ash	0,27	1,00	17	0	0,0	53,9
Permanentgas, raw, dry	112,13	1,58	881	9.556	340,9	
Steam	68,06	2,08	881	0	89,9	
Tar as naphthalene	2,04	3,32	881	39.848	24,2	
Fly char (estimated)	1,00	2	881	33.913	9,9	
Sum Output	183,49					<b>518,8</b>
Failure Input/Output [%]	0,00					1,94

**Table 5-20:** Mass and energy balances, V5\_steam\_ox\_no\_electr\_pellets, for mixed wood pellets and hydrogen/oxygen gasification

If the fly char quantity is set at zero, the error in the energy balance can be further reduced. It is, however, unlikely that with the pellets, in comparison with the chippings, absolutely no fly char will be produced.

On the basis of the mass balances the following balance of materials can be drawn up:

Material	Input		Steam	Oxygen	Air	Nitrogen	Ash		Permanent-gas,dry	condensates (tar and water)		Fly char	
Distribution of mass raw [Ma.-%]	52,72		25,59	21,58	0,11	0,00	0,14		61,11	38,20		0,54	
kg/h	96,73		46,96	39,61	0,20	0,00	0,27		112,13	70,10		1,00	
Composition	raw	dry	raw	raw	raw	raw	raw	dry	raw = dry	raw	dry	raw	raw
kg/h		89,26						0,27	112,13		2,04		1,00
Water [Ma.-%]	7,7		100,0	0,0	0,0	0,0	0		0,00	97,093		0	
Ash [Ma.-%]	0,3	0,3	0,0	0,0	0,0	0,0	100,00	100,0	0,00	0,00	0,00	0,00	0,00
Volatiles [Ma.-%]	0,0	0,0							0,00				
Char [Ma.-%]	0,0	0,0							0,00				
C <sub>fix</sub> [Ma.-%]	0,0	0,0							0,00				
C [Ma.-%]	44,8	48,5	0,0	0,0	0,0	0,0	0,00	0,0	35,60	2,63	90,57	100,00	100,00
H [Ma.-%]	5,6	6,0	0,0	0,0	0,0	0,0	0	0,0	3,66	0,27	9,43	0,00	0,00
N [Ma.-%]	0,0	0,1	0,0	0,0	77,0	100,0	0	0,0	0,19	0,00	0,00	0,00	0,00
O <sub>diff</sub> [Ma.-%]	41,6	45,0	0,0	100,0	23,0	0,0	0	0,0	60,55	0,00	0,00	0,00	0,00
Sum	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,00	100,00	100,00	100,00	100,00
lower heating value h <sub>u</sub> [kJ/kg]	17,114		0	0	0	0	0	0	9,556	0	39,848	33,913	33,913
Water is calculated as H and O													
Ash- balance [kg/h]	0,27		0,0	0,0	0,0	0,0	0,27	0,27	0	0,00	0,000	0,00	0,000
C- balance [kg/h]	43,33		0,00	0,00	0,00	0,00	0,00	0,00	39,92	1,85	1,846	1,00	1,000
H- balance [kg/h]	6,23		5,22	0,00	0,00	0,00	0,00	0,00	4,11	7,76	0,192	0,00	0,000
O- balance [kg/h]	46,84		41,74	39,61	0,05	0,00	0,00	0,00	67,89	60,50	0,000	0,00	0,000
N- balance [kg/h]	0,05		0,00	0,00	0,15	0,00			0,21				
Sum [kg/h]	96,72		46,96	39,61	0,20	0,00	0,27	0,27	112,13	70,10	2,04	1,00	1,00
Element	Sum Input	Sum Output	Failure (regard to Input) [%]										
raw													
Ash- balance [kg/h]	0,27	0,27	0,0										
C- balance [kg/h]	43,33	42,76	1,3										
H- balance [kg/h]	11,45	11,86	3,6										
O- balance [kg/h]	128,23	128,40	0,1										
N- balance [kg/h]	0,20	0,21	2,6										

**Table 5-21:** Balance of materials, V5\_steam\_ox\_no\_electr\_pellets, for mixed wood pellets and hydrogen/oxygen gasification

The nitrogen balance was calculated via the adaptation of the purging air quantity. Under consideration of the only unknown, the fly ash quantity, the C, H and O balances can be completed very well.

## 5.2.6 Comparison of the day tests

In this section the Results will be summarised and compared in graphical form in dependence on the main operating parameters.

This includes

- 1: V1\_steam\_ox\_no\_electr\_chippings
- 2: V2\_steam\_ox\_with\_electr\_chippings
- 3: V3\_steam\_ox\_no\_electr\_time\_chippings
- 4: V4\_air\_no\_electr\_pellets
- 5: V5\_steam\_ox\_no\_electr\_pellets

The following two figures do not take into account the influence of the gas throughput time, the bed circulation rate, and the reactor temperatures on the hydrogen concentrations. The reactor temperatures varied in the day tests as a rule between 900°C and 960°C. This parameter also influences the gas concentration of course, but here a simplified overview of the main parameters is presented:

- Type of input material
  - Gasification agent composition
  - Electric heating
-

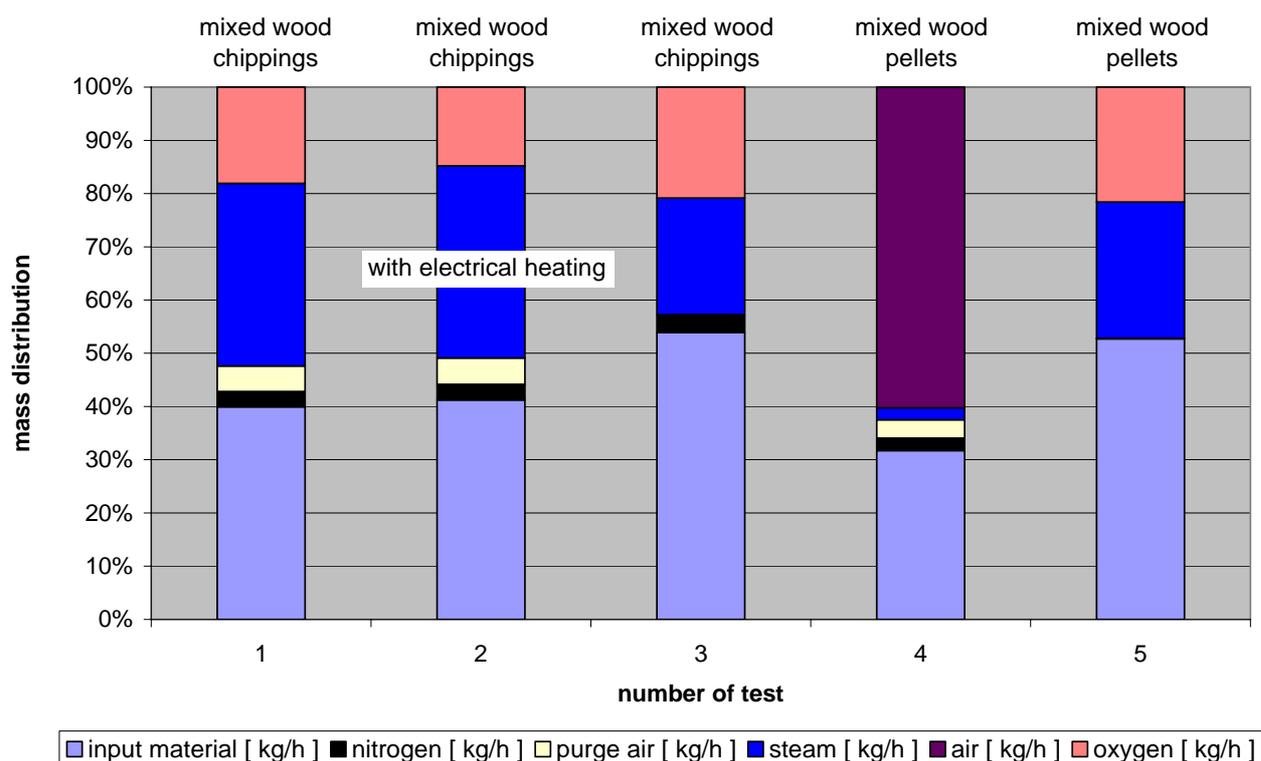


Figure 5-22: Comparison of the tests using the occurring mass flow rates

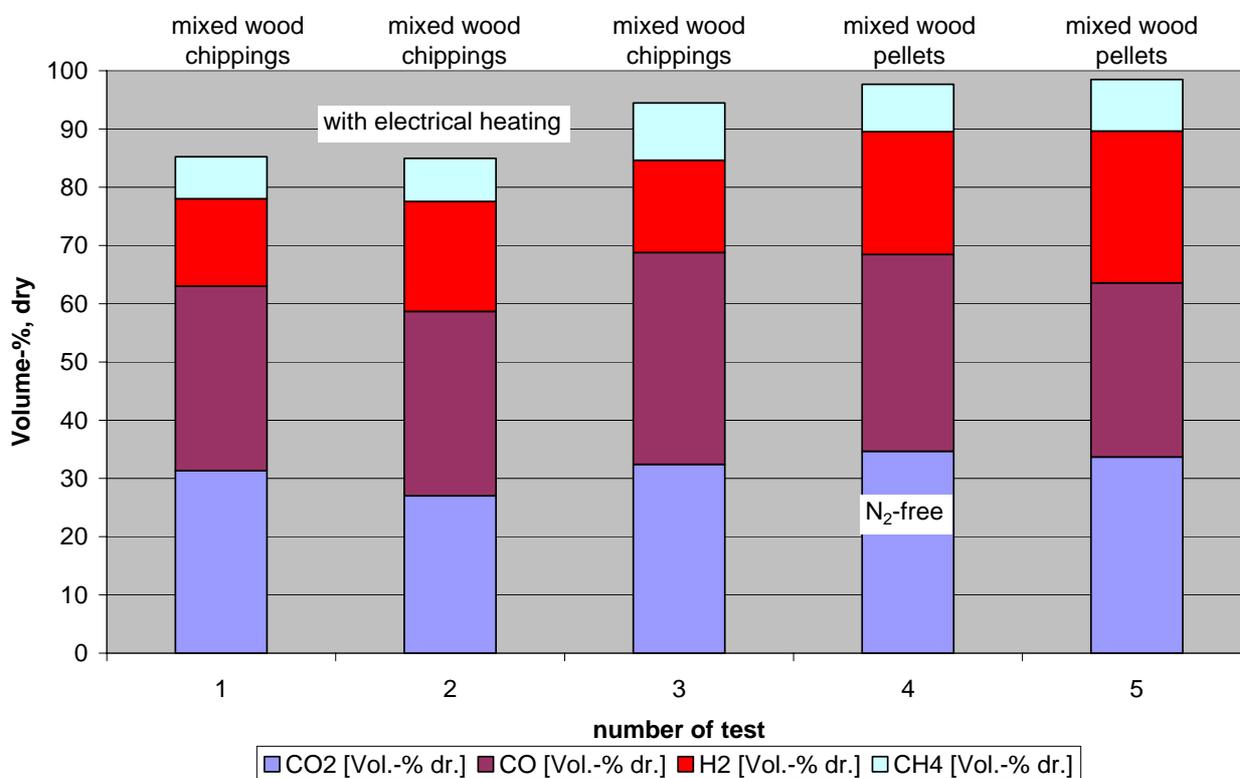


Figure 5-23: Comparison of the tests using the gas concentrations

In a comparison of the test set-ups used, the maximum hydrogen yield was obtained using pellets with steam and oxygen as a gasification agent. The pellets, in comparison with the chippings, give a higher solid material throughput time in the reactor, with which by homogeneous temperature distribution higher concentrations occurred. The methane contents are of similar size between 7 Vol.-% and almost 10 Vol.-% in test 3. With the help of the electric heating, oxygen can be saved, to the advantage of a slightly higher hydrogen content in the synthesis gas. A gasification with steam and oxygen increases the hydrogen concentration considerably compared to a gasification with air (tests 4 and 5).

### 5.2.7 The use of chipboard to heat up the fluidized bed

In order to investigate the influence of the particle size on the behaviour of the input screw and the turnover in the fluidized bed, a test with chipboard pieces to heat up the fluidized bed through combustion was carried out. The pieces of chipboard had an average length of approx. 5 cm (**figure 5-24**).



**figure 5-24:** Chipboard pieces

The heating of the CFB in the combustion mode worked without problems. Instabilities in the fluidized bed could not be determined.

A stoppage of the rotary feeder led to termination of the heating procedure. This was caused by foreign substances, e.g. the wood pieces shown in **Figure 5-25**, which were more than 20 cm long.



**Figure 5-25:** Long wood fragments

From this test one can draw the conclusion that the use of materials of a larger size (up to 5cm in length) is possible in the fluidized bed, as mixed wood chippings or pellets, provided that the material is not contaminated with longer materials. This is particularly of relevance when considering the preparation costs of the biomass.

## 6 Summary and outlook

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In the first test set-ups, orientation tests were carried out using variations of the main operating parameters gasification agent, load relationship (biomass mass flow in relation to the gasification agent quantity), type of input material and size and composition of the bed material. The main results are summarized in the list below:

- The maximum hydrogen yield was obtained with the use of pellets with steam and oxygen as the gasification agent.
- A gasification with steam and oxygen increases the hydrogen concentration significantly compared with a gasification with air; the distribution of the other products is similar, calculated without nitrogen and steam at nearly the same operational parameters.
- The methane content is nearly constant, regardless of the gasification agent.
- The use of oxygen as an operating resource can be reduced by electric heating, with an increase in the hydrogen content in the synthesis gas .
- By using a finer bed material (Nivelsteiner sand N45) high circulation rates and homogeneous temperature distribution in the reactor can be achieved at the start of the test. With a longer test time this bed material is increasingly transported out of the plant.
- Courser bed material (sand S40T) reduces the circulation rate.
- Pellets have a higher solid material throughput time compared with chippings and thus increase the hydrogen yield.
- The tar content is around 1 Ma.-% of the raw gas.
- Higher hydrocarbons including propane can be ignored (only determined for the steam/oxygen gasification).
- The concentrations of unsaturated hydrocarbons are higher than the saturated components (only determined for the steam/oxygen gasification).
- Particles with an equivalent diameter of 1 mm (mixed wood chippings) to 15 mm (chipboard pieces) can be used without difficulty.

The following investigations are planned:

- An increase of the hydrogen content, e.g. through variations of the bed material, an increase in the bed circulation rate with bed materials of suitable particle size distribution.
-

- Furthermore, the influence of the reactor temperature on the methane content should be investigated. For mechanical reasons an increase of the reactor temperature to above 960°C is not possible, therefore this influence can only be tested through a decrease in temperature, e.g. to 850 °C.
  - A determination of the fly ash quantities will be attempted.
  - The use of straw with added lime.
  - The use of sunflower husks and larger wood shavings/chippings.
  - The use of wet biomass (up to 40 Ma.-% moisture).
  - Investigation of tar build-up in dependence on the solid material and gas throughput times, temperature and type of fuel.
  - Investigation of trace components.
  - Test of the gas purification technology (efficiency of the individual systems).
  - Investigation of dust removal and cyclone efficiency.
  - The use of dolomite or donit as bed materials.
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## 7 Literature

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## 7.1 Formula symbols

### 7.1.1 Symbols (Latin)

$C$	carbon proportion	$\left[ \frac{\text{kgC}}{\text{kgBr}} \right]$
$c$	specific heat capacity	$\left[ \frac{\text{kJ}}{\text{kg} \cdot \text{K}} \right]$
$g$	acceleration due to gravity	$9,81 \frac{\text{m}}{\text{s}^2}$
$h$	heating value	$\left[ \frac{\text{kJ}}{\text{kg}} \right]$
	enthalpy	$[\text{kJ}]$
$\dot{H}$	enthalpy flow	$\left[ \frac{\text{kJ}}{\text{h}} \right]$
$\dot{m}$	mass flow	$\left[ \frac{\text{kg}}{\text{h}} \right]$
$p$	pressure	$[\text{Pa}]$
$\dot{Q}$	heat flow	$\left[ \frac{\text{kJ}}{\text{h}} \right]$

### 7.1.2 Symbols (Greek)

$\Delta$	difference	$[-]$
$\varepsilon$	porosity, relative porosity	$\left[ \frac{\text{m}_{\text{leer}}^3}{\text{m}_{\text{ges}}^3} \right]$
$\vartheta$	temperature	$[^{\circ}\text{C}]$
$\xi$	mass concentration	$\left[ \frac{\text{kg}_i}{\text{kg}_{\text{ges}}} \right]$
$\rho$	density	$\left[ \frac{\text{kg}}{\text{m}^3} \right]$
$\psi$	volume concentration	$\left[ \frac{\text{m}_i^3}{\text{m}_{\text{ges}}^3} \right]$

### 7.1.3 Indices (subscript)

$0$	start
$A$	output
$Ad$	additive
$C$	carbon

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CO	carbon monoxide
CO <sub>2</sub>	carbon dioxide
CH <sub>4</sub>	methane
C <sub>2</sub> H <sub>4</sub>	ethene
C <sub>2</sub> H <sub>6</sub>	ethane
C <sub>3</sub> H <sub>6</sub>	propene
C <sub>3</sub> H <sub>8</sub>	propane
C <sub>4</sub> H <sub>10</sub>	butane
C <sub>5</sub> H <sub>12</sub>	pentane
C <sub>6</sub> H <sub>14</sub>	hexane
D	steam
E	input
el	electric
ES	input material
F	fluid
N,N <sub>2</sub>	nitrogen
O,O <sub>2</sub>	oxygen
<i>p</i>	constant pressure ( <i>c<sub>p</sub></i> )
SG	synthesis gas
PG	permanent gas
roh	raw condition
S	solid
tr.	dry
<i>u</i>	lower (heating value <i>h<sub>u</sub></i> )
V	loss
W	water
WS	fluidized bed
zu	added

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