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RENEW

Renewable fuels for advanced powertrains

Integrated Project

Sustainable energy systems

Preliminary Final report

SP2 workpackage 2.2

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Subproject 2 WP 2.2 Gasification Step

Preliminary Final Report
SP2 WP 2.2

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1 Objectives

The objective of subproject 2 “Process optimisation” is directed towards the improvement of the biomass preparation, gasification, gas treatment and Fischer-Tropsch Synthesis steps in the process chain of producing Fischer-Tropsch Biofuels from lingo-cellulosic biomass for transportation.
The focus of work package 2.2 is on the optimization of the gasification step to convert biomass into a raw syngas.. The gasification routes via autothermal fluidised bed gasification at CUTEC and the two step process - fast pyrolysis followed by the pressurized entrained flow gasification - at FZK have been studied.

Whereas wood can be gasified easily the gasification of other kinds of biomass like straw and energy plants involves higher efforts due to higher ash content and other organic compounds and no reliable and durable gasification techniques are available at present. Gas cleaning depends on the preceding gasification step and is more complex and thus higher efforts are needed to clean the gas in order to avoid poisoning of the Fischer-Tropsch catalyst.

2. The two step pyrolysis – gasification process at FZK

Lignocellulose like wood or straw is the most abundant biomass type. Generation and use of synthesis gas from biomass involves complex technology, which can be more economically applied in large central plants, than in many small facilities. Since biomass is distributed over large areas, expensive long-distance transport of low bulk density materials like straw bales, wood chips etc. are required. The problem can be solved by biomass liquefaction. Biomass is first liquefied to pyrolysis oil by fast pyrolysis in many local facilities. The relatively small amount of brittle pyrolysis char is pulverised and suspended in the pyrolysis oil. Such pyrolysis oil/char-slurries have a high density of ~ 1300 kg/m^3 and contain up to 90% of the initial biomass energy. They are easily stored in tanks and can be transported e.g. by rail, to a large central plant for efficient and economic syngas generation and use. An oxygen-blown entrained flow gasifier operated at high pressure is particularly compatible with downstream synthesis of synfuel or chemicals. Slurries are easily pumped into the highly pressurised gasifier chamber and are pneumatically atomised with oxygen. The hot, high-pressure syngas is tar free, thus obviating expensive efforts for tar removal and syngas compression.

2.1 Fast pyrolysis of lignocellulosic biomass

Advantages of fast pyrolysis: Unique characteristic of the Karlsruhe BTL process is a pumpable, hot bioslurry feed for a highly pressurised entrained down-flow gasifier with a cooling screen, suited for high-ash materials. Slurries are prepared by suspending char powder in the liquid condensates obtained by pyrolysis of lignocellulosics like wood or straw. Only fast pyrolysis (FP) can generate sufficient liquid and sufficiently low char and gas yields to allow complete char suspension with condensate/char-mass ratios ≥ 2. Up to 90% of the initial bioenergy can thus be concentrated in the dense slurry. Biomass liquefaction and densification are favourable not only for use as gasifier feed, but also for storage and transport. Larger and therefore more economic BTL-plants can be supplied by cheap electro rail transport of slurries or pre-slurries from many regional FP plants up to ca. 500 km distance.

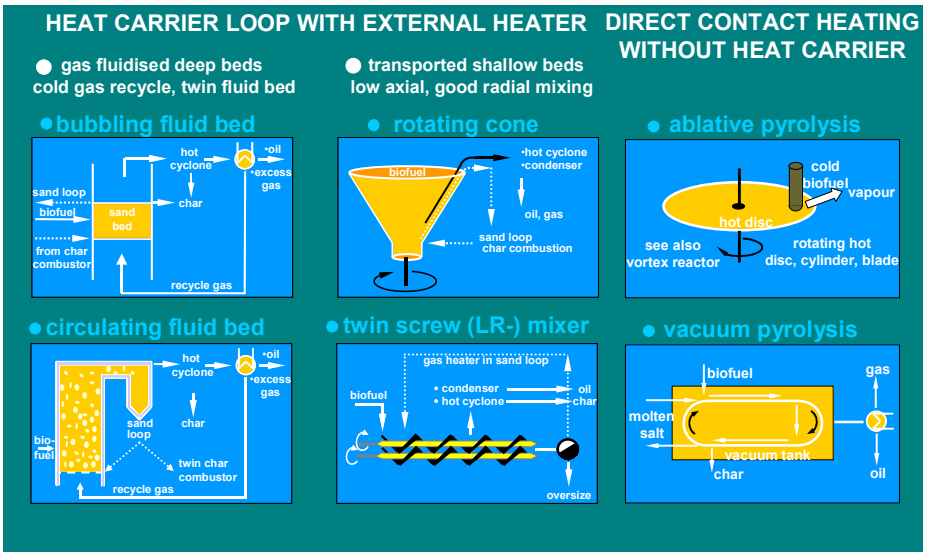

Fig. 1: Reactor types for fast pyrolysis

Choice of a FP reactor type: From the many types of FP reactors being developed or used for biomass or other materials, see fig. 1, we have selected the twin screw or LR mixer reactor mainly for two reasons: (1) A mechanically transported flat bed reactor with quick cross current product vapour removal are favourable design characteristics in accordance with a theoretical reaction analysis; (2) Large-scale technical experience with non-biomaterials coal, vacuum resids or oil shale has been accumulated since 50 years and is expected to reduce development time and cost for biomass:

- The twin screw mixer reactor combines low axial with good radial mixing of fine < 3 mm low-moisture biomass particles with a large amount of heat carrier grains at about 500°C. Pyrolysis product gas, vapours and char dust are blown out in a mixed product stream and are recovered successively: first the char dust in a hot cyclone then the condensates, leaving the gas. The heat carrier grains are separated and reheated from ca. 500°C to max. 600°C prior to recycle in a closed loop with heat exchanger (FZK) or an open (Lurgi) loop with pneumatic lift.

– Residence time (is 1 or few seconds) and thus reactor size and cost is kept short, especially the time of vapour contact with the heat carrier.

– The twin screw mixer is the only FP reactor type with experience obtained from large-scale technical applications.

Suitability of the twin screw mixer reactor for FP of lignocellulosic biomass: Within the RENEW project, the principal suitability of the LR-mixer reactor type has been tested in two different lab-scale facilities at 3-20 kg/h throughput with many different biofeedstocks: wheat straw, rice straw, softwood, hardwood, wheat bran, rapeseed press cake etcetera. The high condensate and low char and gas yields which have been obtained are shown in fig. 2; they are comparable with results reported for other suitable FP reactor, and qualify the LR-mixer also for the FP of biomass. Quality and yield of pyrolysis condensates from wood was usually good. Biofeedstocks with more ash like straw, resulted in poorer condensate quality and lower yields, with tendency to immediate separation into a heavy organic phase and an aqueous solution with much acetic acid - and sometimes plus an additional light upper layer phase. Such poor condensates are not well suited as liquid biofuel for burners, turbines or internal combustion engines. Poor condensate qualities not suited as motor or turbine fuels are still suited for slurry production and subsequent gasification. This new application as a gasifier feed might be a game change for FP of poor-quality biomass. Lower quality requirements can result in further FP process simplifications and corresponding investment cost reductions.

FP design variants with a twin screw mixer reactor: A major objective within the RENEW tasks are investigations of potential further FP- and LR-technology improvements, especially in view to reliability, availability, simplicity and cost. Beside minor LR-mixer reactor modifications, the variants differ mainly in the reactor periphery. Two different versions of the heat carrier loop are being tested: (1) The conventional **open-loop mode** with a pneumatic lift of 1 mm heat carrier sand (SiO_2, SiC etc.) with hot flue gas from pyrolysis gas combustion for reheating to 500-600°C. (2) A more flexible **closed-loop mode** with a hot bucket elevator and larger steel shot or ceramic grains as heat carrier, which are reheated in an additional large and expensive, but closed heat exchanger.

A **0.5 t/h pilot facility** with the **pneumatic lift open loop version** is being built (with substantial FNR / federal ministry of agriculture financial support) with the long technical experience of LURGI company, Frankfurt. The pilot plant is now almost ready for start-up. Pilot facility design has been supported by RENEW with an experimental campaign in LURGI's mini LR-plant in Frankfurt (results see fig. 2).

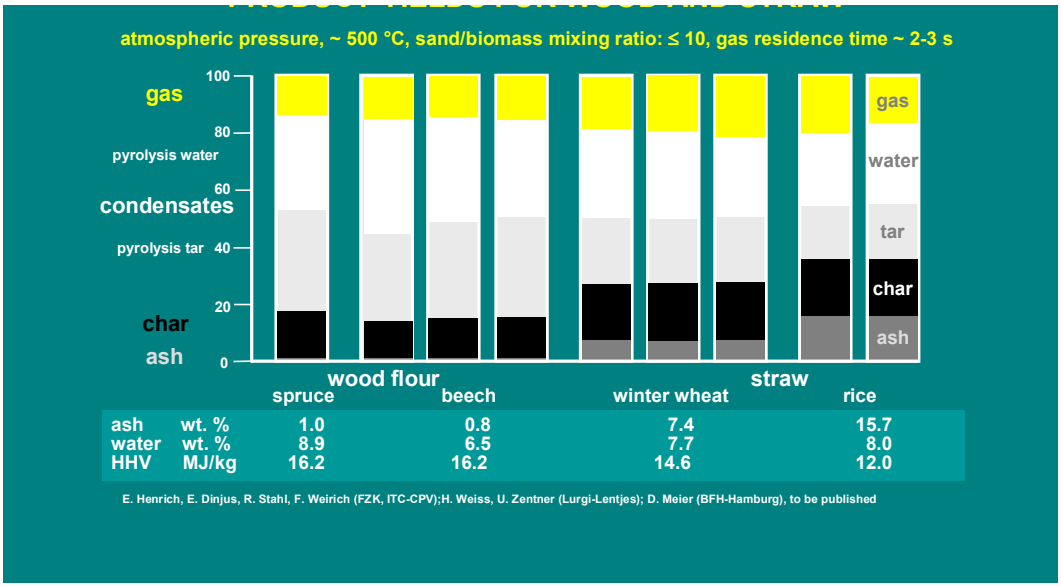

Fig. 2: Fast pyrolysis in the LR-mixer reactor, product yields for wood and straw

The **bucket elevator version** is still under investigation in the **Karlsruhe process development unit** (PDU). By replacing the initial 1 mm heat carrier sand by 1.5 mm steel shot, considerable improvements of operating reliability and plant availability has been achieved as well as about doubling of the throughput to almost 20 kg/h; the heat carrier circulation rate in the closed loop could be raised to > 1 t/h. Damage of downstream slurry preparation equipment e.g. like pumps by wear-out with attritted fine sand is now avoided. Corresponding changes in the available pilot plant would permit an experimental determination of technical design data also for the bucket elevator version.

Various **versions for pyrolysis product recovery** are also being tested in the Karlsruhe PDU. Many experiments have been conducted with a simultaneous common separation of char dust plus tar at 100°C in a single step. The condensing tar is soaked into the extended pore system of the char particles; solidification during cooling down thus does not create operating problems. The char, tar, water and gas yields for this FP product recovery mode are summarised in fig. 3. Pyrolysis chars are highly porous and tars which solify already at higher temperatures can be handled much easier in form of such solid char crumbs soaked with tar. An additional aqueous condensate was then recovered in a second condensation step downstream. These investigations are still continuing.

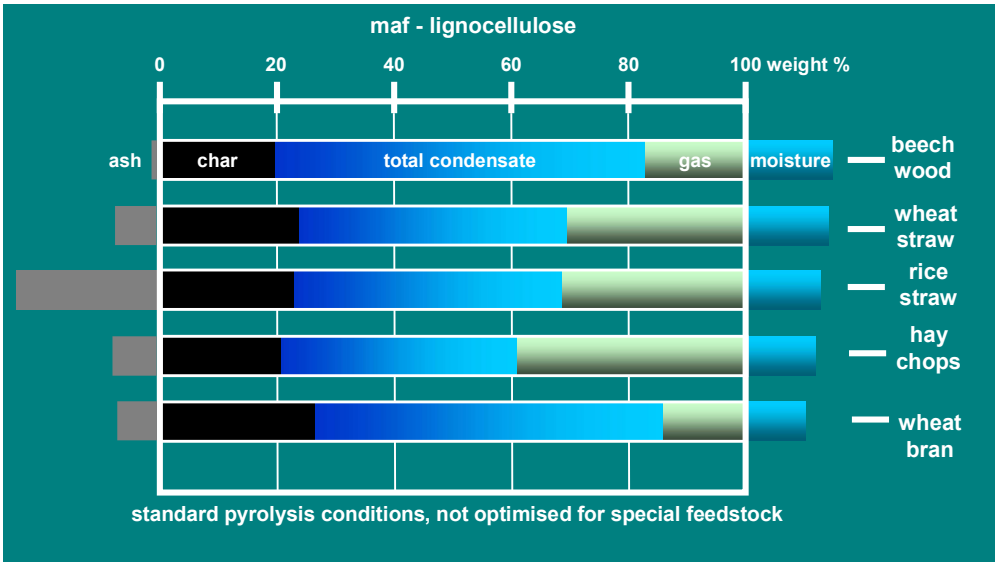

Fig. 3: Pyrolysis mass yields for various biofeedstock 5-15 kg/h twin screw mixer reactor with 1.5 mm steel balls as heat carrier

FP-modelling: Modelling means the prediction of products when educts, equipment design and operating conditions are known. Organic chemistry at typical FP temperatures around 500°C with a participation of solid reactants belongs to the most complex chemical processes. We have proposed and checked a rather simplistic model which just predicts the gas, tar, and water and char yields for a moisture and ash free ligno-cellulose part with < 2% wt ash on the one hand and > 2% ash content on the other hand. FP of low ash ligno-cellulosics like wood results in about 16% char and gas respectively and about 10 %wt reaction water; high-ash biomass like straw yields about 1.5 times more char, gas and reaction water with a correspondingly reduced condensate yield. More complex models do not necessarily result in better predictions.

2.2 Slurry preparation

One of the characteristics of the bio-slurry gasification of Forschungszentrum Karlsruhe is the combination of flash-pyrolysis and entrained flow gasification by using the mixture of the pyrolysis products as the feed for gasification. The so-called bio-slurry contains ~90% of the original biomass energy and should be appropriate for storage, transport, pumping and atomisation.

The dry pyrolysed biomass char is highly porous and light. It can absorb liquids until 80% in weight without becoming flowable. The decisive parameter for the mixing of flowable slurries is the empty volume of the dry bulk material, which is composed of the inner porosity of single particles and the outer porosity, i.e. the gap between single particles and agglomerates. The biomass chars examined here have a loose bulk density of 100-180 kg/m3 with 85-89% free space volume and a tap density of 160-270 kg/m3 with 78-84 % free space volume. The loose dry bulk char is a flowable powder and can easily be shoveled by a spoon.

Joggling or vibration results in a significant volume decrease and the material becomes impenetrable and inflexible.

For mixing a flowable slurry the solid volume / liquid volume should not be higher than the solid volume / gaseous volume of the loose bulk. However, first experiments with the experimental solid to liquid ratio of pyrolysis products of biomass lead to a wet clump. The situation can be drastically improved by reducing the void volume by deagglomeration. The necessary energy is transferred by hydrodynamic stress, provided by the shear rate of a fast rotating mixer. A related concept is the reduction of char porosity by wet or dry milling, which is an alternative way to achieve flowability within the experimental pyrolysis yields.

In the Karlsruhe bio-slurry gasification concept flash-pyrolysis is carried out by the twin-screw-LR-mixing reactor. While the straw is pyrolysed, the emerging brittle char is milled by abrasion from ~6 mm to ~600 µm. This is a positive side effect of the pyrolysis process, but still too large for a high carbon conversion by entrained flow gasification. In laboratory scale we succeeded in a particle size reduction up to 120 µm only by intensive mixing of the pyrolysis products. Also in laboratory scale we realised by addition of 3 % starch and wheat flour, respectively, a total suppression of sedimentation. A slurry of this kind should be the appropriate choice for long term storage, but the technical realisation in medium scale is a little laborious, as low temperature heat (80°C) is needed for gelification.

The Forschungszentrum Karlsruhe has been testing and improving the colloid mill technology from MAT Mischanlagentechnik GmbH, Germany, for the production of bio-slurries and verified its suitability. Extensive experience with colloid mixers of laboratory scale (500 ml, 2 l and 5 l batchwise) up to pilot scale (250 kg batchwise, 1000 kg/h continuous) has been gained. A rotor disk and perforated paddles are pressing the slurry via a small gap and exert high shear forces. The deagglomerating and size-reducing effect of the mixer increases with the slurry viscosity, but also the power which is dissipated into the slurry and, as a consequence, the temperature. The higher temperature and deagglomeration in turn lead to a decrease in viscosity. As a result, the dependence of particle properties, solid content, mixing power requirement, viscosity and slurry temperature is highly complex and needs a thorough assessment in each process design which is planned to be realised.

For mixing slurry with the actually installed 1000 kg/h continuous mixer special care has to be taken for the correct amount of solid which is dosed by a software controlled process balance and rotary-vane feeder. The ability to fall regularly into the rotary-vane feeder has to be verified (or improved) for each new feed material, because it is closely related to bulk density, angle of repose, particle size distribution and particle geometry, which are different for each pyrolysis process arrangement, biomass species, harvesting conditions etc. For a reliable slurry quality a downstream stirrer vessel is necessary for clearing concentration differences caused by systematic short-term dosing irregularities. Favourable are 1-2 more stirrer vessels for realisiation of a stirrer vessel cascade to equalise also medium-term differences in the feed composition (e.g. caused by changing the storage tank of the liquid phase). The gasifier can be controlled easier, if the slurry composition does not change

stepwise with feed changes, but gradually. So far, more than 40 t of bio-slurries were produced for the gasification experiments of Forschungszentrum Karlsruhe.

2.3 Entrained Flow Gasification

In the second step the resulting slurry is transported from many small pyrolysis plants to a large central highly pressurised entrained flow gasifier. There the slurry is used for the production of raw syngas. After gas cleaning the generated synthesis gas can be used for catalysed synthesis of chemicals or liquid fuels. Two experimental campaigns in the pressurised entrained flow pilot gasifier at Future Energy, Freiberg (formerly Noell company), proved the feasibility on a reliably large 3 MW(th) scale. The gasifier, fitted with a special inner cooling screen in a pressure resistant steel shell allows the gasification of different feedstocks with a large and fluctuating ash content and composition expected in the large spectrum of herbaceous biomass species. Compared to direct bio-oil combustion in diesel engines or turbines, the purity requirements for the gasification of bio-oil-slurry are very low. Pyrolysis oil (wood tar) and charcoal dust from beechwood pyrolysis for commercial charcoal production have been used for slurry preparation. Beechwood tar with a room temperature viscosity of 0.16 Pas, had a density of 1184 kg/m3 and a LHV of 19 MJ/kg. Charcoal was pulverised into different sizes between 10 and 100 um and had a LHV 31 MJ/kg. Stable slurries with different weight percent of charcoal dust have been prepared by mechanical mixing, without any stabiliser. To simulate the melting behaviour of straw slag, 3% straw ash and 0.3% KCl have been added. Slurry properties have been as follows: room temperature viscosity 2 – 5 Pas, density 1250+ kg/m3, LHV 21 – 22 MJ/kg. A slurry stream of ~0.35 – 0.6 t/h was transferred with a screw pump into the gasifier chamber at 25 bar and pneumatically atomised with pure oxygen in a special nozzle. Crude material and energy balances have been derived from the known feed composition. With an O2-stoichiometry of λ = 0.4 to 0.5, gasification temperatures of 1100 – 1600°C and more than 99% carbon conversion have been attained. A molten slag layer drains down on the inner screen walls and protects the SiC-liner from corrosion. A suitable slag viscosity is adjusted with the O2-flow (temperature) and – if desirable – suitable inorganic additives in the feed to modify the slag composition. As the hot pressurised syngas is tar-free expensive efforts for tar removal or syngas compression can be avoided. The measured raw syngas composition indicates an approximate equilibration of the homogeneous shift reaction CO + H2O ⇆ CO2 + H2, as expected. Two experimental results for the composition of the dry syngas are shown in the following table. The relatively high energy loss of the cooling screen and the use of flue gas for cleaning of unused piping at the pilot gasifier result in high amounts carbon dioxide and nitrogen. In a large technical gasifier the concentrations will be much lower. The expected syngas composition for a technical gasifier is given in the last line.

CO	H_2	CO_2	N_2	CH_4	H_2S	tar		
45	29	18	9	< 0.1	29 ppmv	–	volume%	$\lambda(O_2)$ = 0.46
47	27	16	9	< 0.1	17 ppmv	–	volume%	$\lambda(O_2)$ = 0.42
58	30	10	2	< 0.1	17 ppmv	–	volume%	$\lambda(O_2)$ = 0,3 -0,4 expected

2.4 Economic considerations

Energy efficiency and polygeneration: Biosynfuels produced via the biosyngas route must compete economically with the global annual production of 2 Gt conventional gasoline, diesel and jetfuel from crude oil. An oil refinery manufactures motor fuels for ca. 0.5 €/t (without tax) at the present crude oil price of about 60 $/bbl. Compared to a large modern > 10 GW oil refinery, biosynfuel plants are somewhat smaller, use more complex technology and are therefore about twice as expensive at the same capacity. An oil refinery converts about 80% of the crude oil energy into liquid motor fuels. Only about 40% of the initial bioenergy can be converted into biosynfuels, but the rest is not lost. This is because the inevitable reaction heats from the complex multistep chemical process can not be converted into main or side-products, but only in electricity, high-p steam and process heat. These energy side-products can be consumed directly in the process or exported with credit. We assume complete own consumption in the process plant without credit for energy export.

This includes use of electricity generated in the large central biosynfuel plant by distribution via the grid e.g. for rail-transport of slurries or biomass diminution in the decentralised pyrolysis plants etcetera. The inevitable polygeneration is thus used to create a self-sustained system supplied completely with bioenergy.

Plant configurations: Decisive characteristic of the Karlsruhe BTL process is a pumpable bioslurry feed for a highly pressurised entrained down-flow gasifier. Feed slurry can be prepared by fast pyrolysis of various biofeedstock either directly on-site of a large central gasifier or off-site. On-site truck delivery of bulky biomass like straw becomes too expensive at larger distance between harvest field and plant, the dense road traffic near a large plant becomes excessive, but also the diesel consumption – which must subtracted from the own production. Input capacities of integrated plants will thus be limited to about 1 Mt/y input of daf ligno-cellulose like wood or straw. The output of max. 0.2 Mt/a is too small in view of the global consumption.

Preparation of bioslurries or simple pasty or crumby slurry preforms in off-site regional pyrolysis plants followed by rail transport – with electricity consumption – of the compact densified pyrolysis product mix in silo wagons (for quick unloading), permits an economic supply of even huge central biosynfuel plants. Rail transport costs do not depend much on the distance in the 100 to 1000 km transport distance range and allow a flexible even "diluted" spot-like pyrolysis plant distribution over large areas (e.g. whole France or Germany). Transport cost estimates for German conditions are summarised in the diagram of fig. 4.

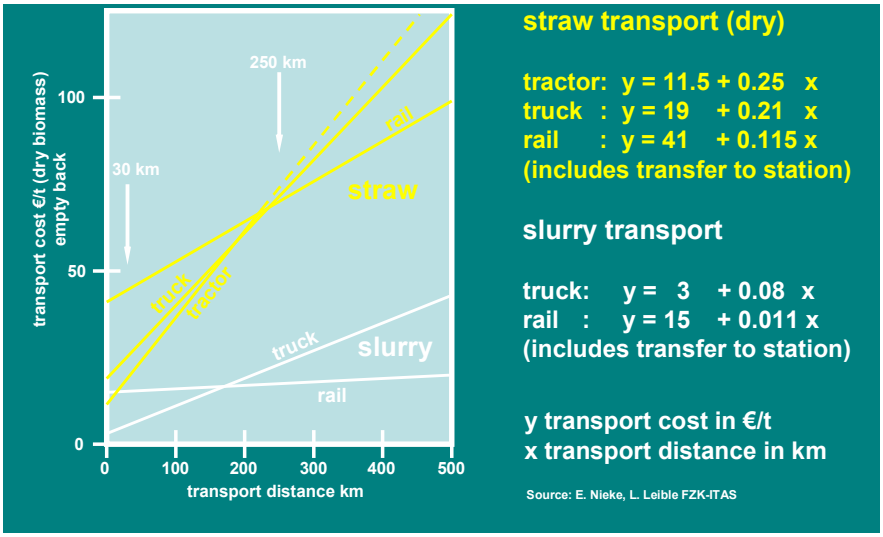

Fig. 4: Biomass transport costs

Large CTL and GTL plants with a synfuel output capacity up to the 1 Mt/y range are already in operation, even larger ones are planned or under construction for economic reasons e.g. in Qatar and Nigeria. The tail end section-after generation of a clean, conditioned syngas-in such plants is not different in a BTL plant; the principal technology is available, further development aims at potential further improvements. Even the detailed engineering of the tail-end steps in BTL, CTL and GTL plants will be the same at the same capacity and does not need engineering efforts for scale-down in case of smaller plants.

Only the head-end steps for syngas generation in the known CTL and GTL-plants need to be adjusted to biomass. The essential steps are preparation of a pumpable bioslurry feed by fast pyrolysis either after direct truck transport of biomass to the on-site pyrolysis plant or in off-site plants followed by rail transport of the densified bioslurry pre-forms. Different biofeedstocks are thus equalised by conversion to a pumpable bioslurry feed for a highly pressurised entrained flow gasifier.

Methodology: Our cost estimates are based on a relatively simple standard method described in textbooks: (1) Onken, Behr; "Chemische Prozesskunde", Thieme press, Stuttgart 1996, chap. 5 and (2) in more detail in Peters, Timmerhaus, West "Plant design and economics for chemical engineers", McGraw Hill 5th ed., mainly chap. 6. Costs are subdivided into only five different contributions: (1) Feedstock, (2) utilities, (3) capital and capital related, (4) personal and personal related, (5) varia.

(1) **Feed costs**, biomass without transport and technical oxygen, are almost independent from the plant size (cost degression exponent ~ 1). Biomass transport costs are composed of a constant value for loading and unloading plus a component about proportional to the distance, which is about the square root of the plant capacity (distance squared equals about area).

(2) **Utilities** (energy): To a very first crude approximation, the complex biosynfuel production is considered to be self-sufficient energetically. The inevitable energy side products in form of electricity, high-p steam and heat are assumed to be completely consumed in the total biosynfuel process; nothing is assumed to be left for export and credits.

(3) **Capital and capital related costs** for BTL-plants are assumed to scale with a degression exponent of 0.7; this halves the specific investment costs for a capacity increase of an order of magnitude.

(4) **Personal and personal related costs** are a low share in the large plants which are typical for energy supply; degression exponent for personal ~ 0.25. A realistic average salary including overhead cost is ~ 60 k€ per anno et capita.

(5) **Varia:** e.g. like product packing and transport, waste water and off-gas cleaning etcetera contribute little, usually < 10% of the sum of (1) to (4).

Separate estimates for the pyrolysis and the gasification / synthesis plants:

Fast pyrolysis: To profit from the economy of scale, also local or regional pyrolysis plant should be as large as reasonably possible. A maximum delivery radius with tractors of local farmers in central EU is 20 – 30 km. An estimated 45% percentage of surplus cereal straw plus forest residues especially by-products from the stem harvest in rural EU areas within 30 km radius corresponds to about 200 000 t/a airdry lignocellulose (LHV 4 kWh/kg). This is sufficient for a FP plant with 100 MW biomass input (ca. 20 t/h dry maf ligno-cellulose LHV ca. 5 kWh/kg).

Investment costs for FP plants have been reported in the literature and have been supplement by own estimates. The different types of FP reactors contribute only about 10% of the total FP investment, larger cost differences are therefore caused by the periphery and scale–up limitations, requiring parallel lines especially for a very large integrated central design. Investment cost reduction by learning can be expected from the beginning for the many decentral plants supplying a large central bio-synfuel plant. A very first crude estimate for a 100 MW FP plant located at an industrial site (no greenfield site) with rail access, including biomass storage and preparation, pyrolysis and product recovery and pre-slurry preparation for transport is ~ 20 M€. There is still technical potential for further cost reductions, e.g. by separation and use of few percent of valuable chemical constituents in the bio-oil, especially chiral compounds. Likely, the many relatively simple FP plants together will be more expensive than the large single biosynfuel plant they supply.

Gasification / synthesis: To design a large optimum – e.g. in view to energy recovery and use – BTL-plant requires several 100 person-years only for engineering. Even a simplified cost estimate on the basis of own engineering activities is far beyond our capabilities. Therefore our cost estimates are based on specific costs reported in the literature by large experienced companies like SASOL, SHELL, EXXONMOBILE etc. for their large planned GTL facilities. Capacity adjustment of specific investment costs with a degression exponent of 0.7 results in 20 – 25 000 US$ per bbl and day synfuel output equal to ca. 3000 t/d = 125 t/h · 8000 h/a = 1 Mt/a for a reference plant with 1 Mt/y synfuel output.

Different from methane gasification in GTL plants, bioslurries require more O_2, generate more slag and need a more efficient syngas cleaning. A BTL front end-plant section is therefore more expensive. Our cost estimate relies on the GTL-front end, which includes a cryogenic air separation unit (ASU); we add the "over the fence" delivery cost of an estimated 8 c€ per m^3 (STP) oxygen to the feed cost and keep the plant front end investment costs unchanged. A BTL front-end section thus becomes about 1.5 times more expensive then the corresponding GTL-plant section. The tail-end sections in BTL, GTL or CTL plants are identical in design and cost.

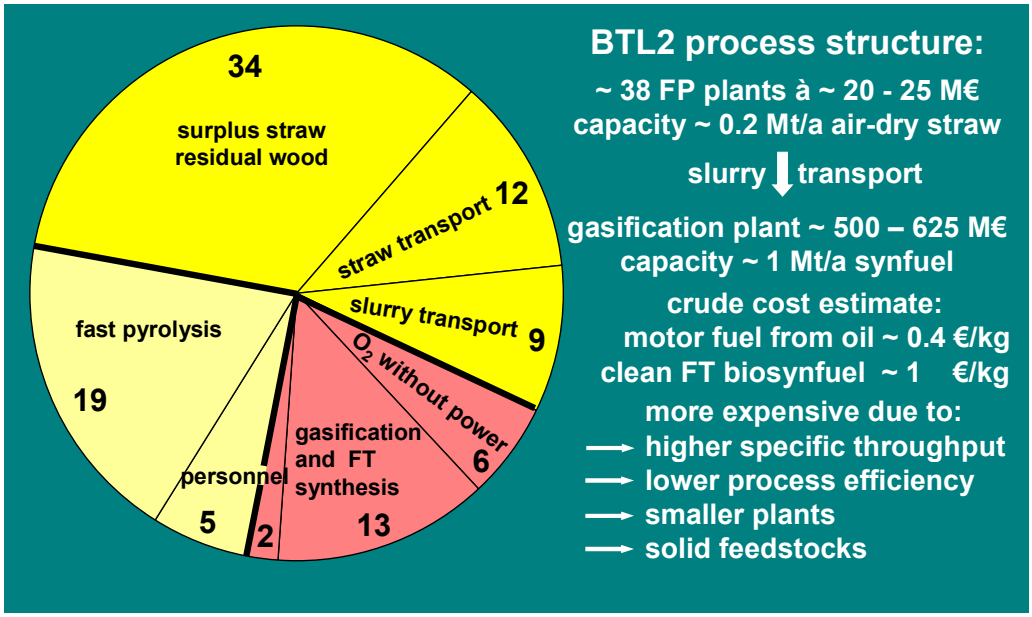

Fig. 5: Breakdown of biosynfuel production costs in %

Biosynfuel cost breakdown: Within the RENEW project, cost estimates for a biosynfuel production plant with 0.5 GW biomass input had to be presented. This is a violation of the Karlsruhe bioslurry concept, which has been designed for at least order of magnitude larger capacities, last not least also in view to coal as a likely future co-feedstock. About 0.2 Gt/a synfuel are considered to be too small for a significant market contribution and do not expose the unique cost advantages of the Karlsruhe bioslurry concept. The cost breakdown in fig. 5 is therefore given for a reasonably large 1 Mt/y biosynfuel production complex. For this capacity and central EU conditions, biofeedstock plus transport costs contribute about half and do not depend much on capacity. Pyrolysis contributes ca. a quarter as well as syngas generation and use including oxygen for the gasifier. The technology investment costs scale with a degression exponent of ca. 0.7. Absolute cost estimates for biosynfuel result in ca. 1 €/kg, for untaxed motor synfuel. At the present crude oil price of ca. 60 $/bbl the conventional motor fuel from a modern oil refinery is about 0.5 €/kg.

Conclusion: Plants for only biosynfuel production may turn out to use a too simple technology for the future. Biorefineries designed for flexible and complete conversion of different biofeedstocks into a versatile spectrum of useful products (polygeneration: organic chemicals and fuels plus electricity high pressure steam and heat) are a more desirable configuration of various local and central facilities. In this context, the Karlsruhe bioslurry process can be considered as the backbone of a large thermochemical biorefinery. Thermochemical biorefineries will be complemented by biochemical ones.

3. The ArtFuel process at CUTEC

3.1 Objectives of the ArtFuel-Process

The basic idea of the ArtFuel-Process was to create an operation with the ability of decomposition various biomass(es) for the Fischer-Tropsch Synthesis. The minerals of the plants should be collected in an unmelted form to have the chance of recycling them as fertilizer.

At CUTEC several facilities were built up:

1. **a** circulating fluidized bed (Abr.: CFB) with a power of 0,4 MW_{therm} (Fig. 1) for the gasification of various types of biomass.

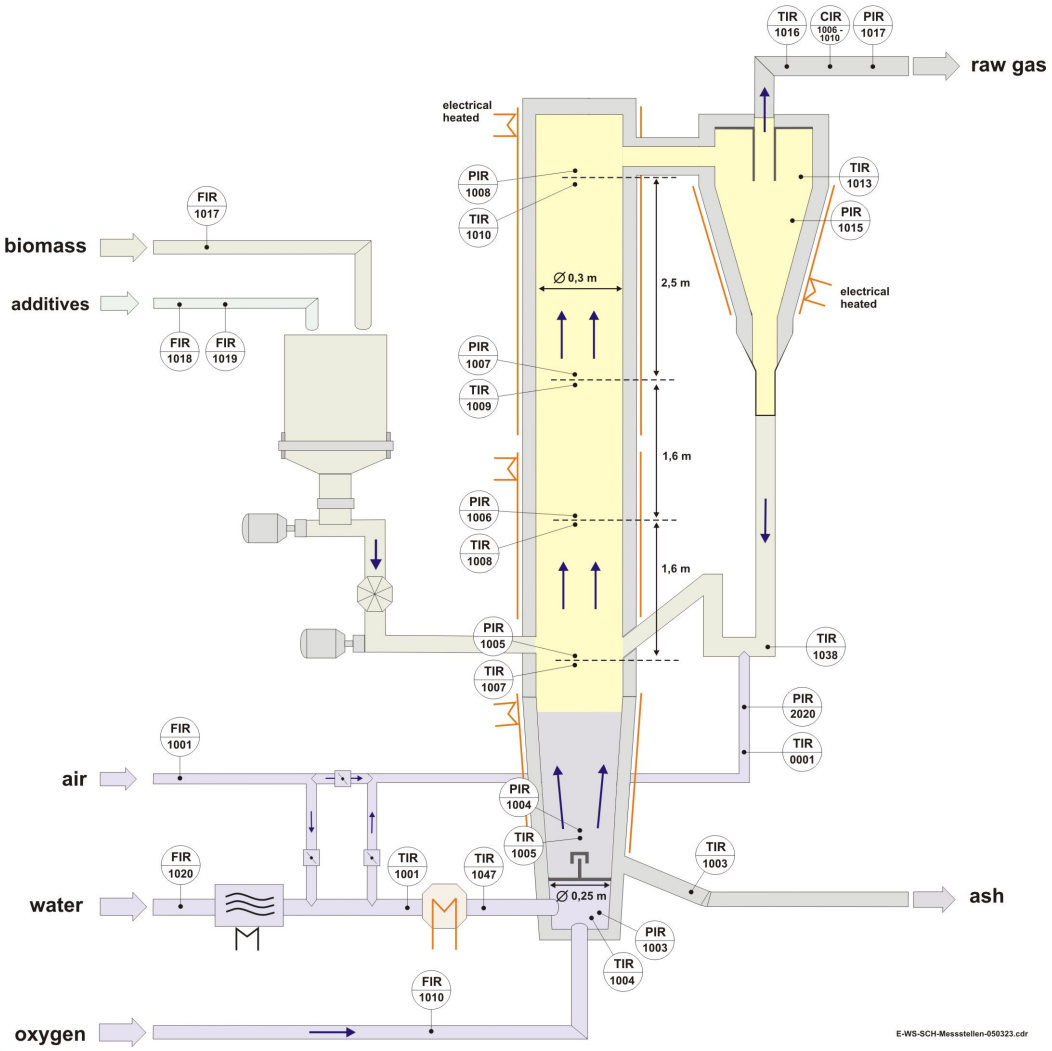

Figure 1: CFB-Reactor in the ArtFuel-Plant

A gas cleaning route (Fig. 2) consisting of ***2.*** hot gas filter (Abr.: HGF) with ceramic filter elements, ***3.*** quenching system, ***4.*** water scrubber ***5.*** biodiesel scrubber and **6.**

a filter with activated charcoal. Within RENEW results of the parts 1 and 2 are included in WP 2.2 whereas those of part 2 to 6 are subjected to WP 2.3.

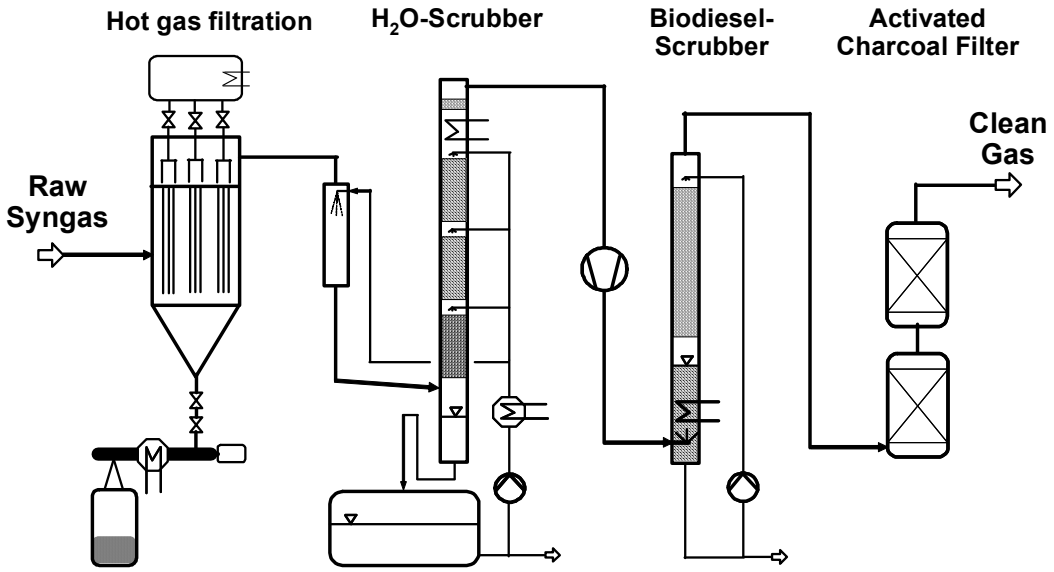

Figure 2: Gas-Cleaning-Route

The CFBs are state of the art in commercial plants for firing processes especially of waste sludge and municipal waste. They are able of working with a wide range of materials. For gasification and especially BtL-production the process parameters have to be adjusted to the specicial conditions.

3.2 Course

At the beginning of the project it was decided not to use an existing small fluidized bed reactor of 0,05 MW_{therm} but to complete the started work of building a big CFB as described above. So a delay in the time schedule happened. After starting of operation is was possible to overhaul the delay because the machine worked with a high availability and efficiency.

3.3 Experimental parameters for the gasification process

In the task list of WP 2.2 a lot of parameters for the gasification had to be examined:

1. shape and size of the particles,
2. kind of biomass,
3. process parameters temperature, steam flow, steam-oxygen ratio and bed material. (The pressure is fixed because the gasification system was designed for nearly athmospheric pressure.)

3.4 Results

The results have been described in detail in the following deliverables:

Del. 2.2.1: *Report on experimental results and operating conditions regarding the variation of biomass* in ***May 2005.***

Del. 2.2.7: *Variation of bed materials and their influence onto the composition of the raw synthesis gas* in **June 2007.**

Del. 2.2.12: *Composition of raw synthesis gas as function of different feedstocks* in **July 2006.**

Del. 2.2.14: *Closed Circles: Biomass ash as raw material for mineral manure production* in **August 2007.**

Despite the fact that the paper form of some deliverables will be published in the future the results of the experiments can be summarized:

a. Form of the input

Tested were shavings, chips, waste wood, pellets and sun flower husks (Fig. 3). With the technical plant the highest efficiency regarding the yield of synthesis gas was archieved with pellets if the same material (the experiments to this point were made with wood) is used.

Figure 3: Forms of input since 2004

b. Kind of biomass

Within RENEW the follwing types of biomass have been studied: wood, wheat straw, rape shred, sugar beet residue and sunflower husks.

Under the same process parameters the quality of the synthesis gas depends very much on the kind of biomass (Fig. 4). But a conclusion “what is the best biomass” is not possible, because

for every biomass an optimum of process parameters regarding steam flow, steam-oxygen ratio and process temperature can be found.

Remarkable is the fact that high H_2 : CO - ratios are possible without an extra shift reactor.

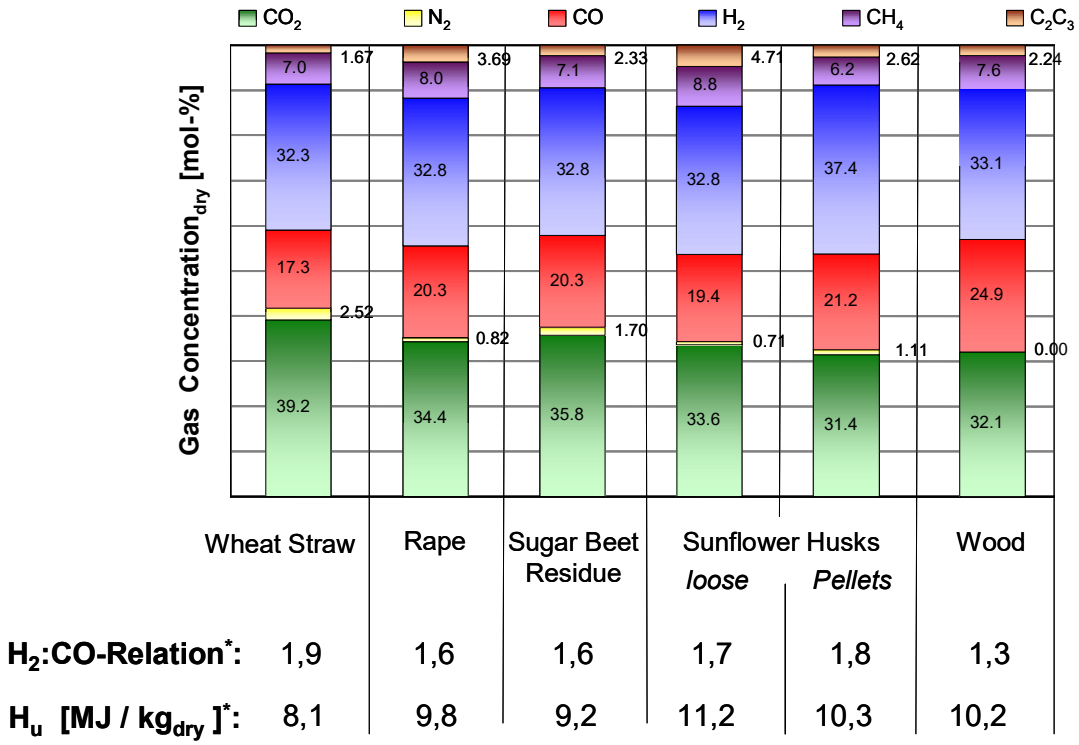

Fig. 4: Gas composition for different biomasses and similar process parameters

c. Process parameters

The operation parameters influence different characteristics of the raw synthesis gas. In general it could be observed:

a) increasing of **steam flow** leads to increased H_2 : CO -ratios

b) raising of the process **temperature** can increase the H_2 : CO -ratio. The effect depends from the kind of biomass and the temperature level (Fig. 5).

c) the variation of **steam flow - oxygen ratio** is included in a) and b). The possibilty to increase the steam flow is limited to the fact, that the duration time in the reactor should be at least 2 sec; better are 3 seconds.

d) testing of different **bed materials** revealed the highest yield of H_2 with lime and dolomit. Sand, alumina and olivin had significant lower values (Fig. 6).

With optimised process parameters the tar concentration in the raw synthesis gas is under 10 $g/m_{is,\ dry}$.

The H_2:CO -ratio of 1,75 as ideal value for an iron based FTS-KAT is easy to get without any shift reactor. A value of 2,1 as ideal for an cobalt based FTS-KAT demands a fine tuning of the process parameters but it is possible.

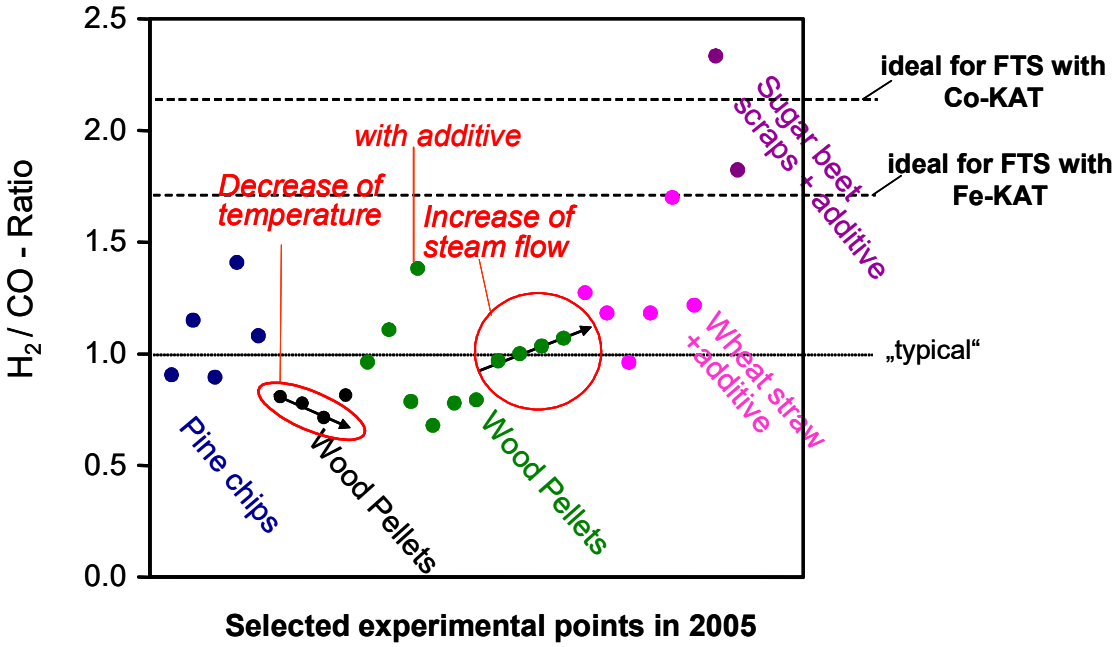

Fig. 5: Variation of process parameters

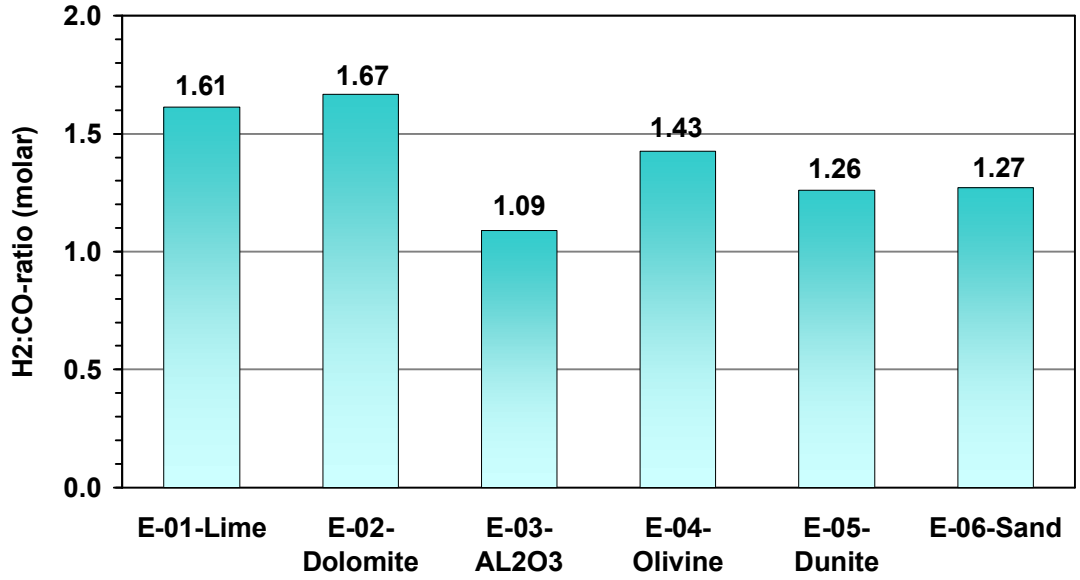

Fig. 6: Influence of the bed material

d. Filter ash of HGF

REM photos reveal that the ash is not melted. The particles are very fine in the µm-range (Fig. 7). Leaching tests after the method of DEV-S4 gave high concentrations of alkali and earth alkali metal in the water. Also the elements chlorine and sulphur were easy to dissolve.

Unfortunately the element phosphor is fixed in the ash under the conditions of the testing method. Some investigations to improve the situation have to be made.

Point X1

Mg	Ca	K	Na	C	Si
[Atom-%]					
0,3	1,5	0,6	0,1	93,8	3,7

Point X2

Mg	Ca	K	Na	C	Si
[Atom-%]					
0,3	0,7	1,1	0,1	73,9	23,9

Point X3

Mg	Ca	K	Na	C	Si
[Atom-%]					
2,4	16,6	8,0	0,4	0	72,6

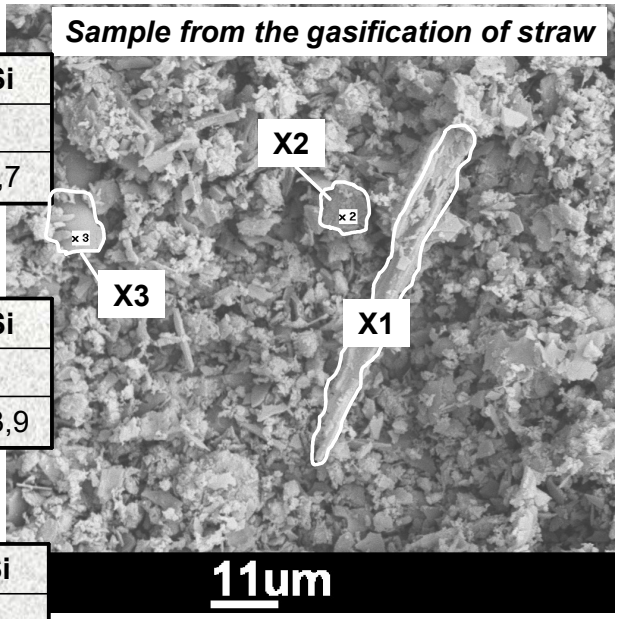

Fig. 7: REM-photo and EDX-analysis of HGF-ash

3.5 Conclusion

The obtained results at the gasifier and the HGF were more promising than expected. Especially the qualification of the CFB for the wide range of biomasses, the high H_2:CO-ratios and the low tar content offer the chance to commercialize the ArtFuel process in the future.