



Efficient use of biomass residues for combined production of transport fuels and heat

Esa Kurkela | Minna Kurkela | Sanna Tuomi |
Christian Frilund | Ilkka Hiltunen

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Esa Kurkela, Minna Kurkela, Sanna Tuomi, Christian
Frilund & Ilkka Hiltunen

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Teknologian tutkimuskeskus VTT Oy

PL 1000 (Tekniikantie 4 A, Espoo)

02044 VTT

Puh. 020 722 111, faksi 020 722 7001

<https://www.vtt.fi>

Teknologiska forskningscentralen VTT Ab

PB 1000 (Teknikvägen 4 A, Esbo)

FI-02044 VTT

Tfn +358 20 722 111, telefax +358 20 722 7001

<https://www.vttresearch.se>

VTT Technical Research Centre of Finland Ltd

P.O. Box 1000 (Tekniikantie 4 A, Espoo)

FI-02044 VTT, Finland

Tel. +358 20 722 111, fax +358 20 722 7001

<https://www.vttresearch.com>

Preface

This report summarises the research results of the BTL2030 project. The target of this project was to develop a medium-scale biomass conversion technology (corresponding to 100-150 MW biomass input), which can be integrated to various energy-intensive industries and district heating power plants. The main focus of the project was in the production of Fischer-Tropsch liquids, which can be further processed to high-quality transport fuels at large central refineries. In addition, concepts producing renewable synthetic natural gas (SNG) and hydrogen were designed and evaluated in the project.

The central activity of the project was the development and testing of the new gasification and gas cleaning process, which was implemented at the Bioruukki Piloting Centre of VTT Finland during years 2016-2018. The process was based on VTT's low-pressure, low-temperature steam gasification technology, simplified gas purification and small-scale industrial synthesis. In addition to the experimental development and testing activities, the BTL2030 project included design and feasibility studies for the combined production of FT liquids or synthetic natural gas and heat at plants integrated to district heating power plants or forest industry sites. This report summarises the studies on the effects of operating conditions of the gasification process on the efficiencies of FT liquid production. Detailed results of the techno-economic assessment work are presented in separate publications.

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Espoo, March 2019
Authors



Project website: www.vtt.fi/sites/BTL2030/en

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Abstract

List of symbols

BFB	Bubbling Fluidized-Bed
BtL	Biomass to Liquid
CapEx	Capital Expenditures
CFB	Circulating Fluidized-Bed
CCU	Carbon Capture and Utilization
DFB	Dual Fluidized-Bed
Dol + S	Dolomite + Sand
ELPI	Electrical Low Pressure Impactor
FT	Fischer-Tropsch
HVO	Hydrotreated Vegetable Oil
KOSEK	Kokkolanseudun Kehitys Oy
LHV	Lower Heating Value
nd	Not determined
PAC	Polycyclic Aromatic Compounds
P2G	Power to Gas
P2L	Power to Liquids
PSA	Pressure Swing Adsorption
SNG	Synthetic Natural Gas
STP	Standard Temperature and Pressure
TEM	Trade and Economics
toe	Tonnes of oil equivalent
WGS	Homogeneous Water-Gas Shift Reaction

1. Introduction

The EU Low Carbon Roadmap [1] requires decarbonization of all economic sectors, including transport. The targets for the transport sector are some 30 % reduction of greenhouse gases by 2030 and 60 % by 2050. To reach these goals, both improvements in efficiency as well as introduction of low-carbon energy to transport are needed. These low-carbon energy options include advanced liquid and gaseous biofuels, low-carbon electricity and hydrogen as well as so-called CCU-fuels from low-carbon electricity (Power to Gas - P2G and Power to Liquids - P2L). Some of these alternative energy carriers require new refuelling infrastructure and vehicles, whereas liquid drop-in fuels can be used directly in current vehicles without modifications.

As biomass is a limited resource in Europe, priority should be given to solutions that have the highest resource efficiencies, including well-to-wheel efficiency in transport and overall efficiency of biomass utilization for fuels, by-product energy and added value products. Gasification in combination with synthesis technologies represents a flexible production pathway, which not only can deliver fuels for road transport, but also renewable aviation kerosene and biobased chemicals and plastics.

Biomass gasification combined with Fischer-Tropsch synthesis is recognized as an attractive option for producing liquid transportation fuels from renewable feedstocks. In the past few decades, R&D activities in the field of gasification were largely focused on the development oxygen-blown gasification technology for large-scale production of liquid transportation fuels [2], [3]. These complex multi-step processes have high investment costs (> 500 M€) and also require large scale (> 300 MW fuel input) and heat integration with an energy-consuming process to become feasible - thereby markedly hindering the commercialization of such technologies [4].

In the attempt to accelerate the penetration of sustainably produced biofuels in the fuel market, VTT has developed a BTL concept that has substantially lower investment costs (200-300 M€) and is suited for intermediate scale (100-150 MW fuel input) corresponding to 30-50 ktoe/a production of transportation liquids. The proposed concept combines a steam-blown dual fluidised-bed gasifier operated at close to atmospheric pressure with a simplified gas clean-up train. The production concept is based on double integration benefits, as illustrated in Figure 1. Biomass

conversion to FT product (or to methanol or synthetic natural gas SNG) is realized in medium-scale units that are located close to biomass sources and heat integrated to industrial sites or district heating networks. Final refining of FT wax into high-quality transportation liquids, primarily biodiesel, takes place in existing oil refineries, where the economies of scale can be fully exploited and the product portfolio can be tailored according to the market needs.

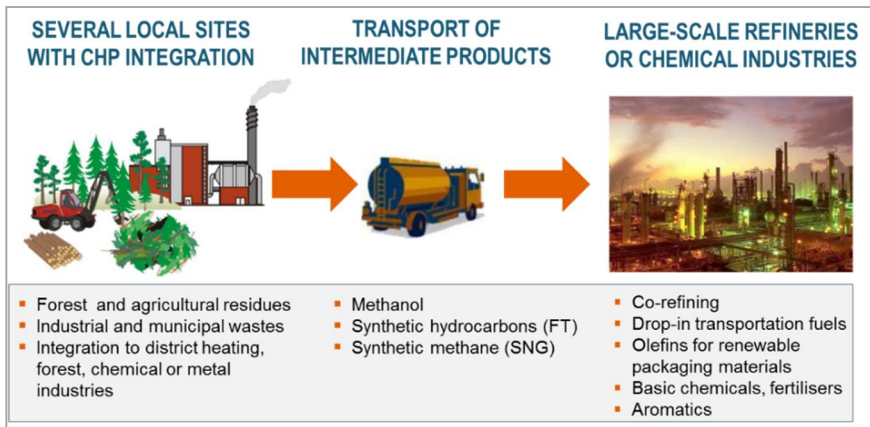


Figure 1. Integrating production of fuels and chemicals from biomass and residues to existing industries to improve competitiveness.

2. Objectives and tasks of the BTL2030 project

The aim of the project was to develop a new gasification process for combined production of transportation fuels and heat and to demonstrate its technical feasibility with various biomass feedstocks in pilot-scale tests. In addition, techno-economic studies were carried out to examine the economic feasibility of the process in different integration cases. Finally, preliminary ideas and plans for follow-on industrial projects were created in co-operation with the participating companies.

The project targets were set as follows:

- The operation of the Dual Fluidized-Bed (DFB) gasifier with various biomass feedstocks has been demonstrated and the measured performance data is used to estimate the performance of an industrial plant.
- Hot gas filtration, catalytic reforming and sulphur removal has been developed and validated in pilot-scale test runs.
- Production concepts for FT, H₂ and SNG have been developed and evaluated in different cases of energy integration. The economic feasibility and the potential of this biofuel production route have been reliably estimated.
- Proposal(s) for industrial follow-on projects have been prepared and preliminary plans for their financing have been made.

The BTL2030 project was divided into five work packages:

- WP1: Pilot-scale development of the new gasification process
- WP2: Feasibility studies
- WP3: Planning of follow-on projects
- WP4: International co-operation and review of competing technologies
- WP5: Reporting, publications, seminars and information exchange

Approximately 80 % of the project resources were used for the experimental development activities of WP1, where the gasification and gas cleaning process was developed and tested at the DFB gasification pilot plant of VTT. WP2 focused on preliminary concept design and evaluation of industrial-scale production of FT liquids, synthetic natural gas and hydrogen. In WP3, different alternatives for follow-on industrial demonstration were preliminary planned in collaboration with the industrial project partners. In WP5, previous activities on DFB gasification were reviewed and the status and challenges of plasma gasification technology assessed. In addition, the project participated in Finnish and European activities focused on renewable transport fuels.

3. Principle of the production concept

The target of this BTL2030 project was to develop a medium-scale BTL concept, which can be integrated to various energy-intensive industries and district heating power plants. The target plant produces main products (FT wax, hydrogen or SNG) with 50-70 % efficiency and by-product heat so that the overall biomass conversion efficiency is close to 80 %. The target plant size corresponds to 50-150 MW biomass input. Ideal sites for the integrated gasification-FT plants are district heating power plants, larger mechanical forest industries, pulping industries, food industries and even chemical and metal industries, which have a need for process steam.

Examples of promising integration options include:

- Integration to district heating power plants offer steam cycle integration and a good sink for the by-product heat. Biomass drying can also be realised at least partly by using low-temperature heat, which cannot be used for district heating. There are several suitable sites for the whole capacity range of 10-50 kt/a. Required heat loads are correspondingly 10-40 MW.
- Integration to pulping plants offer many potential benefits for relatively large FT units in the scale of 30-50 kt/a. Typically around 150 MW of bark is available at modern pulping plants with 1 Mt/a pulp production capacity. In addition, forest residues are available in the same region as the wood raw material. FT off-gas can be used to replace oil in limekilns and there is plenty of waste heat available from the plant that can be utilized for drying. Power plants with steam cycle also exist.
- Integration to smaller bioeconomy industries, such as mechanical wood industry and food industry, is an interesting alternative for smaller production units. Steam demand of 5-15 MW can be found e.g. from sawmills, particleboard mills, dairies, breweries, olive oil plants etc.
- A large chemical and metal industry complex is also a potential site where good energy integration exists. In a previous project, VTT made a conceptual design of a gasification-FT plant integrated to Kokkola industrial park, where there is a large zinc plant and production of fertilisers and other chemicals. This concept was granted with an innovation award in the Biorefinery contest organised by the Finnish Government.

The main idea of the developed concept is illustrated in Figure 2, which shows the basic process designed for the production of FT liquids. In this BTL concept, biomass residues are gasified in a dual fluidized-bed gasifier operated at low temperature. Raw product gas is filtered at gasifier exit temperature, tars and hydrocarbon gases are catalytically reformed and the reformed gas is purified from sulfur compounds using sorbents. Clean syngas is used in synthesis processes, which are designed to the target scale of 30-50 kt/a. Synthesis off-gases and by-

product heat are used to cover the on-site demand and surplus is sold to surrounding industrial co-generation or district heating power plants.

The main attributes of the process are:

- Indirectly-heated atmospheric-pressure steam gasification
- Catalytic reformer operated with air or oxygen
- No need for a separate shift reactor
- Sulphur control with adsorbents
- Compression to 26 bar in one step
- No CO₂ removal or minor removal with pressurized water scrubber
- FT synthesis with a small recycle ratio or operated as once-through
- The majority of synthesis off-gases combusted in the oxidizer to provide heat for gasification

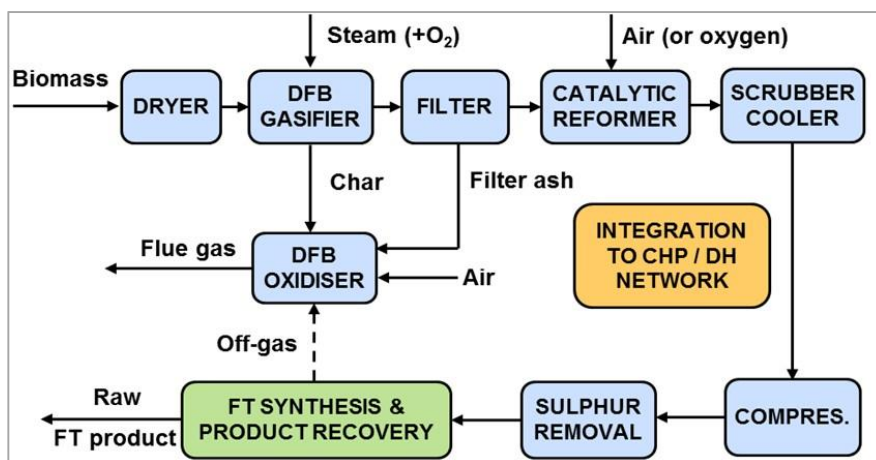


Figure 2. Medium-scale Low CapEx process for combined FT liquids and heat production.

The main advantages of this production concept over earlier proposed gasification-FT concepts are the following:

- Pressurized O₂-blown gasifier is replaced by a steam gasifier operated at low pressure and at a lower temperature of 720 – 820 °C, and in spite of this, all biomass carbon is effectively utilized.
- Low content of heavy tars is reached already in the gasifier and in the filter cake due to the catalyzing effect of calcium-based bed material on steam gasification and reforming of tars.
- Fuel feeding into a low-pressure gasifier is easier, which is important especially in the case of straw and other low-bulk density biomass feedstocks.

- The gasifier design is similar to the air-blown CFB gasifier – existing CFB gasifier can even be converted into a DFB gasifier (which may be an interesting alternative as the first industrial unit).
- The H₂/CO molar ratio of syngas after the reformer can be adjusted in the range 1.8-3 without an additional shift conversion unit.
- The capital-intensive and complex Rectisol-type of wet acid gas removal is replaced by simpler gas cleaning system based on adsorbents.
- Highly effective and compact synthesis technology has a high CO conversion with a small recycle ratio. This results in high biomass-to-FT-wax efficiency but it also allows 10-30 % inert gases in the product gas (CO₂+N₂+CH₄), which simplifies the required gas cleaning system.
- Similar idea of simplified production concept can also be designed for hydrogen using PSA technology for hydrogen separation.

4. Development of the gasification process

The experimental activities of the project focused on the front-end gasification and gas cleaning technologies of the BTL process, while the techno-economic evaluations included the complete process chain from biomass drying up to the production of renewable fuel components and heat.

The key elements of the developed gasification process include:

- Gasification of biomass residues in a DFB gasifier
- Hot filtration of raw product gas
- Catalytic reforming of tars and light hydrocarbon gases
- Sorbent-based sulphur removal and ammonia scrubbing

The first three processing steps were developed and tested at the full flow of the DFB pilot plant, while sulphur removal was studied using a small slip-stream test facility.

4.1 Gasifier development

Background know-how for the gasifier development was created already in previous projects of VTT. In 2005-2006, bench-scale steam gasification experiments were carried out as part of the process studies of the UCG project [5], [6]. These studies indicated that rather high carbon conversion efficiencies of the order of 75-85 % could be achieved already at low gasification temperatures. At that time, however, the main target of the development was large-scale BTL technology, and thus the activities of 2007-2011 focused on pressurized steam-oxygen gasification (see [2] - [6]).

In 2012-2013, when it became evident, that large-scale BTL plants will not be realised in the near future, VTT re-evaluated the smaller scale concepts and decided to construct a new DFB (Dual Fluidized-Bed) gasification pilot plant. The construction and commissioning work was completed in 2016, and the pilot plant (Figure 4) was ready for the needs of the BTL2030 project.

Two different reactor designs were tested. In 2016-2017, the DFB process consisted of a CFB (Circulating Fluidized-Bed) gasifier reactor and a BFB (Bubbling Fluidized-Bed) oxidiser. This initial design was concluded to be suitable for smaller plant sizes (50-100 MW) and especially for cases where part of the heat required in the gasifier was provided by mixing a small amount of oxygen with the gasification steam. At larger plant sizes and in steam-alone operation, the estimated diameter of the industrial-scale BFB oxidiser would be unrealistically large. Consequently, at the end of 2017, the BFB oxidizer was replaced by a CFB reactor, which had a smaller diameter but the reactor tube was higher. In 2018, this version of the DFB process (realized with two CFB reactors) was successfully tested.

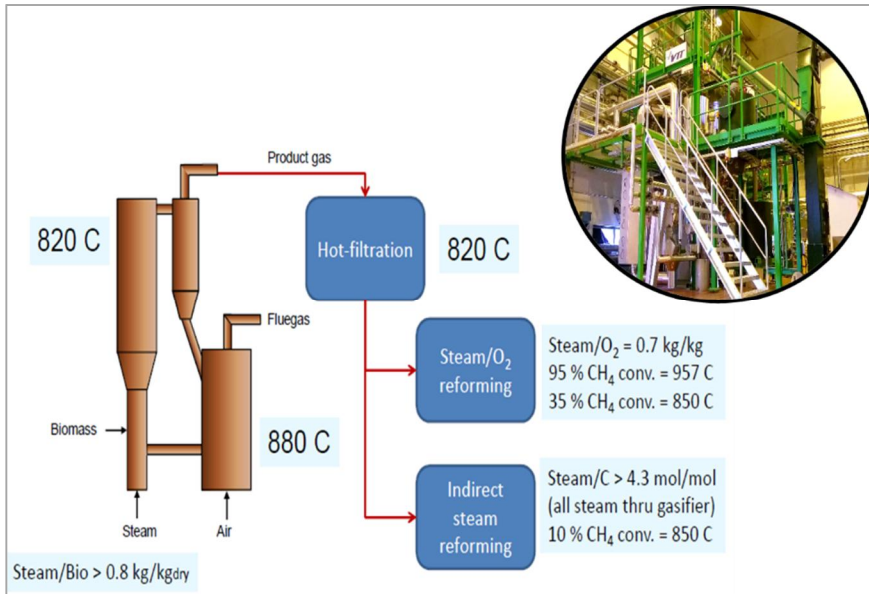


Figure 3. The principle of VTT's Dual Fluidized-Bed (DFB) steam gasification pilot plant at Bioruukki in the early test runs of 2016.

4.1.1 CFB gasifier and BFB oxidizer tested in 2016-17

The initial design of the 200 kW_{th} DFB gasification pilot plant is illustrated in Figure 4, and the key dimensions and design data of the gasifier and the oxidizer are given in Table 1. Biomass feedstock was fed into the CFB gasifier reactor at ca. 2 m above the distributor plate. Fluidization gases were preheated and introduced to the reactor through the inclined distributor plate. A mixture of dolomite and sand was used as bed material. Fresh bed material (make-up) could be fed both to the gasifier and the oxidiser, and bottom ash was removed from both reactors. In the gasifier, fine-grained biomass particles were mixed with the uprising flow of recycling bed material and consequently pyrolysed in the upper part of the gasifier. Larger biomass particles, on the other hand, first fell down into the bottom bed and were elutriated once pyrolysis and some gasification reactions had taken place. The charcoal particles and recycling bed material particles were separated from the raw gas in a uniflow-type cyclone and the separated solids were fed into the lower part of the BFB oxidizer.

The BFB oxidizer was fluidized with air and the heat released from charcoal combustion was used to heat the recycling bed material to 30-120 °C above the gasifier temperature. At this small pilot plant, it was also possible to provide additional heat by electric heaters, which could be controlled, either so that the plant

was approaching adiabatic operation (only heat losses were compensated), or that part of the heat required in gasification was provided externally by the heaters. The second operation mode simulated plant designs, where the oxidizer was partly fuelled by synthesis off-gases. The targeted operation mode could be approached by controlling the set points of the electrical heaters. The actual realized mode of heating at different operating periods was calculated afterwards from the material and energy balances.

The heated bed material left the oxidizer from an opening located at the upper part of the bubbling bed. The recycling lines were realised with fluidized L-valves. Small amounts of air or nitrogen were used to support the recycle flow.

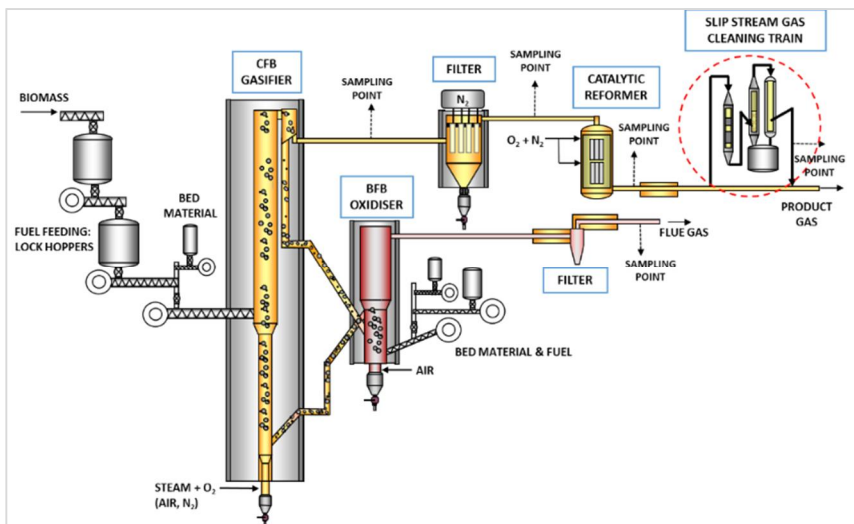


Figure 4. DFB pilot (CFB +BFB) at VTT's Piloting Centre Bioruukki in 2016-2017.

Table 1. Reactor dimensions and main operation conditions of the DFB gasifier used in the experiments of 2016 - 2017.

	CFB gasifier	BFB oxidizer
Reactor i.d. (mm)		
Lower part	102	320
Upper part	320	420
Reactor height (m)		
Total height	8.2	2.5
Height from distributor to gas outlet pipe	7.9	2.1
Height of the lower part	1.8	1.0
Fuel feeding point - distance from the gas distributor plate	2.3	-
Temperature range (°C)		
Lower bed section	750 - 850	840 - 920
Freeboard	750 - 830	850 - 900
Gas velocities, m/s		
Fluidizing velocity at the bottom of bed	4 - 5	0.3 - 0.4
Gas velocity at the top of reactor	4 - 5.3	0.25 - 0.3

The test programme of 2016-2017 is summarized in Table 2. After the test of 2016, some modifications were made to the test facility. The slip-stream gas cleaning train was constructed and taken into operation in test campaigns carried out in 2017. In the flue gas line of the oxidizer, the original cyclone cleaning was replaced by a filter, and flue gas cooling was also improved to reach filtration temperature of ca. 300 °C. This was necessary in order to be able to control the pressure of the oxidizer by using a control valve in the flue gas line. In the initial design realized with only cyclone cleaning, the operation of the valve was disturbed by ash deposits especially in the straw gasification tests, where the ash content of flue gas was higher and the ash was more sticky than in the case of wood gasification.

Table 2. Test runs carried out in 2016-2017 with the CFB gasifier and BFB oxidiser.

	2016	2017
Test runs	Two test weeks 16/20 and 16/24	Three test weeks 17/07, 17/10 and 17/17
Total gasification hours	166	200
Number of set points	12	16
Feedstocks	Crushed wood pellets Forest residues Bark Crushed straw pellets	Crushed wood pellets Bark Used wood Crushed straw pellets
Gasification agents	Steam Steam + O ₂ Steam + air	Steam Steam + O ₂ Steam + air
Gas cleaning	Filtration at 620 - 750 °C Catalytic reforming Lab-scale R&D on desulphurization	Filtration at 670 - 720 °C Catalytic reforming Sulphur removal by ZnO based sorbents in a slip stream

The operation experiences of the DFB process realized with the CFB gasifier and the BFB oxidizer can be summarized as follows:

- It was easy to maintain stable bed inventories in both reactors and the recycling loops worked fluently and without problems. The pressures measured from the top of both reactors were kept relatively close to each other by a control valve located in the flue gas line of the oxidizer. When this pressure difference was smaller than $\pm 20\text{-}30$ mbar, there were no signs of major gas leakages from the gasifier to the oxidizer or vice versa.
- The massive bed reservoir of the BFB oxidizer stabilized the recycling during normal operations as well as in the case of process disturbances. The gasifier bed inventory seemed to stabilize rapidly even after short process disturbances.
- The temperature and the fluidization velocity of the oxidizer could be controlled to some extent based on the incoming char load and targeted temperature levels. With clean wood, for example, the carbon conversion in the gasifier was high, and then the oxidizer was operated with as low fluidizing velocity as possible, while with bark more charcoal entered into the oxidizer and thereby the air flow was also increased.
- Very good and stable operation periods (set points) could be realized with crushed wood pellets, bark and forest residues simulating different operation methods (from steam alone operation to operation with a mixture of steam and oxygen/air).
- The experiences with used wood were also good after some initial challenges related to the handling and feeding of this low-bulk density crushed feedstock were solved.
- Wheat straw pellets with high alkali content could also be used in the process without any signs of ash sintering in the gasifier or in the oxidizer, but the flue gas line of the oxidizer was gradually blocked by straw ash after a few hours of operation. The flue gas cooling and cleaning system should have been reconstructed in a similar way as is used in straw combustion units. This was, however, not realized because the main focus of the project was in using woody residues and these modifications could not be done within the project budget.

4.1.2 CFB gasifier and CFB oxidizer tested in 2018

After the DFB test runs of 2017, it was decided to carry out the following plant modifications:

- Replacement of the previous filter vessel of the gasification line with a new vessel, which can be operated at higher temperatures and which has better insulation and heating.
- Replacement of the BFB oxidizer with a CFB oxidizer in order to study this gasification process alternative.
- Replacement of the small flue gas filter with a larger filter unit in the flue gas line as now the fuel gas flow from the oxidizer was also higher.
- Modifications to the slip-stream gas cleaning facility in order to study the two-stage sulphur removal process, where the bulk of sulphur is removed with an active coke sorbent and the polishing step is done using ZnO-based sorbents.
- The automation system of the pilot plant was also partly renewed, as the system became more complicated, when also the slip-stream facility was included in the automation system.

The modified DFB test rig used in 2018 is illustrated in Figure 5 and the dimensions and operating conditions are summarized in Table 3. The gasifier reactor was not changed, but the oxidiser was replaced by a far higher reactor tube, which had the same diameter as the upper part of the gasifier. In this design, the recycling solid flows were taken from the bottom of both cyclones and fed back to the lower parts of the CFB reactors. The recycling flow rate was not controlled, and thus, all solid material that was separated in the gasifier cyclone ended up in the oxidizer and vice versa. Stable recycle flow could be maintained, when the gas velocities in the upper parts of the reactors were relatively close to each other, and when rather high bed levels were maintained in order to have some reserve to compensate fluctuations in flow conditions.

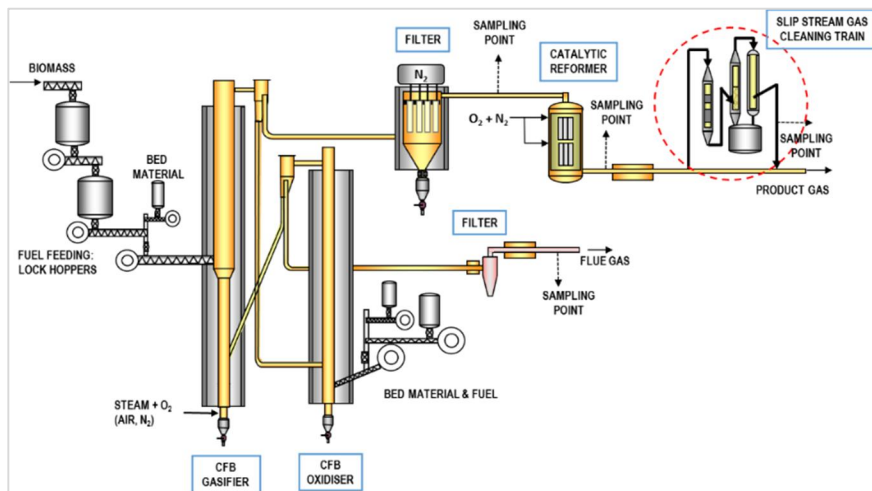


Figure 5. Modified CFB+CFB test rig at VTT's Piloting Centre Bioruukki in 2018.

These rather extensive plant modifications realised in autumn 2017 took longer than anticipated and required also more resources. Consequently, only two test runs could be carried out within the BTL2030 project timeframe and budget in 2018:

- DFB 18/23: Hot commissioning of the plant and the first gasification trial without detailed measuring programme.
- DFB 18/26: Successful test week with 60 hours of continuous operation and 6 measured set points.

Table 3. Reactor dimensions and main operation conditions of the DFB gasifier.

	CFB gasifier	CFB oxidizer
Reactor i.d., (mm)		
Lower part	102	150
Upper part	150	150
Reactor height (m)		
Total height	8.2	6.3
Height from distributor to gas outlet pipe	7.9	6.0
Height of the lower part	1.8	-
Fuel feeding point - distance from the gas distributor plate	2.3	-
Temperature range (°C)	720 - 780	780 - 820
Gas velocities, m/s		
Fluidizing velocity at the bottom of bed	5 - 6	3 - 4
Gas velocity at the top of reactor	3 - 4	3 - 4

The operation experiences of the DFB process realized with two CFB reactors can be summarized as follows:

- Maintaining stable bed inventories in both CFB reactors was more challenging than in the previous DFB system where the large bed inventory of the BFB oxidiser acted as a stabilizing reservoir. The gas velocities of both CFBs had to be kept rather close to each other, which means that the air flow rate (and temperature) of the oxidizer could not be as freely varied as in the previous process design.
- However, with correct velocities, the recycling loops operated smoothly and the use of large bed inventories stabilized the beds during process fluctuations. This could be further improved simply by adding small fluidized reservoirs into both recycling loops.
- Due to the higher air flow rates and high air ratios used in the oxidizer, it was not possible to reach as high operation temperatures as were used in the DFB tests carried out with the BFB oxidizer. Consequently, the whole process was operated at a lower temperature than originally planned. In real industrial plants, the temperature levels can be raised by using additional fuel, such as FT off-gases or external solid feedstocks in the oxidizer.

- It was interesting to notice that the DFB process could be operated successfully also at a gasification temperature as low as 720 °C. In spite of the relatively high tar loading, the product gas could be filtered and reformed without signs of increasing pressure drop or soot formation.

4.1.3 Basic DFB operation alternatives studied in the project

Based on the realised gasification test programme, it was concluded that the DFB steam gasification process can be designed with three different operating principles as illustrated in Figure 6.

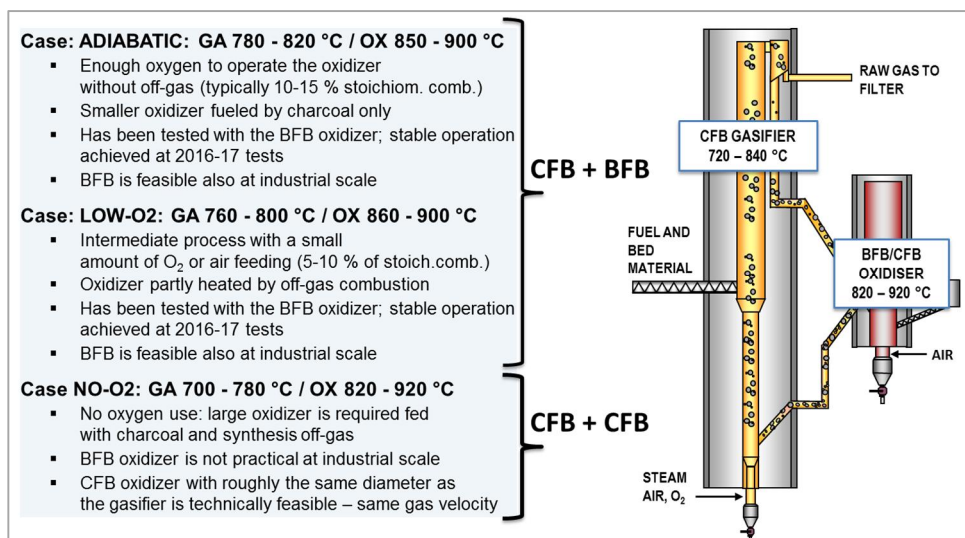


Figure 6. Basic versions of the DFB gasification process.

Cases “ADIABATIC” and “NO-O₂” represent the two basically different operation modes, while the case “LOW-O₂” is an intermediate case. **In the ADIABATIC case**, the gasification process is operated adiabatically, which means that no external fuel sources are used in the oxidiser. The heat required in steam gasification is provided partly by combusting the charcoal in the oxidiser and partly by feeding oxygen to the gasifier. In this case, the required oxygen feed corresponds to 10-15 % of the oxygen consumption of stoichiometric combustion, which is 50-70 % less than needed in direct steam/oxygen-blown gasifiers realised with a single gasifier reactor. In this operation mode, synthesis off-gases are not required in the oxidiser and thus, the synthesis can be realised with more efficient syngas recycling or the off-gases are available for generating more heat and power.

In the NO-O₂ case, all heat required for the endothermic gasification reactions is generated in the oxidiser and provided to the gasifier via bed material recycling. The two principle alternatives to reach this mode are to reduce the carbon conversion in

the gasifier so that enough charcoal is fed from the gasifier to the oxidiser or to use supplementary fuel in the oxidiser. In our case, when the gasifier is operated at 750-820 °C with woody feedstocks, it is easy to achieve 75-85 % carbon conversion already in the gasifier. When advanced FT synthesis or PSA hydrogen are the back-end processes, the most natural supplementary fuel for the oxidiser is the off-gas retrieved from the FT synthesis or PSA units. Another option would be to use additional biomass or waste as the external oxidiser feedstock.

The challenge of the NO-O₂ case is that the required temperature difference between the oxidiser and the gasifier in the tested DFB system was of the order of 100-150 °C. This means that the oxidiser should be operated at a rather high temperature, which might be problematic and lead to bed sintering or deposit formation if the feedstock is particularly rich in alkali metals. On the other hand, relatively high oxygen feed rate is required in the ADIABATIC case. **The intermediate LOW-O₂ case** was used to lower the required temperature difference as now a small part of heating was supplied by exothermic combustion reactions in the gasifier. In this operation mode, typically 5-7 % of the oxygen requirement of stoichiometric combustion was used. This amount of oxygen can be provided either as pure O₂ or as air depending on the back-end gas cleaning and synthesis (limited by the maximum allowed inert gas content).

It can be concluded that, when the DFB gasification system is equipped with the possibility to add oxygen or air to the gasification steam, the operation conditions can be more flexibly controlled, for example, when the feedstock quality is changing. The preliminary design studies of industrial-scale production units (100-150 MW biomass input) revealed that the DFB process design with a BFB oxidiser is feasible at large scale units only if part of the heat is provided by feeding oxygen/air also to the gasifier. Without the oxygen or air feed, the bed diameter of the industrial-scale BFB oxidiser becomes too large as the fluidization velocity needs to be low in bubbling bed mode. Therefore, the NO-O₂ case can best be realised by using two CFB reactors, which will in this case have rather identical diameters.

4.2 Hot filtration and reforming

4.2.1 Filter unit

Following gasification, the raw gas was led into the filter unit, where dust and condensed alkali metals were separated using special temperature and corrosion resistant metal filters. Filtration is needed to achieve efficient reforming and to protect the reformer from dust deposits.

In the tests of this report, the raw gas was filtered using 12 one-meter-long RHT filter elements that were organised in four clusters (3 elements in each cluster). Sintered metal filter elements of GKN (GKN Sinter Metals Filters GmbH) were applied successfully. The filters were pulse cleaned with nitrogen at regular intervals to detach the accumulated filter dust. The elements in one cluster were pulse cleaned simultaneously. The filter arrangement is depicted in Figure 7.

In 2016-2017 tests, the filter unit was operated in the temperature range of 600-750°C and with a face velocity of 3-5 cm/s. The filter material showed excellent performance and corrosion resistance. Practically 100 % particulate removal was achieved. The filter elements of GKN operated well and no filter breakages or other major problems occurred. Cumulative operation hours of filter elements was around 300 hours. However, the filter baseline pressure drop increased at some set points where the dust content was low and the tar content was high. To prevent increasing pressure drop tendency, an additional dust feeder was installed to the raw gas line before the filter and the recycling of filter dust was tested at some set points.

A method for regenerating sticky filter cakes was also developed and successfully used at the DFB facility, where it took only 2-3 hours to carry out the regeneration cycle and return back to stable gasification conditions. This regeneration procedure included the following steps:

- The gasifier operation mode was switched to steam/air-blown operation and the fuel feed rate was reduced. The oxidiser air feed rate was also reduced but bed material recycling was kept in operation.
- The biomass feed rate was reduced to a level that corresponds to combustion with a low excess air ratio - steam was used to prevent too high temperatures. In addition, at this test rig also nitrogen was available to dilute the gas flow.
- Controlled oxidation of the gasifier tube, cyclone and gas lines and finally oxidation of the filter cake. The O₂ content of the gas was kept below 1-3 % during this procedure.
- When oxygen appeared in the gas analysis downstream the filter unit, the procedure could be stopped. When the filter was pulse cleaned, it could be noticed that the pressure drop of the filter was reduced back to the original level.
- After filter regeneration, the process was returned back to the original gasification mode and the test was continued.
- As the test rig is equipped with a nitrogen purging system, this procedure could also be realised simply by stopping the fuel feeding and fluidizing the gasifier with a mixture of N₂, air and steam.

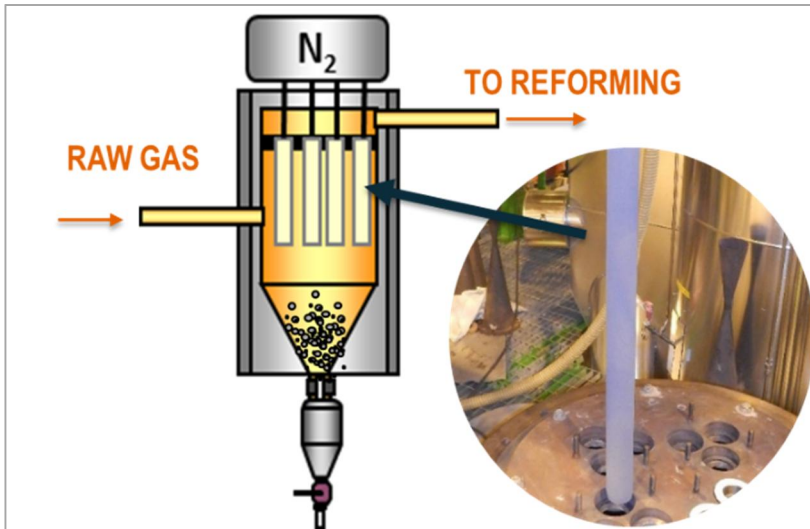


Figure 7. The candle filter arrangement and a photograph of one sintered metal filter element.

The steel structure of the original filter vessel had a maximum operation temperature of 700-750 °C and thus, it had been operated rather close to its upper limits. At the end of 2017, it was decided to construct a new filter vessel, which can be more safely operated at higher temperatures. The new filter vessel was installed and operated in the tests carried out in 2018. The same filter elements, as used in 2016-17 tests, were mounted back to the new filter vessel, which had the same internal dimensions as the original unit. In the tests of 2018, the filter elements got some 100 additional gasification hours totaling now to ca. 400 hours. In addition to these effective gasification hours, the GKN metal filters had experienced some 200-300 hours of other operations, including preheating and shutdown periods and intermediate hot stops between some test periods [2].

The main findings regarding hot gas filtration are:

- Hot filters operated well without filter breakages or other major problems.
- The dust content after the filter was very low (~ 1 ppm indicated by ELPI measurements).
- The CFB+BFB gasifier was more sensitive to tar-induced problems than the CFB+CFB design, where the dust content of raw gas was much higher and protected the filters from the formation of a sticky soot cake (i.e. filter blinding).
- Low-pressure steam gasification seems to be less sensitive to filter blinding than high-pressure steam-oxygen gasification. Even the last set

points carried out in 2018 at a low gasification temperature of 710 °C could be realized with stable filter pressure drop.

- Feeding of additional dust and online regeneration of filter elements by operating the gasification process as a combustor for a short period were successfully tested.

4.2.2 Catalytic reformer

Following filtration, the particulate-free gas was introduced into a two-stage catalytic reformer that was operated autothermally using a mixture of oxygen and nitrogen (simulating air reforming). In the reformer, tars and light hydrocarbon gases were catalytically reformed to carbon monoxide and hydrogen in the temperature range of 850 – 950 °C. The principal design of the reformer is presented in Figure 8. The design details of the reformer are proprietary information and subject to ongoing patenting.

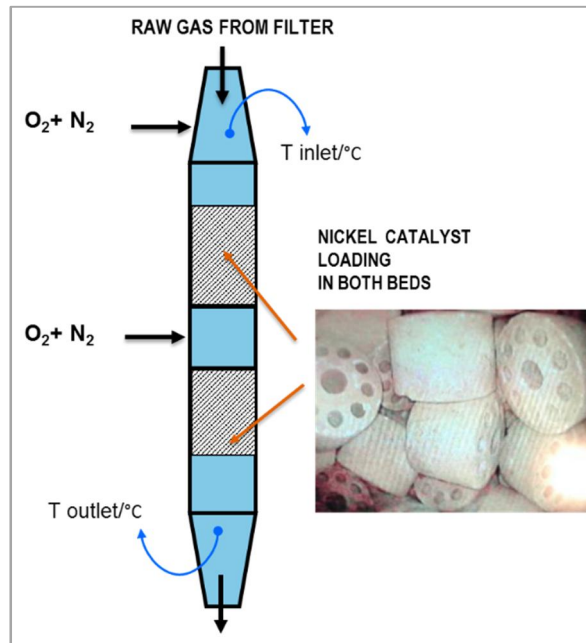


Figure 8. The structure of the two-stage reformer used in the gasification test runs.

The main findings regarding the two-stage reformer are:

- The reformer design seemed to work well: Heavy tars and C₂-hydrocarbons were effectively decomposed in the reformer.
- No signs of soot formation was detected and the pressure drops in both reformer beds remained stable throughout the tests.

- Methane and benzene conversions in the reformer were inadequate but can be improved for example by adding a third reformer bed and/or using a more effective catalyst in the final bed.
- Reforming of low-pressure gas derived from steam gasification is less sensitive to soot formation but requires the use of lower space velocities than reforming of raw gas derived from higher pressure O₂-blown gasifier tested previously [2].

4.3 Results of the DFB gasification tests

4.3.1 Feedstocks and bed materials

Table 4 presents average results for the proximate and ultimate analyses of the feedstocks used in the DFB tests in 2016-2018. Photographs of the feedstocks are presented in Figure 9 and the ash composition of woody feedstocks is illustrated in Figure 10.

All set points selected for the detailed results evaluation of this report were carried out either with crushed wood pellets or crushed bark pellets. When the inhomogeneous woody residues were at first pelletized and then crushed (to below 7 or 10 mm sieve), the resulting feedstock was rather homogeneous and easy to handle in the feeding systems of this small pilot plant. In addition, the fuel feed rate remained constant resulting in very stable operating conditions. The particle size distribution of the two main feedstocks are shown in Appendix A.

Table 4. Feedstock analyses as used in the gasification campaigns at VTT Bioruukki.

	Wood pellets	Crushed bark	Forest residues	Used wood	Straw
LHV MJ/kg d.b.	19.0	18.9	19.7	18.8	17.2
Moisture (avg), wt%	7.4	10.3	9.3	14.2	7.7
Proximate analysis, wt% d.b.					
Volatile matter	83.5	77.1	76.0	80.7	75.2
Fixed carbon	16.1	19.7	21.3	18.5	16.8
Ash, wt% d.b.	0.4	3.2	2.7	0.85	8.0
Ultimate analysis, wt% d.b.					
C	50.8	51.7	52.1	50.2	45.6
H	6.0	5.9	5.9	6.2	5.8
N	0	0.3	0.4	0.6	1.6
Cl	nd	0.012	0.018	0,01	nd
S	0.01	0.04	0.04	0.015	0.15
O as difference	42.7	39.0	38.9	42.2	39.0
Ash	0.4	3.2	2.7	0.85	8.0



Figure 9. Photographs of the feedstocks.

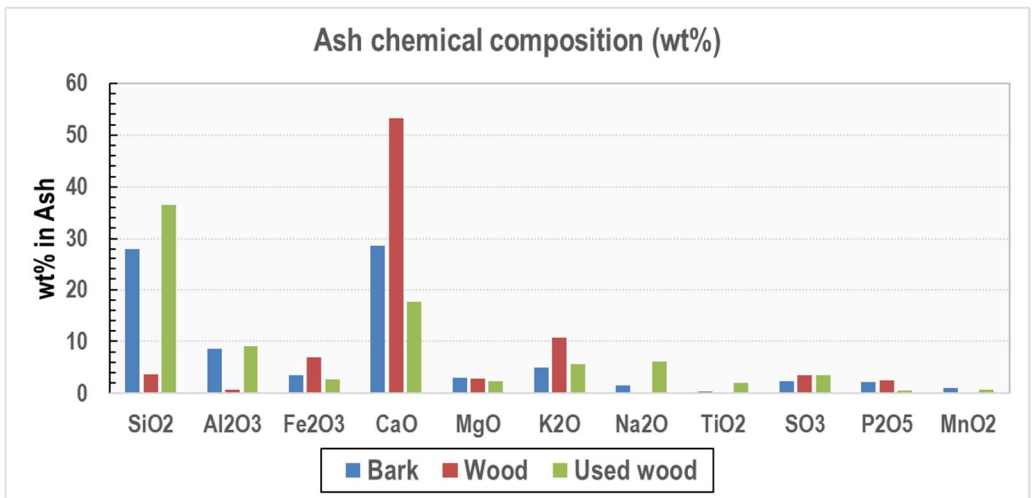


Figure 10. Ash composition of wood, bark and used wood (CuO₂, Cr₂O₃, ZnO, NiO contents < 0.3 %).

A mixture of dolomite and sand was used as bed material in the DFB test campaigns. The particle size of the bed material was selected in a way to ensure good fluidisation and to achieve suitable degree of bed material recycling in different operating conditions. Fresh bed material was fed either to both reactors or only to the oxidizer. Used bed material was removed from the bottom of both reactors.

Being a fragile material, most of the dolomite was attrited in the process and entered finally either to the flue gas or to the raw product gas and was then removed by the filter units. Most of the sand was removed as bottom ash from both reactors. This type of operation with a mixture of hard sand and soft and porous dolomite particles has turned out to be an efficient way of preventing ash agglomeration in fluidised-bed gasifiers as described in our previous reports [7], [8] and [9]. The chemical composition of the bed materials is given in references [3] and [10].

4.3.2 Operating conditions and gasifier performances at selected set points

The gasification tests were realised during six successful test weeks and a total of 34 set points were measured. These set points represent stable operation periods of 2-24 hours carried out with different feedstocks in different operating conditions. Seven representative set points were selected for detailed process performance studies. Three of them (A, B, C) were realized with crushed clean wood pellets and four (D, E, F, G) with crushed bark as the feedstock. Set points A-E are from the 2016-17 test runs, where the oxidiser was a BFB reactor: The last two set points F and G are from 2018 test run, where CFB oxidiser was in use. The gasifier reactor dimensions and design were the same in all test runs. The dimensions of the filter and the reformer were also the same in all test runs, although a new filter vessel was installed in 2018. Thus, the gasification results are comparable with each other. The used sampling and analytical methods are described in more detailed in previous reports [11], [12], [13]. Tar sampling was carried out according to the standard method described in the European Tar Protocol [14].

Set point A with clean wood and **Set point D** with bark represent “NO-O₂” operation mode, although a small amount of oxygen/air was used also at these set points in order to keep these feeding systems operational for the purpose of other set points of the specific test runs. At these two set points, the additional heating of the system was accomplished by operating the electrical heaters of the reactor tubes and recycling loops in a way that the system was partly electrically heated.

Set points B, C and E were planned to represent “ADIABATIC” operation, and some 10 % of the oxygen of stoichiometric combustion was fed into the gasifier. However, as the carbon conversion efficiency of the gasifier was increased significantly, the char flow to the oxidiser was reduced and, in fact, part of the heating was again provided by electrical heaters.

Set points F and G were realised in 2018 using the CFB oxidiser. The operation temperatures of the gasifier and the oxidizer were lower than in previous set points carried out with the BFB oxidiser. This was partly due to the fact that the air flow rate of the oxidizer had to be kept higher than required for charcoal combustion in order to keep the gas velocities in both reactors close to each other (for maintaining stable bed material recycling). Another reason was that with higher gas velocities and much higher recycling lines, the electrical heaters were not sufficient in providing the required additional heat. In future experiments, the CFB oxidizer must be equipped with a possibility to use supplementary feeding of additional solid or

gaseous feedstock. The oxygen feed rate corresponded to ca. 5 % of that of stoichiometric combustion. However, this limitation had a positive consequence as the gasifier was now operated at a lower temperature than planned. Possibly the most important result of the last test campaign was the fact that this DFB process could also be operated at gasification temperatures of 710-730 °C and still there were no problems in hot gas filtration or reforming.

When the results of the last two set points are compared, the only difference in the operation conditions is the fact that in set point G only sand is used as the bed material. The results clearly indicate that under the conditions of set point F, the calcium loop is contributing to the heat transfer between the gasifier and the oxidiser. The temperature difference at set point F is clearly lower than in set point G. In the oxidiser, calcium carbonate is calcined to calcium oxide. This reaction consumes heat and reduces the oxidiser temperature. When the gasifier is operated at close to 700 °C, the calcium oxide coming from the oxidiser is reacting back to carbonate, and the calcination heat is released, which provides heat for the gasification. This principle of calcium looping gasification did not take place in our previous tests because also the gasifier was operated at higher temperatures where the equilibrium of calcium is on the side of calcium oxide.

The material and energy balance calculations for the DFB gasification process are somewhat more complicated than those of direct air or steam-O₂ blown gasifiers described in our earlier reports [2], [3], [11]. This also means that possibilities for inaccuracies increase. This is due to the fact, that the products are distributed in two gas streams (i.e. gasification gas and flue gas), the flow rate of which cannot be accurately measured.

The main principles of the calculation method were as follows:

- The average analysis results for the set point samples were used for solid input and output streams and the mass flows were based on actual weighed amounts over the whole set point period.
- Average analyses of raw gas and flue gas composition and contents of tars, nitrogen species were used.
- The average values collected by the automation system of the plant were used for the input gas streams, temperatures and pressures.
- The material balance calculation is started by calculating the elemental material balances for the oxidizer using the following principles:
 - Nitrogen balance of the oxidizer is used to calculate the flue gas mass flow. Carbon balance is used to calculate the (net) flow of carbon from the gasifier to the oxidizer. The analyses of gasifier filter fines are used to describe the composition of char coal (N, S, O and ash content) and to define the input flow from the gasifier to the oxidizer.
 - The fly ash and bottom ash samples are also analyzed and the elemental flows are taken into account in the elemental balances.

- The net heat flux to the circulating bed material (from oxidizer to gasifier) is calculated as a difference between the energy input and output of the oxidizer (heat of combustion of the input charcoal + enthalpy of the combustion air - flue gas enthalpy - heat of combustion of solid output streams).
- The oxygen balance is a free balance and its accuracy indicates the quality of the measurements and the calculated material balances.
- Secondly, the elemental material balances and energy balance of the gasifier are calculated according to the following principles:
 - The charcoal flow rate from the gasifier to the oxidizer and its elemental composition is taken from the oxidizer calculations.
 - Carbon balance is used to calculate the dry gas flow rate taking into account other carbon flows which are either measured directly or calculated from the oxidizer balance.
 - Oxygen and nitrogen balances of the gasifier are free balances, the accuracy of which can be used to estimate the reliability of the measurements and the calculation procedure.
 - The energy balance is calculated using LHV values for the solid output and input streams and calculated LHV for the gas and tars. The difference between the input and output energy flow is used to estimate how much of the required energy is provided by electrical heaters or whether the system has been operated with heat losses.

Finally, various conversions, product yield and other specific performance indicators are calculated using the results of the material and energy balances. These are defined and calculated according to the principles used in our previous studies [2], [3], [7].

Table 5. Main operating conditions of the gasifier and the oxidiser in selected set points with wood and bark.

Test run	A	B	C	D	E	F	G
GASIFIER							
Fuel	Wood	Wood	Wood	Bark	Bark	Bark	Bark
Moisture content, wt%	7.3	7.3	7.6	8.7	8.7	13.5	10.3
Fuel feed rate, g/s	8.6	7.8	11.1	7.8	7.8	7.45	7.45
Bed material feed rate, g/s	0.49	0.49	0.89	0.41	0.41	0	0
Bed material	Dol+S (70/30)	Dol+S (70/30)	Dol+S (70/30)	Dol+S (70/30)	Dol+S (70/30)	Dol+S (70/30)	Sand
Air feed, g/s	0.7	0.7	0.2	0.7	0.7	2.8	2.8
Steam feed, g/s	7.5	7.2	7.0	7.5	7.2	6.5	6.5
Oxygen feed, g/s	0.2	1.1	1.4	0.2	1.1	0	0
Purge N ₂ feed, g/s	2.2	2.1	2.5	1.8	1.8	2.0	2.0
Steam-to-fuel ratio (daf-basis)	0.95	1.00	0.69	1.10	1.06	1.01	1.01
O ₂ +Air feed, % of stoich. combustion	1.8	10.7	9.9	1.9	10.5	4.9	4.9
Average gasifier bed-T, °C	785	828	838	783	826	712	712
Average gasifier upper part T, °C	781	825	829	781	825	735	723
P-freeboard, bar (abs)	1.02	1.02	1.02	1.02	1.02	1.02	1.02
Fluidising velocity, m/s	4.2	4.5	4.5	4.2	4.5	3.9	3.9
Gas velocity at gasifier top, m/s	4.4	4.5	5.3	4.0	4.4	3.3	3.3
Dry gas, vol-%, measured							
CO	16.7	13.6	15.4	12.7	11.9	9.0	9.3
CO ₂	21.9	26.7	27.3	23.2	27.4	21.1	22.0
H ₂	32.9	33.3	33.3	35.2	33.0	28.1	26.2
N ₂ (calc. as difference)	19.2	18.0	14.2	18.8	18.8	32.7	32.9
CH ₄	6.65	6.19	7.12	6.81	6.27	6.43	6.64
C ₂ H ₂	0.13	0.05	0.04	0.09	0.05	0.05	0.06
C ₂ H ₄	2.12	1.80	2.17	2.52	2.21	2.26	2.38
C ₂ H ₆	0.45	0.34	0.46	0.52	0.38	0.39	0.41
C3-C5	0.08	0.02	0.02	0.10	0.03	0.07	0.10
Benzene, g/m ³ n (dry gas)	10.1	10.4	13.1	12.8	13.2	11.7	11.5
Sum of tars, g/m ³ n (dry gas)	14.6	7.6	9.9	16.4	10.7	20.5	21.7
Heavier PAC, g/m ³ n (dry gas)	6.5	2.4	5.4	8.0	3.8	8.2	8.0
Ammonia, mg/m ³ n (dry gas)	311	nd	490	2238	2242	1910	nd
H ₂ O content in wet gas, vol-%	43.1	41.3	35.2	45.4	42.8	42.4	44.5
Wet gas flow rate, g/s	18.6	18.8	22.2	16.6	18.1	16.6	16.7
C conversion to gas and tar, %	82.0	90.8	90.5	78.0	86.3	74.6	74.3
Chemical efficiency (LHV of gas + tars/ LHV of fuel)	81.8	82.7	82.7	79.4	78.3	77.0	74.0

OXIDIZER							
Air feed, g/s	9.4	8.2	8.0	9.6	8.0	14.8	16.0
Purge N ₂ feed, g/s	0.7	0.7	0.7	0.7	0.7	0.1	0.1
Average oxidizer bed-T, °C	896	858	882	921	883	805	843
Average oxidizer upper part T, °C	871	852	867	883	858	793	816
P-freeboard, bar (abs)	1.02	1.02	1.02	1.02	1.02	1.02	1.02
Fluidising velocity, m/s	0.37	0.31	0.31	0.39	0.31	2.35	2.61
Dry gas, vol-%, measured							
CO ₂	15.80	8.50	11.12	15.60	11.20	11.37	9.97
N ₂ (calc. as difference)	80.01	80.14	78.77	80.30	80.42	78.04	79.38
O ₂	3.40	10.50	9.28	3.30	7.60	9.70	9.74
Ar	0.79	0.76	0.83	0.81	0.78	0.89	0.91
SUMMARY FOR THE DFB PROCESS							
Oxidizer T (bed) - Gasifier T (bed), °C	111	30	44	137	57	93	130
Calculated heat input from char combustion, % of the input energy of biomass	10	3	4	11	5	11	11
Calculated electrical heating as % of the energy input of feedstock	4	5	2	3	2	7	4

The temperatures of the gasifier, the oxidizer and the filter during one whole test run of week 17 in 2017 is presented in Figure 11. The change from set point A to B by increasing the oxygen feed of the gasifier can be clearly seen as an increasing gasifier temperature and decreasing oxidiser temperature as more charcoal is reacting already in the gasifier. By comparing set points A and D, which were carried out at the same gasification temperature and with only a minimal oxygen feed, the reduced carbon conversion in the gasifier is clearly seen as higher temperature in the oxidiser. The last operation period of week 17/2017 was carried out with crushed used wood, which was less homogeneous than crushed wood and bark pellets. The variation in fuel feeding can also be seen in the variation of temperatures. Especially the oxidizer temperature was rather sensitive to the variation in fuel feed rate and particle size.

Generally, this version of DFB gasification realised with a CFB gasifier and a BFB oxidizer could be rather smoothly operated with different target operation modes. In addition, the feedstock could be smoothly changed without severe process disturbances.

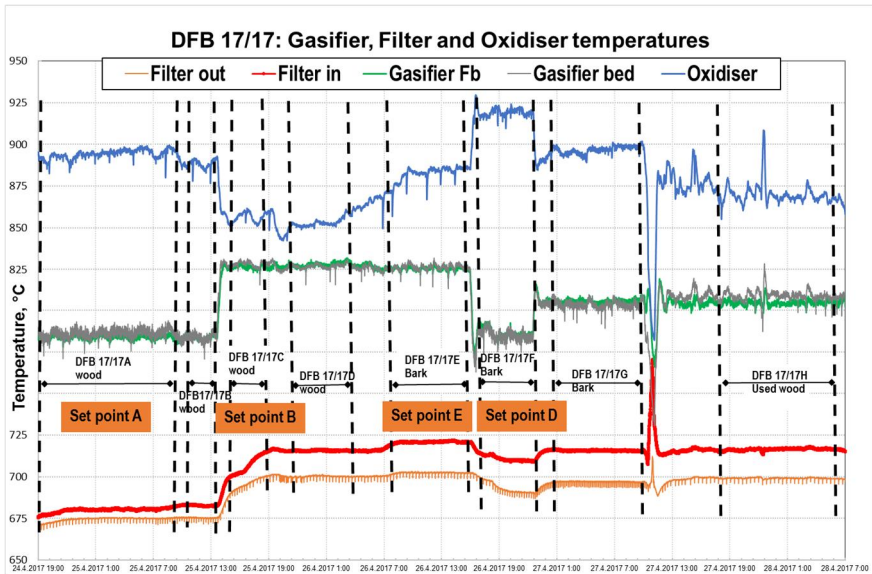


Figure 11. An example of the variation of the process temperature during the test runs of week 17/2017.

The temperature differences between the oxidiser and the gasifier beds at different set points are illustrated in Figure 12 and the carbon conversions of the gasification stage in Figure 13. In gasification of crushed clean wood pellets, the carbon conversion efficiency was already 82 % in set point A and increased to above 90 %, when the oxygen feed rate was increased in set points B and C. This can be seen clearly also in the bed temperatures. In set points B and C, the difference was only 30-44 °C, which was not enough for providing the required heat for the gasifier in spite of the oxygen feed. More charcoal is produced in gasification of bark and this can be seen both in the temperatures of Figure 12 and in the carbon conversions of the gasifier shown in Figure 13. The carbon conversions in set points D and E are ca. 4 %-point lower than conversions determined under similar operating conditions with clean wood (set points A - C).

When the gasification temperature was lowered to ca. 710 °C in the last two set points, the carbon conversion of bark gasification decreased to 74-75 %. This indicates that at higher temperatures also steam gasification reactions play some role, while the last set points were close to pyrolytic gasification assisted by a little bit of combustion.

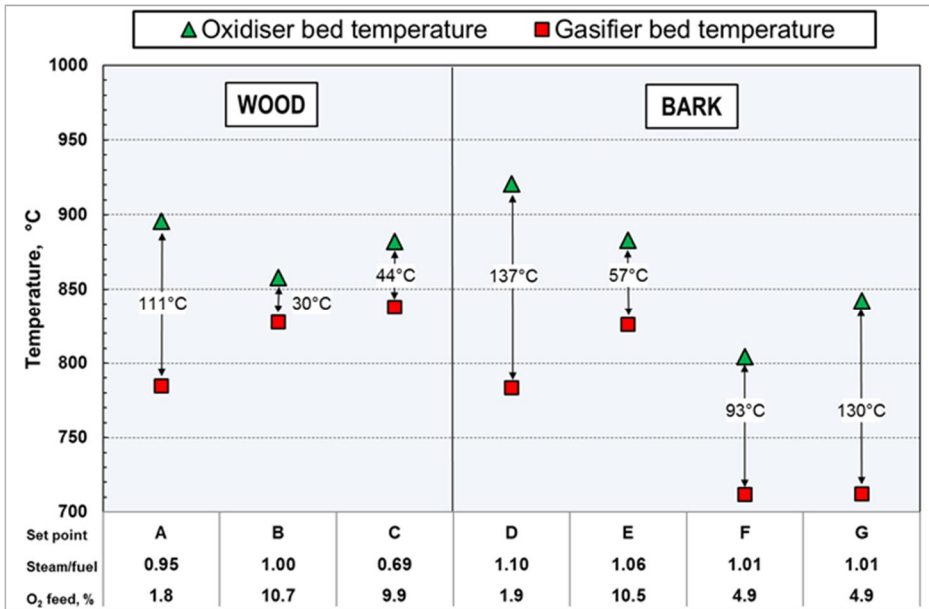


Figure 12. Temperature difference between the gasifier and the oxidiser.

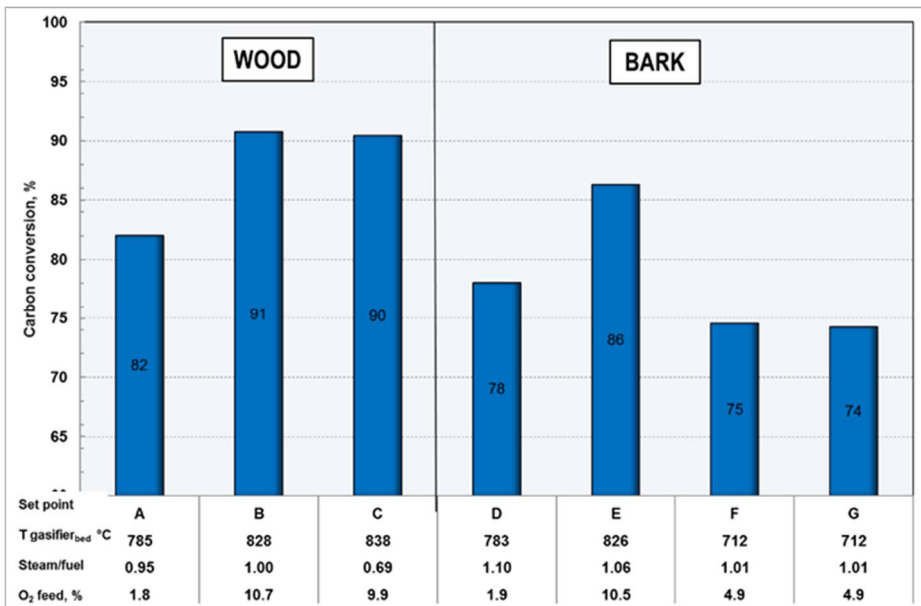


Figure 13. Carbon conversion in the gasifier.

The calculated chemical efficiencies of the gasification stage are illustrated in Figure 14. Here the chemical efficiency is defined as a ratio of the output energy content of the raw gas (incl. tars and ammonia) divided by the energy input in the form of biomass feedstock. Both energy contents are calculated based on lower heating values. It is interesting to notice that the increased feeding of oxygen (from set point A to B and from D to E) does not clearly improve the chemical efficiency of gasification in spite of clearly increased carbon conversions. It is also interesting to notice that the chemical efficiency is also rather high at the “chemical looping” set point F in spite of the very low gasification temperature. The calcium looping is also transforming part of the CO₂ from the gasifier to the oxidiser and thus affecting the ratios of main gas components.

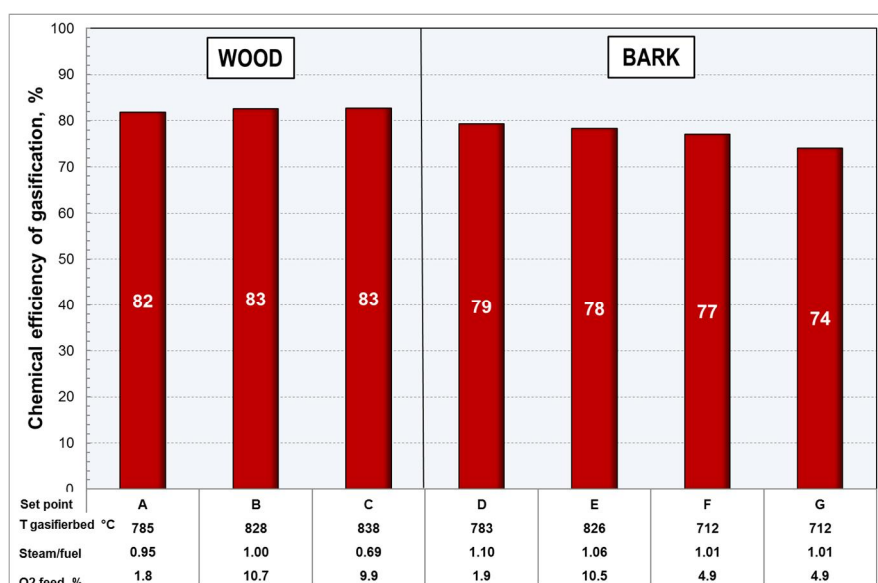


Figure 14. Chemical efficiency of the gasifier.

The measured tar and benzene contents are illustrated in Figure 15 and examples of detailed tar analyses for the two extreme cases (set points B and F) are presented in Appendix B. The sampling point was located after the hot filter unit. Thus, these tar concentrations are the final outcome of the tar decomposition reactions taking place in the gasifier as well as on the filter cake. The tar results were generally in good agreement with the conclusions of our previous direct fluidized-bed experiments [15]. In set points B, C and E, the tar composition was also similar to that of air or steam-oxygen-blown CFB gasification [2], [3], [7]. The effect of increased temperature on tar composition and total contents can be clearly seen when comparing the results of set points A and B. The increase of gasification temperature from 780 °C to 830 °C results in efficient decomposition of light tar components (from pyridine to indene), and also the amount of heavier unidentified

pyrolyzates is reduced drastically. On the other hand, the concentrations of most stable components, like benzene, naphthalene and some heavier polyaromatic compounds are decreasing only slightly.

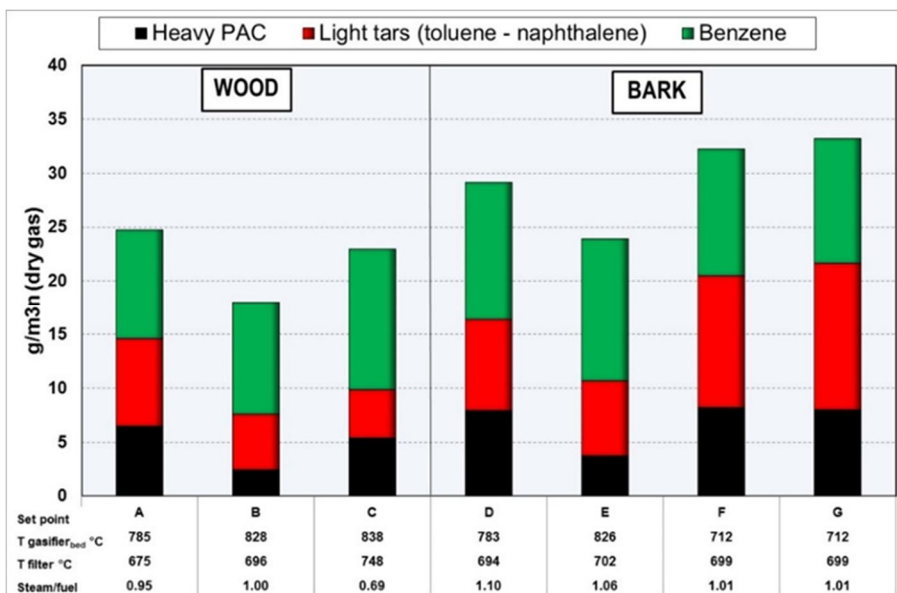


Figure 15. Tar contents measured from raw gas after the filter unit.

4.3.3 Operation conditions and performance of the filter unit

The operating conditions of the filter unit are summarized in Table 6. The particulate content of raw gas at filter inlet is calculated based on the weighed filter dust batches removed from the bottom of the filter during each set point period. It can be noticed, that the dust loading was highest in the last two set points, where both reactors were CFBs and similarly designed recycling cyclones were used to separate the solids from both gas streams. As this filter unit was originally designed for an air-blown CFB gasifier and for filtration temperature of the order of 400-600 °C, the face velocities (3-5 cm/s) were slightly too high due to the increased volume flow of the raw gas. In spite of this, the filtration unit could be operated in most cases with stable pressure drop.

A typical pressure drop curve for the filter unit over one-week-long test run is presented in Figure 16 (the last test run carried out in week 18 of 2018). This test run was started with a period of clean wood gasification and the filter unit was operated at 690 °C. Under these conditions, the pressure drop seemed to increase continuously indicating that the pulse cleaning was not effective enough to remove the filter dust cake. When the feedstock was switched to crushed bark, the operation of the filter stabilized and the dust cake could be removed without tendencies of

increasing pressure drop. Similar differences between clean wood and bark or forest residues was noticed also in other test runs. Evidently, as a result of lower carbon conversion of bark in gasification, a higher amount of fine charcoal particles is carried over to the filter. These porous char particles (like active coke) seem to adsorb heavy tars and soot thereby preventing the formation of a sticky filter cake, which would ultimately lead to filter blinding. Similar phenomena was also observed in earlier pressurized gasification tests described in [7] and [16].

Table 6. Operating conditions of the filter and the reformer in gasification experiments.

Test run	A	B	C	D	E	F	G
T before filter, °C	680	707	757	711	721	700	700
T after filter, °C	675	696	748	694	702	nd	nd
Particulate content at filter inlet, g/m ³ n	7.8	4.3	13.2	12.7	10.9	21.5	23.2
Filter face velocity, cm/s	3.7	3.8	4.6	3.5	3.7	3.3	3.3
Pressure drop, mbar	41.2	54.0	51.8	60.0	55.0	55.0	57.0

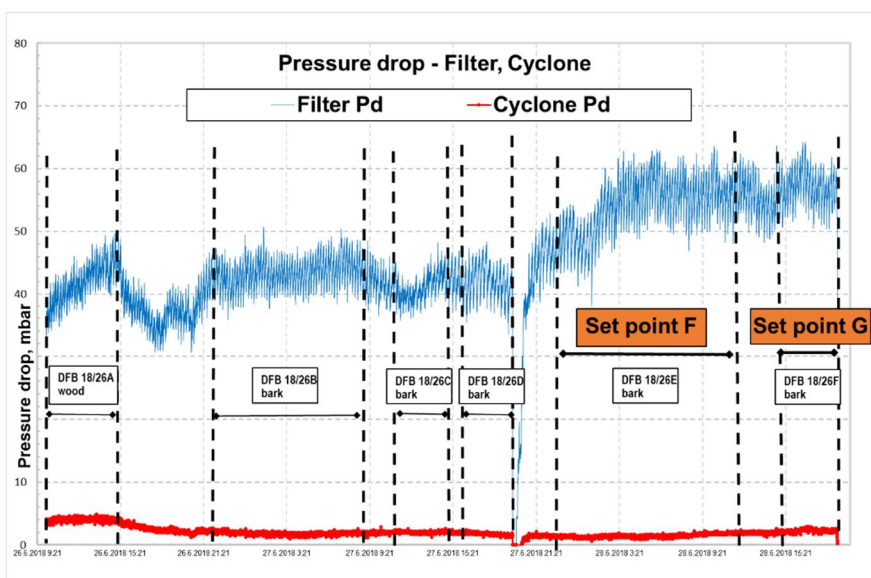


Figure 16. Pressure drops across the raw gas filter and the recycling cyclone in the test run of week 26/2018.

The effect of the hot filter unit on tar concentrations was investigated in the test runs of week 20/2016, where tars were measured both before and after the filter unit. The results for two different filtration temperatures are presented in Figure 17. The tar results represent an average of 2-5 samples (each representing 10-12 minutes sampling period). When the filter was operated at 635 °C, the tar content remained unchanged, while at higher filtration temperature of 734 °C the tar concentrations were clearly reduced in the filter. Parallel bench-scale tests carried out in a bubbling fluidised-bed test rig at VTT clearly demonstrated that when the filtration temperature is raised to above 800 °C, the tar contents are significantly reduced due to continued tar reactions on the filter cake and/or in the filter matrix [10].

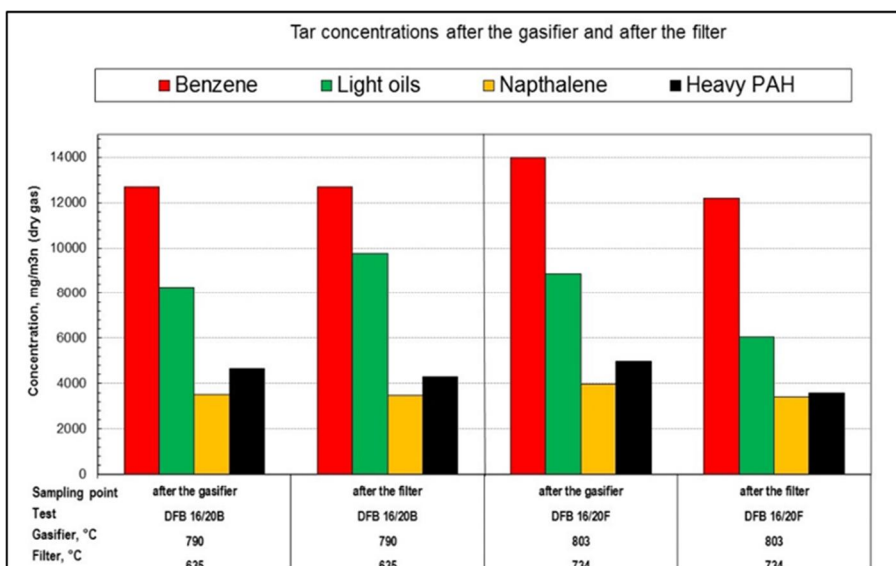


Figure 17. Tar concentrations measured before and after the hot filter unit in test runs of week 20/2016.

4.3.4 Operation conditions and performance of the catalytic reformer

The operating conditions, measurement results and calculated performances of the catalytic reformer in different set points are presented in Table 7, and typical reformer temperatures and pressure drops over one week of operation are illustrated in Figure 18-19. The online gas analysis of CO, H₂, CO₂ and CH₄ measured before and after the reformer unit in test run of week 18/2018 is presented in Figure 20.

The reformer was operated with the same nickel catalyst loading and with similar operating principles in all set points presented in this report. The oxygen feed rate was adjusted so that the maximum temperatures at the top of both beds remained below 1000 °C and the outlet temperature after the second bed was in the range 905 - 915 °C. The space velocity (calculated for the total bed volume) varied in the range 1400 - 2100 1/h (STP). As can be seen from Figure 18, the reformer operation was stable and the oxygen feed required hardly any adjustment when the gasifier was operating steadily. When the operation conditions in the gasifier were changed, the oxygen feed to the reformer bed was adjusted accordingly in order to maintain the operation temperatures more or less constant. The pressure drop curves shown in Figure 19 were typical to all test runs of this project. The pressure drops remained stable and there were no signs of soot or deposit formation in the catalyst beds.

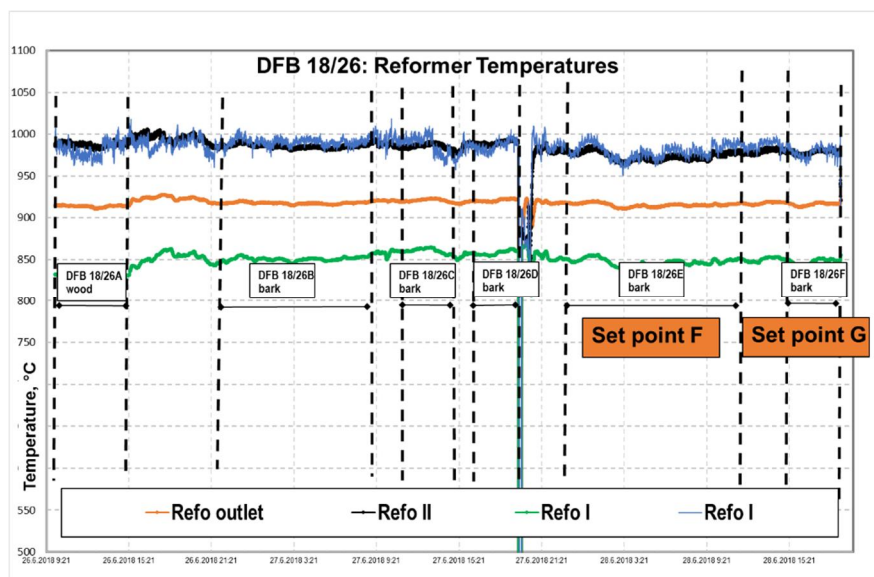


Figure 18. Reformer temperatures during the test run of week 18/2018.

Table 7. Operating conditions and measurement results of the reformer.

Test run	A	B	C	D	E	F	G
T at reformer outlet, °C	906	908	912	913	913	915	916
T _{avg} in Reformer I, °C	858	870	884	892	891	912	914
T _{avg} in reformer II, °C	920	923	938	935	932	940	943
Pd - reformer I, mbar	10,4	11.3	12.5	9.9	10.7	13.2	13.7
Pd - reformer II, mbar	13.2	14.2	15.5	12.5	13.5	15.8	16.4
Space velocity (1/h, STP)	1760	1680	2120	1520	1580	1460	1420
Dry gas after reformer, vol-%,							
CO	17.7	18.1	20.6	16.1	16.5	13.1	13.3
CO ₂	17.0	19.1	19.1	17.7	19.9	17.3	17.9
H ₂	42.5	39.2	37.5	41.0	37.9	34.5	34.1
N ₂ (calc. as difference)	21.5	22.3	20.8	22.6	23.3	32.4	31.8
CH ₄	1.27	1.25	1.99	2.56	2.29	2.76	2.85
C ₂ H ₂	0.0	0.0	0.0	0.0	0.0	0.0	0.0
C ₂ H ₄	0.01	0.01	0.02	0.04	0.03	0.06	0.07
C ₂ H ₆	0.0	0.0	0.01	0.0	0.0	0.01	0.01
C ₃ -C ₅	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H ₂ S	nd	nd	nd	nd	nd	nd	nd
COS	nd	nd	nd	nd	nd	nd	nd
Benzene, mg/m ³ n (dry gas)	82	73	194	1032	700	1088	1308
Sum of Tars, mg/m ³ n (dry gas)	8,2	6.9	13.3	86.0	43.3	128.1	174.0
Heavier PAC, mg/m ³ n (dry gas)	0.0	0.0	0.0	0.0	0.0	12.0	18.0
Ammonia, mg/m ³ n (dry gas)	157	nd	190	809	582	605	nd
HCN, mg/m ³ n (dry gas)							
H ₂ O in wet gas, vol%	28.5	31.2	26.4	32.9	32.6	31.8	33.1
Wet gas flow rate, g/s	20.8	21.2	26.0	18.9	20.5	18.9	18.8
Kp-shift from gas composition	1.02	0.91	0.97	0.92	0.95	0.98	0.93
Calculated equilibrium temperature of shift reaction, °C	810	844	823	839	831	818	839
Methane conversion, %	70.3	71.5	60.4	44.2	48.3	38.8	38.2
C ₂ -C ₅ Hy conversion, %	100	100	100	100	100	100	100
Benzene conversion, %	98.8	99.0	97.9	88.0	92.5	86.7	83.6
Tar conversion, %	99.9	99.9	99.8	99.2	99.4	99.1	98.8
Heavy PAC conversion, %	100	100	100	100	100	99.8	99.7
Ammonia conversion, %	32.6	nd	48.3	52.4	65.5	56.7	nd

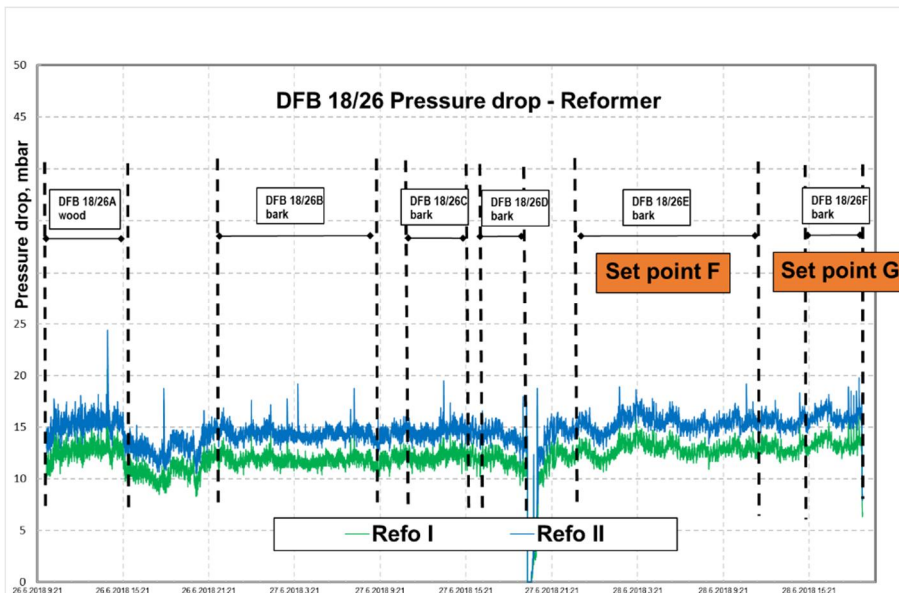


Figure 19. Pressure drops of the reformer beds in the test run of week 18/2018.

Figure 21 illustrates measured tar concentrations after the reformer in different set points, while ammonia concentrations are presented in Figure 22, respectively. The tar results - methane and benzene conversions in particular - clearly show the effect of sulphur on the nickel catalyst performance. Being a catalyst poison, sulphur causes catalyst deactivation and thus, higher tar conversions were achieved with clean wood than with bark that contains more sulphur. The calculated tar conversion efficiencies are presented in Figure 23 and the conversion of methane, C₂-hydrocarbon gases and ammonia in Figure 24. The concentration of sulphur gases were not measured in all set points. When bark was used as feedstock, H₂S and COS concentrations measured after the hot filter unit were in the range 140 - 180 ppm-v and 3-5 ppm-v, respectively. With clean wood, sulphur species were measured only in one set point: H₂S concentration was 20 ppm-v and COS concentration 0.5 ppm-v. Complete conversion of C₂-hydrocarbon gases and heavy tars (heavier than naphthalene) was achieved in all set points of this report, but the conversion of benzene and methane was lower than achieved in previous pressurized steam/oxygen-blown gasification experiments [2] in spite of the lower volumetric space velocities applied. Higher methane and benzene conversions could be achieved if a third reformer bed was incorporated and a more effective catalyst was used in the final reforming stage.

The composition of reformed product gas is illustrated in Figure 25. The concentrations are presented for dry gas on purge nitrogen-free basis (purge nitrogen extracted).

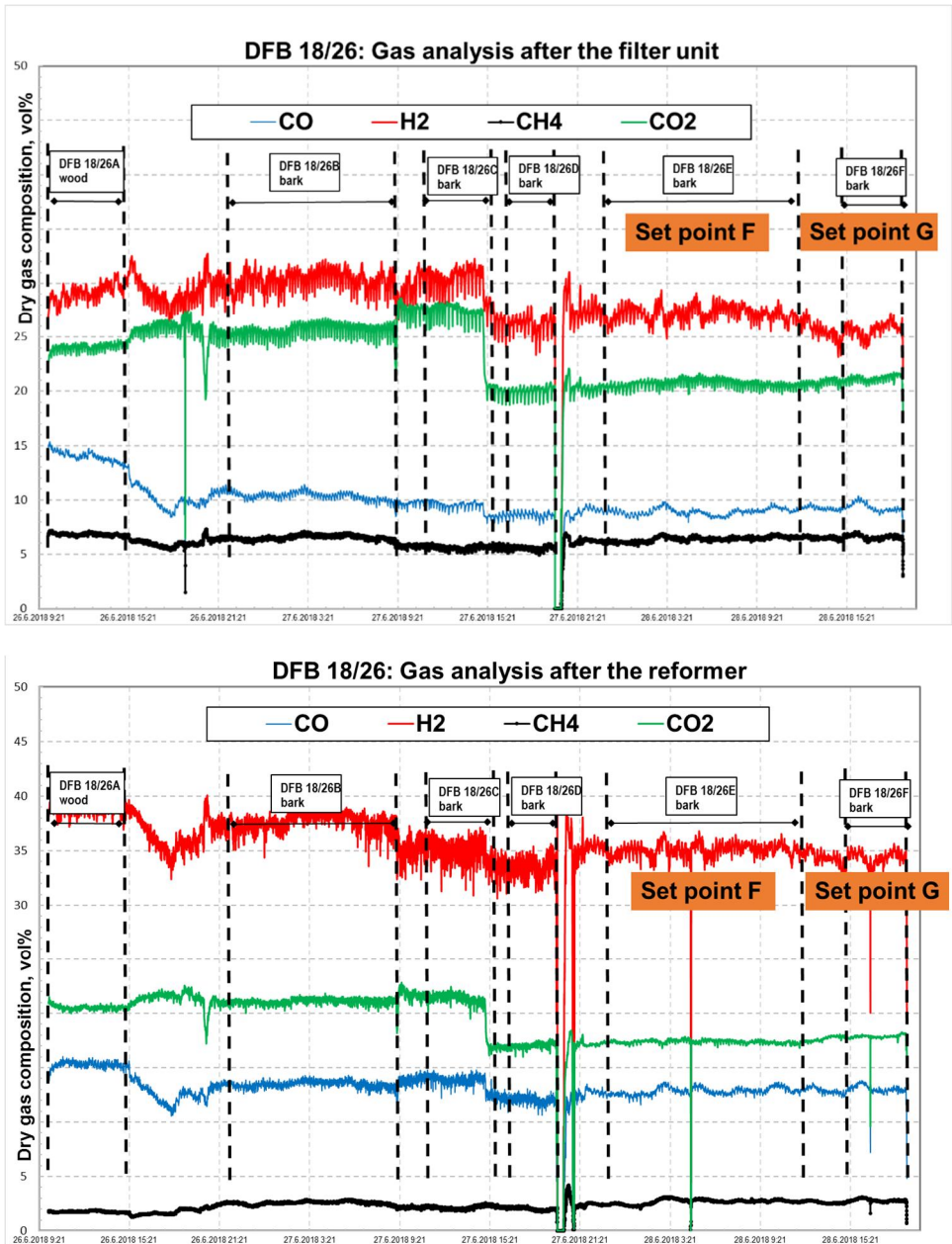


Figure 20. Measured concentrations of the main gas components before and after the catalytic reformer in test run of week 18/2018.

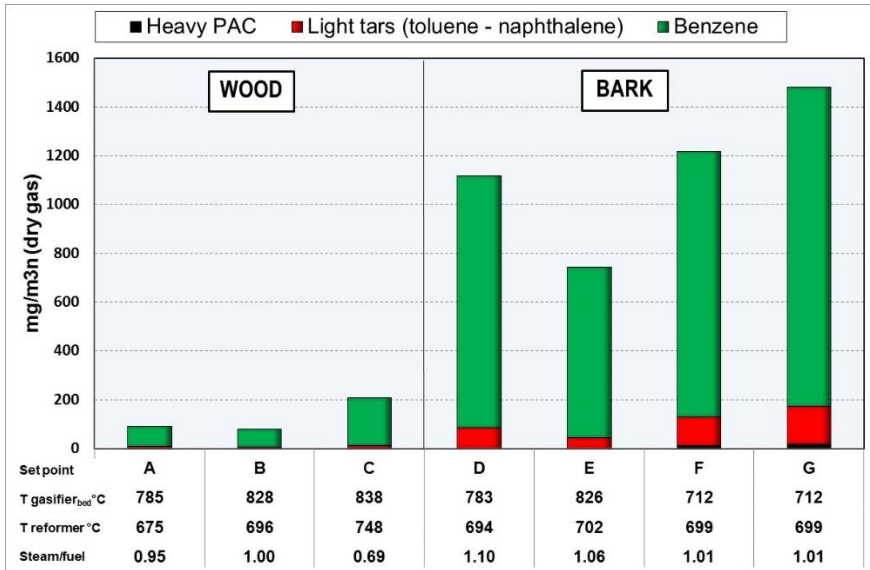


Figure 21. Measured tar contents after the reformer at different set points.

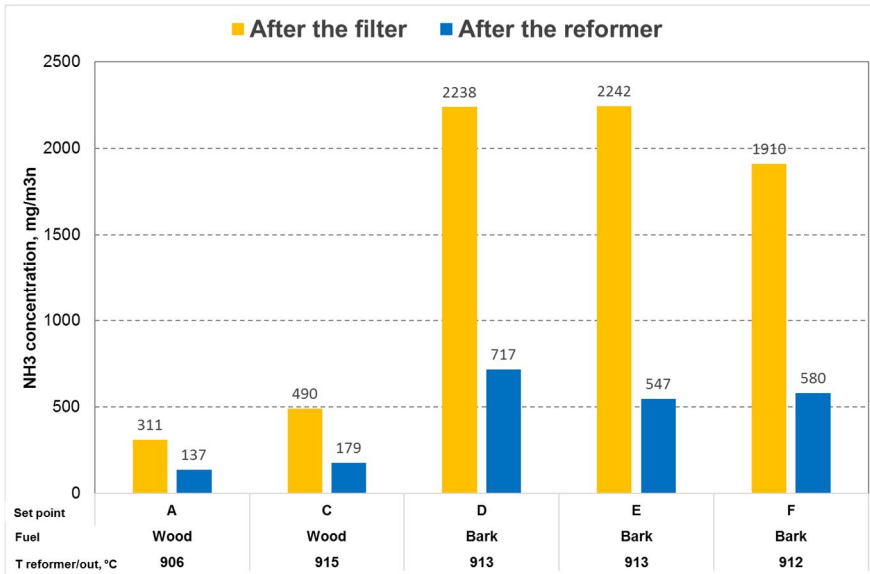


Figure 22. Concentration of nitrogen compounds after the filter and the reformer in gasification experiments.

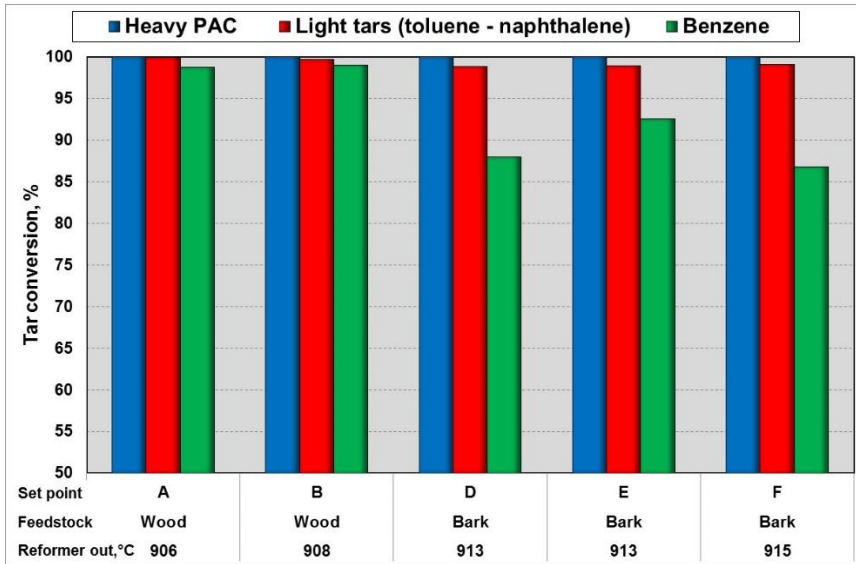


Figure 23. Tar conversion efficiencies in the reformer (from measured concentrations and material balances as $100 \times (\text{mass flow in} - \text{mass flow out}) / \text{mass flow in}$).

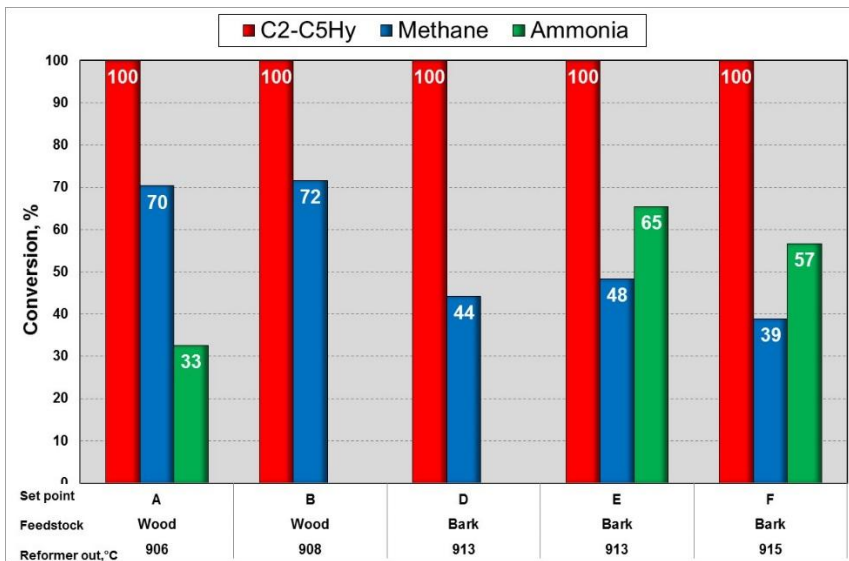


Figure 24. Conversion efficiencies of methane, C₂-hydrocarbon gases and ammonia at selected set points.

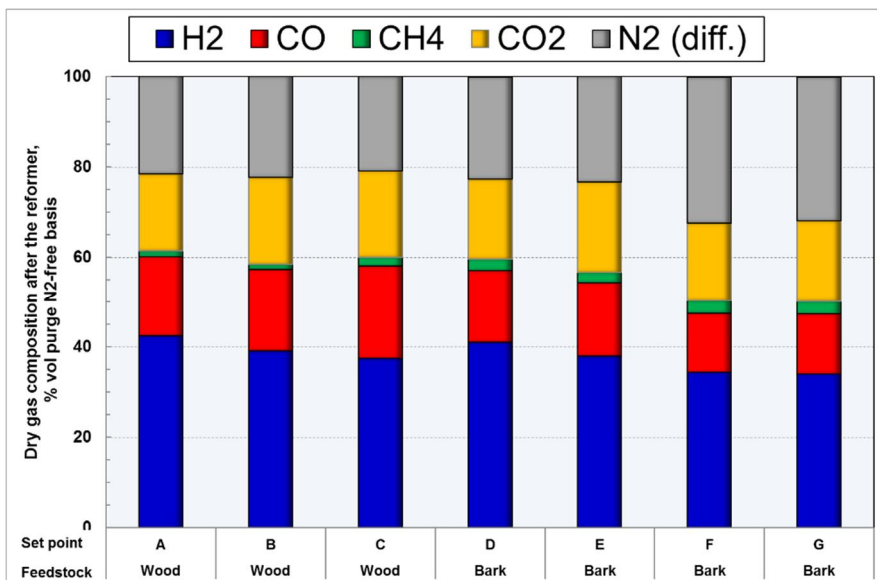


Figure 25. Dry gas composition after the reformer (vol-%, purge N₂-free basis)

4.4 Sulphur removal studies in slip-stream

Two alternative sorbent-based desulphurization schemes were tested in DFB test campaigns. The first sulphur removal concept was based on using low-cost, non-regenerable ZnO-based sorbents. In this concept, the used sorbent containing ZnO and ZnS would be sold to a zinc production plant, where the captured sulphur is recovered as sulphuric acid. The used ZnO sorbent would be a clean and good-quality feedstock for the zinc plant. A summary of the slip-stream tests with ZnO-based sorbents is presented in Figure 26.

Based on the results, the main challenge related to ZnO-based sulphur removal was the low adsorption capacity of the sorbent. The pores of the sorbent particles were blocked by the ZnS product and the adsorption was limited to the particle surfaces only thereby increasing the sorbent consumption. Consequently, ZnO-based sorbents are more suited for the final polishing step with inlet sulphur contents below 5 ppm-v.

In 2018, sorbent-based sulphur removal studies were continued as part of the EU project COMSYN [17]. The second sulphur removal concept, which is being evaluated in COMSYN, was based on a combination of activated carbon and ZnO-based sorbents: the bulk of sulphur is removed using active coke, while final removal to sub-ppm level is done using ZnO sorbents. Laboratory-scale tests were first carried out to study the adsorption capacity of different active coke products, and the slipstream facility was modified accordingly to test this two-stage sulphur removal approach. The new concept was successfully taken into operation in the

last test run of BTL2030 project in week 18 of 2018 and the results are summarized in Figure 27.

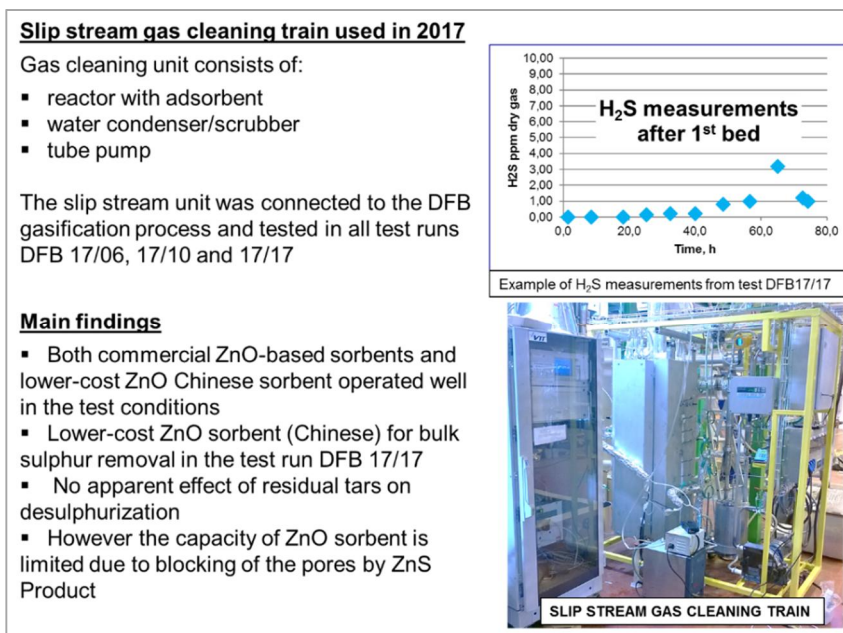


Figure 26. The slip stream final gas cleaning tests carried out in the 2017 DFB test runs to study the sulphur removal by ZnO sorbents.

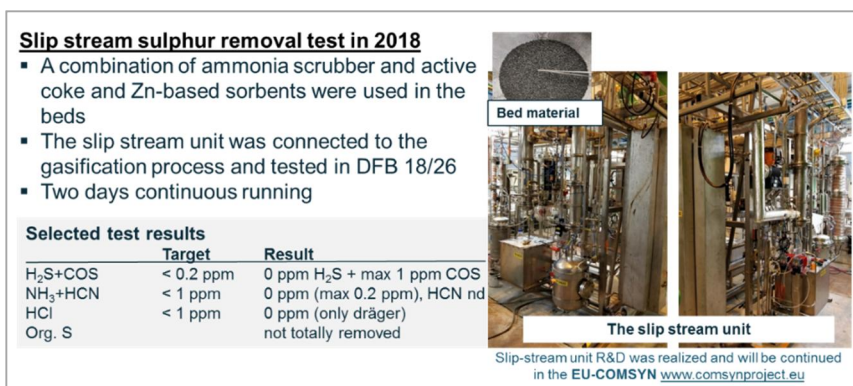


Figure 27. Summary on the slip stream final gas cleaning tests carried out in the DFB test runs of week 18 of 2018.

5. Estimated performances of industrial-scale production plants

5.1 Comparison of different gasification process designs

The experimental results obtained in DFB gasification test campaigns were introduced into a flowsheet process simulation model in order to evaluate the effects of different process parameters on the overall production efficiencies in industrial-scale BTL plants. The process model was set up as follows:

- Gasification performance was modelled to correspond the experimental data with respect to carbon conversion, yields of hydrocarbon gases and WGS ratio ($\text{H}_2 \cdot \text{CO}_2 / (\text{CO} \cdot \text{H}_2\text{O})$).
- Heat losses were assumed to be ca. 1 % of the biomass energy input (LHV basis).
- The need for external energy in the oxidizer was calculated from the energy balance of the gasifier and oxidizer - off-gases retrieved from the FT synthesis unit was used as supplementary feedstock in the oxidiser.
- Gas filtration was realized at gasifier exit temperature without intermediate gas cooling.
- Reformer was estimated based on three catalyst beds resulting in considerably higher conversions than achieved in DFB test runs:
 - Outlet temperature 900 °C
 - Tar + benzene conversion 99,5 %, C_2H_y conversion 100 %, $\text{CH}_4/\text{NH}_3/\text{HCN}$ conversions 80 %
 - Air or O_2 + steam were used as input gases
 - WGS equilibrium was assumed at the outlet temperature
- The FT synthesis performance was estimated with a process model used previously in VTT's UCG projects [6], [18].
 - CO conversion max. 90 %, C_{5+} selectivity 91 %
 - In some cases, the CO conversion had to be lowered in order to provide a sufficient amount of off-gas to the oxidizer (and/or to maintain the LHV of off-gas > 3.5 MJ/m³n to guarantee stable combustion)
 - The C_{5+} selectivity used in the model is probably too high for the small-scale FT systems as e.g. Velocys reports 88-89 % vs. 91% used in this FT model [19]. This results in too high FT efficiencies.
 - However, the comparison of the effects of gasification process parameters are still qualitatively valid.
- Energy and material balances were calculated for gasification of bark. The initial moisture content of the feedstock (before drying) was set to 50 % and after drying to 11 % (drying with 55 % efficiency).

- The power consumption of the production concepts was not estimated. Instead, it was assumed to amount to 10 % of the biomass energy input in all cases. In reality, power consumption is mainly borne from compression of the syngas from atmospheric pressure to FT synthesis pressure and from the oxygen separation in the concepts, where oxygen is used.
- The results are shown for 100 MW biomass input to the dryer.

The studied process alternatives, their key operating parameters and the results of calculations are presented in Figure 28.

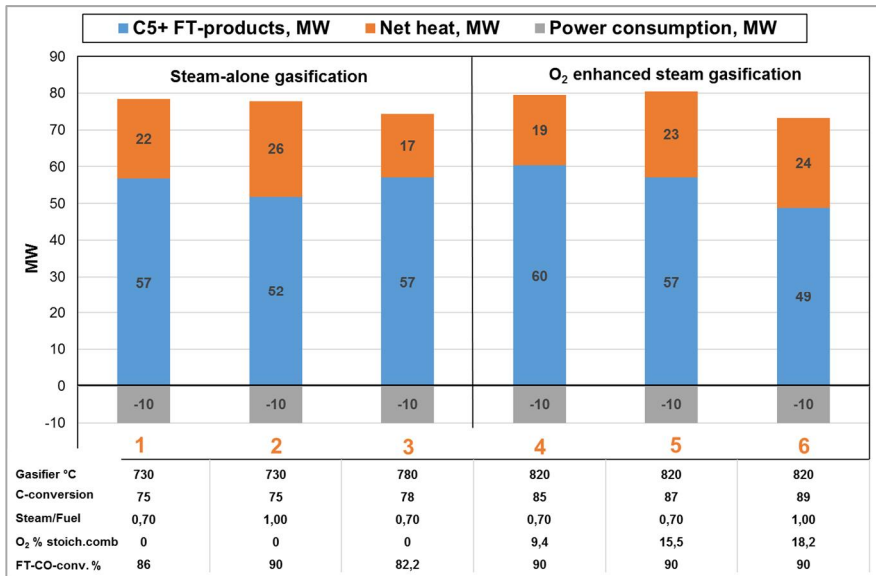


Figure 28. The effect of DFB gasification process parameters on the production of FT liquids and heat.

The **three cases of “Steam-alone gasification”** were realized so that air was used in the reformer and the key gasification parameters were the gasification temperature and the steam-to-fuel ratio. The first two cases carried out at 730 °C gasifier temperature and 850 °C in the oxidizer were modelled according to the results obtained in week 18/2018. In the first case, the steam-to-fuel ratio is 0.7, while in the second case it is increased to 1, which was also the case in the actual test runs. In the first case, the CO conversion of the FT unit had to be reduced to 86 % in order to maintain LHV > 3.5 MJ/m³n in the FT off gas. The reduction of CO conversion in the FT unit is not needed in the second case, because the H₂/CO molar ratio is 2.37, which means that larger part of hydrogen is not consumed, and consequently the LHV of off-gas is high enough. In the third case, the gasification temperature is raised to 780 °C corresponding to the tests of 2016-17, and the oxidizer is operated at 900 °C. It is interesting to notice that conversion efficiency to

FT liquids is the same in the first and the third case. Higher carbon conversion seems to compensate the increased demand for gasifier heating. In the third case, the CO conversion of the FT unit had to be limited to 82.2 % in order to produce enough off-gas for heating the oxidizer. The design of the first case is illustrated in the upper diagram of Appendix C.

Cases 4, 5 and 6 represent cases where also **some oxygen is introduced into the gasifier** to supply part of the heat required in gasification. In all these cases, the reformer is operated with a mixture of oxygen and steam instead of air. The oxygen feed in case 4 corresponds to 9 % of stoichiometric combustion, which is not sufficient to reach adiabatic operation, and thus part of the FT off-gas is led into the oxidiser also in this alternative. Cases 5 and 6 represent adiabatic operation, where there is no need for leading FT off-gas into to the oxidizer. An example of these process alternatives is illustrated in the lower diagram of Appendix C. The advantages of concepts 4-6 are that the syngas H₂/CO ratio is close to the ideal 2, and due to the lower inert gas content, the off-gas LHV is always high enough (4.1-5.4 MJ/m³n). Thus, there is no need for limiting the FT conversion efficiency.

In the adiabatic cases 5 and 6, the required steam-to-fuel ratio has a prominent effect on the yields of FT liquids. If this ratio is increased from 0.7 to 1.0, more products must be combusted in the gasifier with higher oxygen feed rates, and this reduces the chemical efficiency of the gasification stage rather dramatically.

The following conclusions can be drawn from these process evaluation studies:

1. High conversion efficiencies of biomass to FT liquids can be reached both with the steam-alone operation and with the oxygen-enhanced steam gasification.
2. In steam-alone gasification, similar conversion efficiencies to FT products can be achieved by operating the gasifier at 730 or at 780 °C. The low-temperature process alternative is not very sensitive to the steam-to-fuel ratio.
3. In the adiabatic operation mode with oxygen-enhanced gasification, the steam-to-fuel ratio has a critical role on the yields of FT products.
4. Finally, practical operation issues and process economics will define the optimum operation conditions.

5.2 Design of process concepts and feasibility studies

The technical and economic feasibility of the biofuel production concepts based on the DFB gasification process were assessed in WP2 of the BTL2030 project. These studies were initiated by designing industrial-scale production concepts for all target final products. The evaluation was based on mass and energy balances calculated using Aspen Plus process simulation software. Calculations were carried out for a 100-150 MW (fuel input) BTL plant that uses crushed bark as feedstock. The feedstock properties and the process design parameters concerning the front-end process (gasification, filtration and reforming), which are embedded in the Aspen Plus simulation model, were derived from DFB gasification experiments carried out

in 2016-2018. The performances of the back-end synthesis processes were taken from literature.

TEA studies of the BTL2030 project included the following concepts:

1. Bio-methane from solid biomass
 - Comparison of low-CapEx configuration with earlier SNG assessments
2. Refinery hydrogen
 - Case 4A: 150 MW biohydrogen + 150 MW hydrogen from natural gas
 - Case 4B: 300 MW gasifier for 50/50 biomass/fossil waste (energy basis)
 - Case 4C: 300 MW hydrogen from natural gas
3. FT fuels integrated to district heating steam power plant
 - Off-gases combusted in the oxidiser
 - Different dryer configurations: belt drying vs. highly efficient two-stage drying concept

The results of these studies will be presented and discussed more detailed in separate publications [20].

6. Preliminary plans for demonstration activities

6.1 General plan for industrial implementation

In Finland, more than 90 % of CO₂ emissions in transport sector originate from road transport. The longer-term national target set by the Finnish Government is to achieve practically CO₂ neutral passenger car traffic and to increase the share of biofuels in heavy-duty road transport to 70 % by 2050 [21].

In a study financed by the Ministry of Trade and Economics (TEM), VTT and VATT evaluated the effects of different CO₂ reduction alternatives on the national economy of Finland [22]. The target was set to reduce the emissions in road transport by 30-40 % by 2030 compared to the reference year of 2005. These different alternatives included electric cars, biofuels and hydrogen-based fuel cell cars. The main conclusion of the study was that drop-in biofuels produced from domestic raw materials, mainly wood residues, have a positive effect on the Finnish economy, while the overall effect of other alternatives is negative. In the case of domestic biofuels, the higher fuel costs of transport are compensated by the positive effects - the reduced oil bill and the increased employment, especially in the front end of the biofuel production chain. In addition, the technology export potential will be significant. According to Prof. Nils-Olof Nylund [23], the transport system of 2030 will require both electric cars and biofuels especially in heavy-duty traffic. The Swedish study "Fossilfri fordonsflotta" reached similar conclusions [24].

Figure 29 illustrates different alternatives and complementary pathways to reach the required production of 1 Mtoe of biofuels by 2030. Compared to the status in 2014, additional production capacity of 600 000 toe/a is required. The biotechnical routes and the HVO technologies have quite limited domestic raw material basis, while only thermochemical conversion routes can utilize the extensive resources of wood residues. In order to have a significant contribution to the national targets set for 2030-50, the BTL2030 project concepts should be ready for industrial implementation at around 2030. A general plan for the follow-on demonstration and flagship projects illustrated in Figure 30 was presented already in the BTL2030 project work plan. The principal idea of this plan is to lower the technical risks associated with the flagship plant by realizing an intermediate demonstration stage. In WP3 of the BTL2030 project, different demonstration alternatives were preliminary planned together with the industrial project partners.

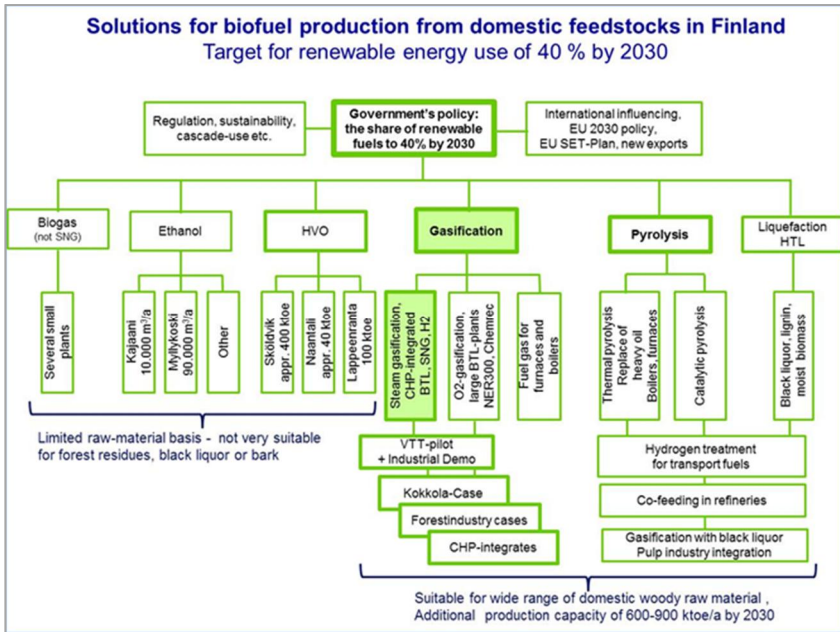


Figure 29. Different alternatives for the production of biofuels for the transport sector.

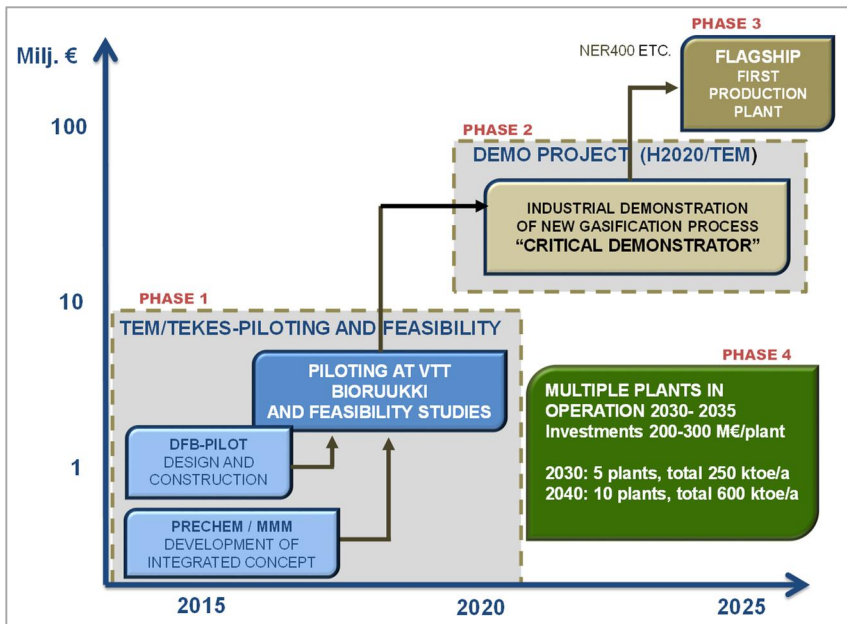


Figure 30. General plan for the industrial implementation of the BTL2030 process.

6.2 Industrial demonstration plant with synthesis testing in slip-stream

Two different alternatives for realizing the demonstration step were considered. The first alternative "Case A" is illustrated in Figure 31. In Case A, the gasification and hot filtration stages of the process are demonstrated at an existing site, where the product gas can be combusted in a boiler or an industrial kiln and a 1-5 MW slip-stream is extracted from the main product gas stream for final gas cleaning and synthesis testing. The basic idea is that after the demonstration period, the gasification unit would continue operation as a commercial fuel gas plant replacing fossil fuels. This type of demonstration has the following advantages:

- Major part of the investment can be realized as a commercial project and will be paid back by replacing fossil fuels in heat generation.
- The main gas line can be built with minimal risks as the plant can be converted back to air-blown CFB gasifier, which is a commercially proven technology.
- The size of 10-50 MW is sufficiently large for follow-on scale-up to 150 MW but, on the other hand, small enough to allow testing different biomass residues and waste-derived feedstocks.
- The total investment is reasonably low as the final gas cleaning and synthesis operation is demonstrated at a smaller scale.
- The demonstration can be divided into two or more projects, which may have different financing structures (combination of investment support and demonstration funding).

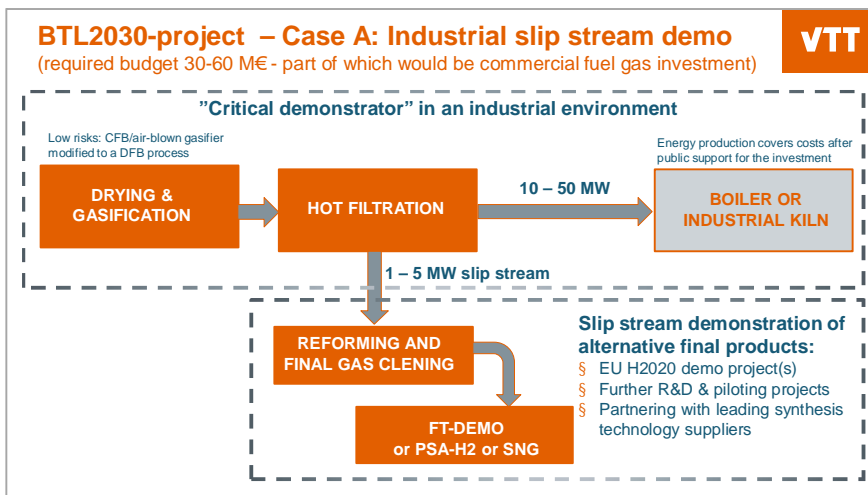


Figure 31. Case A: Industrial demonstration of the gasification and filtration technologies and slip-stream testing of final gas cleaning and synthesis.

In the WP3 of the BTL2030 project, two potential sites were already identified for this type of demonstration. The first alternative was the Kokkola industrial park, where the gasification plant would produce 12 MW filtered product gas to replace heavy fuel oil in industrial drying kilns. The second alternative was to integrate the gasification plant with the Vuosaari power plant site of Helen. In this case, a 30-50 MW gasifier would provide fuel gas for district heat production. On a system level, this heat would then replace coal-based district heat in the large district heating network of Helsinki.

It was concluded that this demonstration alternative can be realized if the following requirements are met:

- The investor of the primary fuel gas plant exists and has also interest in follow-on synthesis applications.
- Other parts of the BTL value chain are also represented in the consortium by companies, who are willing to invest on the project and can foresee profitable business from BTL.
- The fuel gas investment is economically acceptable for the investor, which will require that there is a real need to replace fossil energy and that some investment support can be received from the Ministry of Economic Affairs and Employment (TEM).
- Public financing is also required to the development and demonstration of the final gas cleaning and synthesis technologies in the slip-stream facility. This could be sought from national or EU RTD financing sources.

6.3 Demonstration of the whole production concept at VTT Bioruukki

If the required financing of 50-60 M€ cannot be secured for the demonstration “Case A”, a lower cost “Case B” was also preliminary planned (see Figure 32). In this case, VTT can be the main contractor who will construct and operate a small semi-industrial demonstration plant designed for 1-2 MW biomass input. The whole production concept, including fuel synthesis, would be designed for this biomass input capacity. The estimated cost for this alternative is 15-25 M€, which could be feasible also for the Innovation Actions of Horizon 2020. Industrial partners representing the biomass-to-biofuel value chain could then take part in the project with reasonable individual shares of 1-5 M€.

One of the advantages of this alternative is that the complete production concept would be operated as an integrated plant. The small scale of operation also enables testing and further development of various process alternatives. Moreover, the costs for preparing different feedstock qualities would be smaller than in “Case A” realised at larger scale. The disadvantage of “Case B” is, of course, the challenges related to scaling up from 1-2 MW to industrial scale of 100-150 MW.

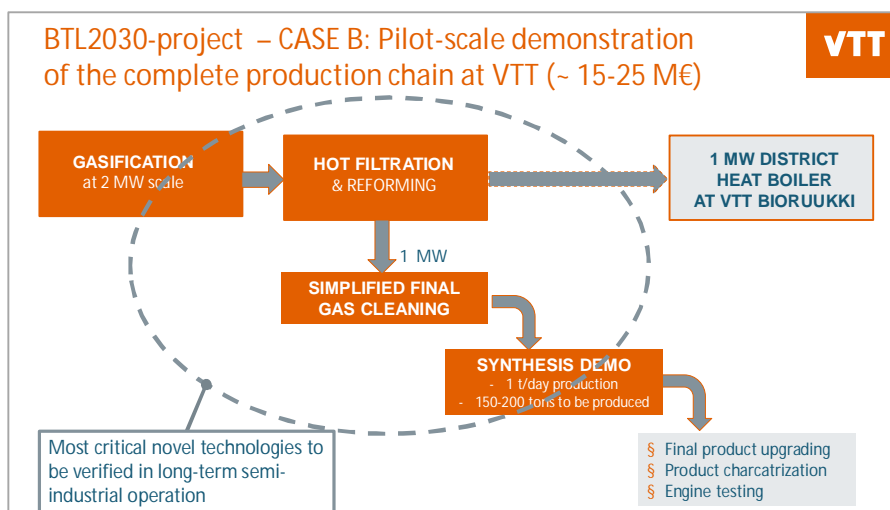


Figure 32. Case B: Demonstration of the complete integrated production concept at semi-industrial scale at the Bioruukki Piloting Centre of VTT.

7. Conclusions

The main conclusions of the BTL2030 project were the following:

- A new gasification process targeted for intermediate-scale production of synthesis gas was developed and tested at a small pilot plant.
- Different gasification operation modes, including steam-alone operation and oxygen-enhanced gasification, can be designed for different target applications.
- The DFB gasification process with a circulating fluidized-bed reactor as the gasifier can be realized both with a bubbling fluidized-bed and a circulating fluidized-bed oxidizer.
- Hot filtration of the product gas was successfully realized at gasifier exit temperature and the novel sintered metal filters operated reliably throughout the test runs.
- Technically the gasification process is ready for industrial-scale demonstration – however, a follow-up industrial project is required before commercial plants can be realized.
- Production concepts for FT liquids, bio-SNG and H₂ were designed and evaluated. Investment support is required to improve the economic competitiveness against the present HVO-based biofuels – this situation is, however, expected to change after 2030, as the demand for renewable fuels significantly increases and the raw material basis for HVO becomes the limiting factor.
- No follow-on demonstration projects were initiated immediately after the project, but plans for alternative routes for industrial implementation were created and assessed – including the preparation an IA proposal to H2020 together with Helen and the detailed economic assessment of a plant in Kokkola (within a parallel project of KOSEK).
- The development of gasification-based BTL technologies continues in three ongoing Horizon 2020 RIA projects (FLEXCHX, COMSYN and REDIFUEL).

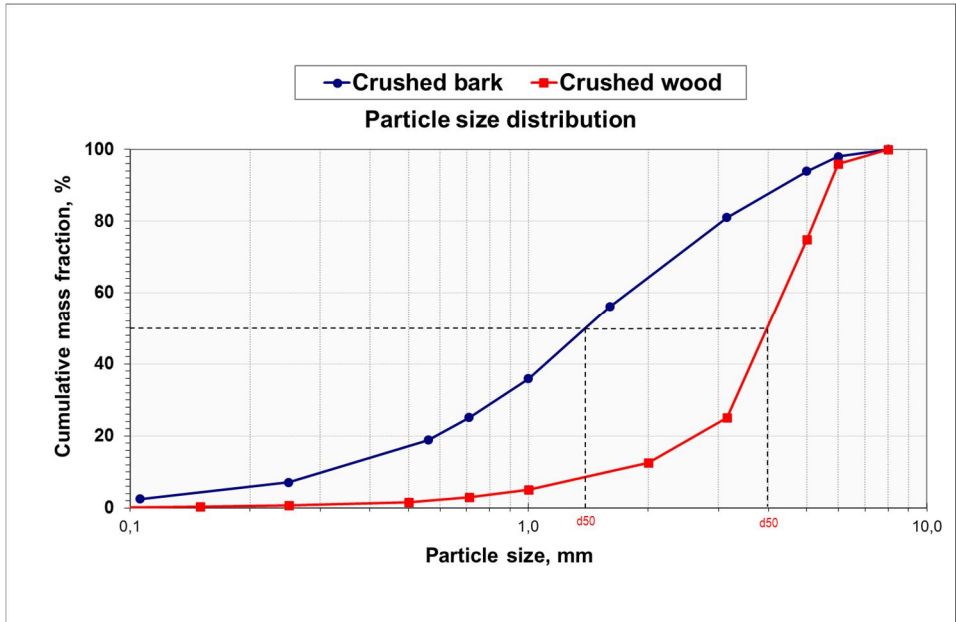
References

- [1] *The European Commission, A roadmap for moving to a competitive low-carbon economy in 2050. (2011). <http://eur-lex.europa.eu/legal-content/EN/TXT/PDF/?uri=CELEX:52011DC0112&from=EN>.*
- [2] Kurkela, E., Kurkela, M., Hiltunen I., "Steam-oxygen gasification of forest residues and bark followed by hot gas filtration and catalytic reforming of tars: Results of an extended time test," *Fuel Processing Technology*, Vols. 141, Part 1, pp. 148-158, 2016.
- [3] Kurkela, E., Kurkela, M., Hiltunen, I., "The effects of wood particle size and different process variables on the performance of steam-oxygen blown circulating fluidized-bed gasifier," *Environmental Progress and Sustainable Energy*, vol. 33 (3), pp. 681-687, 2014. <https://doi.org/10.1002/ep.12003>.
- [4] Simell, P., Hannula, I., Tuomi, S., Nieminen, M., Kurkela, E., Hiltunen, I., Kaisalo, N., Kihlman, J., "Clean syngas from biomass - process development and concept assessment," *Biomass Conversion and Biorefinery*, vol. 4, pp. 357-370, December 2014.
- [5] Kurkela, E., Simell, P., McKeough, P., Kurkela, M., "Synteesikaasun ja puhtaan polttokaasun valmistus," *VTT Publications 682*, p. 54 + app. 5 p., Espoo 2008, <https://www.vtt.fi/inf/pdf/publications/2008/P682.pdf>.
- [6] McKeough, P., Kurkela, E., "Process evaluations and design studies in the UCG project 2004-2007," *VTT Research Notes 2434*, p. 45, Espoo 2008, <https://www.vtt.fi/inf/pdf/tiedotteet/2008/T2434.pdf>.
- [7] Laatikainen-Luntama, J., Kurkela, E., "Air-blown CFB gasification of forest residues, demolition wood and crushed wheat straw pellets," *22nd International Conference on Fluidized Bed Conversion*, June 2015, Turku, Finland.
- [8] Kurkela, E., Kurkela, M., Moilanen, A., "Fluidised-bed gasification of high-alkali biomass fuels," *Science in Thermal and Chemical Biomass Conversion. Bridgwater, A. V.; Boocock, D. G. B. (eds.)*, vol. 1, pp. 662-676, Vancouver Island, Canada 30 Aug.-2 Sept. 2004.
- [9] Kurkela, E., Moilanen, A., Kangasmaa, K., "Gasification of biomass in a fluidised bed containing anti-agglomerating bed material," *European Patent Office, Bulletin 2003/41*, 2003, p. 10.
- [10] Tuomi, S., Kurkela, E., Simell, P., & Reinikainen, M., "Behaviour of tars on the filter in high temperature filtration of biomass-based gasification gas," *Fuel* 139, pp. 220-231, 2015 <https://doi.org/10.1016/j.fuel.2014.08.051>.
- [11] Kurkela, E., Ståhlberg, P., Laatikainen, J., "Pressurized fluidized-bed gasification experiments with wood, peat and coal at VTT in 1991 - 1992. Part

1. Test facilities and gasification experiments with sawdust," *VTT Publications 161*, p. 55+app. 2 p, Espoo, VTT. 1993.
- [12] Kurkela, E., Ståhlberg, P., "Air gasification of peat, wood and brown coal in a pressurized fluidized-bed reactor. 2. Formation of nitrogen compounds," *Fuel Processing Technology*, vol. 31, pp. 23-32, 1991.
- [13] Leppälahti, J., Kurkela, E., "The behaviour of nitrogen compounds and tars in fluidized-bed air gasification of peat," *Fuel* 70, pp. 491-497, 1991.
- [14] Biomass gasification – tar and particles in product gases – sampling and analysis, TC BT/TF 143 WI CSC 03002.4. Technical Specification, 2004.
- [15] Kurkela, Esa, "Formation and removal of biomass-derived contaminants in fluidized-bed gasification processes," *VTT Publications 287*, p. 47+app. 87 p., Espoo, 1996.
- [16] Ranta, J., Nieminen, M., Kurkela, E., "The significance of porosity of gasification filter dust.," *Pajares, J.A. & Tascon, J.M.D. Coal Science. 8th International Conference on Coal Science Proceedings. Elsevier Science*, pp. 1907-1910, 1995.
- [17] Compact Gasification and Synthesis process for Transport Fuels (Horizon 2020 EU funded international project), "www.comsynproject.eu," [Online].
- [18] McKeough, P., Kurkela, E., "Detailed comparison of efficiencies and costs of producing FT liquids, methanol, SNG and hydrogen from biomass," *15th European Biomass Conference, Berlin, Germany, 7-11 May, 2007*.
- [19] LeViness, S., Deshmukh, S., Richard, L., Robota, H., "Velocys Fisher-Tropsch Synthesis Technology -New advances on State-of-the-Art," *Topics in Catalysis*, vol. 57, no. 6-9, pp. 518-525, Apr. 2014.
- [20] Tuomi, S., Kurkela, E., Hannula, I., Berg, C.G., "The effect of biomass drying on the overall efficiency of combined production of transportation fuels and heat by steam gasification and ft-synthesis.," *EUBCE 2018. ETA-Florence Renewable Energies*, pp. 521-527, 2018 <https://doi.org/10.5071/26thEUBCE2018-2BO.14.1>.
- [21] Jääskeläinen, S. (LVM), "Tarve liikenteen päästöjen vähentämiseen ja käyttövoimavaihtoehdot maalla, merellä ja ilmassa (in Finnish)," VTT 2G 2020 Biofuels seminaari 26.5.2015, VTT Bioruukki, Espoo 2015.
- [22] Nylund, N-O., Tamminen, S., Sipilä, K., et al., "Tieliikenteen 40 %:n hiilidioksidipäästöjen vähentäminen vuoteen 2030: Käyttövoimavaihtoehdot ja niiden kansantaloudelliset vaikutukset," (*How to Reach 40% Reduction in Carbon Dioxide Emissions from Road Transport by 2030: Propulsion Options and their Impacts on the Economy*) *Research Report VTT-R-00752-15.*, pp. 105+ app. 204 p., Espoo 2015..
- [23] Nylund, Nils-Olof, "Tieliikenteen vaihtoehtoiset käyttövoimaratkaisut vuoteen 2030: Bio, sähkö vai molemmat? (In Finnish)," VTT 2G 2020 Biofuels seminaari 26.5.2015, VTT Bioruukki, Espoo 2015.

[24] "www.regeringen.se/rattsdokument/kommittedirektiv/2012/07/dir.-201278,
Fossiloberoende fordonsflotta - ett steg på vägen mot nettonollutsläpp av
växthusgaser," Dir. 2012:78, 2012. [Online].

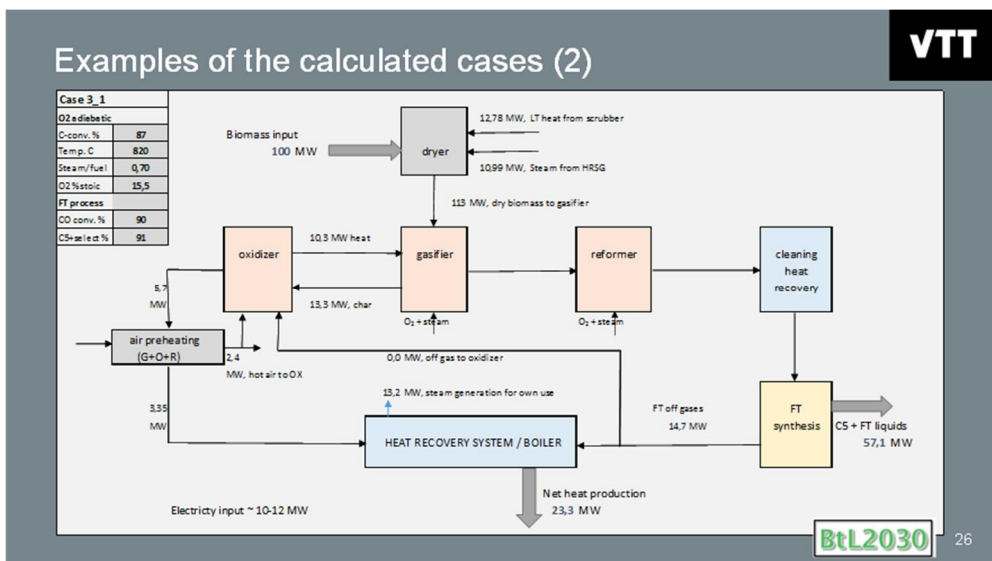
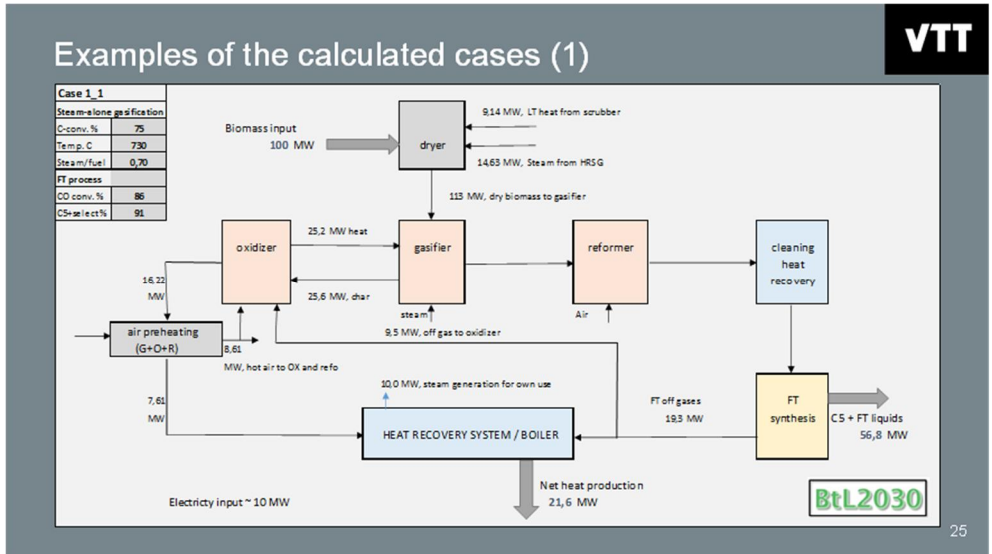
Appendix A: Particle size distributions of feedstocks



Appendix B: Examples of detailed tar analyses

GC ANALYSES FOR LIGHT TAR				
Set point: Fuel	After the filter	After the reformer	After the filter	After the reformer
	B Wood	B Wood	F Bark	F Bark
Concentration [mg/m3n]:				
Benzene	10410	72,7	11696	1088
Pyridine	0	0	146	0
1H-Pyrrole	10	0	0	0
Toluene	1892	0	3930	13
Ethylbenzene	18	0	4	0
m-Xylene	85	0	365	0
Ethynylbenzene	6	0	28	0
Styrene	470	0	961	0
o-Xylene	0	0	29	0
Benzaldehyde	0	0	0	0
Phenol	104	0	482	0
Benzonitrile	0	0	35	0
4-Methylstyrene	39	0	191	0
Indene	655	0	1558	0
o-Cresol	0	0	0	0
m+p-Cresol	0	0	10	0
Naphthalene	2464	6,9	3943	103
Quinoline	0	0	35	0
Quinazoline	0	0	0	0
Isoquinoline	0	0	0	0
1H-Indole	0	0	52	0
2-Methylnaphthalene	183	0	588	0
1-Methylnaphthalene	93	0	334	0
Biphenyl	145	0	210	0
2-Ethyl-naphthalene	0	0	0	0
1,6 Dimethylnaphthalene	0	0	22	0
Acenaphthylene	426	0	962	0
Acenaphthene	27	0	52	0
Dibenzofuran	68	0	116	0
Bibenzyl	0	0	4	0
2-Methyl-1-Naphthol	0	0	51	0
Fluorene	144	0	451	0
Phenanthrene	382	0	862	12
Anthracene	117	0	294	0
Carbazole	0	0	3	0
1-Phenylnaphthalene	0	0	11	0
2-Methylantracene	0	0	23	0
4H-Cyclopenta(def)Phenanthrene	31	0	110	0
Fluoranthene	111	0	247	0
Pyrene	90	0	262	0
2,3-Benzofluorene	6	0	55	0
1,1 Binaphthyl	0	0	0	0
1,2 Benzantracene	23	0	87	0
Chrysene	25	0	96	0
2,3 Benzantracene	0	0	0	0
Benzo(b)fluoranthene	6	0	36	0
Benzo(e)pyrene	0	0	33	0
Benzo(a)pyrene	7	0	70	0
Perylene	0	0	0	0
Indeno(1,2,3-cd) pyrene	0	0	16	0
Benzo(ghi)perylene	0	0	14	0
Anthanthrene	0	0	5	0
Coronene	0	0	0	0
TARS	7626	6,9	16784	128,1
TARS+BENZENE	18036	79,6	28480	1216,3
UNKNOWN COMPOUNDS	669	0	3734	0
C-%	92	92,4	92,3	92,4
H-%	7	7,6	7,3	7,6
N-%	0	0	0,1	0
O-%	0	0	0,3	0
Identified-%	96	100	88,4	100

Appendix C: Examples of the process concept evaluated by the flow sheet model



Title	Efficient use of biomass residues for combined production of transport fuels and heat
Author(s)	Esa Kurkela, Minna Kurkela, Sanna Tuomi, Christian Frilund & Ilkka Hiltunen
Abstract	<p>VTT Technical Research Centre of Finland Ltd has developed a new technique based on gasification, which offers a sustainable way to turn forest industry by-products, such as bark, sawdust and forestry waste, into transport fuels and chemicals. The new approach uses gasification to transform biomass into intermediate products – liquid hydrocarbons, methanol or methane – in production units that are integrated with communal district heating plants or forest industry power plants. The intermediate products are processed further in oil refineries to produce high-quality renewable fuels or chemicals. VTT developed and piloted the new gasification process and evaluated its competitiveness in the course of a recently concluded project called BTL2030. The distributed generation process developed by the project team makes efficient use of the energy content of biomass. Approximately 55% of the energy content is turned into transport fuels and another 20–25% can be used to provide district heating or to produce steam for industrial processes. The process is based on VTT's low-pressure, low-temperature steam gasification technology, simplified gas purification and small-scale industrial synthesis. Thanks to the small-scale approach, the heat generated by the process can be used throughout the year, and the process can be fuelled with local residues and wastes.</p> <p>The process development was realized using the pilot-scale Dual Fluidized-Bed (DFB) gasification test rig located at the Bioruukki Piloting Centre of VTT. Different DFB gasification alternatives were tested, including operation with two CFB (Circulating Fluidized-Bed) reactors and with a combination of a CFB gasifier and a BFB (Bubbling Fluidized-Bed) oxidizer. Both steam-alone gasification and gasification with a mixture of steam and a small amount oxygen/air were tested. The oxidizer was fluidized with air. The raw gasifier gas was purified in a hot filter unit and the hydrocarbon gases and tars were reformed in a catalytic reformer. Final gas cleaning was tested in a smaller slip-stream test facility. The DFB gasification process was operated for a total of 400 hours under gasification conditions using forest residues, bark, crushed wood pellets, used wood and crushed straw pellets as feedstock.</p> <p>This report is mainly focused on presenting the results and experiences of the experimental process development. In addition, the effect of the gasification process performance on the overall efficiencies of industrial production plants was estimated using an Excel-based flowsheet model. The results of the more detailed techno-economic evaluations carried out in the project is presented in separate publications.</p>
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