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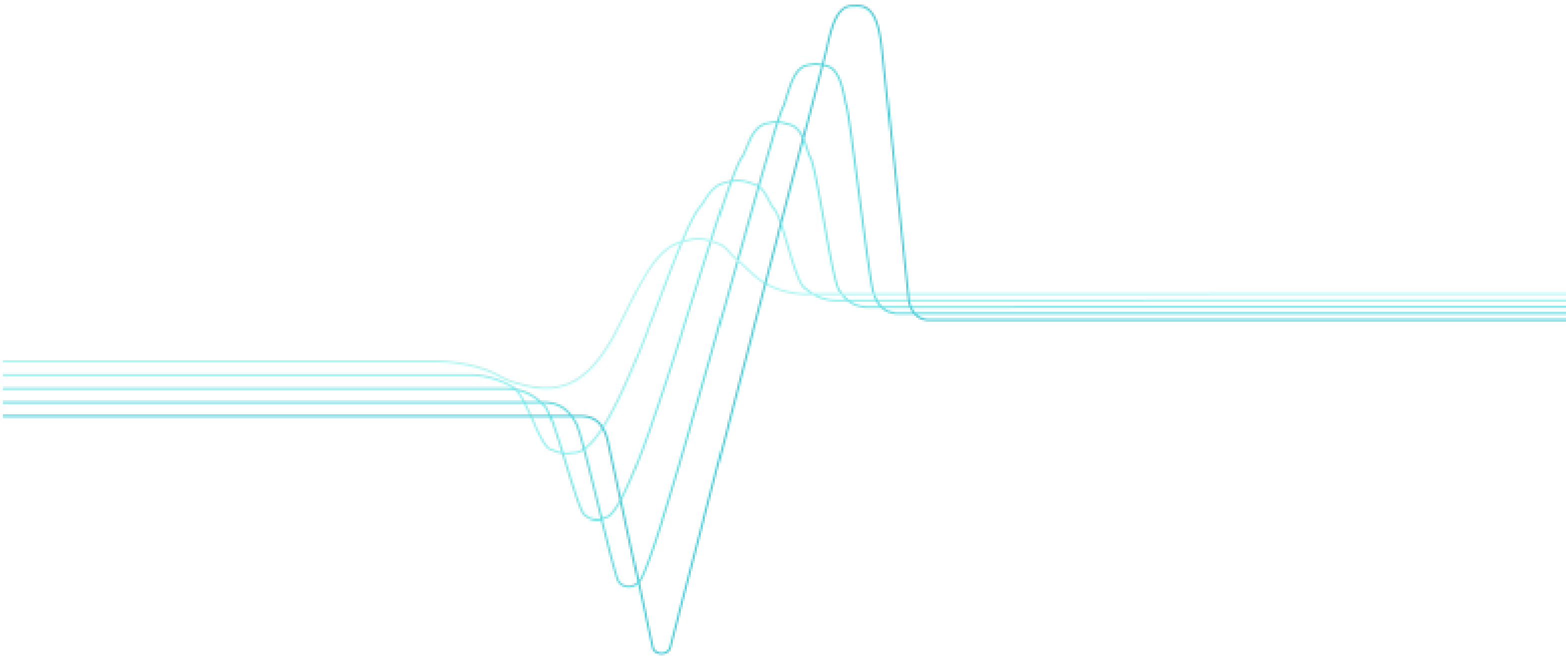

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Keywords synthesis-gas, biomass, process-evaluation

Abstract

The results of the process-evaluation work carried out in the project, Development of Ultra-Clean Gas (UCG) Technologies for Biomass Gasification, are presented in the publication. The UCG project was directed towards the development of innovative biomass gasification and gas-cleaning technologies for the production of ultra-clean synthesis gas.

The process-evaluation work in the UCG project covered a number of topics, including:

- process-configuration screening
- production-chain screening
- techno-economic evaluations of plants producing FT liquids, methanol, SNG or hydrogen; processes simulated with Excel-based codes
- techno-economic evaluations of competing technologies / production chains
- evaluations of benefits of integration; including novel concepts
- design studies of front-end operations by Foster-Wheeler and Pohjolan Voima.

The results of the work have proved useful in:

- developing and improving the UCG concept for bio-syngas production
- planning the experimental R&D work in the UCG project
- providing input data for sustainability analyses, CO_2-balance calculations, etc. (carried out elsewhere)
- compiling business plans and developmental strategies.

Preface

The research described herein was carried out during the period 2004–2007 as part of the project entitled “Development of Ultra-Clean Gas (UCG) Technologies for Biomass Gasification”. This research project, generally known as the UCG project, was led by Esa Kurkela of VTT Technical Research Centre of Finland. The main funding organisation was Tekes – the Finnish Funding Agency for Technology and Innovation. Additional funding was provided by VTT and by the following industrial companies: Neste Oil, Foster-Wheeler, Andritz, Metsä-Botnia, M-real, Pohjolan Voima, Stora-Enso, UPM and Vapo. As well, two of these companies – Foster-Wheeler and Pohjolan Voima – directly participated in the work of this subproject. The authors particularly wish to acknowledge the work contributions of these two companies.

On the basis of the results of this subproject, the following five papers were published:

McKeough, P. and Kurkela, E., Comparison of the performances and costs of alternative applications of bio-syngas, 14th European Biomass Conference, Paris, France, October 17–21, 2005.

McKeough, P. and Saviharju, K., Advances and possibilities in the utilisation of black liquor and other pulping by-products, ABTCP-PI 2005 Conference, Sao Paulo, Brasil, October 17–20, 2005.

McKeough, P. and Saviharju, K., Enhancing production of energy, power and fuels in conjunction with chemical recovery and simultaneously reducing green-house gas emissions, 2007 International Chemical Recovery Conference, Quebec City, Canada, May 9 – June 1, 2007.

Saviharju, K. and McKeough, P., Integrated forest biorefinery concepts, PulPaper 2007, Helsinki, Finland, June 5–7, 2007.

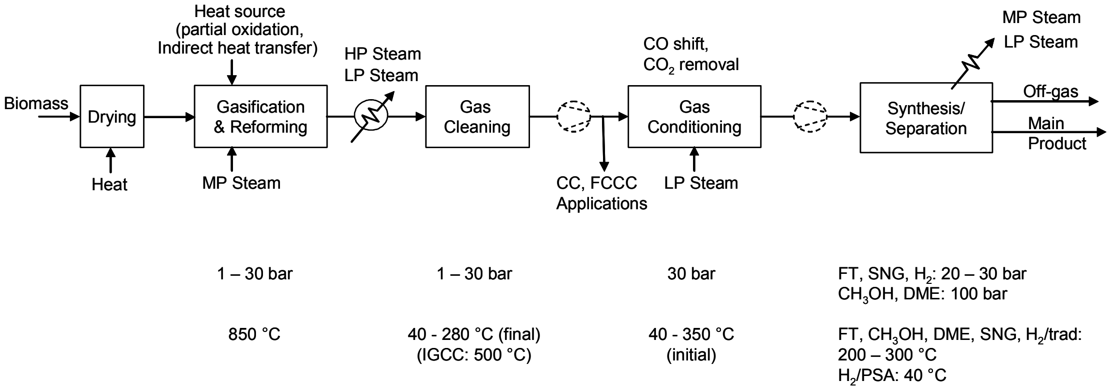
McKeough, P. and Kurkela, E., Detailed comparison of efficiencies and costs of producing FT liquids, methanol, SNG and hydrogen from biomass, 15th European Biomass Conference, Berlin, Germany, May 7–11, 2007.

These publications are alluded to in the following publication.

Contents

1. Introduction

A general layout of a process for producing and converting biomass-derived synthesis gas is given in Figure 1. Ranges of process conditions are indicated. A variety of alternative synthesis-gas derivatives may be produced.



Abbreviations:		
	CC	Combined-Cycle energy production
	FCCC	Fuel-Cell Combined-Cycle energy production
	FT	Fischer-Tropsch liquid fuels
	SNG	Synthetic Natural Gas (CH_4)
	CH_3OH	Methanol
	DME	Di-Methyl Ether
	H_2/trad	Traditional process for H_2 production from syngas
	H_2/PSA	H_2 production based on separation by Pressure Swing Absorption

Figure 1. General layout of process for producing and converting biomass-derived synthesis gas.

The process-evaluation work was carried out in five stages as outlined below:

Stage 1, Autumn 2004:

- Process-configuration screening, mainly on basis of published information and data
- Development and application of a syngas-conversion model
- Production-chain screening for Finnish conditions: plant capacity, benefits of integration with paper mill, alternative chain based on pyrolysis oil as an intermediate.

Stage 2, Spring 2005:

- Development of Excel-based simulation codes for methanol and FT plants and procurement of more reliable cost data
- Application of Excel-based codes: comparison of different process variants for the gasification-reforming train; reforming and shift reactions based on chemical equilibrium.

Stage 3, Autumn 2005:

- Improvement of Excel-based codes: reforming reactions from VTT experimental data; performance data for FT synthesis from the literature
- Application of Excel-based codes: effect of gasification pressure on costs; comparison with Choren process; comparison with Chemrec black-liquor gasification process.

Stage 4, Spring 2006:

- Providing input data to Foster Wheeler for preliminary design of gasifier and of primary gas-cleaning train (cooler, reformer, shift-reactor)
- Detailed evaluations of benefits of integration with pulp and paper manufacture.

Stage 5, Autumn 2006:

- Development of Excel-based simulation codes for SNG and hydrogen plants and detailed comparison with FT and methanol plants
- Evaluation of modified black-liquor processing to facilitate integration with stand-alone pulp mills
- Design studies by Foster Wheeler (pre-treatment, gasifier and primary gas-cleaning) and Pohjolan Voima.

The presentation of results in this publication basically follows the chronological order of the activities (above). However, in some cases, later results are introduced at an earlier stage to improve the logical structure of the presentation.

2. Screening of alternative process configurations

The main issues addressed when screening and narrowing down various process configurations are indicated in Figure 2. These different issues interrelate with each other, which renders the screening of alternative process configurations a rather challenging task. For example, a regenerative absorption operation becomes an attractive option for removing impurities such as H_2S, if the simultaneous removal of CO_2 is beneficial for the synthesis step.

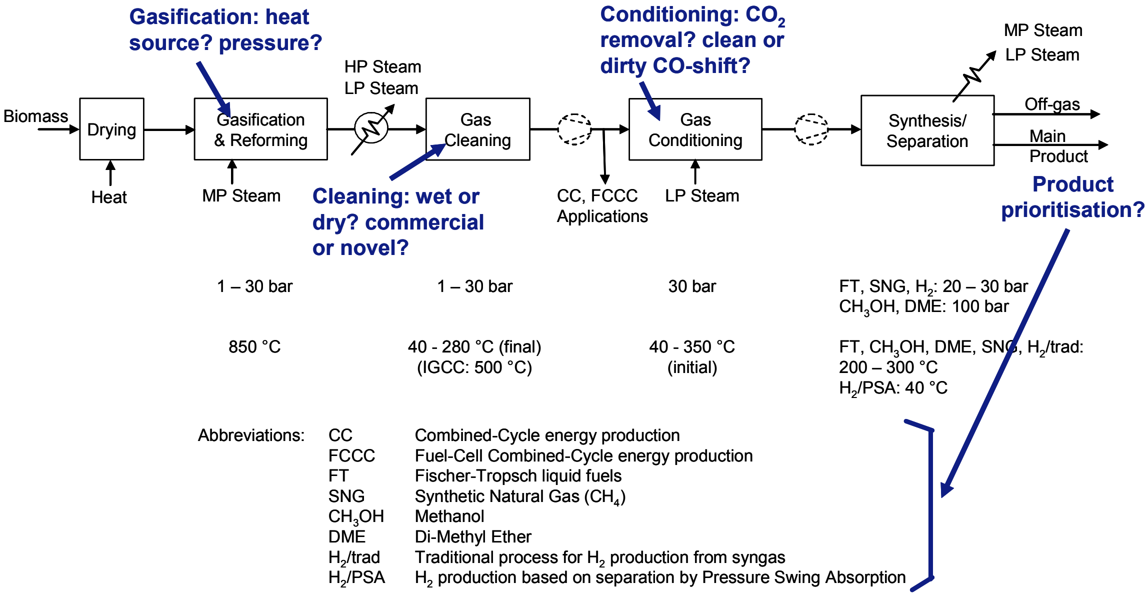


Figure 2. General process layout with the main issues addressed when screening alternative process configurations.

2.1 Comparison of alternative sources of heat for gasification

The alternatives evaluated were:

- partial oxidation with air
- partial oxidation with oxygen
- indirect heat transfer (twin reactors; circulating bed).

Three initial deductions were made:

- Employing air as the oxidant for partial oxidation introduces a large amount of nitrogen into the synthesis-gas stream, which has negative impacts on downstream gas-conditioning and synthesis steps. Thus, only if an air-blown gasification train, beginning at the gasifier and ending at the compressor, is significantly more economical than an oxygen-blown train should further consideration be given to the use of air as an oxidant in the gasifier.

- Indirect twin-bed gasification systems are challenging to operate and the challenge increases steeply with increase in pressure. In consultation with VTT's gasification experts, it was concluded that operating pressures greater than 5 bar would not be realistic for the twin-bed system.

- For the indirect system, it was not assumed that all heat would be supplied indirectly. The reason is that the UCG concept for synthesis-gas production includes an autothermal gas-reforming step using air or oxygen as the oxidant. When partial oxidation is utilised in one part of the gasification train, it makes no sense to rigidly deny the use of any oxidant in the primary gasification step. In the UCG evaluation work, it was assumed that the gasification-reforming process exploiting indirect heat transfer would receive about two thirds of the required gasification heat indirectly, the rest being generated by partial oxidation.

Key results of the assessment of alternative gasification heat sources are given in Figure 3. For various cases, the estimated costs of supplying oxidant to the gasification train (to gasifier and reformer) and of compressing the cooled product gas (before gas conditioning and synthesis) are given as a percentage of the production costs of the syngas derivative – in this case, Fischer-Tropsch (FT) liquids. The error bars indicate the uncertainty attached to the estimates of the equipment costs for these particular compressors. However, this uncertainty does not measurably affect the conclusions because all the cases are similarly affected. For example, if the high end of the error bar is the reality for one case then the high ends of the error bars will be the realities for all the cases in Figure 3.

In accordance with the initial deductions (above), the cases in Figure 3 exploiting indirect heat transfer employ (1) partial oxidation to cover about one third of the total heat requirement for gasification and reforming and (2) are restricted to a maximum gasification pressure of 5 bar. From the results in Figure 3, the following conclusions can be drawn:

- In terms of the sum of the costs presented in Figure 3, there are no significant differences between the oxygen-blown and the corresponding air-blown processes. It was initially deduced that, only if an air-blown gasification train is significantly more economical than an oxygen-blown train should further consideration be given to the use of air as an oxidant in the gasifier. Thus, in the light of the results given in Figure 3, it was decided to exclude air-blown gasification as an alternative to be further explored and developed in the UCG project.

- In terms of the costs given in Figure 3, configurations employing indirect heat transfer in addition to partial oxidation offer significant cost benefits in comparison to those employing partial oxidation only. On the other hand, twin-bed gasifiers are more complicated and expensive to build and operate than single-bed gasifiers based on partial oxidation, only. In parallel with this study, VTT and Foster-Wheeler gasification experts carried out an evaluation of alternative gasifier reactor technologies. The outcome of this evaluation, reported in detail in the other part of the final report of the UCG project, was that preference should be given to the oxygen-blown single-bed gasifier. However, it should be emphasised that this decision does not exclude the twin-bed gasifier as a subject of a longer-term R&D effort. In fact, one can envisage that a successful oxygen-blown single-bed gasifier could, at a later stage, be modified into a twin-bed gasifier, thus enabling part of the heat load to be satisfied by indirect heat transfer.

- According to the estimates of Figure 3, considerable cost savings could be attained by increasing gasification pressure. However, there are other impacts of elevated pressure on costs. These impacts, which offset some of the differences evident in Figure 3, were taken into account at a later stage in the project after results of certain bench-scale experiments had become available. Results of the later evaluation of the effects of gasification pressure on costs are presented in the next section.

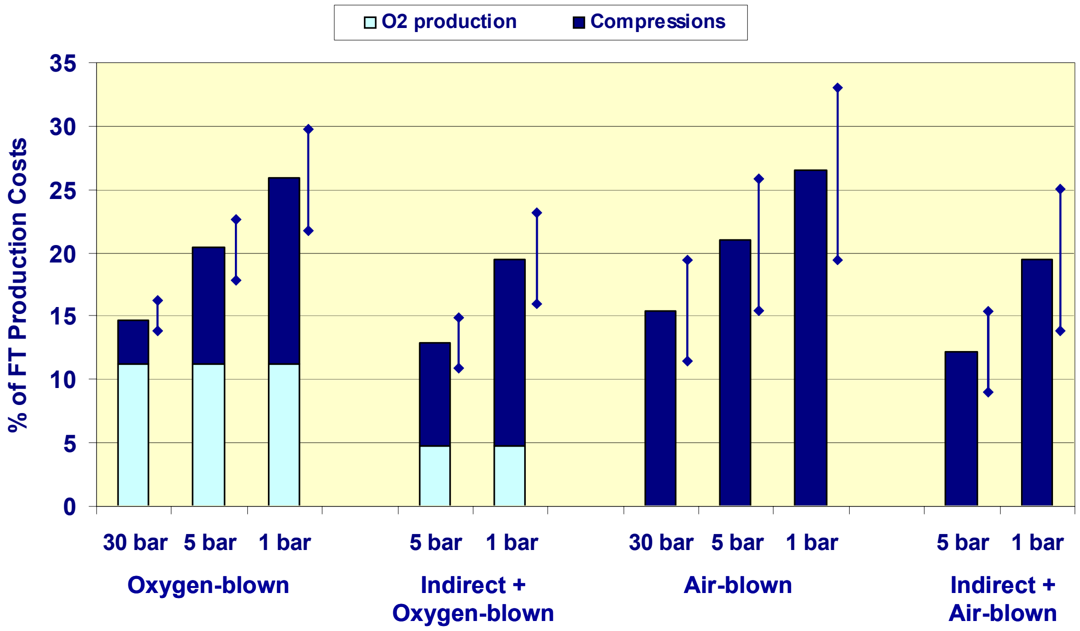


Figure 3. Estimated contributions to costs of FT liquids of (1) costs of producing O_2 *(when applicable) and (2) costs of compression of oxidant to gasifier pressure and of product gas to 30 bar for down-stream processing. Error bars indicate uncertainty attached to equipment costs for compressors.*

2.2 Evaluation of impact of gasification pressure on costs

The initial results (Figure 3) indicated a significant saving in costs should the gasification pressure be increased from 1 bar to 30 bar. However, later on, results of bench-scale experimental investigations of the gas-reforming and shift operations indicated that pressure would have a considerable impact on the costs of those operations, as well. The shift process in question is the so-called "dirty shift" process, which may be applied to a synthesis gas before removal of sulphur-containing impurities.

For most pieces of equipment in the gasification train, the effect of elevating the pressure on the capital costs would not be large: wall thickness would increase but vessel volume would decrease. For the reforming and shift operations, however, an increase in pressure will lead to a significant increase in the costs of the equipment. This is a consequence of the fact that, according to the results of bench-scale experiments performed in the UCG project, the required volumes of the reforming and shift reactors will not decrease measurably with increase in pressure. So, overall, the dependence of the production costs on pressure will be smaller than that indicated in Figure 3.

In the later evaluation, the effect of pressure on the equipment costs of the reforming and shift operations was taken into account. In this evaluation, the costs of producing methanol from biomass were estimated in each case. By that time, VTT had compiled detailed Excel-based models of synthesis-gas based processes for producing either methanol or FT liquids from biomass (refer Section 3.1, below). The effect of pressure on the cost of a piece of equipment can be described by the following formula: $K_2 = K_1*(P_2/P_1)^n$, where K_2 is the cost at pressure P_2, K_1 is the cost at P_1, and n is the pressure exponent. For many pieces of equipment in the gasification train, the effect of pressure on the equipment cost will be minor; i.e. the exponent n will have a value near zero. On the basis of the experimental results, it was estimated that, for the reforming and shift steps, the value of n would be somewhere in the range 0.2 to 0.4. For example, if the pressure were doubled, an exponent of 0.2 would mean a cost increase of about 15%, while an exponent of 0.4 would mean an increase of about 30%. These increases may seem moderate in view of the above conclusion that the volumes of these reactors would not measurably decrease with increase in pressure. On the other hand, these reactors have rather complicated internals, which lower the dependency of the cost on pressure. In this study, n was varied over the range 0.2–0.4. Another key variable was the level of equipment costs for compressors. The range of uncertainty was the same as in the earlier study (Figure 3).

The results obtained for one value of the pressure exponent (0.2) and for one level of compressor cost (highest) are shown in Figure 4. The estimated dependency on gasification pressure is quite moderate, with only limited benefit to be obtained by increasing the pressure beyond about 5 bar. The case in Figure 4 can be considered as a best case when seeking cost savings through elevation of pressure. For either a higher value of the pressure exponent or a lower level of compressor cost, the curve levels out, or even begins to swing upward, after a pressure of about 5 bar. These results were used as one of the bases for selecting a pressure range of 5–10 bar for the UCG gasification process. The other support for this pressure range came from the evaluation of alternative gasifier reactor technologies, referred to above and reported in the other part of the final report of UCG project.

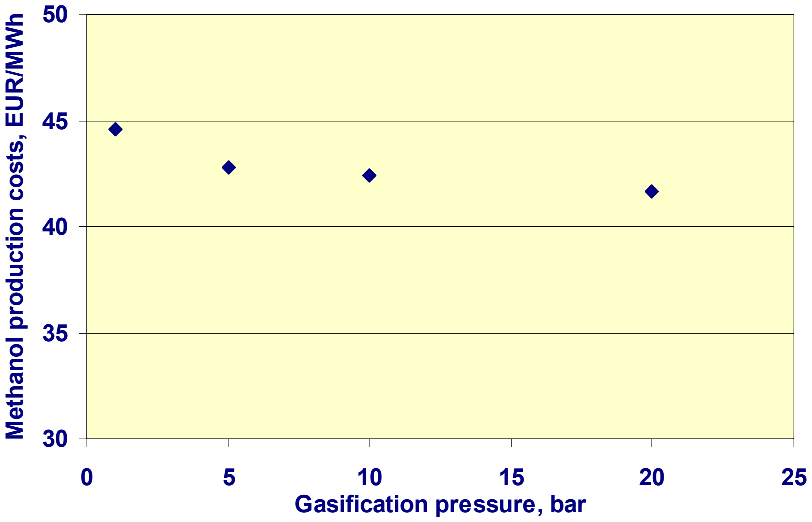
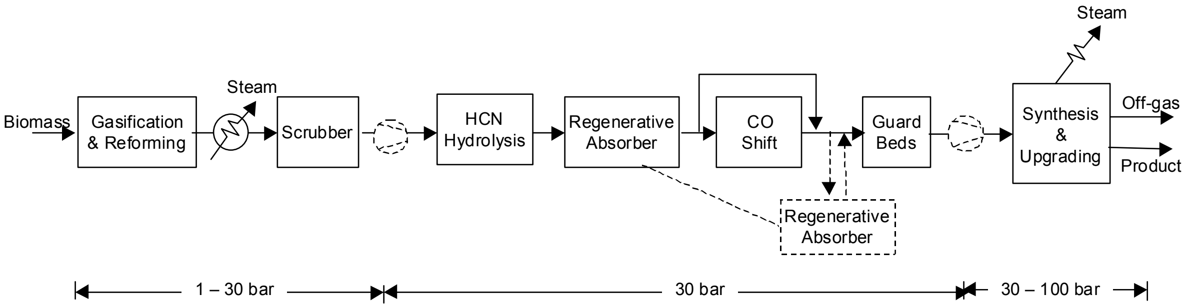


Figure 4. Estimated production costs of methanol as a function of pressure. Pressure exponent of 0.2 for equipment costs of gas-reforming and shift steps. Highest level of estimated equipment costs for compressors.



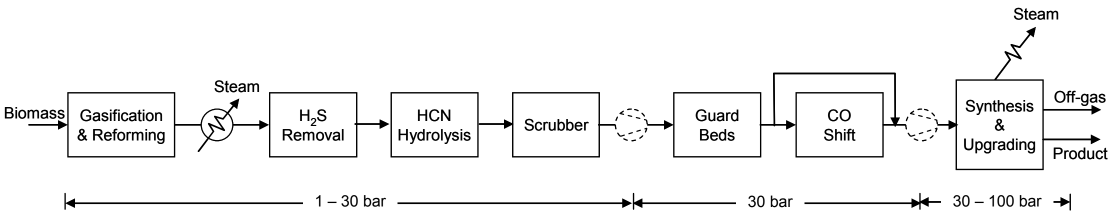
Scrubber:	Absorbs NH_3 to final level and HCl to guard-bed level
HCN Hydrolysis:	Converts HCN (and COS) to NH_3 (and H_2S)
Regenerative Absorber:	Commercial solvent-based or solvent-amine-based acid-gas removal process, e.g. Rectisol, Selexol, Sulfinol; removes H_2S, HCN (and COS) to guard-bed level, as well as most of the CO_2
Applicability:	Applicable as such to all applications Particularly suitable for SNG and H_2/trad, where CO_2 removal is a must at some point, and for FT-liquids, CH_3OH and DME, where inerts removal is favourable FCCC: NH_3 removal (scrubber) unnecessary

Figure 5. Process scheme showing main steps of gas-cleaning alternative based on regenerative absorption (commercial processes).

2.3 Gas-cleaning alternatives

Commercial wet gas cleaning systems are available for all the synthesis-gas applications considered in the UCG project. These are based on regenerative absorption of acid-gas components, in particular of CO_2 and H_2S. Figure 5 depicts the general scheme for gas cleaning based on regenerative absorption. These systems can handle coal-derived gases, which are more demanding than biomass-derived gases. On the other hand, these commercial systems have relatively high investment and operating costs. For this reason, two other types of clean-up systems, proposed elsewhere, were evaluated when screening alternative process configurations in the UCG project, namely:

- wet cleaning based on simple scrubbing at several different pHs
- mid-temperature dry cleaning.

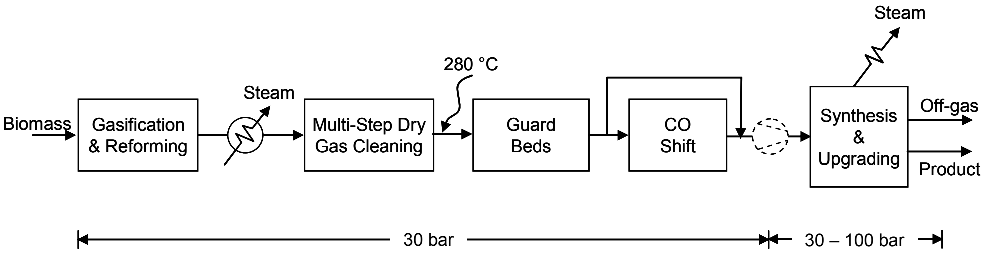


H_2S Removal:	H_2S content lowered from 100 ppm level to 1 ppm level e.g. absorption on solids either in bed (350 °C) or in slurry phase (20 – 200 °C)
HCN Hydrolysis:	Converts HCN (and COS) to NH_3 (and H_2S)
Scrubber:	Absorbs NH_3 to final level and HCl to guard-bed level
Applicability :	Once-Through FT, CH_3OH, DME, but high partial pressure of CO_2 makes synthesis less efficient SNG, H_2/trad: not promising (H_2S, HCN removal in conjunction with mandatory CO_2 removal) H_2/PSA: applicable, H_2S Removal and HCN Hydrolysis not required (H_2S-resistant shift catalyst) FCCC: applicable, NH_3 removal unnecessary, HCN?

Figure 6. Process scheme showing main steps of gas-cleaning alternative based on novel wet-gas cleaning without regenerative absorption of CO_2.

The first of these, proposed by research groups from the Netherlands (e.g. Hamelinck 2004), was shown to be unworkable. The basic flaw in this scheme was the unwarranted assumption that H_2S could be absorbed by alkali without CO_2 being simultaneously absorbed to a significant degree. A large extent of CO_2 absorption means an exorbitant consumption of alkali. VTT modified the scheme to include a separate step for selectively removing H_2S. This scheme is shown in Figure 6. Note that, because CO_2 is not removed at all from the product gas in this scheme, the range of applicability is restricted in comparison to that of the alternative based on regenerative absorption (Figure 5). Furthermore, the presence of a significant amount of CO_2 in the final synthesis gas would adversely affect the synthesis process in all cases. In particular, it should be mentioned that one of the main applications being explored in the UCG project; viz., synthesis of FT liquids with recycle of FT off-gas to the reformer, would not be feasible with the gas-cleaning option of Figure 6.

The second of the above alternatives – mid-temperature dry cleaning – has been proposed by researchers at the Gas Technology Institute (GTI) in the USA (Gas Technology Institute 2002). In the present evaluation, this scheme, shown in Figure 7, was deemed to be technically highly uncertain, restricted in application and providing only a moderate improvement in thermal efficiency. Furthermore, this last mentioned improvement is only realised when the gasification step operates at the same pressure level as the synthesis step. Thus, this gas-cleaning option is not compatible with the preferred pressure level of the UCG gasification process; viz. 5–10 bar (refer previous section). As in the previous case, it should also be mentioned that one of the main applications being explored in the UCG project is synthesis of FT liquids with recycle of FT off-gas to the reformer and that this application would not be feasible with the gas-cleaning option of Figure 7.



Multi-Step Dry Gas Cleaning:
Sorbent-based process for removal of H_2S and HCl to below 1 ppm level; sub-processes for NH_3 and HCN removal largely undeveloped as yet

Applicability:
Once-Through FT, but high partial pressure of CO_2 makes synthesis less efficient
CH_3OH, DME: not promising because gas has to be cooled before compression (100 bar) in any case
SNG, H_2/trad: not promising (H_2S, HCN removal in conjunction with mandatory CO_2 removal)
H_2/PSA: not promising because H_2S removal is unnecessary and PSA temperature is low (40 °C)
FCCC: applicable as such, if FC tolerates HCN

In general, Dry Gas Cleaning probably not competitive when gas has to be, in any case, cooled below 280 °C, e.g. when gasifying at lower pressures (gas compression required)

Figure 7. Process scheme showing main steps of gas-cleaning alternative based on novel mid-temperature dry-gas cleaning.

Some preliminary cost data were obtained for the three gas-cleaning alternatives. These are presented in Figure 8. The data for regenerative absorption (Rectisol-type process) and for dry gas cleaning were obtained directly from a GTI report (Gas Technology Institute 2002). Bearing in mind the technical uncertainties of dry gas cleaning, as well as its limited applicability, there would seem to be insufficient economic incentive to develop this option (Figure 8). That, at least, was the conclusion drawn in the UCG project. The other results in Figure 8 relate to two options for the H_2S removal step in the novel wet gas cleaning scheme depicted in Figure 6. These results suggest that there would be economic justification for further R&D work on this novel gas-cleaning alternative. However, in view of the technical uncertainties still attached to the attainment of the required gas purity with this alternative, as well as its restricted applicability, it was, in the end, a relatively easy decision to choose wet gas cleaning based on regenerative absorption (Figure 5) as the base case for the UCG synthesis-gas process. It should be emphasised that this decision does not preclude the possibility of further developing the novel wet gas cleaning technology (Figure 6) at a later date.

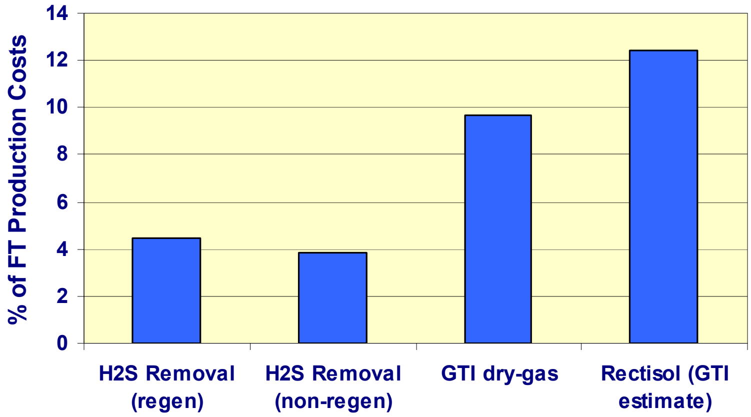


Figure 8. Estimated contributions to production costs of FT liquids of costs of certain gas-cleaning options.

2.4 Development and application of syngas-conversion model

When the UCG project was initiated, a number of syngas derivatives were considered as potential products of interest. These derivatives are listed in Figure 1. It should be mentioned that one of the key industrial companies participating in the project – Neste Oil – had, through internal evaluations, identified FT-derived fuels as the preferred final products for the transportation sector. In keeping with the objectives of the UCG project, comparisons of the efficiencies and costs of producing alternative syngas derivatives were nonetheless carried out as part of the process-evaluation work. A relatively simple syngas-conversion model was developed to facilitate the initial comparisons. The structure of the model is presented in Figure 9. In the publication by McKeough and Kurkela (2005), the model and its application are described in detail and certain results are presented and discussed. Examples of the results are given in Figures 10 and 11. Figure 10 compares the estimated production costs of various syngas derivatives as a function of feedstock price. The results in Figure 11 show how the model can be applied to provide useful information about the sensitivity of the production costs to key parameters. In the case of Figure 11, the key variable is the amount of residual CH_4 after gas reforming. From these results, it can be concluded, for example, that, for a FT process employing a reforming loop, the production costs would be relatively insensitive to the amount of residual CH_4 over the range examined.

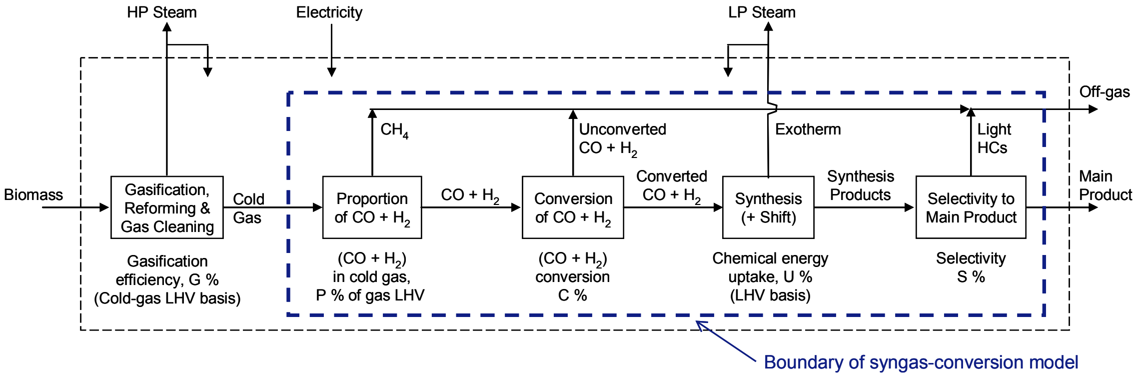


Figure 9. Structure of syngas-conversion model.

The investment-cost estimates, on the basis of which the production costs in Figures 10 and 11 were calculated, apply to plants employing mature technology and not first-of-a-kind plants. The estimate of the front-end section was largely based on in-house data, while the estimates for the syngas-conversion sections were based both on quotes from manufacturers and on published cost data (e.g. Hamelinck 2004).

2.5 Production-chain screening for Finnish conditions

Production-chain screening mainly focussed on the question of whether synthesis-gas production and conversion could be competitive at the scale of 200–400 MW_{feed}. This was the scale that appeared to suit Finnish conditions for the following reasons:

- procurement of the feedstock (biomass residues) at a competitive price
- feasible integration of the syngas process with pulp and paper mills
- realisation of fluidised-bed gasification as one or two trains, only.

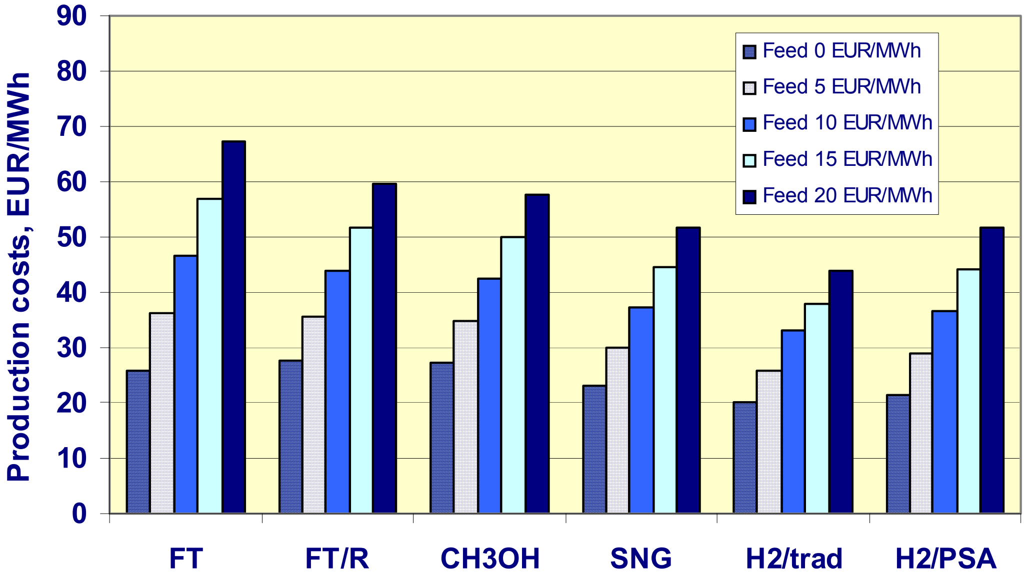


Figure 10. Estimated production costs of alternative syngas derivatives as a function of feedstock price. FT: once-through FT synthesis; FT/R: FT process with reforming loop; CH3OH: methanol; SNG: synthetic natural gas; H2/trad: hydrogen by traditional method; H2/PSA: hydrogen by separation using pressure swing adsorption. Electricity price: 30 EUR/MWh$_e$*. Interest on capital: 10% for 20a. Capacity: 260 MW*$_{feed}$*. LHV basis.*

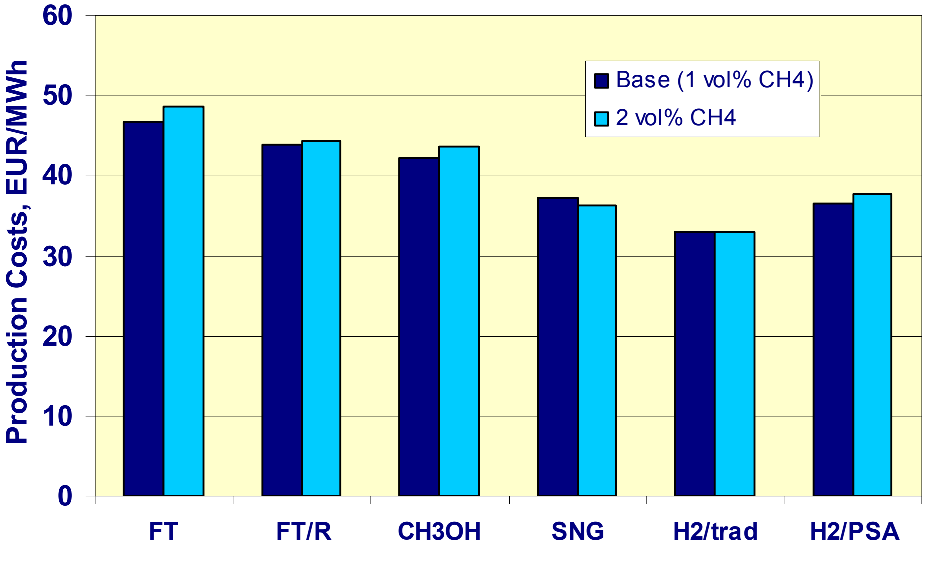


Figure 11. Estimated production costs of alternative syngas derivatives as a function of amount of residual CH_4 *after gas reforming. Process alternatives as in Figure 10. Feedstock price: 10 EUR/MWh. Electricity price: 30 EUR/MWh*$_e$*. Interest on capital: 10% for 20a. Capacity: 260 MW*$_{feed}$*. LHV basis.*

The investigations made in the UCG project did not reveal any reasons to doubt the competitiveness of the scale 200–400 MW_{feed}. Dutch studies had estimated, for a constant biomass feedstock cost, only moderate reduction in the production costs of FT liquids with increasing plant capacity above 300 MW_{feed} (Hamelinck 2004). In practice, this reduction will be offset by increasing feedstock costs.

The often encountered preconception that competitiveness would necessitate very large capacities derives primarily from the experiences of another sector. When producing FT liquids from natural gas, very large scales are needed in order to compete directly with petroleum-derived fuels. But the biomass-derived fuel market is a separate market having its own set of rules.

The work in this area also included an evaluation of the competitiveness of an alternative production chain based on the utilisation of biomass-derived pyrolysis oil as an intermediate. In this chain, tailored to Finnish conditions, pyrolysis oil would be produced in a number of small distributed plants, after which the oil would be transported to a central processing facility where it would be gasified in an entrained-flow gasifier having a capacity of about 300 MW_{feed}. The conclusion of this assessment was that, even when pyrolysis-oil production is efficiently integrated with heat and power production, costs of FT liquids would be about 20% higher than those of FT liquids produced directly from biomass residues in a syngas plant of similar capacity; i.e. about 300 MW_{feed}. However, it should be mentioned that this cost difference could be reduced by increasing the capacity of the centralised gasification process. There are no technical obstacles to the firing of pyrolysis oil in very large entrained-flow gasifiers. This approach is being seriously pursued in Germany (Stahl et al. 2005).

3. Development and application of detailed Excel-based simulation codes for methanol and FT plants

3.1 First phase: evaluation of variants of the gasification-reforming process

Detailed process schemes were compiled for methanol and FT plants and then the mass and energy flows were calculated in detail within the framework of the Excel code. For production of FT liquids, two configurations were simulated: (1) FT process with a once-through synthesis; (2) FT process in which FT off-gas is, to a high extent, recycled to the reformer. A block diagram of the FT process with reforming loop is given in Figure 12. During this and later phases of the work, the investment-cost estimates were continually upgraded as more reliable cost data came to hand.

The simulation codes were first applied to a comparison of several variants of the gasification train, including incorporation of two ideas for utilising higher-moisture feedstocks directly in the gasification stage. One of the key parameters in the gasification train is the temperature of the gas-filtration step prior to gas reforming. In view of the fact that the gas-reforming temperature will be similar to the gasification temperature, the closer the gas-filtration temperature is to the temperature of the gas exiting the gasifier, the more thermally efficient will be the overall gasification-reforming process. In the so-called UCG Target Configuration, the gasifier exit temperature is 750 ºC and the same temperature is applied in the subsequent gas-filtration step; i.e. no gas cooling is required between the two steps. Use of a gas-filtration temperature as high as 750 ºC has not yet been proven for this type of application. In the so-called UCG Conservative Configuration, the gasifier exit temperature is 850 ºC and the filtration temperature is 550 ºC. In respect of drying prior to gasification, the base case is drying down to a moisture content of 15% using low-pressure steam. However, because (1) steam is required as a reactant in the gasification, gas-reforming and shift steps and (2) the biomass-drying step contributes significantly to the investment and operating costs of the plant, the question of whether feedstock with higher moisture content could be fed to the gasifier naturally arises. Two novel ideas aimed at exploiting this possibility were evaluated in this phase of the project. These are discussed in more detail in the following paragraphs.

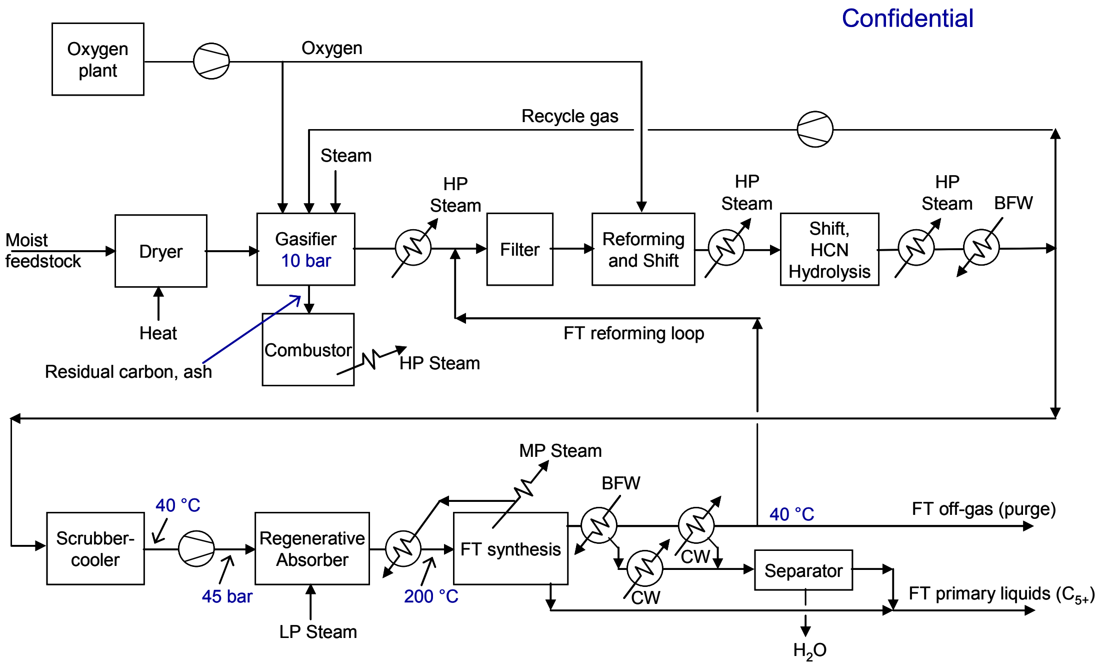


Figure 12. Block diagram of FT process with reforming loop.

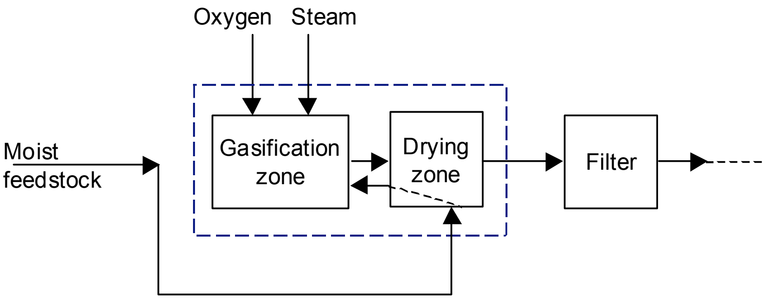


Figure 13. One principle for utilisation of higher-moisture feedstock in the gasifier (Wet-Feed Concept).

The first concept for feeding higher-moisture feedstock to the gasifier is depicted in Figure 13. The idea is to utilise the sensible heat of the product gas of the gasifier to partly dry the feed material before it enters the actual gasification zone. This would cool the product gas but the penalty would be minimal as long as the gas temperature remains above the (pre-set) temperature of the subsequent gas-filtration step. So this concept would mainly be of interest for a gasification-reforming process in which the intermediate gas-filtration temperature is significantly lower than the gasification temperature. The water evaporation occurring between the feeding point and the gasification zone does not consume any of the heat generated in the gasification zone through partial oxidation; i.e. it does not consume any oxygen. However, in this concept, the extra water fed into the gasification chain with the feedstock can not be exploited as reactant steam in the gasification zone. It can, however, improve the performance of the downstream gas-reforming and shift operations.

Table 1. Comparison of estimated methanol production costs for variants of the gasification-reforming process. As-received moisture content of feedstock: 50%. Gas reforming and shift modelled as equilibrium reactors. HW: secondary heat in the form of hot water. Unless otherwise stated, feedstock drying is by low-pressure steam. Feedstock price: 10 EUR/MWh. Electricity price: 30 EUR/MWh$_e$. Interest: 10% for 20 a. LHV basis.

Variant		**Moisture content of feed to gasifier, %**	**Filtration temperature, °C**	**Gasification pressure, bar**	**Methanol production costs, EUR/MWh**
1	Target	15	750	10	42.5
2	Target + part drying by HW: 50% → 40%	15	750	10	42.1
3	Conservative	15	550	10	45.4
4	Wet-Feed	50	550	10	47.9
5	Wet-Feed	50	550	30	45.7
6	Wet-Feed + part drying by HW: 50% → 35%	35	550	10	42.5

In the second concept, the higher-moisture feedstock is fed in the normal way to the gasifier. The idea is to employ a sufficiently high pressure in the gasification train to allow part of the heat of condensation of the extra feed water to be recovered as low-pressure steam. Referring to the block diagram in Figure 12, it is in the heat-exchange section before the scrubber-cooler where the heat of condensation can be converted to low-pressure steam if the pressure is sufficiently high. In effect, the gasification train would then work as a high-efficiency feedstock dryer; i.e. a dryer with low net input of heat. As in the previous concept, the extra steam originating from the water fed into the gasification chain with the feedstock can improve the performance of the downstream gas-reforming and shift operations. In this case, depending on the location of the biomass feeding point, the extra steam may even have a positive impact on the gasification reactions. However, preliminary calculations showed that significant recovery of the heat of condensation of water vapour as steam would not begin until the pressure had reached a level around 30 bar. Thus, this concept is not compatible with the preferred pressure level of the UCG gasification process; viz., 5–10 bar. Furthermore, to benefit from the concept, all the low-pressure steam produced in the syngas plant must be fully and efficiently exploited. In other words, the syngas plant must be integrated with an energy demanding industrial facility, such as a paper mill. Despite these limitations, the concept may be of interest as part of later-generation advanced process.

The main results of this phase of the evaluation work are given in Table 1. The main variables were the gas-filtration temperature and the application or not of the so-called Wet-Feed Concept; viz., the concept depicted in Figure 13. In each case, costs of producing methanol were estimated. The production costs for the Conservative Configuration are estimated to be about 7% higher than those for the Target Configuration (Case 3 vs. Case 1; Table 1). In later evaluations it was shown that, in the case of a FT plant, the application of a reforming loop decreases the relative difference between these two configurations in terms of both performance and costs.

For a gas-filtration temperature of 550 ºC, the production costs for the Wet-Feed concept are only moderately higher (< 6% higher) than those for the base concept, even though the moisture content of the gasifier feed material is very high (50%) in this Wet-Feed case (Case 4 vs. Case 3). Elevating the pressure of Case 4 to a high level (30 bar, Case 5) lowers the production costs to near those of the base case (Case 3). Case 5 incorporates the second of the above two novel ideas pertaining to the feeding of higher-moisture feedstocks. In other words, part of the condensation heat of water is recovered as low-pressure steam in Case 5.

Cases 2 and 6 both utilise secondary heat for the initial drying stage. In each case, the amount of secondary heat corresponds to that made available by the biomass-to-methanol process itself. When utilisation of secondary heat is combined with the Wet-Feed concept (Case 6), production costs can be reduced to the level of the process that is based on the Target Configuration and exploits secondary heat for drying (Case 2). It should be borne in mind that the Wet-Feed case (Case 6) employs a lower, technically more certain gas-filtration temperature (550 ºC vs. 750 ºC).

Overall, the results presented in Table 1 confirm that the Wet-Feed concept could be a promising way of offsetting the penalty of intermediate gas filtration at a temperature significantly below the gasification and reforming temperatures. However, so far, no simple and reliable ways to put this principle into practice have been devised. In the case of a circulating fluidised-bed gasifier, the novel drying step would need to be realised as a separate stage. Although VTT has preliminary ideas about the arrangement of this stage, there are a number of unresolved technical issues that would have to be addressed, requiring a significant R&D effort. Thus, the Wet-Feed concept was put “on hold” in the UCG project.

3.2 Second phase: detailed evaluations of the benefits of integration

Later in the project, results of bench-scale gas-reforming and shift experiments became available. These results were used to improve the modelling of these steps in the Excel based simulation codes. The modelling of the FT synthesis step was improved on the basis of fairly reliable published data that were uncovered at about the same time. The improved simulation codes were applied to generate performance and cost data for methanol and FT plants integrated with pulp and paper mills.

Methanol and FT plants produce considerable amounts of high-grade by-product energy in the form of steam and fuel gas. Figure 14 presents the same block diagram of a FT process as in Figure 12 but complemented with the energy flows for one example plant. Obviously, the conversion plant will benefit if the high-grade by-product energy can be fully utilised, e.g. in a pulp and paper mill. As discussed earlier (Section 2.5), the minimum capacity of a competitive synthesis-gas plant in Finland is considered to about 200 MW_{feed}, so, in general, a significant amount of solid biomass residues will need to be imported to the mill and a significant amount of by-product energy will be produced. When the site includes a kraft pulp mill, a precondition for effective integration is that the total demand for process heat on the site is clearly larger than the amount which can be provided by the recovery boiler. The ideal situation for integration of a FT plant is one where the power boiler of the mill is due for replacement or for refurbishing. The FT plant can provide most of the process heat that was earlier provided by the power boiler. Maximum integration benefit can be obtained when the feed input of the to-be-replaced power boiler is at least half of that of the proposed FT plant. For FT plants of 200 MW_{feed} capacity, this condition can be met by many of today’s integrated pulp and paper mills.

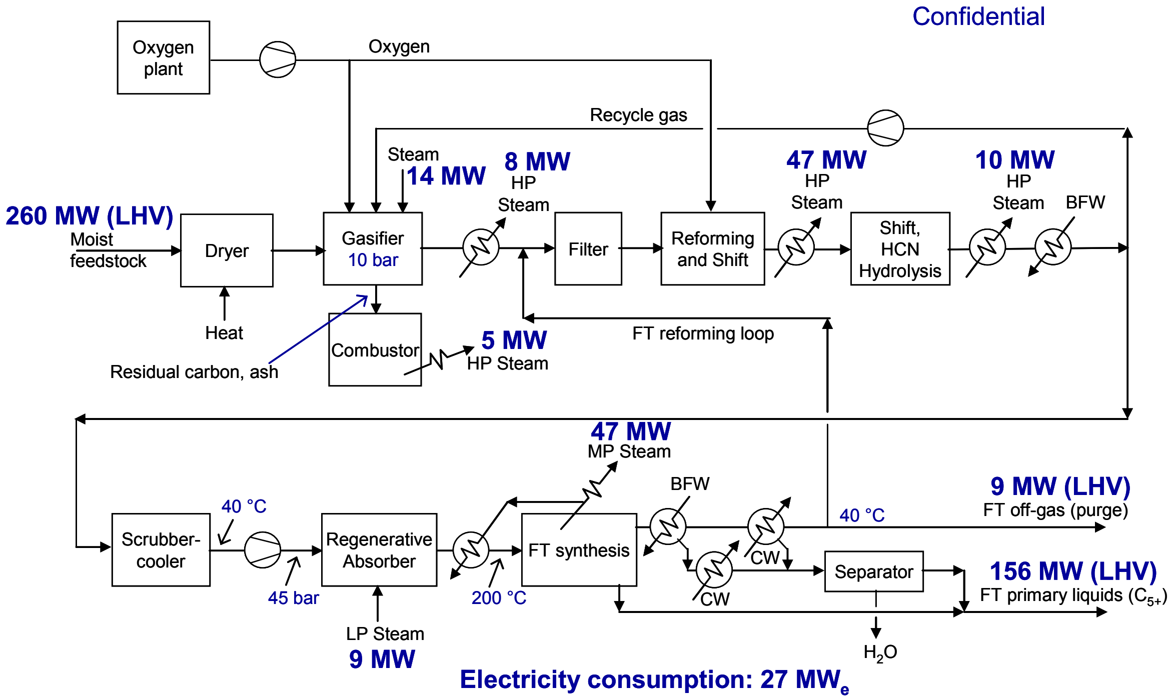


Figure 14. Block diagram of FT process with reforming loop, showing main energy flows for one example plant. Gas-filtration temperature 600 ºC. In case of steam production or consumption, MW means heat transferred from or to process stream. Pressure of high-pressure steam 100 bar.

With respect to the generation of high-pressure steam, the UCG base case incorporates a superheater after the gas reforming stage. Because of the elevated pressure on both sides of the heat-transfer surface, the design of this superheater differs considerably from that of the superheater of a conventional boiler. A noteworthy feature of the UCG process concept (Figure 14) is that the hot gas stream after the reformer is essentially free of particulate matter. It should be mentioned that generation of saturated high-pressure steam in conjunction with cooling of pressurised synthesis gas is a very widely applied industrial operation. Compact and reliable boilers are readily available. Thus, an alternative approach for the present purposes would be to generate high-pressure saturated steam in the gasification train and then superheat this steam in a separate chamber fired, for example, by FT off-gas. Final optimisation of the steam system can not be carried out until reliable estimates of the equipment costs of alternative heat-exchanger configurations are available.

Figure 15 depicts key features of the integration of a biomass-fired synthesis-gas process with an integrated pulp and paper mill. The synthesis-gas process could provide fuel gas to the mill for applications other than heat and power production based on combustion of the gas in gas-fired boilers or gas turbines. Figure 15 shows optional utilisation of fuel gas for final superheating of recovery-boiler steam and for fuelling the lime kiln. Paper mills may have other gas-fired equipment as well; e.g. gas-fired paper dryers. The preferred source of the fuel gas would usually be the off-gas of the synthesis process but there would also be the possibility to use part of the cleaned product gas from the gasifier. The FT synthesis offers great flexibility in terms of the amount of off-gas produced. The once-through synthesis typically converts about two thirds of the synthesis gas to liquids and about one third to off-gas. By recycling the off-gas to the gas-reforming step, the ratio of net liquids to net gas can be increased significantly. Clearly, the benefits of integration of the synthesis-gas process would be two-way. The synthesis-gas process would benefit from the full utilisation of its by-product energy in the pulp and paper mill while the mill would benefit from the availability of competitively priced biomass-derived fuel gas.

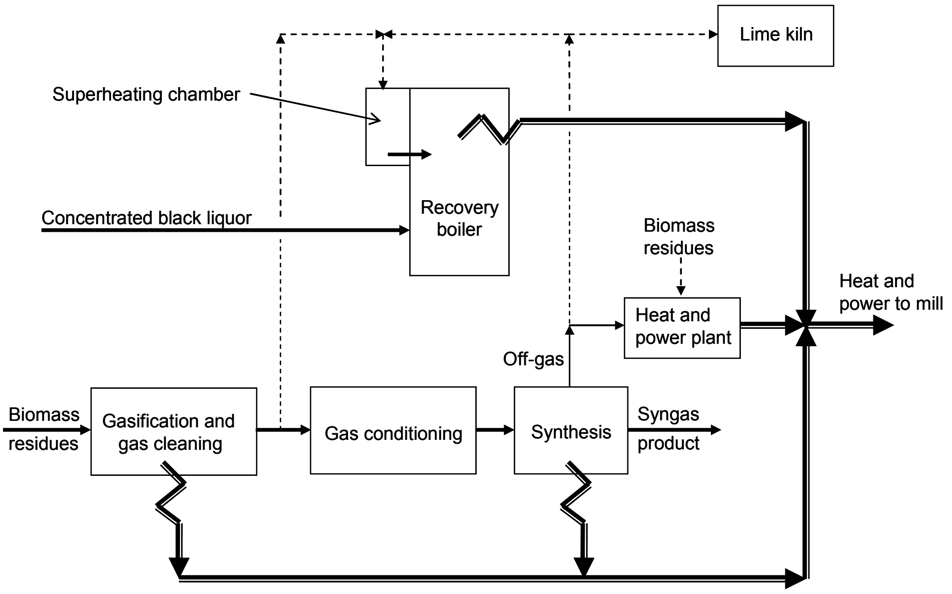


Figure 15. Principles of integration of biomass-fired synthesis-gas process with an integrated pulp and paper mill.

Figures 16, 17 and 18 explain how the incremental energy balance is arrived at when a FT plant and a new (smaller) power boiler replace the existing large biomass-fired power boiler of an integrated pulp and paper mill. In this example, the feed input of the FT plant is 260 MW (LHV), while that of the existing biomass-fired boiler is assumed to be at least half of this; i.e. at least 130 MW (LHV). As put forward above, full utilisation of the by-product energy of the FT plant is possible under these conditions. Other key features of this FT case are (1) a gas-filtration temperature of 600 ºC, (2) recycling of a large part of the FT off-gas to the reforming step and (3) drying of the biomass residues from the as-received moisture content of 50% to a moisture content of 30% by secondary heat and then from 30% to the final moisture content of 15% by low-pressure steam. Figure 16 shows certain energy flows before the integration of the FT plant and Figure 17 certain flows after the integration. From the numbers in Figures 16 and 17, the net changes in energy flows resulting from the integration can be calculated. These incremental flows are given in Figure 18, together with some information about the calculations themselves.

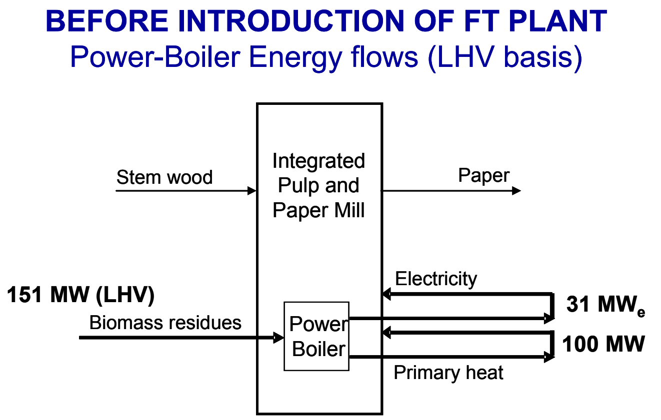


Figure 16. Example of integration with pulp and paper mill. Energy flows of to-be-replaced power boiler before introduction of FT plant.

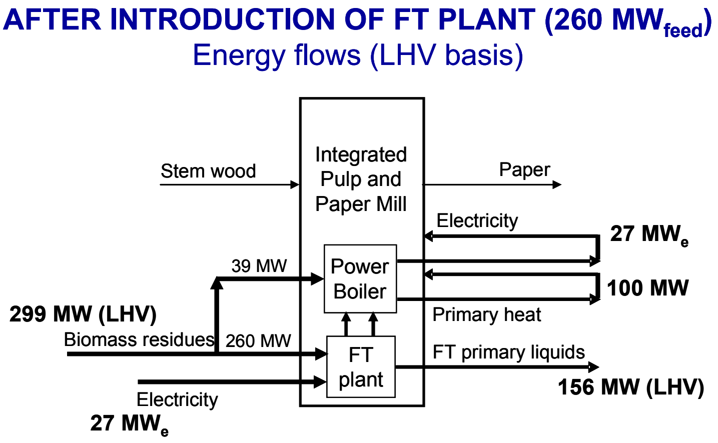


Figure 17. Example of integration with pulp and paper mill. Energy flows of FT plant and new (smaller) power boiler. Gas-filtration temperature 600 ºC. Configuration of FT plant as depicted in Figure 14. Drying of biomass residues from 50% moisture to 30% by secondary heat and from 30% to 15% by low-pressure steam.

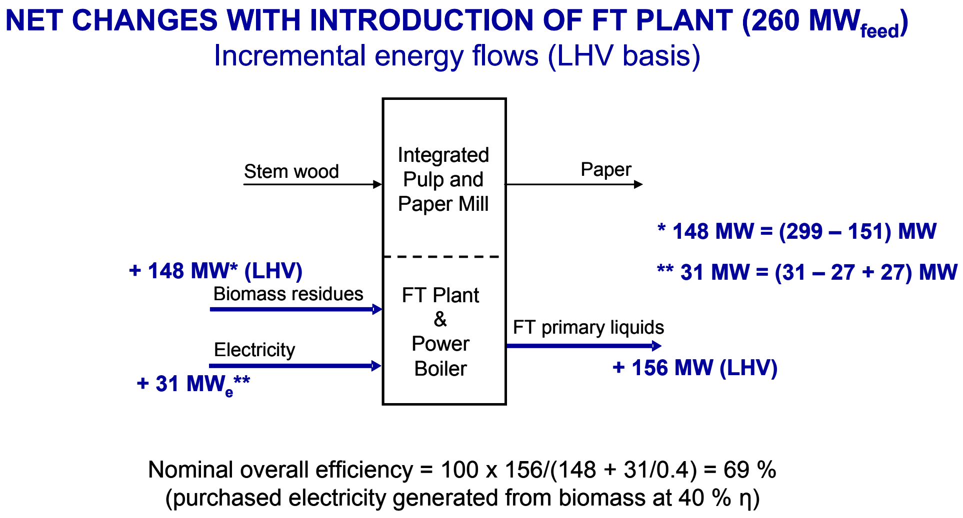


Figure 18. Example of integration with pulp and paper mill. Incremental energy flows upon introduction of FT plant and new (smaller) power boiler. FT case as in Figure 17.

The benefits of fully integrating the FT plant with a pulp and paper mill are evident in Figure 18. The effective feedstock-conversion efficiency is very high: more than 1 MWh of FT liquids can be produced from a net input of 1 MWh of biomass. On the other hand, a significant net input of electricity is required. A nominal overall thermal efficiency has been calculated and is presented in Figure 18. As elsewhere (Ekbom et al. 2003), this number has been calculated assuming that the required increment in electricity is produced from (additional) biomass at an efficiency of 40%. The overall efficiency is sensitive to the source of heat used for drying the biomass feedstock. It would decrease from 69% to 64%, should only primary heat be used for drying, while it would increase to 72%, should all the drying be accomplished with secondary heat, only.

As mentioned above, the FT process is flexible in terms of the ratio of output of FT liquids to output of FT gas. By employing a once-through FT configuration combined with utilisation of the off-gas in a combined-cycle heat and power plant, the required net electricity input can be reduced significantly. The incremental energy balance for such a case is given in Figure 19. Compared to the case of Figure 18, the case of Figure 19 exhibits, in addition to the smaller electricity input, (a) a lower feedstock-conversion efficiency and (b) a lower output of FT liquids (for the same feed input to the FT plant), but (c) a higher nominal overall efficiency. The higher overall efficiency is primarily a result of the fact that the combined-cycle heat and power plant has a high electricity-generating efficiency and a very high overall efficiency, whereas purchased power is assumed to be produced in a condensing power plant at 40% efficiency.

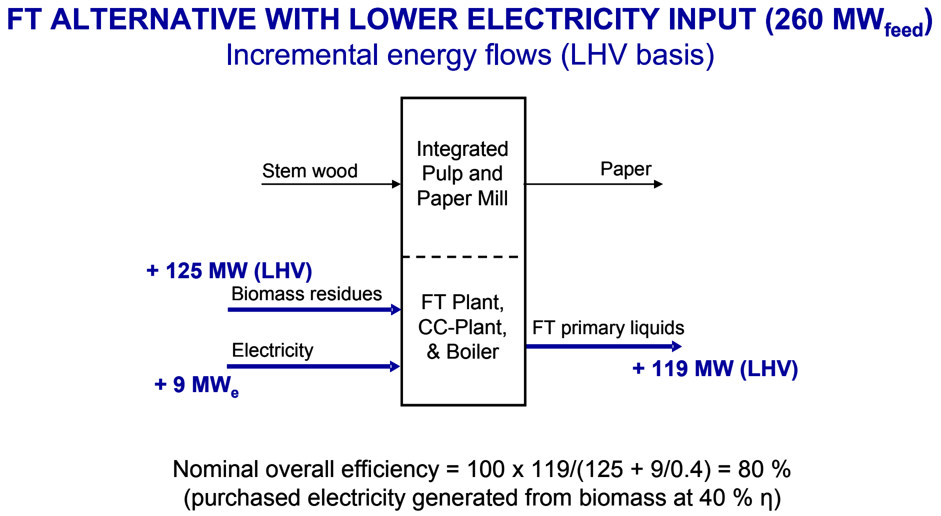


Figure 19. Incremental energy flows upon introduction of FT plant, combined-cycle plant and new (smaller) power boiler. Gas-filtration temperature 600 ºC. Once-through FT synthesis; off-gas fired in combined-cycle plant. Drying of biomass residues from 50% moisture to 30% by secondary heat and from 30% to 15% by low-pressure steam.

For the FT plant of the case examined in Figure 18, the total capital requirement was estimated at 245 MEUR. A breakdown of this investment estimate, which assumes mature technology, is provided in Table 2. Foster-Wheeler, one of the companies participating in the UCG project, carried out an independent design study of the front-end of a full-scale biomass-fired syngas process (250 MW_{feed}) based on the UCG concept. The scope extended from the feedstock receiving yard up till, but not including, the scrubber (refer Figure 12). A two-stage drying operation was included. VTT supplied Foster-Wheeler with (1) the mass and energy flow data (obtained from the Excel-based simulation codes) and (2) the key equipment-design parameters for the reforming and shift stages. Foster-Wheeler utilised in-house data for the design of feedstock handling equipment, gasifier, gas filters, gas coolers and all auxiliary equipment. The estimate of equipment costs was also based on Foster-Wheeler in-house data, either existing or newly acquired.

Table 2. Breakdown of estimate of total capital requirement (TCR) for 260 MW_{feed} plant producing FT primary liquids. Gas-filtration temperature 600 ºC. FT reforming loop incorporated. The TCR estimate includes the indirect project costs on top of the installed equipment costs. The estimate applies to mature technology.

	MEUR, as of year 2005
O_2 plant and compressor	30
Two-stage dryer	19
Gasification process, including feeding system, gasifier, filter, reformer, shift, HCN hydrolysis, gas coolers	120
Final gas processing, including scrubber, syngas compressor, Rectisol	41
FT-process, including gas heater and coolers	35
Total capital requirement	**245**

When estimating equipment costs, Foster-Wheeler considered both a first-of-a-kind plant and a plant based on mature technology. In the latter case, Foster-Wheeler came up with the following investment estimates: 18 MEUR for the two-stage dryer and 120 MEUR for the gasification section. On face value, these estimates are virtually identical to VTT's estimates (Table 2). However, VTT's gasification section incorporates a step for hydrolysing HCN in conjunction with the shift step. This step was not included in the scope of Foster-Wheeler design study. In addition, VTT's total capital requirement (TCR) covers all indirect costs, including start-up costs, interest during construction, etc. Not all such indirect costs are included in the estimates of Foster-Wheeler. All in all, it was concluded that VTT's estimates are 10–15% on the low side.

Figure 20 presents estimates of the costs of producing FT primary liquids, as a function of feedstock price, for four cases covering two different sources of heat for biomass drying – primary heat or secondary heat – and two degrees of integration of the FT plant with the pulp and paper mill – fully integrated or non-integrated. The assumed FT plant capacity is the same as in the cases of Figures 18 and 19, above; viz. 260 MW (LHV) of feedstock input. The fully integrated cases assume replacement of a large existing biomass-fired power boiler and, with the exception of the heat source for biomass drying, correspond to the case of Figure 18. The non-integrated cases assume that the FT plant is on a site adjacent to the pulp mill, thus benefiting from certain savings in infrastructure costs.

The unit used for the product cost in Figure 20, EUR/MWh, is convenient because EUR/MWh is roughly equivalent to eurocent/litre when the fuel is a hydrocarbon liquid. Note that the latter unit should not, however, be widely applied because it complicates comparisons with non-hydrocarbon and/or non-liquid bio-fuels, such as ethanol, methanol, DME, methane and hydrogen.

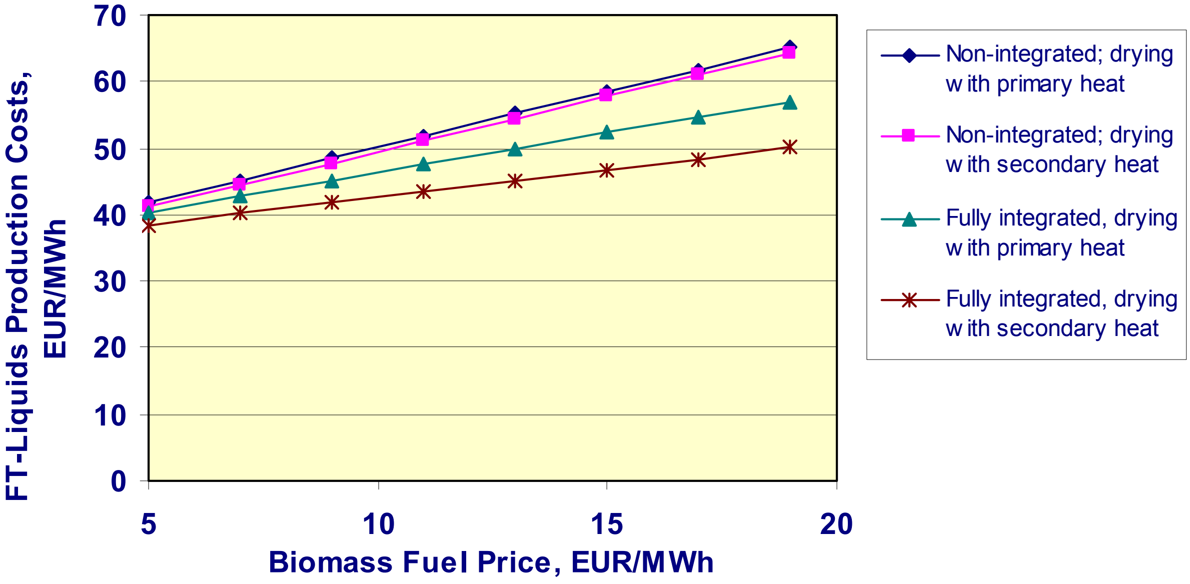


Figure 20. Estimates of the production costs of FT primary liquids for several cases. Feed input to FT plant: 260 MW (LHV) of biomass residues with 50% moisture content. Gas-filtration temperature 600 ºC. FT reforming loop incorporated. Electricity price: 30 EUR/MWh$_e$. Capital requirements according to VTT's estimates. Interest on capital: 10% for 20a. LHV basis.

It needs to be borne in mind that the production costs in Figure 20 refer to FT primary liquids. Upgrading based on mild hydrocracking and hydrotreatment is required to convert these liquids into marketable and/or blendable automotive fuels. Within the UCG project, a rough estimation of upgrading costs was made. According to Dutch sources (Hamelinck 2004), upgrading would cost about 4 EUR/MWh. VTT gained access to a confidential study of the costs of co-processing primary FT liquids in a petroleum-oil refinery. According to this study, the effective upgrading costs would be around 4.5 EUR/MWh, of which 1.5 EUR/MWh would be due to hydrogen. Costs will be higher if the final upgrading is carried out on the same site as the biomass-FT conversion plant. In this case, the cost of hydrogen (extracted from a gas stream of the biomass-FT plant) will be higher – perhaps 0.5 EUR/MWh higher – and the capital-related costs will be higher – perhaps 2 EUR/MWh higher. Thus, a rough estimate of the costs of final upgrading of FT liquids on the site of the biomass conversion plant is 7 EUR/MWh. Overall, it was decided that, at this stage, the costs of final upgrading would be presented as a range: 5–10 EUR/MWh.

This phase of the evaluation work also included a comparison of the performance and costs of the UCG process concept with those estimated for the German Choren process. The Choren process is one of the more advanced biosyngas processes – a 45 MW$_{feed}$ demonstration plant is in the start-up phase. The Choren gasification-reforming process is, in principle, an efficient one. Although gas reforming is based on thermal decomposition at high temperatures, the reforming step is realised in such a way that the final temperature of the gas exiting the gasifier is in the range 800–1000 °C. This, in turn, leads to an overall thermal efficiency of the gasification-reforming process which is similar to those of processes, such as the UCG process, that are based on catalytic gas reforming.

The results of detailed calculations confirmed that the efficiency of the Choren process, presuming it performs as claimed, would be similar to that of the UCG concept. On the other hand, the Choren process is a more complicated process, incorporating two, perhaps even three, gasification stages. Compared to the UCG process, the degree of technical uncertainty is clearly higher. Another key question is: how readily can the Choren process be scaled up from present capacity of 45 MW_{feed}?

4. Options for black-liquor processing

4.1 Comparison of biomass-fired syngas processes with black-liquor-fired syngas processes

The UCG evaluation work included a comparison of the estimated performance and costs of the biomass-fired UCG syngas process with those reported for the most advanced black-liquor-fired syngas process, namely that based on the Swedish Chemrec gasification process. A detailed assessment of the latter process had indicated relatively high efficiencies and low production costs for methanol manufactured from black liquor (Ekbom et al. 2003). Details of the comparison carried out in the UCG project are given in the publication by McKeough and Saviharju (2005). The thermal efficiency of a biomass-to-methanol plant, when integrated with a paper mill, was shown to be similar to that of a black-liquor-to-methanol plant based on the Chemrec process. This confirms that the high efficiency reported for the black-liquor-to-methanol process (Ekbom et al. 2003) is a consequence of the process being integrated with the pulp mill and not, as implied in Chemrec publications, a consequence of the type of feedstock: black-liquor vs. biomass.

The comparison carried out in the UCG project (McKeough & Saviharju 2005) did, however, conclude that the black-liquor-fired syngas process has the potential of being somewhat more economical than the biomass-fired alternative. However, other key factors speak in favour of the biomass-fired process. These include (a) a greater market potential in terms of the number of potential sites, (b) less interaction with the pulp-mill chemical-recovery cycle and so a smaller availability risk for the mill and (c) considerably less technical uncertainty that needs to be addressed before commercialisation of the technology. The larger potential market is a consequence of the fact that the application of biomass-fired technology is not restricted to the pulp mill. Not only can the technology be integrated with any industrial facility having a high demand for process heat; e.g. a paper mill based on purchased fibre, but also, in the longer term, the technology can be applied for stand-alone production of transportation fuels in dedicated biomass conversion plants.

4.2 Modifications to black-liquor processing to facilitate integration of biomass-fired syngas processes

The stand-alone market pulp mill is not, in general, a favourable mill with which to integrate a biomass-fired synthesis-gas process, because the amount of process heat produced by the recovery boiler is usually more than enough to cover the demand of the stand-alone pulp mill itself. To create a sink for the by-product energy of the FT plant, it would be necessary to avoid burning all the black liquor in the recovery boiler. Two ways of realising this are described in more detail in the following.

The first is based on gasifying part – say one third – of the black-liquor stream. The principles of this concept are shown in Figure 21. After the black-liquor-derived gas has been scrubbed, it is combined with the scrubbed gas from the biomass-fired gasifier. The hydrogen sulphide entering the main synthesis-gas processing line with the black-liquor-derived gas is removed in the gas-conditioning step and absorbed into cooking liquor. A feature of the concept is that, if necessary, the recovery boiler can handle the whole black-liquor stream. As a consequence of this, the concept imposes far less risk on the availability of the pulp mill than a process in which the recovery boiler is entirely replaced by a black-liquor gasification process.

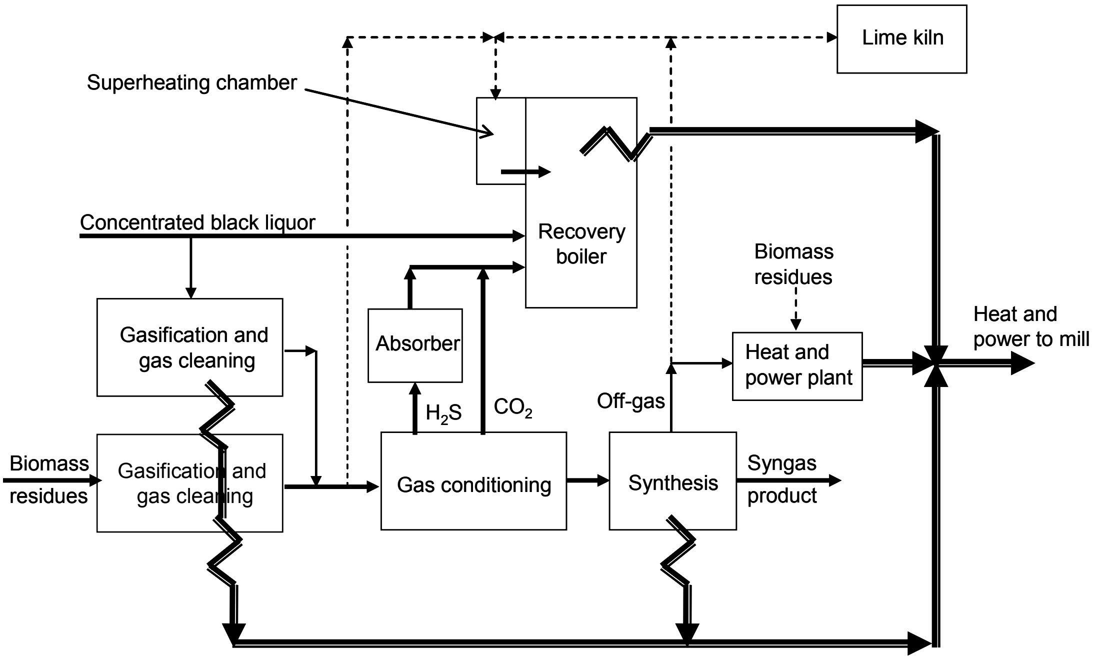


Figure 21. Principles of integration of a biomass-fired synthesis-gas process with a pulp mill wherein part of the black-liquor stream is gasified in a separate reactor and the scrubbed gas is combined with the scrubbed gas from the biomass-fired gasifier.

The second approach is based on removing lignin from black liquor by acidulation. One such lignin-removal process, LignoBoost^{TM}, has been developed in recent years by the Swedish Pulp and Paper Research Institute (STFI-Packforsk). VTT and STFI have developed a concept wherein a lignin-removal process is applied in conjunction with a biomass-fired synthesis-gas process. The key features of this concept are shown in Figure 22. According to the concept of Figure 22, the main option for the utilisation of the recovered lignin is co-gasifying in the biomass-fired synthesis-gas process.

However, export of the lignin for utilisation in higher-value applications would be another option. A key feature of the concept is the utilisation of a CO_2 rich gas stream from the synthesis-gas conditioning step as the acidulation agent in the lignin-removal process. This gas stream is otherwise a waste stream having a small negative value. When the lignin-removal process is applied in a mill in the absence of a synthesis-gas process, the costs associated with the procurement of the acidulation agent form a considerable part of the total costs. Thus, the presence of the synthesis-gas process has the potential to significantly reduce the costs of lignin removal. Obviously, application of the concepts depicted in Figures 21 and 22 is not restricted to the stand-alone market pulp mill. For example, the concepts could be attractive options for an integrated pulp and paper mill if the existing power boiler of the mill is not ready for replacement.

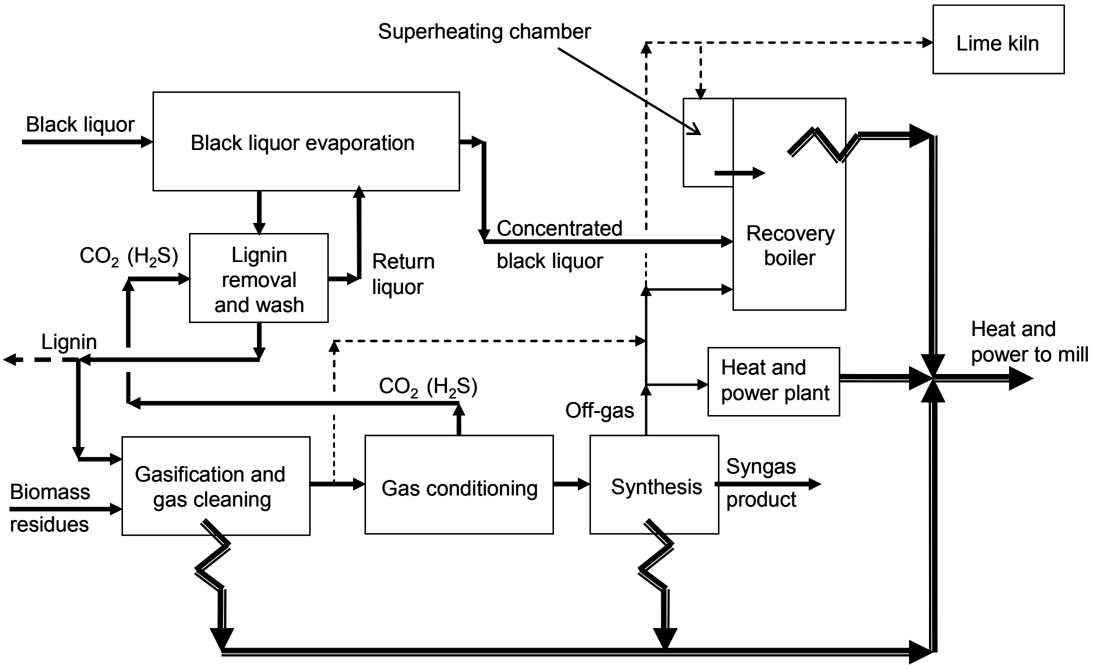


Figure 22. Principles of integration of a biomass-fired synthesis-gas process with a pulp mill according to the VTT-STFI concept wherein lignin is simultaneously removed from black liquor by acidulation.

On the basis of the evaluations of the UCG project, it was concluded that, with regard to the future processing of black liquor, the recovery boiler is in a strong position. On the other hand, black-liquor gasification and lignin removal from black liquor are seen as promising add-on technologies that would facilitate integration of biomass-fired syngas processes with certain types of pulp and paper mills.

The integration of biomass-fired syngas processes with various types of pulp and paper mills has been the subject of the publications by McKeough and Saviharju (2007) and

Saviharju and McKeough (2007). Note that the four separate publications, so far mentioned, present performance and cost data which differ somewhat from those presented in this publication proper. In all cases, one reason for the differences is that the data in the publications have been derived for the Target Configuration of the UCG gasification-reforming process. In the Target Configuration, the gas-filtration temperature is 750 ºC, whereas, in this publication, the main case evaluated employs a gas-filtration temperature of 600 ºC. Secondly, in some of the cases considered in the publications, the system for drying the biomass feedstock differs from that assumed in this publication. Thirdly, data presented in the two publications of 2005 were derived using earlier versions of the Excel-based simulation codes.

5. Detailed comparison of the performances and costs of producing FT liquids, methanol, SNG and hydrogen

In this phase of the work, Excel-based simulation codes were compiled for two additional syngas derivatives: for synthetic natural gas (SNG) and for hydrogen. The estimated performances and costs of these processes were then compared with those derived earlier for processes producing FT liquids and methanol. The process flow diagrams and the results of the comparison are presented in detail in the publication by McKeough and Kurkela (2007). Key results of the evaluation are shown in Figures 23, 24 and 25.

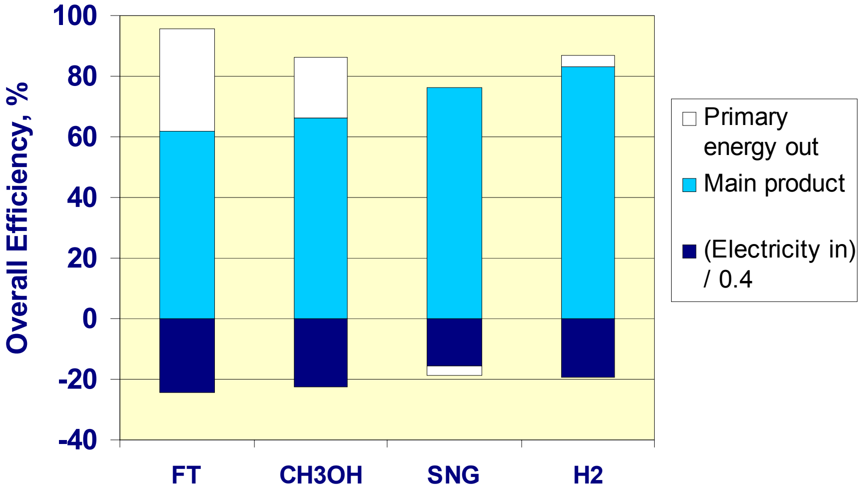


Figure 23. Breakdown of estimated overall efficiencies (LHV basis). FT: FT primary liquids. Target Configuration of gasification-reforming process.

Estimates of the overall efficiencies of the various alternatives are shown in Figure 23, with the breakdown into the contributions of (1) the main product, (2) the high-grade by-product energy and (3) the electricity input (negative contribution). As before, for the purposes of calculating the overall efficiency, electricity input is assumed to be produced from (additional) biomass at an efficiency of 40%. The detailed calculations have brought into light unexpected differences in the overall efficiencies of the various processes. The reasons for these differences are elucidated in the publication by McKeough and Kurkela (2007). Of the four alternatives, the FT process has the lowest efficiency of conversion of the feedstock into the main product (feedstock-conversion efficiency) but the highest overall efficiency (Figure 23). The SNG and hydrogen production processes do not yield significant amounts of high-grade by-product energy and so will not benefit from integration with pulp and paper manufacture.

The estimated production costs are presented in Figure 24. Also included for comparison are corresponding estimates obtained with the simple syngas-conversion model, described earlier (Section 2.4). The production-cost estimates based on the detailed process calculations display only a relatively small dependence on the end-product (Figure 24). The differences are smaller than would be expected on the basis of the feedstock-conversion efficiencies and are largely explained by compensating differences in the overall thermal efficiencies (Figure 23). Because the simple syngas-conversion model, developed earlier (McKeough & Kurkela 2005), does not fully take into account the latter differences, it is not surprising that it predicts a greater dependence on the end product. When comparing the various end-products in terms of their overall competitiveness as transportation fuels, the significant differences in distribution and end-use costs need to be taken into account. These costs increase considerably in the following order:

FT-derived fuels < methanol < SNG < H_2.

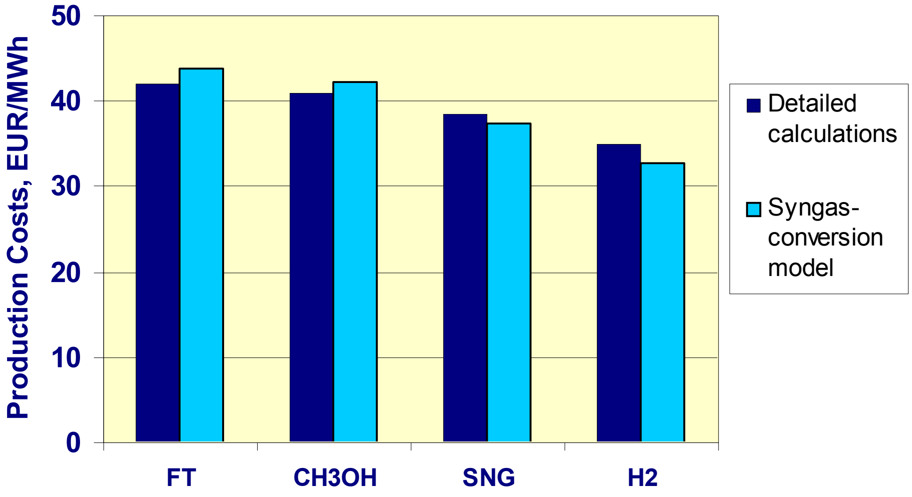


Figure 24. Estimated production costs of four syngas derivatives. Feedstock price: 10 EUR/MWh. Plant capacity: 260 MW_{feed}. Electricity price: 30 EUR/MWh_e. Interest on capital 10%, 20 a. FT: FT primary liquids. LHV basis. Target Configuration of gasification-reforming process.

Reasonably accurate estimates of distribution and end-use costs can only be derived for two of the end-products considered herein: FT liquids and SNG. After upgrading, the former can be distributed and utilised in the same way as conventional petroleum-derived automotive fuels. SNG can be distributed via natural-gas pipeline grids and then via special compression and filling stations. In several European countries, natural gas is already used an automotive fuel, so fairly reliable information about the costs of distributing and utilising the gas as an automotive fuel are available. Estimates of the equivalent biomass-to-pump costs for these two alternatives are given in Figure 25. According to these results, the biomass-FT route is clearly more competitive than the biomass-SNG route. Because both the distribution costs and the extra end-use costs will be considerably higher for H_2 than for SNG, the results of Figures 24 and 25 do not bode well for the biomass-H_2 route.

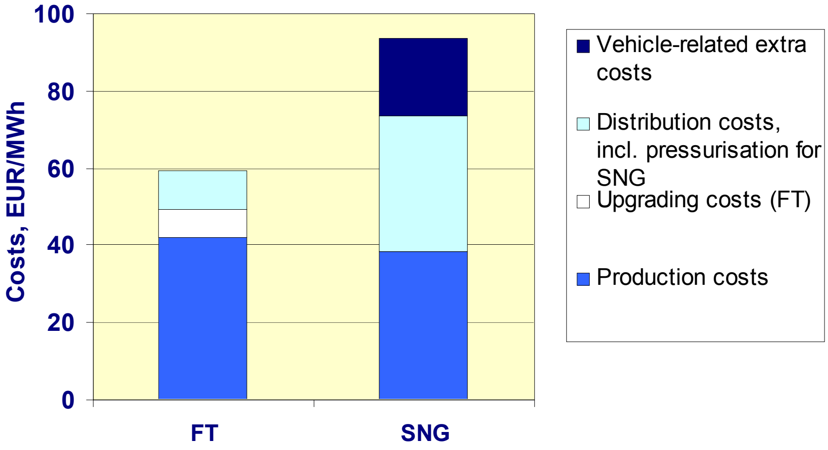


Figure 25. Estimates of equivalent biomass-to-pump costs for FT-derived fuels and SNG. Production cost estimates from Figure 24. Feedstock price 10 EUR/MWh. LHV basis.

6. Techno-economic comparison of a peat-fired syngas processes with the biomass-fired syngas process

The feedstock data applied for peat in this evaluation are given in Table 3. The as-received moisture content was assumed to be 50%.

Table 3. Peat fuel data (dry-matter basis).

LHV	**20.9 MJ/kg**
C	54%
H	5.8%
O	33%
S	0.2%
N	2%
Ash	5%

Wood-fired and peat-fired syngas processes were compared using one UCG case, only:

- Target Configuration for the gasification-reforming process (temperature of gas exiting gasifier 750 °C; filtration temperature 750 °C)
- synthesis of FT primary liquids using a once-through configuration.

A plant capacity of 260 MW_{feed} was applied.

The initial cost comparison between peat-fired processes and wood-fired processes was made with the assumption of the same price for the feedstock, viz. 10 EUR/MWh. In Section 6.2, the expected price differential between peat and wood fuels (to the advantage of peat) is taken into consideration.

One expected difference between peat-fired and wood-fired processes is that the peat-fired process will have a lower extent of carbon conversion in the gasification step. Based on experience obtained with atmospheric-pressure CFB gasifiers, the carbon conversion for the peat-fired process was set at a value 4% lower than that of the wood-fired process (94% vs. 98%).

6.1 Results and discussion pertaining to the comparison made at the same feedstock price for peat and wood

The predicted levels of NH_3 and H_2S in the gas from the peat-fired gasifier are 16 000 ppm and 700 ppm, respectively. As expected, these are much higher than the levels that will be encountered with most woody feedstocks. The relatively high level of H_2S raises the question of whether this level would be too high for the S-resistant catalysts planned for the UCG reforming and dirty-shift stages. However, VTT catalyst experts are fairly confident that the catalysts in question can withstand levels of H_2S of the order of 1 000 ppm.

The high level of NH_3 would not create any additional technical challenges for the UCG process as it presently stands. NH_3 can be readily removed in the scrubbing stage. On the other hand, if NH_3 is removed, as such, from the main process stream, not only would this mean the creation of a significant waste stream, but also it would signify a significant loss of hydrogen. Thus, it would be advantageous to reform and/or decompose NH_3 in the peat-fired gasification-reforming process. The extent of decomposition of NH_3 in the main gas-reforming and shift stages currently planned for the UCG process appears to be limited. However, VTT researchers believe that an additional catalytic step (copper-based catalyst at about 400 °C) will be effective. VTT is planning to conduct research on this step in the future.

Two alternative peat-fired cases were evaluated:

- case assuming no reforming or decomposition of NH_3 exiting the gasifier
- case assuming complete reforming and/or decomposition of NH_3 exiting the gasifier.

Table 4 compares key results obtained for the two peat-fired cases with those obtained for the corresponding wood-fired process. In all three cases it was assumed that the feedstock is dried from 50% moisture to 15% moisture using secondary or waste heat. The differences between the three cases are relatively small. The peat-fired process has the advantage of a somewhat lower throughput of dry matter for a given feed-energy input. (The LHV of peat dry-matter is about 10% higher than that of wood.) This mainly shows up as a lower investment cost. However, the advantage of the higher heating value of peat is more than offset by the lower extent of carbon conversion during gasification. Product yields are lower than those of the wood-fired process. The high-pressure steam yield is higher for the peat-fired process because non-gasified carbon is combusted to produce high-pressure steam. The low-pressure steam yield of the peat-fired process is lower, mainly because of a higher feed steam demand in the gasification step. At the same feedstock price (10 EUR/MWh), estimated production costs are a few percent higher for the peat-fired process.

Table 4. Key estimated performance and cost data of peat-fired FT processes in comparison with those of a reference wood-fired process. Feedstock price 10 EUR/MWh. Electricity price: 30 EUR/MWh$_e$.

Case	**Peat; zero NH_3 reforming**	**Peat; complete NH_3 reforming**	**Wood**
Feed-rate, MW (LHV)	260	260	260
FT-liquids (C_{5+}), MW (LHV)	117.1	120.1	125.5
Off-gas, MW (LHV)	62.6	64.2	67.1
HP steam (transferred heat), MW	49.0	49.3	44.2
LP steam (transferred heat), MW	21.8	22.6	30.8
O_2 requirement, t/d	586	598	575
Electricity requirement, MW	19.2	19.5	19.8
Investment, MEUR	201	202	210
Production costs of FT liquids (C_{5+}), EUR/MWh	44.7	43.6	42.6

On the basis of these results and bearing in mind the uncertainty attached to the present performance and cost estimates, it is concluded that, for current purposes in the Finnish process development work, sufficiently accurate cost estimates for peat-fired processes can be obtained directly from the results of evaluations of the corresponding wood-fired processes. For example, the results presented in Figure 20, derived for woody feedstocks, can be applied as such to peat.

6.2 Comparison of production costs at typical current prices for peat and wood

Figure 26 is a duplicate of Figure 20 with example feedstock prices superimposed. In accordance with the conclusion of the previous section, it is assumed that, at the same feedstock price, the production costs for the peat-fired plant are nearly the same as those for the wood-fired plant. On the figure are marked typical current (delivered) fuel prices in Finland: 8.5 EUR/MWh for peat and 12.5 EUR/MWh for forestry residues. It can be seen that, at these feedstock prices, the predicted production costs for the peat-fired plant are about 15% lower than those for the corresponding wood-fired plant. However, it must be borne in mind that peat is currently classified as a fossil fuel in the EU, so that, on the transportation fuel market, peat-derived fuel has to compete directly with conventional petroleum-derived fuels – at least for the time being.

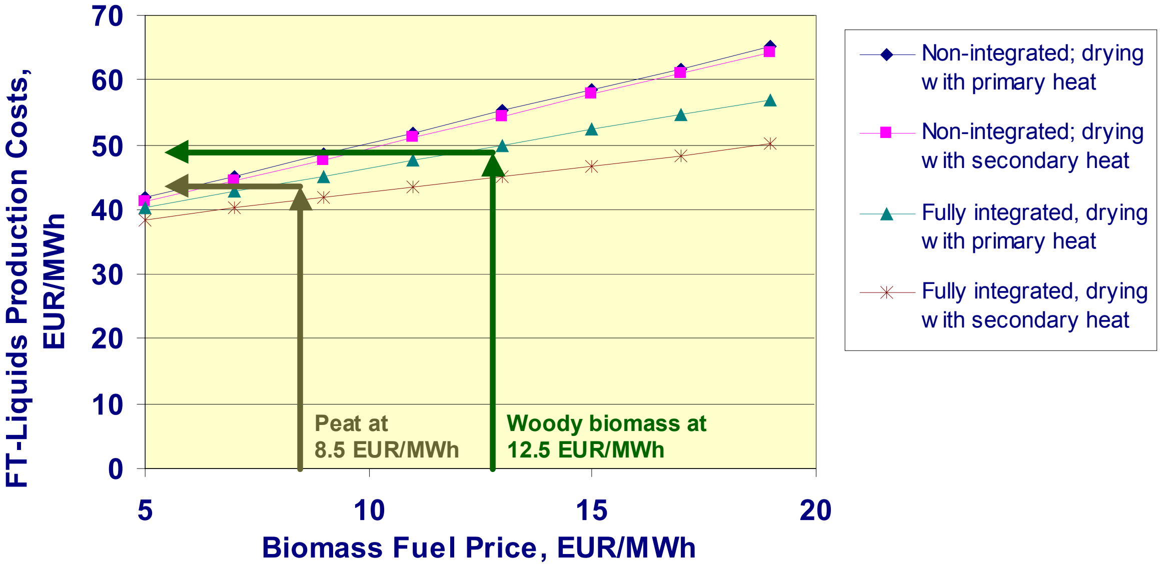


Figure 26. Estimates of the production costs of FT primary liquids for several cases. Feed input to FT plant: 260 MW (LHV) of peat or woody residues with 50% moisture content. Gas-filtration temperature 600 ºC. FT reforming loop incorporated. Electricity price: 30 EUR/MWh$_e$. Interest on capital: 10% for 20a. LHV basis.

7. Concluding remarks

The process-evaluation work in the UCG project covered a number of topics, including:

- process-configuration screening
- production-chain screening
- techno-economic evaluations of plants producing FT liquids, methanol, SNG or hydrogen; processes simulated with Excel-based codes
- techno-economic evaluations of competing technologies / production chains
- evaluations of benefits of integration; including novel concepts
- design studies of front-end operations by Foster-Wheeler and Pohjolan Voima.

The results of the work have proved useful in:

- developing and improving the UCG concept for bio-syngas production
- planning the experimental R&D work in the UCG project
- providing input data for sustainability analyses, CO_2-balance calculations, etc. (carried out elsewhere)
- compiling business plans and developmental strategies.

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Title

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Abstract

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