

**FEASIBILITY PHASE PROJECT FOR
BIOMASS-DERIVED ALCOHOLS
FOR AUTOMOTIVE AND
INDUSTRIAL USES**

ALTENER "BAL-Fuels Project"

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FINAL REPORT

**Ecotrafic R&D AB
Nykomb Synergetics AB**

Stockholm in September 1997

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Preamble.

This final report covers work under Phase 2 in the outline of the project (May 1996 - June 1997) and is based on the technology reviews and selections in phase 1 (Nov. 1995 - April 1996) and reported in the Interim report (Sept. 1996.)

The work in phase 2 has mainly been an effort to more closely analyse:

- two technologies for biomass gasification to synthesis gas,
- the enzyme route to hydrolyse celluloses to fermentable sugars, and
- executed engine tests.

A preliminary engineering study on methanol production from biomass, based on the HTW-gasifier finally chosen, was assigned to the engineering company Krupp-Uhde, Dortmund, with participation from Haldor Topsoe, Copenhagen. Nykomb Synergetics co-ordinated this work, completed the study's energy balances and wrote the chapter on methanol production.

The Swedish Ethanol Foundation supplied available input data from a recent study on the ethanol route to make an up-date possible.

Volvo Truck Co participated in the planning and analysis of the engine tests, which were carried out on the test bench at the engine laboratory at Luleå University, Sweden.

Ecotraffic wrote the summary, the chapters on ethanol production and LCA-considerations, and compiled this report.

Stockholm in September 1997.

Bengt Sävbark

Managing Director

Abbreviations and acronyms

ASU	Air Separation Unit (to produce pure oxygen)
ATR	Auto-Thermal Reformer
BAL	Bio-Based Alcohols (work name)
CASH	Canada, America, Sweden, Hydrolysis
CFB	Circulating Fluidised Bed
DEM	Deutsche Mark
DME	Dimethylether
DOE	Department of Energy (USA)
DS	Dry Solids, i. e. moisture-free solids
HTW	High Temperature Winkler (gasifier)
IEA	International Energy Agency
KFB	Swedish Transport & Communications Research Board
LCA	Life Cycle Analysis
LHV	Lower Heating Value
MJ	Megajoule
MM	Millions
NOx	Nitrogen Oxides
NREL	National Renewable Energy Laboratories (USA)
PAH	Polycyclic Aromatic Hydrocarbon
R&D	Research and Development
SEK	Swedish Crown
SRF	Short Rotation Forestry
SSEU	Swedish Ethanol Development Foundation
SSF	Simultaneous Saccharification and Fermentation
Syngas	Synthesis gas (mainly hydrogen, H ₂ , and carbonmonoxide, CO)
USD	US Dollar

SUMMARY

The BAL-project has comprised

- ◆ reviews and up-dating of routes to motor alcohols, methanol by gasification/synthesis and ethanol by hydrolysis /fermentation from lignocellulosic feedstocks (wood biomass),
- ◆ heavy duty engine testing of alcohol-containing diesel oils, and
- ◆ discussion on possible actors and measures for implementation of bio-alcohol production via erection of a demonstration plant.

RAW MATERIAL

The average raw material (forestry and forest industry residues, short rotation forestry hardwoods) will have an *energy content (LHV) of just above 19 MJ/kg dry wood and contain slightly above 60 wt-% of cellulose and hemicellulose.*

METHANOL PRODUCTION

For methanol production via gasification the oxygen/steam-blown, pressurised gasifier was preferred to the atmospheric, indirectly heated gasifier (pyroliser) with heat transferred from a separate air-blown combustor. The latter (Battelle Columbus) was judged as less mature process with unproven elements, convincing engineering data could not be obtained, and claimed efficiencies could not be substantiated.

The preliminary engineering study by Krupp Uhde/Haldor Topsoe/Nykomb was therefore based on the HTW-gasifier, optimally operating at a pressure of 27 bar. For an energy self-sustained plant (i.e., no other external energy supply other than the feedstock) *49 % of the energy content of the dry feedstock (LHV-basis) can be converted to methanol and another 12 - 21 % be recovered as hot water for district heating.* Conversion of process feedstock to methanol is 57 % but needed auxiliary fuel (electricity) to make the plant energy self-sustained reduces the figure to 49 %. However, it should be possible to augment this figure to about 55 % by improved technology and power/heat management.

Investment costs for the 1,000 t/d methanol plant studied have been estimated to ECU 0.9 per litre/year or ECU 1,600 per kW methanol capacity. At typical feedstocks costs of ECU 9.5/MWh and 12 % annuity a methanol cost of ECU 0.25 per litre (ECU 16 per GJ) is estimated without credits for district heat.

A demonstration plant could consist of one of the three gasification trains foreseen for the commercial plant studied in this report.

ETHANOL PRODUCTION

For ethanol production the enzymatic hydrolysis is assumed to have had a proven breakthrough and micro-organisms for fermentation of also pentoses have become available. Thus, today's in pilot-scale proven figures on conversion from theoretical to actual yield of ethanol of nearly 70 % are assumed to raise to 85 % by further development, resulting in a future practical yield of 30 wt-% of ethanol or nearly 41 % on energy basis (LHV). By-products at the conversion are sufficient to make the ethanol plant energy self-sustained and to produce a small surplus of other saleable products such as solid "lignin" fuel or electricity, bringing the total energy yield to somewhat above 50 %. Raw materials enriched in cellulose/hemicellulose or selected, cellulose/hemicellulose-rich feedstocks should be preferred, and the remainders be used as fuels.

However, the processes have yet to be proven in continuously operating pilot and demonstration plants.

Investment costs have been estimated in the range of ECU 1.5 - 0.6 per litre/year of ethanol with figures at the high end for the moderately sized plants considered in Europe.

Production costs for ethanol from average wood feedstocks are for Nordic conditions (cost ECU 9.5 per MWh) estimated to be in the range of ECU 0.41 - 0.47 per litre (ECU 19 - 22 per GJ). Claimed figures on lower costs emerge from estimates for big plants with cheaper, selected, cellulose/hemicellulose-rich feedstocks.

ENGINE TESTING

Engine testing (11 lit. heavy duty diesel engine) with 12 vol-% methanol (water-free) emulsified in diesel oil showed in comparison with pure diesel oil a considerable reduction of emissions of nitrogen oxides (NOx) and particles at a slight reduction (4 - 5 %) of power and torque but with an improved efficiency (2 - 3 %). Parallel tests with 15 vol-% ethanol (azeotrope) emulsified in diesel oil did not produce the reduction of NOx emissions recorded for methanol, and earlier tests elsewhere could not be reproduced.

Alcohol only fuels are effective in reducing emissions of NOx, particles and polycyclic aromatic hydrocarbons (PAHs) and their biological effects. In these respects, the tests also show that the environmentally classified oils in Sweden (very low contents of PAHs and sulphur) are superior to European standard diesel oils.

Alcohols used to produce stable emulsions in diesel oil need not be water-free but may contain a few percent water, for instance the ethanol/water azeotrope or crude methanol. Blended fuels must be standardised on energy content.

LIFE CYCLE ANALYSIS CONSIDERATIONS

It is quite obvious from the studies reported that both methanol and ethanol can be produced by conversion from wood with use of almost only renewable energy, i. e. fuels used are by-products or part of the biomass feedstock.

Energy usage in other steps in the chain from feedstock production to end use in vehicles (forestry, agriculture, transports) is today mainly based on fossil resources. In the whole chain *about ten times as much transport fuel energy in form of methanol or ethanol can be obtained for each unit input of fossil energy in a properly designed system.*

This ratio can be considerably lower when fossil auxiliary fuels and electricity are used. Such systems should not be accepted other than during transitions to fully renewable systems.

INDUSTRIAL ACTORS

The efforts for commercial demonstration will focus on selecting raw materials, a sit, a process configuration, product slate, and commercial investor partnership and alliances, which will meet among others the following criteria:

- Robust investment object in terms of product flexibility.
- Sustainable long term operation.
- Plant scale to meet commercial demonstration objective.
- Transport fuel produced, which will be compatible with major automotive actors views.

The ideal candidate are sites where substantial volumes of forestry material is at hand or can be transported at low costs, i. e., in conjunction with pulp mills or similar. Also, the ideal site should provide a possibility of heat and utility integration with the commercial demonstration site.

Ecotraffic, Nykomb, and SSEU have during 1996 and 1997 had several contacts with and studies for national and local energy companies, authorities and transport companies as preparation for entering demonstration stages.

To further raise the interest from potential partners, it is of great importance that the European Commission will indicate interest in continued support to proceed to a level of knowledge, where commercial demonstration plant investors can be attracted.

These future actions are also in line with earlier presentations to the Commission of continuation of this "BAL-Fuels" project.

1 BIOMASS GASIFICATION, METHANOL PRODUCTION

1.1 Methanol Production via Battelle Columbus Technology

Reference is made to the Phase 1 documentation in which the reasoning behind the selection of the Battelle Columbus technology during the technology selection phase is given. As a consequence of the selection in Phase 1 work was started according to the Battelle Columbus route.

1.1.1 Concept Principle

The overall concept is illustrated in Figure 1.1, Biomass based methanol production (Battelle Columbus route). The gasifier is operated at atmospheric pressure and is provided with the necessary heat for gasification indirectly via circulating hot sand. Due to the atmospheric pressure in the gasifier the syngas must be compressed later in the process chain to reach the pressure of the methanol synthesis.

Certain gas conditioning can be done in the tar decomposition step downstream the gasifier. Methane (CH_4) reforming must however be done to reach maximum overall yield. Inert gas for fuel feeding becomes less of a problem in an atmospheric system. Syngas compression instead becomes a major energy consumer. As the sand heating Circulating Fluidised Bed (CFB) is using air as oxidant and the heat input to gasification is indirect via the hot sand no air separation unit to produce pure oxygen is required.

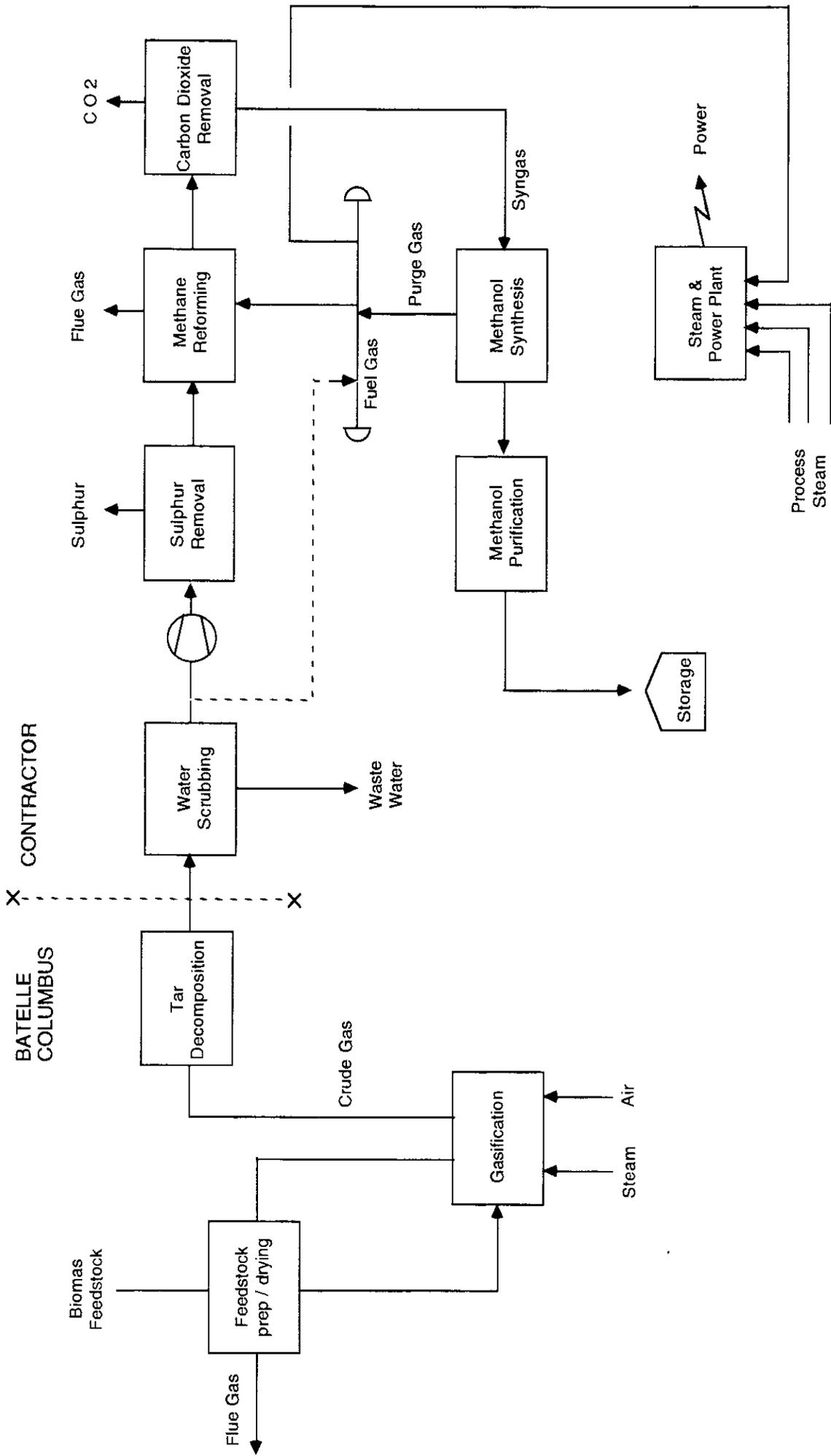
1.1.2 Process Description

(Due to the termination of the work on this process route the description of the process is only preliminary).

Gasification

The Battelle Columbus gasification system is a process operated at atmospheric pressure and consists of two CFB units. The primary CFB is the gasifier in which the fast pyrolysis of the biomass takes place. The heat necessary for the reactions is brought to the biomass fuel by hot sand.

This creates an efficient heat transfer and the retention time in the gasifier is therefore low, about half a second resulting in that the gasifier volume can be kept comparably small. However, the gasifier reactions create a considerable amount of char. This flows with the sand to the second CFB in where the char is combusted with air to heat up the sand and the sand is thereafter recycled to the gasifier. The hot flue gases from the second CFB are used for steam production and for fuel drying.



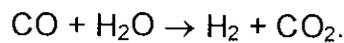
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SCALE	NTS
DRGN#	1000
PROJECT	
TITLE	BIOMASS TO METHANOL

Figure 1.1. Biomass based methanol production (Battelle Columbus route).

Tar decomposition

The hot crude gas from the gasifier is fed to a hot gas processing step in which:

- a) Tars are decomposed,
- b) Carbon monoxide (CO) is shifted with water (H₂O) to hydrogen (H₂) and carbon dioxide (CO₂) according to the reaction:



This unit makes the downstream water scrubbing less severe as the scrubber water only contains traces of non-converted tars. The process step is also claimed to carry out the shift reaction according to b) above, thus making a downstream shift reactor unnecessary.

Water scrubbing

After cooling, the crude gas is scrubbed with water to make sure that the gas is free of traces of tars and particulates.

Syngas compression

Before further processing to methanol syngas the crude gas is compressed. This is done downstream of the water scrubber as the compression work is at its minimum with this chosen process configuration.

Gas purification / CH₄ reforming

The crude gas has to be cleaned from sulphur components, most of its CO₂-content and has its methane (CH₄) content reduced. The latter process step can also be handled within the methanol loop. In the methanol loop the methanol synthesis loop purge stream, which has a much higher concentration of CH₄, can be put through a CH₄ reformer unit (see Chapter 1.2.3, Auto-Thermal reforming).

Methanol synthesis

After the crude gas has been adjusted to syngas quality the syngas is fed to the methanol synthesis and purification units.

1.1.3 Plant Performance

The process work was initiated in November 1996 with the provision of the Design Basis to Battelle Columbus. In principle, it was a desire to make the gasification part as free-standing as possible and Battelle Columbus therefore used sensible heat from the gasification unit to dry the feedstock.

A couple of iterations were made to narrow in on desired gas quality. During this process it became evident that the gasifier performance differed considerably to the negative from the values originally presented.

The deviations were most clear in the following three areas:

- a) Content of non-desirable hydrocarbon components in the crude gas,
- b) Cold gas efficiency (heating value of the crude gas divided by heating value of the feedstock),
- c) The H₂/CO ratio in the crude gas.

On these three issues background information had previously been provided by Battelle Columbus, namely:

- a) Gas Clean-up for Combined Cycle Power Generation Using a Hot Gas Conditioning Catalyst (August 21 - 24, 1995), see Attachment 1,
- b) R. Katovsky's thesis from Princeton University, pages 67 - 74 (June, 1993), see Attachment 2,
- c) Hot-Gas Conditioning of Biomass Derived Synthesis Gas (August 30, 1993), see Attachment 3.

The initial performance data presented by Battelle Columbus could not be substantiated as the project moved forth towards final technology selection. The Battelle Columbus process was subject to a significant downward revision of its net efficiency in primary energy to methanol conversion.

It was also clear that the process configuration had to be supplemented with a process unit for gas conditioning additional to what had been claimed in the initial discussions with Battelle Columbus.

As a consequence of this it was decided to base the project work on the second technology alternative as presented in Chapter 1.2.

Note:

Battelle Columbus has been asked to further comment the above results but has so far declined.

In addition, Princeton University has been asked to comment. They have agreed to that the quoted table in Katovsky's thesis is incorrect but claims that the overall result presented in the thesis is nonetheless correct.

Battelle Columbus has also agreed to that said table contains the described error.

Gas Cleanup for Combined Cycle Power Generation Using a Hot Gas Conditioning Catalyst

Presented at:

Second Biomass Conference of the Americas
Portland OR, August 21-24, 1995



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wood weight exit the gasifier as condensable tar material. In conventional scrubbing systems, these tars are quenched producing a waste water stream laden with tars that must be cleaned before discharge from the plant. In addition to the waste water cleanup issue, wet scrubbing systems can leave a small fraction (10 to 30% by weight of the original 0.5 to 1%) of the tars in the gas as a fine aerosol mist that is difficult to remove, but can create problems in downstream equipment such as compressors or turbines. We have observed this in testing with the compressor-turbine system installed in the Battelle PRU.

Experimental Evaluation of DN34

Battelle has identified and tested a low cost, disposable cracking catalyst (DN34) that can be used to condition the product gas prior to its use in a power generation system. This catalyst has been evaluated by Battelle at the PRU scale in slip-stream and full scale tests, and has been studied at the microscale at NREL to better understand catalyst activity and selectivity (Rejai, 1992). DN34 is a very effective high-temperature water gas shift catalyst. Near equilibrium concentrations of H₂ are obtained when raw product gas is shifted over DN34. This makes conditioned product gas better suited for advanced power generation cycles that use fuel cells while simplifying gas processing.

DN34 Testing with the Battelle PRU

DN34 was evaluated under actual biomass gasification conditions in two separate reactor systems in the Battelle gasification PRU. These systems were, a 6-inch diameter slip stream reactor and a 36-inch diameter reactor designed to process the complete output of the PRU gasifier. The following sections discuss results of tests in both of these reactor systems.

Tar Concentration Measurements

Tar concentration in the product gas was measured by sampling using a modified method 5 (MM5) train. The MM5 train consist of a series of 5 impingers place in an ice bath followed by a dry gas meter to measure the quantity of gas sampled. After sampling, the impingers are rinsed with toluene to remove the tars and water collected. Toluene and water are removed from the samples by heating in an oven at 65C overnight. Tar concentration at the inlet and outlet of the catalyst bed was then compared to determine tar destruction efficiencies.

Catalyst Testing Results

Catalyst DN34 was operated at conditions similar to the gasifier conditions (815C and 17 kPa) to simulate adiabatic operation of the catalyst system in both the slip stream and "full-flow" catalyst chambers. Data generated on tar destruction showed that high levels of tar destruction were possible with the catalyst and that following gas cooling in the PRU scrubbers, essentially all tars were removed from the product gas (Paisley 1993, 1994).

Tar concentrations measured in the product gas from the Battelle gasifier are typically $1.6 \times 10^{-2} \text{ kg/m}^3$. These tars are highly aromatic in character and are relatively insoluble in

water. As discussed above, these tars can be removed by conventional water based scrubbing, however such scrubbing methods can leave significant amounts of tar behind in the product gas. Table 1 below shows tar removal rates as the tar laden product gas from the Battelle PRU passes through the venturi and spray tower scrubbers in the PRU.

DN34's relatively poor activity for methane steam reforming (< 20% conversion) is actually a benefit in power applications. This is because the water-gas shift is moderately exothermic (equation 2) and the total concentration of tar is quite small compared to the amount of CO and H₂O undergoing the shift. While the steam reforming reactions for the tar compounds are endothermic (equation 1), the small amount of tar results in a relatively small overall heat demand. In a system where CH₄ is totally steam reformed (e.g. using as supported Ni catalyst) there will be a significant heat requirement in the syngas conditioning reactor (see equation 3). The use of autothermal reforming by adding a small amount of O₂ has been suggested to avoid adding external heat to the catalytic hot-gas conditioning reactor.

Table 1. Tar Removal Efficiencies - PRU Scrubber

Gasifier Temperature, C	Tar Production, kg/m ³	Scrubber Outlet, kg/m ³
~700	3.2×10^{-2}	5.3×10^{-3}
~800	2.3×10^{-2}	1.5×10^{-3}

By providing hot gas conditioning of the product gas using DN34, significant improvement in the quality of the product gas can be realized. Table 2 shows the data generated using DN34 both in the slip stream reactor and in the "full flow" catalyst reactor in the PRU.

Table 2. Hot Gas Conditioning Test Results

Gasifier Temperature, C	Tar Production, kg/m ³	Catalyst Unit	Catalyst Outlet, kg/m ³
~800	2.3×10^{-2}	Full Flow	1.4×10^{-3}
815	1.9×10^{-2}	Slip Stream	1.4×10^{-3}
815*	1.2×10^{-1}	Slip Stream	3.6×10^{-3}

* Green wood feedstock

When using the "full flow" catalyst chamber, the gas exiting the reactor was passed through the existing PRU scrubbers prior to compression (Paisley, 1994). This process removed all remaining tars from the gas resulting in no measurable residual tars leaving the scrubber system. Figures 3 and 4 show the MM5 impingers at the exit of the gasifier and

at the exit of the scrubber with the DN34 catalyst system in use. Figure 4 illustrates the cleanliness of the product gas possible using the DN34 catalyst system. The cleanliness of the product gas is further illustrated by measurements made of the water discharged from the PRU scrubber. Without the catalyst chamber in line the scrubber discharge water (single pass) showed an organic carbon content of 1200 mg/l but with the catalyst chamber in line this value was cut by over 90% to 110 mg/l.

Water Gas Shift Activity of DN34

A second benefit of hot-gas conditioning, when applied to advanced power generation cycles, such as those including fuel cells is that the hydrogen content of the gas increased. Fuel cell applications require high hydrogen content fuel gases. By using catalyst DN34, the hydrogen content of the product gas can be raised to a level so that no further water gas shift reaction is necessary in the fuel cell system. The product gas leaving most biomass gasifiers, including the Battelle gasifier, has a H₂ to CO ratio much less than 2:1 and in some cases less than 1:1. The ratio can be adjusted by the water gas shift reaction shown by equation (2) earlier.

This reaction requires the presence of a catalyst to enhance reaction rates. Catalyst DN34 is effective in enhancing the water gas shift reaction to produce a gas with a high H₂ to CO ratio as shown in Table 3 below. With a hydrogen content in excess of 60% no further conversion of CO to H₂ would be necessary for fuel cell applications.

Table 3. Water Gas Shift Results Catalyst DN34, Battelle PRU

COMPONENT	RAW PRODUCT	DN-34 OUTLET
H ₂	24.64	60.45
CO ₂	18.24	31.84
C ₂ H ₄	4.61	ND
C ₂ H ₆	0.46	ND
C ₂ H ₂	0.56	ND
N ₂	8.40	3.36
CH ₄	10.46	2.46
CO	32.64	1.89

NOTE: All analyses dry basis.

These water gas shift results are of a preliminary nature, but show the significant potential for DN34 as a means of achieving gas compositions suitable for supplying fuel cell power systems.

Thermodynamic equilibrium calculations were done to determine the approach to equilibrium of tar destruction, the water-gas shift and methane steam reforming reactions. Equilibrium compositions were calculated for temperatures between 700C and 1000C in 5° C increments for various measured inlet gas compositions taken during PRU catalyst testing. The predicted H₂/CO ratios were between 2.3 and 5.7 at 815C with the exact value depending on the CH₄ steam reforming activity assumed for DN34 and the inlet gas composition (especially the steam content). Equilibrium calculations where all reactions are kinetically allowed gives the result that the concentration of CH₄ is essentially zero at T ≥ 800C. To simulate the low CH₄ steam reforming activity of DN34, the calculations were done substituting argon for methane in the inlet gas. We are presently investigating possible reasons for the differences between the equilibrium predictions and the measured H₂/CO ratios shown in Table 3.

Microscale DN34 Tests with Model Systems

Two types of microscale experiments have been done with DN34 at NREL; catalyst testing in a 25 gram fixed bed microreactor system that used gas chromatography for product analysis, and testing with a second smaller (4 gram) fixed bed reactor interfaced with a molecular beam mass spectrometer (MBMS). In both experiments, H₂, and a 3:1:1 mixture of CO, CO₂ and CH₄ were used to formulate a synthetic product gas. Two high pressure liquid pumps feed water and benzene for steam and "tar" feed. Benzene was used as a tar surrogate because it is present in the tar in significant quantities and is one of the more difficult components to destroy. DN34 was tested using 50 vol% steam using a gas hourly space velocity (GHSV) of 1950 m³_{gas}/m³_{catalyst}/hr, at 765C. The inlet H₂/CO ratio for the DN34 test was 0.68. The measured H₂/CO ratio exiting the microreactor was 2.6 (Jacoby et al 1995). This was in excellent agreement with the thermodynamic equilibrium value of 2.5 obtained from the low methane steam reforming activity calculation.

The catalytic chemistry of tar destruction over DN34 was examined at NREL in a series of microscale experiments where product analysis was done using the molecular beam mass spectrometer (MBMS). The temperature, inlet gas flow rate and composition were similar to those used described above, except that a smaller reactor was close coupled to the sampling port of the MBMS to reduce the condensation of the high molecular weight polycyclic aromatic hydrocarbons used in these studies. Lower conversions of benzene, naphthalene, phenanthrene and pyrene were observed compared to toluene, phenol, m-cresol, 1-methylnaphthalene, and 1,4-dimethylnaphthalene. This work is discussed in detail elsewhere (Gebhard et al., 1994a).

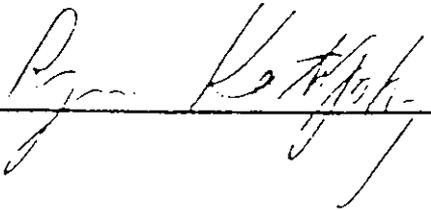
The microreactor experiments with toluene steam reforming over DN34 have shown that steam dealkylation is an important reaction path. In the case of toluene, benzene and methane are formed. Similarly, substituted naphthalenes will yield naphthalene and methane, water, etc., depending on the identity of the substituent group. The observation that substituted aromatic hydrocarbons are more easily destroyed over DN34 can be explained by higher rates of steam dealkylation than the rates of total steam reforming to produce additional syngas. This chemistry is consistent with observations from real time analysis of the Battelle PRU product gas before and after conditioning with DN34. In that work, the NREL transportable molecular beam mass spectrometer (TMBMS) showed that

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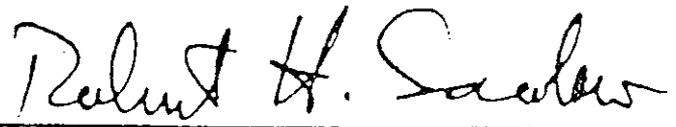
Submitted in partial fulfillment of the requirements for the ^{degree} of Master of Science in Engineering from Princeton University

PU/CEES Report No. 279
June 1993

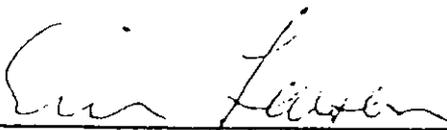
Prepared by:



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Professor Robert Socolow
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Dr. Eric Larson
Co-Advisor



Dr. Robert Willi
Co-Advisor

Table 3-5: Operating characteristics of the gasifiers evaluated in this thesis. n/a = not available.

	Directly heated gasifiers			Indirectly heated gasifiers		
	IGT ^d	Shell-coal ^f	Shell-bio ^e	WM ^a	MTCI ^c	BCL ^b
Bed type	Bubbling-fluidized	Entrained	Entrained	Rotor kiln	Bubbling-fluidized	Fast-fluidized
Feedstock characteristics						
Type	wood	coal	wood	wood	wood	wood
Dry, ash free composition	CH _{1.52} O _{0.68}	CH _{0.91} O _{0.11}	CH _{1.52} O _{0.68}	CH _{1.5} O _{0.67}	CH _{1.63} O _{0.66}	CH _{1.46} O _{0.60}
HHV (GJ/dry tonne)	19.28	28.21	19.28	20.93	19.40	20.12
Moisture % (after drying)	15	5	11	45	20	10
Additional inputs						
Steam (kg/kg dry feed)	0.3	0.028	0.03	0	1.37	0.314
Oxygen (kg/kg dry feed)	0.3	0.8	0.45	0	0	0
Air (kg/kg/dry feed)	0	0	0	0	2.10	1.46
Reactor characteristics						
Exit temperature (°C)	982	1371	1085	600	697	927
Pressure (MPa)	3.45	2.43	2.43	1.5	0.101	0.101
Throughput (dry kg/m ² -s)	1.9	n/a	n/a	0.18	0.07	2.7
Solids residence time	minutes	~1 second	~1 second	~1 hour	minutes	~1 second
Product gas characteristics						
Yield (kmol/dry tonne)	82.0	92.4	79.3	90.3	138.6	58.3
Molecular weight (kg/kmol)	22.52	20.53	20.08	21.82	17.65	21.22
HHV (MJ/kg wet)	8.67	12.62	10.33	11.11	9.35	14.22
Composition ^g (mole % wet [dry])						
H ₂ O	31.8 [0]	2.1 [0]	18.4 [0]	42.0 [0]	49.5 [0]	30.8 [0]
H ₂	20.8 [30.5]	31.8 [32.5]	30.7 [37.6]	12.0 [20.7]	25.3 [50.0]	14.6 [21.1]
CO	15.0 [22.0]	64.3 [65.7]	39.0 [47.8]	4.0 [6.9]	11.2 [22.1]	32.4 [46.8]
CO ₂	23.9 [35.0]	1.7 [1.7]	11.8 [14.4]	22.0 [37.9]	9.9 [19.4]	7.8 [11.3]
CH ₄	8.2 [12.0]	0 [0]	0.1 [0.1]	20.0 [34.5]	4.0 [8.0]	10.3 [14.9]
C ₂ +	0.3 [0.5]	---	---	---	0.2 [0.4]	4.2 [6.1]
Net carbon conversion ^h (%)	96.2	>99	100	98.5	83.5	80.3
Cold gas efficiency ⁱ (%)	83.0	80.4	85.3	85.9	87.8	87.4

- (a) WM is the Wright-Malta gasifier. Based on Hooverman (1979), Coffman (1981), and Coffman (1991).
- (b) BCL is the Battelle-Columbus Laboratory gasifier. Based on Paisley (1991). In all cases, "kg dry feed" refers to the total biomass feed for the combustor and gasifier beds.
- (c) MTCI is the Manufacturing and Technology Conversion International gasifier. Data are based on Durai-Swamy, et al. (1991a).
- (d) IGT is the Institute of Gas Technology. Data is based on OPPA (1990).
- (e) Shell-bio is the Shell entrained-bed gasifier operating on biomass. For a complete description of the assumptions used to estimate the performance of this system see Appendix 3A.
- (f) From Synthetic Fuels Associates (1983).
- (g) Small quantities of N and S containing compounds and Argon are not shown.
- (h) This is the net carbon conversion to synthesis gas. For the BCL and MTCI gasifiers, carbon converted to gas in the combustors is counted as part of the feed carbon but not as carbon converted to gas.
- (i) Defined as the higher heating value of the product gas divided by the higher heating value of all energy inputs to the gasifier, including the combustors in the BCL and MTCI cases and the external heating in the WM case. The energy for raising steam or drying biomass is not included.

only the HTW (High Temperature Winkler) technology has been commercially applied to producing chemicals; since 1988, an HTW unit has operated successfully on peat at a Finnish ammonia plant (Larson and Katofsky, 1992).¹⁶

A number of circulating fluidized-bed biomass gasifiers have also been developed (Figure 3-14). Lurgi (Germany), Ahlstrom (Finland), Studsvik (now TPS) and Gotaverken (Sweden) have several atmospheric pressure units operating commercially (Larson et al., 1989). Lurgi (with Rheinbraun/Uhde) and Ahlstrom are also developing pressurized CFB gasifiers, of which there are no commercial units. These gasifiers are being developed primarily for heat and power generation, but if they are proven commercially, there is no reason why they cannot be adapted to fluid fuels production, provided they are oxygen-blown units.

(iii) Entrained-bed Gasifiers

The third basic type of directly heated gasifier is the entrained-bed gasifier (Figure 3-7). This design was originally developed for coal, and unlike fluidized- or fixed-bed gasifiers, there is no operating experience with biomass. In entrained-bed coal gasification, pulverized coal is fed into the reactor dry or in a water slurry, where it reacts with a large amount of oxygen (approximately 0.8-1.0 kg/kg dry coal). The resulting high operating temperatures (1300-1800°C) completely gasify the coal and produce a tar-free gas that is close to chemical equilibrium with virtually no CH₄ and no higher hydrocarbons. At the high temperatures characteristic of entrained-bed oxygen-blown coal gasification, the ash melts and is removed as slag.

¹⁶ Air blown gasification can be used for ammonia synthesis since it requires nitrogen in the feed.

temperature as a dry feed system.²⁰ The pressurized operation of the Shell gasifier is also a desirable characteristic.

The operating conditions for the Shell-biomass gasifier shown in Table 3-5 have been selected based on the parametric analysis carried out earlier in this chapter, where it was shown that for a moisture content on 11%, the optimum syngas composition corresponded to an oxygen consumption rate of 0.45 kg/kg dry feed (see Figure 3-8). At lower temperatures, there is more methane in the synthesis gas, CO+H₂ production is less than the maximum, and some tars may be produced. At higher temperatures, there is virtually no methane in the product gas so that additional oxygen consumption would only oxidize some of the CO and H₂ to CO₂ and H₂O.

3.4.2 Indirectly Heated Gasifiers

Although there are no commercial facilities using indirect biomass gasification, pilot scale units have successfully operated using a variety of heat transfer methods (Figure 3-15). Operating conditions vary from atmospheric pressure to 1.5 MPa and exit temperatures of 600-930°C (see Table 3-5). Because of their low operating temperatures, indirectly heated units typically produce gases that contain significant quantities of hydrocarbons (mostly methane) and some tars. Low temperature operation also means that low carbon conversion is a potential problem, and each gasifier described here uses a different method for ensuring high carbon conversion rates. For fluidized-bed units, steam is the fluidizing medium, although recycled product gas can also be used.

²⁰ A slurry feed is practical for coal gasification because the high temperatures must be moderated in some manner. With biomass, gasification temperatures would not be high enough to warrant the use of a slurry to moderate the temperature. Rather, you want to minimize the energy required to achieve a certain temperature. Thus a slurry would achieve the opposite effect with biomass.

(i) *The Battelle-Columbus Laboratory (BCL) Gasifier*

The Battelle-Columbus Laboratory (BCL) gasifier is an atmospheric pressure, twin bed, fast-fluidized-bed unit (Feldmann et al., 1988; Paisley, 1991; Wyman et al., 1992) (Figure 3-16) that resembles fluid catalytic crackers commonly used in the petrochemical industry (Refining Handbook, 1990). The term fast-fluidized-bed has been used to describe the BCL system since the specific throughput rate of the BCL gasifier is significantly higher than a conventional fluidized-bed like the IGT gasifier (see Table 3-5). To date, the BCL gasifier has operated at a dry feed rate of up to 500 kg/hr, and has over 10,000 hours of operating experience on a variety of biomass feedstocks (Paisley, 1993).

In one bed, biomass (dried to 10% moisture) is pyrolyzed in steam at temperatures up to 930°C. Ash, char and sand (the bed material) are entrained in the product gas, separated using a cyclone, and sent to a second bed where the char is burned in air to re-heat the sand. The heat is transferred between the two beds by circulating the hot sand back to the gasification bed. This is similar to char recirculation systems used with commercial fluidized-bed coal combustors (Paisley, 1993). This system allows one to provide heat by burning some of the feed, but without the need to use oxygen because combustion and gasification occur in separate vessels, and is possible with biomass because of its high volatile content. As a way of producing higher temperatures in the gasification bed, some additional biomass can be burned directly in the combustor to augment the heat provided by the combustion of char alone. Ash is rejected from the combustor, and no char is produced. Tars and oils (approximately 1% of the feed) are collected from the quench water and recycled to the combustor. Therefore, the BCL system is characterized by nearly 100% overall carbon conversion. An advantage of this system is the inherently high heat transfer rates of the direct mixing of hot sand and

biomass. Therefore, although the system as a whole is indirectly heated, biomass heating rates are high.

If the BCL gasifier were to operate in *balanced mode*, where the heat available for gasification is that produced by burning char alone in the combustor, gasification temperatures would be low (approx. 800°C) with correspondingly low carbon conversion in the gasifier (<70%). Using a heat and material balance model developed at BCL, it was found that by increasing the gasification temperature, the overall production of synthesis gas increases, even though some additional biomass must be burned in the combustor (Paisley, 1991). Table 3-5 presents operating data for this type of configuration. A gasification temperature of 927°C results in good carbon conversion in the gasifier (88%) while limiting the amount of biomass which must be burned in the combustor to less than 10% of the total feed. Thus, overall carbon conversion to synthesis gas is 80.3%, and because of the rapid indirect heating, CO₂ accounts for only 11.3% by volume of the synthesis gas on a dry basis. Higher temperatures are possible, but would result in more heat leaving the combustor as waste heat in the flue gases²¹ rather than being transferred to the gasifier by the circulating sand.

The BCL gasifier illustrates one way in which high overall carbon conversion can be achieved with indirect heating; design a reactor that actually relies on poor carbon conversion for its operation. To accomplish this requires two basic conditions, which can be explained in terms of the earlier discussion of gasification fundamentals: avoid high peak temperatures and keep solids residence times very short compared to the characteristic reaction time of the char-steam reaction. The first condition is satisfied by the use of indirect heating, and the second is by design: the BCL gasifier is designed to

²¹ The combustor bed was assumed to have an exit temperature 167°C higher than the gasifier operating temperature (Paisley, 1991).

have residence times on the order of 1 second (Paisley, 1993). As a result, the char produced during pyrolysis is not gasified, but is entrained within the product gas. Pyrolysis is the only step that occurs to any significant extent in the gasifier. Since temperature and heating rate are the most important variables affecting pyrolysis, it is not surprising that during the testing of the BCL gasifier, changing the fluidizing medium from steam to nitrogen had a negligible effect on both carbon conversion and the inert free gas composition (Feldmann, et al., 1988). Furthermore, since less char is produced at higher temperatures and heating rates, it is also not surprising that carbon conversion in the gasification bed was found to be directly proportional to the bed temperature. Since char gasification need not occur in the gasification bed, the specific throughput rate is very high, which helps reduce the capital cost of the BCL gasifier.

In summary, the heat and carbon balance of the BCL system is quite different from all other gasifiers discussed here. Not only is the endothermic char-gasification step bypassed, but essentially all the char that is produced is burned, as is some of the feed. One would therefore expect there to be a significant amount of waste heat produced in the BCL process, a result that is evident in the analysis of Chapter 6. When coupled with the high cold gas efficiency of the BCL gasifier, this high grade waste heat, which can be used to generate a fair amount of electricity, makes it the most efficient gasifier included in this thesis.

(ii) The Manufacturing and Technology Conversion International (MTCI) Gasifier

A second indirectly heated gasifier is being developed by Manufacturing and Technology Conversion International (MTCI), a small US company (Figure 3-17). The MTCI gasifier combines an atmospheric pressure bubbling fluidized-bed with pulse-combustion-enhanced in-bed heat exchange tubes that provide the heat for gasification (MTCI, 1990; Durai-

Hot-Gas Conditioning of Biomass Derived Synthesis Gas

To be presented at

The First Biomass Conference of the Americas

August 30, 1993

by

Mark A. Paisley and Robert D. Litt



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Other commercial cracking catalysts might be potential candidates, but testing such as conducted during this program will be necessary to determine their effectiveness before commercial applications are planned.

The DN34 catalyst, on the other hand, showed a high level of tar destruction during the tests run. C₂+ hydrocarbons were essentially eliminated from the product gas during all tests. Methane cracking was initially high, reducing incoming levels by 2/3. The methane cracking activity tended to reduce with time to a point where only 12 percent of the incoming methane was consumed in the reactions.

Water Gas Shift Reactivity

The other reaction of interest for a hot-gas conditioning catalyst is the water gas shift reaction. During these tests, ICI 46-1 catalyst showed very little shift activity. As shown in Table 2, the H₂ to CO ratio was increase from 0.64:1 to slightly over 1:1 while the DN 34 catalyst showed an exit ratio of 2.8:1 with the same feed gas and reactor conditions.

Table 2. Water Gas Shift Data DN-34 and ICI-46-1

Inlet H ₂ /CO Ratio	Steam Conc. -%	DN-34 Outlet H ₂ /CO	ICI-46-1 Outlet H ₂ /CO
0.70	25	1.74	
0.64	35	2.80	1.15
0.73	40		1.31
0.59	45	3.06	
0.55	65	12.25	

Water gas shift reactivity with DN 34 remained high throughout the tests with the catalyst. The results with the DN34 catalyst are shown in Table 3 below. Comparing these results with the water concentrations in the product gas, shown in Table 2, shows that the H₂ to CO ratio can be adjusted based on the inlet steam concentration. At higher steam concentrations in the product gas, a higher H₂ to CO ratio can be realized at the outlet of the catalyst chamber. Low steam concentration tests (25 to 35 percent) provide ratios of 1:1 to 2:1 while a high steam concentration (47 percent) provides a product gas with a ratio of as high as 12:1. The relationship between inlet steam concentration and the final H₂ to CO ratio is illustrated in Figure 3.

**OUTLET H₂ TO CO RATIO
VS. INLET STEAM CONCENTRATION
DN34 CATALYST**

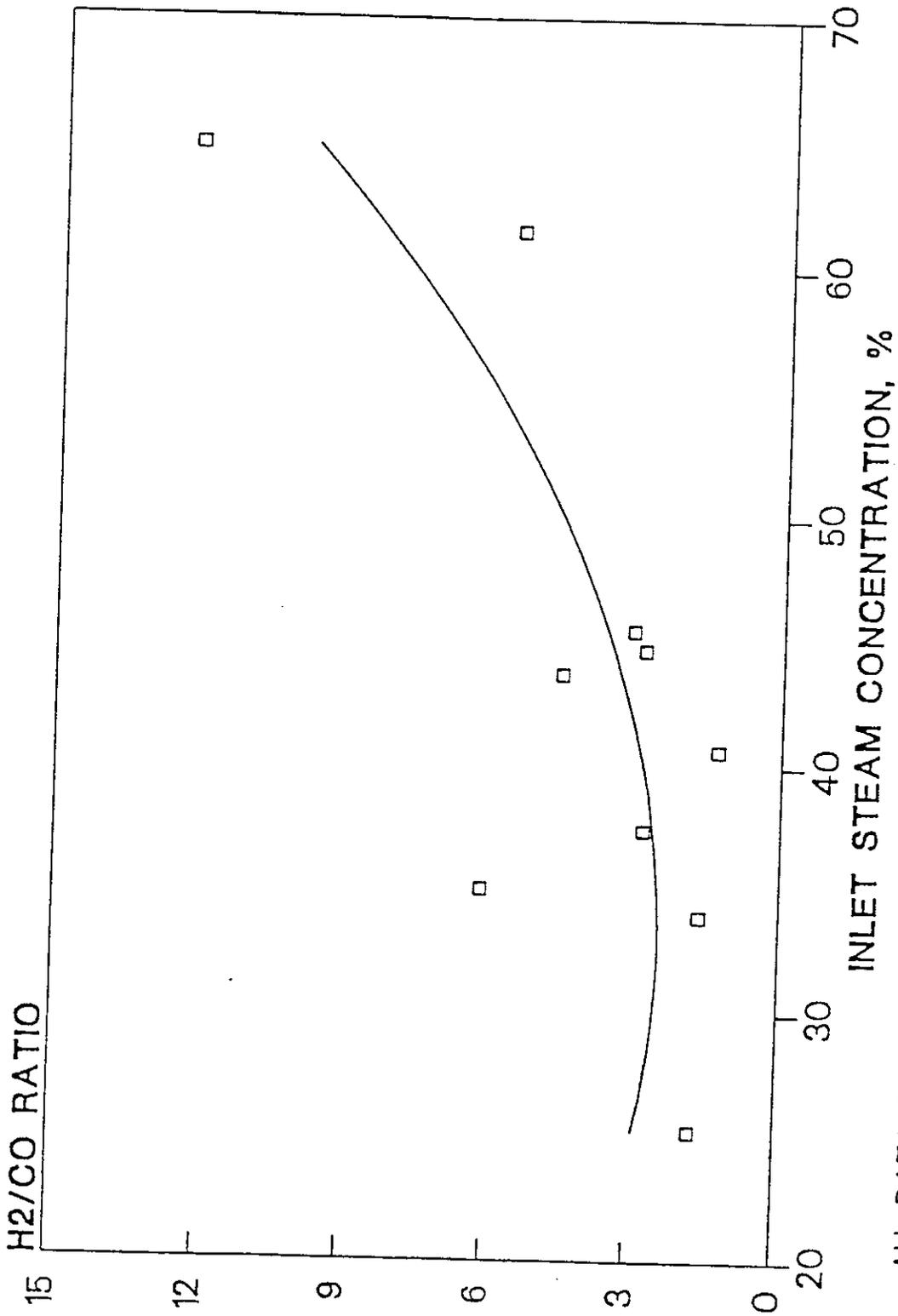


Figure 3. Outlet H₂ to CO Ratio versus Inlet Steam Concentration DN34 Catalyst

Table 3. Product Gas Analyses DN-34 Catalyst Tests Nitrogen Free

	Typical Inlet Gas Conc. -%	25% H ₂ O Feed Gas Outlet Gas Conc. -%	35% H ₂ O Feed Gas Outlet Gas Conc. - %	65% H ₂ O Feed Gas Outlet Gas Conc. - %
H ₂	26.1	36.73	43.12	55.49
CO ₂	15.2	25.42	27.64	22.31
C ₂ H ₄	5.2	1.68	--	0.39
C ₂ H ₆	0.4	0.54	0.23	--
C ₂ H ₂	0.6	--	--	--
CH ₄	14.7	14.52	13.62	7.28
CO	37.8	21.10	15.39	14.53

DN 34 Performance over a Range of Conditions

The initial success with DN34 led to a decision to examine its performance through a factorially designed experimental matrix during the switch grass testing. This matrix was developed jointly by Battelle and NREL to examine high and low temperatures, high and low space velocities, and high and low steam levels. The baseline matrix used is shown below in Table 4.

Table 4. Baseline Test Matrix
DN34 Catalyst Testing

Chamber	Temperature C°	Space Velocity	Steam Concentration -%
A	815	1500	58
B	650	2500	40
A	650	1500	58
B	815	2500	40
A	650	2500	58
B	815	1500	40
A	815	2500	58
B	650	1500	40

1.2 Methanol Production via HTW Technology

1.2.1 Concept Principle

The overall concept is illustrated in Figure 1.2, Simplified block flow diagram methanol from biomass. The methanol production plant comprise a gasification plant, the design of which is supplied by the engineering company Krupp Uhde, Germany, and a methanol synthesis plant, which is supplied by the engineering company Haldor Topsøe, Denmark.

The gasifier is operated at an elevated pressure of 27 bar a and uses oxygen/steam as gasification agent. The feeding system of wood biomass uses inert gas taken as nitrogen (N₂) produced by the air separation unit and carbon dioxide produced by the carbon dioxide/hydrogen sulphide removal unit. The gasification unit is followed by a gas conditioning step.

The syngas produced syngas from the gasification plant is additionally compressed in the methanol synthesis plant in order to meet demands of the methanol synthesis unit.

In the methanol synthesis plant an Auto-Thermal Reformer (ATR) is fed by a recycled purge gas stream from the methanol synthesis unit, thus making a gas stream loop which maximises the methanol yield. Produced distilled methanol is then taken to storage. However, two alternative concept configurations for the methanol synthesis plant can be considered:

- A steam reforming step in the main crude gas stream. This case is most unlikely, since the volume is considerably larger than in the purge stream and the methane concentration is much lower.
- An Once Through Methanol (OTM) concept with a non-conditioned crude gas stream fed once through a methanol synthesis reactor. The non-reacted gases are fed to a combined cycle for power generation. In such a case the plant may be a net electricity producer.

1.2.2 Study objectives

The main objectives of this part of the study are:

- Development of an appropriate process configuration for the production of methanol from wood,
- Preliminary conceptual design of the plant working by the selected concept,
- Rough estimate of plant performance and of capital investment cost as a basis for the assessment of production cost.

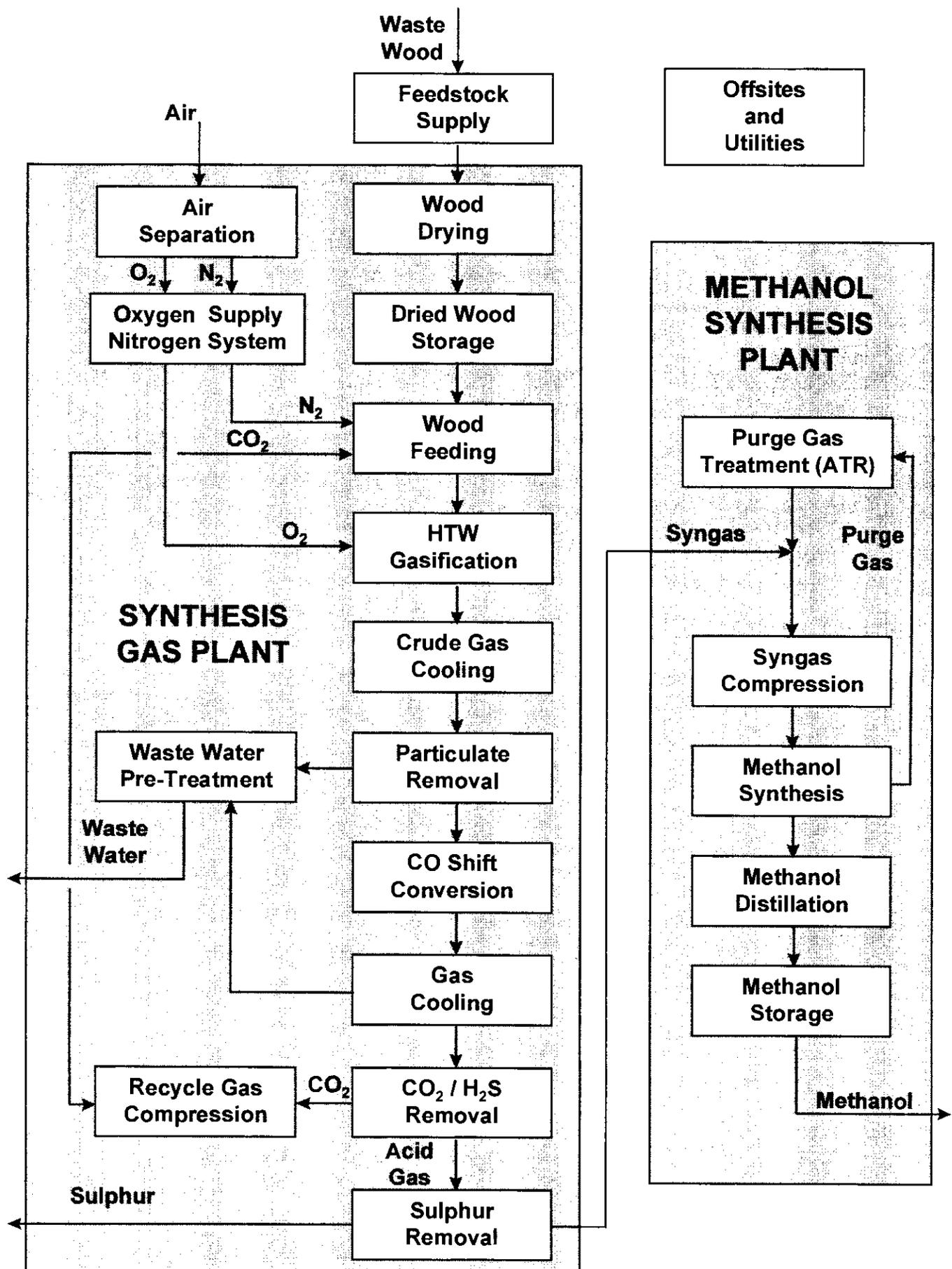


Figure 1.2. Simplified block flow diagram methanol from biomass.

1.2.3 Process Description

Overall plant description

The prepared wood feedstock is gasified in a fluidized-bed reactor at high temperature and high pressure to produce a crude gas. The crude gas mainly consists of a mixture of carbon monoxide (CO), hydrogen (H₂), carbon dioxide (CO₂), methane (CH₄) and water vapour (H₂O). The produced crude gas is then cooled for heat recovery and scrubbed with water for particulate removal.

In the following step the gas passes through a CO-shift conversion unit where the carbon monoxide content is reduced and an equivalent amount of hydrogen is produced. The gas is further cleaned from sulphur in a sulphur removal step. The clean fuel gas, containing an acceptably low concentration of sulphur, and virtually no metals or particulate is then utilised as make-up synthesis gas in a methanol synthesis unit.

A purge gas stream is led to an auto-thermal reformer where the methane is converted with oxygen and steam back to make-up synthesis gas quality and recycled to the methanol synthesis unit. The produced methanol is upgraded to fuel-grade quality in a distillation unit and then stored in tanks that are connected to a distribution line for tank trucks.

Steam is produced in the crude gas cooling unit and in the methanol synthesis and in the ATR unit. This steam is utilised as a gasification agent with oxygen or as a heat source in different plant units. Additional steam is produced in a boiler by wood burning. Surplus super-heated steam of high pressure not consumed in the plant is expanded in a steam turbine that generates electric power. This power is consumed by various units within the plant.

Process sequence plant configuration

The process sequence and the interaction of the individual plant sections within the synthesis gas plant complex are represented schematically in the overall block flow diagram in Figure 1.2. The gasification part of the plant is configured as three trains up to the particulate removal unit, thereafter one common train.

Wood drying

From the wood preparation and storage unit the wood chips may be transported to the drying unit by means of belt or chain conveyors or other conveying devices. (This feeding system is not part of this study.) The drying unit consists of three identically equipped drying trains and one condensate system which serves all three trains.

The feedstock is dried in a fluidized bed using immersed heating surfaces. Dust contained in the vapours produced in the drier is removed (by cyclone or filter). After cleaning, the vapours are re-circulated for fluidisation in the drier and, after re-compression, are used as a heating agent in the submerged heat exchanger mentioned.

The hot vapour condensate can be used for preheating the raw feedstock. The vapour condensate obtained has a high quality and is utilised as make-up water for Boiling Feed Water (BFW).

Dried wood storage

The dried wood chips are withdrawn from the wood cooling unit of the drying facility and conveyed to the top of the dried wood silo by means of the chain conveyor. At the discharge section of the chain conveyor, the material passes through a star feeder and drop into a silo. The silo which is filled completely during normal operation has a storage capacity of about 1000 m³.

In case of disturbances in the drying unit or in a upstream facility this quantity enables running of the gasification unit for a period of approx. 5 h based on normal load. The silo will be designed as a shock resistant vessel (3 bar g). To prevent the silo from exceeding its design strength in the event of dust explosions, explosion doors are placed on the silo roof.

To detect smouldering fires early, the silo atmosphere will be monitored for its CO content; temperature measuring elements serve for the same purpose. In case of a fire, the silo can be isolated and blanketed up with nitrogen. The star feeders at the inlet and outlet are designed and constructed as explosion pressure shock resistant units. They also serve as flame barriers.

Wood char is used for the start up of the gasifiers. The char is delivered pneumatically into the char silo. The conveying gas (N₂ or air) is discharged into atmosphere after cleaning in the silo filter. The transportation of the char into the individual charge bins is effected via conveying lines by the conveying bin, which operates by the principle of pressure vessel conveying.

Low Pressure (LP) N₂ is used as conveying gas, which is discharged out of a charge bin to the atmosphere via the pressure balance pipe to a lock hopper, the lock hopper, the lock hopper exhaust gas line and the silo filter. The silo has a capacity of approx. 60 m³, thus the quantity required for starting up all 3 gasifiers at the same time can be stored.

Similar protection measures against the dangerous effects of dust explosions are provided for the char silo as they are included in dried wood silo.

Wood feeding

The wood feeding unit constitutes the link between the dried wood storage unit and the HTW gasifier. In the wood feeding unit the pressure of the wood is increased from ambient pressure to gasification pressure and the wood is transported into the gasifier. Both functions are performed by two identically arranged lock hopper systems per gasifier.

The wood feeding system of a gasifier consists of two complete trains, with the exception of the CO₂-buffer tank which serves for both trains. A lock hopper system essentially comprises four vessels: one feed bin, 2 lock hoppers and one charge bin. In addition, the valves arranged between these bins, a star feeder, a screw conveyor and a charging screw are required. The process proceeds as follows.

The feed bin is continuously charged with wood from the dried wood storage by means of the chain conveyor. The wood is fed at ambient pressure from the feed bin via a gravity line into one of the two lock hoppers arranged in parallel where the pressure is subsequently raised to gasification pressure by means of CO₂.

Thereafter, the wood is fed into charge bin which is under gasification pressure at all times. From the charge bin, the wood is continuously fed into the gasifier by means of the star feeder being the metering device and the downstream screw conveyor and the charging screw. Before the wood lock hopper is recharged, it must be de-pressurised to ambient pressure by discharging the CO₂ into the atmosphere via silo filter.

The two lock hoppers are charged and discharged alternately. This concept and by use of the CO₂-buffer tank results in considerable energy savings for CO₂ compression. Furthermore, the flow rate of lock hopper exhaust gas, to the atmosphere is reduced accordingly.

HTW-gasification

For the gasification of the wood feedstock, the High-Temperature Winkler (HTW) process will be used. In this process, gasification takes place in a fluidized bed at a pressure of 27 bar a. This pressure was found to be optimal pressure for this methanol producing plant. The gasification system mainly comprises a gasifier and a cyclone with a gravity pipe.

The gasifier is a pressure vessel provided with a refractory lining. The lower section of the gasifier is cone-shaped to allow for the increase in the gas volume during the gasification process. The upper section of the gasifier is cylindrical.

The gasifier can be subdivided into two zones: The highly expanded fluidized-bed, which has a relatively uniform temperature (in the order of 700 °C when gasifying wood), is the lower zone. The superposed zone is termed "free board" or "post-gasification zone".

The gasification agent (oxygen) is not only admitted to the fluidized-bed but also into the free board in order to decompose undesirable reaction by-products (tars) and to gasify entrained char by raising the temperature (approx. to 900 °C).

The crude gas which leaves the gasifier at the top contains a considerable amount of entrained particulate. The major portion of this particulate is separated from the crude gas in the hot cyclone and returned into the fluidized bed of the gasifier through a gravity pipe. A crude gas analysis on dry basis reveals the following gas composition:

• CO	35.6 mol %
• H ₂	28.6 mol %
• CO ₂	28.8 mol %
• CH ₄	6.7 mol %
• Trace gases	0.3 mol %
<hr/>	
Total	100.0 mol %.

As is the case with the gasifier, the cyclone and the gravity pipe are provided with a refractory lining. The gasification system comprising the gasifier, hot cyclone and gravity pipe is enclosed by a water jacket which is supplied with boiler feed water and operated at atmospheric pressure. The water jacket serves:

- To maintain the entire system at a uniform surface temperature level in order to avoid inadmissible stresses due to differing thermal expansion.
- To protect the pressure shells from superheating in the case of refractory fractures. Such an incident is, however, not very likely to occur.

The wood feedstock is fed into the fluidized-bed by means of a water-cooled screw conveyors. Gasification residues, termed bottom product, are withdrawn at the bottom of the gasifier. A cooling screw serves for discharging and cooling the bottom product. Due to the small amount obtained, the removal system only consists of one vessel and one cooling screw.

Crude gas cooling

The purpose of the crude gas cooling system is to cool down the crude gas produced in the gasifier from approx. 900 °C to approx. 350 °C. The heat recovered is used for the generation of high pressure (85 bar) saturated steam. The crude gas cooler is a convective boiler of the fire-tube type arranged horizontally.

The steam drum is located above the cooler to render possible natural circulation via downcomers and risers. The steam pressure is kept at 85 bar with a pressure control station. In order to keep the water in the steam drum at a normal level during varying gasifier conditions, the steam drum is provided with an automatic three-element feed water controller.

Particulate removal

The purpose of the particulate removal unit is to remove the dust and noxious components such as chlorides and ammonia by wet scrubbing and to saturate the crude gas as much as possible with water vapour which is necessary for the performance of the following CO-shift conversion. To achieve this, the crude gas is brought into contact with water which is mainly collected as condensate from the process.

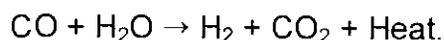
The crude gas leaving the crude gas cooler enters the quench cooler through a dip pipe. As a result of coming into contact with process water, the superheated crude gas is saturated with water and most of the dust is precipitated. The crude gas leaves the quench cooler at the top and flows to a venturi scrubber. By injecting process water most of the remaining dust is precipitated.

The scrubber (also having the function of a separator) is located downstream of the venturi. The crude gas separated from the water phase, then flows upward through a packing and is scrubbed in counter-current by process water. The gas which is now almost dust-free passes through a second packing where it is scrubbed again with a small quantity of boiler feed water (flowing in counter-current) in order to remove residual chlorides.

The dust-free gas leaves the scrubber at the top and passes through a final separator where entrained liquid is separated and recycled to the sump of the scrubber. After de-pressurising and cooling the dust-laden waste water (blackwater) is fed to the waste water pre-treatment section.

CO-shift conversion

The crude syngas contains too much CO and insufficient H₂ for the methanol synthesis process. In the catalytic CO-shift conversion process, the water gas shift reaction causes CO to be converted to H₂ according to the following formula:

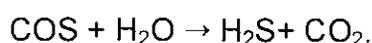


This is an equilibrium reaction and is influenced by the following parameters, amongst others:

- Variable stoichiometry of the dry gas due to the gasification of different wood types under different conditions.
- In particular, varying water vapour quantities in the crude syngas.
- Varying reactivity of the catalyst (between start of run and end of run).
- Varying residence times of the gas as a result of changing flow rates.

As the methanol synthesis gas must still contain a high CO portion, only partial conversion of the gas is required. For this reason and in order to maintain the CO / H₂ ratio constant at these changing parameters, it is necessary that the process gas quantities be controlled by means of a by-pass.

For this purpose, the crude syngas is split up, the major part fed over a catalyst and converted there until nearly the equilibrium is reached. In addition, most of the carbon-oxy-sulphide (COS) is hydrolysed to hydrogen sulphide (H₂S) in the presence of the catalyst according to the reaction formula:



Downstream of the reactor, the converted and non-converted gas streams are mixed to form a product gas at a CO : H₂ ratio that is required for the methanol synthesis process. The water-vapour-saturated crude syngas from the crude gas scrubber is split into a main and a by-pass stream. The main stream is heated up to the reaction temperature of approx. 320 °C in the gas/gas heat exchanger and passes through the CO-shift conversion reactor.

The hot converted gas leaving the CO-shift conversion reactor is utilised to superheat HP steam produced in the crude gas cooling section and to preheat the reactor feed in the said gas/gas heat exchanger. Subsequently, the converted gas is combined with the by-pass stream.

Gas cooling

The heat released during cooling down to 100 °C is used for boiler feed water preheating. The hot condensate is collected in a hot condensate drum and is pumped back directly to the scrubber in the particulate removal section. Then the gas is cooled down against cooling water. After separation of the condensate, the gas is further cooled with cold gas and refrigerant from the CO₂ / H₂S removal unit to the inlet conditions of the CO₂ / H₂S removal section.

Downstream the last chiller the gas is scrubbed with cooled make-up water to remove traces of ammonia. The cold process condensate and the ammonia wash water are collected in a cold condensate drum, from where they are fed to the waste water pre-treatment unit.

CO₂ / H₂S removal

The CO₂ / H₂S removal unit is based on GENOSORB 1753[®], solvent. This physical agent is produced by Hoechst AG, Gendorf works (now Clariant) and exhibits identical properties like the well-known SELEXOL[®], solvent by Union Carbide. The system is designed to remove CO₂, H₂S and benzene (C₆H₆) from the cooled syngas stream.

The inlet syngas is cooled down via gas/gas heat exchanger and syngas chiller to a temperature of 5 °C. The cold gas is washed with water to remove traces of ammonia, this scrubber being part of the gas cooling unit. In the absorption column the H₂S fraction is removed from the syngas down to a residual concentration of 1 ppm by volume at temperatures below 0 °C. Here also the benzene proportion of the gas is scrubbed out to a value below the specified lean gas concentration.

The H₂S enrichment in the CO₂ / H₂S laden solvent is achieved by recycling of a flashed gas stream via a compressor to the absorber feed stream. The CO₂ / H₂S rich solution is regenerated in the desorber. Heat required for the thermal stripping process is supplied by low pressure steam to the reboiler of the column. The overhead CO₂ / H₂S off-gas stream is fed to a H₂S conversion unit (for instance: incinerator, Linde Clinsulf or Topsøe Wet gas Sulphuric Acid process). The CO₂ fraction of the syngas is removed in the absorption column. The regeneration of the CO₂-rich solution is effected by flash and stripping regeneration.

The cold stripping process is based on nitrogen as a stripping agent, which is supplied by the air separation unit. The stripper off-gas contains traces of benzene, which can be removed in a separate activated carbon filter unit. The CO₂ / H₂S unit also includes a compressor-type refrigeration unit (refrigerant: ammonia) in order to cool the process gas to the low temperature level required.

Recycle gas compression

For the pressurising of the lock hoppers, for the basic fluidisation in the HTW-gasifiers and the like, CO₂ will be used at the following pressure levels:

- High Pressure (HP) CO₂ (60 bar a) for pressurising of the lock hopper systems.
- Medium Pressure (MP) CO₂ (32 bar a) for pressure maintaining in the charge bins and for the basic fluidisation in the HTW gasifiers.

Within the CO₂ / H₂S removal unit a part stream of the absorbed CO₂ is flashed and, after removal of benzene traces (in a activated carbon filter unit), fed to the recycle gas compression unit. The HP CO₂ is compressed in a two stage reciprocating compressor with inter-cooling stages. The MP CO₂ is produced by a two stage turbo compressor with inter-coolers.

The both compressors will be delivered as package units including related facilities. In general, recycle CO₂ discharged from the compressors is fed to the respective pipelines without final cooling. The CO₂ stream for the basic fluidisation in the HTW-gasifier is preheated up to 200 °C in a shell and tube heat exchanger with the aid of condensing MP steam.

Waste water pre-treatment

The blackwater which leaves the quench cooler is cooled down in two steps. In the first heat exchanger the water is cooled against the recycled greywater (filtered blackwater) and then with the aid of cooling water to the inlet temperature of the filter press. The solids are removed in a plate-and-frame type filter press. Two filter presses are installed in parallel: One in operation, while the second is being cleaned.

The filter cake is discharged via a belt conveyor into a container for transportation to the steam boiler for incineration. The filtrate is collected in a filtrate drum and recycled to the scrubber via the circulating water heat exchanger where it is heated up by the hot blackwater.

The surplus water is mixed with the cold process condensate from the gas cooling section, alkalisied with caustic solution (aqueous solution 20 % by wt.) and heated up in the stripper feed pre-heater to inlet conditions of the waste water stripper.

In the waste water stripper all the dissolved gases (mainly ammonia or NH₃, H₂S and CO₂) are removed from the water. The feed tray is located in the upper part of the column. The stripping steam is reboiled bottom product generated in the stripper reboiler which is heated by means of condensing LP steam.

The column overhead product is cooled down in a quench cooler to approximately 70 °C to 80 °C to avoid the formation of ammonia salts. The cooling is effected by injecting cold water (in counter-current to the column overhead product) into the quench cooler.

The cooled column overhead product is fed to a combustion facility (e.g. steam boiler). The quench water is a circulating water stream. This means, that the warm water from the sump of the quench cooler is re-cooled with the aid of cooling water in the circulating water cooler and then fed back to the top of the quench cooler.

The surplus water (condensed water vapour) is pumped to the top of the waste water stripper by means of a pump. To cool the stripper bottom product, it is used as a heating medium in the stripper feed pre-heater and is then cooled with cooling water in the bottom cooler product. Thereafter it is eventually admitted to a biological treatment unit (outside battery limits) for further treatment.

Air separation

To supply the required quantity of oxygen with a purity of 99.5 % by vol., a simple cryogenic air fractionation process will be provided. Ambient air is compressed to approx. 6 bar with the aid of a four-stage turbo compressor. The compressed process air is pre-cooled to approx. 8 °C by direct contact with cold water and then passed through molecular sieves, where it is freed from CO₂, hydrocarbons and the remaining water vapour.

The purified air now enters the separation unit (cold box) and passes through the main heat exchanger in counter-current to the cold products, oxygen and nitrogen. By this measure, the air is cooled to liquefaction temperature and is fed to the pressure column. In this column, the air is separated by rectification into a pure nitrogen and a crude oxygen fraction.

An air part stream is withdrawn from the central part of the main heat exchanger at a temperature of approx. -90 °C and expanded in an expansion turbine in order to cover the cold requirements of the process. The cold air, which is at almost atmospheric pressure, is fed to the low-pressure column. In the low-pressure column, the crude oxygen from the pressure column is subjected to final separation into the pure components, nitrogen and oxygen.

The purity of the oxygen is determined by the number of trays used. Both products are withdrawn in the gaseous state and heated by transferring their cold to the process air in the aforementioned main heat exchanger. A small nitrogen part stream can be withdrawn in the liquid state in order to charge the liquid nitrogen storage tank.

Oxygen supply and nitrogen system

Oxygen from the air separation unit is compressed to the required pressure (i.e., 30 bar) with the aid of a six-stage turbo compressor. The major portion of the compressed oxygen is used as an oxidising agent in the HTW gasification unit, whereas a minor part stream is fed to the ATR (Auto-Thermal Reformer) for the treatment of the purge gas of the methanol synthesis loop.

Nitrogen produced in the air separation unit with a high purity is used as an inert gas for the following purposes within the plant:

- Cleaning the gas filter systems in the wood drying unit and dried wood storage sections.
- Pneumatic conveyance of gasifier bottom product (char) in the HTW-gasification unit.
- Stripping of solvent in the regeneration section of the CO₂ / H₂S removal unit.
- Purging and blanketing, in particular during abnormal cases of operation.

For the permanent consumers (filter cleaning, pneumatic conveyance, stripping) a pressure of approx. 6 bar is required. Nitrogen from air separation unit is compressed to this pressure level in a turbo compressor and then distributed to the places of consumption.

A small portion of nitrogen can be obtained from the air separation unit in liquid state in order to fill up a liquid nitrogen storage which ensures that the nitrogen supply to the plant is maintained during abnormal cases of operation.

Methanol synthesis and distillation

The methanol synthesis plant comprise the following units:

- syngas compression,
- methanol synthesis,
- purge gas treatment (ATR),
- methanol distillation,
- methanol storage.

The produced syngas of the synthesis gas plant is further compressed in a syngas compressor to meet the pressure requirements of the methanol synthesis. The methanol synthesis takes place in a Boiling Water Reactor (BWR) loop; in order to prevent a high level of methane from building up in the loop, the purge gas is recycled through an Auto-Thermal Reformer (ATR).

This is done to maximise the total methanol yield by reforming the purge gas to syngas quality. Consequently, this reduces the feedstock requirement substantially. The synthesised methanol is further distilled and thus upgraded to fuel-grade quality in a two column distillation. The produced methanol has a purity of 99.85 % and is taken to storage tanks. Only the Haldor Topsøe ATR unit is here further elaborated as the rest of the methanol unit is of conventional type.

Auto-Thermal Reforming

The Auto-Thermal Reforming (ATR) unit is a stand-alone process, in which the entire hydrocarbon conversion is completed in a single reactor based on a combination of partial oxidation and adiabatic steam reforming. The reactor features a refractory-lined pressure vessel with a burner, a conical combustion zone, and a catalyst bed (Christensen, Primdahl, 1994), (Holm-Larsen *et al*, 1996).

Unlike the tubular steam reformer, no pressure-bearing or high-alloy internals are required, and catalyst loading and unloading is straight forward and fast.

The burner is the key element in the ATR. It mixes the hydrocarbon and oxidant feedstocks, and by internal combustion in the flame, heat is provided for conversion of the hydrocarbons. Careful design of the burner nozzle is required to ensure a flow pattern which both mixes efficiently and protects the refractory and burner itself from the hot flame core.

The chemical reactions involved are combinations of combustion and steam reforming and the reactor space is divided into three reaction zones:

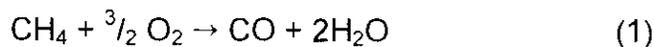
1. The combustion zone,
2. The thermal zone,
3. The catalytic zone.

The combustion zone is the turbulent diffusion flame where hydrocarbon and oxygen are gradually mixed and combusted. Normally, the principle of "mixed-is-burnt" is assumed, because the exothermic combustion reactions are very fast. Combustion in the ATR is a sub-stoichiometric process with an overall oxygen to hydrocarbon ratio of 0.55 to 0.60.

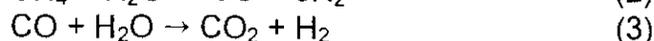
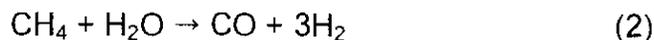
Methane combustion takes place through many radical reactions, but in a simplified model it is considered as one molecular reaction, i.e., the highly exothermic combustion of methane to carbon monoxide and water with an oxygen to methane ratio of 1.5 (1).

This reaction (1) was selected because carbon monoxide is the primary combustion product, which is converted to carbon dioxide by a slow secondary reaction. Excess methane will be present at the combustion zone exit after all the oxygen has been converted. Combustion of higher hydrocarbons can be simplified by similar reactions.

Combustion zone:



Thermal and catalytic zones:



The thermal zone is part of the combustion chamber where further conversion occurs by homogenous gas-phase reactions. The main reactions a) Thermal-methane reforming (2) and b) Water-gas shift reaction (3), but various pyrolysis reactions involving higher hydrocarbons also occur.

A fixed-bed catalyst follows the combustion chamber. In the catalytic zone, final hydrocarbon conversion takes place through heterogeneous catalytic reactions. At the catalytic-zone exit, the syngas will be in equilibrium with the methane reforming (2) and shift reactions (3). Furthermore, the catalyst will destroy any soot precursors formed in the combustion and thermal zones.

The catalyst bed is exposed to high temperatures (1,100 °C to 1,400 °C) as the gas leaves the combustion chamber. A high-thermal-stability catalyst and carrier system are thus required.

Medium pressure steam is generated when cooling the ATR effluent. The ATR is capable of operating without carbon formation at a much lower steam to carbon ratio than tubular steam reforming, hence, the MP steam generated by the cooling of the ATR effluent is sufficient both for process steam for medium pressure steam consuming units and for power generation in a steam extraction high pressure steam turbine.

Steam and electric power

The synthesis gas plant and the methanol synthesis plant forms part of an overall complex for the production of methanol. It is intended to make the overall plant complex self-sufficient for energy. This is accomplished by the incorporation of a wood-based steam boiler unit in conjunction with steam turbines generating electric power.

Gasification residue as well as waste gas streams from the synthesis gas plant is incinerated in said steam boiler as additional fuels. This consumption of process steam and power are shown in Appendix 1. As can be seen in Appendix 2, the overall plant is self-sufficient in terms of steam at all pressure levels and in addition produces a surplus of available heat as hot water for distant heating.

STEAM AND POWER BALANCE

Appendix 1

Consumption Data

Gasification syngas plant

Gasification syngas plant

Unit	Designation	Pressure [bar]	Temp [°C]	Flow [kg/h]	Power [kW]	BFW [kg/h]
2	Wood drying	6	sat	5000	9150	0
3	Dried wood storage	6	sat	3000	245	0
4	Wood feeding	6	sat	6000	270	30000
5	HTW gasification	35	300	4125	36	0
6	Crude gas cooling	N/A	N/A	N/A	0	65560
7	Particulate removal	N/A	N/A	N/A	75	0
8	CO Shift	35	sat	27825	0	0
9	Gas cooling	N/A	N/A	N/A	16	N/A
10	CO ₂ / H ₂ S removal	6	sat	11400	8560	0
11	Recycle gas compression	35	sat	640	1380	0
12	Waste water pretreatment	6	sat	5900	115	0
14	Air separation	35	sat	550	8200	N/A
15	O ₂ Supply system	N/A	N/A	N/A	3700	N/A
	Total	N/A	N/A	64440	31747	95560

HTW gasification: Steam consumed in reaction
 CO Shift: Steam consumed in reaction

Methanol synthesis plant

Designation	Pressure [bar]	Temp [°C]	Flow [kg/h]	Power [kW]	BFW [kg/h]
Distillation	6	sat	62500	25	0
BWR and ATR WHS	25	250	9300	9875	86000
Gas cooling after ATR	N/A	N/A	0	0	N/A
Effluent cooling after ATR	N/A	N/A	0	0	N/A
Total	N/A	N/A	71800	9900	86000

BWR and ATR WHS: 51 bar, 230 °C

[N/A = Not Applicable]

Production Data

Gasification syngas plant

Unit	Designation	Pressure [bar]	Temp [°C]	Flow [kg/h]	Condens. [kg/h]	BFW [kg/h]	Heat [kW]
2	Wood drying	N/A	N/A	N/A	5000	62000	0
3	Dried wood storage	N/A	N/A	N/A	3000	0	0
4	Wood feeding	N/A	N/A	N/A	6000	30000	0
5	HTW gasification	N/A	N/A	N/A	0	0	0
6	Crude gas cooling	N/A	N/A	N/A	0	0	0
7	Particulate removal	N/A	N/A	N/A	0	0	0
8	CO Shift	N/A	400	64911	0	0	0
9	Gas cooling	N/A	N/A	N/A	N/A	N/A	39100
10	CO ₂ / H ₂ S removal	N/A	N/A	N/A	11400	0	0
11	Recycle gas compression	N/A	N/A	N/A	640	0	0
12	Waste water pretreatment	N/A	N/A	N/A	5900	0	0
14	Air separation	N/A	N/A	N/A	550	N/A	0
15	O ₂ Supply system	N/A	N/A	N/A	N/A	N/A	0
	Total	N/A	N/A	64911	32490	92000	39100

Gas cooling: 195 °C to 40 °C

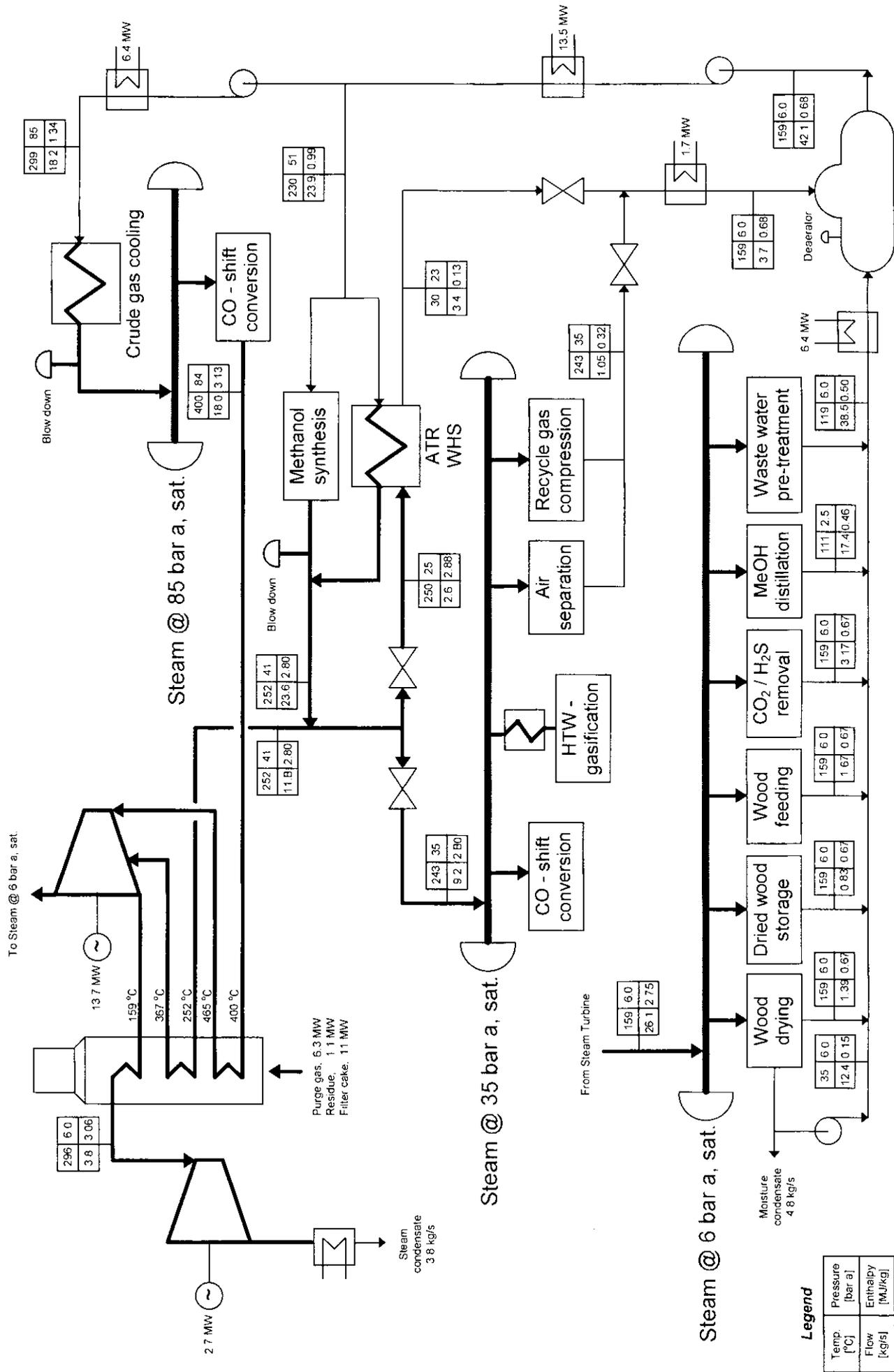
Methanol synthesis plant

Designation	Pressure [bar]	Temp [°C]	Flow [kg/h]	Condens. [kg/h]	BFW [kg/h]	Heat [kW]
Distillation	N/A	N/A	N/A	62500	0	0
BWR and ATR WHS	41	sat	85000	12100	0	0
Gas cooling after ATR	N/A	N/A	N/A	N/A	0	47500
Effluent cooling after ATR	N/A	N/A	N/A	N/A	N/A	8141
Total	N/A	N/A	85000	74600	0	55641

Distillation: 2.5 bar, 111 °C
 BWR and ATR WHS: 23 bar, 30 °C + excess water 1500 kg/h, 31 bar, 30 °C
 Effluent cooling after ATR: 260 °C to 125 °C

STEAM DIAGRAM

Appendix 2



Note: The power/heat boiler necessary to make the plant self-sufficient in power is not part of this diagram.

State of reference H₂O (l), 1 bar a, 20 °C

1.2.4 Plant performance

Basis for design

This methanol production plant will consume 177 - 187 t/h or about 1.3 million t/year of unprocessed wood biomass feedstock (50 % moisture) with an estimated availability of 85 %. Thus, the plant will produce 1,000 t/d or 390 million litre/year of fuel-grade methanol.

The plant site is yet to be defined, however, is to be located in Sweden at a site where substantial volumes of forestry material is at hand - i.e., in conjunction with pulp mills, preferably with integrated sawmills. Man power needed for operation of the plant is estimated as 60 - 70 people.

The High Temperature Winkler (HTW) fluidized-bed gasification technology forms the basis for the synthesis gas production plant. This process, supplied by Krupp Uhde, Germany has attained industrial maturity for the gasification of reactive feedstocks such as brown coal and peat under pressure.

Two commercial-scale gasification plants (gasification pressure 10 - 13.5 bar) and advanced pilot plants (gasification pressure up to 25 bar) have been in successful operation for many years.

The methanol synthesis plant comprise a methanol synthesis with a purge gas stream led to an Auto-Thermal Reformer (ATR), thus making a loop. This process is supplied by Haldor Topsøe, Denmark and is commercially available. Consequently, the methanol production plant will ensure high reliability and availability.

Mass and energy balances

The total plant mass and energy balances have been calculated. This has been achieved by combining and integrating performance data for the various plant units as provided by the respective technology suppliers. Hence, given mass and energy data are listed in a steam and power balance in Appendix 1.

Furthermore, a steam diagram has been configured from the steam and power balance data and with steam data taken from Wester, 1991. This steam diagram is illustrated in Appendix 2. Using the steam diagram the respective heat and power producers and consumers have been identified. The mass and energy balance has then been calculated. This calculation is in a simplified form given below.

The total heat consumed in the heating of the make-up water and the total heat produced by various units (e.g. gas cooling, utility boiler etc.) has been calculated. Moreover, a rough pinch-analysis has been made in order to make sure that no cooling and heating curves are crossed and that an appropriate temperature difference is present to overcome the overall heat transfer resistance. Thus, the calculated heat duty available has been calculated as,

- Consumed heat: $6.4 + 1.7 + 13.5 + 6.4 = 28.0$ MW,
- Produced heat: $8.1 + 37.7 + 19.5 + 12.6 + 6.5 = 84.4$ MW,
- Available heat: $84.4 - 28.0 = 56.4$ MW.

The available heat is taken as hot water at a temperature >100 °C. As is shown later, additional heat can be produced in a combined heat and power plant unit, when operated in back-pressure mode. The additional heat totals 49 MW.

- Range of total heat available $= 56 - 105$ MW.

The total power produced and consumed has been calculated as,

- Produced power: $2.7 + 13.7 = 16.4$ MW,
- Consumed power: $31.7 + 9.9 = 41.6$ MW,
- Power shortage: $41.6 - 16.4 = 25.1$ MW.

This power shortage can be compensated for in several ways. For example, as in the case of Sweden, low electricity prices give an incentive of purchasing power at the grid. However, in order for a fair comparison between alternative technologies in producing biomass-derived alcohol fuels this methanol production plant is designed to be energy self-sufficient.

Consequently, an additional combined heat and power plant unit is needed and to be located within the plant. Therefore, it has been chosen that additional biomass is to be burned in a power boiler and a (condensing or not condensing) steam turbine will generate electricity.

Thus, with an assumed electric efficiency of 40 % of an advanced combined heat and power plant in condensing mode or with 28 % in back-pressure mode the net additional biomass power needed can be calculated.

- Condensing mode:

$$\text{Needed additional biomass: } 25.1 \cdot 0.40^{-1} = 63 \text{ MW.}$$

In this mode there is no production of available heat. This biomass power corresponds to an added biomass feedstock input of

$$\begin{aligned} 63 \cdot 3600 \cdot 0.50^{-1} \cdot 19.2^{-1} &= 23,625 \text{ kg/h.} \\ [\text{MJ/s}] \cdot [\text{s/h}] \cdot [\text{kg total/kg dry}] \cdot [\text{kg/MJ}] &= [\text{kg/h}] \end{aligned}$$

- Back-pressure mode, special case - integration with co-generation units:

$$\text{Needed additional biomass: } 25.1 \cdot 0.28^{-1} = 90 \text{ MW.}$$

In this mode an additional 49 MW of available heat is produced. This corresponds to an additional biomass feedstock input of

$$\begin{aligned} 90 \cdot 3600 \cdot 0.50^{-1} \cdot 19.2^{-1} &= 33,750 \text{ kg/h.} \\ [\text{MJ/s}] \cdot [\text{s/h}] \cdot [\text{kg total/kg dry}] \cdot [\text{kg/MJ}] &= [\text{kg/h}] \end{aligned}$$

The total input of biomass feedstock as received is thus 153,000 kg/h + an additional 23,625 kg/h or 33,750 kg/h, depending on mode of the combined heat and power unit. Thus, the total input equals 177 - 187 t/h. The results of the calculated total energy balance and methanol yield is shown in Table 1.1.

Wood consumption, 50 % moisture [t/h]	177 - 187
<u>Input</u>	[MW]
Biomass feedstock (LHV)	408
Additional biomass (LHV)	63 - 90
Total	471 - 498
<u>Output</u>	[MW]
Methanol	231
Heat	56 - 105
Losses	162 - 184
Total	471 - 498
Net Methanol Output	231 MW
Gross Methanol Yield	57 %
Net Methanol Yield	46 - 49 %
Total Energy Yield	61 - 67 %

Table 1.1. Total mass and energy balance summary and methanol yield.

The total plant net methanol efficiency has been calculated from the steam and power balance and by the steam diagram. This efficiency can be shown in a Sankey diagram, see Figure 1.3.

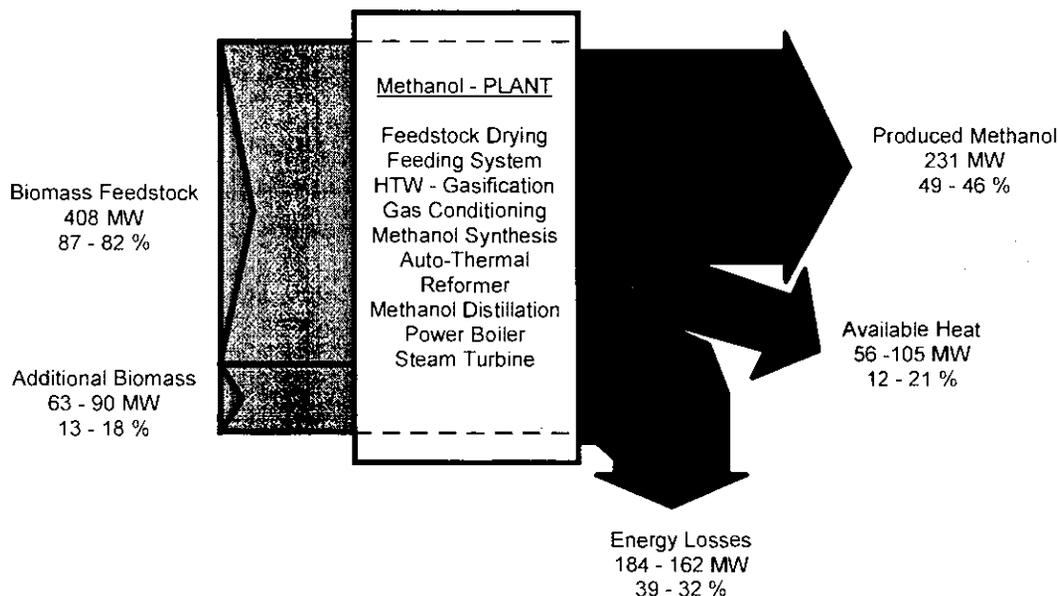


Figure 1.3. Sankey diagram, methanol production from biomass.

1.2.5 Environmental performance

The crude gas from the HTW wood gasification unit contains trace components in the form of sulphur and nitrogen compounds and higher hydrocarbons. Their concentration depends on the feedstock composition and on the gasification conditions. The total sulphur content of the wood feedstock is approx. 30 mg S per MJ net calorific value (as received conditions).

The sulphur will react in the gasifier with hydrogen, carbon and carbon oxide mainly to gaseous H_2S and COS . A certain portion of the sulphur in the wood feedstock is combined in the ash during gasification and found again in the gasifiers' bottom product and the entrained dust. The sulphur compounds are completely removed from the process gas in the CO_2 / H_2S removal unit.

A certain portion of the nitrogen contained in the wood will be converted to NH_3 , whereas a minor portion will be obtained as hydrogen cyanide (HCN). Ammonia is dissolved in the quench water of the wet scrubbing system and in the process condensate obtained during process gas cooling.

In the waste water pre-treatment section, the ammonia is separated from water by stripping, HCN is largely decomposed by catalytic effects in the CO-shift conversion unit. Different volatile organic compounds from benzene to heavy molecular tars may form in the fluidized-bed wood gasification system. The definition of tar is the sum of organic components heavier than benzene.

Summarising the experience gathered in biomass gasification, it can be stated that most of the tar components, such as phenols, toluene's, pyridine and cresols, are decomposed by using high gasification temperature (≥ 900 °C).

The solid residues of gasification will be incinerated as additional fuels in a steam boiler in order to utilise the energy contained. During combustion, a minor portion of the combined sulphur will be released and exhausted to the atmosphere in the form of SO₂. In conclusion, the plant will exhibit good environmental performance with very little emissions.

1.2.6 Investment estimate

The investment cost estimate as well as the calculated methanol production cost is based on 1st quarter of 1997 and was based on cost information given by respective supplier or vendor. Some costs were assessed using in-house information and applying accepted factoring methods as degression exponents, scale-up factors etc.

Moreover, the capital investment cost was calculated with an interest of 8 % and a project lifetime of 15 years, hence, an annuity of some 12 %. Thus, the estimated investment cost for the biomass to methanol production plant is 417 MMUSD or USD1,800/kW. An investment cost break-down is illustrated below in Figure 1.4.

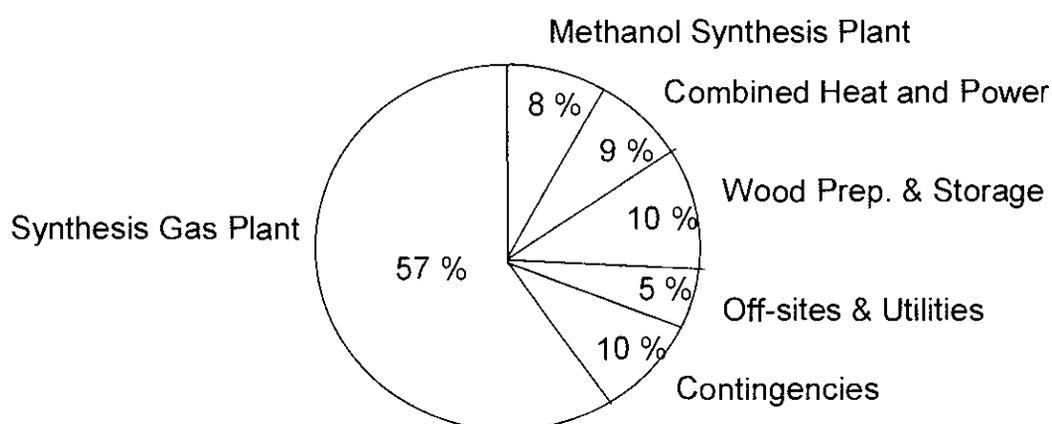


Figure 1.4. Investment cost breakdown.

Methanol production costs

The calculation of the methanol production cost is shown below. In the calculated cost there has been no credit taken for the available production of heat for distant heating. Thus, for the production of 1,000 t/d of fuel-grade methanol the calculated production cost per year is USD0.28/litre or ECU 0.25/litre.

General

Feedstock Input (a)	177 t/h
Equivalent Full Load Hours (b)	7,500 h/year

Investment Cost

Plant Capital	
Gasification Synthesis Plant (c)	240 MMUSD
Methanol Synthesis Plant (d)	35 MMUSD
Combined Heat and Power Unit (e)	38 MMUSD
Wood Preparation & Storage (f)	42 MMUSD
Off-sites & Utilities (g)	21 MMUSD
Contingencies (h)	42 MMUSD
Total	417 MMUSD

Operating Cost

Operation (i)	9 MMUSD/year
Maintenance (j)	13 MMUSD/year
Feedstock Cost (k)	38 MMUSD/year
Total	60 MMUSD/year

Yearly Cost

Capital Cost (l)	49 MMUSD/year
Operating Cost	60 MMUSD/year
Total	108 MMUSD/year

Estimated Investment & Production Costs in US and EU currencies

Daily Production	1,000 t/day
Yearly Production (m)	390 million litres/year

Plant Investment Cost	1,807 USD/kW
Plant Investment Cost	1.07 USD/litre
Methanol Production Cost	0.28 USD/litre

Plant Investment Cost	1,595 ECU/kW
Plant Investment Cost	0.94 ECU/litre
Methanol Production Cost	0.25 ECU/litre

Exchange rates (Postgirot, 26 September 1997):

1 USD = 7.50 SEK
1 DEM = 4.29 SEK
1 DEM = 0.57 USD
1 ECU = 8.50 SEK (set rate by Authors).

- a) Total wood input, taken as received @ 50 % moisture, and when the combined heat and power unit operates in condensing mode.
- b) Equivalent full load hours is based on an availability of 85 % per year.
- c) The gasification synthesis plant cost is estimated by Krupp Uhde as 421 million DEM and comprise specified equipment, bulk material, engineering and licence fees, civil works and erection, and package units.
- d) The methanol synthesis plant cost is estimated by Haldor Topsøe as 60 million DEM and comprise synthesis loop, the distillation section and the Auto-Thermal Reforming unit.
- e) The installed cost for a condensing-extraction steam turbine co-generation system including a waste heat boiler is assumed as USD2,500/kW_e.
- f) The cost for a wood preparation & storage including machines is assumed as 10 % of the total investment cost.
- g) Off-sites and utilities include buildings, emergency power system, flare system, waste water treatment facilities, sewer systems, fire water system, chemicals dosing system, utility boiler and extraction steam turbine. The cost is taken as 5 % of the total investment cost.
- h) Contingencies is set to 10 % of the total investment cost.
- i) Operation cost comprise staffing, chemicals, catalysts, overhead, insurance, land lease, and real estate tax. The cost is calculated as a staff cost of 63 men, with a salary of USD46,666/year and 40 % overhead; catalyst replacement cost of 2 million DEM; and 1 % of the total investment cost.
- j) Maintenance cost is assumed as 3 % of the total investment cost.
- k) Fuel cost is the cost for wood as received including delivery is assumed as USD10.7/MWh. The wood has a LHV of 9.6 MJ/kg @ 50 % moisture.
- l) Capital cost is calculated with an 8 % interest and a 15 years project lifetime, thus an annuity of some 12 %.
- m) The total production of methanol per year in litre is calculated with a production of 1,000 t/day and yearly availability of 85 % and a methanol density of 795 kg/m³ @ 15 °C, 1 bar (Perry *et al*, 1984).

1.2.7 Plant reliability and availability

This process plant configuration is mainly based on proven and commercially available technologies and equipment with guaranteed performance from competing vendors. The simple and straight-forward process configuration with a minimum of integration between plant units and heat exchange between process streams contributes to an easy and reliable operation of the facility.

Thus, this methanol producing plant will exhibit good reliability and availability. Adequate maintenance record for the gas generation section, air separation, gasification and gas treatment, are well documented from commercial operation. In addition, gas CO-shift conversion are widely used within the chemical and petroleum industries.

1.3 Conclusions and possible improvements

Advanced biomass gasification holds the promise to change the way that biomass residues are used in the generation of steam and power and in the production of chemicals.

The commercialisation of such technologies involved will provide users with the opportunity to replace fossil fuels combustion processes with a clean, efficient, and totally renewable alternative. With current environmental pressures and the ageing of existing systems, the need for economically viable alternatives is crucial.

1.3.1 Conclusions

The goal with this study has been to create a straight forward system - biofuel to methanol. This approach effects certain parameters in the concept and introduces more costly solutions (i.e., more energy demanding) than necessary if the concept was to be more open. A more open system on the other hand would require integration with other plants such as a combined cycle plant (gas turbine/steam turbine cycle).

With the proposed process scheme the methanol can be produced at an energy efficiency of 57 %. When the plant is made to produce its own electric power needs the biomass to methanol efficiency decreases to 49 % at the same time as the plant will produce heat suitable for district heat system corresponding to 12 - 21 % of the fuel input.

Total thermal efficiency is therefore 61 - 67 %. The required extra power is in this presented case produced in a combined heat and power plant. The overall conclusion is that methanol can be produced from biomass in a free-standing, self-sufficient plant at a net biomass to methanol efficiency of 49 %.

Certain improvements in fuel drying and gas purification could presumably bring this to around 55 %. This is achieved with today's technology, however, with the knowledge that the gasifier so far has not been tested in full scale/full pressure with biomass as fuel. Other process steps in the sequence are all tested and operate commercially in other part of the process industry.

1.3.2 Possible improvements

When biomass gasification matures improvement in gasifier efficiency will occur. This is also true for the methanol synthesis where small improvements can be expected. Overall, however, no major gains can be expected as nature puts up limits for what actually can be done. There is, however, a greater potential for improvements in the management of the heat released in the exothermal processes and in the way electricity is used in the process.

Electricity savings can be expected in:

- The process of fuel drying,
- The gas clean-up section,
- Fine tuning of the gasifier vs. methanol synthesis pressure,
- The methanol synthesis loop.

Steam consumption savings can be expected in:

- The gas clean-up section,
- The shift section, as certain catalysts can be operated with a lower steam to dry gas ratio,
- The methanol purge gas handling, if the purge is handled differently,
- The methanol distillation section.

If the straight biomass to methanol concept was extended to become a combined methanol/power/heat plant a number of advantages would be gained. The principle configuration of such a concept is compared with the process concept of this report in Figures 1.5 and 1.6 (see below).

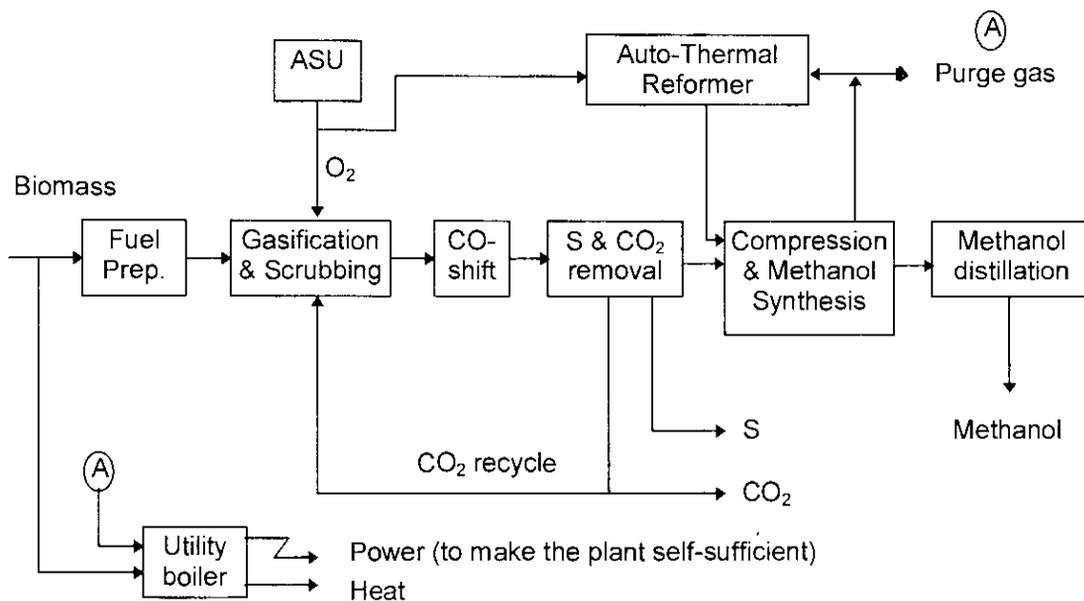


Figure 1.5. Process concept of the report.

In the present concept the methane content of the gas must be converted to syngas in order to gain maximum methanol yield of the feedstock. This is most efficiently done via the introduction of an Auto-Thermal Reformer (ATR) on the purge stream from the methanol synthesis (a purge, although smaller is still needed to purge inerts and this is shown as feed to the utility boiler).

The ATR produces H_2 and CO out of the CH_4 in the gas but also some CO_2 . The methanol catalyst operates at its optimum at a certain ratio between H_2 , CO and CO_2 . Moreover, CO_2 should not be higher than about 3 mol %. As CO_2 is formed in the ATR the CO_2 in the syngas from the H_2S / CO_2 removal unit must be kept very low, around 0.2 mol %.

This puts extra pressure on the gas clean-up unit which increases both power and steam demands. The deletion of the ATR will therefore decrease utility consumption in the gas clean-up unit by a fair amount.

The alternative, integrated solution, see Figure 1.6, then calls for optimum usage of the purge gas. This is best done by a gas turbine combined with a steam cycle. The heat recovery section of the gas turbine will also be an important high temperature zone where saturated steam flows can be superheated to high temperatures.

The purge stream may need to be supplemented by an extra amount of fuel gas (see dotted line of Figure 1.6) to reach a convenient flow which suits an existing gas turbine.

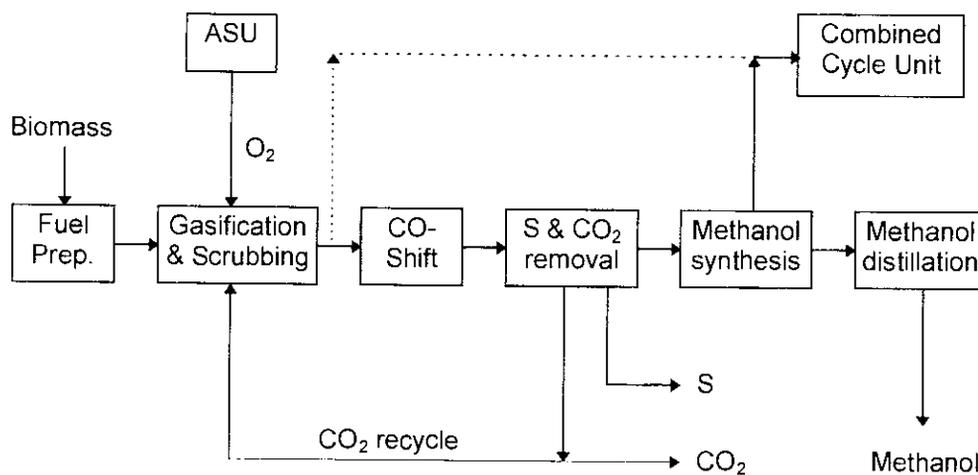


Figure 1.6. Alternative concept.

The amount of recycled gas in the methanol synthesis loop also needs to be optimised to find a balance between methanol and power production. Minimum recycle is no recycle which leads to the so called Once Through Methanol concept (OTM). In such a case the upstream shift unit can be found to be unnecessary.

As the methanol syngas is only brought once through the synthesis reactor a highly active catalyst is desirable in such a case and at least one methanol synthesis supplier has developed such a methanol catalyst (in a liquid slurry) for this purpose. The non-reacted gas is brought to a gas turbine/steam turbine cycle as previously described.

The clearly positive effect of a combined system as described above is that the sensible heat of various flows can be more powerfully utilised. This will lead to a higher overall thermal efficiency of the complex and a higher yield of valuable products in the form of methanol and power.

1.4 References

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2. CELLULOSE HYDROLYSIS - ETHANOL PRODUCTION

For accurate and correct assessment of processes for ethanol production from wood, detailed, although preliminary, engineering studies are required. Unfortunately, few such studies have been published in sufficient detail on the latest developments of the enzymatic route in the U.S. This process developed by NREL and Amoco was offered by a group, Swan Biomass (earlier including the engineering company Stone & Webster), for agricultural and hardwood feed-stocks. Negotiations for commercial application have been conducted by SSEU (Swedish Ethanol Foundation) and a very preliminary process proposal is available (SWAN 1996).

Previous studies (DOE 1993, IEA 1993, Lund 1993) are reviewed herein. The most detailed engineering study for the CASH-process (SSEU 1991) with all acid hydrolysis is summarised for comparison, as this is the most comprehensive study available.

Each process comprise:

- pre-treatment (acid hemicellulose hydrolysis),
- enzyme production,
- cellulose hydrolysis,
- fermentation of simple sugars,
- distillation,
- solids upgrading.

The principles of these are presumed to be well-known. To some extent, details can be found in the referenced source literature. This report focuses on the energy picture of the processes in stand-alone, energy self-sufficient plants, as true comparisons can only be made on such basis in judging the long term prospects on the very big motor fuel market. In reality, combinations by integration with other kinds of production (electricity, district heating, industrial process heat) will initially prove beneficial and be attempted and used to lower costs. However, they will probably be limited in scale and are considered special cases. There are several examples of such studies (Nutek 1995).

Development work on concentrated acid hydrolysis goes on in Australia (Apace Research, Ltd) in co-operation with TVA in USA, but no published results are available. Other developments such as hydrolysis with supercritical water (SCW) are studied at laboratory level and should be watched. Improved heat management by using supercritical extraction techniques to recover ethanol from dilute solutions (instead of distillation) is another interesting development.

2.1 The CASH study

The CASH-process is based on development work carried out in Canada (SLR, St. Lawrence Reactor, and Bio-Hol), USA (TVA, Tennessee Valley Authority)

and in Sweden (SSEU), and includes a two-stage dilute acid hydrolysis. The engineering study (principal process steps *Figure 2.1*) is based on softwood feedstock (pine sawdust) and ferments only hexoses. The wood-DS (dry substance) contains 44 % cellulose and 22 % hemicellulose and is built by sugars, of which nearly 90 % are hexoses. Such a feedstock may theoretically yield 37.5 wt-% of ethanol (i. e., 476 litres per ton DS) or 53 % in energy terms based on the lower heating value (LHV).

Conversion efficiency is assumed to be 95 % in the first hemicellulose hydrolysis stage (and 92 % sugar recovery in the washing) and 65 % in the second cellulose hydrolysis stage with 50 % yield of hexoses (and 90 - 92 % sugar recovery in the washing). 59 % the theoretical content of hexoses in the feedstock is converted and recovered for fermentation and 55 % of the potential pentoses.

Ethanol yield in the fermentation from hexoses is 90 % of the theoretical and loss at the distillation are less than 1 %. No pentoses are converted to ethanol. On overall, an ethanol yield of 17.7 wt-% is obtained or 47 % the theoretical.

The "stand-alone", energy self-sustained plant converts in 330 days 100,000 t/yr DS to 22,395 m³ of 100 % ethanol (in product with 5 wt-% water) and yields a surplus of nearly 28,000 t/yr of lignin-DS as solid fuel with 15 % moisture for sale. Dissolved solids are anaerobically fermented to methane, which is used as internal plant fuel together with part of the lignin to produce steam for heat and power needs.

The following mass and energy data were extracted from the study, *Table 2.1* (purchased electricity has been transformed to self-generation with back-pressure steam turbine with methane and lignin as fuels).

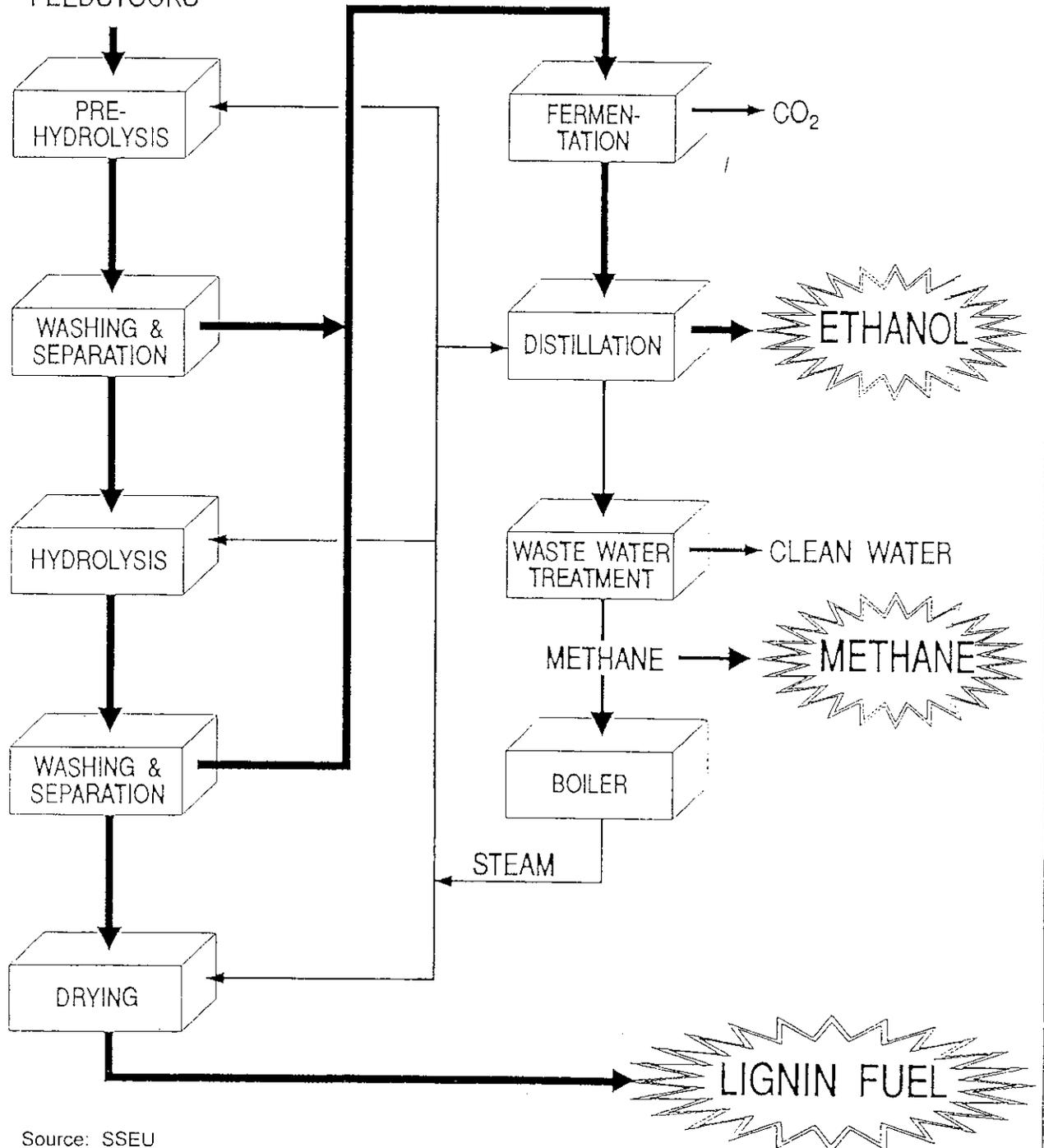
The yield of ethanol, nearly 25 % on energy basis, can be expected to increase to approximately 32 to 33 % by development work based on increasing cellulose conversion and by finding micro-organisms capable of fermenting pentoses as well. The total yield of energy products, somewhat above 61 % in the table, will decrease as less unconverted cellulose is available in the solid fuel product and as process energy needs increase. The total efficiency could be improved if the added power needed (above what is produced with a back-pressure turbine) was purchased from more energy efficient power plants.

Obviously, co-location with a co-generation power plant or other big power and heat consuming plants, such as pulp and paper mills, could be advantageous for an ethanol plant due to integration effects. The surplus lignin fuel might be a potential powder fuel for gas turbines, as it has been found to have low content of alkali salts compared to the original wood feedstock.

Figure 2.1

THE CASH PROCESS

LIGNOCELLULOSIC
FEEDSTOCKS



Source: SSEU

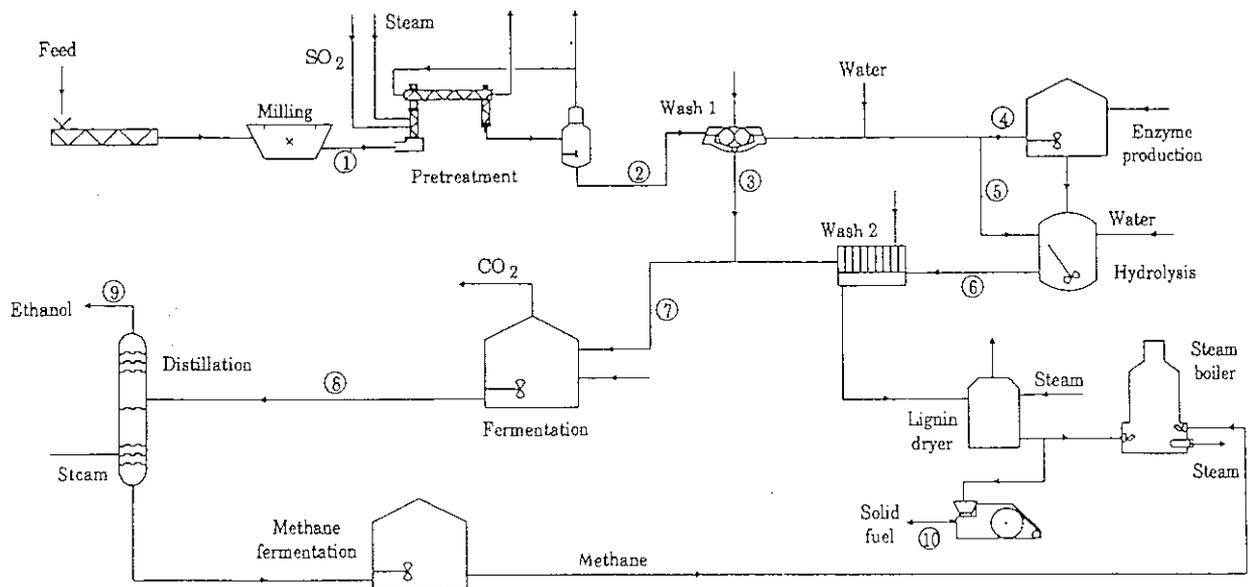
Table 2.1

	Mass or volume t/yr or m ³ /yr	Energy TJ/yr	Energy yield, % (LHV-basis)
<u>IN</u>			
Pine wood, DS, mass	100,000	1,900	
(Steam internally produced		641.5)	
(Electricity, net internally produced		77.5)	
Chemicals, mass	3,255		
<u>OUT</u>			
Ethanol (100 %), vol. (produced as azeotrope)	22,395	473	24,9
Lignin-DS, mass (for sale)	27,920	670	36,3
(Lignin-DS, internal use	6,290	151)	
(Methane, internal use, energy		469)	
Hot water >100 °C	none		0

2.2 The Lund university studies

At the Lund Institute of Technology (University of Lund) R&D work on production of ethanol from wood has been accomplished over a period of 15 years, focusing on pretreatment/hydrolysis with enzymes (Department of Chemical Engineering) and development of micro-organisms (genetically modified yeasts) capable of fermenting both pentoses and hexoses (Department of Applied Microbiology). On basis of this work preliminary engineering studies have been carried out for technical and economic evaluation, including comparison of the enzymatic route with the CASH dilute acid process (principal process steps, *Figure 2.2 and 2.3*).

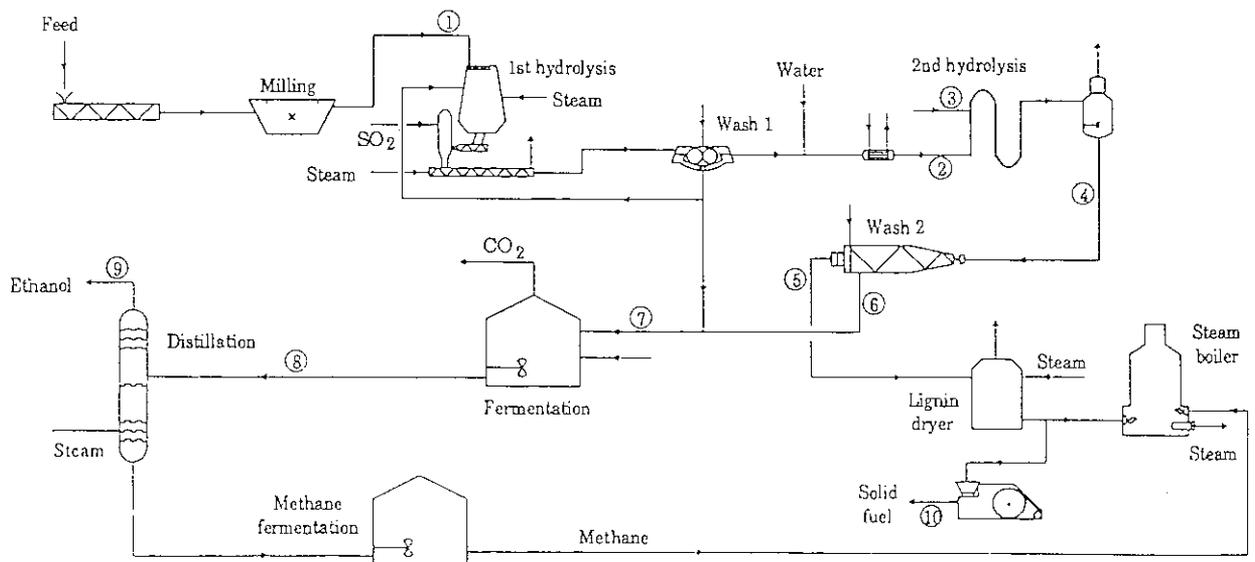
The mass and energy data in *Table 2.2* has been extracted from the studies for a stand-alone plant processing in 300 days 100,000 t/yr of wood-DS (presumed to be pine sawdust containing 38.7 % cellulose and 22.9 % hemicellulose, and consisting of sugars, of which 87 % are hexoses) . With this raw material the theoretical maximum yield of ethanol is 35 wt-% (i. e. 445 litres per ton DS) or 49 % of the energy content (LHV-basis) in the feedstock. This assumes 100 % conversions and recoveries in all process steps, including pentose fermentation



Stream	1	2	3	4	5	6	8	9	10
Fibrous	13.89	7.61	0.16	0.44	7.01	3.75	0.53	0.02	2.72
Hexoses		3.12	2.77	0.02	0.32	4.21	6.77	0.06	
Pentoses		0.84	0.74		0.09	0.09	0.83	0.80	
Ethanol							3.26	3.23	
Water	9.26	12.10	24.39	2.47	39.37	63.66	104.38	102.43	0.48
Other		1.92	1.71	0.02	0.26	0.61	2.28	2.72	

Figure 2.2 Schematic flowsheet for the enzymatic process. Flow rates in ton/h.

Source: Lund



Stream	1	2	3	4	5	6	7	8	9	10
Fibrous	13.89	9.69		6.23	6.16	0.06	0.13	0.15		3.98
Hexoses		0.12		3.01	0.15	2.87	5.09			
Pentoses		0.03		0.03		0.03	0.65	0.65		
Ethanol								2.47	2.45	
HCl			0.08	0.08						
Water	9.26	45.11	0.14	33.99	18.98	52.93	74.76	75.76	0.13	0.70
Other		0.02		0.12	0.01	0.21	0.55	0.92		

Figure 2.3 Schematic flowsheet for the CASH process. Flow rates in ton/h.

and assumes no degradation and formation of by-products at hydrolysis or fermentation.

The assumptions of input data (resting on different sets of experiments) and of process layout for the CASH-process (Fig. 2.2) differ somewhat from the CASH-study above and therefore results are not strictly comparable. 61 % of the theoretical content of hexoses are converted and recovered for fermentation and 2 % remains in the solid fuel product. Of the potential pentoses 54 % are converted and recovered for fermentation. In the fermentation 100 % of the hexoses are converted to ethanol (of which 1 % is lost in the distillation) but none of the pentoses.

In the enzyme process 6 % of the pretreated (by acid hydrolysis) and washed wood is diverted to production of enzymes (several different cellulases) for the following hydrolysis and breakdown to simple sugars (Fig. 2.3). 81 % of the theoretical content of hexoses in the feedstock is converted and recovered for fermentation and 2.5 % is lost with the solid fuel. The remainder is not converted or has formed by-products; 69 % of the potential pentoses are recovered for fermentation. Nearly 94 % of the hexoses, but none of the pentoses is fermented to ethanol. The remainder, together with a small amount of pentoses, is consumed to build yeast biomass. The surplus of which is anaerobically converted to methane together with unfermented sugars and other dissolved organic matter.

The yield of ethanol in the CASH-process agrees closely with the previous study (2.1), but the surplus of solid "lignin"-fuel is somewhat higher.

The enzyme route yields about 25 % more ethanol (32.4 % in energy terms, which is 66 % of the theoretical) but considerably less solid fuel product for sale, which would be lower even if electricity was purchased rather than produced internally from lignin. The total energy yield is somewhat lower in the enzyme process because of the different input data and feedstock consumption for the enzyme production. Similar results have been published in an earlier Canadian study on the IOGEN enzymatic process.

The results of the R&D work on pentose fermentation was not incorporated in the Lund studies but should be incorporated in coming studies. The work seems to make the development of genetically modified yeasts likely, which are capable of producing ethanol from both pentoses and hexoses, and which are considered robust and easily handled for industrial use. Before saying that an industrial process is ready for commercialisation, operation of a pilot plant is considered necessary to tie together the different process steps (now studied separately) into a continuously running plant.

Table 2.2

	Mass or volume t/yr or m ³ /yr		Energy TJ/yr		Energy yield, % (LHV-basis)	
	CASH	ENZYME	CASH	ENZYME	CASH	ENZYME
IN						
Wood-DS mass	100,000	100,000	1,920	1,920		
(Steam internal production	263,500	172,000	450	342)		
(Electricity, net internal production			69	105)		
Chemicals mass	6,255	3,060				
OUT						
Ethanol, vol 100%	23,270	29,460	491	622	25.6 (24.7) ⁺	32.4 (30.7) ⁺
Lignin-DS mass	32,930	26,077	734 ^x	377 ^x	38.2 (42) ⁺	19.6 (26.4) ⁺
(Lignin-DS internal use	12,605	9,156	281 ^x	204 ^x)		
(Methane intern.use	6,465	6,682	323	334)		
Hot water >100 °C	none	none				

^x 22.3 MJ/kg DS assumed

⁺ if electricity is purchased and included in input as such.

2.3 The DOE/NREL-study

The NREL-process (DOE 1993) rests upon enzymatic hydrolysis, following an acid pre-treatment, and simultaneous saccharification (hydrolysis) and fermentation (SSF) and uses micro-organisms also capable of fermenting pentoses. This is important as it is applied to hardwood and herbaceous feedstocks, which have high contents of hemicellulose-pentosans as compared to softwoods.

The wood raw material consists of debarked hardwood chips containing 46.2 wt-% of cellulose, 24.0 % pentosans, and 24 % lignin on wood-DS basis. The theoretical ethanol yield is 509 litres of pure ethanol (100 %) per ton wood-DS, assuming 100 % conversions and recoveries. As shown in *Table 2.3* on mass and energy balances, based on the study data, the actual yield was 343 litres/ton. Surplus of lignin fuel, not consumed internally to cover the steam and electricity needs of the plant, can be converted to electricity for sale.

The NREL process (*Figure 2.4*) has formed the basis for preliminary engineering studies in a later IEA-study (2.4) and also was the basis for a recent offer (SWAN) for commercial application, exemplified in a (wheat) straw project in Sweden (2.5).

In the acid pre-hydrolysis step 80 % of the pentosans are converted to pentoses (but only a few percent of the cellulose to hexoses), the remainder is left unconverted or forms by-products (furfural). A slipstream of the hydrolysate is used for enzyme production (cellulases) and the main stream is fermented to ethanol by using pentose fermenting bacteria. After this step the ethanol-cellulose stream joins the enzyme stream for the simultaneous cellulose hydrolysis and hexose fermentation to ethanol (SSF), which takes 7 days. 72 % of the cellulose is converted to ethanol and 13 % is unreacted, the remainder forms by-products or is consumed to grow micro-organisms. After distillation to recover ethanol, and separation of solids the surplus of this biomass is anaerobically fermented (together with other dissolved organic matter) to methane, which is used as an internal fuel to produce steam. The solids are used for production of additional steam and surplus steam is used to produce electricity for sale in condensing turbine.

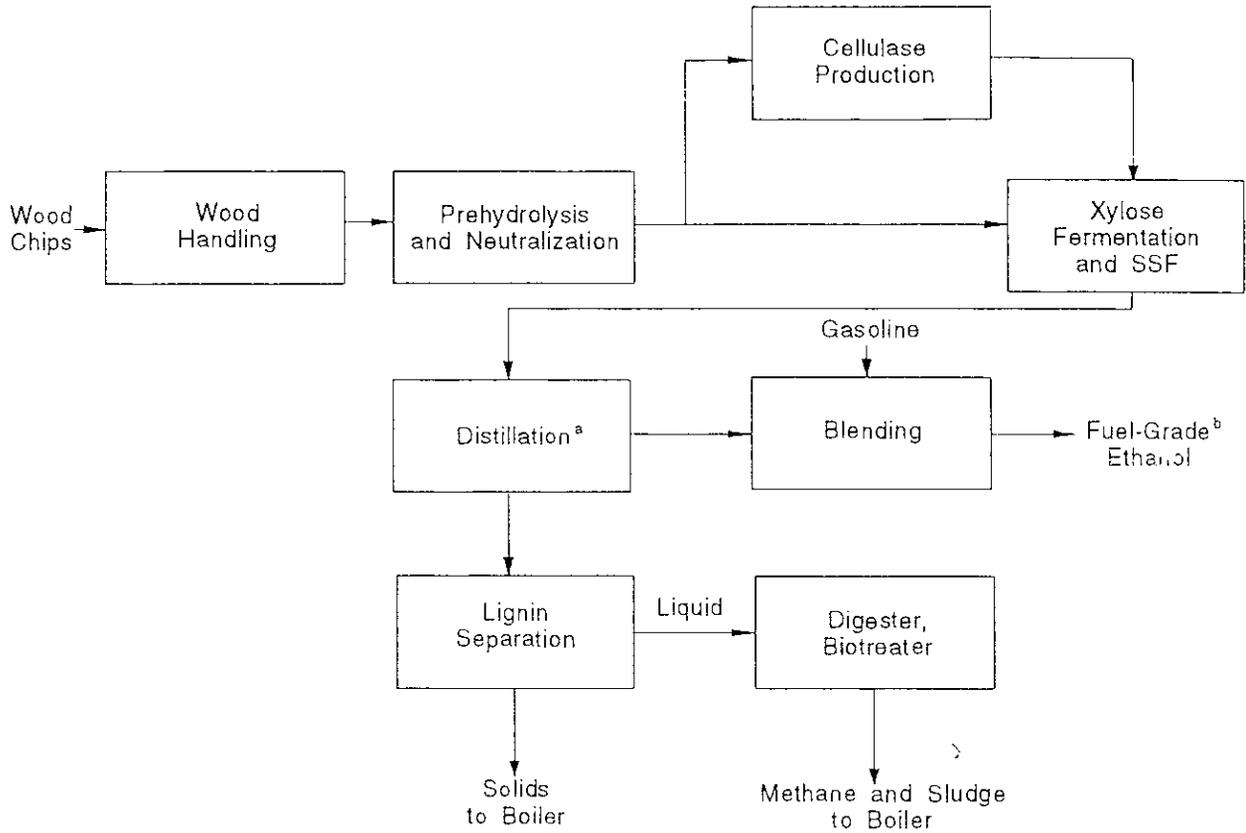
Table 2.3 summarises overall mass and energy balances based on the DOE-study.

If surplus lignin was not converted to electricity (in a condensing turbine) but sold as solid "lignin" fuel, that would correspond to about 1,000 TJ/yr or somewhat more than 9 % of input wood-DS energy (LHV), giving a total energy yield of just above 47 %.

The present ethanol yield is 67 - 68 % of the theoretical. In the future R&D is assumed to push this yield to nearly 90 % by improved conversion to sugars, and conversion times can be considerably shortened. The ethanol energy yield would in turn increase to nearly 51 % (LHV) for the chosen feedstock with no change in surplus electricity (or lignin fuel).

The study also describes a case for a much larger plant for 1.1 million m³/yr of ethanol (azeotrope). However, conversion and yield figures are the same as for the smaller plant.

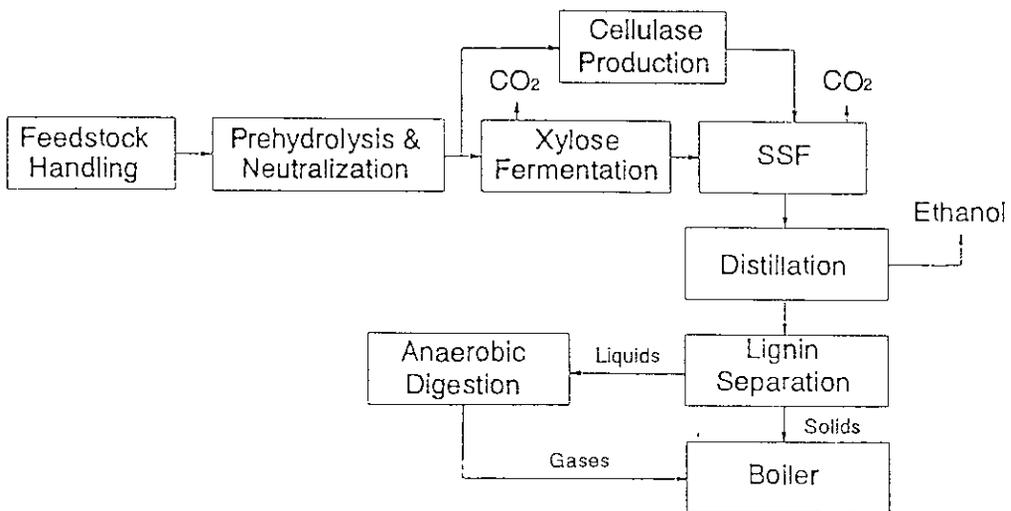
Figure 2.4 — Overview of Wood-to-Ethanol Process



^a Separation to azeotropic ethanol.

^b 90.3 percent (of weight) ethanol, 4.7 percent water, 5 percent gasoline.

Source: DOE



Schematic of the NREL SSF-xylose fermentation process.

Table 2.3

	Mass or volume t/yr or m ³ /yr	Energy TJ/yr	Energy yield, % (LHV-basis)
<u>IN</u>			
Wood-DS, mass	580,480	11,030 ^x	
(Steam, mass internal production	959,090		
(Electricity internal production		655)	
Chemicals, mass	38,250		
<u>OUT</u>			
Ethanol (100 %), volume (produced as azeotrope)	199,255	4,204	38.1
(Lignin-DS, mass internal use	260,770	4,425)	
(Methane, mass internal use	20,100	1,005)	
(Electricity, internal production and use		655)	
Electricity, surplus, 13.3 MW		385	3.5

^x assumed 19 GJ/t DS (not given in report).

2.4 The IEA/NREL study

The IEA-Agreement Motor Fuels group has evaluated (IEA 1996) the NREL wood-to-ethanol process layout, using the same input data on conversions and recoveries as in the DOE-study, but using a different feedstock (red oak with a content of nearly 76 wt-% cellulose and hemicellulose, 46 and 29.6 % respectively). The theoretical yield of ethanol from this feedstock is 43 wt-% (i. e., 549 litres per ton).

Unfortunately, no details are given (only boundary conditions) on mass, steam and power balances or on how intermediate products (methane, lignin) are used as internal fuels. The feedstock composition is very different from what is used in the studies referred to above. Comparisons with the other processes

are therefore uncertain. The following overall data (*Table 2.4*) is given for a stand-alone, energy self-sustained plant at a throughput of 1,818 t/d wood-DS during 335 d/yr.

Table 2.4

	Mass or volume t/yr or m ³ /yr	Energy TJ/yr	Energy yield, % (LHV-basis)
<u>IN</u>			
Red oak wood, DS mass	609,000	11,800 ^x	
Chemicals			
<u>OUT</u>			
Ethanol (100%) volume	219,000	4,620	39
Electricity (13.3 MW)		385	3.3

^x assuming LHV = 19.4 MJ/kg DS (the report states 18.44 MJ/kg as HHV!).

The ethanol yield is about 66 % of the theoretical assuming that the hemicellulose consists of more than 90 % pentosans. The content of cellulose and hemicellulose in the feedstock is extremely high (compare with 62 - 66 % in the previous studies).

In the future it is assumed that R&D will result in improved performance (advanced SSF), pushing the ethanol yield from pentoses from 90 % to 95 % and from cellulose from 72 % to 90 %. The ethanol production will then increase to 280,000 m³/yr and the yield of ethanol in energy terms will increase to 50 % (LHV). Surplus power will, however, decrease somewhat to 10.6 MW.

2.5 The SWAN Biomass study

SWAN Biomass/Stone & Webster has presented a study to Sweden (Report SSEU 1996-12-27, and pers. comm. from JL 1997) describing an ethanol project based on (wheat and) wheat straw. The wheat straw part of this is of interest as it rests upon the NREL process referred above in the latest version (*Figure 2.5*). Fermentation is accomplished by using a proprietary, modified yeast capable of fermenting both hexoses and pentoses in one fermentation stage, simultaneous with the enzymatic cellulose hydrolysis. Process guarantees can be given, following validation test runs with the actual feedstock in the 1 t/d pilot unit in Colorado.

The available study report (*Enclosure 2*) is unfortunately brief and incomplete as it does not include proprietary know how and does not give steam and power

Wheat Straw to Ethanol Block Flow Diagram

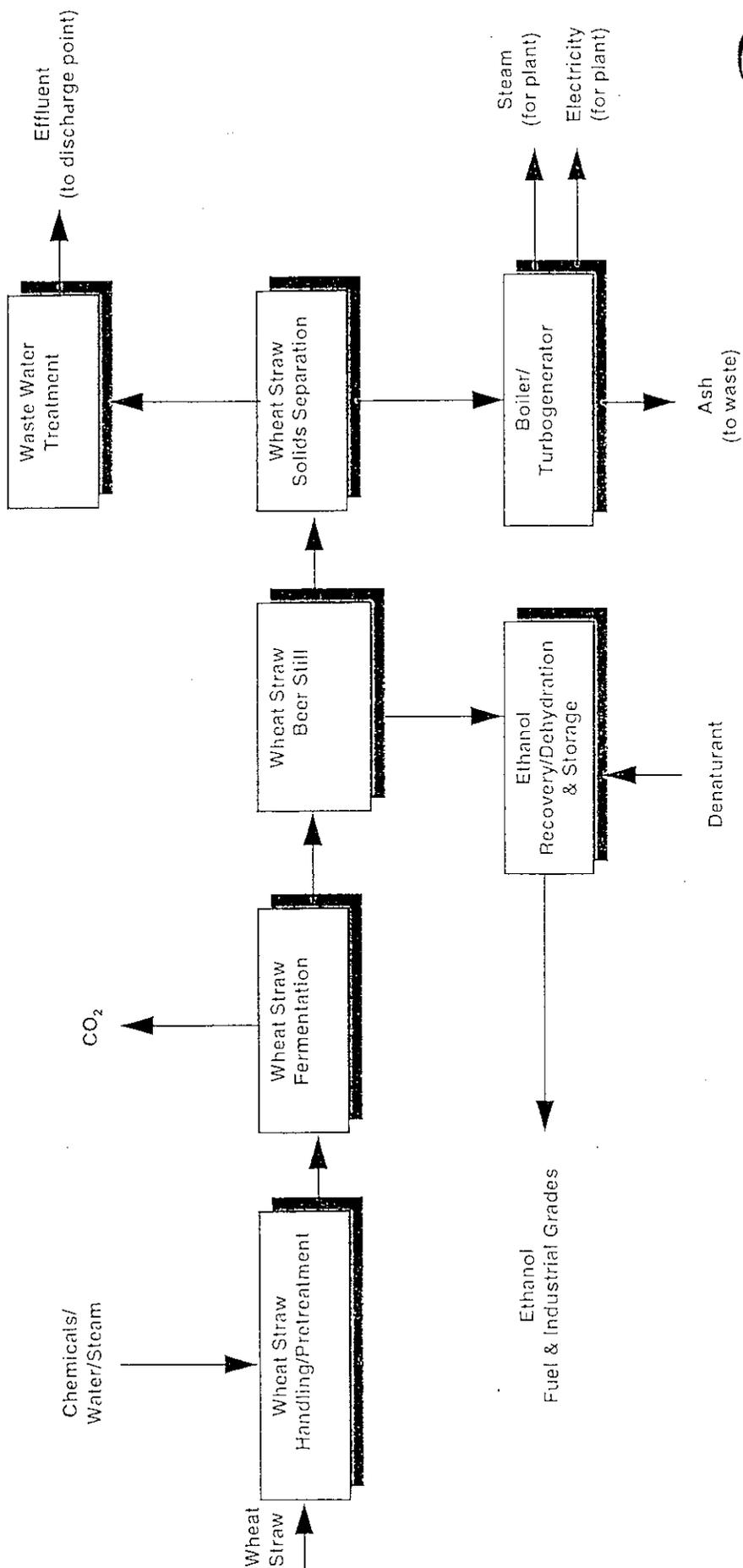


Figure 2.5



balances or information on how intermediates are used internally. The following balances (*Table 2.5*) have been constructed from the study data (straw case) and the supplementary information (pers. comm. JL 1997) for a stand-alone, energy self-sustained plant producing 50,000 m³/yr ethanol (60 % as absolute, 40 % as azeotrope) from about 158,000 t/yr straw-DS in 330 days. The straw contains 38 wt-% hexosans and 19 % pentosans, which corresponds to a theoretical ethanol yield of 415 litres per ton.

Table 2.5

	Mass or volume t/yr or m ³ /yr	Energy TJ/yr	Energy yield, % (LHV-basis)
<u>IN</u>			
Straw-DS, mass	158,170	2,690	
(Electricity, internally produced)		152	
(Steam, internally produced)	280,000	366)	
Chemicals, mass	13,350		
Enzymes, mass	5,740		
<u>OUT</u>			
Ethanol (100 %), volume	49,515	1,045	38.8
(Methane, internal use, mass)	2,511	125)	
(Lignin-DS, internal use mass)	25,750	515)	
Lignin-DS, surplus for sale mass	31,250	625	23.2

The ethanol yield (313 litres per ton straw-DS) is 75 % of the theoretical and corresponds in energy terms to nearly 39 % of the energy content of the feedstock (LHV-basis). Surplus lignin is 23 % of the feedstock energy, and the total yield of energy products is nearly 62 %.

2.6 Summary and discussion

Figures from the studies referred to above, which are of interest from yield and efficiency point of view have been summarised in the *Table 2.6* for stand-alone, energy self-sustained plants. Most important with regard to the ethanol yield is, of course, the content of hemicellulose and cellulose (*Figure 2.6*). This is also the most important cost factor.

The quality of the feedstock therefore is very important, specifically its content of bark and other non-cellulose compounds, which yield little or no ethanol. Most studies have been based on selected feedstocks of pure hardwood with about 70 wt-% convertibles or higher, while real world's large scale raw materials would range around 60 % due to the presence of bark and the

Table 2.6

Study	SSEU	LUND		DOE	IEA	SWAN
Process	CASH	CASH	ENZ.	NREL	NREL	NREL
Feedstock	Pine	Pine	Pine	Hardwood	Red oak	Straw
Wt-% content of hemi.+cell.	66	61.6	61.6	70.2	75.6	57.2
Theor. ethanol yield, lit./ton	476	445	445	509	549	415
Actual ethanol yield, lit./ton	224	233	295	343	360	313
% of theor.	47	52	66	67	66	75
Energy yield ethanol, % (LHV)	25	26	32	38	39	39
"Lignin", % (LHV)	23	36	38	20	9	9

inclusion of softwoods, such as in forestry residues. Short rotation forestry (SRF) species such as willow typically have about 60 % convertible hemicellulose and cellulose.

Conversion efficiency (theoretical to actual ethanol yield) is about 67 % when applying the enzyme route with a top value of 75 %. This 75 % was taken from the latest study based on straw. This corresponds to an ethanol energy yield of 39 % (LHV).

In all cases methane and the solid residue are enough to make plants energy self-sufficient, and provide a surplus of solid fuel, to bring the total yield of energy products to the range of 50 - 60 % (there is a remarkable difference between the SWAN-study and the DOE/IEA-studies, which cannot be explained until detailed engineering calculations are available). The total yield will be lower as the conversion to ethanol increases, and the yield will be higher the better the heat management is in the plant.

With improved conversion efficiency to ethanol of 85 % (based on nearly 90 % conversion at hydrolysis and fermentation together and a few percent consumption for enzyme production and losses) an average feedstock in the future would be expected to yield 370 litres of ethanol per ton dry wood (about 41 % on LHV-energy basis). An overall conversion efficiency to ethanol of 85 % is considered to be the practical maximum in future plants. Yield figures used in the studies are typically about 80 % of this practical maximum.

Costs of production have not been a special focus in this report, since no new information has been published on basis of new engineering studies. Often quoted figure on future production costs (technique extrapolated as described above) for ethanol from wood is USD 0.26 per litre (USD 12.3/GJ (LHV) at mid-1990 cost level; IEA 1993). The estimate rests upon a selected wood feedstock with very high content of cellulose + hemicellulose (76 %), low cost of ~~feedstock with very high content of cellulose + hemicellulose (76 %), low cost of~~ feedstock from huge SRF plantations and large plants (200,000 m³/yr of ethanol or more). These are not typical for European conditions and cannot be said to represent the future production costs here.

With more normal lignocellulosic feedstocks and costs and moderate plant sizes Swedish evaluations (KFB 1997) indicate ethanol production costs to be not below the range of SEK 3.5 - 4 per litre (USD 0.47 - 0.53 per litre at current exchange rate, USD 1 = SEK 7.50) or higher (IEA 1996, Brooks 1996) and above what will be obtained from plants based on surplus wheat grains.

Investment costs expressed as cost per litre/year of ethanol or per kW ethanol capacity are for the smaller European plants SEK 12 -13 per litre/year (about USD 1.7 per litre/year) and about USD 0.7 per litre/year in the US estimates for the bigger plants. Expressed as cost per kW ethanol capacity the investment is estimated to about SEK 16,000/kW (USD 2,100/kW) for the small plants and about USD 1,000/kW for the bigger US plants. For transformation to ECU the rate is assumed to be ECU 1 = USD 1.13.

To summarise: Ethanol production costs from average wood feedstock (à ECU 9.5 per MWh wood-DS; 12 % annuity) are estimated to be in the range of ECU 0.41- 0.47 per litre of ethanol (as azeotrope) in plants with a capacity of the magnitude 100,000 m³ per year.

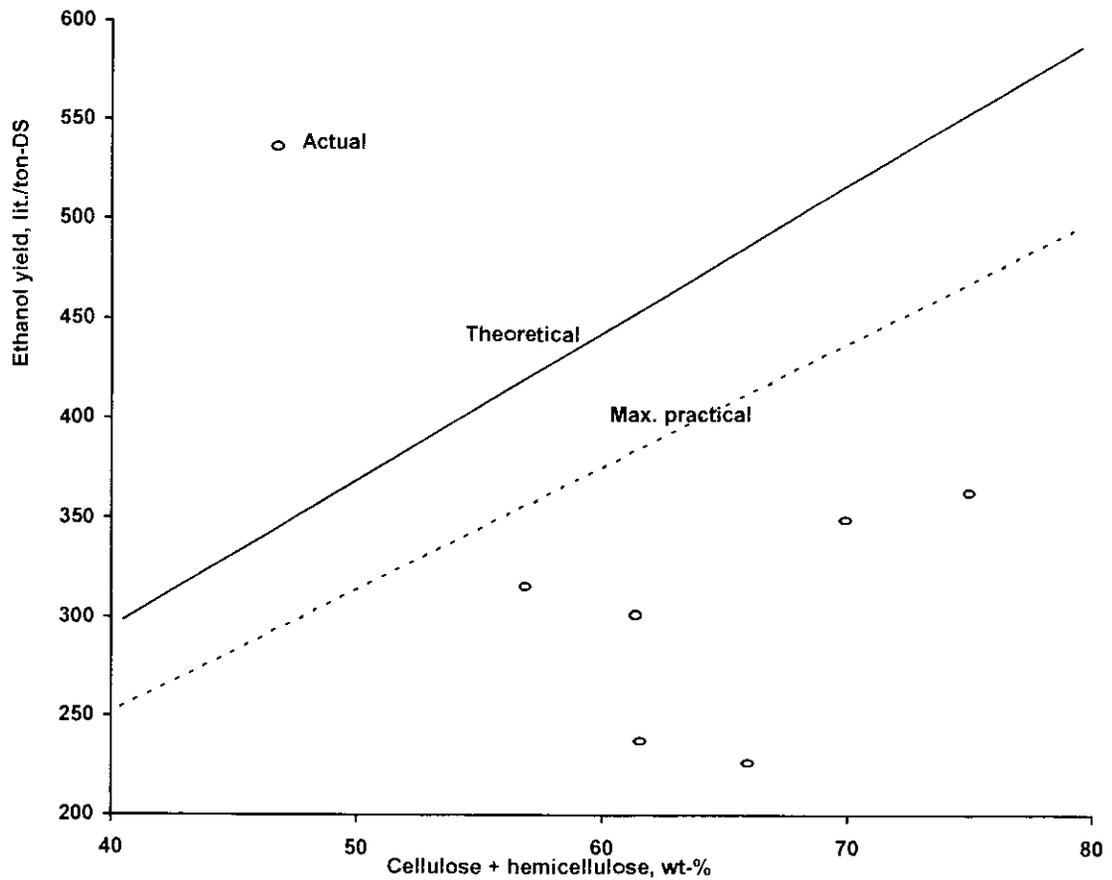
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Figure 2.6 Theoretical and actual ethanol yields



2.7 References

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3. ENGINE TESTS WITH ALCOHOL CONTAINING FUELS

The results of the tests (*Enclosure 3*) with alcohols emulsified in diesel oil are commented by Volvo Truck Corp. (VTC) on the following pages. The earlier experiences of VTC on ethanol fuels from the interim report are included.

The tests at Luleå University confirm that blending of alcohols can lead to lowered emissions of NOx and particles, although the NOx reduction did not materialise with ethanol contrary to earlier test results from Australia and Sweden. The causes may have been not quite successful emulsification or engine to engine variability. Therefore additional tests are required for clarification.

There was an evident beneficial effect of the ethanol emulsions on emissions of particles, PAHs (particle bound and semi-volatile), and on the biological activities (mutagenic activity at Ames test and dioxin receptor binding activity). Corresponding data on PAHs and on biological activity for methanol emulsions are not available as no such sampling was accomplished. True low emission fuel was the ethanol only fuel (with denaturants and ignition improver), as considerable reductions of NOx, particles, PAHs, and biological effects were recorded. Again, no comparative data for methanol are available, as no engine for this fuel was tested.

The superiority of the Swedish environmentally classified diesel oil (MK 1, characterised above all by very low content by PAHs as triaromatics+) over the European standard grade are quite evident concerning emissions of NOx and particles and the corresponding effect on biological effects. The beneficial effect of oxidative, catalytic clean-up, made possible by also the very low content of sulphur of the MK 1 (<20 ppm), is also notable.

Alcohols for emulsification in diesel oils need not to be nearly water-free as is the case when blending with gasoline. The ethanol/water azeotrope or crude methanol might well be used for emulsification. The blends of diesel oil and alcohols must be standardised regarding the energy content to reasonably maintain the performance of the engines.

Receiver(s):

Ecotraffic R&D, Stockholm Sweden
Archives 24600, VTC

1. Introduction

This report is a part of the ALTENER project "Feasibility Phase Project For Biomass Based Alcohols for Automotive Fuels and Industrial Uses" with contact no XVII/4.1030/AL/124/95/SWE and summarize the engines tests at Luleå Technical University with methanol/diesel blends on a heavy duty truck engine.

This work was performed by Luleå Technical University.

The results are compared with previous results with ethanol/diesel blend.

2. Engine specification

The tests was performed with a Scania engine with the following specification:

Type: DSC11-24

Cylinder:6

Cylinder volume:11.0209 litre

Compression ratio: 17:1

Injection pump: EDC

Power:180kW

Turbo charged and inter cooled

3. Fuels

Diesel fuel EC1 (environmental class 1)

Methanol/Diesel EC1 blend with 11.6 vol-% methanol.

Ethanol/Diesel EC1 blend with 15 vol-% ethanol.

The alcohol concentration in the mixture was adjusted in order to obtain the same calorific value in both blends.

4. Performed tests.

- Full load tests
- Performance and emissions, regulated and unregulated according to a 22-mode cycle. The modes for the ECE-R49 cycle was included.

5. After treatment

Catalyst from Johnson Matthey type DF-07 was used.

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Performance and emission tests with alcohol/diesel blends at Luleå UniversityIssue No
3

Receiver(s):

6. Results

6.1. Regulated emissions

		13-mode cycle values g/kWh (ECE R49) ref.1		Modal driving cycle * g/kWh. Ref.2	
		Diesel EC1	MeOH/EC1	Diesel EC1	EtOH/EC1
CO	without cat.	0.81	0.75	1.4	1.8
	with cat.	0.13	0.20	1.6	1.6
HC	without cat.	0.35	0.37	0.68	0.80
	with cat.	0.19	0.19	0.77	0.81
NO _x	without cat.	5.89	5.33	6.9	7.1
	with cat.	5.94	5.34	6.9	7.1
NO ₂	without cat.	0.38	0.33	0.55	1.06
	with cat.	0.45	0.16	0.79	0.72
NO	without cat.	5.54	5.02	6.26	5.94
	with cat.	5.51	5.20	6.07	6.35
PM	without cat.			0.17	0.11
	with cat.			0.12	0.062
BSFC		236	245	362	357

* driving cycle designed for sampling of PM and PAH.

The MeOH/EC1 blend indicated 35% lower particulate emissions compared EC1 diesel. The engine was equipped with catalyst in both cases.

6.2. Unregulated emissions

Formaldehyde and acetaldehyde measured after the catalyst was not detectable.

Ethane and propane after the catalyst was not detectable.

Ethanol emission was not detectable before and after the catalyst.

Measurements indicates very low methanol emissions

Previous results with the same engine and EtOH/EC1 blend shows a 35% reduction of total PAH emissions when comparing with EC1 diesel fuel.

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Receiver(s):

7. Discussion

Engine performance

Comparison with the full load data indicate, as expected 4 to 5% loss of power and torque with the MeOH/EC1 blend but slightly improved efficiency.

Emission results

It was expected that both blends should give similar influence on the emissions.

The results indicates that this is not the case.

The emission levels with the EtOH blend are strange.

The reason can be either measurement errors during the tests or the influence of the driving cycle.

VTC is not confident with the EtOH results. Some NOx and smoke reduction were obtained at earlier tests with a 7 lit Volvo engine (ref.3).

The MeOH blend gives similar emissions with respect to CO and HC, as EC1 diesel fuel. In the contrary the EtOH blend indicates higher emissions compared with the same fuel.

MeOH caused a reduction of total NOx with 10%. On the contrary EtOH increase the NOx with approximately 3%.

I general, blending alcohols in the diesel fuel will decrease the cetane number of the blend which will cause an increase in NOx.

This should be compensated by the higher heat of evaporation of the blend which will lower the combustion temperature therefore the NOx production will decline.

The behaviour of the two blends differs. Further tests are necessary in order to investigate and explain the reason of the contradiction.

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Receiver(s):

General comments about the fuel specification

Alcohols are corrosive to many of metals. Corrosion products in the injection system can increase the wear. Fuel parameters which affect the corrosion ability are the water content and the acid number.

The lower viscosity and the poor lubricity will also cause increased wear in a conventional injection equipment.

Elastomers and plastic components can be attacked. The correct choice of fuel system materials is very important.

The effects from the alcohol/diesel blend will probably be less critical than with the pure grades.

The risk for cavitation increase with the blends. The injection system have to be adapted to the fuel specification.

The water contamination (pick-up) and phase separation is another problem. Pure grades should have lowest possible water content. EtOH is more hygroscopic than MeOH.

The boiling range of the diesel fuel will change when blending alcohol with diesel.

Lowering the boiling point by adding alcohol will increase the risk for vapour lock. This will also affect the warm start ability of the engine.

Further, blending diesel with alcohols will decrease the flash point and affect the fire safety. Workshops classified as fire class 1, for gasoline and alcohol vehicles have to be used.

References

Ref. 1 Nordström, F. (Dept. of Environment Technology, Engine Laboratory, Luleå University). Investigation of exhaust gas emissions from Scania DSC11-24 engine using an emulsion of methanol in diesel oil. November 1996.

Ref. 2 Westerholm, R. et al (Stockholm University), Haupt, D. (Luleå Technical University). Chemical and biological characterisation of exhaust emissions from ethanol and ethanol blended diesel fuels in comparison with neat diesel fuels. Report as KFB-meddelande 1997:17

Ref. 3 Gjirja, S. Olsson, E. (Chalmers University of Technology). Experimental investigation on emission's performance of the Volvo TD73 engine running with MK1 and Etamix E15. August 1994.

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Appendix

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Reg. No

24624-61456

Subject

Emission Data on Alcohol Fueled Heavy Duty Diesel Engines.

Issue No

1

Receiver(s):

Ecotrafic R&D AB, Stockholm, Sweden
Archives 24600

1. Introduction

This report is part of the ALTENER project "Feasibility Phase Project for Biomass Based Alcohols for Automotive Fuels and Industrial Uses" with contract no XVII/4.1030/AL/124/95/SWE and covers previous in house results at Volvo Truck Corporation (VTC) regarding alcohol fueled engines for trucks and busses.

2. Summary

VTC have developed ethanol versions of the 7 litre truck diesel engine and the 10 litre bus diesel engine. The ethanol fuel have ignition improver - 7 % by volume Beraid - and the converted engines have diesel type combustion system. The prototype engines are used in ethanol fuel fleet tests supported by the Swedish authority Kommunikationsforskningsberedningen (KFB).

The table below shows the emission results in the ECE R49 test cycle.

Exhaust emissions, ECE R49, g/kWh

	NOx	HC	CO	PM
Truck engine A7B260	3,8	0,20	0,16	<0,05
Bus engine AH10A245	3,9	0,14	0,02	<0,05

3. Background

An important part of VTC:s long-term environmental work is the search for optimum alternative fuel concepts.

The main criteria that must be considered are:

- * Life cycle energy efficiency.
- * The potential to be manufactured from major renewable sources.
- * Cost effectiveness.
- * Life cycle emissions.

Important projects, among recent work on alternative fuel concepts within VTC, are the conversion of a 7 litre truck diesel engine to ethanol fuel (diesel process) version, called A7B260 and a similar project on a 10 litre bus engine called AH10A245.

The engines are used in minor field test programs with five trucks and two busses and the projects are supported by the Swedish authority Kommunikationsforskningsberedningen (KFB), (Project no 92-230-742 and 94-141-742).

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Emission Data on Alcohol Fueled Heavy Duty Diesel Engines.Issue No
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Receiver(s):

4. Project data

4.1. Fuel

Ethanol with ignition improver Beraid (polyethylene glycol) at the range of 7 % by volume (9,5 % by weight).

4.2. Engine specifications

4.2.1. Truck engine A7B260

- 6 cylinder, straight
- 6,7 litre swept volume
- Turbocharged
- Inter-cooled
- Direct injection
- Compression ratio 22
- Max power 192 kW at 2400 rpm
- Peak torque 1040 Nm at 1440 rpm
- Bore x Stroke 104,78 x 130 mm

4.2.2. Bus engine AH10A245

- 6 cylinder, horizontal
- 9,6 litre swept volume
- Turbocharged
- Inter-cooled
- Direct injection
- Compression ratio 23
- Max power 180 kW at 2000 rpm
- Peak torque 1050 Nm 1250 rpm
- Bore x Stroke 120,65 x 140 mm

4.2.3. Technical solutions

The most important technical changes on the ethanol engines compared to the diesel fuel versions are:

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Emission Data on Alcohol Fueled Heavy Duty Diesel Engines.Issue No
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Receiver(s):

- Increased compression ratio
- Modified piston bowl
- Geometrical and material changes in the fuel injection system
- Adjusted turbocharger size
- An added fuel delivery pump on the bus engine
- Exhaust oxidation catalyst
- Cold start aid system with electric inlet air heater element and exhaust pressure governor (also available on diesel version)

5. Results

The engine performance and emission goals were reached. The emission results are given below.

5.1. Regulated exhaust emissions

Exhaust emissions, ECE R49, g/kWh

	NOx	HC	CO	PM
Truck engine A7B260	3,8	0,20	0,16	<0,05 ¹⁾
Bus engine AH10A245	3,9	0,14	0,02	<0,05 ¹⁾
Aim	4,0	0,3	0,15	0,05

1) PM was not measured but is well below the 0,05 level.

5.2. Unregulated exhaust emissions

Exhaust emissions, ECE R49, g/kWh

	Aldehyde ²⁾
Truck engine A7B260	0,05
Bus engine AH10A245	0,03
Aim	0,05

2) Formaldehyde plus acetaldehyde.

4. LIFE CYCLE ANALYSIS CONSIDERATIONS

It is quite obvious from the studies on both methanol and ethanol production that conversions from lignocellulosic feedstocks can be accomplished by using only renewable energy, i. e. fuels which are by-products or part of the feedstock.

In the conversion step of the chain from feedstock to accomplished transport work ("life cycle") thus no fossil energy is used except what might have been used for production of input chemicals and for the production of plant equipment. These inputs are, however, with certainty small in comparison to the energy contents of the products.

Energy usage occurs in the earlier steps of the chain, e. g. energy inputs in forestry and agriculture (silviculture, SRF cultivation, grains cultivation), harvesting, and transports to the conversion plants. Energy inputs for the manufacture of fertilisers, pesticides, and engine fuels (and equipment) are today based on fossil fuels. In biomass feedstock preparation the ratio of biomass energy output and inputs are in the range of 4 - 35 with the low figure for grains/straw production and harvesting (intensive agriculture) and the high value for recovery of trees and tree residues (extensive silviculture). Short rotation forestry (SRF) has intermediate values, typically a ratio about 20.

In theory much of the energy inputs, which are fossil-based today, may be bio-based in the future, for instance nitrogen fertilisers and engine fuels, the ratio will then be greatly increased. It is therefore important to clearly define the system and its limits when presenting an LCA.

When including the efficiencies at the conversion to transport fuels such as methanol and ethanol, the whole chain (distribution to consumers included) can be illustrated as in *Figure 4.1 and 4.2 (see ref.)* for cases when fossil based chemicals and fuels are used in preparation of feedstocks to the plant and at motor fuel distribution. *About ten times as much transport fuel energy in form of methanol or ethanol can thus be obtained for each unit of fossil energy input in a properly designed system.* This ratio can be increased by the developments to improve efficiencies in the conversion step. It can, however, also be considerably lower when fossil auxiliary fuels and electricity are used at the production. Such systems should not be accepted other than during transitions to fully renewable systems.

REFERENCE: The Life of Fuels - Motor Fuels from Source to End Use.
Ecotraffic AB. Multiclient study. Stockholm March 1992.
ISBN: 91-88370-09-7.

METHANOL

(TREE RESIDUES)

Energy
(% of fuel in tank)

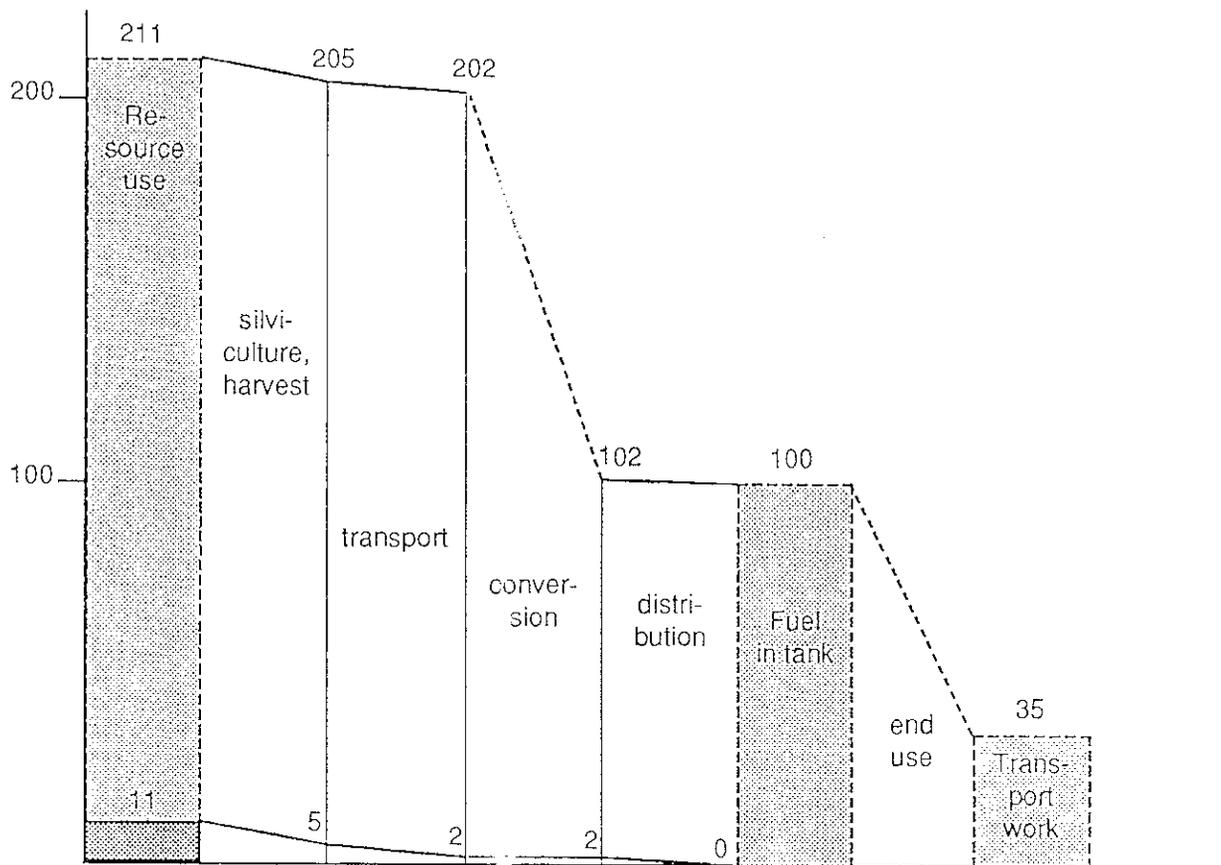


Figure 4.1 Energy turnover in the fuel chain tree residues to methanol use in heavy duty diesel engine operated in bus cycle.

-  Shaded box = state of fossil energy
-  Shaded box = state of renewable energy
-  Solid line = alteration of fossil energy use
-  Dotted line = alteration of renewable energy use

ETHANOL

(TREE RESIDUES; CASH-PROCESS)

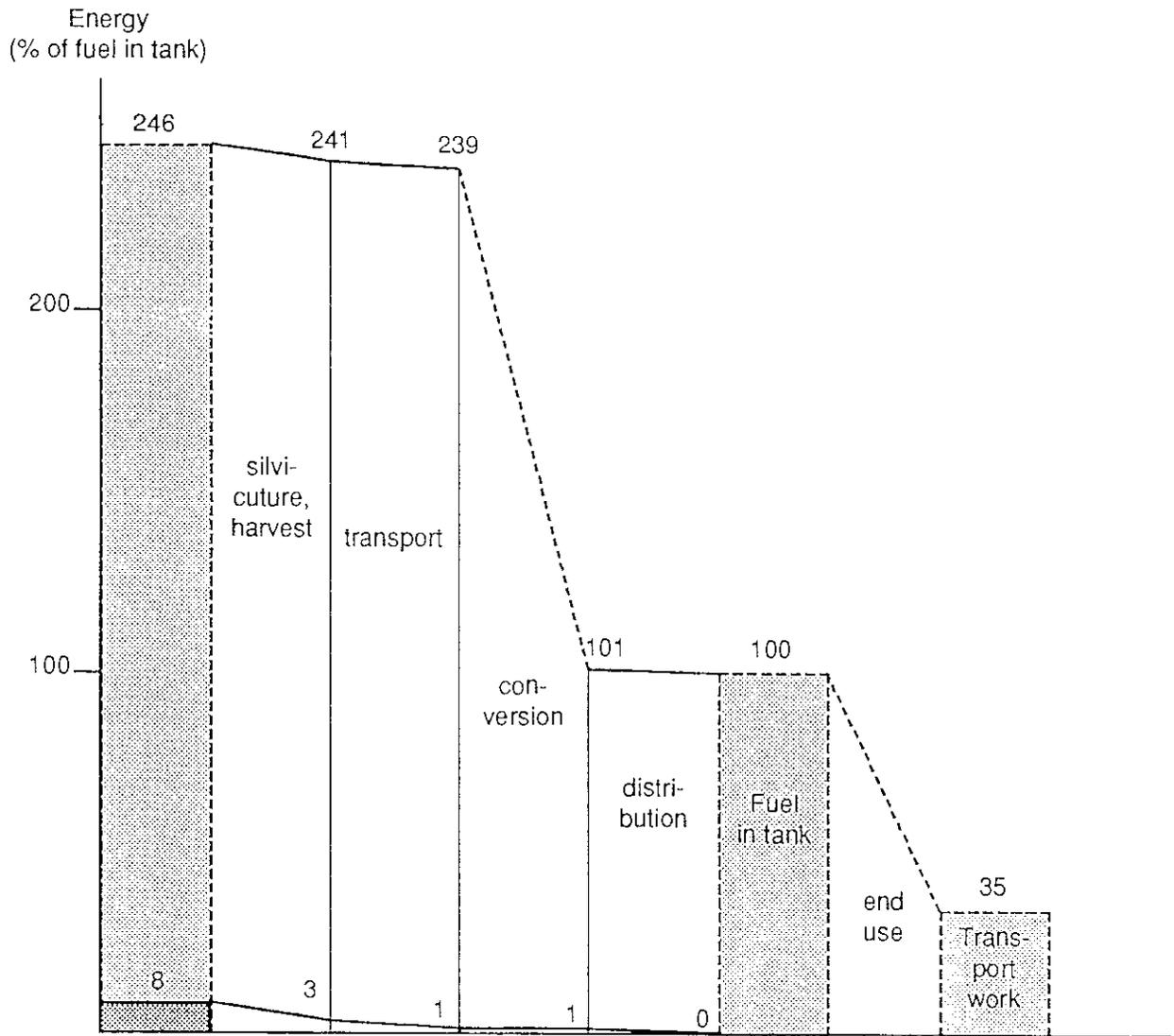


Figure 4.2 Energy turnover in the fuel chain tree residues to ethanol use in heavy duty diesel engine operated in bus cycle.

-  Shaded box = state of fossil energy
-  Shaded box = state of renewable energy
-  Solid line = alteration of fossil energy use
-  Dotted line = alteration of renewable energy use

5 INDUSTRIAL ACTORS FOR A DEMONSTRATION PLANT

5.1 Commercial demonstration - objects and establishment criteria

In the current report, technically viable routes for the conversion of biomass to transportation fuels have been documented. The report demonstrates that the primary interest is to be focused on biomass gasification followed by conversion of the syngas produced into methanol and/or its derivatives and analogues.

The subsequent phase of efforts will clearly focus on selecting a raw materials base, a site, a process configuration, product slate, and a commercial investor partnership and alliances which will meet the following criteria:

- Robust investment object in terms of product flexibility.
- Sustainable long term operation.
- Plant scale minimum to meet commercial demonstration objective.
- Transport fuel produced which will be compatible with major automotive actors views.
- Overall economics to provide a commercial return after allowing for compensation for small scale and short term market aberrations in pricing of competing fossil-based fuels.
- Possibility to organise a substantial fleet demonstration for the biomass derived fuel.

5.2 Work tasks to establish the Commercial Demonstration Phase

Subsequent work will focus on analysing and selecting project specifications, as they also are embodied in a technical and commercial prospectus for investors. The project specifications shall meet the stated criteria and can be discussed under the four headings - site, process configuration, product slate, and partners/financing.

An overriding consideration is that the selection of project specifications shall be made in such a way that a successful commercial demonstration is as far as possible assured - i.e., a commercial return on the partnership's risk capital shall be likely in view of the parameters at hand. A synopsis on the development of the documentary basis for the Commercial Demonstration Phase will deal with issues as follows.

5.3 Outline of a synopsis for a planning document for the Commercial Demonstration Phase

5.3.1 Site

Even for a fairly modest demonstration plant, the logistic task of securing biomass feedstock is a significant one.

The ideal candidate sites are sites where substantial volumes of forestry waste material is at hand or can be transported at low costs - i.e., in conjunction with pulp mills, preferably with integrated sawmills. At such sites the logistic investment is largely in place and cross haul costs would be minimised on the way from the logging/harvesting operation to the plant.

Also, the ideal site should provide a possibility of heat and utility integration with the commercial demonstration site.

In view of economics and long term robustness, co-production of power shall also be a possibility at the selected site.

Consequently the site will be calculated based on the total optimum of above factors.

5.3.2 Process Configuration

The current report demonstrated one technically viable process route to methanol with reasonable investment and efficiency characteristics. This issue should be revisited before the final selection is made to check on budget offers from different vendors and technology options. In gasification the primary option is HT Winkler with a possible interest in alternatively pursuing new reforming approaches. As for the synthesis loop, conventional catalytic methanol synthesis should be contrasted to the new liquid phase synthesis process as is currently being demonstrated in the Tennessee Eastman Chemicals plant in the US.

A demonstration plant could consist of one of the three gasification trains foreseen for the commercial plant studied in this report. Utilisation of the moth-balled Finnish HTW-gasifier might be another alternative.

For the route to ethanol via enzymatic hydrolysis and simultaneous saccharification and fermentation, including pentose fermentation, a continuously operating pilot plant in Europe seems desirable, before a commercial demonstration can be considered.

5.3.3 Product Slate

Very likely, process steam/electricity will be interesting co-products to enhance overall economics. The final selection of the primary transport fuel product is a

fundamental decision which requires substantial documentation and consultations with government and industry policy and technical bodies in the European Union as well as in the member states. In this context the selection of methanol and various forms of DME (Dimethylether) formulations is a work task.

Blending and stand-alone applications of the new fuel, environmental performance will be important considerations in this respect. Methanol will be considered as the main product line which also will meet the targets for liquid fuels for fuel cells.

5.3.4 Partnership

Based on the assumption that a 50 % public sector investment grant for the Commercial Demonstration Phase may be available, the establishment of a commercial partnership to build, own, and operate a plant for biomass-based transportation fuels production is deemed possible. Ideally, process, automotive industry, raw materials, and site related expertise should be included in such a partnership.

Ecotrafic has during 1996 and 1997 made studies including production of alcohol's for Vattenfall AB, County Administrative Boards of Jämtland and of Västernorrland and for Trollhättan Energi AB who have shown interest to go on by further studies and could be partners in the future. Also Nykomb Synergetics has indications of interest from their assignors Växjö Energi AB and Air Products.

5.3.5 Deliverables from the planning effort for the Commercial Demonstration Phase

The planning effort for the next phase should be documented in a core document where the issues identified in the above synopsis discussion are detailed and rationales for normative decisions are documented. Upon completion of this core document an independent sequel in terms of an investors' prospectus shall be produced together with a short-list of candidate partners/partnerships, sites, and vendors. Based on solicited responses from the short listed actors to the prospectus, decisions on grants and the formation of final partnerships may be taken.

To further raise the interest from the partners, it is of great importance that the European Commission will indicate interest in continuing to support proceeding studies to be able to reach the level of knowledge, where demonstration commercial plant investors can be attracted.

These future actions are also in line with earlier presentation to the European Commission of the continuation of this "BAL Fuels" study.