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The MethaHydro-process - preliminary design and cost evaluation

TNO Institute of
Environmental Sciences,
Energy Research and
Process Innovation

Laan van Westenenk 501
P.O. Box 342
7300 AH Apeldoorn
The Netherlands

Phone +31 55 - 549 34 93
Fax +31 55 - 541 98 37

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Author(s)
Hans Oonk (TNO-MEP)
Jasper Vis (UU-NW&S)
Ernst Worrell (UU-NW&S)
André Faaij (UU-NW&S)
Jan-Willem Bode (UU-NW&S)

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Intended for
Ministry of VROM/DGM/LE
Attn. Drs. R.A. v.d. Wijngaart
P.O. Box 30945
2500 GX The Hague

Samenvatting

Deze studie betreft het ontwerp en de evaluatie van een 'MethaHydro'-proces, waarmee uitgaande van biomassa en aardgas simultaan verschillende energiedragers kunnen worden geproduceerd, bijvoorbeeld: methanol, elektriciteit en waterstof voor injectie in het aardgasnet (Methane Enriched Hydrogen - MEH). De fossiele koolstof, welke wordt geïntroduceerd in het systeem in de vorm van aardgas, wordt afgescheiden en ondergronds opgeslagen. Daardoor zijn de geproduceerde energiedragers energie-neutraal. Het MethaHydro-proces bestaat uit bekende en bewezen deelprocessen voor het produceren, reinigen, bewerken en verwerken van synthesegas.

Massa- en energiebalansen van het MethaHydro-proces zijn opgesteld voor verschillende product-verhoudingen methanol, elektriciteit en MEH. De energetische input van het ontworpen MethaHydro-proces is ongeveer 730 MW. De energetische efficiency van de omzetting naar energiedragers is 60-70%, afhankelijk van de geproduceerde mix. De benutting van de energiedragers uit het MethaHydro-proces resulteert in een reductie van CO_2-emissies. Bij deze schaalgrootte is dit ongeveer 1 Mton CO_2 per jaar.

Technische haalbaarheid
Het MethaHydro-proces bestaat geheel uit deelprocessen, die reeds op praktijkschaal bewezen zijn. Om deze reden is het MethaHydro-proces op dit moment technologisch haalbaar. Voor specifieke onderdelen van het MethaHydro-proces zijn verdere verbeteringen mogelijk. Voorbeelden hiervan zijn de biomassavergasser, de autotherme reformer, de integratie van de vergasser en de autotherme reformer, CO_2-afscheiding en methanolsynthese onder de specifieke MethaHydro-condities. Omdat een deel van deze ontwikkelingen autonoom zullen verlopen (onafhankelijk van specifieke MethaHydro-ontwikkelingen) kunnen in de toekomst nog verbeteringen van het MethaHydro-proces worden verwacht.

Kosten
De kosten voor de energiedragers uit het MethaHydro-proces hangen af van de methode waarop de kapitaalskosten voor produktie en conditionering van het synthese gas worden verdeeld. Als deze worden toegedeeld op basis van de energie-inhoud van de produkten, dan staan in tabel S.1 de produktiekosten weergegeven van een MethaHydro-plant, waarvan de energetische output voor 50% uit methanol, voor 25% uit elektriciteit en voor 25% uit MEH bestaat. De tabel geeft ook de kosten van de normale brandstoffen, welke kunnen worden vervangen door de MethaHydro-produkten. De kosten van MethaHydro methanol zijn 40 tot 120 % hoger dan de prijzen van benzine (omgerekend naar prijs per kilometer). Als de accijns op methanol even hoog wordt als de accijns op benzine, dan zal dit resulteren in een prijsstijging van 17 tot 37 cent op een totale prijs van ongeveer 2 gulden. De produktiekosten van MethaHydro elektriciteit zijn 50 tot 150% hoger dan de gemiddelde prijs voor elektriciteitsproductie. Methane Enriched Hydrogen

wordt 2,5 tot 3,5 keer zo duur als aardgas. Vergeleken met de prijs van traditionele methanolproductie, is MethaHydro methanol 10 tot 50% duurder, afhankelijk van de geproduceerde brandstof-mix en de biomassaprijs.
Wanneer de allocatie van de kapitaalskosten op een andere manier gebeurt, veranderen de kosten in tabel S.1 enigszins.

Table S.1 *Produktiekosten voor een 50% MeOH MethaHydro-proces en standaard brandstofkosten (cursief) in het jaar 2020.*

	Scenario	
	Global Competition	European Co-ordination
MethaHydro methanol (*f*/l ge)	0,63 - 0,72	0,58 - 0,67
Benzine (f/l)	*0,46*	*0,30*
MethaHydro elektriciteit (*f*/kWh)	0,16 - 0,18	0,14 - 0,18
Elektriciteit (f/kWh)	*0,11*	*0,07*
MethaHydro MEH (*f*/m^3 nge)	0,60 - 0,68	0,54 - 0,63
Aardgas (*f*/m^3)	0,25	0,19

De kostenberekeningen zijn gebaseerd op een prijs voor biomassa van *f* 4-12/GJ

ge: benzine equivalent (1 liter methanol is 0.5 liter benzine, onder aanname de de efficiency van een verbrandingsmotor gelijk is voor methanol als voor benzine)

MEH: Methane Enriched Hydrogen, een additief voor aardgas

nge: aardgas equivalent (1 kJ_{th} komt overeen met 31 m^3 aardgas)

De kosten voor methanolproductie met het MethaHydro-proces zijn aanzienlijk lager dan de kosten voor methanol, uitgaande van alleen biomassa (*f* 0,89-1,40/l ge). Ook de waterstof uit het MethaHydro-proces is goedkoper dan waterstof uit biomassa (*f* 25-35/GJ). Echter, waterstofproductie uit aardgas, waarbij CO_2 wordt afgescheiden en opgeslagen is met *f* 16-18/GJ de goedkoopste wijze van productie van CO_2-neutrale waterstof. Voor wat betreft CO_2-neutrale opwekking van elektriciteit is gebruik van aardgas met CO_2-verwijdering en opslag de goedkoopste optie met ongeveer *f* 0,12/kWh, vergeleken met MethaHydro elektriciteit en *f* 0,14-0,21/kWh voor elektriciteit op basis van biomassa.

De kosten voor CO_2-emissiereductie kunnen worden gedefinieerd als het verschil in kosten van de CO_2-neutrale brandstof als geproduceerd met het MethaHydro-proces en de normale brandstof, gedeeld door de hoeveelheid CO_2 die wordt vermeden door toepassing van de CO_2-neutrale brandstof. Voor de 50% methanol plant en bij het 'Global Competition' scenario resulteert dit in gemiddeld *f* 110-160 per ton vermeden CO_2. Bij het 'European Co-ordination' scenario zijn de gemiddelde kosten voor CO_2-emissiereductie *f* 140-180 per ton vermeden CO_2.

Summary

In this study a MethaHydro-process is designed, suitable for the large scale production of methanol, Methane Enriched Hydrogen (MEH, an additive to the natural gas grid) and electricity. The process uses biomass and natural gas as feedstock. The fossil carbon introduced in the form of natural gas is recovered during the process and stored as carbon dioxide. As a result the MethaHydro products are carbon dioxide neutral upon production and use. The MethaHydro-process consists of well-known unit-operations for synthesis gas generation, synthesis gas treatment and synthesis gas utilisation.

The performance of the MethaHydro process has been estimated for different product mixes. Upon utilisation of the energy carriers produced by the MethaHydro process, the use of fossil fuels is mitigated and carbon dioxide emissions are reduced. The energetic input of the MethaHydro process is about 730 MW. The overall energetic efficiency is 60-70%. Carbon dioxide emission reduction of the process at this scale is in the order of 1 Mtonne per year.

Technological feasibility
The MethaHydro process designed can be considered as technologically feasible, since all unit-operations involved are demonstrated in real-scale applications. Technology development and improvement of such a MethaHydro process is still possible for specific details of the process. Examples are the biomass gasification, the autothermal reformer, the integration of both, carbon dioxide recovery and the methanol production at MethaHydro-conditions. Since part of these developments will be achieved autonomously, significant improvements in the MethaHydro-process might be expected in future.

Costs
The costs of MethaHydro methanol as transport fuel will be 40 to 200% higher than those of gasoline. If the levy imposed on methanol would be equal to the one on gasoline, the price difference for the consumer would be ¢10-35 on a total price of approximately $1.10. The production costs of MethaHydro electricity will be 50 to 150% higher than the average for the Dutch electricity generation. Methane Enriched Hydrogen would be 3 times as expensive as natural gas.

The costs of methanol production using the MethaHydro process is significantly cheaper than production from biomass only. Converting the MEH from the MethaHydro process into pure hydrogen is cheaper than hydrogen from biomass. However, a natural gas based process with CO_2-recovery is the cheapest option to produce CO_2-lean hydrogen. Concerning CO_2-lean electricity production, a natural gas based process with CO_2-recovery is the cheapest option, where the MethaHydro process and biomass based electricity production are about equal. Depending on biomass and natural prices, the average CO_2-mitigation costs for the MethaHydro-plant are $65-105/tonne CO_2.

The MethaHydro-process - preliminary design and cost evaluation

- executive summary-

Hans Oonk[1], Jasper Vis[2], Ernst Worrell[2], André Faaij[2], Jan-Willem Bode[2]

1. Introduction

Use of fossil fuels is the main cause of carbon dioxide emissions. Fossil fuels are consumed in most sectors in society: industry, power generation, in households and in road transport. In recent years much attention is paid to the production of carbon-lean energy carriers, either from renewable or fossil resources. Examples of technologies are hydrogen produced from fossil fuels while separating and storing the carbon dioxide released (Blok *et al.*,1997); electricity produced while separating carbon dioxide produced (Hendriks, 1995) and methanol produced from biomass.

Possible even more attractive as production of the individual products is the combined production of a mix of energy carriers (e.g., methanol, hydrogen, electricity), starting from a mix of inputs (e.g., natural gas, oil residues, oil and biomass). An example of such a 'MethaHydro'-process is depicted in Figure 1.

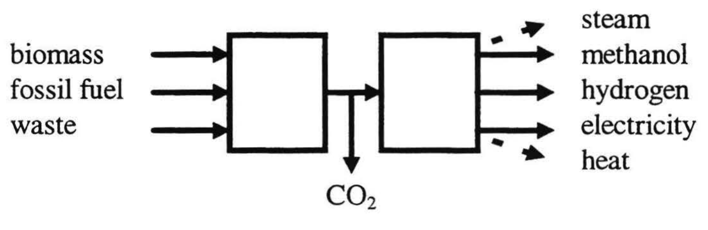

Figure 1 Schematic representation of the MethaHydro-process

Besides production of methanol, other chemical products may also be produced, for example a mixture of high alkanes, that in turn can be converted to gasoline and diesel. Besides, methanol can also be converted to an number of components as MtBE and ethene. In such a way a MethaHydro-process is also suited as a basis for a more sustainable chemical industry.

[1] TNO-MEP, P.O. Box 342, 7300 AH Apeldoorn, The Netherlands, phone +31 55 549 34 93, fax +31 55 549 3410

[2] Utrecht University, Padualaan 14, 3584 CH Utrecht, The Netherlands, phone +31 30 253 76 00, fax +31 30 253 76 01

2. Process design

The lay-out of the MethaHydro process depends on a number of factors. The most important ones are the scale of process, the mix of fossil resources used, the mix of products to be produced and the flexibility required. In this study a MethaHydro-process is designed, suitable for processing a mixture of biomass and natural gas, and producing a mixture of methanol, methane enriched hydrogen (MEH) and electricity at a large scale. The process is able to produce different ratios of these components. The resulting MethaHydro process is depicted in figure 2.

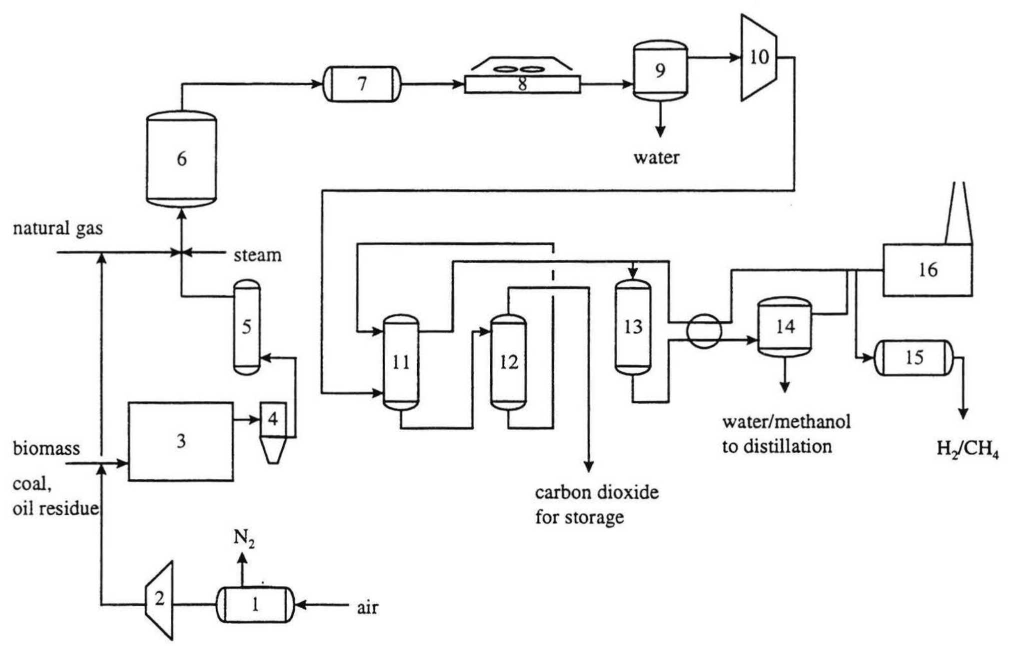

Figure 2 The MethaHydro-process

Oxygen is produced and compressed in a cryogenic oxygen plant ((1) and (2). Biomass is dried and sized to the requirements of the IGT-gasifier (3). The resulting product gas consists notably of the synthesis gas products CO, CO_2 and H_2, but contains also some methane and higher hydrocarbons. Impurities in the product gas (particles and alkali metals) might be separated simultaneously in a wet scrubber (5), possibly preceded by a cyclone (4) to separate the larger particles. The clean product gas is mixed with natural gas, steam and oxygen and fed to an autothermal reformer (6). This autothermal reformer is a combination of a steam-reformer (converting CH_4 with steam in an endothermic reaction) and a partial oxidation (converting CH_4 with oxygen in an exothermic reaction). The product of the autothermal reformer is a near equilibrium synthesis gas, containing large amounts of CO and H_2, smaller amounts of CO_2 and unconverted CH_4. Besides that, H_2O is present in the gas as a result of excess steam introduced in the autothermal reformer. The product gas is cooled, while exchanging heat with the feed of the autothermal reactor, and subjected to a water-gas shift (7), where CO is converted to CO_2. After being shifted, the product gas is cooled further and water is subsequently separated in a knock-out drum (9). The product gas is compressed to 80 bar (10) after which part of the carbon dioxide is separated using Selexol (11). At regeneration (12) carbon dioxide is released and compressed before transport and underground storage. Since about 5 vol% carbon dioxide is allowed in the feed-gas for the methanol synthesis, carbon dioxide partial pressure in the process gas after carbon dioxide separation is rather high (4 bar[3]). The combination of this high partial pressure of carbon dioxide and the high pressure of the process gas, makes carbon dioxide absorption and subsequent regeneration of the process fluid cheap compared to other systems for carbon dioxide recovery. After carbon dioxide recovery, the synthesis gas is fed to the methanol synthesis (13 and 14). The products of the methanol synthesis are a water-methanol mixture for further product make up, and a purge gas stream, consisting notably of hydrogen, but also containing some unconverted methane from the autothermal reformer and unconverted CO and CO_2 from the methanol synthesis. This purge gas stream can be used as a fuel gas for electricity generation (16). Besides that the purge may be injected in the natural gas distribution grid. In the latter case, the carbon oxides are converted to methane in a methanation reactor (15).

This MethaHydro-process consists of well-known unit-operations for synthesis gas generation, synthesis gas treatment and synthesis gas utilisation. Carbon dioxide recovery and storage is an integral part of the MethaHydro-process. When the amount of carbon fed into the process as natural gas equals the amount of carbon recovered and stored as carbon dioxide, the mix of energy carriers produced can be regarded as carbon dioxide neutral.

The MethaHydro-process described above is flexible. Three degrees of freedom exist: (i) the ratio of biomass and natural gas as input in the system; (ii) the amount of CO_2- recovered and (iii) the conversion of carbon oxides to methanol. As a result, different ratios of methanol, electricity and MEH might be produced in a carbon dioxide neutral way. Table 1 gives some examples of product mixtures.

3. Material and energy balances, overall efficiency

The material and energy balances for the MethaHydro-process are calculated for 4 situations. In the first situation the conversion to methanol is maximised. In the second situation about 75% of the energy output is as methanol. In the third case this is about 50% and in case 4 about 33% of the energy is produced as methanol. The feed and product characteristics are summarised in table 1.

[3] For comparison, when recovering 90% of CO_2 from an atmospheric flue gas, the carbon dioxide partial pressure after separation is 0,015 bar. When recovering 90% CO_2 from a 10 bar fuel gas, the carbon dioxide partial pressure is 0.3 bar.

Table 1 *Results of the material and energy balance calculations.*

	90% methanol	70% methanol	50% methanol	30% methanol
feed				
biomass (dried, tonne day^{-1})	1790	1340	980	680
natural gas (tonne day^{-1})	640	790	920	1020
products				
methanol (tonne day^{-1})	1800	1370	920	530
(automotive fuel 1000 l ge y^{-1})[2]	*(1130)*	*(870)*	*(580)*	*(330)*
electricity (MW_e)	23	68	106	144
CH_4/H_2 (tonne day^{-1})	18/8	23/38	25/64	30/93
(as MEH, 1000 m^3 nge day^{-1})[3]	*(63)*	*(180)*	*(290)*	*(390)*
energy in (MW)	732	732	732	732
energy out (MW)	462	442	424	412
overall efficiency (%)	63	60	58	56
CO_2 emission reduction (Mtonne y^{-1})[4]	1,06	1,11	1,14	1,18

1: dried biomass contains 15% water

2: ge: gasoline equivalent (1 litre MeOH equals 0.5 litre gasoline)

3: MEH: Methane Enriched Hydrogen as an additive to the natural gas grid; nge: natural gas equivalent (1 kJ_{th} equals 31 m^3 of natural gas)

4: CO_2 emission reduction is calculated, assuming 1,4 kg CO_2 emission reduction per kg methanol produced; 0,15 kg CO_2 MJ_e^{-1} and 0,05 kg CO_2 per MJ_{th} Methane Enriched Hydrogen.

4. The economics of the MethaHydro plant

Capital costs

The capital costs of the MethaHydro plant are calculated separately for all the main parts of the installation. These cost calculations are based on manufacturers' data or on one or more reference plants from literature and are scaled according to an appropriate quantity and scale factor for that specific part. In the case manufacturers' data for separate items are used, the capital costs are increased with 25 percent for utilities and auxiliaries (Williams *et al.* 1995). In the case of reference plants from literature, utilities and auxiliaries are assumed to be included in the total capital costs. Table 2 gives an overview of the capital costs for the MethaHydro plant for each of the product mixes, based on the estimations described above. For the flexible plant, each part is priced at the highest costs for the other product mixes. For reasons of comparison, the table also shows the capital costs of a methanol plant which converts natural gas in methanol. This reference plant is assumed to be equipped with an autothermal reformer.

Table 2 *Capital costs of MethaHydro plant (10^6 US\$).*

Installed hardware	Product mix (%MeOH of energy output) 90%MeOH	70%MeOH	50%MeOH	30%MeOH	flexible	'Normal' MeOH plant
Oxygen plant	44	46	46	48	48	53
Biomass gasifier	115	94	75	58	115	
Autothermal reformer	23	26	28	29	29	35
Shift reactor	34	35	35	36	36	
Syngas compressor	47	50	51	54	54	53
Carbon dioxide removal	14	16	18	19	19	
Methanol reactor	30	30	30	30	30	33
Methanation reactor	1	2	4	5	5	
Combined cycle	18	54	85	123	123	
Utilities/auxiliaries	29	25	26	27	27	43
Total costs installed hardware	306	293	279	272	329	217

Operating and maintenance costs

The operating costs of the MethaHydro plant strongly depend upon the price of natural gas and biomass. For natural gas, the shadow prices from the 'Global Competition' and the 'European Co-ordination' scenarios for the year 2020 of \$0,14/m^3 and \$0,11/m^3, respectively are used (CPB, 1996). The future price of imported biomass is highly uncertain and will fluctuate depending on the time of the year and the total demand for biomass. Therefore, a price range of \$2,3-6,8/GJ has been used for the biomass price. Table 3 shows the annual operating and maintenance (O&M) costs of the different MethaHydro plants. The O&M costs for the flexible plant depend on the actual input and output.

Table 3 *Operating & maintenance costs (10^6 US\$/year) of the MethaHydro plant in the year 2020 under the Global Competition scenario (in brackets the costs which are different under the European Co-ordination scenario).*

	Product mix (%MeOH of energy output) 90%MeOH	70%MeOH	50%MeOH	30%MeOH	'Normal' MeOH plant
Variable costs					
Biomass	25-74	18-55	13-40	9-28	
Natural gas	50 (40)	62(50)	72 (57)	80 (64)	102 (80)
CO_2 storage	1	1	1	1	
Catalysts and chemicals	3	3	3	3	2
Fixed costs					
Labour	2	1	1	1	1
Maintenance	10	10	9	9	7
Overhead	8	8	7	7	5
Total operating costs	98-147 (88-137)	103-140 (92-128)	107-134 (92-119)	110-129 (94-113)	116 (94)

Total costs

Table 4 shows the total unit costs of the MethaHydro plant in 2020 under the 'Global Competition' scenario. Hardware is depreciated over 25 years at a 15 percent capital charge rate. The costs for the production of the synthesis gas are allocated over the three products according to their energetic contents at the split-off point. For reasons of comparison, the last column of Table 4 shows the costs of a normal methanol plant with an autothermal reformer, calculated with the same assumptions.

Table 4 *Production costs of MethaHydro energy carriers in the year 2020 under the Global Competition scenario (all costs in \$/GJ).*

	Product mix (%MeOH of energy output)												ref.
	90% MeOH			70% MeOH			50% MeOH			30% MeOH			MeOH plant
	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH
Capital													
Share of syngas production	5,1	5,1	5,1	4,6	4,6	4,6	4,2	4,2	4,2	3,9	3,9	3,9	2.9
Methanol production	0,4	0,0	0,0	0,5	0,0	0,0	0,7	0,0	0,0	1,2	0,0	0,0	0.3
MEH production	0,0	0,0	0,2	0,0	0,0	0,2	0,0	0,0	0,1	0,0	0,0	0,2	
Electricity production	0,0	2,0	0,0	0,0	2,0	0,0	0,0	2,0	0,0	0,0	2,0	0,0	
Labour & maintenance	1,3	1,3	1,3	1,2	1,2	1,2	1,1	1,1	1,1	1,0	1,0	1,0	0.8
Biomass[1]	2-5	2-5	2-5	1-3	1-3	1-3	1-2	1-2	1-2	1-2	1-2	1-2	
Gas	3,3	3,3	3,3	3,8	3,8	3,8	4,2	4,2	4,2	4,4	4,4	4,4	6.0
CO_2-storage	0,0	0,0	0,0	0,0	0,0	0,0	0,1	0,1	0,1	0,1	0,1	0,1	
Catalysts/chemicals	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0.1
Total Production Costs	12-15	14-17	12-15	11-14	13-15	11-13	11-13	13-14	11-12	11-12	12-13	10-11	10.1
Total production costs under the EC[2] scenario	11-15	13-16	11-14	11-13	12-15	10-13	10-12	12-13	10-11	10-11	11-12	9-10	9.2

[1] The costs for biomass are based on a price range of \$2.3-6.8/GJ

[2] European Co-cordination

In determining the costs of the MethaHydro fuels, the joint costs of production (notably the synthesis gas production and conditioning) have to be allocated. In this study two methods are used: allocation according to the energy content, and allocation according to the market prices of the products. The costs of the cheapest alternative CO_2-neutral fuel have been used as an approximation for the market price.

Table 4.2 shows the production costs in the year 2020 of the fuels from a MethaHydro plant which delivers 50% of its output in the form of methanol. The table also shows the costs of the regular fuels that would be replaced. The costs of MethaHydro methanol as transport fuel will be 40 to 200% higher than those of gasoline. If the levy imposed on methanol would be equal to the levy on gasoline, the price difference for the consumer would be ¢10-35 (depending on allocation rules for joint costs) on a total price of approximately \$1.10. A lower levy on methanol could make its end-use price equal or lower than the price of gasoline. The production costs of MethaHydro electricity will be 50 to 150% higher than the average for the Dutch electricity generation. Methane Enriched Hydrogen will be 3 times as expensive as natural gas.

Table 5 *Production costs for the fuels of a 50% MeOH MethaHydro plant and regular fuels (in italics) in the year 2020.*

	Scenario	
	Global Competition	**European Co-cordination**
MethaHydro Methanol ($/l ge)		
-allocation on energetic content	0.36-0.41	0.33-0.38
-allocation on value	0.48-0.55	0.44-0.51
Gasoline ($/l)	*0.26*	*0.17*
MethaHydro electricity ($/kWh)		
-allocation on energetic content	0.09-0.10	0.08-0.10
-allocation on value	0.07-0.08	0.07-0.08
Electricity ($/kWh)	*0.06*	*0.04*
MethaHydro MEH ($/m^3 nge)		
-allocation on energetic content	0.34-0.39	0.31-0.36
-allocation on value	0.24-0.28	0.23-0.26
Natural gas ($/m^3)	*0.14*	*0.11*

The cost calculations are based on a biomass price of $2.3-6.8/GJ

ge: gasoline equivalent (1 litre MeOH equals 0.5 litre gasoline)

MEH: Methane Enriched Hydrogen as an additive to the natural gas grid

nge: natural gas equivalent (1 kJ_{th} equals 31 m^3 of natural gas)

The costs of methanol production using the MethaHydro are significantly cheaper than production from biomass only, which costs $0.51-0.80/l ge (adapted from Williams *et al*, 1995). When the joint costs are allocated according to the value of the products, converting the MEH from the MethaHydro process into pure hydrogen is cheaper than hydrogen from biomass ($14-20/GJ, adapted form Williams *et al*, 1995), and is comparable to hydrogen from a natural gas based process with CO_2-recovery. Process with CO_2-recovery is the cheapest option ($9-10/GJ, Blok *et al*, 1997). Concerning electricity production, a natural gas based process with CO_2-recovery (¢7/kWh Hendriks *et al*, 1992) about as costly as electricity from the MethaHydro process, provided that the joint costs are allocated according to the value of the products. Biomass based electricity production is the most expensive option (¢8-12), adapted from Solantausta *et al*, 1996. Compared to traditional methanol production, CO_2-neutral methanol from the MethaHydro-plant is about 10-50% more expensive, depending on configuration chosen and the biomass price.

Cost of carbon dioxide mitigation

The costs of CO_2-mitigation may be defined as the cost difference of the CO_2-neutral fuel as produced by the MethaHydro plant and the normal fuel, divided by the specific CO_2-emission upon use of the normal fuel. For the 50%-MeOH plant under the Global Competition scenario this results in $45-135 per tonne mitigated CO_2 for methanol used as transportation fuel, $35-80 per tonne CO_2 for the electricity and $65-150 per tonne CO_2 for the MEH supplied to the national gas distribution grid. The average CO_2-mitigation costs for this plant are $65-90/tonne. Under the European Co-ordination scenario, the average CO_2-mitigation costs are

\$80-105/tonne: \$80-160/tonne for methanol, \$45-90 for electricity and \$70-150 for MEH.

5. Conclusions

The MethaHydro-process
The lay-out of a MethaHydro-process depends on a number of factors. The most important factors are: the capacity of the process, the mix of fossil fuels used, the mix of energy carriers produced and the flexibility in production. In this study a MethaHydro-process is designed, suitable for processing both of biomass and natural gas, and producing a flexible mixture of methanol, methane enriched hydrogen and electricity at a large scale. The overall energetic efficiency of the MethaHydro-process is about 55-65%.

Technological feasibility
The MethaHydro process as described in this study can be considered as technologically feasible, since all unit-operations involved are demonstrated in real-scale applications. The technologies however are never combined in the way presented here, so a demonstration of technology is recommendable. Development and improvement is required for:
- the technology for large-scale biomass gasification;
- the integration of gasification and autothermal reforming, this implies either adapting biomass gasification in such a way that the process gas is optimal for autothermal reforming or adapting autothermal reformer in such a way that the process gas from the gasifier can easily be handled without further purification;
- the carbon dioxide separation and subsequent regeneration of the absorber has to be demonstrated at process conditions and may be subject to further improvements;
- the development of methanol-synthesis suitable for operating at high stoichiometric ratios;
- development of technology for H_2 -separation from the product gas in order to adjust the stoichiometric ratio before methanol synthesis;
- the dynamics of the process, since a flexibility for hydrogen and electricity production on a few hours scale might be an advantage of a MethaHydro process.

Costs
The production costs of CO_2-neutral methanol are about 40-200% more expensive than regular fuels. MEH will be about even be 3 times as expensive as natural gas. Compared to traditional methanol production, MethaHydro methanol is only slightly more expensive. When evaluated on its costs for carbon dioxide emission reduction, the MethaHydro-process is a relative cheap way for reducing carbon dioxide emissions, compared to other biomass based options. H_2 or electricity production from fossil fuels while separating and storing carbon dioxide proves to be more cheap than the MethaHydro-process.

6. References

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Table of contents

1. Introduction

General
Use of fossil fuels is the main cause of carbon dioxide emissions. Fossil fuels are consumed in most sectors in society: industry, power generation, in households and in road transport. An option for reduction of carbon dioxide is the use of energy carriers that produce less CO_2. Several examples are:
- electricity with less carbon dioxide emissions may be produced in various ways. This can be done by using renewable energy sources, such as biomass, wind or solar-energy. Although costs of renewable sources are still high, a substantial reduction in costs is already achieved, and it can be expected that further reduction will occur in future. Another option is to separate the carbon dioxide from a fossil fuel based process. (Hendriks, 1995). The separated carbon dioxide can subsequently be stored in empty gas-fields or in aquifers.
- natural gas can partially be replaced by hydrogen. This hydrogen can be produced either from a renewable source or starting from natural gas, while the carbon dioxide produced is separated and stored (Blok *et al.*,1996).

Road transport is an important consumer of fossil fuels. In the Netherlands about 16% of carbon dioxide emissions are produced by this sector (van Amstel *et al.* 1994). In Table 1.1, an overview is presented of energy use, carbon dioxide emissions and fuel consumption in Dutch transport.

Table 1.1. Energy use, carbon dioxide emissions and fuel consumption in Dutch transport (van Amstel et al., 1994).

Dutch traffic (1994)	
- total energy use	384 PJ
- carbon dioxide emissions	25.9 Mtonne
- gasoline consumption	3788 million kg
- diesel consumption	4139 million kg
- LPG-consumption	800 million kg

For reduction of environmental effects of road transport (emissions of VOC and NO_x), much attention is given to the use of alternative fuels, such as methanol and hydrogen. Methanol (as methanol or as a methanol-derivative, e.g. MTBE) may be introduced on a short term, without significant adaptations of cars and infrastructure. Introduction of hydrogen takes much more effort and is expected not to be feasible for large-scale use in the transport sector on short or medium term. It might however be used in specific applications such as public transport (e.g. buses). Both fuels are especially of interest, when fuel-cells are used to convert chemical energy into electricity and subsequently into motion.

The MethaHydro process
Both methanol and hydrogen are nowadays produced from fossil fuels, such as natural gas or coal. During production or upon use, the fossil carbon is released as

carbon dioxide. In order to produce fuels with low or zero carbon dioxide emissions, new processes have to be developed and implemented, which produce these alternative fuels with no or negligible carbon dioxide emissions. Hydrogen might be produced by electrolysis of water, using electricity from renewable sources, but this option is still very costly, even in the long term (Williams *et al.*,1995). Methanol or hydrogen production from waste or biomass is an alternative. A possibly less expensive option is the combined production of a mix of energy carriers (e.g., methanol, hydrogen, electricity and their co-products steam and heat), starting from a mix of inputs (e.g., natural gas, oil residues, oil and biomass). An example of this, the MethaHydro-process, is depicted in Figure 1.1.

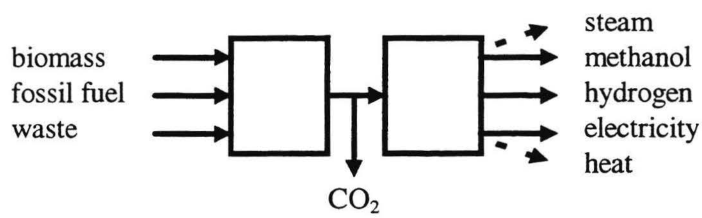

Figure 1.1 In- and output of the MethaHydro-process

Besides methanol, other hydrocarbons may be produced as well, for example a mixture of high alkanes, that in turn can be converted to gasoline and diesel. In this way the difficulties of introducing a new fuel in the transport sector could be avoided.

The MethaHydro-process is CO_2-neutral, when the carbon input as biomass equals the sum of the carbon output as methanol and other hydrocarbons and the carbon-output in rest-emissions. In such a case the amount of carbon that enters the system in fossil fuels, equals the amount of carbon that is separated and stored as carbon dioxide.

Fuels from the MethaHydro-system may be competitive (i) when produced on a large scale, (ii) when the system is flexible with regards to types of biomass, waste, fuels and residues used (thus having the opportunity to use the least expensive combination), (iii) when opportunities exist for large-scale import of biomass, and (iv) when costs for CO_2 separation and storage are low.

The objective of this study is to make a preliminary design of a MethaHydro process and subsequently evaluate the technological and economic feasibility of such a process.

The results of this study are presented in this report. In Chapter 2 describes the selection of the scale of the process, the starting materials used, the mixture of energy carriers used and the flexibility required. Chapter 2 also contains material and energy balances of the selected processes. In chapter 3, the economics of the processes are assessed and evaluated. Conclusions are drawn in chapter 4.

2. Selection and design of the MethaHydro process

2.1 Selection

A MethaHydro-process, as defined in Chapter 1, may be compiled in a number of ways. Alternatives differ with respect to:
- the capacity of the process;
- the mix of fossil resources used and the flexibility of this mix;
- the technology for synthesis gas generation;
- the method for utilisation of the synthesis gas, the mix of products and its flexibility;
- the lay-out of required water-gas shift and carbon dioxide removal and synthesis gas utilisation.

For all aspects choices are made, ultimately resulting in one choice for one MethaHydro process. It has to be noted that this MethaHydro-process is not the most beneficial process in all cases. For example when a process is designed on a substantially smaller scale, when other fossil fuels are used or when another product mix is produced, another MethaHydro-process might be preferred.

Capacity
The capacity of a chemical process is determined by a number of aspects:

1) availability of resources
For a MethaHydro process in The Netherlands, biomass will be the most limited resource. Three sources of biomass could be distinguished in the Dutch context:
- biomass wastes and residues. The available potential of biomass wastes and residues in the Netherlands amounts approximately 100 PJ, but this includes streams like organic domestic waste and sludge from waste water treatment, which necessitates additional gas treatment after gasification. Costs range between -10 to +5 $/GJ (the higher values are found for straw and to a lesser extent thinnings) (TEB, 1995; Faaij 1997).
- cultivated biomass (energy crops). Biomass is currently not cultivated on any significant scale in the Netherlands. The potential depends on the available land. Current developments in agriculture such as phasing out of subsidies may lead to continuing supply of agricultural land. In 2015 this could lead to a maximum of almost 400,000 ha available for energy crops with an energetic potential of maximally 90 PJ, when utilised for growing Miscanthus or Short Rotation Forestry. However, depending on agricultural policy and lack of support for energy crops this potential can be far lower. The costs of cultivated biomass in the Netherlands amount approximately 4-8 $/GJ, depending on yield, costs of land and farmers incomes (TEB, 1995; Faaij 1997).
- imported biomass. Import of biomass could theoretically make a very large contribution of biomass to the Dutch energy system. Depending on the country

of origin the costs of imported biomass are estimated to be 2.3-6.8 $/GJ in the current situation. Imported biomass will most likely be wood, which gives a relatively clean product gas (TEB, 1995; Faaij 1997).

If biomass is supplied from within the Netherlands, it is advisable to limit the scale of the MethaHydro plant to about 200 tonne methanol per day. To supply a 2000 tonne per day plant (see below), large scale inland transport of many different small biomass streams would be necessary, with accompanying costs.

It should be realised that the price of imported biomass will fluctuate depending on the time of year and depending on the total demand. On the Swedish wood market, which is well developed, considerable increase in wood prices are observed in the heating season, which might make import of biomass unattractive during parts of the year. Furthermore, a high demand for wood in certain regions such as the Baltics (both for export as for the indigenous market) might lead to considerable price-rises. Such mechanisms are also observed in the Netherlands itself where new biomass conversion capacity claims parts of the, limited, potential and more capacity is planned, leading to price-rises for e.g. thinnings and waste wood.

2) market-size for products
For the MethaHydro-process it is assumed, that methanol demand is high, because of its use as a fuel in traffic and that injection of H_2 in natural gas distribution grids is generally accepted. So the market-size for the products is not considered as a limiting factor for the plant capacity.

3) economy of scale
The standard gas phase methanol process, coal gasification and methane reforming have large economies of scale, so a large scale process has relatively low capital costs. This forms a rationale to design the MethaHydro process at the scale of the largest methanol plants operated to-date: about 2000 tpd methanol or its energetic equivalent in Methane Enriched Hydrogen or electricity[4].

At a large scale, oxygen production is relatively cheap. This might favour the use of oxygen-based (more efficient) processes for synthesis gas preparation, such as auto-thermal reforming and directly heated biomass gasification (see Section 2.4). In this way direct emissions of carbon dioxide in the production of synthesis gas are avoided.

With regard to biomass, the choice for a large scale plant implies the use of imported biomass. The amount of biomass available in The Netherlands (either cultivated or from waste and residues) does not suffice for a large scale MethaHydro plant. Imported biomass will most likely be wood, which has the advantage of a relatively clean product gas.

4 The use of 2000 tpd carbon dioxide neutral methanol corresponds with a CO_2-emission reduction of 1 Mtonne per year.

Fossil resources

A number of fossil fuels might be used as hydrogen supplier in the MethaHydro process. The options considered in this study are:

- natural gas, at this moment most frequently used as feedstock in methanol and hydrogen production. The advantage of natural gas is the large amount of carbon dioxide produced (per mole of carbon dioxide about 3.5 moles of hydrogen). A disadvantage of natural gas is its limited reserves. Currently, the world natural gas reserves are estimated to suffice for about 65 years at present consumption rate. The price of natural gas is higher than the price of coal.
- heavy oil residues might be an alternative. Per mole of carbon dioxide, about 2.5 mole of hydrogen is produced. When heavy oil residues are used in the MethaHydro process, more carbon dioxide has to be separated and stored compared to the use of natural gas. Since oil residues are heavily contaminated with a.o. sulphur, and heavy metals, use of oil residues requires an expensive synthesis gas cleaning. According to Gary and Handwerk (1994) the price of oil residues has historically been circa 70% of the crude oil price. Prices are expected to drop in the near future, because (i) the sulphur content of crude oil is slowly increasing, (ii) the market for high sulphur fuel oil is steadily reducing and (iii) environmental regulations are becoming more stringent (Gulli 1995, Emsperger and Karg 1996).
- coal produces 1.4 mole of hydrogen per mole of carbon dioxide. The advantages of coal are its low price and its abundance. At present consumption rates, expected reserves suffice for over 200 years. The disadvantage of coal, when used in the MethaHydro-process, is the large amount of carbon dioxide that has to be separated and stored.
- combinations of the resources above.

In this study natural gas is chosen as fossil resource. In a first estimate, the operating costs for use of methane in the MethaHydro-process are slightly less than the those for the use of coal and oil residues. The high costs of methane are compensated by its high efficiency in hydrogen generation and the relative small amount of carbon dioxide that has to be recovered.

Synthesis gas generation

Different options exist to convert the feedstocks of the MethaHydro process into synthesis gas. For natural gas the relevant processes are: steam-reforming, (the process most used in commercial methanol production); partial oxidation and autothermal reforming, a relatively new process (for a more detailed description of this technology, see Appendix 1). The latter two processes use oxygen, while steam-reforming is air-based. The choice of air-based versus oxygen-based processes depends the availability of oxygen. For a normal methanol synthesis, when no cheap oxygen is available, oxygen-based processes become favourable at production capacities of about 1500 tonne methanol per day (Westerterp, 1990).

Gasifiers used to convert biomass in synthesis gas are classified according to the operating pressure, the heat supply mechanism and the bed type. Most interesting options for the MethaHydro process are the pressurised directly heated fluid-bed gasifier (e.g. IGT) and the atmospheric indirectly heated fluid-bed gasifier (e.g. BCL) as described in Appendix 1. The former uses oxygen and has a high efficiency. The latter uses air and has a lower efficiency, but might be an interesting option if waste is used as feedstock, because it fits well with atmospheric low temperature gas cleaning.

Since the MethaHydro-process in this study is designed at a large-scale, oxygen based processes are preferred. An additional advantage of oxygen based processes is that they do not emit carbon dioxide, so no additional biomass-carbon is required to compensate for such a carbon leak. Autothermal reforming is used for conversion of the natural gas to synthesis gas; a pressurised directly heated gasifier (e.g. IGT) will be used for biomass conversion, because of its high efficiency and its relative clean product gas. It has to be noted that this choice of synthesis gas generation depends on scale and fossil resources used. For small-scale processes at sites where no cheap oxygen is available air-based processes may be preferred; when coal is used as a fossil resource multi-feedstock gasifiers might be considered.

Utilisation options

The synthesis gas can be used in various ways:

- Hydrogen might be produced either by shift of CO to CO_2 and subsequent carbon dioxide separation, by hydrogen removal using membranes or by pressure swing adsorption (see Appendix 1). The product hydrogen might be used in industrial processes or added to the natural gas grid, thus reducing carbon dioxide emissions from the use of natural gas. It is estimated that hydrogen can be added to the natural gas grid to a maximum of 5 percent without adaptations to the grid or appliances (CE, 1997). Another option is the use of hydrogen in niche markets like public transportation (e.g. by using fuel cells).
- The synthesis gas might be used as a fuel gas in electricity generation, e.g. in a combined cycle unit.
- Conversion of synthesis gas to methanol, higher alcohols, amines and alifates (through the Fischer-Tropsch process) are well-known processes (see appendix 1). These chemicals in turn might be converted to other organic components. In principle, the synthesis gas from the MethaHydro-process might replace all crude oil as a starting point in carbon-chemistry, thus being the basis for a more sustainable chemical industry. Methanol might be used as an automotive fuel, either directly or indirectly after conversion to MTBE, thus contributing to carbon dioxide emission reduction from traffic. The alifates produced in a Fisher-Tropsch process might also be processed to a mixture of gasoline, diesel and liquefied petrol gas (LPG).
- Combinations of the options mentioned above in a fixed or more flexible way are also possible.

The MethaHydro-process designed in this study aims at the production of automotive fuels, electricity and hydrogen for injection in the natural gas grid. The ratio in which the various energy carriers are produced depends on a.o., market demands and prices. Since demands and prices might change in future, a more flexible-output MethaHydro-process is preferred.
With respect to automotive fuels, production of methanol is preferred over the production of alifates through a Fischer-Tropsch synthesis. The reasons are the higher energetic efficiency (79% vs. 66%) and the lower capital and operating costs of the methanol synthesis.

Lay-out of the process
Synthesis gas production, water-gas shift, synthesis gas utilisation and carbon dioxide removal might be integrated in various ways. A number of options, and the selection of the most feasible one is described in Appendix 2. The ultimate choice of the lay-out depends on a number of aspects, such as the scale of the process, the mix of raw materials used, the mix of utilisation options required and the flexibility in this. In this study a large-scale, natural gas based process is preferred, producing a mix of methanol, electricity and H_2 for addition to the natural gas grid. The product mix is preferably flexible to some extent. These features of the process have lead to the lay-out described in the next section.

2.2 Description of the selected MethaHydro-process

The selected MethaHydro-process is depicted in figure 2.1. For a more detailed description of unit-operations see Appendix 1.

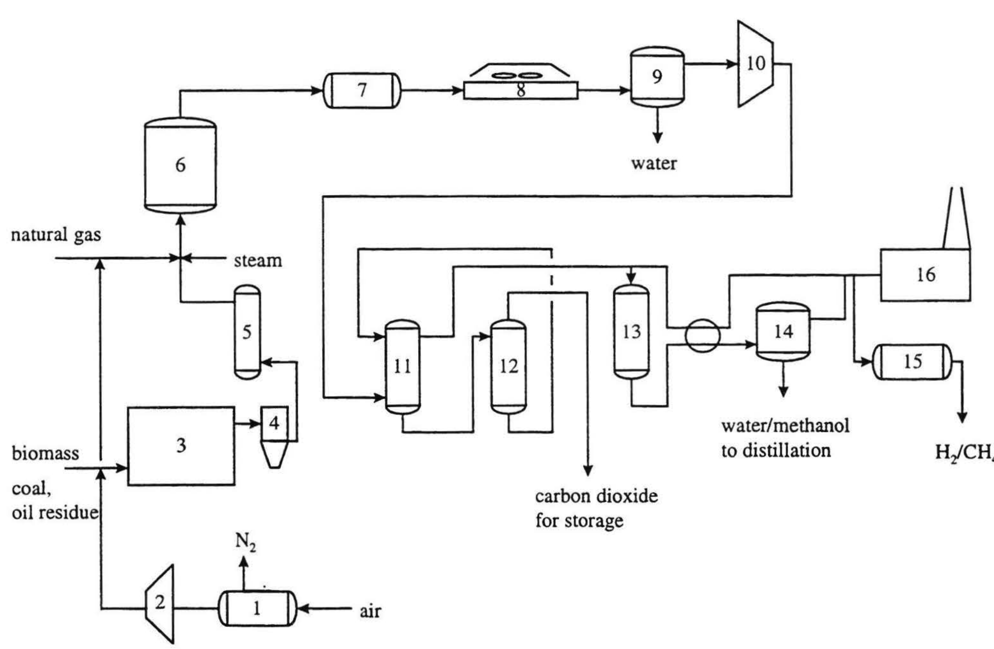

Figure 2.1 The MethaHydro-process

- *oxygen production and compression*

Oxygen (99.5% purity) is needed for the gasifier and the auto thermal reformer (ATR). It is produced (1) and compressed (2) in a cryogenic oxygen plant.

- *gasifier*

Biomass is dried in a steam dryer and sized to the requirements of an IGT-gasifier (3), which operates at a pressure of 30 bar and a temperature of 980 °C, using pure oxygen and steam as the oxidising agents. The resulting product gas consists notably of CO, CO_2 and H_2, but also contains some methane and higher hydrocarbons.

- *product-gas make-up*

Depending of the nature on the biomass, the product gas is contaminated by particles and alkali metals. For product-gas make-up several possibilities exist. When alkali metals are a problem in the subsequent autothermal reformer and water-gas shift, alkali metals and particles might be separated simultaneously in a wet scrubber (5), possibly preceded by a cyclone (4) to separate the larger particles. When alkali metals are of no concern, product gas make-up might simply constitute of a cyclone and a high-temperature dust-filters.

- *autothermal reformer*

The clean product gas is mixed with natural gas, steam and oxygen and fed to an autothermal reformer (6). An autothermal reformer combines steam-reforming(converting CH_4 with steam in an endothermic reaction) with partial oxidation (converting CH_4 with oxygen in an exothermic reaction). The reaction is performed in such a way that the heat generated by the partial oxidation provides the heat required for the steam-reforming. For this purpose a O_2 to CH_4 ratio of about 0.58 is required and a steam to CH_4 ratio of about 1.9 is used (Christensen and Primdahl, 1994). The autothermal reactor operates at 950 °C and 30 bar. The product of the autothermal reformer is a near equilibrium synthesis gas, containing large amounts of CO and H_2 and smaller amounts of CO_2 and unconverted CH_4. Besides that, H_2O is present in the gas as a result of excess steam introduced in the autothermal reformer.

- *water-gas shift*

The product gas is cooled, while exchanging heat with the feed of the autothermal reactor, and subjected to a water-gas shift (7). Here CO is converted to CO_2. The conversion of CO is determined by the equilibrium of the water-gas shift and increases with decreasing temperatures. In the MethaHydro-systems described here, shift-conversion proceeds at temperatures between 410 and 650 °C.

- *water separation*

After being shifted, the product gas is cooled further, while exchanging heat with the feed of the autothermal reformer. At 60 °C, the water in the product gas starts to condense. An air-cooler (8) is used for further cooling of the product gas and simultaneous condensation of water. The liquid water is subsequently separated in a knock-out drum (9).

- *compression*

The product gas is compressed to 80 bar in a four-stage compressor with intermediate cooling (10). This compressor is operated using steam generated in the methanol synthesis.

- *CO_2-separation*

At 80 bar, carbon dioxide is separated from the synthesis gas. The amount of carbon dioxide separated is in agreement with the amount of fossil carbon introduced in the process as natural gas. After carbon dioxide separation, the concentration of carbon dioxide in the product gas is about 5 volume percent, being the maximum carbon dioxide concentration with respect to deactivation of the methanol synthesis catalyst (Ladebeck, 1993). As a result, the carbon dioxide partial pressure after carbon dioxide separation is rather high (4 bar[5]). The combination of this high remaining partial pressure of carbon dioxide and the high pressure of the process gas, makes physical carbon dioxide absorption and subsequent regeneration of the process fluid cheap compared to other systems for carbon dioxide recovery.

In this MethaHydro-system carbon dioxide is physically absorbed in Selexol (11). The absorption fluid is subsequently regenerated upon pressure reduction to 3 bar, the pressure at which carbon dioxide is subsequently released (12). The carbon dioxide is subsequently compressed to 80 bar for transport and underground storage (Hendriks 1995).

The choice for Selexol is determined by the relatively high energy efficiency of this physical absorption process compared to other options like chemical absorption (Oudhuis, 1992). It has to be noted that existing experiences with carbon dioxide recovery are not easily translated to the MethaHydro process, and other absorber fluids and other ways of regeneration may have specific advantages over the methods proposed here. For example, if excess heat from the methanol production is available, this might be used for the regeneration of the absorber fluid. This might release carbon dioxide at higher pressures, thus reducing the costs for carbon dioxide recovery.

- *methanol synthesis*

After carbon dioxide recovery, the synthesis gas is fed to the methanol synthesis (13). As a result of the unfavourable equilibrium composition, the conversion of CO and CO_2 to methanol is incomplete (about 35% in a single pass). To overcome this problem the unconverted gas is partially recycled after separation of product methanol and water (14). The fraction of the unconverted gas that is recycled depends on the product mix chosen for the MethaHydro plant. The higher the methanol output, the higher the recylce ratio has to be.
Since the conversion of synthesis gas to methanol is exothermic, steam is produced

[5] For comparison, when recovering 90% of CO_2 from an atmospheric flue gas, the carbon dioxide partial pressure after separation has to be 0,015 bar. When recovering 90% CO_2 from a 10 bar fuel gas, the carbon dioxide partial pressure has to be 0.3 bar.

in the methanol synthesis. This steam is used for distillation of the crude methanol, oxygen compression, synthesis gas compression and the drying of biomass.

- *product make-up*

The water-methanol mixture obtained in the methanol synthesis is treated in a two-stage distillation. In the first step some of the light co-products from the methanol synthesis (alkanes from a Fischer-Tropsch-like side-reaction) are separated. In the second step methanol is separated from water and other heavy side products.

- *methanation*

The purge of the methanol synthesis consists notably of hydrogen, but also contains some unconverted methane from the autothermal reformer and unconverted CO and CO_2 from the methanol synthesis. The purge gas stream can be injected in the natural gas distribution grid as long as the hydrogen content remains below 5 percent. The introduction of CO in the gas grid might be considered a health problem. This may be solved by treating the purge gas in a methanation reactor (15). This methanation reactor is in principle a steam-reform reactor, where at low temperature (300 °C), high pressure (80 bar) and absence of water, equilibrium favours the formation of methane from CO, CO_2 and H_2.

- *electricity generation*

The purge from the methanol synthesis is also used as fuel gas for electricity generation with a combined cycle (16). The purge has a low caloric value compared to natural gas normally used as fuel in combined cycles. This might necessitate adaptations to the gas turbine. Another problem might be the control of NO_x-formation.

- *possible H_2-separation*

The stoichiometric ratio of the synthesis gas after carbon dioxide separation and before the methanol synthesis is rather high, especially in the cases where relative large amounts of MEH and electricity are produced (the 50%MeOH and the 30% MeOH-case). This has two disadvantages:
- the methanol synthesis might be inhibited, because of the relative high H_2-concentration and the relative low concentrations of carbon oxides;
- large part of the hydrogen passes the methanol synthesis without being converted. As a result capacities of the methanol synthesis and accompanying equipment (80 bar compressor, carbon dioxide separation, the decompression stage) are relative large.

This problems might be overcome, when a method becomes available to separate part of the H_2 from the synthesis gas, preferably between water separation and the 80 bar compressor. When the H_2 produced is of a sufficient quality, it might be used in industrial applications as well, resulting in high revenues compared to the use of hydrogen as MEH.
In theory two types of systems are available for H_2-separation: PSA and membranes (see appendix 1). Regarding the complex gas mixture that has to be treated

and the prerequisite of no or negligible pressure drop of the synthesis gas[6], H_2-separation is considered as not technologically demonstrated at this moment and is not incorporated in the MethaHydro-design. H_2-separation however is recognised as an option to improve the overall performance of the MethaHydro-project.

2.3 Flexibility of the MethaHydro process

The MethaHydro process as described in Section 2.2 is highly flexible. Three parameters exist, that can be used to adapt process performance and overall product composition:

- the ratio of biomass and natural gas used as a feedstock in the process.
- the conversion of CO and CO_2 to methanol. The conversion of CO and CO_2 to methanol can be adjusted with the recycle ratio of the methanol synthesis.
- the amount of carbon dioxide recovered and stored. The amount of carbon dioxide recovered can be adjusted by increasing or decreasing the amount of Selexol recycled in the absorber-system;

These three parameters give rise to two degrees of freedom:

- *the ratio of carbon-containing fuels (the sum of methanol and CH_4) and non-carbon containing fuels (H_2 and most of the electricity[7]) produced.* This ratio is connected to the ratio of natural gas and biomass used as feedstock. When this ratio increases, more CO_2 has to be recovered to keep the system CO_2-neutral. This means the amount of non-carbon containing fuels increases as well.
- *the ratio of methanol and CH_4-produced.* This factor can be adjusted by adapting the conversion of carbon oxides to methanol. When this conversion is almost complete, the MEH and the combined cycle fuel gas will contain virtually no methane. When the conversion of carbon oxides to methanol decreases, the remaining carbon oxides are transferred to methane in the methanation reactor or released as CO_2 from the stack of the combined cycle. It has to be noted, that the production of large amounts of methane is unfavourable, since this leads to an overall conversion of a mixture of methane and biomass to methane.

In the next section, the flexibility of the system is further illustrated by the presentation of material and energy balances for four different configurations. These configurations range from a process where about 30% of the energetic output of the system proceeds as methanol to a system where 90% of the energetic output is in the form of methanol. When the capacities of all unit operations are sized sufficiently large, the whole range of production possibilities might be achieved with the same plant.

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6 This implies that in case of PSA, hydrogen has to be adsorbed in a selective way on the bed and subsequently released at low pressures. In case of membranes, H_2 must be in the permeate of the system, so the membrane has to be highly permeable to H_2, compared to CO_2, CO and CH_4.

7 This refers only to the electricity generated by the H_2 in the fuel gas of the combined cycle. The electricity generated by the CO and CH_4 in the fuel gas belongs to the 'carbon-containing fuels'.

2.4 MethaHydro material and energy balances

In this section, material and energy balances for four different configurations of the MethaHydro process are presented. In these configurations respectively 90, 70, 50 and 30 percent of the energetic output is in the form of methanol. The remaining output is split evenly in MEH and electricity[8].

2.4.1 Assumptions and design considerations

The material and energy balances of the MethaHydro process are based on the following assumptions and design considerations:

- *gasifier*

The composition of the product gas of the biomass gasifier is assumed to be as indicated in table 2.1.

Table 2.1 Composition of the gasifier product gas (Katofsky, 1993).

component	concentration (vol%)
H_2O	31,8
H_2	20,8
CO	15,0
CO_2	23,9
CH_4	8,2
C_2^+	0,3

- *autothermal reformer*

The autothermal reformer is designed, assuming an O_2 to C ratio of 1,9 and a steam to C ratio of 1,9 (Christensen and Primdahl, 1994). Operating conditions are assumed to be 950 °C and 30 bar. The product gas is assumed to be at equilibrium.

- *water-gas shift*

The product gas from the water-gas shift reactor is assumed to be at equilibrium. The amount of carbon dioxide in the product gas is adapted in such a way, that after carbon dioxide removal, the process gas contains about 5% carbon dioxide. This amount can be adjusted by choice of an adequate temperature of the water-gas shift or the creation of a bypass around the water-gas shift in combination with a water gas-shift operating at a sufficient low temperature (about 400 °C, which is still high for a water-gas shift reactor).

- *CO_2-absorption*

The amount of CO_2-C to be recovered and stored is assumed to be equal to the amount of natural gas-C MethaHydro-process.

8 This apportionment is based on the energetic content of the final products, i.e. the heating value of the methanol and MEH and the electrical power of the electricity.

- *methanol synthesis*

The methanol synthesis is designed at 300 °C and 80 bar, assuming that equilibrium is obtained after the methanol synthesis. The overall methanol conversion depends on the stoichiometric ratio of the synthesis gas feed to the methanol reactor. In normal methanol synthesis, the stoichiometric ratio is chosen to be 2.2. This gives a 97% conversion of carbon oxides to methanol using a recycle ratio of 3.5. Because the MethaHydro-process is used for the production of a mix of energy carriers (e.g. methanol, MEH and electricity), excess H_2 has to be present in the synthesis gas, besides the amount of H_2 required to convert the carbon oxides to methanol. Therefore, synthesis gas with a high stoichiometric ratio has to be converted in the methanol reactor. The lower the amount of methanol produced in a certain configuration, the higher the stochiometric ratio. As illustrated in figure 2.2, this requires higher recycle ratios to achieve the same conversion of carbon oxides. Concerning the design of the MethaHydro methanol synthesis, two possibilities exist: (i) the recycle ratio is increased to increase the conversion of carbon oxides to methanol, at the cost of increased volumes of the methanol synthesis equipment; (ii) a drop in conversion of carbon oxides to methanol is accepted and excess carbon oxides are converted to methane in the subsequent methanation reactor respectively released as CO_2 from the stack of the combined cycle. In this methanol synthesis design, the latter option is chosen, and the maximum recycle ratio is assumed to be 4 for all configurations. This means the methanol reactor and recycle loop have the same dimensions for all configurations.

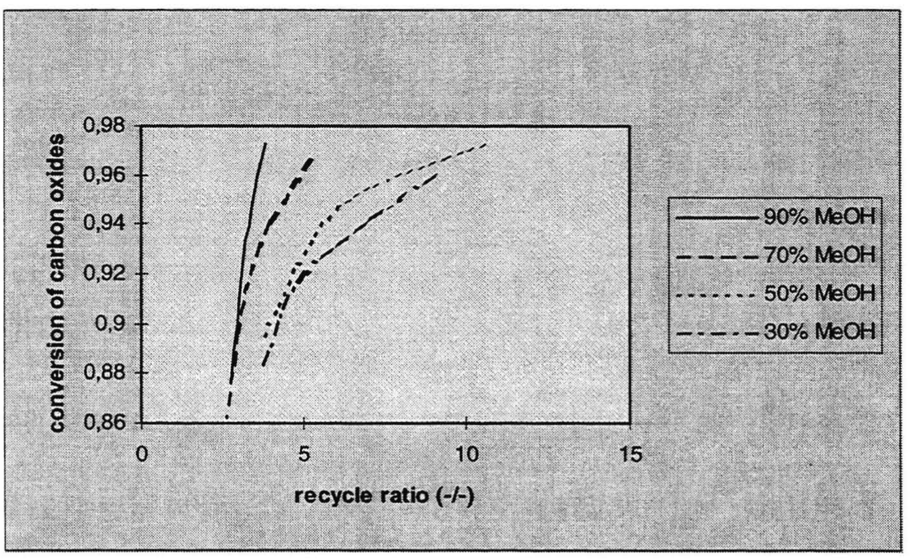

Figure 2.2 Conversion of carbon oxides to methanol as a function of methanol reactor throughput (the four lines represent the conversion for the syngas stochiometry of the four MethaHydro configurations considered here).

- *Methanation and electricity generation*

Conversion of carbon oxides in the methanation reactor is assumed to proceed completely. The efficiency of electricity generation is assumed to be 50%. The way in which the purge gas is divided over both utilisation option assumed to be such, that the output of the electricity generation in MW_e equals the output of the methanation reactor in MW_{th}.

2.4.2 Results of the calculations

Table 2.2 summarises the feed and product characteristics of the MethaHydro configurations considered in this study.

Table 2.2 Results of the material balance calculations.

	90% methanol	70% methanol	50% methanol	30% methanol
feed				
biomass (dried, tonne day^{-1})	1790	1340	980	680
natural gas (tonne day^{-1})	640	790	920	1020
products				
methanol (tonne day^{-1})	1800	1370	920	530
(automotive fuel 1000 l ge y^{-1})^2	*(1130)*	*(870)*	*(580)*	*(330)*
electricity (MW_e)	23	68	106	144
CH_4/H_2 (tonne day^{-1})	18/8	23/38	25/64	30/93
(as MEH, 1000 m^3 nge day^{-1})^3	*(63)*	*(180)*	*(290)*	*(390)*
energy in (MW)	732	732	732	732
energy out (MW)	462	442	424	412
overall efficiency (%)	63	60	58	56
CO_2 emission reduction (Mtonne y^{-1})^4	1,06	1,11	1,14	1,18

1: dried biomass contains 15% water

2: ge: gasoline equivalent (1 litre MeOH equals 0.5 litre gasoline)

3: MEH: Methane Enriched Hydrogen as an additive to the natural gas grid;
nge: natural gas equivalent (1 kJ_{th} equals 31 m^3 of natural gas)

4: CO_2 emission reduction is calculated, assuming 1,4 kg CO_2 emission reduction per kg methanol produced; 0,15 kg CO_2 MJ_e^{-1} and 0,05 kg CO_2 per MJ_{th} Methane Enriched Hydrogen.

Detailed results of the material balances are presented in Appendix 4.

2.4.3 Steam production and utilisation

The material balance of the MethaHydro-process as described before, is a preliminary design, based on overall material balances of unit-operations. A design of heat-exchangers networks, steam generation and consumption and optimisation of this network is out of the scope of this preliminary design. In this study it is assumed that the MethaHydro-process produces surplus steam (see table 2.3), so no steam or electricity import (e.g. from the combined cycle, resulting in a drop of the overall efficiency of the MethaHydro) is required to operate the MethaHydro process. This paragraph produces a first estimate of a steam balance of the MethaHydro-process as a further justification of this assumption. When upon further elaboration a MethaHydro-process proves to be an overall steam-consumer, the overall efficiency might be reduced somewhat, but the error introduced will be minor (for comparison, per m^3 natural gas, 12,5 kg steam might be produced, where the plant consumes the energetic equivalent of 25 m^3 s^{-1} natural gas).

The methanol synthesis is highly exothermic, and the heat of reaction usually is converted to steam. Normally about 1,5 to 2 kg steam are produced per kg methanol (Oonk and Visser, 1991). Besides, an amount of steam is generated and superheated, while cooling down the process gas (and when a steam reformer is applied, the off-gases) from the reformer. The latter amount depends more on the energetic input into the system and is less dependent on the energy mix produced.

The major steam consumers in the MethaHydro-process are:

- the drying of the biomass prior to gasification. For this purpose about 1 tonne of steam is required per tonne of dry biomass produced. Biomass may also be dried in a rotary drum dryer using the hot off-gases from the combined cycle. In such a case about 14 tonnes of waste gases (with a temperature in excess of 350 °C) are required per tonne of dry biomass produced. In case of the 30% and 50% methanol process, the amount of waste gases suffice for this purpose, so this drier might be an alternative. In case of the 70 and 90% process, the amount of waste gases produced is too small to dry the biomass in this way;
- the autothermal reformer and the biomass gasifier, where steam is used as a co-reactant (17-28 kg s^{-1}, see table 2);
- compressors for various gas streams in the process: O_2-compression; synthesis gas compression to 80 bar (about 7 MW); CO_2-compression to 100 bar for storage (3-5 MW). Part of the compression energy can be recovered from the decompression of the purge gas after methanol generation; About 1,1 kg s-1 steam required per MW compression power, steam consumption is about 12 kg s^{-1}.
- the distillation of the methanol product. The amount of steam used here depends on the qualitity of methanol required. Assumed a low to moderate quality is required when utilising methanol as a fuel, about 1 kg steam per kg methanol is required.

Table 2.3 gives an overview of steam generation and consumption in the various MethaHydro-process.

Table 2.3 Steam generation and consumption in the MethaHydro-process(in kg s^{-1}).

	90%	70%	50%	30%
steam generation	85	72	60	48
steam consumption				
- drying	21	15	12 (0)	8 (0)
- gasifier/reformer	17	21	23	28
- compression	12	12	13	14
- distillation	+ 21	+ 16	+ 11	+ 6
	71	64	59 (47)	56 (48)

the numbers between brackets refer to application of a rotary drum drier

2.5 Side products - other environmental effects

The MethaHydro-process produces a number of side-poducts and generates some environmental effects. Below the most important effects are described per unit-operation.

- *oxygen production*

During oxygen production N_2 is formed as a side product. The amount of N_2 produced is about 50-60 kg s^{-1} and is almost independent on the product mixture produced.

- *biomass drying and gasification:*

When biomass is dried using steam, low pressure steam is produced. Total steam production is about 1,7 kg steam per kg of dry biomass. Gasification results in the prodution of ashes, that might be re-used in several purposes.

- *product gas purification*

The biomass gasifier product gas contains a number of pollutants, that are not allowed in the autothermal reformer and the subsequent methanol synthesis. These pollutants will consists notably of particles and are separated prior the autothermal reformer. When a dry system is used, this results in an ash-like side-product, that might be used e.g. in the cement industry. When a wet system is used, e.g. a scrubber for simultaneous separation of particles and other components, a contaminated water stream is obtained.

- *autothermal reformer*

The main waste stream from the autothermal reformer consist of spent catalyst. At the moment no experience exists with the amount of spent catalyst produced and options for catalyst regeneration.

- *water separation*

After the autothermal reformer the synthesis gas is cooled and subsequent water is separated (15-20 kg s^{-1}). This water may contain some of the impurities coming from biomass gasification, and not removed before autothermal reforming. Examples of possible impurities in this water stream are ammoniumcarbonate of ammoniumchloride.

- *methanol synthesis and product make-up*

Methanol production produces steam. This steam is utilised in other parts of the MethaHydro-process (see chapter 2.4.3). The metals from the spent catalyst in methanol synthesis are generally recovered, so they represent no environmental burden. The light ends of methanol distillation consist notably of light alifates and olefins. They might be incinerated, e.g. in the combined cycle of the MethaHydro-process. The heavy ends of methanol synthesis (water contaminated by various hydrocarbon components-methanol, higher alcohols, other oxygen containing organic components, higher alifates and olefins, app. 1,5 kg s^{-1}) generally represent some difficulties in treatment.

- *combined cycle*

Exhaust emissions from turbines are generally rather low. The most important pollutant is NO_x. When modern turbines are used, emissions will be less than 65 g NO_x GJ_{th}^{-1}, resulting in total NO_x-emissions of about 90 and 560 tonne per year in the 90% MeOH and 30% MeOH-case, respectively.

3. The economics of the MethaHydro plant

3.1 Introduction

This chapter presents an analysis of the production costs of carbon dioxide-neutral fuels with the MethaHydro plant described in Chapter 2, following the approach of Williams *et al.* (1995). Section 3.2 gives a cost estimate of the hardware needed and Section 3.3 an estimate of the operation and maintenance (O&M) costs. In Section 3.4, the total costs per unit product are presented and compared with the production costs of the regular fuels which should be replaced by the MethaHydro products. All costs are expressed in 1994 US dollars. Appendix 5 contains more detailed information from the spreadsheet used for the cost calculations.

3.2 Capital costs

The capital costs of the MethaHydro plant are calculated separately for all the main parts of the installation. These cost calculations are based on manufacturers' data or on one or more reference plants from literature and are scaled according to an appropriate quantity and scale factor for that specific part. In the case manufacturers' data for separate items are used, the capital costs are increased with 25 percent for utilities and auxiliaries (Williams *et al.* 1995). In the case of reference plants from literature, utilities and auxiliaries are assumed to be included in the total capital costs.

- *oxygen plant*

The capital costs of the oxygen plant which supplies both the biomass gasifier and the autothermal reformer are based on a cost curve of Williams *et al.* (1995) for cryogenic on-site plants, which is in turn based on manufacturers' data. The capital costs are scaled according to the oxygen production rate with a scale factor 0.7. The plant produces 99.5% purity oxygen at 37 bar and includes 20 minutes gaseous oxygen storage.

- *biomass gasifier*

The pressurised biomass gasifier is the least developed part of the MethaHydro installation. Until now biomass gasifiers have not been built at a scale larger than 200 tonne per day of dry biomass, whereas in the MethaHydro plant up to 1500 tonne per day will have to be gasified. No technical complications are expected if biomass gasifiers are scaled up, but the costs of a first-of-a-kind large scale gasifier will be relatively high. On the long term, if several pressurised biomass gasifiers have been built, learning effects might reduce the capital costs of such installations dramatically. The costs of biomass gasifiers reported in the literature seem to depend strongly upon the assumption of the maturity of the technology at the time the gasifier will be built. Figure 3.1 shows the capital cost estimation for pressurised biomass gasifier systems from two sources, both including receiving, storage and sizing facilities, a steam dryer, the gasifier itself and gas make-up. The capital cost curve of Solantausta *et al.* (1996) is based on the costs of scaling up small-scale plants which are or could be built on short term. The curve of Williams *et al.* (1995) is based on a long-term perspective for large-scale gasifiers. Both studies assume a scale factor of 0.7 according to the biomass capacity of the gasifier.

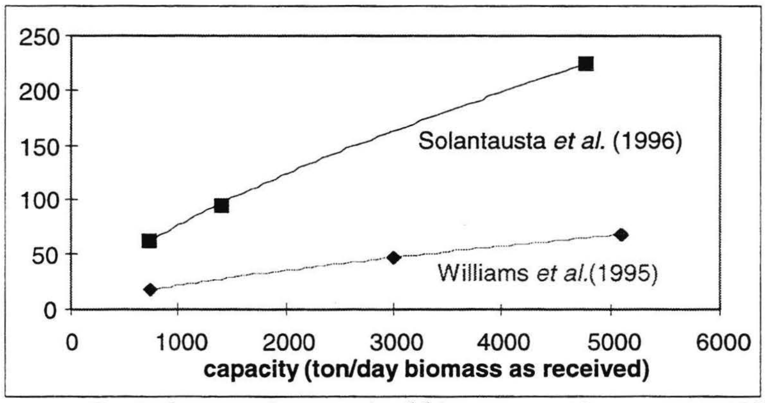

Figure 3.1 *Capital costs of pressurised biomass gasifier systems (10^6 US$)*

In this study the capital costs are assessed of the third large scale (>1000 tpd) biomass gasifier to be built in The Netherlands. To this end an adapted version of the curve of Solantausta et al. (1996) is used. In analogy to experiences in the chemical industry, assumed a capital cost reduction of 15 percent is assumed if the cumulative production capacity is doubled (Solantausta et al. 1996). This means the capital costs for the third plant are 70 percent of those for first plant.

- *autothermal reformer (ATR)*

The costs for an autothermal reformer are based on a unit with a capacity of 1.10^6 Nm^3/day of natural gas. For this unit the costs are estimated at 20 million US dollars including feed preparation and heat recovery section (Holm-Larsen 1997). The ATR is scaled according to the methane input with a scale factor of 0.8.

- *shift reactor*

The shift reactor described in Hendriks (1995), has a capacity of 6355 kmol/h $CO+H_2$. The estimated costs are 30 million US dollars, including water separation. This estimate is used in this study with a scale factor of 0.7 according to the amount of $CO+H_2$ which enters the shift reactor.

- *syngas compressor*

The syngas compressor, the recycle compressor of the methanol synthesis and the steam expander used to drive the compressors are taken together as one cost item. The cost are scaled with a factor 0.8 according to the compressor power. The reference system is taken from van Dijk *et al.* (1995), who describe a unit with plant investments of 53 million US dollars for a system with a syngas compressor of 6700 kW.

- *CO_2-removal*

The investments for the Selexol CO_2-removal unit are based on the cost estimate for CO_2-removal from coal gas (at 25 bar) by Hendriks (1995) and are scaled according to the quantity of CO_2 removed with a factor 0.7. Hendriks describes a unit removing 9936 kmol/h CO_2 requiring an investment of 49 million US dollars, including compression of the carbon dioxide to 80 bar. The costs of a high pressure system as described in Chapter 2, are expected to be equal or lower than those of the system described by Hendriks. The amount of Selexol needed will be lower because of the higher solubility of CO_2 under these conditions, the costs for the absorber however will be higher. This is confirmed by the system of Oudhuis (1992), which removes both H_2S and CO_2 (9850 kmol/h) at a pressure of 78 bar at an investment 34 million US dollars.

- *methanol reactor*

The capital costs of the methanol reactor and methanol distillation unit are calculated for the 90% methanol product mix on the basis of the ICI process as described by van Dijk *et al.* (1995). Their system produces 2130 tonne per day grade AA Methanol at an investment of 35 million US dollars. The costs are scaled according to the methanol production rate with a scale factor 0.8. In the configurations of the MethaHydro plant for other product mixes, the methanol reactor has the same size as in the 90%-MeOH case (see Chapter 2). Therefore, the capital costs for the methanol reactor are equal for all product mixes.

- *methanation*

Since the methanation reaction is a reversed reforming reaction, the costs of the reactor are derived from the steam reformer described by van Dijk *et al.* (1995), which produces 11,844 kmol syngas per hour and requires an investment of 19 million US dollars. The costs are scaled according to the gas flow to the methanation reactor with a scaling factor 0.8.

- *combined cycle*

The costs for the combined cycle are derived from Solantausta *et al.* (1996), who give an investment of \$830 per kW_e.

Table 3.1 gives an overview of the capital costs for the MethaHydro plant for each of the product mixes, based on the estimations described above. For the flexible plant, each part is priced at the highest costs for the other product mixes to allow production of all four considered product mixes with the flexible plant. For reasons of comparison, the table also shows the capital costs of a methanol plant which converts natural gas in methanol. This reference plant is assumed to be equipped with an autothermal reformer.

Table 3.1 *Capital costs of MethaHydro plant (10^6 US$).*

	Product mix (%MeOH of energy output)					'Normal' MeOH plant
Installed hardware	**90%MeOH**	**70%MeOH**	**50%MeOH**	**30%MeOH**	**flexible**	
Oxygen plant	44	46	46	48	48	53
Biomass gasifier	115	94	75	58	115	
Autothermal reformer	23	26	28	29	29	35
Shift reactor	34	35	35	36	36	
Syngas compressor	47	50	51	54	54	53
Carbon dioxide removal	14	16	18	19	19	
Methanol reactor	30	30	30	30	30	33
Methanation reactor	1	2	4	5	5	
Combined cycle	18	54	85	123	123	
Utilities/auxiliaries	29	25	26	27	27	43
Total costs installed hardware	306	293	279	272	329	217

- *additional capital costs*

Contingencies are estimated at 20 percent of total installed hardware costs, owners costs and fees 20 percent and start-up costs at 5 percent (Williams *et al* (1995).

- *assumptions for capital depreciation*

The whole MethaHydro plant is assumed to have an economic lifetime of 25 years. In the calculation of the unit costs (Section 3.4), a capital charge rate of 15% will be used. Also, the effect of using a 10% rate instead on the total costs per unit will be shown.

3.3 Operating and maintenance costs

- *labour*

The labour costs of the MethaHydro plant are assumed to be linearly related to the fuel inputs. Biomass requires more labour than natural gas because receiving, storage and handling are more complicated. Labour costs for the whole methanol plant are $920 annually per tonne/day of dry biomass (Solantausta *et al.* 1996, Williams *et al.* 1995) and $0.27 annually per m^3/day of natural gas (Williams *et al.* 1995).

- *maintenance*

Maintenance costs are set at an annual 3 percent of the installed hardware costs (Williams *et al.* 1995) for all parts, excluding the biomass gasifier for which 5 percent is assumed because of the corrosive conditions in this part of the installation.

- *overhead*

Direct overhead costs of the MethaHydro plant are assumed to be 45 percent of the annual labour costs, while general overhead is set at 65 percent of the annual labour and maintenance costs (Williams *et al.* 1995).

• *biomass*
As discussed in Section 2.2, it is not possible to give one price for imported biomass. Therefore, in the cost calculations in this section, a price range of \$2-7/GJ is used for biomass[9].

• *natural gas*
For the natural gas price, the shadow prices from the Global Competition and the European Co-ordination scenarios for the year 2020 of \$0.14/m^3, respectively \$0.11/m^3 are used.

• *catalysts*
The catalysts used in the ATR, shift reactor, Selexol unit and methanol and methanation reactors are taken together as one cost item. For this item the estimation of 3 million per year of Williams *et al.* (1995) for a comparable plant is used.

• *CO_2-storage*
The costs of CO_2-storage are considered as operating costs in this study because it is an off-site activity. For carbon dioxide storage the cost estimate of \$1/tonne CO_2 by Hendriks (1995) is used, which is based on storage in a depleted natural gas field at a depth of 2 km and with a flow rate of 6 m^3/s. These costs exclude transport costs.

Table 3.2 shows the annual operation and maintenance (O&M) costs of the different MethaHydro plants. The O&M costs for the flexible plant depend on the actual input and output. In Section 3.4 the unit costs of the flexible plant will be analysed using the O&M costs for the product mixes shown in Table 3.2. Table 3.2 also shows the O&M costs for the 'normal' methanol plant.

9 In fact, a price range of 4-12 Dutch guilders per GJ has been used (\$2.3-6.8/GJ).

Table 3.2 *Operating & maintenance costs (10^6 US$/year) of the MethaHydro plant in the year 2020 under the Global Competition scenario (in brackets the costs which are different under the European Co-cordination scenario)*

	Product mix (%MeOH of energy output)				'Normal' MeOH plant
	90%MeOH	70%MeOH	50%MeOH	30%MeOH	
Variable costs					
Biomass	25-74	18-55	13-40	9-28	
Natural gas	50 (40)	62(50)	72 (57)	80 (64)	102 (80)
CO_2 storage	1	1	1	1	
Catalysts and chemicals	3	3	3	3	2
Fixed costs					
Labour	2	1	1	1	1
Maintenance	10	10	9	9	7
Overhead	8	8	7	7	5
Total operating costs	98-147 (88-137)	103-140 (92-128)	107-134 (92-119)	110-129 (94-113)	116 (94)

3.4 Total costs

Table 3.3 shows the total unit costs for each of the MethaHydro products in 2020 under the Global Competition scenario. As described in Section 3.2, hardware is depreciated over 25 years at a 15 percent capital charge rate. The natural gas price is $0.14/m^3 and the biomass price range is $2.3-6.8/GJ.
The allocation of the joint costs (i.e. costs of syngas production) to the three products is based on the energetic contents of the products at the split-off point. The costs after the split-off point are directly allocated to the related product. For reasons of comparison, the last column of Table 3.3 shows the costs of a normal methanol plant with an ATR, calculated with the same assumptions.

Table 3.3 Production costs of MethaHydro energy carriers in the year 2020 under the Global Competition scenario (all costs in $/GJ).

	Product mix (%MeOH of energy output)												ref.
	90% MeOH			70% MeOH			50% MeOH			30% MeOH			MeOH plant
	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH
Capital													
Share of syngas production	5,1	5,1	5,1	4,6	4,6	4,6	4,2	4,2	4,2	3,9	3,9	3,9	2.9
Methanol production	0,4	0,0	0,0	0,5	0,0	0,0	0,7	0,0	0,0	1,2	0,0	0,0	0.3
MEH production	0,0	0,0	0,2	0,0	0,0	0,2	0,0	0,0	0,1	0,0	0,0	0,2	
Electricity production	0,0	2,0	0,0	0,0	2,0	0,0	0,0	2,0	0,0	0,0	2,0	0,0	
Labour & maintenance	1,3	1,3	1,3	1,2	1,2	1,2	1,1	1,1	1,1	1,0	1,0	1,0	0.8
Biomass[1]	2-5	2-5	2-5	1-3	1-3	1-3	1-2	1-2	1-2	1-2	1-2	1-2	
Gas	3,3	3,3	3,3	3,8	3,8	3,8	4,2	4,2	4,2	4,4	4,4	4,4	6.0
CO2-storage	0,0	0,0	0,0	0,0	0,0	0,0	0,1	0,1	0,1	0,1	0,1	0,1	
Catalysts/chemicals	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0,2	0.1
Total Production Costs	12-15	14-17	12-15	11-14	13-15	11-13	11-13	13-14	11-12	11-12	12-13	10-11	10.1
Total production costs under the EC[2] scenario	11-15	13-16	11-14	11-13	12-15	10-13	10-12	12-13	10-11	10-11	11-12	9-10	9.2

[1] The costs for biomass are based on a price range of $2.3-6.8/GJ

[2] European Co-ordination

As can be seen in this table, the costs of CO_2-neutral methanol produced with the MethaHydro plant are 10 to 50 percent higher than the costs of "normal" methanol, depending on the MethaHydro configuration chosen and the biomass price.

Flexible plant

In Table 3.4, the production costs of the flexible MethaHydro plant are shown. These are based on the capital costs for the flexible plant as shown in Table 3.1 and the O&M-costs for the different product mixes as shown in Table 3.2.

Table 3.4 Production costs of the flexible MethaHydro plant in the year 2020 ($/GJ).

	Product mix (%MeOH of energy output)											
	90% MeOH			70% MeOH			50% MeOH			30% MeOH		
Scenario	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH	Electr.	MEH	MeOH	Electr.	MEH
Global Competition	13-16	27-30	13-17	12-15	17-19	12-15	12-14	15-16	12-13	12-13	13-14	11-12
European Co-ordination	12-15	26-30	13-16	12-14	16-18	11-14	11-13	14-16	11-13	11-12	12-14	10-11

MEH: Methane Enriched Hydrogen

From the table, it can be concluded that the production costs of electricity are extremely high if the flexible plant produces 90 or 70 percent methanol. The high costs of a combined cycle unit require a high load to provide economic electricity generation. The production costs of methanol and MEH are much less sensitive for the product mix chosen in the flexible plant. From these results, it can be concluded that the flexibility of a MethaHydro plant should be limited to the ratio between MEH and methanol. To assess whether such flexibility is worth 10 percent higher production costs, requires more insight in the seasonal variation in the demand for the MethaHydro fuels. The further calculations in this chapter are based on the costs of plants designed for a fixed product mix.

Comparison with other processes
A comparison has been made between the production costs of the MethaHydro fuels and the costs of other CO_2-neutral or -lean energy carriers. Cost estimations from other studies have as much as possible been brought in line with the assumptions in this report. Capital charge rate, life time, biomass price, natural gas price and the capital costs for specific units have been adapted[10].

For the MethaHydro process, the costs from the 50%-MeOH configuration have been used in the comparison. This configuration has lower unit costs than the 70%- and 90%-MeOH configurations and still produces a large amount of methanol. The unit costs of the MethaHydro products are strongly influenced by the rules for the allocation of joint costs to the three products. Therefore, the figures in this section show the unit costs for two allocation principles: the one used above, based on the energetic content of the products and one based on the value of the products using the prices of the cheapest competing CO_2-neutral/lean process.

Figure 3.2 shows the production costs of MethaHydro methanol and the production costs for the production of methanol from biomass only. The latter is based on a study by Williams *et al.* (1995), with the following adaptations: the costs of the biomass gasifier have been estimated on the basis of Solantausta *et al.* (1996) as described in Section 3.2; a biomass price of \$2.3-6.8/GJ is used. The cost range for MethaHydro methanol shown in Figure 3.2 comprises the range in natural gas costs, assuming both the Global Competition and European Co-ordination scenario.

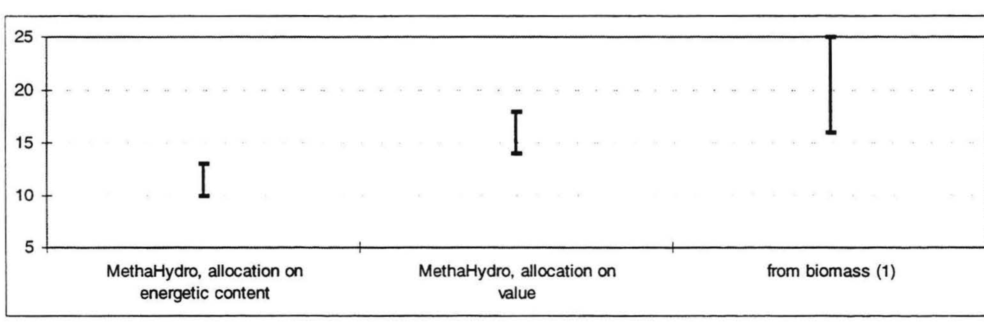

Figure 3.2 Production costs of CO_2-neutral methanol (\$/GJ), (1) adapted from Williams et al. 1995

10 The scale of the processes in other studies has not been adapted.

The costs of methanol from biomass are much higher than costs of MethaHydro methanol. The cost difference is smaller if the joint costs of the MethaHydro plant are allocated according to the value of the products, but still significant. The difference is mainly caused by the fact that the capital costs for the use of natural gas (autothermal reformer and CO_2-removal unit) are much lower than for the use of biomass (dryer, gasifier etc.). The smaller scale (800 tpd methanol) of the design by Williams *et al.* affects the costs of the biomass-based process as well.

In Figure 3.3, the production costs of MethaHydro electricity are compared to the production costs of electricity from biomass and of electricity from natural gas. The production costs of electricity from biomass are adapted from a study by Solantausta *et al.* (1996) which is based on a pressurised gasification combined cycle of 60 MWe. The costs of the biomass gasifier are reduced with 30% as described in Section 3.2, a biomass price of \$2.3-6.8/GJ is assumed, the lifetime has been set to 25 years and the capital charge rate to 15%. The production costs of electricity from natural gas are based on a study by Hendriks *et al.* (1992) which is based on a 600 MWe combined cycle plant with CO_2-recovery using Selexol. It should be noted that this process is not completely CO_2-neutral, since it emits 10% of the CO_2 of a normal natural gas fired combined cycle. The cost estimation has been adapted with regard to the capital charge rate and the gas price. In Figure 3.3, the cost ranges for the MethaHydro electricity and the electricity from natural gas comprise the costs with the natural gas price from the Global Competition as well as the European Co-ordination scenario.

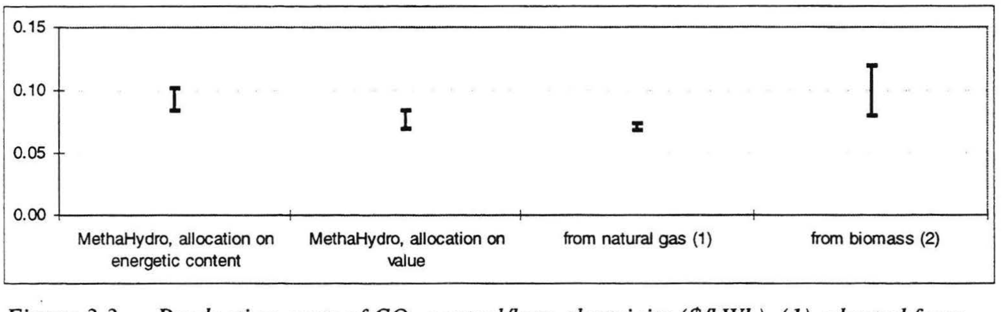

Figure 3.3 Production costs of CO_2-neutral/lean electricity (\$/kWh), (1) adapted from Hendriks et al. 1992, (2) adapted from Solantausta et al. 1996

If the joint costs of the MethaHydro plant are allocated on energetic content, its electricity costs much more than CO_2-lean electricity from natural gas. However, if the joint costs are allocated according to the value of the products, the cost difference becomes much smaller. Electricity from biomass can only be produced at similar costs if the biomass price is low. Note that the costs of the system of Solantausta *et al.* (1996) might decrease if it is operated at a larger scale.

Figure 3.4 shows the production costs of hydrogen from biomass, from natural gas and with the MethaHydro process. The costs of hydrogen from biomass are based on Williams *et al.* (1995) with the same adaptations as described above for their cost estimate of methanol production. The costs of hydrogen production from natural gas are adapted from Blok *et al.* (1997): the capital charge rate has been set to 15% and the gas price from the Global Competition and the European Co-ordination scenario have been introduced. The benefits of the enhanced gas recovery by injection of the CO_2 in depleted gas fields are not taken into account here. It should be noted that this process is not CO_2-neutral: its CO_2-emission is 90% lower than a normal hydrogen plant. The costs hydrogen from the MethaHydro process have been calculated from the production costs of MEH. These costs have been increased by $1/GJ to account for purification (e.g. with Pressure Swing Adsorption, Williams *et al.* 1995).

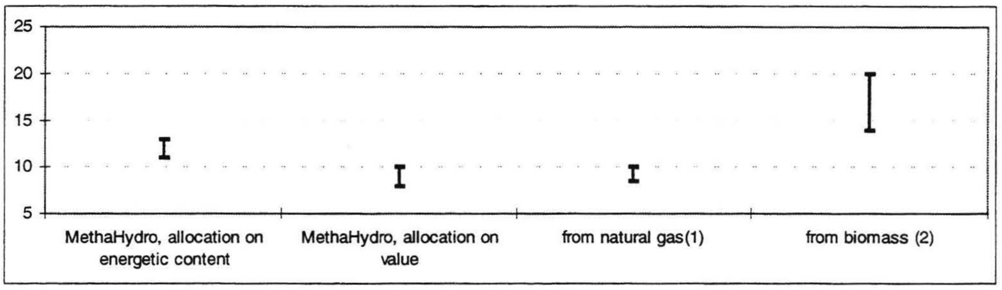

Figure 3.4 Production costs of CO_2-neutral/lean hydrogen ($/GJ), (1) adapted from Blok et al. 1997, (2) adapted from Williams et al. 1995

Hydrogen from the MethaHydro process is cheaper than hydrogen from biomass or natural gas if the joint costs of the plant are allocated according to the value of the products. Hydrogen from natural gas is more attractive if the joint costs of the MethaHydro plant are allocated according to the energetic content of the products. Hydrogen from biomass is under all circumstances more expensive.

The conclusion of the comparison of the MethaHydro products with other CO_2-neutral/lean energy carriers strongly depends on the allocation method used for the joint costs of the MethaHydro plant. If the allocation is based on the energetic content of the products, the methanol from the MethaHydro process is much cheaper than methanol from biomass, but MethaHydro electricity and hydrogen are costlier than competing options. However, if the allocation is based on the value of the products, MethaHydro hydrogen and methanol are cheaper than the competing options and MethaHydro electricity has approximately the same production costs as CO_2-lean electricity from natural gas.

Comparison with regular fuels
Table 3.5 shows a comparison between the specific production costs in the year 2020 of the fuels from the 50%-MeOH MethaHydro plant and the regular fuels that would be replaced.

Table 3.5 Production costs for the fuels of a 50% MeOH MethaHydro plant and regular fuels (in italics) in the year 2020.

	Scenario	
	Global Competition	**European Co-cordination**
MethaHydro Methanol ($/l ge)		
-allocation on energetic content	0.36-0.41	0.33-0.38
-allocation on value	0.48-0.55	0.44-0.51
Gasoline ($/l)	*0.26*	*0.17*
MethaHydro electricity ($/kWh)		
-allocation on energetic content	0.09-0.10	0.08-0.10
-allocation on value	0.07-0.08	0.07-0.08
Electricity ($/kWh)	*0.06*	*0.04*
MethaHydro MEH ($/m^3 nge)		
-allocation on energetic content	0.34-0.39	0.31-0.36
-allocation on value	0.24-0.28	0.23-0.26
Natural gas ($/m^3)	*0.14*	*0.11*

The cost calculations are based on a biomass price of $2.3-6.8/GJ

ge: gasoline equivalent (1 litre MeOH equals 0.5 litre gasoline, assuming equal efficiency for gasoline and methanol fuelled engines)

MEH: Methane Enriched Hydrogen as an additive to the natural gas grid

nge: natural gas equivalent (1 kJ_{th} equals 31 m^3 of natural gas)

Depending on the biomass price and the allocation rules for the joint costs, the costs of MethaHydro methanol as transport fuel will be 40 to 200% higher than those of gasoline. If the levy imposed on methanol would be equal to the levy on gasoline, the price difference for the consumer would be ¢10-20 on a gasoline end-use price of approximately $1.10/l if the joint costs of the MethaHydro plant are allocated on energetic content. If the joint costs are allocated on value, the price difference for the consumer will be ¢20-35. A lower levy on methanol than on gasoline could make the end-use price of methanol equal or lower than the price of gasoline.
If the joint costs are allocated on energetic content, the production costs of MethaHydro electricity will be 50 to 150% higher than the average for the Dutch electricity generation. The absolute price difference for the consumer will be ¢3-6 on a end-use price of around ¢15/kWh. If the joint costs are allocated on value, the costs of MethaHydro electricity will be 50 to 100 higher than the average for the Dutch plant and the price difference will only be ¢2-4.
Production costs of Methane Enriched Hydrogen (MEH) will be approximately 3 times as high as those of natural gas, if joint costs of the MethaHydro plant are allocated on energetic content. If the joint costs are allocated on value, the production costs of MEH will be twice as high as of natural gas. The price difference between MEH and natural gas is not of direct interest for the consumer because MEH can not be used in the current gas grid and appliances. It can only be mixed with natural gas to a maximum of 5 percent, which would only lead to a small price increase for the end-user.

Cost sensitivity

Table 3.6 shows the sensitivity of the production costs of energy carriers from the 50%-MeOH MethaHydro plant for changes in the capital charge rate and the capital costs. All costs are calculated on the basis of allocation of joint costs on value.

Table 3.6 Sensitivity of the production costs of energy carriers from the 50%-MeOH MethaHydro plant in 2020 under the Global Competition scenario, joint costs allocated on value.

	Reference case	10% capital charge rate	25% higher capital costs	25% lower capital costs
Methanol ($/l ge)	0.48-0.55	0.42-0.49	0.43-0.50	0.53-0.60
Electricity ($/kWh)	0.07-0.08	0.06-0.07	0.07	0.09
Methane Enriched Hydrogen ($/m^3 nge)	0.24-0.28	0.22-0.25	0.22-0.26	0.27-0.31

If a capital charge rate of 10% is used instead of 15%, the unit costs of the MethaHydro fuels are reduced with approximately 10 percent. The effect of 25% higher respectively lower capital costs is in the same order of magnitude.

Cost of carbon dioxide mitigation

The costs of CO_2-mitigation may be defined as the cost difference of the CO_2-neutral fuel as produced by the MethaHydro plant and the normal fuel, divided by the specific CO_2-emission of the normal fuel. Table 3.7 shows the CO_2-mitigation costs for the 50%-MeOH plant under the Global Competition and European Co-ordination scenario for the two allocation rules.

Table 3.7 CO_2-mitigation costs ($/tonne) for the 50%-MeOH MethaHydro plant in 2020.

	Methanol	Electricity	MEH	Average
Global Competition				
-allocation[1] on energetic content	45-70	60-80	120-150	65-90
-allocation on value	100-135	35-50	65-85	65-90
European Co-ordination				
-allocation on energetic content	80-100	70-90	120-150	80-105
-allocation on value	130-160	45-60	70-90	80-105

[1] of joint costs

The average CO_2-mitigation costs under the Global Competition scenario are significantly lower than under the European Co-ordination scenario. This difference is explained by the lower prices of regular fuels under the latter scenario, especially for gasoline.

If the joint costs of the MethaHydro plant are allocated on energetic content, the mitigation costs are lowest for methanol and highest for MEH. If the allocation is based on the value of the products, electricity has the lowest mitigation costs and MEH the lowest.

4. Discussion and conclusions

The MethaHydro-process

In a MethaHydro-process fossil fuels (coal, natural gas, oil residues, plastic waste) and biomass are converted to a mixture of carbon-dioxide neutral fuels. Such a MethaHydro process consists of unit operations for synthesis gas generation (gasification and reformers), unit-operations for synthesis gas treatment (removal of e.g., particles, SO_2, alkali-metals; water-gas shift, carbon dioxide removal and underground storage) and unit operations for synthesis gas utilisation (e.g., H_2-purification, electricity generation, methanol synthesis, Fischer-Tropsch synthesis). The lay-out of a MethaHydro-process depends on a number of factors. The most important factors are: the capacity of the process, the mix of fossil fuels used, the mix of energy carriers produced and the flexibility in production.

In this study a MethaHydro-process is designed, suitable for processing both of biomass and natural gas, and producing a flexible mixture of methanol, Methane Enriched Hydrogen and electricity at a large scale.

Typical inputs and outputs of this MethaHydro-process, producing various mixtures of methanol, electricity and Methane Enriched Hydrogen are described in table 4.1.

Table 4.1 Results of the material and energy balance calculations.

	90% methanol	70% methanol	50% methanol	30% methanol
feed				
biomass (dried, tonne day^{-1})	1790	1340	980	680
natural gas (tonne day^{-1})	640	790	920	1020
products				
methanol (tonne day^{-1})	1800	1370	920	530
(automotive fuel 1000 l ge y^{-1})^{2}	*(1130)*	*(870)*	*(580)*	*(330)*
electricity (MW_e)	23	68	106	144
CH_4/H_2 (tonne day^{-1})	18/8	23/38	25/64	30/93
(as MEH, 1000 m^3 nge day^{-1})^{3}	*(63)*	*(180)*	*(290)*	*(390)*
energy in (MW)	732	732	732	732
energy out (MW)	462	442	424	412
overall efficiency (%)	63	60	58	56
CO_2 emission reduction (Mtonne y^{-1})^{4}	1,06	1,11	1,14	1,18

1: dried biomass contains 15% water
2: ge: gasoline equivalent (1 litre MeOH equals 0.5 litre gasoline)
3: MEH: Methane Enriched Hydrogen as an additive to the natural gas grid; nge: natural gas equivalent (1 kJ_{th} equals 31 m^3 of natural gas)
4: CO_2 emission reduction is calculated, assuming 1,4 kg CO_2 emission reduction per kg methanol produced; 0,15 kg CO_2 MJ_e^{-1} and 0,05 kg CO_2 per MJ_{th} Methane Enriched Hydrogen.

Technological feasibility

The MethaHydro process as described in this study can be considered as technologically feasible, since all unit-operations involved are demonstrated in real-scale applications. Technology development and improvement of such a MethaHydro process is still possible for:

- the technology for large-scale biomass gasification;
- the integration of gasification and autothermal reforming, this implies either adapting biomass gasification in such a way that the process gas is optimal for autothermal reforming or adapting autothermal reformer in such a way that the process gas from the gasifier can easily be handled without further purification;
- the carbon dioxide separation and subsequent regeneration of the absorber has to be demonstrated at process conditions and may be subject to further improvements;
- the development of methanol-synthesis suitable for operating at high stoichiometric ratios;
- development of technology for H_2 -separation from the product gas in order to adjust the stoichiometric ratio before methanol synthesis;
- the dynamics of the process, since a flexibility for hydrogen and electricity production on a few hours scale might be an advantage of a MethaHydro process.

Costs

In determining the costs of the MethaHydro fuels, the joint costs of production (notably the synthesis gas production and conditioning) have to be allocated. In this study two methods are used: allocation according to the energy content produced, and allocation according to the market prices of the products.
Table 4.2 shows the production costs in the year 2020 of the fuels from a MethaHydro plant which delivers 50% of its output in the form of methanol. The table also shows the costs of the regular fuels that would be replaced. The costs of MethaHydro methanol as transport fuel will be 40 to 200% higher than those of gasoline. If the levy imposed on methanol would be equal to the levy on gasoline, the price difference for the consumer would be ¢10-35 (depending on allocation rules for capital costs) on a total price of approximately $1.10. A lower levy on methanol could make its end-use price equal or lower than the price of gasoline. The production costs of MethaHydro electricity will be 50 to 150% higher than the average for the Dutch electricity generation. Methane Enriched Hydrogen will be 3 times as expensive as natural gas.

Table 4.2 Production costs for the fuels of a 50% MeOH MethaHydro plant and regular fuels (in italics) in the year 2020.

	Scenario	
	Global Competition	**European Co-cordination**
MethaHydro Methanol ($/l ge)		
-allocation on energetic content	0.36-0.41	0.33-0.38
-allocation on value	0.48-0.55	0.44-0.51
Gasoline ($/l)	*0.26*	*0.17*
MethaHydro electricity ($/kWh)		
-allocation on energetic content	0.09-0.10	0.08-0.10
-allocation on value	0.07-0.08	0.07-0.08
Electricity ($/kWh)	*0.06*	*0.04*
MethaHydro MEH ($/m^3 nge)		
-allocation on energetic content	0.34-0.39	0.31-0.36
-allocation on value	0.24-0.28	0.23-0.26
Natural gas ($/m^3)	*0.14*	*0.11*

The cost calculations are based on a biomass price of $2.3-6.8/GJ
ge: gasoline equivalent (1 litre MeOH equals 0.5 litre gasoline)
MEH: Methane Enriched Hydrogen as an additive to the natural gas grid
nge: natural gas equivalent (1 kJ_{th} equals 31 m^3 of natural gas)

The costs of methanol production using the MethaHydro process are significantly cheaper than production from biomass only. When the joint costs are allocated according to the value of the products, converting the MEH from the MethaHydro process into pure hydrogen is cheaper than hydrogen from biomass and is comparable to hydrogen from a natural gas based process with CO_2-recovery, Concerning electricity production, a natural gas based process with CO_2-recovery is about as costly as electricity from the MethaHydro-process, provided that the joint costs are allocated according to the value of the products. Biomass based electricity production is the most expensive option. Compared to traditional methanol production, CO_2-neutral methanol from the MethaHydro-plant is about 10-50% more expensive, depending on configuration chosen and the biomass price.

The average costs of CO_2-mitigation for the 50%-MeOH MethaHydro plant under the Global Competition scenario are $65-90 per tonne CO_2. In the European Co-ordination scenario average costs are $80-105 per tonne CO_2.

Future expectations
With regard to biomass gasification, autothermal reforming and conversion of synthesis gas to electricity, developments are expected in the next decades. As a result the MethaHydro process may be significantly improved, without further MethaHydro specific development. The learning effect for the construction of biomass might result in substantial reductions in capital costs (see Fig. 3.1), as assumed by Williams et.al. (1995). For the MethaHydro process this might result production costs reduction up to 10-15%.
On the demand side, new applications might be expected for hydrogen (e.g. fuel cells in (public) transportation). In industrial chemistry the interest in synthesis gas

or methanol as a starting material for carbon-chemistry might be growing. As a result changes may be expected for the market of the MethaHydro products as well.

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6. Authentication

Name and address of the principal:
Ministry of VROM/DGM/LE
Attn. Drs. R.A. v/d Wijngaart
P.O. Box 30945
2500 GX The Hague

Names and functions of the cooperators:
Hans Oonk
Jasper Vis
André Faaij
Ernst Worrell
Jan-Willem Bode

Names and establishments to which part of the research was put out to contract:
Utrecht University
Dept. Of Science, Technology and Society
Padualaan 14
3584 CH Utrecht
The Netherlands

Date upon which, or period in which, the research took place:
December 1996 - June 1997

Signature:

Hans Oonk
research coordinator

Approved by:

C. Roos
section leader

Appendix 1 Description of the unit-operations

Contents

1. Synthesis gas production

1.1 Natural gas

Natural gas is the most used feedstock for synthesis gas production. Natural gas can be cracked both by steam reforming and by partial oxidation. In steam reforming (Section 1.1.1), the feedstock is catalytically cracked in the absence of oxygen with the addition of water. The reaction heat is supplied externally. In partial oxidation (Section 1.1.2), cracking takes place without catalyst, reaction heat is generated by direct oxidation of part of the feedstock with oxygen (Fiedler *et al.* 1990). Combinations of steam reforming and partial oxidation also have been proposed and appliedimplying a primary and secondary reformers. The latest development is autothermal reforming (Section 1.1.3),, in which the primary and secondary reformer are integrated.

1.1.1 Steam-reforming

Methane steam-reforming is widely applied for the generation of synthesis gas for methanol production (Kirk-Othmer). Besides that methane steam-reforming is at the moment the most efficient and most widely applied method for generation of hydrogen (Johansen et al., 1992). Steam reforming is performed over a Ni-ctalyst, sensitive to a.o. sulphur in the natural gas feed. Overall reactions in steam-reforming of methane are:

$$CH_4 + H_2O \leftrightarrow CO + 3\ H_2$$

$$CO + H_2O \leftrightarrow CO_2 + H_2$$

The conversion of both reactions are determined by their equilibria, and the final product gas is a mixture of CH_4, H_2O, CO, CO_2 and H_2. The conversion depends on operating temperature and pressure, the steam-to-carbon ratio in the steam-reformer and the composition of the natural gas, and is normally limited to about 75-85% (870 °C, 20 bar).

When pure methane is reformed, the stoichiometric ratio

$$C = ([H_2] - [CO_2]) / ([CO] + [CO_2])$$

of the product is 3. This ratio is independent of the methane conversion and water-gas shift. When the natural gas feed contains carbon dioxide or higher hydrocarbons, this stoichiometric ratio is somewhat reduced.
The steam reform reaction is highly endothermic. For this reason steam-reforming is normally performed in a natural gas operated furnace. In this furnace, 0,37 m^3 of

methane (as natural gas) is required per m^3 of methane converted to synthesis gas. In many applications, this amount of fuel gas is somewhat reduced, because energetic waste streams can be used as a fuel. For example in methanol production, the hydrogen and methane containing waste purge stream can be used (see Section 3.2), and when hydrogen is produced, using PSA-separation technology (see Section 3.1), purge gases from the PSA may be used for this purpose.

1.1.2 Partial oxidation

In partial oxidation, natural gas is transformed to synthesis gas at temperatures of 1200-1400 °C and pressures of 30-80 bar. The overall reaction is:

$$CH_4 + \tfrac{1}{2}\,O_2 \rightarrow CO + 2\,H_2$$

The reaction is not limited by its equilibrium and proceeds almost to completeness. The product mixture consists notably of CO and H_2. Additional hydrogen can be produced using the water-gas shift reaction, at addition of water:

$$CO + H_2O \rightarrow CO_2 + H_2$$

In contrast to the steam-reform reaction, the partial oxidation of methane is an exothermic reaction and requires no catalyst. As a result, partial oxydation may be applied to fossil fuels that have high contents of sulphur.

1.1.3 Autothermal reforming

A new development in transforming methane to synthesis gas is the autothermal reformer. An autothermal reformer is a combination of a steam-reformer (converting CH_4 with steam in an endothermal reaction) and a partial oxidation (converting CH_4 with oxygen in an exothermal reaction). The reaction is performed in such a way that the heat generated by partial oxidation provides the heat required for steam-reforming. For this purpose a O_2 to CH_4 ratio is required of about 0.58 and a steam to CH_4 ratio is used of about 1.9 (Christensen and Primdahl, 1994). The autothermal reactor operates at 950 °C and 30 bar. The product of the autothermal reformer is a near equilibrium synthesis gas, containing large amounts of CO and H_2, smaller amounts of CO_2 and unconverted CH_4. Besides that, H_2O is present in the gas as a result of excess steam introduced in the autothermal reformer.

1.2 Biomass Gasification

Thermo-chemical conversion of biomass can be divided into three basic steps: feedstock preparation, pyrolysis and char gasification and combustion.
After chipping the biomass, it is dried to 10-15% moisture. Drying is an energy intensive step that consumes roughly 10% of the energy content of the feedstock. Usually low grade waste heat is available from elsewhere in the process that is suitable for drying, e.g. exhaust gas of gas-turbine or excess steam of methanol reactor. In the last preparatory step the biomass is brought in a form that meets the feed size and density requirements of the gasifier by densification or grinding (Katofsky 1993).
During pyrolysis at temperatures around 200° C the biomass is decomposed in non-condensable gases (CO_2, H_2, H_2O, CH_4 and other light hydrocarbons), condensable hydrocarbons (tars and oils) and char. At temperatures in excess of 600°C, the volatile pyrolysis products undergo secondary gas-phase reactions that most closely resemble the hydrocarbon cracking reactions used in the petrochemical industry for manufacturing ethylene and propylene. In the final step, the char reacts endothermically with steam at temperatures above 700°C to produce mainly CO and H_2.

Most of the gasifier concepts tried or proposed for use with biomass are modified versions of coal gasifiers. Biomass is more reactive than coal, therefore high gasification efficiencies can be attained at temperatures lower than those required for coal. The heat needed in a gasifier can be supplied in two ways: directly by partial oxidation of the feedstock or indirectly through a heat-exchange mechanism. Directly heated gasifiers can be divided in three types: fixed, fluidised and entrained beds.

In a *fixed-bed gasifier*, the feedstock enters at the top of the reactor and sequentially undergoes drying, pyrolysis, char gasification and char combustion. Fixed-bed gasifiers produce a high fraction of condensable gases. Much of the energy content of these gases can not be recovered if the product gas is used for methanol or hydrogen production. Fixed-bed gasifiers are therefore better suited for 'close-coupled' processes where the hot product gas is burned directly.

In *fluidized-bed gasifiers* the feedstock usually enters through the sidewall of the reactor. The feed mixes in a bed containing an inert material such as sand or a catalytic material such as dolomite. The fluidised state is maintained by the injection of steam and oxygen from below. Since the average bed temperature is higher than in a fixed bed, the product gas contains fewer tars and oils. In this study the oxygen-blown, pressurised bubbling fluidised-bed gasifier designed by the Institute of Gas Technology (IGT) is taken as a representative and well-tested fluidised-bed gasifier.

Entrained-bed gasifiers were originally developed for coal and there is no operating experience with biomass. In entrained-bed coal gasification, pulverised coal is fed into the reactor dry or in a water slurry, where it reacts with a large amount of oxygen. The resulting high operating temperatures (1300-1800°C) completely gasify the coal and produce a tar-free gas which is virtually free of CH_4 and higher hydrocarbons, which makes it very well suited as syngas for methanol production. There are two reasons why there has been little interest in entrained-bed gasification of biomass: biomass does not require such high peak temperatures and grinding biomass to the particle size (125-600]m) required for entrained-bed gasification is capital and energy intensive. Drying biomass from 15% to 8% moisture and grinding it from 3cm to 1.5mm required an additional investment of 16.2 million 1992US$ for 10 t/h wood in the biomass co-firing project of the Dutch utility EPON. This process requires 7.5% of the energy content (van den Broek *et al.* 1995). In the context of this project entrained-bed gasifiers might be of interest for combined gasification of coal and biomass. The Shell pressurised entrained-bed gasifier is selected in this study because it operates using dry feed and uses relatively little oxygen (Katofsky 1993).

Indirectly heated gasifiers operate at much lower temperatures than the directly heated types and therefore the product gas contains significant quantities of hydrocarbons (mostly methane) and some tars. The Batelle-Colombus Laboratory (BCL) gasifier is an atmospheric twin bed fast-fluidised bed unit that resembles fluid catalytic crackers commonly used in the petrochemical industry. In the first sand bed, biomass is pyrolysed in steam at temperatures up to 930°C. Ash, char and sand are separated from the product gas using a cyclone and are sent to a second bed where the char is burned in air to reheat the sand. The heat is transferred between the two beds by circulating the hot sand back to the gasification bed (Katofsky 1993).

Williams et al. (1995) describe the main characteristics of the IGT, BCL and Shell gasifier.

2. Conditioning of synthesis gas

2.1 Removal of particles

For the removal of particles various types of equipment may be used.

Mechanical separators use natural gravitational forces to remove the particles from the off-gas streams. Examples of mechanical separators are gravity settling chambers, cyclones and multi-cyclones. Cyclones are the most widely applied type of particle collection equipment. The particle containing gas enters the cyclone in a tangential way. As a result of the vortex created by the cyclone, the particles are submitted to forces of 5 to 100 times gravity, as a result of which the particles are separated and removed. Cyclones are suited for removal of particles higher than 3-5 mm, while at particles sizes from 1 to 5 mm multi-cyclones might be used above 200 mm. Mechanical separators are operated at temperatures as high as 1000 °C and pressures up to 500 bar (Perry). Experiences of Shell indicate however, that use of cyclones after coal gasification at temperatures higher than 400 °C results in the sintering of particles to larger abrasive particles (Gerrits et al., 1994).

Fabric filters consist of a filter-bags, contained in a baghouse. Usually some kind of cleaning mechanism is used to remove the dust from the bags surface. Examples of filter materials are natural materials such as cotton and wool, plastics based bags such as nylon, polypropene and teflon or mineral-based bags, e.g. fiberglas. Each material has its specific operating conditions, stability in the reducing media and minimum particle size they are able to remove. The maximum temperature depends on the type of cloth used. Natural materials can be used up to 60-80 °C; plastic-based materials can deal with temperatures up to 100-200 °C, where fiberglas may be used up to 225 °C (Vatavuk, 1990). Metal-sheet bags might even be used to temperatures of 450 to 600 °C (Gerrits, 1994). Due to the sheet metal construction of the baghouse, the operating pressure of most fabric filters is limited to near atmospheric conditions (Vatavuk, 1990).

In *electrostatic precipitators* (ESP) particles become charged. As a result of the electrostatic field, particles migrate to collection plates where they are removed from the gas phase. When properly designed, ESPs are effective for particles, with sizes as small as 0,1 mm. ESPs are normally operated at atmospheric pressures and temperatures up to 300-400 °C. At elevated pressure, ESPs might be operated at even higher temperatures 20 bar, 700 °C (Gerrits et al., 1994) and 50 bar, 800 °C (Perry) are reported in literature

In *wet scrubbers*, particles are removed from off-gases, in a collision with droplets or with the water surface. There are many types of wet-scrubbers, ranging from spray towers to cyclone scrubbers, packed bed scrubbers and venturi-scrubbers. The pressure-drop of the gas phase is an indicator of the intensity of contact of the gas and the liquid phase and of the minimal size of particles that are separated. Pressure drop in spray towers is generally low, resulting in separation of only the larger particles (> 5-10 mm), where pressure drop of e.g. venturi-scrubbers is rather high, resulting in separation of particles larger than 0,5 mm.

Ceramic filters might be used for off-gas cleaning at high temperatures. Tube-filters and cross-flow filters are commercially available for cleaning gases at high temperature (800-1000 °C) and moderate pressure (10-20 bar) (Gerrits et al., 1994).

2.2 Cl-, and F-removal

- *wet systems*

Halogens might be removed in a wet-scrubber, using a neutral or slightly caustic aqueous solution. The effluent is usually not regenerated, but is obtained as a waste stream. Such a wet system might be integrated with a wet scrubber for removal of particles. An example of the latter option is the use of a venturi-scrubber (Raas, 1990). Neutral wet systems remove as well part of the NH_3 from the flue-gas.

- *dry systems*

dry systems are based on the injection of mild soda (Raas, 1990), the subsequent formation of Na-salts and ultimately the separation of these salt-particles. Soda injection might be located before or after the removal of particles. The advantage of injection before particle removal is an integration of particle and halogen-removal. The possibilities for utilisation of ashes, however, is dramatically reduced.

2.3 Water-Gas Shift Reaction

The water-gas shift reaction is used to adjust the ratio between CO and H_2 in synthesis gas.

$$CO+H_2O=CO_2+H_2$$

The degree to which the shift reaction is used depends on the desired end-product. If hydrogen is the final product, as much CO as possible is shifted. For methanol synthesis, the shift reaction is performed till the H_2:CO ratio is obtained, preferred by kinetic considerations and requirements for cartalyst deactivation (Ladebeck, 1993).
The shift reaction is exothermic, higher conversion levels are achieved at lower temperatures. The reaction is independent of pressure and is often operated at elevated pressures to accomodate upstream and downstream pressure requirements. Shift reactors have a simple design because they do not require external heating and operate at low temperatures. The water-gas shift is catalysed by e.g. a cobalt-molybdate catalyst, which is not deactivated in the presence of sulphur. To prevent coking problems and to ensure good conversion, steam is usually added to the feed gas to maintain a steam:carbon ratio of at least 3:1.

2.4 Carbon dioxide removal

Carbon dioxide separation and subsequent underground storage is often mentioned as a way for reducing greenhouse gas emissions. The carbon dioxide may be removed from flue gases or from fuel gases, e.g. from a gasification plant.

Carbon dioxide separation is a well established unit operation in industrial chemistry and is part of ammonia and hydrogen production. For these purposes in most cases chemical absorption is used. For other applications (e.g. recovery from CO_2-rich natural gas, upgrading of biogases) and also for the specific purpose of mitigating greenhouse gas emissions other methods for carbon dioxide are developed as well. At the moment a number of options exist for carbon dioxide separation:
- chemical absorption, e.g. in MDEA or some other amine;
- physical adsorption, e.g. in water, methanol, Selexol or Purisol;
- gas-separation membranes;
- gas-absorption membranes;
- adsorption processes, e.g. PSA;
- cryogenic processes

For carbon dioxide recovery from fuel gases from coal gasification, physical adsorption, using selexol is preferred (Hendriks, 1995), where at lower carbon dioxide partial pressures chemical absorption seems to be more effective (Blok, 1993). PSA and gas-separation membranes might be a feasible option, when small gasstreams have to be treated. Gas-adsorption membranes might become an interesting alternative to absorption processes, because they combine advantageous properties of chemical or physical adsorption with the compactness and flexibility of membrane units (e.g. Feron and Jansen, 1995).

For storage of carbon dioxide several options exist (Blok, 1993):
- it might be stored in empty oil or gas fields at pressures of 80 to 110 bar;
- it might be stored in aquifers at about the same pressures;
- ocean disposal at depths of over 1500 m is a third option, especially when deep ocean is present near the main land (as it is the case e.g. in Japan);
- besides several options for reuse of carbon dioxide are discussed in literature, e.g. use in greenhouses to enhance plant growth and various applications in chemical industry. It has to be noted that in many cases reuse is either rather energy consuming (so emitting CO_2) or results in a only short-term disposal, and does not contribute to reduction of carbon dioxide concentrations on medium or long terms (Oonk and Heslinga, 1996).

3. Utilisation of synthesis gas

3.1 H_2-separation

Pressure swing adsorption (PSA) is an alternative for other ways of carbon dioxide removal, especially when high purity hydrogen is needed (> 99.9 vol% pure H_2). The working mechanism of PSA is based on the selective adsorption of molecules as CO, CO_2, CH_4 and H_2O, compared to H_2 on e.g., activated carbon or molecular sieves. A PSA-unit consists of four or more adsorber columns. Each column is op-

erated in a cyclic mode, consisting of (1) adsorption, (2) cocurrent depressurisation, (3) countercurrent blowdown, (4) purge, or (5) repressurisation). The product recovery for four-bed systems is about 70-75%, where product recoveries for poly-bed processes might be up to 80-85%. The remainder of the hydrogen is lost in the CO_2-CO-effluent (Yang, 1987).
PSA is industrially applied for hydrogen removal from products of natural gas steam reforming (containing about 70-75% H_2); removal of hydrogen from the catalytic reformer effluent (65-85% H_2), or the ethylene plant effluent (70-90% H_2). Maximum capacities are about 1 to 2 million Nm^3 H_2 per day (Cirmac, 1996).

3.2 Methanol synthesis

3.2.1 Gas phase process

Methanol is one of the most organic chemicals. Its world-wide production capacity is about 21 million tonne per year (1989) (Fiedler et al., 1995). Its main use is in the production of formaldehyde and the subsequent production of resins.
Methanol is produced from synthesis gas at temperatures of about 200-300 °C, pressures in excess of 50 bar and using a nickel-based catalyst. The overall reactions are:

$$CO + 2\,H_2 \rightarrow CH_3OH$$

$$CO_2 + 3\,H_2 \rightarrow CH_3OH + H_2O$$

The mechanism of methanol synthesis is not clear. Until the beginning of the 1980s it was assumed that methanol was produced on the catalyst surface after adsorption of CO, but recent insight suggest that CO_2-adsorption is of importance as well [ref.]. Recent experiments show that both pathways are possible. For methanol, production a ratio of $CO:H_2=2$ and a ratio of $CO_2:H_2=3$ is required. Both prerequisites can be combined in one stoichiometry-ratio:

$$C = ([H_2] - [CO_2]) / ([CO] + [CO_2])$$

Both methanol producing reactions are equilibria, as a result of this, conversion of synthesis gas to methanol is limited to about 20% per pass. Normally methanol and water are separated after reaction, and the remaining synthesis gas is recycled to the methanol reactor feed (see figure *). A purge, containing CH_4, H_2 and some CO and CO_2 is normally fed to the steam-reformer furnace. Water in the synthesis gas shifts the equilibria to the synthesis gas side. For this reason, the synthesis gas is dried before methanol production. Methanol synthesis is highly exothermic. The reaction heat is used for production of steam. Methanol make-up consists of a two-step distillation: in the first step, light side-products are removed from the methanol-water mixture, in a second distillation step methanol is obtained.

3.2.2 'Once through' methanol (OTM)

In the standard methanol production process the synthesis gas remaining in the reactor effluent is recycled to the reactor after recompression. In 'once-through' methanol production the remaining synthesis gas is not recirculated but used for other purposes such as the generation of electricity. This process configuration eliminates the necessity for excess CO_2 removal recycle gas compression and in the case biomass and/or coal gas shift conversion. Also, in the case of combination with electricity generation, maximum use can be made of the heat liberated during methanol production by raising intermediate and low pressure steam which can be efficiently utilised in the power plant's steam system as injection steam in the gas turbine exhaust.

A high-pressure liquid phase methanol process seems most suited for a 'once-through' configuration because it gives a higher conversion per pass.

3.3 Fischer-Tropsch synthesis/SMDS

The Fischer-Tropsch synthesis, was developed in Germany in the 1920s and 30s as a part of indirect coal liquefaction (production of hydrocarbon fuels from coal). In war-time Germany the Fischer-Tropsch process was essential to fuel Hitlers war-machine. After the second world war, Fischer-Tropsch technology was mainly further developed for coal liquefaction in the 1950s to 1970s in South-Africa, and more recently by Shell to convert natural gas to higher hydrocarbons.

The main reaction of the Fischer-Tropsch process is the conversion of synthesis gas into a mixture of alkanes, olefins and alcohols, according to:

$$2n\,H_2 + n\,CO \rightarrow C_nH_{2n} + n\,H_2O$$

$$(2n+1)\,H_2 + n\,CO \rightarrow C_nH_{2n+1} + n\,H_2O$$

$$2n\,H_2 + n\,CO \rightarrow C_nH_{2n+1}OH + (n-1)\,H_2O$$

Fischer-Tropsch reaction is promoted by iron-based or supported, alkali-promoted catalysts; operating conditions of the Fischer-Tropsch process are temperatures of 220-340 °C, pressures of about 25 bar.

The Fischer-Tropsch reaction yields a mixture of alkanes, olefines and alcohols of different chain lenghts. A products work-up, consisting of destillation, catalytic reforming, alkylation, catalytic polymerisation, C5-C6-isomerisation, hydrogenation or hydrocracking is required for transformation of the Fischer-tropsch product into gasoline and other products. Two types of processes can be distinguished: a mixed output process and an all liquid process. In a mixed output process about half of the product consists of methane rich fraction. In an all liquid output process this methane is reformed to synthesis gas and fed back again to the Fischer-Tropsch reactor bed.

3.4 Combined-Cycle Power Generation

A combined-cycle power generation system consists of a gas turbine generator, a heat recovery steam generator (HRSG) and steam turbine generator. The fuel gas is heated and saturated with water vapour in a saturation system and further heated against circulating hihg-pressure boiler feed water from the HRSG. The lower heating value obtained by this process inhibites the formation of thermal NO_x in the gas turbine. Next, the humidified fuel gas is introduced in the gasa turbine combustor along with air. The hot gas exiting the combustor is supplied to the gas turbine expander, which in turn drives the generator. The turbine exhaust gases flow to the HRSG, which recovers the heat from these gases in the form of superheated high-pressure steam and reheated intermediate pressure steam. This steam is lead to the steam turbines, which are driven by expansion of the steam.

Fluor Daniel (1988) describes a 693 MW_e combined-cycle power system consisting of two General Electrics MS9001F gas turbine generators, two HRSGs and a single reheat steam turbine generator.

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Appendix 2 Lay-out of the synthesis gas production, shift and carbon dioxide removal

Combined synthesis gas production, water-gas shift and carbon dioxide removal might be integrated in various ways. The options are described below. Each option consists of a certain integration of unit-operations. All unit-operations are described in more detail in Appendix 1. In the Figures below various process schemes are depicted. These schemes aim to clarify the options for process integration. For reasons of simplicity, the various aspects of synthesis gas generation (e.g. (i) multi-feedstock gasifier versus two gasifiers in the case of coal/oil residue feedstock, (i) oxygen supply, (ii) BCL-gasification and (iii) compression to 30 bar in the case of or (i) oxygen generation, (ii) steam supply and (iii) autothermal reforming) are left out.

- *option I (Figure A2.1)*

Biomass, and maybe oil residues and coal are converted in synthesis gas in one or two gasifiers. The product of biomass gasification contains a considerable amount of methane and higher hydrocarbons (depending on technology used, up to 10 vol% - see Appendix 1). The conversion to synthesis gas is increased by reforming of the product gas, together with the natural gas in a steam-reformer or an autothermal-reformer at 30 bar, 950 °C. Carbon dioxide concentrations in the product gas are increased by water-gas shift and subsequently water is removed by cooling of the product gas. After compression (e.g., to about 80 bar: the operating pressure of methanol synthesis) carbon dioxide is removed. Since only small part of the carbon oxides have to be removed, carbon dioxide removal is relatively easy.

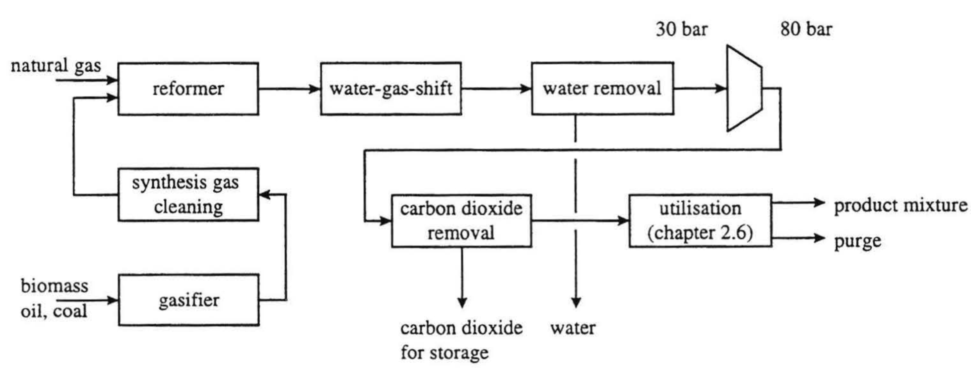


Figure A2.1 The MethaHydro process, option I (heat integration not shown)

- *option II*

When only biomass and natural gas are used and a IGT-gasifier is applied, the amount of carbon dioxide after gasification exceeds the amount of carbon dioxide that has to be removed and stored, in order for the process to be CO_2-neutral. So an alternative to option I comprises carbon dioxide without water-gas shift from the gasification product. The product gas is fed to the reformer, dried and compressed, similar as in option I.

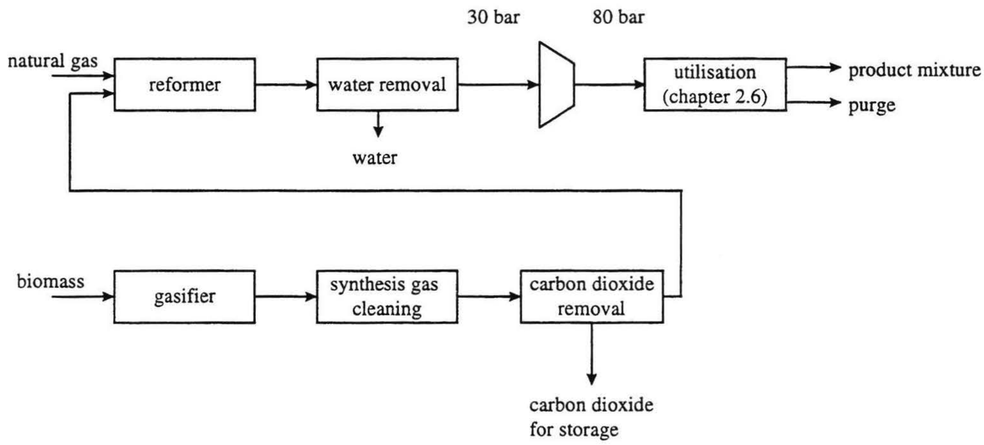


Figure A2.2 The MethaHydro process, option II (heat integration not shown)

- *option III*

Biomass, maybe oil residues and coal are gasified, and after the product gas is cleaned it is mixed with the product gas from the methane reformer. The mixture is shifted, dried and compressed, after which the carbon dioxide is removed.[1] The excess methane and higher hydrocarbons act as an inert in utilisation and end up in the utilisation purge.

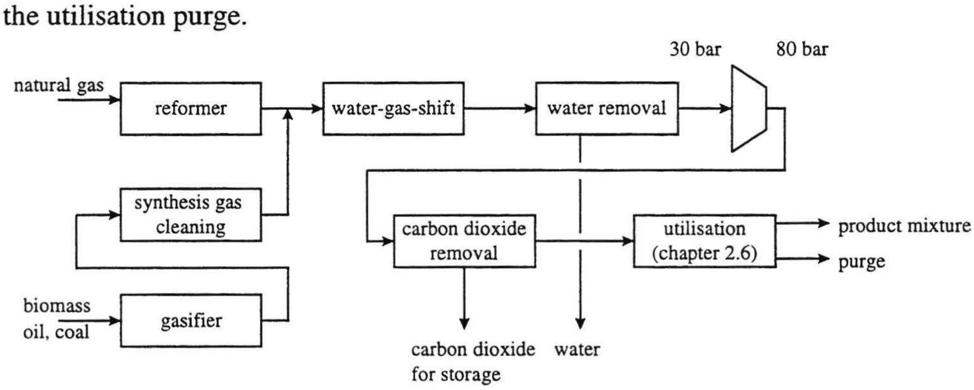


Figure A2.3 The MethaHydro process, option III (heat integration not shown)

1 Water-gas shift and carbon dioxide removal before mixing is an alternative. However, carbon dioxide of the combined stream after compression seems to have advantages: (i) the gas-steam to be treated is smaller in volume; (ii) carbon dioxide partial pressures are higher; (iii) carbon dioxide is obtained at higher pressures. So removal before mixing isn't considered here.

- *option IV*

When biomass and natural gas are used and an IGT-gasifier is applied, an alternative to option III might be removal of carbon dioxide from the gasifier product gas, without a shift required.

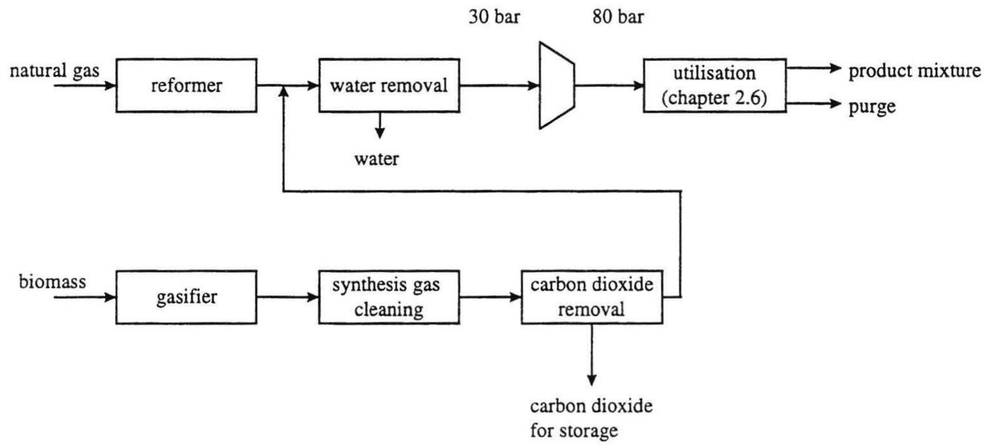


Figure A2.4 The MethaHydro process, option IV (heat integration not shown)

- *option V*

Synthesis gas generation as in option I or II, after which the synthesis gas is dried and compressed. After utilisation, the purge gas is subjected to a water-gas shift and the carbon dioxide is removed. Since a synthesis gas has to be utilised with a low-stoichiometric ratio, this option is only viable in combination with a 'once through' methanol process.

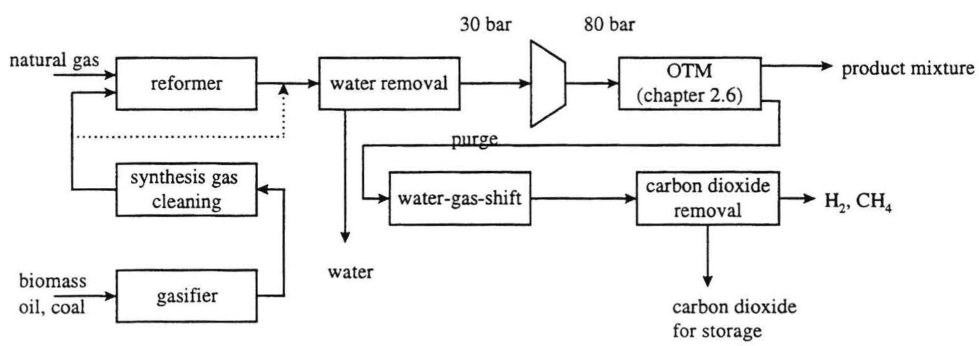


Figure A2.5 The MethaHydro process, option V (heat integration not shown)

- *option VI*

As in option V, only the purge gas is used as a fuel-gas for a STEG installation, after which carbon dioxide is removed from the exhaust stack.

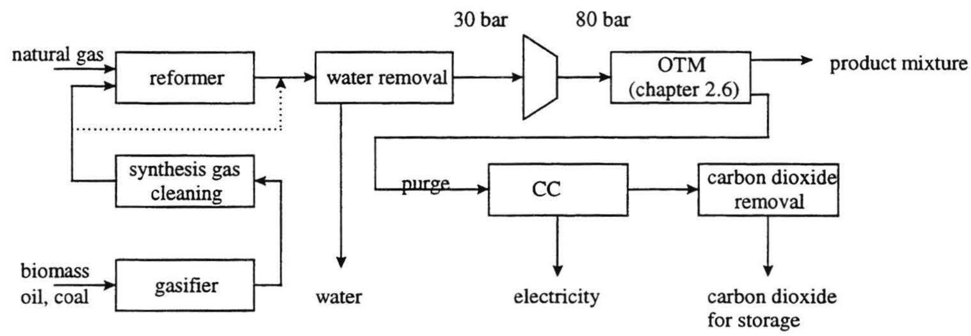


Figure A2.6 The MethaHydro process, option VI (heat integration not shown)

• *option VII*

Hydrogen is produced from natural gas, as nowadays industrially performed, and the product hydrogen is added to the biomass gasifier mixture to adjust the stoichiometric ratio of the synthesis gas. This process is only suited for a combination of biomass and natural gas. When coal or oil residues are used as well, the CO_2-neutrality of the process and product can not be achieved.

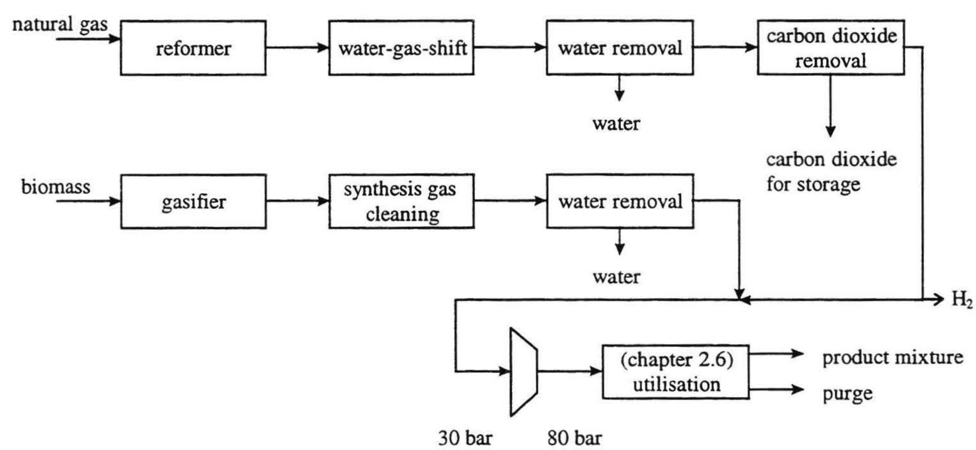


Figure A2.7 The MethaHydro process, option VII (heat integration not shown)

• *evaluation*

The choice of the most efficient method to prepare the synthesis gas, and to combine it with water-gas shift and carbon dioxide separation depends a.o. on the choice of resource mix to be used, the utilisation options chosen and the flexibility of input- and product-mix required. Besides that the ease of carbon dioxide removal and overall conversion and costs are relevant criteria.

• *use of fossil resources*

Table A2.1 indicates to what extent the process is suited for various mixes of fossil fuels.

Table A2.1 Suitability of various processes for mixes of fossil fuels.

	Fossil resource
option I	suited for every mix of fossil fuels
option II	only suited for natural gas
option III	suited for every mix of fossil fuels
option IV	only suited for natural gas
option V	suited for every mix of fossil fuels
option VI	suited for every mix of fossil fuels
option VII	only suited for natural gas

- *utilisation options*

Table A2.2 indicates to what extent the various processes are suited for the various utilisation options.

Table A2.2 Suitability of various processes for utilisation options.

	utilisation option
option I	best suited for production of methanol or Fischer-Tropsch synthesis
option II	b est suited for production of methanol or Fischer-Tropsch synthesis
option III	best suited for production of methanol or Fischer-Tropsch synthesis
option IV	best suited for production of methanol or Fischer-Tropsch synthesis
option V	best suited for once-through methanol in combination with CH_4/H_2 or electricity production
	best suited for once-through methanol in combination with electricity production
option VI	best suited for mixed methanol or Fischer-Tropsch synthesis and H_2-production
option VII	

- *product-mix*

Table A2.3 shows the typical product mix of the process, and indicates what options exist for flexibilities exist in this product mix.

Table A2.3 Typical product mixture and flexibility towards product mixture.

	product mix	**flexibility**
option I	90% MeOH; 10% CH_4/H_2 1)	larger amounts of CH_4/H_2 possible by (i) increasing amount of CH_4 reformed, (ii) increasing the amount of CO_2 recovered and (iii) reducing the methanol synthesis recycle
option II	80% MeOH; 20% CH_4/H_2 1)	slightly larger amounts of CH_4/H_2 possible by reducing the methanol synthesis recycle
option III	90% MeOH; 10% CH_4/H_2 1)	larger amounts of CH_4/H_2 possible by (i) increasing amount of CH_4 reformed, (ii) increasing the amount of CO_2 recovered and (iii) reducing the methanol synthesis recycle
option IV	80% MeOH; 20% CH_4/H_2 1)	slightly larger amounts of CH_4/H_2 possible by reducing the methanol synthesis recycle
option V	20% MeOH; 80% CH_4/H_2 2)	negligible
option VI	20% MeOH; 80% electricity	negligible
option VII	MeOH, H_2 in any ratio desired	full flexibility

1) instead of 'MeOH', it is possible to read 'gasoline/diesel'; instead of 'CH_4/H_2' it is possible to read 'electricity'

2) instead of 'CH_4/H_2' it is possible to read 'electricity'

- *synthesis gas preparation*

Synthesis gas is prepared in a combination of gasifier and reformer. In normal methanol production, costs of synthesis gas production are about 30% of the total capital costs (Williams et al., 1995; van Dijk et al., 1995). So the size of gasifier and reformer are of interest. The size of both gasifier and reformer is about the same for all options, except for option I and II, where the product stream from the gasification plant is treated in the reformer as well. As a result of this, the reformers size will increase by 50%, and its costs will increase by 30%. Total investment costs of the MethaHydro-process will increase by about 10%.

- *compression*

Capital and operating costs for compression are large in methanol synthesis (van Dijk *et al.* 1995) So the volume of the stream that has to be compressed to 80 bar determines large part of the costs of the MethaHydro process. These costs are minimised in option VII and are relatively large (compared to the amount of methanol that is produced) in options V and VI.

- *carbon dioxide removal*

Carbon dioxide removal will comprise significant part of the capital and operating costs of methanol synthesis. For this reason, systems are preferred in which carbon dioxide separation and storage is relatively efficient. These systems may have the following characteristics:

- absorption at high pressure reduces equipment sizes and enables desorption at high pressure as well;
- high carbon dioxide partial pressure (product of concentration and total pressure)
- low degree of separation. 100% separation requires high capital and operating costs, compared to e.g., 60% separation

Table A2.4 shows the characteristics of carbon dioxide recovery.

Table A2.4 Characteristics carbon dioxide recovery with chemical adsorption for the various options.

	pressure (bar)	CO_2-partial pressure (bar)	efficiency of separation[2] (%)
option I	80	15[1]	65[1]
option II	30	10	85
option III	80	15[1]	65[1]
option IV	30	10	85
option V	80	30[1]	75%
option VI	atm.	0.1	>95
option VII	30	6	>98

1) depending on conversion of water-gas shift

2) the fraction of the carbon dioxide that has to be separated

From Table A2.4 it can be read, that in option I, option III and V, carbon dioxide separation and recovery is expected to be relatively cheap. Carbon dioxide becomes available at pressures of 80 bar, so little extra energy is required for compression to 110 bar for storage in aquifers. In system VI and VII carbon dioxide recovery will be expensive.

Conclusions

The preferred option for synthesis gas generation, water-gas shift and carbon dioxide recovery, depends on a number of choices:
- the scale of the process;
- the mix of fossil resources used;
- the mix of products required and its flexibility

small-scale process
When a small-scale process is preferred. Option I, III or V might be applied. Option I has higher capital costs than option II, but the overall conversion to methanol is 10% higher as well, so this process seems to be most attractive on small-scale.

large-scale processes, natural gas based
For these types of processes, option I, II, III and IV might be applied. For the same reason as above, option I and II are preferred over option III and IV. The advantage of II over I is the absence of a water-gas-shift reactor. A disadvantage is a somewhat more difficult carbon dioxide separation and storage. A choice between I and II requires more elaborate calculations, that can be avoided when the steering committee chooses not to use a process, based on natural gas only.

large-scale processes, based on a mix of fossil fuels
For this type of process, options I and III might be used. Again, option I is preferred.

large-scale processes, natural gas based, flexible output
For this type of process option VII is the most attractive one.

Appendix 3 Several possible MethaHydro-processes

small-scale process

When a small-scale process, about 200 tpd methanol (see chapter 2.2 an evaluation of the scale of the process), is preferred, the most attractive MethaHydro-processes are those depicted in Figure A3.1 and A3.2.

The first steps in these processes are the same. Biomass is gasified in an indirectly heated gasifier (1), cleaned (2 and 3) and the product gas is subsequently compressed to 20 bar (4). Natural gas is added and the mixture is converted in a steam-reformer at 20 bar, 870 °C.
Next, in the configuration using traditional methanol synthesis (figure A3.1), the synthesis gas mixture is shifted (6), excess water is removed (7and 8)and the mixture is compressed to 80 bar (9). Subsequently carbon dioxide is removed (10 and 11). Since the steam reformer is fueled with natural gas (resulting in carbon dioxide emissions), the amount of carbon dioxide that is removed here, exceeds the amount of carbon in the natural gas process-feed. After carbon dioxide removal the synthesis gas is fed to a conventional methanol synthesis (12). Product methanol a remaining syngas are separated (13). The syngas is recompressed and recycled to the methanol reactor. The methanol water mixture is fed to a 2-step distillation. The purge of the methanol synthesis contains mainly containing hydrogen and methane with small amounts of carbon oxides, that are converted with excess hydrogen to additional methane (14). Ultimately a methane/hydrogen mixture is obtained, suited for injection in the natural gas distribution grid. The steam from the methanol synthesis is partially used in the process (for thriving compressors and in the methanol/water distillation) and may partially be exported.

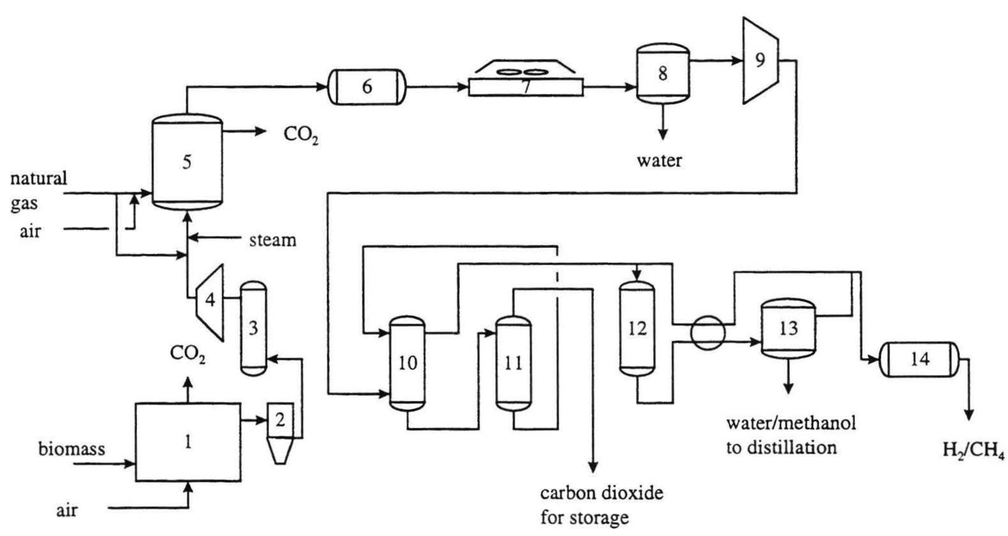


Figure A3.1 Small scale MethaHydro-process, using traditional methanol synthesis

In the 'Once-Through' configuration (Figure A3.2), after the reformer (5), water is removed from the syngas (6 and 7), the gas is compressed to 80 bar (8)and fed to the liquid phase methanol reactor (9). The product methanol is separated from the remaining syngas (10) and sent to the distillation unit. The synthesis gas is shifted (11) and carbon dioxide is removed from the remaining syngas (12 and 13). The resulting gas can be sent to a PSA-unit (14) to produce pure hydrogen.

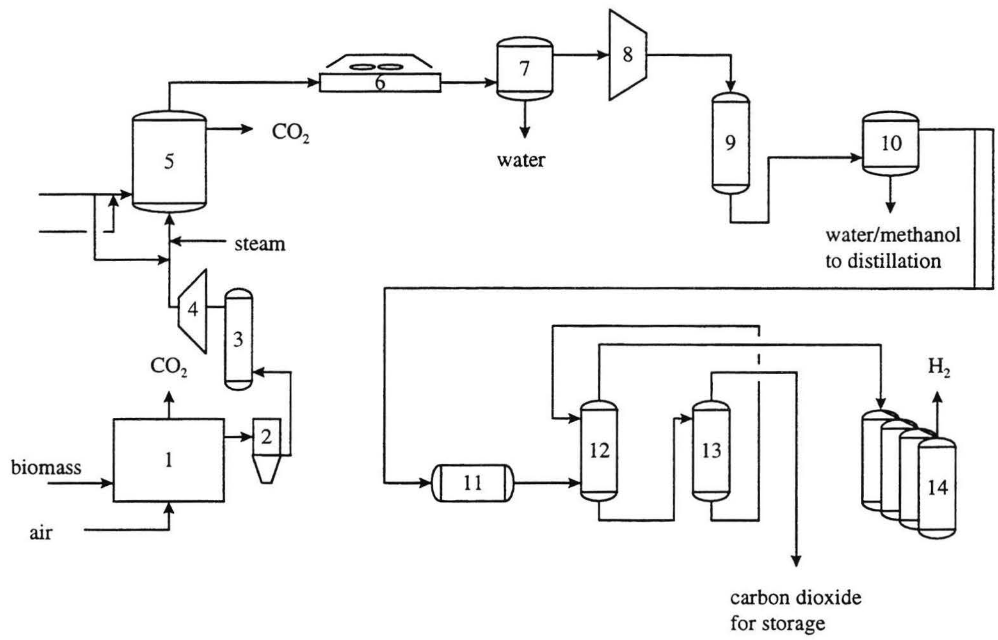


Figure A3.2 Small scale MethaHydro-process, using 'Once-Through' methanol synthesis

large-scale processes, natural gas based
For these types of processes, the most attractive MethaHydro-processes are the ones depicted in Figure A3..3 or A3.4.

In Figure A3.3, oxygen is produced and compressed (1 and 2) and used in biomass gasification in a IGT-gasifier (3) at a pressure of 30 bar. Particles, tars and possibly halogens are removed from the gas (4 and 5), after which carbon dioxide is removed (6 and 7). The product gas is mixed with natural gas and fed to an autothermal reformer at 30 bar (8), water is removed (9 and 10) and the synthesis gas is compressed to 80 bar (11) for methanol synthesis (12 and 13). The methanol purge is after methanation (14) injected in the natural gas grid.

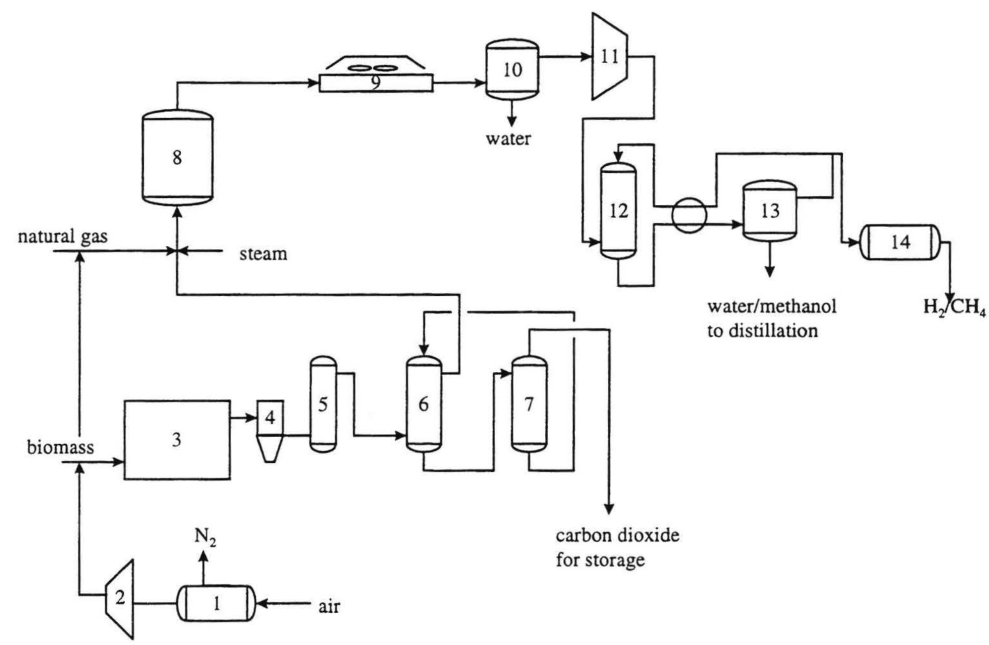


Figure A3.3 Large-scale, natural gas based MethaHydro-process

In Figure A3.4, biomass is gasified in a IGT gasifier and the coal/oil residues in a Texaco entrained-bed gasifier or all these feedstocks are gasified in one single gasifier (3). Particles, tars, sulphur and possibly halogens are removed from the gas (4 and 5) and is mixed with natural gas and fed to an autothermal reformer (6). The product gas is shifted (7) to increase concentrations of carbon dioxide to appropriate levels, after which water is removed (8 and 9). The product gas is compressed (10) and excess carbon dioxide is removed at 80 bar (11 and 12). The synthesis gas is subsequently used as described above at figure A3.3 (13 - 15).

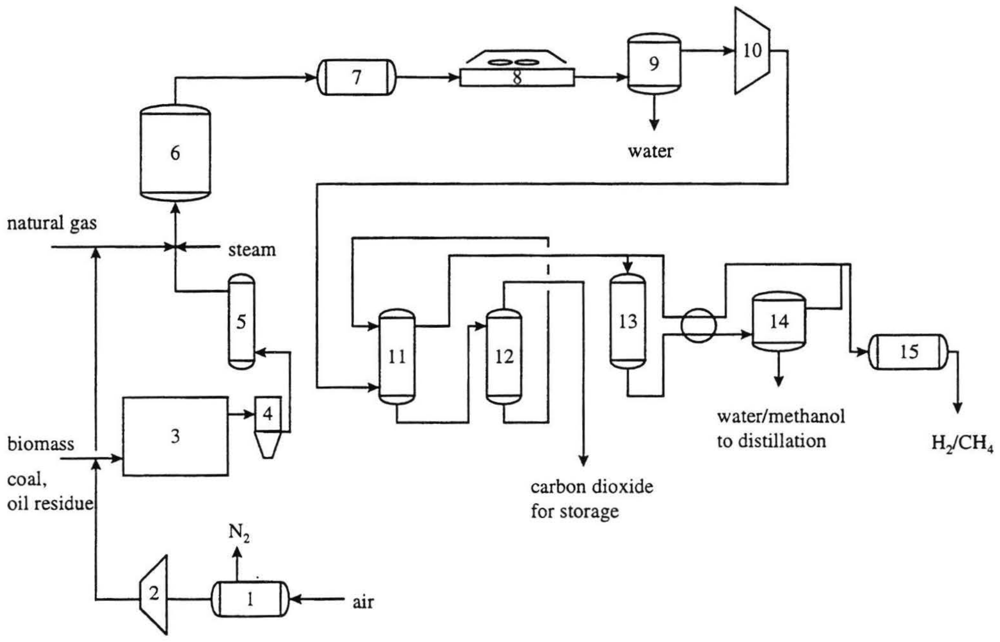


Figure A3.4 Large-scale, MethaHydro-process, based on a mix of fossil fuels

large-scale processes, based on a mix of fossil fuels
For this type of process the most attractive MethaHydro-process is depicted in Figure A3.4, and described above

large-scale processes, natural gas based, flexible output
For this type of process the most attractive MethaHydro-process is depicted in Figure A3.5. Natural gas is converted to hydrogen in the traditional way, by reforming (2), water-gas shift (3), water (4 and 5) and CO_2-removal (6 and 7). The product hydrogen is used to adjust the stoichiometric ratio of the gasifier product gas 8-12), after which the mixture is compressed (13) fed to a methanol synthesis (14 and 15), as decribed at Figure 3.3. The purge is subjected to methanation (15) and may be injected in the natural gas grid.

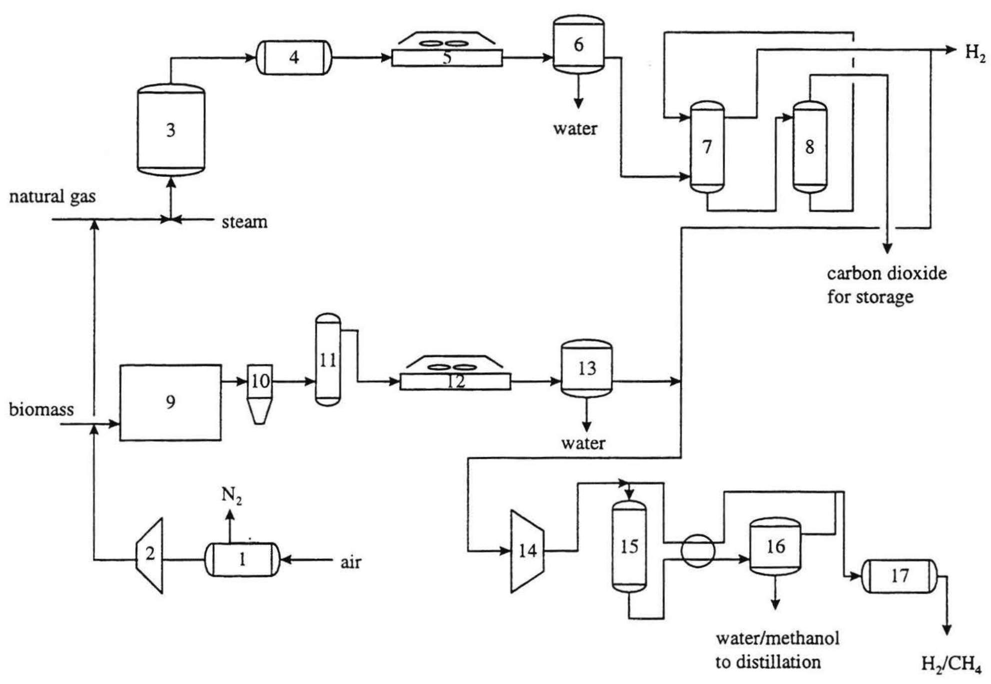


Figure A3.5 Large-scale, natural gas based MethaHydro process with flexible output

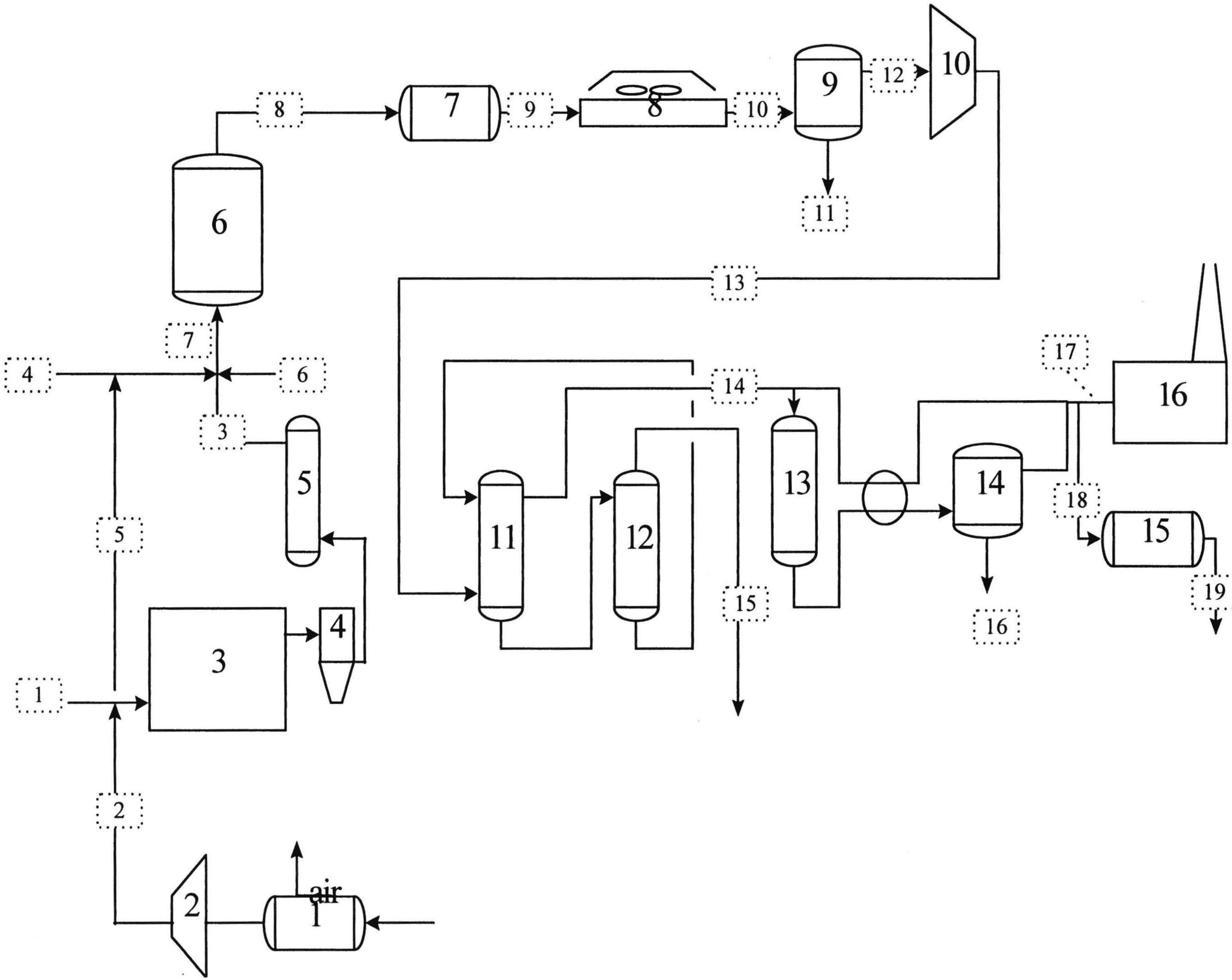
Appendix 4 Material and energy balances

Energetic contents[3)]	
- natural gas	32 MJ m^{-3}
- coal[1)]	27 MJ kg^{-1}
- oil residues[2)]	34-38 MJ kg^{-1}
- methane	35,8 MJ m^{-3}/53 MJ kg^{-1}
- hydrogen	10,8 MJ m^{-3}/121 MJ kg^{-1}
- methanol	20,0 MJ kg^{-1}; 15,8 MJ l^{-1}
- gasoline	44 MJ kg^{-1}; 32 MJ l^{-1}
- diesel	42,5 MJ kg^{-1}; 35,6 MJ l^{-1}
- specific weight methanol	0.79 kg l^{-1}

1) CBS (1996)

2) (Emsperger and Kerg 1996)

3) based on lower heating values



1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
air

MethaHydro M&E balance

Conversion to MeOH: 90%

a) mole-flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomass	(kmole/s)	0,69																			
CH4	(kmole/s)				0,12	0,46			0,58	0,03	0,03	0,03		0,03	0,03	0,03			0,02	0,01	0,01
C2+	(kmole/s)				0,01				0,01												
O2	(kmole/s)		0,17				0,33		0,33												
H2O	(kmole/s)	0,17		0,18	0,46			0,63	1,09	1,13	0,99	1,13	1,13								0,00
CO	(kmole/s)				0,21				0,21	0,73	0,59	0,59		0,59	0,59	0,59		0,07	0,01	0,00	
CO2	(kmole/s)				0,34				0,34	0,39	0,53	0,53		0,53	0,53	0,07	0,46				
H2	(kmole/s)				0,30				0,30	1,39	1,53	1,53		1,53	1,53	1,53			0,11	0,06	0,05
MeOH	(kmole/s)																	0,65			
total	(kmole/s)	1,86	2,17	0,18	4,45	4,46	0,33	0,63	2,86	3,67	3,67	3,81	1,13	2,68	2,68	2,22	0,46	0,72	0,14	0,07	0,07

pressure	(bar)		30	30	30	30	30	30		30	30	30	30	30	80	80	80	80	80	80	80
temp.	(oC)				20	20	20	350		950	420	15	15	15	60	15	15	15	300	300	15

b) mass-flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomassa	(kg/s)	54,22																			
CH4	(kg/s)				1,87	7,36			9,23	0,46	0,46	0,46		0,46	0,46	0,48			0,32	0,16	0,21
C2+	(kg/s)				0,21				0,21												
O2	(kg/s)		16,27				10,47		10,47												
H2O	(kg/s)	9,54		16,27	8,31			11,25	19,53	20,34	17,82	20,34	20,34								0,06
CO	(kg/s)				5,98				5,98	20,44	16,52	16,52		16,52	16,52	16,52		1,96	0,19	0,09	
CO2	(kg/s)				15,15				15,15	17,16	23,32	23,32		23,32	23,32	3,08	20,24				
H2	(kg/s)				0,61				0,61	2,78	3,06	3,06		3,06	3,06	3,06			0,23	0,11	0,10
MeOH	(kg/s)																	20,83			
total	(kg/s)	63,77	16,27	16,27	32,12	7,36	10,47	11,25	61,18	61,18	61,18	63,70	20,34	43,36	43,36	23,14	20,24	22,79	0,73	0,37	0,37

c) volume flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
total	(m3)				1,08	0,34	0,24	0,99		11,43	6,48	2,80		1,97	0,85	0,61	0,13		0,08	0,04	0,02
total	(Nm3)				32,40	10,30	7,33	14,01		82,18	82,18	85,32		60,01	60,01	49,73	10,30		3,14	1,57	1,46
phase		s	g	g	g	g	g	g	g	g	g	g	l	g	g	g	g	l	g	g	g

gas composition

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
CH4	(vol %)				8,10	100,00			20,20	0,79	0,79	0,76		1,08	1,08	1,35			14,29	14,29	20,41
C2+	(vol %)				0,48				0,24												
O2	(vol %)		100,00				100,00		11,45												
H2O	(vol %)			100,00	31,90			100,00	37,98	30,80	26,98	29,67	100,00								5,10
CO	(vol %)				14,76				7,47	19,90	16,08	15,49		22,02	22,02	26,58		9,71	4,76	4,76	
CO2	(vol %)				23,81				12,05	10,63	14,45	13,91		19,78	19,78	3,15	100,00				
H2	(vol %)				20,95				10,60	37,89	41,70	40,17		57,11	57,11	68,92			80,95	80,95	74,49

MethaHydro M&E balance

Conversion to MeOH: 70%

a) mole-flows

		1	2 2a		3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomass	(kmole/s)	0,54																			
CH4	(kmole/s)				0,10	0,57			0,67	0,03	0,03	0,03	0,03	0,03	0,03	0,03			0,02	0,01	0,02
C2+	(kmole/s)				0,01				0,01												
O2	(kmole/s)		0,12				0,40		0,40												
H2O	(kmole/s)	0,13		0,25	0,34			0,96	1,30	1,33	1,06	1,06	1,06					0,08			
CO	(kmole/s)				0,16				0,16	0,70	0,43	0,43		0,43	0,43	0,43			0,01	0,01	
CO2	(kmole/s)				0,25				0,25	0,39	0,66	0,66		0,66	0,66	0,08	0,57				
H2	(kmole/s)				0,22				0,22	1,54	1,81	1,81		1,81	1,81	1,81			0,48	0,24	0,22
MeOH	(kmole/s)																	0,49			
total	(kmole/s)	0,67	0,12	0,25	1,08	0,57	0,40	0,96	3,02	3,99	3,99	3,99	1,09	2,93	2,93	2,35	0,57	0,58	1,53	0,77	0,71
pressure	(bar)		30,00	30,00	30,00	30,00	30,00	30,00		30,00	30,00	30,00	30,00	30,00	80,00	80,00	80,00	80,00	80,00	80,00	80,00
temp.	(oC)				20,00	20,00	20,00	350,00	virtueel	950,00	420,00	15,00	15,00	15,00	60,00	15,00	15,00	15,00	300,00	300,00	15,00

b) mass-flows

		1	2 2a		3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomassa	(kg/s)	13,15																			
CH4	(kg/s)				1,55	9,16			10,71	0,45	0,45	0,45	0,47	0,45	0,45	0,45			0,30	0,15	0,26
C2+	(kg/s)				0,15				0,15												
O2	(kg/s)		3,95				12,80		12,80												
H2O	(kg/s)	2,31		3,95	6,07			17,32	23,44	23,93	19,04	19,04	19,04					1,57			
CO	(kg/s)				4,62				4,62	19,56	11,96	11,96		11,96	11,96	11,96			0,39	0,20	
CO2	(kg/s)				11,10				11,10	17,08	29,04	29,04		29,04	29,04	3,71	25,19				
H2	(kg/s)				0,44				0,44	3,09	3,63	3,63		3,63	3,63	3,63			0,96	0,48	0,44
MeOH	(kg/s)																	15,84			
total	(kg/s)	15,47	3,95	3,95	23,94	9,16	12,80	17,32	63,26	64,11	64,11	64,11	19,51	45,07	45,07	19,74	25,19	17,41	1,65	0,83	0,70

c) volume flows

		1	2 2a		3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
total	(m3)				0,81	0,43	0,30	1,53		12,43	7,04	2,93		2,15	0,93	0,65	0,16		0,28	0,14	0,07
total	(Nm3)				24,16	12,83	8,96	21,56		89,32	89,32	89,32		65,63	65,63	52,73	12,83		11,46	5,73	5,29
phase		s	g	g	g	g	g	g	g	g	g	g	l	g	g	g	g	l		g	g

gas composition

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
CH4	(vol%)				9,00	100,00			22,21	0,70	0,70	0,70	2,68	0,96	0,96	1,19			3,66	3,66	6,93
C2+	(vol%)				0,47				0,17												
O2	(vol%)		100,00				100,00		13,26												
H2O	(vol%)			100,00	31,29			100,00	43,18	33,34	26,53	26,53	97,32					15,00			
CO	(vol%)				15,30				5,47	17,52	10,71	10,71		14,57	14,57	18,14			2,74	2,74	
CO2	(vol%)				23,40				8,37	9,73	16,55	16,55		22,52	22,52	3,58	100,00				
H2	(vol%)				20,55				7,35	38,70	45,51	45,51		61,94	61,94	77,09			93,60	93,60	93,07

MethaHydro M&E balance

Conversion to MeOH: 50%

a) mole-flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomass	kmol/s	0,39																			
CH4	kmol/s				0,06	0,67			0,73	0,03	0,03	0,03	0,03	0,03	0,03	0,03			0,02	0,01	0,02
C2+	kmol/s				0,00				0,00												
O2	kmol/s		0,11				0,43		0,43												
H2O	kmol/s	0,11		0,19	0,25			1,12	1,41	1,38	0,99	0,99	0,99								
CO	kmol/s				0,12				0,12	0,65	0,26	0,26		0,26	0,26	0,26			0,01	0,01	
CO2	kmol/s				0,19				0,19	0,37	0,76	0,76		0,76	0,76	0,09	0,67	0,08	0,01	0,00	
H2	kmol/s				0,16				0,16	1,60	1,99	1,99		1,99	1,99	1,99			0,79	0,39	0,38
MeOH	kmol/s																	0,33			
total	kmol/s	0,50	0,11	0,19	0,79	0,67	0,43	1,12	3,03	4,02	4,02	4,02	1,02	3,03	3,03	2,37	0,67	0,42	0,83	0,41	0,39
pressure	(bar)	·	30,00	30,00	30,00	30,00	30,00	30,00		30,00	30,00	30,00	30,00	30,00	80,00	80,00	80,00	80,00	80,00	80,00	80,00
temp.	(oC)				20,00	20,00	20,00	350,00	virtueel	950,00	420,00	15,00	15,00	15,00	60,00	15,00	15,00	15,00	300,00	300,00	15,00

b) mass-flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomassa	(kg/s)	11,65																			
CH4	(kg/s)				1,02	10,66			11,69	0,43	0,43	0,43	0,43	0,43	0,43	0,43			0,29	0,14	0,30
C2+	(kg/s)				0,11				0,11												
O2	(kg/s)		3,50				13,62		13,62												
H2O	(kg/s)	2,05		3,50	4,52			20,11	25,31	24,81	17,81	17,81	17,81								
CO	(kg/s)				3,25				3,25	18,13	7,25	7,25		7,25	7,25	7,25			0,35	0,17	
CO2	(kg/s)				8,55				8,24	16,28	33,37	33,37		33,37	33,37	4,07	29,30	3,66	0,27	0,14	
H2	(kg/s)				0,33				0,33	3,20	3,98	3,98		3,98	3,98	3,98			1,58	0,79	0,75
MeOH	(kg/s)																	10,66			
total	(kg/s)	13,71	3,50	3,50	17,78	10,66	13,62	20,11	62,55	62,84	62,84	62,84	18,24	45,03	45,03	15,73	29,30	14,32	2,48	1,24	1,05

c) volume flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
total	(m3)				0,59	0,50	0,32	1,77		12,54	7,10	2,95		2,23	0,97	0,65	0,18		0,45	0,23	0,11
total	(Nm3)				17,78	14,92	9,53	25,03		90,10	90,10	90,10		67,93	67,93	53,02	14,92		18,49	9,25	8,84
phase		s	g	g	g	g	g	g	g	g	g	g	l	g	g	g	g	l	l	g	g
gas composition																					
CH4	(vol%)				8,02	100,00			24,08	0,67	0,67	0,67	2,63	0,88	0,88	1,13			2,16	2,16	4,69
C2+	(vol%)				0,47				0,12												
O2	(vol%)		100,00				100,00		14,02												
H2O	(vol%)			100,00	31,63			100,00	46,34	34,26	24,60	24,60	97,37								
CO	(vol%)				14,63				3,83	16,10	6,44	6,44		8,54	8,54	10,94			1,49	1,49	
CO2	(vol%)				24,47				6,17	9,20	18,86	18,86		25,01	25,01	3,91	100,00	20,00	0,75	0,75	
H2	(vol%)				20,77				5,43	39,78	49,44	49,44		65,57	65,57	84,02			95,60	95,60	95,31

MethaHydro M&E balance

Conversion to MeOH: 50%

a) mole-flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomass	kmol/s	0,25																			
CH4	kmol/s				0,04	0,72			0,76	0,03	0,03	0,03		0,03	0,03	0,03			0,02	0,01	0,02
C2+	kmol/s				0,00				0,00												
O2	kmol/s		0,07				0,47		0,44												
H2O	kmol/s	0,07		0,13	0,17			1,38	1,44	1,37	0,88	0,88	0,92					0,09			
CO	kmol/s				0,08				0,08	0,60	0,11	0,11		0,11	0,11	0,11			0,01	0,01	
CO2	kmol/s				0,13				0,13	0,34	0,82	0,82		0,82	0,82	0,11	0,72		0,02	0,01	
H2	kmol/s				0,11				0,11	1,64	2,12	2,12		2,12	2,12	2,12			1,10	0,55	0,52
MeOH	kmol/s																	0,19			
total	kmol/s	0,33	0,07	0,13	0,53	0,72	0,47	1,38	2,96	3,97	3,96	3,96	0,92	3,08	3,08	2,37	0,72	0,28	1,16	0,58	0,55

pressure	(bar)		30,00	30,00	30,00	30,00	30,00	30,00		30,00	30,00	30,00	30,00	30,00	80,00	80,00	80,00	80,00	80,00	80,00	80,00
temp.	(oC)				20,00	20,00	20,00	350,00	virtueel	950,00	420,00	15,00	15,00	15,00	60,00	15,00	15,00	15,00	300,00	300,00	15,00

b) mass-flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
biomassa	(kg/s)	7,56																			
CH4	(kg/s)				0,69	11,44			12,13	0,52	0,52	0,52		0,52	0,52	0,52			0,38	0,19	0,33
C2+	(kg/s)				0,08				0,08												
O2	(kg/s)		2,27				15,16		14,07												
H2O	(kg/s)	1,33		2,27	3,04			24,78	25,92	24,61	15,93	15,93	16,57					1,61			
CO	(kg/s)				2,19				2,19	16,77	3,00	3,00		3,00	3,00	3,00			0,33	0,17	
CO2	(kg/s)				5,55				5,55	14,94	36,18	36,18		36,18	36,18	4,72	31,46		0,79	0,39	
H2	(kg/s)				0,22				0,22	3,27	4,24	4,24		4,24	4,24	4,24			2,20	1,10	1,05
MeOH	(kg/s)																	6,01			
total	(kg/s)	8,90	2,27	2,27	11,76	11,44	15,16	24,78	60,15	60,11	59,86	59,86	16,57	43,94	43,94	12,48	31,46	7,62	3,71	1,85	1,38

c) volume flows

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
total	(m3)				0,40	0,53	0,35	2,19		12,39	7,00	2,91	0,68	2,26	0,98	0,65	0,20		0,63	0,32	0,15
total	(Nm3)				11,86	16,02	10,61	30,83		89,02	88,82	88,82	20,62	68,99	68,99	52,98	16,02		25,89	12,95	12,21
phase		s	g	g	g	g	g	g	g	g	g	g	l	g	g	g	g	l	l	g	g

gas composition

		1	2	2a	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19
CH4	(vol%)				8,10	100,00			25,65	0,81	0,82	0,82	1,05		1,05	1,37			2,06	2,06	3,83
C2+	(vol%)				0,48				0,09												
O2	(vol%)		100,00				100,00		14,87												
H2O	(vol%)			100,00	31,90			100,00	48,73	34,41	22,32	22,32		100,00				32,26			
CO	(vol%)				14,76				2,65	15,07	2,71	2,71	3,48		3,48	4,53			1,03	1,03	
CO2	(vol%)				23,81				4,27	8,55	20,74	20,74	26,70		26,70	4,53	100,00		1,55	1,55	
H2	(vol%)				20,95				3,75	41,16	53,42	53,42	68,77		68,77	89,56			95,36	95,36	96,17

Appendix 5 The detailed economics of the MethaHydro plant

Economics of the MethaHydro-plant under the European Co-ordination scenario

Main mass flows MethaHydro plant

		Product mix (%MeOH)			
Input	Unit	90%MeOH	70%MeOH	50%MeOH	30%MeOH
Biomass	(dry ton/day)	1524	1136	832	575
Natural gas	(Nm3/day)	9,55E+05	1,19E+06	1,37E+06	1,53E+06
Oxygen	(ton/day)	1362	1447	1420	1522
Process					
Feed to ATR	(Nm3/day natural gas equival	1,20E+06	1,37E+06	1,51E+06	1,62E+06
Syngas to shift reactor	(kmol/h CO+H2)	7632	8070	8048	8291
Shift catalyst use	(m3/year)	48	51	51	52
Syngas compressor duty	(kW)	5790	6332	6520	6861
CO_2 removal	(kmol/h)	1656	2061	2385	2653
Selexol use	(ton/year)	12	15	17	19
Syngas to methanation	(kmol/h)	252	921	1478	2123
Output					
Methanol	(ton/day)	1800	1368	916	535
'Ecogas'	(MWth)	45	133	210	296
Electricity	(kWe)	22268	65040	102588	147656

*volume CH4 plus 2 times the volume C2+

Capital Costs (1e6$)

	Product mix (%MeOH)				
Installed hardware	90%MeOH	70%MeOH	50%MeOH	30%MeOH	flexible
Oxygen plant	44	46	46	48	48
Biomass gasifier	115	94	75	58	115
Autothermal reformer	23	26	28	29	29
Shift reactor	34	35	35	36	36
Syngas compressor	47	50	51	54	54
Carbon dioxide removal	14	16	18	19	19
Methanol reactor	30	30	30	30	30
Methanation reactor	1	2	4	5	5
Gas and steam turbine	18	54	85	123	123
Utilities/auxilaries	29	25	26	27	27
Sub-total syngas production	306	293	279	272	329
Sub-total methanol synthesis	30	30	30	30	30
Sub-total 'ecogas' production purification	1	2	4	5	5
Sub-total electricity production	18	54	85	123	123
Total costs installed hardware	337	325	313	307	364
Contingencies	67	65	63	61	73
Owners costs, fees	67	65	63	61	73
Startup	17	16	16	15	18
Total capital requirement	489	472	454	445	527
Working capital	34	34	34	34	34
Land	4	4	4	4	4

				EC	
Total Capital Costs	527	510	492	483	565

Operating costs (1e6 $/year)

	Unit costs		Product mix (%MeOH)			
			90%MeOH	70%MeOH	50%MeOH	30%MeOH
Variable costs						
Biomass	44	$/ton	25	18	13	9
Natural gas	0,11	$/m³	40	50	57	64
CO2 storage	1	$/ton	0,64	1	1	1
Catalysts and chemicals			3	3	3	3
Subtotal			68	72	75	77
Fixed costs						
Labor			2	1	1	1
Maintenance	3%	of installed h	10	10	9	9
Overhead	65%	of labor and	8	8	7	7
Subtotal			20	19	18	17
Total operating costs			88	91	93	94

Annual levelized costs (1e6$/yr)

			Product mix (%MeOH)			
			90%MeOH	70%MeOH	50%MeOH	30%MeOH
Capital	kapitaalonzeker	1,00				
Syngas production			77	74	71	69
Methanol production			5	5	5	5
Ecogas production			0	0	1	1
Electiricity production			3	8	13	19
Labour & maintenance			20	19	18	17
Biomass			25	18	13	9
Gas			40	50	57	64
CO2-storage			1	1	1	1
Catalysts/chemicals			3	3	3	3
Total Production Costs			173	178	182	188

Allocation factor for costs of syngas production (allocation based on energetic content of the process streams at the split-off point)

Methanol	0,86	0,62	0,39	0,22
'Ecogas'	0,04	0,13	0,22	0,26
Electricity	0,09	0,25	0,38	0,52

Levelized costs per product (1e6 $/yr)

	90% MeOH			70% MeOH			50% MeOH			30% MeOH		
	Methanol	Electricity	Ecogas'	Methanol	Electricity	Ecogas'	Methanol	Electricity	Ecogas'	Methanol	Electricity	Ecogas'
Capital												
Share of syngas production	66	7	3	45	19	10	28	27	16	15	36	18
Methanol production	5			5			5			5		
'ecogas' production			0			0			1			1
Electricity production		3			8			13			19	
Labour & maintenance	17	2	1	12	5	2	7	7	4	4	9	5
Biomass	21	2	1	11	5	2	5	5	3	2	5	2
Gas	34	4	2	31	13	6	23	22	13	14	33	17

CO_2-storage	1	0	0	0	0	EC 0	0	0	0	0	1	0
Catalysts/chemicals	3	0	0	2	1	0	1	1	1	1	2	1
Total Production Costs	147	18	8	106	50	22	69	76	37	40	104	44