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Co-production of fuels as an option for Demkolec?

A preliminary review of opportunities for the co-production of liquid or gaseous energy carriers

T.J. de Lange
M. Beeldman
J.H.A. Kiel
H. den Uil
J.M. Veenkamp

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J.H.A. Kiel</td><td rowspan="2">ECN Biomass
ECN Policy Studies</td></tr>
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J.J.C. Bruggink</td><td>Issued by:
H.J. Veringa</td></tr>
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Scope

This study was performed in a joint effort by ECN Biomass, ECN Clean Fossil Fuels and ECN Policy Studies for Novem (contract number 355200/3510) in the framework of the EWAB-programme, which is financed by the Dutch Ministry of Economic Affairs, and Demkolec B.V. (contract number D0058221). The Novem project manager was Ir. P.J. Stollwerk. The contact persons for Demkolec B.V. were Ir. M. van den Boomen (Toptree bv) and Ir. J.Th.G.M. Eurlings.

Abstract

This study comprises a preliminary review of the opportunities for the co-production of liquid or gaseous energy carriers at the 253 MWe coal gasification combined cycle (IGCC) power plant of Demkolec in Buggenum, the Netherlands. Demkolec is actively pursuing various options to optimise the economics of the plant to become a strong competitor on the liberalised power market. In this respect, it has initiated already a substitution of approx. 50% of the coal consumption of this plant by cheaper alternative fuels. Part of this alternative fuel input will provide an extra product income, because this part comprises organic material that is regarded "green". In addition, Demkolec has recognised that fuels co-production may be an opportunity to further improve the economics of the plant.

In the present situation, the plant is operated at partial load (approx. 57%) during off-peak hours since off-peak electricity production is not cost-effective. However, by maintaining the gasification unit at full load during off-peak hours, a surplus of syngas could be produced, which could be converted into liquid or gaseous energy carriers. In this study, four options for off-peak co-production have been examined from a technical and economic point of view, viz. co-production of hydrogen, methanol, Fischer-Tropsch transportation fuels and Substitute Natural Gas (SNG). For the co-production of methanol, a further distinction has been made between the Lurgi process including prior adjustment of the H_2/CO ratio (CO-shift) and Air Products' Liquid Phase Methanol (LPMEOH^{TM}) process without this prior adjustment, and between the production of Fuel Grade and AA Grade methanol. For each co-production option, either a specific process configuration has been considered for which specifications of the process streams and costs were available from other studies or only a general configuration with rough estimates for the process stream specifications and costs. This means that for all options other process configurations are possible and that the selected ones have not been optimised for the Demkolec plant. It has been assumed that the gasifier is operated on the new fuel mixture.

Hydrogen and Lurgi methanol are identified as technically the most mature options, whilst LPMEOH methanol is the most flexible one. All co-production options, except for Fischer-Tropsch transporations fuels, are found to be financially viable, but the viability appears to be strongly dependent on the product market price. The co-production of Lurgi (AA Grade) Methanol proves to yield the highest net yearly income for the given best estimates of the product market price. The economics of hydrogen and SNG co-production are quite comparable. The co-production of Fischer-Tropsch transportation fuels is not financially viable at the assumed fossil crude oil price of 15 US$/barrel; a price level of at least 23-29 US$/barrel is required.

Apart from the viability of the co-production options, two alternative modes of plant operation have been considered. Firstly, it appears that electricity production during off-peak hours can be made cost-effective if the plant switches to the new fuel mixture, through a combination of lower fuel cost and a green premium on the biomass component of the feed. Consequently, full-load power generation may become an attractive alternative. The calculated additional net yearly income is fairly large compared to the values for the different co-production options and no additional investments or risks are involved. However, given the present situation on the liberalised electricity market, continuous full-load power generation is probably not the most profitable alternative.

A second alternative is to operate the Demkolec plant in a regulating power mode. This mode of operation enforces the opportunities for fuels co-production. With a sufficient flexibility of the co-production process, load changes in the electricity production can be matched by changing the load on the co-production section. It is recommended to evaluate the alternative of regulating power mode operation in combination with fuels co-production more thoroughly, as it represents the most likely future for the Demkolec plant and it offers important opportunities to further increase its overall financial-economic performance.

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SUMMARY

At the end of the eighties, the Dutch Electricity Generating Board (SEP) took the decision to build a power plant based on coal gasification in combination with a combined cycle unit. The main reason for applying this new technology was the aim to diversify the fuel mix within the boundaries of increasing environmental restrictions. This 253 MWe Integrated Gasification Combined Cycle (IGCC) unit, the Demkolec plant in Buggenum, has been taken into operation by the end of 1993. However, already during the building stage, it became clear that the market position of the Demkolec plant was changing in a rapid way. Liberalisation and privatisation of the power sector and the overcapacity on the electricity market made it difficult for Demkolec to compete with other power plants based on more mature technologies.

In order to offer a future perspective, and to avoid that the gasification part of the plant will have to be broken down in favour of using natural gas, Demkolec is actively pursuing various options where maximum advantage is made of the presence of the gasification block. This has already resulted in a broader fuel input by including a wide range of relatively cheap, lower quality (import) coals as well as several opportunity fuels, while the co-gasification of several biomass streams (e.g. paper and sewage sludge, and chicken manure) is currently being tested. Use of cheap feedstocks, in particular biomass, will significantly improve the economics of the plant, also taking into account the national and European government policy to aim for a steep increase in renewable energy production and a limitation of net CO_2 emissions. In an earlier study conducted by ECN[1], an analysis has been made of the financial viability of this option in comparison with other options for large-scale power generation from renewable sources. The results of this study were positive, although it was recognised that several uncertainties still exist with regard to the technical feasibility and the benefits for producing green electricity.

In addition to switching to cheaper feedstocks, both Demkolec and the Netherlands Agency for Energy and the Environment (Novem) have recognised that the co-production of liquid or gaseous products may be an opportunity to further improve the economics of IGCC plants. Furthermore, driven by Dutch government policy goals, Novem is promoting the development of options for green liquid or gaseous fuels production and of cost-effective options for CO_2 removal and storage. In this context, Demkolec and Novem have asked the Netherlands Energy Research Foundation (ECN) to make a first assessment of the techno-economic prospects of the co-production of different liquid or gaseous products.

This report gives an overview of four options that have been explored on their technical and financial viability. These options are:

- **Co-production of hydrogen**, which is supplied into the existing natural gas network.
- **Co-production of methanol**. Two methanol processes have been considered, viz. the Lurgi process including prior adjustment of the H_2/CO ratio (CO-shift) and Air Products' once-through Liquid Phase Methanol (LPMEOH^{TM}) process without this prior adjustment, and two methanol product qualities (Fuel Grade and AA Grade).
- **Co-production of Fischer-Tropsch transportation fuels**, involving once-through Fischer-Tropsch synthesis, hydrocracking and product fractionation.
- **Co-production of Substitute Natural Gas (SNG)**, which is supplied into the existing natural gas network.

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[1] De Lange, T.J. and J.H.A. Kiel, *Kosten-batenanalyse 'groene Demkolec' in vergelijking met andere duurzame energieopties.* ECN report ECN-CX--00-046, Petten, the Netherlands, 2000 (in Dutch).

In a first step, these options have been defined and evaluated in technical terms. Starting-point formed the operating mode for the Demkolec plant in 2000 (base case). This consisted of full-load (460 MWth) power generation during peak hours and operation at minimum load (260 MWth), as determined by the gas turbine, during off-peak hours. Power generation was not cost-effective during off-peak hours due to the low selling price for the electricity produced. Therefore, it has been assumed that in principle the co-production section will be in operation during off-peak hours only and involves 200 of the 460 MW total thermal plant capacity. In addition, it has been assumed that the gasifier is operated on a mixture of 24% "green fuels" (biomass), 21% opportunity fuels and 55% coal (energy basis) in agreement with the earlier study.

The hydrogen co-production technology selected for this study is conventional. The process information and the economic data are mainly based on a SEP report concerning co-production of hydrogen and power. Hydrogen is produced with a purity of 99.9%.

For the co-production of methanol, two processes have been considered, viz.:

- The Lurgi process including prior adjustment of the H_2/CO ratio (CO-shift);
- Air Products' Liquid Phase Methanol (LPMEOH^{TM}) process without this prior adjustment.

The Lurgi process is a conventional process of which many plants have been built around the world. The process information and the economic data on the co-production of methanol using the Lurgi process are based on a SEP report concerning co-production of methanol and power. The LPMEOH process is an innovative process of which a demonstration plant is running since 1997. The LPMEOH process has been developed especially to handle directly the CO-rich syngas produced by the gasification of coal, petroleum coke, residual oil, wastes or other hydrocarbon feedstocks. In case of co-production of power and methanol, it is claimed to have the flexibility to operate in a daily electricity demand load-following manner. The process information and the economic data on the co-production of methanol using a once-through LPMEOH process are based on information from Air Products. With both processes, two product qualities can be produced, viz. Fuel Grade Methanol (95% pure for the Lurgi process and 98% pure for the LPMEOH process) and AA Grade Methanol (99.85% pure).

The once-through Fischer-Tropsch transportation fuels co-production option is based on the Shell Middle Distillate Synthesis (SMDS) process and includes units for hydrocracking and product fractionation. Process information and economic data have been obtained from Shell.

Finally, the data for the SNG option have been generated in a simultaneous, but separate study by GASTEC. To enable a direct comparison with the other options, the GASTEC data have been interpreted and adapted to obtain suitable input data for the ECN evaluation approach.

The financial evaluation of the different options has been conducted on the basis of additional costs and additional income compared with the base case in which only electricity is produced. Therefore, the results do not provide information on the viability of the Demkolec plant as a whole, but solely on the financial viability of the co-production options. For most co-production options, the calculations have been based on the assumption that during off-peak hours a fixed fraction of the syngas (260 MWth due to the minimum load requirements of the gas turbine) is used for electricity production and the remaining part (200 MWth.) for the production of liquid or gaseous fuels. During peak hours, the plant is used for electricity production only and the co-processing plant is turned down as far as possible.

For the new fuel mixture, as described above, the following fuel prices have been assumed:

• Coal	4.50 NLG/GJ
• "Green" alternative fuels	0.92 NLG/GJ
• "Non-green" alternative fuels	0.48 NLG/GJ

The electricity that is being produced is estimated to be sold for 0.065 NLG/kWh during peak hours and 0.03 NLG/kWh during off-peak hours. The additional premium of the green electricity, valid for peak hours as well as off-peak hours, is estimated to be 0.0877 NLG/kWh (including 0.0427 NLG/kWh REB and 0.045 NLG/kWh green label premium). The following product prices have been assumed:

• Hydrogen	10 NLG/GJ
• AA Grade Methanol	12.6-25.2 NLG/GJ
• Fischer-Tropsch transportation fuels	8.7-11.0 NLG/GJ
• SNG	8.8-13.9 NLG/GJ

For hydrogen, this is the present market price. The AA Grade Methanol market price was ranging during the past year between 96 $/tonne in 1999 to a maximum of 230 $/tonne in the USA during the month of September of 2000. In Europe the fourth quarter 2000 contract list price FOB Rotterdam showed a price of approximately 224 $/tonne (information from Methanex Corporation's web-site). Due to the large fluctuations on the market, the price for AA Grade Methanol has been assumed to range between 100 $/tonne and 200 $/tonne in the coming years, corresponding with 12.6-25.2 NLG/GJ at an exchange rate of 2.50 NLG/$. The price for Fuel Grade Methanol is not available on the open market but one can assume that it is directly linked with the price for AA Grade Methanol with an offset for the variable costs of upgrading the quality. The price range for Fischer-Tropsch transportation fuels is based on the assumption that these transportation fuels have a 135-170% higher value than fossil crude oil. The price of fossil crude oil is assumed to be 15 $/barrel. For Fischer-Tropsch transportation fuels, this means a price ranging from 8.7 NLG/GJ to 11.0 NLG/GJ. The price range for SNG is equal to the current commodity price range for natural gas and corresponds to 0.28-0.44 NLG/m_n^3.

With respect to the main results and conclusions of this evaluation, the limitations of the applied evaluation approach (a preliminary review as stated in the subtitle of the report) should be noticed clearly. For each co-production option either a specific process configuration has been considered for which specifications of the process streams and costs were available from other studies (hydrogen, Lurgi and LPMEOH methanol and SNG to some extent) or only a general configuration with rough estimates for the process stream specifications and costs (Fischer-Tropsch transportation fuels). This means that for all options other process configurations are possible and that the selected ones have not been optimised for the Demkolec plant.

With this in mind, the following main results and conclusions can be drawn from the technical and financial evaluation of the different co-production options:

1. From a viewpoint of technical maturity, hydrogen and Lurgi methanol are the preferred co-production options. Both are mature technologies that are demonstrated on a wide scale. Air Products' LPMEOH process is an innovative process of which a demonstration plant is running since 1997. Also, the Fischer-Tropsch transportation fuels and SNG processes are regarded to be innovative processes, although all individual process steps are mature technology.

2. In terms of operating flexibility, the (once-through) LPMEOH process design without prior adjustment of the H_2/CO ratio in a CO-shift section is the most attractive option. The absence of a CO-shift section in combination with the robustness of the slurry bubble column reactor for methanol synthesis enable rapid ramping and extreme stop/start actions; the process can be decreased to 0% capacity during peak hours. However, the size of the once-through LPMEOH process units is rather large since the total syngas stream of 460 MWth has to be processed. The flexibility of the hydrogen, Lurgi methanol, Fischer-Tropsch transportation fuels and SNG co-production options is limited mainly by the CO-shift section, the CO-shift + methanol synthesis sections, the Fischer-Tropsch synthesis section and the methanation section respectively. These processes cannot be hold on a hot

stand-by; their minimum capacity amounts to 10% (SNG) or 20% (hydrogen, Lurgi methanol and Fischer-Tropsch Transportation fuels).

3. A comparison of the financial viability of the different co-production options is given in the table below. From this table it can be seen that all co-production options, except for Fischer-Tropsch transporations fuels, are found to be financially viable, but the viability appears to be strongly dependent on the product market price. The co-production of Lurgi (AA Grade) Methanol proves to yield the highest net yearly income for the given best estimates of the product market price. The lower financial viability of the Fuel Grade Methanol options compared to the AA Grade Methanol options is because the assumed difference in product market price is larger than the difference in production cost. The economics of hydrogen and SNG co-production are quite comparable. The 99.9% purity of the hydrogen is probably higher than required for supply into the natural gas network. The production of hydrogen with a lower purity is expected to yield a higher net yearly income. The co-production of Fischer-Tropsch transportation fuels becomes financially viable at an average product price of 16.8 NLG/GJ, which corresponds to a fossil crude oil price of 23-29 US$/barrel.

Co-production option	Product market price (best estimate, NLG/GJ)	Product price for which zero yearly income (NLG/GJ)	Net yearly income at product market price (MNLG/year)	Δ yearly income per Δ NLG/GJ product price (MNLG/year)
Hydrogen	10.0	6.8	8.8	2.8
Lurgi Fuel Grade methanol	12.6	10.2	6.0	2.5
LPMEOH Fuel Grade Methanol	12.6	6.4	9.3	1.5
Lurgi AA Grade methanol	18.9	12.8	15.4	2.5
LPMEOH AA Grade Methanol	18.9	10.1	12.8	1.5
FT transportation fuels	9.8	16.8	-15.4	2.2
SNG	10.0	5.6	8.8	2.0

4. For the given best estimate product market prices, a 25% change in investment costs leads a change in net yearly income ranging from 15% for the SNG option to 65% for the Lurgi Fuel Grade Methanol option. A 25% change in fuel costs leads to a change in net yearly income ranging from 13% for the Fischer-Tropsch transportation fuels option to 35% for the Lurgi Fuel Grade Methanol option.

In addition, the following more general conclusions can be drawn:

5. In the different cases no investment subsidies like IEA and VAMIL have been taken into account. These subsidies by the Dutch government could (with certain limitations) amount up to approximately 20% of the investment costs. They are especially applicable for investments in energy saving measures, renewable energy or environmental measures in the energy sector. Most options for co-production, but especially investments to enable the replacement of coal by biomass as a fuel for the Demkolec plant could benefit from these subsidies.

6. Because of present energy and environmental policy, the ‘greenness’ of electricity and fuels represents an extra value for the product. For electricity, this value is rather clear and is strongly related to the regulating energy tax. For the other products, however, the value of this greenness is still less clear. To stimulate the implementation of fuels (co-)production options, it is important that this issue will be clarified. In this study, it has been assumed that the extra value for greenness for the produced fuels can be equalised to that for electricity on a primary input basis.

7. In all co-production options, except for the LPMEOH methanol option, a concentrated CO_2 process stream becomes readily available. For the Lurgi methanol and Fischer-Tropsch transportation fuels options, this is a concentrated CO_2 purge stream from the Selexol CO_2-removal unit, whilst for the hydrogen and SNG options, this is the purge gas stream from the PSA unit. The costs for conditioning, pressurising, transportation and storage in aquifers, empty gas fields or coal beds show large uncertainties, just as the height of the extra income that could be generated by producing CO_2-neutral products. Therefore, the costs and extra income for CO_2 removal and storage have not been included in the financial evaluation. To give a rough indication, the yearly CO_2-reduction in case of CO_2-removal and injection amounts to 0.3 Mtonne/year and 0.2 Mtonne/year for the hydrogen and Lurgi methanol options respectively. The availability of concentrated CO_2 at the co-production plant could also offer other opportunities by looking into possibilities to sell it to industries that need it as a feedstock.

8. In the present situation, with the Demkolec power plant being operated mainly on coal, the revenues for the produced electricity during off-peak hours are lower than the production costs. However, if the plant switches to the new fuel mixture, then electricity production during off-peak hours becomes cost-effective. Consequently, continuous full-load power generation may become an attractive option from a financial point of view. In fact, at the current off-peak electricity price of 0.03 NLG/kWh the calculated net yearly income of 15 MNLG/year is fairly large compared to the values for the different co-production options. Even when the off-peak electricity price would be 0 NLG/kWh, the off-peak electricity production would still be cost-effective due to the income from the partial "greenness" of the fuel mixture. In addition, continuous full-load power generation does not involve any additional investments or risks. However, given the present situation on the liberalised electricity market, continuous full-load power generation is probably not the most profitable alternative.

9. Due to the flexibility of operation of the Demkolec plant, the difference between full load (100%) and minimum off-peak load (57%) can be sold as regulating power capacity (at a higher price). In fact, the Demkolec plant is operating to a considerable extent in such a mode already, both during off-peak hours as well as during peak hours. This mode of operation enforces the opportunities for fuels co-production. With a sufficient flexibility of the co-production process, load changes in the electricity production can be matched by changing the load on the co-production section. The gasifier can be operated continuously at full load. In this way, it will be possible to get the high prices for the electricity being produced in a regulating power mode and, at the same time, it can lead to more operating hours of the co-production section, thus increasing its financial viability. It is recommended to evaluate this alternative more thoroughly as it represents the most likely future for the Demkolec plant and it offers important opportunities to further increase its overall financial-economic performance.

1. INTRODUCTION

At the end of the eighties, the Dutch Electricity Generating Board (SEP) took the decision to build a power plant based on coal gasification in combination with a combined cycle unit. The main reason for applying this new technology was the aim to diversify the fuel mix within the boundaries of increasing environmental restrictions. This 253 MWe Integrated Gasification Combined Cycle (IGCC) unit, the Demkolec plant in Buggenum, has been taken into operation by the end of 1993. However, already during the building stage, it became clear that the market position of the Demkolec plant was changing in a rapid way. Liberalisation and privatisation of the power sector and the overcapacity on the electricity market made it difficult for Demkolec to compete with other power plants based on more mature technologies.

In order to offer a future perspective, and to avoid that the gasification part of the plant will have to be broken down in favour of using natural gas, Demkolec is actively pursuing various options where maximum advantage is made of the presence of the gasification block. This has already resulted in a broader fuel input by including a wide range of relatively cheap, lower quality (import) coals as well as several opportunity fuels, while the co-gasification of several biomass streams (e.g. paper and sewage sludge, and chicken manure) is currently being tested. Use of cheap feedstocks, in particular biomass, will significantly improve the economics of the plant, also taking into account the national and European government policy to aim for a steep increase in renewable energy production and a limitation of net CO_2 emissions. In an earlier study conducted by ECN (De Lange and Kiel, 2000), an analysis has been made of the financial viability of this option in comparison with other options for large-scale power generation from renewable sources. The results of this study were positive, although it was recognised that several uncertainties still exist with regard to the technical feasibility and the benefits for producing green electricity.

In addition to switching to cheaper feedstocks, both Demkolec and the Netherlands Agency for Energy and the Environment (Novem) have recognised that the co-production of liquid or gaseous products may be an opportunity to further improve the economics of IGCC plants. Furthermore, driven by Dutch government policy goals, Novem is promoting the development of options for green liquid or gaseous fuels production (GAVE, 2000) and of cost-effective options for CO_2 removal and storage. In this context, Demkolec and Novem have asked the Netherlands Energy Research Foundation (ECN) to make a first assessment of the techno-economic prospects of the co-production of different liquid or gaseous products.

This report gives an overview of four options that have been explored on their technical and financial viability. These options are:

- **Co-production of hydrogen**, which is supplied into the existing natural gas network;
- **Co-production of methanol**;
- **Co-production of Fischer-Tropsch transportation fuels**;
- **Co-production of Substitute Natural Gas (SNG)**, which is supplied into the existing natural gas network.

The first three co-production options have been evaluated in detail by ECN, whilst the information on the SNG option has been obtained from a study conducted simultaneously, but separately, by GASTEC for Demkolec (Van Rens, 2000).

2. METHODOLOGY

In the study, first a broad range of possible co-production schemes has been identified and characterised briefly, using information available at ECN and Demkolec through earlier projects and from open literature. From this inventory, four options have been selected for a more detailed techno-economic assessment. These options are:

- **Co-production of hydrogen**, which is supplied into the existing natural gas network.
- **Co-production of methanol**. Two methanol processes have been considered, viz. the Lurgi process including prior adjustment of the H_2/CO ratio (CO-shift) and Air Products' once-through Liquid Phase Methanol (LPMEOH^{TM}) process without this prior adjustment, and two methanol product qualities (Fuel Grade and AA Grade).
- **Co-production of Fischer-Tropsch transportation fuels**, involving once-through Fischer-Tropsch synthesis, hydrocracking and product fractionation.
- **Co-production of Substitute Natural Gas (SNG)**, which is supplied into the existing natural gas network.

The definition and technical evaluation of these options is given in Chapter 3. The data for the first three options were generated by ECN, relying heavily on results from earlier and parallel R&D activities concerning these processes. The data for the SNG option were generated simultaneously, but separately, by GASTEC (Van Rens, 2000). Steady-state system calculations were performed to determine overall mass and energy balances, and an assessment was made of the current development status of the processes involved. Then, an estimate was made of the additional investment costs and the additional operating and maintenance (O&M) costs.

Starting-point for the definition and evaluation of the four options formed the operating mode for the Demkolec plant in 2000. This consisted of full-load (460 MWth) power generation during peak hours and power generation at minimum load (260 MWth), as determined by the gas turbine, during off-peak hours. Power generation was not cost-effective during off-peak hours due to the low selling price for the electricity produced. Therefore, it has been assumed that in principle the co-production section will be in operation during off-peak hours only and involves 200 of the 460 MW total thermal plant capacity. In addition, it has been assumed that the gasifier is operated on a mixture of 24% "green fuels" (biomass), 21% opportunity fuels and 55% coal (energy basis) in agreement with the earlier study conducted by ECN (De Lange and Kiel, 2000).

Subsequently, the technical and financial data for the four options have been used for a financial-economic evaluation. This evaluation, described in Chapter 4, has been conducted on the basis of *additional costs and additional income* compared with the base case in which only electricity is produced as described above. Therefore, the results do not provide information on the viability of the Demkolec plant as a whole, but solely on the financial viability of the co-production options. The evaluation includes sensitivity analyses for key parameters and an evaluation of the possibilities for CO_2-removal and storage. In addition, two alternative scenarios are discussed briefly, viz.:

- Continuous power generation at full load, since power generation may become cost-effective during off-peak hours when switching to new the fuel mixture.
- Liquids/gaseous fuels co-production in case the Demkolec plant is operated in a regulating power mode. In the course of the study, this mode of operation emerged as a more likely scenario. In fact, the Demkolec plant is operating in this way already.

In Chapter 5, the main conclusions drawn from the different evaluations are given.

3. TECHNICAL DEFINITION AND EVALUATION OF CO-PRODUCTION OPTIONS

3.1. Hydrogen

The hydrogen co-production technology selected for this study is conventional. The process information and the economic data are mainly based on a SEP report concerning co-production of hydrogen and power (Lensink - van der Veen, 1996b). According to this SEP report, the hydrogen co-production unit consists of a saturator/desaturator system, a CO-shift reaction section based on a Co/Mo catalyst, a Selexol CO_2-removal unit and a Pressure Swing Adsorption (PSA) unit. Hydrogen is produced with a purity of 99.9%.

The design of the hydrogen co-production unit, as considered in this study, differs from the SEP design in one aspect: instead of using both Selexol and PSA, only PSA is used to remove carbon dioxide and to bring the hydrogen purity to 99.9%. This option has been made possible by developments in PSA technology (Steinberg and Cheng, 1989) and leads to approx. 25% reduction in overall investment costs. The block scheme of the hydrogen co-production section is shown in Figure 1.

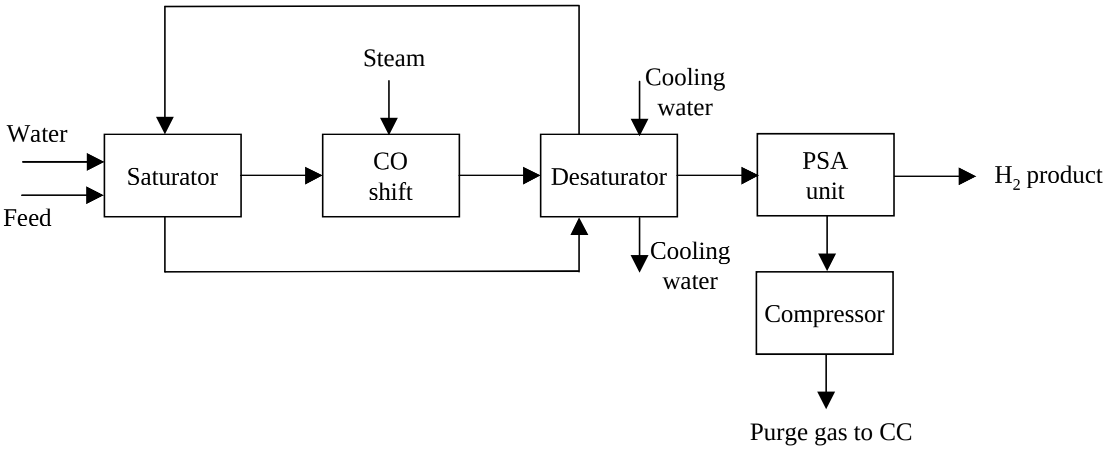

Figure 1. *Block scheme of the hydrogen co-production section*

CO-shift reaction section and saturator/desaturator system

Cleaned syngas (feed) enters the saturator at 37.5 °C. In the saturator, the syngas is saturated with water at about 200 °C. 70% of the syngas that leaves the saturator is fed to the first shift reactor and is preheated with the reactor effluent. Medium pressure steam is added to increase the steam/dry gas ratio. The effluent of the first shift reactor is cooled with the feed, is mixed up with the remaining syngas and is fed to the two following shift reactors. The effluent of the last, low temperature, shift reactor is fed to the desaturator, where it is cooled first to 140 °C with the saturator bottom stream. Further cooling to 40 °C is conducted with cooling water.

Pressure Swing Adsorption

The hydrogen-rich product stream from the desaturator is purified to 99.9% in the PSA unit. The PSA purge gas contains combustible components. It is therefore pressurised by a purge gas compressor and fed to the combined cycle for electricity production.

Maturity of the technology

The technology used is a conventional and proven way of producing hydrogen from syngas (Steinberg and Cheng, 1989; Badin et al., 1999; Kwon et al., 1999).

Flexibility of the process

The hydrogen co-production section is not very flexible because of the thermal characteristics of the adiabatic shift reactors. Especially rapid start/stop operations are problematic, because the reactors cannot be held on a hot standby and they always need a humid condition. Therefore, it is assumed that hydrogen co-production has to be maintained at 20% capacity during peak hours. It is assumed that the heat and mass balances can be scaled down linearly in the same ratio. It should be noted that this leads to a lower electricity production during peak hours compared to the base case, which will have to be accounted for in the financial-economic evaluation.

Operating hours

The assumed number of operating hours for the co-production plant is 3650 off-peak hours per year and 7000 hours per year in total, corresponding to approx. 80% availability. This is lower than the 85% availability mentioned in the SEP report on which the given data are based.

3.1.1. Process analysis hydrogen co-production

Specification of process streams

The specification of the process streams entering and leaving the hydrogen co-production section is given in Appendix A. All streams entering and leaving the co-production section are summarised in Table 1. This table also gives the streams on a tonne/h basis, a tonne/3650h (yearly off-peak production) basis, a tonne/4320h basis and a tonne/7000h (yearly full production) basis.

Table 1. *Process streams, major enthalpy flows and electricity entering and leaving the hydrogen co-production section*

Process Streams												
Name	kg/s		t/h		t/t H2		t/y (3650 h/y basis)		t/y (4320 h/y basis)		t/y (7000 h/y basis)	
	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
Syngas	18,85		67,9		12,72		247688		293154		475017	
22 bar steam	7,13		25,7		4,81		93694		110892		179686	
Make-up water	3,42		12,3		2,31		44973		53228		86249	
Cooling water	46,98		169,1		31,69		617253		730557		1183773	
Hydrogen product		1,48		5,3		1,00		19477		23053		37354
PSA-purge to CC		27,92		100,5		18,84		366877		434221		703599
Cooling water		46,98		169,1		31,69		617253		730557		1183773
Total	76,38	76,38	275,0	275,0	51,53	51,53	1003607	1003607	1187831	1187831	1924726	1924726

Major enthalpy flows (MWth)												
	MWth		MWh/h		MWh/t H2		MWh/y (3650 h/y)		MWh/y (4320 h/y)		MWh/y (7000 h/y)	
Syngas	221,49		221,5		41,5		808425		956821		1550404	
22 bar steam	3,57		3,6		0,7		13013		15402		24956	
Hydrogen product		176,22		176,2		33,0		643189		761254		1233513
PSA-purge to CC		21,49		21,5		4,0		78439		92837		150430
Balance		27,35		27,3		5,1		99811		118132		191418
Total	225,05	225,05	225,1	225,1	42,2	42,2	821438	821438	972222	972222	1575360	1575360

Electricity Usage (MWe)						
	MWe	MWh/h	MWh/t H2	MWh/y (3650 h/y)	MWh/y (4320 h/y)	MWh/y (7000 h/y)
Electricity	3,48	3,48	0,65	12701	15032	24357

Steam Consumption translated to Electricity Usage (note 1)						
	MWe	MWh/h	MWh/t H2	MWh/y (3650 h/y)	MWh/y (4320 h/y)	MWh/y (7000 h/y)
Electricity	1,25	1,25	0,23	4555	5391	8735

note 1: electricity generating efficiency from steam is taken to be 35%

note 2: 100% of production capacity during off-peak hours (3650 h/y) and 20% of production capacity during peak hours (3350 h/y) is equivalent to 4320 h/y operating on 100% of production capacity

note 3: t = tonne

3.1.2. Investment and O&M costs hydrogen co-production

Investment costs

The total investment costs for the hydrogen co-production plant are 67.8 $MNLG_{2000}$. The build-up of the investment costs is given in Table 2. The costs were calculated using the SEP data (Lensink - van der Veen, 1996b) and using (Kwon et al., 1999). The SEP cost basis is year-1995 NLG for a 1398 MWth turnkey plant based in the Netherlands. The costs were converted into year-2000 NLG and adjusted for the capacity: the investment costs were divided by the number of parallel trains in the original SEP design and scaled down using the 0.65-rule of thumb.

Table 2. *Investment costs for hydrogen co-production*

Process section	Investment costs (MNLG)
Saturator/CO-shift section/desaturator	18.1
PSA unit	32.8
Purge gas compressor	16.9
Total	67.8

Operating and maintenance costs

The operating and maintenance costs were calculated as 4% of the investment costs, just as in the SEP report. Added to this were the yearly costs of the CO-shift catalyst, which were estimated to be 0.3 MNLG/year (based on Hendriks, 1994). The yearly PSA-bed costs were assumed to be negligible (Haldor Topsoe, 2000). This resulted in total O&M costs of 3.0 MNLG/year.

3.2. Methanol

For the co-production of methanol, two processes have been considered, viz.:

- The Lurgi process including prior adjustment of the H_2/CO ratio (CO-shift);
- Air Products' once-through Liquid Phase Methanol (LPMEOH^{TM}) process without this prior adjustment.

The Lurgi process is a conventional process of which many plants have been built around the world. The LPMEOH process is an innovative process of which a demonstration plant is running since 1997. The LPMEOH process has been developed especially to handle directly the CO-rich syngas produced by the gasification of coal, petroleum coke, residual oil, wastes or other hydrocarbon feedstocks. In case of co-production of power and methanol, it is claimed to have the flexibility to operate in a daily electricity demand load-following manner (Daimond, 1999).

With both processes, two product qualities can be produced, viz. Fuel Grade Methanol (95% pure for the Lurgi process and 98% pure for the LPMEOH process) and AA Grade Methanol (99.85% pure). This yields four methanol co-production cases, which are described in the paragraphs below.

3.2.1. Lurgi Process

The process information and the economic data on the co-production of methanol using the Lurgi process are based on a SEP report concerning co-production of methanol and power (Lensink - van der Veen, 1996a). The Fuel Grade Methanol co-production unit consists of a saturator/desaturator system, a CO-shift reaction section based on a Fe/Cr catalyst, a CO_2-

removal section based on Selexol and a methanol synthesis section. For producing AA Grade Methanol, a distillation section is added. The block scheme of the methanol co-production unit is given in Appendix B.

CO-shift reaction section and saturator/desaturator system
Cleaned syngas (feed) enters the saturator/CO-shift/desaturator section at 37.5 °C. 67% of the syngas is fed to the saturator. In the saturator, the syngas is saturated with water of about 200 °C. The syngas leaving the saturator is preheated with the reactor effluent and is fed to the CO-shift reactor. Medium pressure steam is added to increase the steam/dry gas ratio. The effluent of the reactor is cooled with the feed and is mixed with the remaining syngas. To protect the catalyst of the methanol synthesis against high sulphur concentrations, the stream passes through a ZnO guard bed. The effluent of the guard bed is fed to the desaturator, where it is cooled first to 140 °C with the saturator bottom stream. Further cooling to 40 °C is conducted with cooling water.

Carbon dioxide removal
The hydrogen-rich gas is fed to the CO_2-removal unit. The unit consists of an absorber, in which the carbon dioxide is absorbed by the solvent Selexol, and a regeneration section. The main part of the carbon dioxide present in the gas is removed. The remaining product gas contains 1.3% carbon dioxide. The purge gas is vented to the atmosphere. Next to water and carbon dioxide, the purge gas contains also traces of carbon monoxide, hydrogen, argon and nitrogen.

Methanol synthesis
The syngas from the CO_2-removal unit is brought to 98.7 bar and 150 °C in a two-stage compressor with intercooling. Next, the gas is fed to a second compressor, together with the recycle stream. The outlet pressure is 101.2 bara. The stream is heated to 230 °C with the effluent of the methanol reactor and is fed to the reactor entrance. The reactor is a cooled multi-tube reactor that is operated isothermally at 255 °C. The reaction takes place in the gas phase inside the tubes over a catalyst. The heat of reaction is removed by generating steam on the shell side.

Distillation of methanol
The Fuel Grade Methanol from the methanol synthesis is reduced in pressure in two steps. The stream is then fed to the distillation section. In the distillation unit, methanol is purified to AA Grade Methanol. The recovery of the distillation section is 97.7%. The section consists of three columns: a pre-run column and two purification columns.

Maturity of the technology
The Lurgi methanol process with prior adjustment of the H_2/CO ratio is a conventional and proven way of producing methanol. Up to 1999, twenty-nine methanol plants have been built using Lurgi's methanol technology (Hydrocarbon Processing, 1999).

Flexibility of the process
The methanol synthesis section is rather flexible and can be held on a hot standby by injecting steam into the cooling jacket. The amount of steam injected to keep the reactor hot is taken to be small with respect to the amount of steam that is produced while the section is on stream. In case of AA Grade Methanol production, also the distillation section does not pose a problem, because the distillation train is fed from an intermediate Fuel Grade Methanol storage. By increasing the size of this tank, it can serve as a buffer between peak and off-peak hours. This would enable the reduction of the size of the methanol distillation train to a capacity half of that of the methanol reactor.

However, the flexibility of the Lurgi methanol co-production is limited by the limited flexibility of the CO-shift section in a similar way as explained for the hydrogen co-production option.

Therefore, it is assumed that also the Lurgi methanol co-production has to be maintained at 20% capacity during peak hours and that the heat and mass balances can be scaled down linearly in the same ratio. It should be noted that this leads to a lower electricity production during peak hours compared to the base case, which will have to be accounted for in the financial-economic evaluation.

Operating hours

The assumed number of operating hours for the co-production plant is 3650 off-peak hours per year and 7000 hours per year in total, corresponding to approx. 80% availability. This is lower than the 85% availability mentioned in the SEP report on which the given data are based.

3.2.1.1. Process analysis Lurgi Fuel Grade Methanol co-production

Specification of process streams

The specification of the process streams entering and leaving the Lurgi Fuel Grade Methanol co-production section is given in Appendix B. All streams entering and leaving the co-production section are summarised in Table 3. This table also gives the streams on a tonne/h basis, a tonne/3650h (yearly off-peak production) basis, a tonne/4320h basis and a tonne/7000h (yearly full production) basis.

Table 3. *Process streams, major enthalpy flows and electricity entering and leaving the Fuel Grade Methanol co-production section*

Process Streams

Name	kg/s		t/h		t/t MeOH		t/y (3650 h/y basis)		t/y (4320 h/y basis)		t/y (7000 h/y basis)	
	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
Syngas	18,56		66,8		2,26		243862		288626		467681	
22 bar steam	4,75		17,1		0,58		62367		73815		119607	
5 bar steam	0,03		0,1		0,00		369		437		708	
Make-up water	25,64		92,3		3,13		336884		398723		646079	
Cooling water	76,04		273,8		9,27		999214		1182631		1916301	
Fuel grade Methanol		8,20		29,5		1,00		107780		127564		206702
CO2 to atmosphere		13,00		46,8		1,58		170769		202115		327501
purge MEOH cycle to CC		2,72		9,8		0,33		35798		42369		68654
Condensate		0,03		0,1		0,00		369		437		708
Cooling water		76,04		273,8		9,27		999214		1182631		1916301
Water to make-up		25,02		90,1		3,05		328766		389115		630511
Total	125,01	125,01	450,1	450,1	15,24	15,24	1642696	1642696	1944232	1944232	3150376	3150376

Major enthalpy flows (MWth)

	MWth		MWh/h		MWh/t MeOH		MWh/y (3650 h/y)		MWh/y (4320 h/y)		MWh/y (7000 h/y)	
Syngas	218,1		218,1		7,4		795939		942043		1526459	
22 bar steam	2,4		2,4		0,1		8662		10252		16612	
5 bar steam	0,0		0,0		0,0		51		61		98	
Fuel grade Methanol		156,8		156,8		5,3		572156		677182		1097285
purge MEOH cycle to CC		18,1		18,1		0,6		65939		78043		126459
Balance		45,6		45,6		1,5		166557		197131		319425
Total	220,5	220,5	220,5	220,5	7,5	7,5	804652	804652	952356	952356	1543169	1543169

Electricity Usage

	MWe	MWh/h	MWh/t MeOH	MWh/y (3650 h/y)	MWh/y (4320 h/y)	MWh/y (7000 h/y)
Electricity	7,22	7,22	0,24	26359	31197	50551

Steam Consumption translated to Electricity Usage

	MWe	MWh/h	MWh/t MeOH	MWh/y (3650 h/y)	MWh/y (4320 h/y)	MWh/y (7000 h/y)
Electricity	0,84	0,84	0,03	3050	3609	5849

note 1: electricity generating efficiency from steam is taken to be 35%

note 2: 100% of production capacity during off-peak hours (3650 h/y) and 20% of production capacity during peak hours (3350 h/y) is equivalent to 4320 h/y operating on 100% of production capacity

note 3: t = tonne

3.2.1.2. Investment and O&M costs Lurgi Fuel Grade Methanol co-production

Investment costs

The total investment costs for the Lurgi Fuel Grade Methanol co-production plant are 100.1 $MNLG_{2000}$. The build-up of the investment costs is given in Table 4. The costs were calculated using the SEP data (Lensink - van der Veen, 1996a). The SEP cost basis is year-1995 NLG for a 1398 MWth turnkey plant based in the Netherlands. The costs were converted into year-2000 NLG and adjusted for the capacity: the investment costs were divided by the number of parallel trains in the original SEP design and scaled down using the 0.65-rule of thumb.

Table 4. *Investment costs for Lurgi Fuel Grade Methanol co-production*

Process section	Investment costs (MNLG)
Saturator/CO-shift section/desaturator	10.5
Selexol CO_2-removal unit	48.3
Methanol synthesis section	41.4
Total	100.1

Operating and maintenance costs

The operating and maintenance costs were calculated as 4% of the investment costs, just as in the SEP report. Added to this were the CO-shift catalyst costs and the methanol catalyst costs. The yearly ZnO guard bed costs were taken to be negligible (Shell, 2000b). The CO-shift catalyst costs are 0.2 MNLG/year (estimate based on Hendriks, 1994). The methanol catalyst costs are 1.9 MNLG/year (estimate based on Udengaard, 1993). The resulting total O&M costs are 5.1 MNLG/year.

3.2.1.3. Process analysis Lurgi AA Grade Methanol co-production

Specification of process streams

The specification of the process streams entering and leaving the Lurgi AA Grade Methanol co-production section is given in Appendix B. All streams entering and leaving the co-production section are summarised in Table 5. This table also gives the streams on a tonne/h basis, a tonne/3650h (yearly off-peak production) basis, a tonne/4320h basis and a tonne/7000h (yearly full production) basis.

Table 5. *Process streams, major enthalpy flows and electricity entering and leaving the Lurgi AA Grade Methanol co-production section*

Process Streams												
Name	kg/s		t/h		t/t MeOH		t/y (3650 h/y basis)		t/y (4320 h/y basis)		t/y (7000 h/y basis)	
	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
Syngas	18,97		68,29		2,33		249248		295000		478009	
22 bar steam	4,85		17,46		0,60		63744		75445		122249	
5 bar steam	70,67		254,42		8,67		928616		1099074		1780907	
Make-up water	26,20		94,34		3,21		344324		407529		660347	
Cooling water *	77,72		279,80		9,54		1021281		1208749		1958620	
AA grade Methanol		8,15		29,34		1,00		107101		126761		205400
CO2 to atmosphere		13,28		47,82		1,63		174540		206579		334734
Impurities to CC		0,23		0,84		0,03		3059		3620		5866
Purge MEOH cycle to CC		2,78		10,02		0,34		36588		43305		70170
Condensate		70,67		254,42		8,67		928616		1099074		1780907
Cooling water		77,72		279,80		9,54		1021281		1208749		1958620
Water to make-up		25,57		92,06		3,14		336027		397709		644435
Total	198,42	198,42	714,30	714,30	24,34	24,34	2607212	2607212	3085796	3085796	5000133	5000133

*) Cooling water for distillation not included

Major enthalpy flows (MWth)												
	MWth		MWh/h		MWh/t MeOH		MWh/y (3650 h/y)		MWh/y (4320 h/y)		MWh/y (7000 h/y)	
Syngas	222,88		222,88		7,60		813517		962847		1560169	
22 bar steam	2,43		2,43		0,08		8853		10478		16979	
5 bar steam	35,34		35,34		1,20		128974		152649		247348	
AA grade Methanol		155,80		155,80		5,31		568670		673056		1090600
Purge & Impurities to CC		22,88		22,88		0,78		83517		98847		160169
Balance		81,96		81,96		2,79		299158		354071		573727
Total	260,6	260,6	260,64	260,64	8,88	8,88	951344	951344	1125975	1125975	1824496	1824496

Electricity Usage						
	MWe	MWh/h	MWh/t MeOH	MWh/y (3650 h/y)	MWh/y (4320 h/y)	MWh/y (7000 h/y)
Electricity	7,57	7,57	0,26	27639	32713	53006

Steam Consumption translated to Electricity Usage						
	MWe	MWh/h	MWh/t MeOH	MWh/y (3650 h/y)	MWh/y (4320 h/y)	MWh/y (7000 h/y)
Electricity	13,22	13,22	0,45	48240	57095	92515

note 1: electricity generating efficiency from steam is taken to be 35%

note 2: 100% of production capacity during off-peak hours (3650 h/y) and 20% of production capacity during peak hours (3350 h/y) is equivalent to 4320 h/y operating on 100% of production capacity

note 3: t = tonne

3.2.1.4. Investment and O&M costs Lurgi AA Grade Methanol

Investment costs

The total investment costs for the Lurgi AA Grade Methanol co-production plant are 114.5 $MNLG_{2000}$. The build-up of the investment costs is given in Table 6. The costs were calculated using the SEP data (Lensink - van der Veen, 1996a). The SEP cost basis is year-1995 NLG for a 1398 MWth turnkey plant based in the Netherlands. The costs were converted into year-2000 NLG and adjusted for the capacity: the investment costs were divided by the number of parallel trains in the original SEP design and scaled down using the 0.65-rule of thumb.

Table 6. *Investment costs for Lurgi AA Grade Methanol co-production*

Process section	Investment costs (MNLG)
Saturator/CO-shift section/desaturator	10.6
Selexol CO_2-removal unit	49.0
Methanol synthesis section	42.0
Methanol distillation section	12.9
Total	114.5

Operating and maintenance costs

The operating and maintenance costs were calculated as 4% of the investment costs, just as in the SEP report. Added to this were the shift catalyst costs and the methanol catalyst costs. The yearly ZnO guard bed costs were taken to be negligible. The shift catalyst costs are 0.2 MNLG/year (estimate based on Hendriks, 1994). The methanol catalyst costs are 1.9 MNLG/year (estimate based on Udengaard, 1993). The resulting total O&M costs are 6.7 MNLG/year.

3.2.2. LPMEOH Process

A demonstration plant of Air Products' Liquid Phase Methanol (LPMEOH^{TM}) process is running since 1997 at Eastman Chemical Company's chemicals-from-coal complex at Kingsport, Tennessee (Diamond et al., 1999). A process flow diagram of the demonstration unit is given in Appendix C. This Appendix also shows the slurry bubble column reactor, which is the heart of the LPMEOH process.

Conventional methanol reactors use fixed beds of catalyst pellets and operate in the gas phase. The LPMEOH reactor uses catalyst in powder form, slurried in an inert mineral oil. The mineral oil acts as a temperature moderator and a heat removal medium, transferring the heat of reaction from the catalyst surface via the liquid slurry to boiling water in an internal tubular heat exchanger. The reactor has the capability to remove heat and maintain a constant uniform temperature over the entire length of the reactor. It can achieve a much higher syngas conversion per pass than its conventional counterparts.

Furthermore, because of the LPMEOH reactor’s unique temperature control capabilities, it can directly process syngas that is rich in carbon oxides (carbon monoxide and carbon dioxide), as produced by oxygen-blown gasification of coal, petroleum coke or other hydrocarbon feedstocks. For these feedstocks, gas phase methanol technology as applied, e.g., in the Lurgi process generally requires stoichiometric adjustment of the H_2/CO ratio in the syngas by the water-gas shift reaction, to increase the hydrogen content, and subsequent carbon dioxide removal. Alternatively, temperature moderation could be achieved by recycling large quantities of hydrogen rich gas, utilising the higher heat capacity of hydrogen, as compared to carbon monoxide.

Another feature of the LPMEOH process is that a high quality methanol product (generally greater than 97%) is produced directly from syngas rich in carbon oxides. Further, the process allows withdrawing spent catalyst slurry and adding fresh catalyst on-line periodically. This facilitates uninterrupted operations and allows perpetuation of high productivity in the reactor.

In either base-load or cycling operation, partial conversion of between 20% and 33% of the IGCC plant’s syngas is optimal, and conversion of up to 50% is feasible. The required degree of conversion of syngas, or the quantity of methanol relative to the power plant size, determines the design configuration for the LPMEOH plant. In its simplest configuration, syngas at maximum available pressure from the IGCC power plant's gasifier system passes once-through the LPMEOH plant and is partially converted to methanol without recycle, CO-shift, or CO_2 removal. The unreacted gas is returned to the IGCC power plant's gas turbine. If greater syngas conversion is required, different plant design options are available.

Flexibility of the process

A characteristic feature of the LPMEOH reactor is its robust character and flexibility. The slurry reactor is suitable for rapid ramping, idling, and even extreme stop/start actions. The thermal moderation provided by the liquid inventory in the reactor acts to buffer sharp transient operations. This characteristic is especially advantageous in the environment of electricity demand load following in IGCC facilities. Since there is no prior adjustment of the H_2/CO ratio, there are no limitations of a CO-shift section.

Maturity of the technology

The first full-scale plant based on the LPMEOH technology is now being demonstrated on a 300-tonne/day basis at Eastman Chemical Company's chemicals-from-coal complex at Kingsport, Tennessee. The plant started operating in 1997 and a demonstration test plan is running until 2001. Simulation of IGCC co-production will be performed in 2000-2001. This includes testing of syngas compositions of different gasifiers (amongst others Shell gasifier syngas composition) and testing of rapid ramping and stop/start operations (hot and cold standby).

Limited syngas conversion: more than 260 MWth syngas to the combined cycle

For the configuration considered, the syngas conversion of the LPMEOH process is limited by the hydrogen content of the CO-rich gas. 36% conversion on a LHV basis is theoretically possible. However, due to design considerations a conversion of 30% is preferred. Because of this limitation, the minimum flow of syngas to the combined cycle increases to about 320 MWth. This is 23% more than the preferred minimum of 260 MWth, which is the minimum load for the combined cycle itself.

A higher methanol production is feasible. Firstly, measures could be taken in the gasifier. By adding water to the gasifier, the water-gas shift reaction will increase the hydrogen content of the syngas. However, this will also increase the CO_2 content resulting in a larger Sulfinol make-up in the desulphurisation unit. Secondly, a CO-shift reactor could be added to the LPMEOH configuration, but this would seriously limit the flexibility of the co-production unit as explained for the hydrogen and Lurgi methanol co-production options. Neither of these options is further dealt with here.

Operating hours

The LPMEOH plant can achieve an availability of over 99.7%. This was shown at the LPMEOH demonstration plant at Eastman Chemical's site at Kingsport, Tennessee (Tijm et al., 1999).

3.2.2.1. Process analysis LPMEOH Fuel Grade Methanol co-production

Specification of process streams

The specification of the main process streams entering and leaving the once-through LPMEOH Fuel Grade Methanol co-production section is given in Appendix C. The specification is restricted to the incoming syngas and the leaving methanol, since detailed information on other streams (steam, cooling water, etc.) and on electricity usage is not available. The main process streams are summarised in Table 7.

Table 7. *Main process streams and enthalpy flows entering and leaving the LPMEOH Fuel Grade Methanol co-production section*

Process Streams

Name	kg/s		t/h		t/t MeOH		t/y (3650 h/y basis)		t/y (7000 h/y basis)	
	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
Syngas	39,15		140,94		6,82		514417		986553	
fuel grade Methanol		5,74		20,65		1,00		75385		144574
Balance (including purge and steam)		33,41		120,28		5,82		439032		841979
Total	39,15	39,15	140,94	140,94	6,82	6,82	514417	514417	986553	986553

Major enthalpy flows (MWth)

	MWth		MWh/h		MWh/t MeOH		MWh/y (3650 h/y)		MWh/y (7000 h/y)	
Syngas	460,00		460,00		22,27		1679000		3220000	
fuel grade Methanol		110,15		110,15		5,33		402053		771061
Purge to CC		318,78		318,78		15,43		1163562		2231489
Steam production		19,51		19,51		0,94		71212		136570
Balance		11,55		11,55		0,56		42173		80879
Total	460,0	460,0	460,00	460,00	22,27	22,27	1679000	1679000	3220000	3220000

Steam Production translated to Electricity Production

	MWe	MWh/h	MWh/t MeOH	MWh/y (3650 h/y)	MWh/y (7000 h/y)
Electricity	6,83	6,83	0,33	24924	47800

note 1: electricity generating efficiency from steam is taken to be 35%
note 2: t = tonne

3.2.2.2. Investment and O&M costs LPMEOH Fuel Grade Methanol co-production

Investment Costs

The total investment costs for the LPMEOH Fuel Grade Methanol co-production plant are 68.0 $MNLG_{2000}$. The costs were calculated from (Drown et al., 1997) and are backed up by previous personal communication with Air Products (Air Products, 1999). The cost basis provided by Air Products is in US dollar, location US Gulf Coast for a turnkey 340 tonne methanol/day plant. The figures were converted into year-2000 US dollar. Scaling down the investment costs was done by the 0.65-rule of thumb. The US dollars were converted to Dutch guilders (NLG) using a 1996 exchange rate of 1.69 NLG/US dollar. Next, the costs were raised with an inflation figure of 9% over the period 1996 to 2000. No location factor was applied to account for the Buggenum location instead of the US Gulf Coast site.

Operating and maintenance costs

The operating and maintenance costs were calculated from data given by (Drown et al., 1997). This includes all O&M costs, including catalyst cost. The resulting total O&M costs are 3.9 MNLG/year.

3.2.2.3. Process analysis LPMEOH AA Grade Methanol co-production

Specification of process streams

The specification of the main process streams entering and leaving the once-through AA Grade Methanol co-production section is given in Appendix C. The specification is restricted to the incoming syngas and the leaving methanol, since detailed information on other process streams (steam, cooling water, etc.) and on electricity usage is not available. The main streams are summarised in Table 8. This table also gives the streams on a tonne/h basis, a tonne/3650h (yearly off-peak production) basis and a tonne/7000h (yearly full production) basis.

Table 8. *Main process streams and enthalpy flows entering and leaving the LPMEOH AA Grade Methanol co-production section*

Process Streams										
Name	kg/s		t/h		t/t MeOH		t/y (3650 h/y basis)		t/y (7000 h/y basis)	
	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
Syngas	39,15		140,94		7,02		514417		986553	
AA grade Methanol		5,58		20,08		1,00		73292		140559
Balance (including purge)		33,57		120,86		6,02		441125		845994
Total	39,15	39,15	140,94	140,94	7,02	7,02	514417	514417	986553	986553

Major enthalpy flows (MWth)										
	MWth		MWh/h		MWh/t MeOH		MWh/y (3650 h/y)		MWh/y (7000 h/y)	
Syngas	460,00		460,0		22,91		1679000		3220000	
Steam	22,03		22,0		1,10		80428		154245	
AA grade Methanol		107,11		107,1		5,33		390970		749805
Purge to CC		321,82		321,8		16,03		1174646		2252746
Balance		53,10		53,1		2,64		193812		371694
Total	482,0	482,0	482,0	482,0	24,01	24,01	1759428	1759428	3374245	3374245

Steam Consumption translated to Electricity Consumption					
	MWe	MWh/h	MWh/t MeOH	MWh/y (3650 h/y)	MWh/y (7000 h/y)
Electricity	7,71	7,71	0,38	28150	53986

note 1: electricity generating efficiency from steam is taken to be 35%
note 2: t = tonne

3.2.2.4. Investment and O&M costs LPMEOH AA Grade Methanol co-production

Investment Costs

The investment costs for the LPMEOH AA Grade Methanol co-production plant including a methanol distillation section are 86.8 $MNLG_{2000}$. The costs were calculated from (Drown et al., 1997) and are backed up by previous personal communication with Air Products (Air Products, 1999). The cost basis provided by Air Products is in US dollar, location US Gulf Coast for a turnkey 340 tonne methanol/day plant. The figures were escalated to year-2000 US dollar. Scaling down the investment costs was done by the 0.65-rule of thumb. The US dollars were converted to Dutch guilders (NLG) using a 1996 exchange rate of 1.69 NLG/US dollar. Next, the costs were raised with an inflation figure of 9% over the period 1996 to 2000. No location factor was applied to account for the Buggenum instead of the US Gulf Coast site.

Operating and maintenance costs

The operating and maintenance costs were calculated from data given by (Drown et al., 1997). This includes all O&M costs, including catalyst cost. The resulting total O&M costs are 4.2 MNLG/year.

3.3. Fischer-Tropsch transportation fuels

The once-through Fischer-Tropsch transportation fuels co-production section consists of a saturator/CO-shift/desaturator section, a Selexol CO_2-removal section, a hydrogen production section, and a 'Fischer-Tropsch' section. A schematic representation of the Fischer-Tropsch transportation fuels co-production unit is given in Figure 2.

For the saturator/CO-shift/desaturator section and the CO_2-removal section, the process information and the economic data are based on the SEP report on the co-production of methanol and power (Lensink - van der Veen, 1996a), just as for the Lurgi methanol co-production option. For the hydrogen production section, the information is based on the SEP report on the co-production of hydrogen and power (Lensink - van der Veen, 1996b). For the Fischer-Tropsch section, the information is based on data provided by Shell (Shell, 2000a) and data reported in open literature.

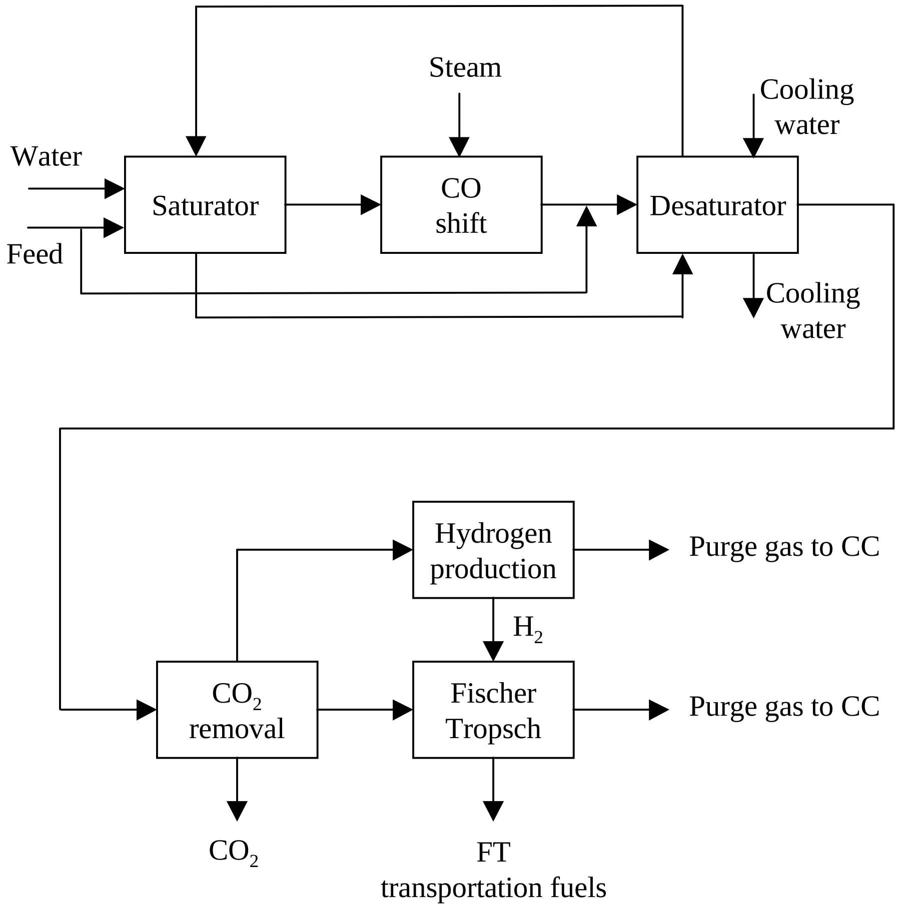

Figure 2. *Block scheme of the Fischer-Tropsch transportation fuels co-production section*

CO-shift reaction section and saturator/desaturator system

The H_2/CO-ratio of the syngas from the gasifier is 0.5, whereas the Fischer-Tropsch synthesis requires an H_2/CO-ratio of 2 (Shell, 2000a). In the CO-shift section, CO is converted with steam into H_2 to shift the H_2/CO-ratio to the required value.

Cleaned syngas (feed) enters the saturator/CO-shift/desaturator section at 37.5 °C. 67% of the syngas is fed to the saturator. In the saturator, the gas is saturated with water of about 200 °C. The syngas leaving the saturator is preheated with reactor effluent and is fed to the shift reactor. Medium pressure steam is added to increase the steam/dry gas ratio. The effluent of the shift reactor is cooled with the feed and is mixed with the remaining syngas. To protect the catalyst of the Fischer-Tropsch synthesis against high sulphur concentrations, the stream passes through a ZnO guard bed. The effluent of the guard bed is fed to the desaturator, where it is cooled first to 140 °C with the saturator bottom stream. Further cooling to 40 °C is conducted with cooling water.

Carbon dioxide removal

The gas leaving the desaturator contains approximately 30 vol.% inert gases (CO_2, N_2, Ar and H_2O). The Fischer-Tropsch synthesis requires that the amount of inert gases is reduced to ~10 vol.% (Shell, 2000a).

The amount of inert gases is reduced by carbon dioxide removal. This section consists of an absorber, where the solvent Selexol absorbs the carbon dioxide, and a regeneration section. The major amount of the carbon dioxide is removed. The remaining product gas contains 10 vol.%

of inert gases. The purge gas is vented to the atmosphere. Besides water and carbon dioxide, the purge gas contains also traces of carbon monoxide, hydrogen, argon and nitrogen.

Hydrogen production

In the Fischer-Tropsch section, the syngas is first converted into high molecular waxes that are subsequently converted into liquid fuels by hydrocracking. For hydrocracking an amount of hydrogen equal to 1 wt.% of the Fischer-Tropsch wax produced is required (Shell, 2000a).

A small part of the gas leaving the carbon dioxide removal section is routed to the hydrogen production section. Hydrogen is separated from the gas by pressure swing adsorption. The hydrogen concentration of the feed gas is ~60 vol.%. It has been assumed that 75% hydrogen recovery is accomplished. The PSA purge gas contains combustible components. Therefore, the purge gas is pressurised by a compressor and fed to the combined cycle.

Fischer-Tropsch section

The once-through Fischer-Tropsch section is based on the Shell Middle Distillate Synthesis (SMDS) process (Shell, 1992). In the first step, the actual Fischer-Tropsch synthesis, the syngas is converted into high molecular waxes. The waxes are subsequently converted into transportation fuels by hydrocracking. Various distillation sections separate the products of the hydrocracking process (mainly gas oil, kerosene and naphtha).

For the mass balance the following product yields have been used (Shell, 2000a):
Fischer-Tropsch transportation fuels : 18 wt.%
Process water : 22 wt.%
Purge gas : 60 wt.%
Furthermore, it has been assumed that 20% of the chemical energy converted is released in the form of heat, or the production of 4 MJ Fischer-Tropsch transportation fuels is accompanied by the production of 1 MJ heat.

Maturity of the technology

All process steps are mature technology. The CO-shift reaction, CO_2-removal by the Selexol process and H_2 production by Pressure Swing Adsorption (PSA) are used in a large number of commercial plants. Only one commercial SMDS plant has been built up to now in Bintulu, Sarawak, Malaysia. The capacity of this plant is 12500 barrels per day (Tijm et al., 1994).

Flexibility of the process

The Fischer-Tropsch transportation fuels co-production section is not very flexible with respect to rapid start/stop operations and ramping. It has been assumed that the co-production can be turned down to 20% of its capacity during peak hours for electricity production, but it has to be verified whether this is possible. It should be noted that this leads to a lower electricity production during peak hours compared to the base case, which will have to be accounted for in the financial-economic evaluation.

3.3.1. Process analysis Fischer-Tropsch transportation fuels co-production

All streams entering and leaving the once-through Fischer-Tropsch transportation fuels co-production section are summarised in Table 9.

Table 9. *Process streams, major enthalpy flows and electricity entering and leaving the Fischer-Tropsch transportation fuels co-production section*

Process streams										
Name	kg/s		t/h		t/t FT product		t/y (3650 h/y basis)		t/y (7000 h/y basis)	
	IN	OUT	IN	OUT	IN	OUT	IN	OUT	IN	OUT
Syngas	29.59		106.51		9.13		388765		745578	
22 bar steam	6.83		24.60		2.11		89802		172223	
5 bar steam	0.04		0.16		0.01		572		1098	
Make-up water	0.89		3.21		0.28		11722		22480	
FT products		3.24		11.67		1.00		42596		81691
Purge FT to CC		10.81		38.90		3.33		141987		272304
Purge PSA to CC		0.43		1.54		0.13		5630		10797
CO_2 to atmosphere		18.87		67.95		5.82		248013		475642
Water FT-process		3.96		14.26		1.22		52062		99845
Condensate		0.04		0.16		0.01		572		1098
Total	37.36	37.36	134.48	134.48	11.52	11.52	490861	490861	941378	941378

Major enthalpy flows										
	MWth		MWh/h		MWh/t FT product		MWh/h (3650 h/y)		MWh/h (7000 h/y)	
Syngas	346.4		346.4		29.69		1264528		2425122	
22 bar steam	3.4		3.4		0.29		12472		23920	
5 bar steam	0.0		0.0		0.00		79		152	
FT products		145.9		145.9		12.50		532452		1021141
Purge FT to CC		142.1		142.1		12.18		518724		994813
Purge PSA to CC		4.3		4.3		0.37		15804		30310
Balance		57.6		57.6		4.93		210099		402930
Total	349.9	349.9	349.9	349.9	29.98	29.98	1277080	1277080	2449194	2449194

Electricity usage					
	MWe	MWh/h	MWh/t FT product	MWh/h (3650 h/y)	MWh/h (7000 h/y)
Electricity	3.16	3.16	0.27	11517	22087

Steam consumption translated into electricity					
	MWe	MWh/h	MWh/t FT	MWh/h (3650 h/y)	MWh/h (7000 h/y)
Electricity[*]	1.20	1.20	0.10	4393	8425

Note 1: electricity generating efficiency from steam is taken to be 35%
Note 2: t = metric ton

3.3.2. Investment and O&M costs Fischer-Tropsch transportation fuels co-production

Investment costs

The total investments costs for the Fischer-Tropsch transportation fuels co-production unit are 168.9 MNLG. The costs for the different sections are given in Table 10.

Table 10. *Investment costs for Fischer-Tropsch transportation fuels co-production*

Process section	Investment costs (MLNG)
Saturator/CO-shift section/desaturator	14.2
Selexol CO_2-removal unit	61.5
Hydrogen production section	0.8
Fischer-Tropsch section (incl. Fischer-Tropsch synthesis, hydrocracking and product fractionation)	92.3
Total	168.9

The investment costs for the CO-shift reaction, CO_2-removal and hydrogen production sections are based on the SEP reports on the co-production of methanol and power (Lensink - van der Veen, 1996a), and on the co-production of hydrogen and power (Lensink - van der Veen, 1996b). The costs were converted to year 2000 NLG and scaled with an exponent of 0.65.

The costs for a complete natural-gas based 20000 Barrel Per Day (BPD) Fischer-Tropsch plant (including syngas production) are approximately 30000 $/BPD (Corke, 1998). The investment costs for the Fischer-Tropsch section, including synthesis, hydrocracking and product fractionation, have been estimated to be about 30% of the total investments costs, or 9000 $/BPD for a 20000 BPD plant. The costs for the Fischer-Tropsch section have been scaled with an exponent of 0.66 (Corke, 1998) down to ~2500 BPD, the size considered in this study.

Operating and maintenance costs

The operating and maintenance costs were estimated to be 4% of the investment costs, just as for the co-production of hydrogen and methanol. Data on the costs of Fischer-Tropsch catalysts are confidential and not reported. As a rough estimate, the costs for catalysts were assumed equal to the catalysts costs for the methanol co-production section, i.e. 0.2 MNLG/year for the CO-shift catalyst, 1.9 MNLG/year for the Fischer-Tropsch synthesis catalyst and negligible costs for the ZnO guard bed. This resulted in total O&M costs of 8.9 MNLG/year.

3.4. Substitute Natural Gas (SNG)

The definition and technical evaluation of the SNG co-production option was conducted separately by GASTEC (Van Rens, 2000). An English translation of the summary of the final report by GASTEC is included as Appendix D. Unfortunately, there are many substantial differences in starting-points and evaluation approach between the GASTEC study and the ECN evaluations. This makes it impossible to compare the results as presented in Appendix D directly with the results for the other options.

In this paragraph, a short description of the SNG option is given. The differences will be discussed in more detail in Paragraph 4.2.5. The SNG co-production section according to the GASTEC design is shown schematically in Figure 3.

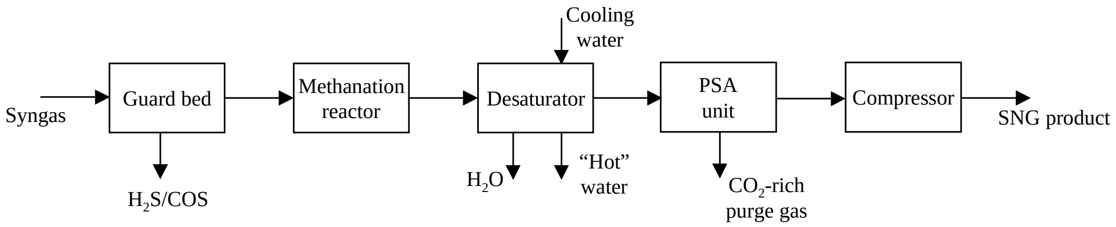

Figure 3. *Block scheme of the SNG co-production section*

Cleaned syngas (feed) enters the co-production at 130 °C. It is assumed that the syngas is saturated with water and that the sulphur content has been brought down to 1 ppm in an extra scrubber column. The syngas is first passed through a guard bed to remove remaining sulphur compounds down to below 100 ppb. Leaving the guard bed, the syngas is preheated with the methanation reactor effluent and steam is added to increase the steam/dry gas ratio. In the subsequent methanation reactor, methanation and the CO-shift reaction occur simultaneously. Excess steam has to be added to avoid Boudouard-C formation. The exothermal methanation reaction is controlled by transferring the heat of reaction to boiling water in an internal heat exchanger.

The effluent of the methanation reactor is rich in H_2O, which is removed mainly by condensation through water-cooling. This results in a hot-water byproduct stream, the heat of which is converted into electricity in an Organic Rankine Cycle (ORC) system. Following water condensation, the methane-rich stream is sent to a Pressure Swing Absorption (PSA) unit for further drying and CO_2-removal to adjust the Wobbe-index to the requirements of the natural gas network. For the layout and costing of the PSA unit, it is assumed that CO_2 instead of N_2 is applied for inertisation and coal feeding to the gasifier, leading to only 1 instead of 6.8 vol.% N_2 in the syngas, and that the remaining CO_2-concentration in the SNG product stream is 10 vol.%. For the purge gas of the PSA unit it is mentioned that, after combustion in a catalytic afterburner, the gas might be used for CO_2-fertilisation in greenhouses. However, in the further evaluation this purge gas with a total heating value of 18 MW (LHV) is not taken into account at all.

The SNG product stream is finally compressed to the required pressure level of the gas network, which amounts to 41 or 18 bar depending on the network selected.

The process flow diagram, as taken from the GASTEC report, is included in Appendix D as well.

Maturity of the technology

This issue is not addressed in the GASTEC report.

Flexibility of the process

In the GASTEC report, it is stated that the SNG co-production has to be maintained at minimal 10% capacity during peak hours.

4. FINANCIAL EVALUATION

4.1. General assumptions and input data

The financial evaluation of the five different options has been conducted on the basis of additional costs and additional income compared with a base case in which only electricity is produced. Therefore, the results do not provide information on the viability of the Demkolec plant as a whole, but solely on the financial viability of the co-production options. The base case is the operating mode for the Demkolec plant in 2000. This consisted of full-load (460 MWth) power generation during peak hours and operation at minimum load (260 MWth), as determined by the gas turbine, during off-peak hours. However, also for the base case it is assumed that switching to the new fuel mixture, described in Paragraph 4.1.1, has been accomplished already.

For all the co-production options, except for LPMEOH methanol co-production (see Paragraph 3.2.2) and the SNG co-production (see Paragraph 4.2.5), the calculations have been based on the assumption that during off-peak hours a fixed fraction of the syngas (260 MWth due to the minimum load requirements of the gas turbine) is used for electricity production and the remaining part (200 MWth.) for the production of liquid or gaseous fuels. During peak hours, the plant is used for electricity production only and the co-processing plant is turned down as far as possible.

An option to further increase the attractiveness of the Buggenum plant could be to remove the CO_2 that is emitted into the atmosphere by the co-production plant and to store it in aquifers, empty gas fields or in coal beds. This option has not been included in the calculations. However, in a separate paragraph some general remarks are made on this option.

4.1.1. The fuel mix of the plant

Demkolec is planning to switch from gasifying solely fossil fuels to other fuels that are of less interest to other conversion plants, and are therefore expected to be available on the longer run at relative low prices. It has been assumed that these fuels can be co-gasified without considerable pre-treatment. The fraction of alternative "green" fuels is envisaged to be 24% of the energy input. The fuel mix used in all cases (base case as well as the co-production cases) consists of the following fractions:

• Fossil fuels (55%)	- coal - gas
• "Green" alternative fuels (24%)	- chicken manure - plastic/paper pellets (25%) - wood fraction of vegetable, fruits and garden waste - industrial wastewater treatment sludge - sewage sludge
• "Non-green" alternative fuels (21%)	- petroleum cokes (in coal blend) - plastic/paper pellets (75%) - sea harbour waste oil

For the plastic and paper pellets fraction, it has been assumed that 25% can be regarded as "green" alternative fuel and 75% as "non-green" alternative fuel.

4.1.2. Fuel and product prices

The following fuel prices have been assumed:

Coal	4.50 NLG/GJ
"Green" alternative fuels	0.92 NLG/GJ
"Non-green" alternative fuels	0.48 NLG/GJ

These prices were provided by Demkolec. The price of the "green" alternative fuel is based on a limited percentage of industrial wastewater treatment sludge and sewage sludge. Due to good results with recent experiments, the fraction of these sludge streams within the alternative "green" fraction is expected to be raised substantially. In that case, the fuel price could drop from 0.92 NLG/GJ to -2.49 NLG/GJ. This would of course influence the results in the base case situation as well as in the co-production scenarios substantially. For a sensitivity analysis on a change in fuel prices, see Paragraph 4.3.7.

For the calculations, it has been assumed that the Demkolec plant will be operated as a base load plant. In that case, the electricity that is being produced is estimated to be sold for 0.065 NLG/kWh during peak hours and 0.03 NLG/kWh during off-peak hours. In an additional paragraph, some remarks are made on the possibility of operating the Demkolec plant in a regulating power mode. In that case, the price for the electricity can be substantially higher, but the volume of the electricity being produced will most probably be much smaller. The additional premium of the green electricity, valid for peak hours as well as off-peak hours, is estimated to be 0.0877 NLG/kWh (including 0.0427 NLG/kWh REB and 0.045 NLG/kWh green label premium). The following product prices have been assumed:

Hydrogen	10 NLG/GJ
AA Grade Methanol	12.6-25.2 NLG/GJ
Fischer-Tropsch transportation fuels	8.7-11.0 NLG/GJ
SNG	8.8-13.9 NLG/GJ

For hydrogen, this is the present market price. The AA Grade Methanol market price was ranging during the past year between 96 \$/tonne in 1999 to a maximum of 230 \$/tonne in the USA during the month of September of 2000. In Europe the fourth quarter 2000 contract list price FOB Rotterdam showed a price of approximately 224 \$/tonne (information from Methanex Corporation's web-site). Due to the large fluctuations on the market, the price for AA Grade Methanol has been assumed to range between 100 \$/tonne and 200 \$/tonne in the coming years, corresponding with 12.6-25.2 NLG/GJ at an exchange rate of 2.50 NLG/\$. The price for Fuel Grade Methanol is not available on the open market but one can assume that it is directly linked with the price for AA Grade Methanol with an offset for the variable costs of upgrading the quality. The price range for Fischer-Tropsch transportation fuels is based on the assumption that these transportation fuels have a 135-170% higher value than fossil crude oil. The price of fossil crude oil is assumed to be 15 \$/barrel. For Fischer-Tropsch transportation fuels, this means a price ranging from 8.7 NLG/GJ to 11.0 NLG/GJ. The price range for SNG is equal to the current commodity price range for natural gas and corresponds to 0.28-0.44 NLG/m_n^3.

4.1.3. Valuation and amount of green electricity and green liquid/gaseous fuel

Because of present energy and environmental policy, the ‘greenness’ of electricity and fuels represents a potential extra value for the product. For electricity, this value is rather clear and is strongly related to the regulating energy tax. For the other products, the value of greenness is less clear and some assumptions have to be made. In principle, it is assumed that the extra value for greenness for the produced fuels is equalised to that for electricity on a primary input basis.

The value for greenness in the case of electricity amounts to 0.0877 NLG/kWh (24.4 NLG/GJe). Calculating with an estimated general efficiency of the total installed large-scale electric power generating capacity in the Netherlands of 46%, the value for greenness of primary fuels amounts to 11.2 NLG/GJ. This is the extra value for greenness for the produced fuels that has been assumed.

In all cases, it is assumed that 24% of the output can be regarded as green.

4.1.4. Remaining items (option independent)

Financial issues

The results of the financial analysis are expressed as yearly cost or income. Because almost all data used are time independent it did not make much sense to use a cash flow analysis model producing results in terms of an Internal Rate of Return. The capital costs are annuity based, using an interest rate of 6.5% and an economic lifetime of 15 years.

In the different cases no investment subsidies like EIA and VAMIL have been taken into account. These subsidies by the Dutch government could (within certain limitations) amount up to approximately 20% of the investment costs.

Technical issues

In most cases the co-production unit consumes steam that otherwise would have been used in the combined cycle section to generate electricity with a total efficiency of 35%. This leads to a reduction in electricity produced. When the co-production unit produces steam, then this is assumed to lead to an increase in electricity production.

The number of full-load equivalent operating hours during a year amounts to 3350 peak hours and 3650 off-peak hours.

During off-peak operation in the base case (no co-production), the gasifier output is being reduced from 460 MWth to 260 MWth. The efficiency of the gasification plant drops from 77.8% under full-load conditions to 73.3% at a thermal output of 260 MWth. The efficiency of the combined cycle drops from 54% (full load) to 45%. Hence, the overall plant efficiency drops from 42% to 33%. During off-peak operation with co-production, the gasifier output is being kept at 460 MWth. The co-production capacity is designed such that still a minimum of 260 MWth is sent to the combined cycle.

4.2. Option-specific assumptions and input data

In this paragraph, the specific assumptions and input data for the different co-production options are presented. These data are taken from Chapter 3. In the different tables, the net electricity consumption by the co-production section is mentioned. This consumption includes the electricity equivalent of the steam that is produced or consumed by the co-production section, assuming a 35% conversion efficiency of steam into electricity.

4.2.1. Hydrogen

An important assumption for the hydrogen co-production option is that the co-production plant cannot be shut down completely during peak hours. It has been assumed that the production is decreased to 20% of the full-load capacity. This leads to a lower electricity production during peak hours compared to the base case, which has been accounted for in the financial-economic evaluation.

The option-specific assumptions that have been used in the financial evaluation are given in Table 11.

Table 11. *Option-specific input data for Hydrogen co-production*

Item	Value
Investment costs [NLG]	67,800,000
Input to combined cycle - peak hours [MWth]	420
Input to combined cycle - off-peak hours [MWth]	260
Secondary product - peak hours [tonne/year]	3576
Secondary product - off-peak hours [tonne/year]	19477
Lower Heating Value of secondary product [GJ/tonne]	120
Net electricity consumption by co-production section - peak hours [MWh]	3167
Net electricity consumption by co-production section - off-peak hours [MWh]	17256
CO_2 production by co-production section - peak hours [tonne/year]	-*
CO_2 production by co-production section - off-peak hours [tonne/year]	-*
O&M costs [% of total investment costs/year]	4.4

* In this option, there is no (selective) CO_2-removal; the purge gas from the PSA unit is sent to the combined cycle.

4.2.2. Methanol - Lurgi process

Also in case of methanol co-production with the Lurgi process, it has been assumed that the co-production is decreased to 20% of the full-load capacity during peak hours, which leads to a lower electricity production during peak hours compared to the base case.

The option-specific input data for Lurgi Fuel Grade Methanol and Lurgi AA Grade Methanol co-production are presented in Tables 12 and 13 respectively.

Table 12. *Option-specific input data for Lurgi Fuel Grade Methanol co-production*

Item	Value
Investment costs [NLG]	100,100,000
Input to combined cycle - peak hours [MWth]	420
Input to combined cycle - off-peak hours [MWth]	260
Secondary product - peak hours [tonne/year]	19784
Secondary product - off-peak hours [tonne/year]	107780
Lower Heating Value of secondary product [GJ/tonne]	19.8
Net electricity consumption by co-production section - peak hours [MWh]	5397
Net electricity consumption by co-production section - off-peak hours [MWh]	29409
CO_2 production by co-production section - peak hours [tonne/year]	31346
CO_2 production by co-production section - off-peak hours [tonne/year]	170769
O&M costs [% of total investment costs/year]	5.1

Table 13. *Option-specific input data for Lurgi AA Grade Methanol co-production*

Item	Value
Investment costs [NLG]	114,500,000
Input to combined cycle - peak hours [MWth]	420
Input to combined cycle - off-peak hours [MWth]	260
Secondary product - peak hours [tonne/year]	19660
Secondary product - off-peak hours [tonne/year]	107101
Lower Heating Value of secondary product [GJ/tonne]	19.8
Net electricity consumption by co-production section - peak hours [MWh]	13929
Net electricity consumption by co-production section - off-peak hours [MWh]	75879
CO_2 production by co-production section - peak hours [tonne/year]	32039
CO_2 production by co-production section - off-peak hours [tonne/year]	174540
O&M costs [% of total investment costs/year]	5.9

4.2.3. Methanol - LPMEOH process

In using the once-through LPMEOH concept only a limited amount of the syngas can be converted into methanol. This conversion percentage is too low to reduce the thermal power after the co-production unit to "only" 260 MWth. Therefore this option differs from the others in the thermal input to the combined cycle during co-production. This is not 260 MWth, but 320 MWth.

The option-specific input data for LPMEOH Fuel Grade Methanol and LPMEOH AA Grade Methanol co-production are presented in Tables 14 and 15 respectively.

Table 14. *Option-specific input data for Fuel Grade Methanol co-production using the LPMEOH process*

Item	Value
Investment costs [NLG]	68,000,000
Input to combined cycle - peak hours [MWth]	460
Input to combined cycle - off-peak hours [MWth]	320
Secondary product - peak hours [tonne/year]	0
Secondary product - off-peak hours [tonne/year]	75385
Lower Heating Value of secondary product [GJ/tonne]	19.8
Net electricity consumption by co-production section - peak hours [MWh]	0
Net electricity consumption by co-production section - off-peak hours [MWh]	-24924
CO_2 production by co-production section - peak hours [tonne/year]	-*
CO_2 production by co-production section - off-peak hours [tonne/year]	-*
O&M costs [% of total investment costs/year]	5.7

* In this option, there is no CO_2-removal; the purge gas from the co-production section is sent to the combined cycle.

Table 15. *Option-specific input data for AA Grade Methanol co-production using the LPMEOH process*

Item	Value
Investment costs [NLG]	86,800,000
Input to combined cycle - peak hours [MWth]	460
Input to combined cycle - off-peak hours [MWth]	320
Secondary product - peak hours [tonne/year]	0
Secondary product - off-peak hours [tonne/year]	73292
Lower Heating Value of secondary product [GJ/tonne]	19.8
Net electricity consumption by co-production section - peak hours [MWh]	0
Net electricity consumption by co-production section - off-peak hours [MWh]	28150
CO_2 production by co-production section - peak hours [tonne/year]	-*
CO_2 production by co-production section - off-peak hours [tonne/year]	-*
O&M costs [% of total investment costs/year]	4.8

* In this option, there is no CO_2-removal; the purge gas from the co-production section is sent to the combined cycle.

The data on net electricity consumption show that in the LPMEOH process for the co-production of Fuel Grade Methanol the consumption is negative, meaning that the co-production process actually produces steam.

4.2.4. Fischer-Tropsch transportation fuels

Also in case of once-through Fischer-Tropsch transportation fuels co-production, it has been assumed that the co-production is decreased to 20% of the full-load capacity during peak hours, which leads to a lower electricity production during peak hours compared to the base case.

The option-specific input data for Fischer-Tropsch transportation fuels co-production are presented in Table 16.

Table 16. *Option-specific input data for Fischer-Tropsch transportation fuels co-production*

Item	Value
Investment costs [NLG]	168,900,000
Input to combined cycle - peak hours [MWth]	420
Input to combined cycle - off-peak hours [MWth]	260
Secondary product - peak hours [tonne/year]	8519
Secondary product - off-peak hours [tonne/year]	42596
Lower Heating Value of secondary product [GJ/tonne]	43
Net electricity consumption by co-production section - peak hours [MWh]	3182
Net electricity consumption by co-production section - off-peak hours [MWh]	15910
CO_2 production by co-production section - peak hours [tonne/year]	49603
CO_2 production by co-production section - off-peak hours [tonne/year]	248013
O&M costs [% of total investment costs/year]	5.2

4.2.5. SNG

As mentioned in Paragraph 3.4, it is impossible to compare the results as presented for the SNG option in Appendix D directly with the results for the other options, because of many substantial differences in starting-points and evaluation approach between the GASTEC study and the ECN evaluations. Main differences as well as some comments regarding the GASTEC approach are given below.

- Different syngas inlet conditions are applied for the co-production section. In the ECN evaluations dry, clean syngas is fed to the co-production section at 37.5 ^{o}C, whilst in the GASTEC approach saturated, clean syngas is fed at a temperature level of 130 ^{o}C. Compared to the ECN evaluation, the steam for the saturation and temperature increase has to be withdrawn from the combined cycle, but this is not taken into account.
- In the GASTEC study, the purge gas from the PSA unit is not utilised or processed. However, this purge gas cannot be vented to atmosphere due to the presence of combustibles (methane, carbon monoxide, hydrogen). In the ECN evaluations, these purge gas streams are compressed and fed into the combined cycle.
- If operation of the co-production section has to be maintained at a certain minimum load during peak hours, this leads to lost income due to lower electricity production compared to the base case. According to GASTEC, the minimum load for the SNG co-production is 10% (20 MWth syngas), but the lost income is not taken into account.
- In the GASTEC study, also additional green benefits due to extra syngas production during off-peak hours are not included.
- In the GASTEC study, the syngas production costs for the co-production are fixed at 0.244 NLG/ m_n^3 SNG. This value is based on operating the gasifier on coal only. In the ECN evaluations, on the contrary, it has been assumed that the gasifier is operated on the fuel mixture as specified in Paragraph 4.1.1 and that the efficiency of both the gasifier and the combined cycle are load dependent (see Paragraph 4.1.4).
- In the GASTEC study, an Organic Rankine Cycle system is applied to produce 4.7 MWe from the 160 ^{o}C hot water purge stream. However, the investment costs estimated by GASTEC for this system are considered unrealistically low. GASTEC uses specific investment costs of approx. 1200 NLG/kWe, whilst 3000-6000 NLG/kWe is considered a realistic range.

- In the GASTEC study, it is assumed that CO_2 instead of N_2 is used for inertisation and coal feeding to the gasifier, leading to only 1 instead of 6.8 vol.% N_2 in the syngas. It is mentioned that this has a positive effect on the SNG yield at limited additional costs. However, these additional costs are not included and the technical implications are not substantiated.
- The electricity prices assumed in the GASTEC study are 0.085 NLG/kWh (peak hours) and 0.025 NLG/kWh (off-peak hours), whilst in the ECN evaluations these values are 0.065 NLG/kWh and 0.030 NLG/kWh respectively.
- In the GASTEC study, the capital costs are accounted for, using an interest rate of 6.0% and an economic lifetime of 10 years, whilst these values are 6.5% and 15 years respectively in the ECN evaluations.

In an attempt to enable a direct comparison of the SNG option with the other options, the GASTEC data as given in (Van Rens, 2000) have been interpreted and adapted to obtain suitable input data for the ECN evaluation approach. The following adaptations have been made:

- Syngas inlet conditions in agreement with the other options (37.5 ^{o}C, dry).
- Compression of the purge gas stream from the PSA unit and feeding it to the combined cycle.
- Operation of the gasifier on the fuel mixture specified in Paragraph 4.1.1.
- No utilisation or processing of the hot water purge stream. With the higher investment costs, as mentioned above, application of an ORC system is not cost-effective. This may require the changes in the plant design to avoid the production of the large 160 ^{o}C water purge stream. However, this is beyond the scope of this study.

This leads to a set of option-specific input data for SNG co-production as specified in Table 17.

Table 17. *Option-specific input data for SNG co-production*

Item	Value
Investment costs [NLG]	28,630,000
Input to combined cycle - peak hours [MWth]	441.8
Input to combined cycle - off-peak hours [MWth]	278.4
Secondary product - peak hours [tonne/year]	4559
Secondary product - off-peak hours [tonne/year]	49673
Lower Heating Value of secondary product [GJ/tonne]	36.63
Net electricity consumption by co-production section - peak hours [MWh]	2169
Net electricity consumption by co-production section - off-peak hours [MWh]	23634
CO_2 production by co-production section - peak hours [tonne/year]	-*
CO_2 production by co-production section - off-peak hours [tonne/year]	-*
O&M costs [% of total investment costs/year]	7.0

* In this option, there is no CO_2-removal; the purge gas from the co-production section is sent to the combined cycle.

4.3. Results

4.3.1. Hydrogen

In Figure 4, the net yearly income that can be generated with the co-production of hydrogen is presented. The evaluation of the hydrogen option shows that hydrogen is cost-effective at a price of approx. 6.8 NLG/GJ (816 NLG/tonne). At the assumed market price for hydrogen of 10 NLG/GJ, the net yearly income would be 8.8 MNLG/year.

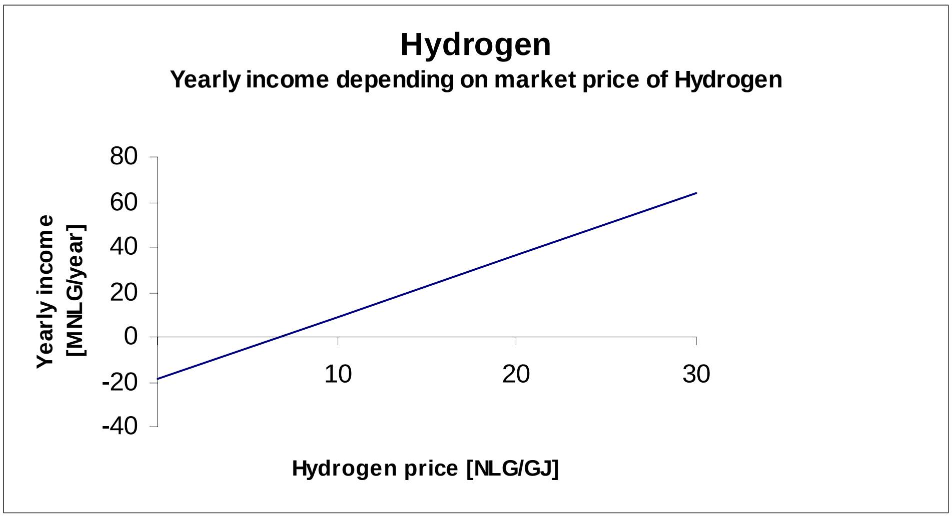

Figure 4. *Net yearly income versus hydrogen market price for the hydrogen co-production option*

Figure 4 shows that the net yearly income increases with approximately 2.8 MNLG/year per NLG/GJ increase in product price. At the product price of 10 NLG/GJ, the built-up of the 8.8 MNLG/year result is based on a gross income of 27.7 MNLG/year and total costs of 18.9 MNLG/year. The costs comprise capital cost (38.1%), fuel cost (46.0%) and O&M cost (15.9%).

It is likely that the cost-effectiveness of the hydrogen co-production increases, if hydrogen with a lower purity is produced. Since it is foreseen to supply the hydrogen into the existing natural gas network, a purity of 99.9% is not required. In particular, this would lead to lower investment costs for the PSA unit, which amount to approx. 50% of the total investment costs in the 99.9% purity case. As an extreme example, leaving out the PSA unit completely would lead to the production of 54% pure hydrogen. If this product, with a Lower Heating Value of 6.9 GJ/tonne, also can be sold at 10 NLG/GJ, the net yearly income would increase from 8.8 to 17.1 MNLG/year.

4.3.2. Methanol - Lurgi process

In Figure 5, the net yearly income that can be generated with the co-production of methanol using the Lurgi process is presented for the case of Fuel Grade Methanol production and AA Grade Methanol production. The evaluation of the Lurgi process shows that methanol co-production is cost-effective at a price of approx. 10.2 NLG/GJ (202 NLG/tonne) for Fuel Grade Methanol and approx. 12.8 NLG/GJ (253 NLG/tonne) for AA Grade Methanol. This latter value is at the lower end of the market price range of 12.6-25.2 NLG/GJ. At an assumed average market price for AA Grade Methanol of 18.9 NLG/GJ, the net yearly income would be 15.4

MNLG/year. The yearly income increases with approx. 2.5 MNLG/year per NLG/GJ increase in product price. The built-up of the 15.4 MNLG/year result is based on a gross income of 42.9 MNLG/year and total costs of 27.5 NLG/year. The costs comprise capital cost (44.2%), fuel cost (31.5%) and O&M cost (24.3%).

Due to lower investments, lower O&M costs and lower electricity consumption, the Lurgi process for Fuel Grade Methanol production results in a higher net yearly income assuming the same methanol market price. The market price for Fuel Grade Methanol may be approx. 2.6 NLG/GJ lower than the price for AA Grade Methanol in order to obtain the same net yearly income.

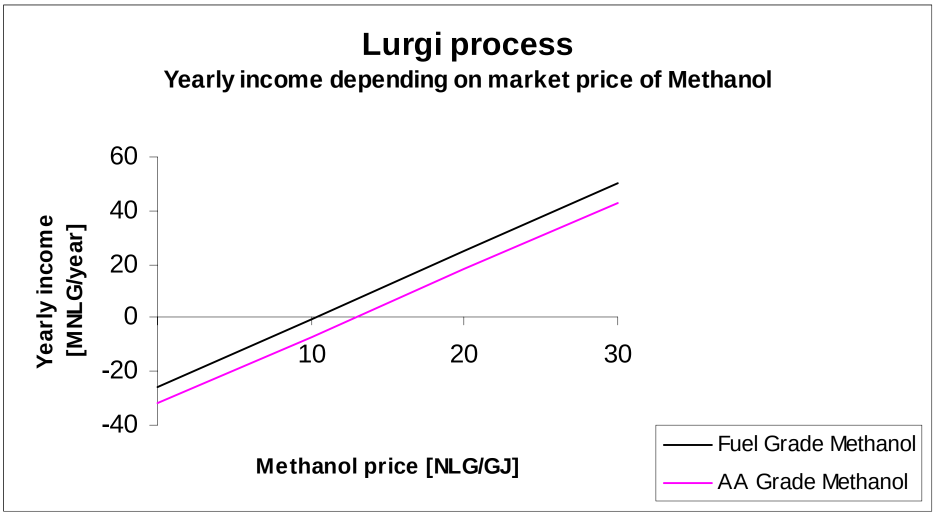

Figure 5. *Net yearly income versus methanol market price for the Lurgi methanol co-production option; note that the market price for Fuel Grade Methanol is expected to be lower than for AA Grade Methanol*

4.3.3. Methanol - LPMEOH process

The LPMEOH methanol co-production option can be described in the same way as the Lurgi methanol co-production option.

LPMEOH methanol co-production is cost-effective at a price of approx. 6.4 NLG/GJ (127 NLG/tonne) for Fuel Grade Methanol and approx. 10.1 NLG/GJ (200 NLG/tonne) for AA Grade Methanol. Therefore, in the case the market price for Fuel Grade Methanol may be approx. 3.7 NLG/GJ lower than the price for AA Grade Methanol in order to obtain the same net yearly income. At an assumed average market price for AA Grade Methanol of 18.9 NLG/GJ, the net yearly income would be 12.8 MNLG/year. The yearly income increases with approx. 1.5 MNLG/year per NLG/GJ increase in product price. The built-up of the 12.8 MNLG/year result is based on a gross income of 34.9 MNLG/year and total costs of 22.1 NLG/year. The costs comprise capital cost (41.6%), fuel cost (39.4%) and O&M cost (19.0%).

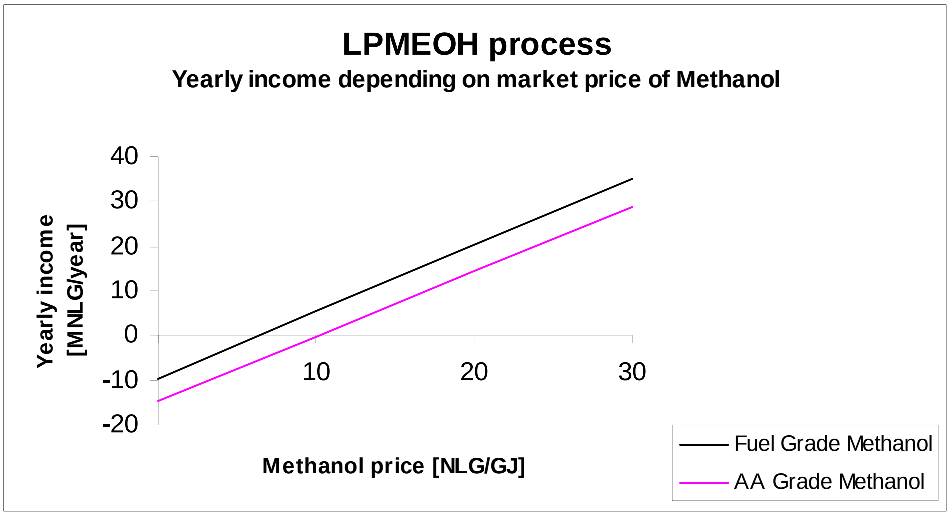

Figure 6. *Net yearly income versus methanol market price for the LPMEOH methanol co-production option; note that the market price for Fuel Grade Methanol is expected to be lower than for AA Grade Methanol*

Great care should be taken when comparing the financial-economic results for the LPMEOH process with those for the Lurgi process. Not only the reactors for the methanol synthesis, but also the selected plant designs are different. The selected Lurgi process configuration includes a CO-shift section for prior adjustment of the H_2/CO ratio in the syngas, which requires maintaining the co-production at minimal 20% capacity during peak hours. The selected LPMEOH process configuration, on the contrary, has been designed to process the syngas rich in carbon oxides directly. This, together with the robustness of the slurry bubble column reactor, makes the process very flexible and enables a decrease to 0% capacity during peak hours. However, the size of the LPMEOH process units is rather large since the total syngas stream of 460 MWth is processed.

It appears that the Lurgi process can also be operated without the prior CO-shift section (Haldor Topsoe, 2000), and that the LPMEOH process can be operated with such a section. However, it was beyond the scope of this study to derive detailed process designs including a specification of the process streams and costs for these alternative configurations.

4.3.4. Fischer-Tropsch transportation fuels

A proper reference to analyse the Fischer-Tropsch transportation fuels price is the fossil crude oil price. This fossil crude oil price is assumed to be 15 $/barrel. Estimates of the higher value of Fischer-Tropsch transportation fuels compared to fossil crude oil range from 135% to 170%. Under these assumptions the Fischer-Tropsch transportation fuels co-production option cannot be regarded as financially viable. In Figure 7, the financial viability is shown as a function of the fossil crude oil price.

Assuming the price of the Fischer-Tropsch transportation fuels to be 135% of the fossil crude oil price, the net yearly income becomes positive at a fossil crude oil price of approx. 29 $/barrel. When the price of the Fischer-Tropsch transportation fuels is assumed to be 170% of the fossil crude oil price, this value is approx. 23 $/barrel. In both cases, this corresponds with an average Fischer-Tropsch transportation fuels price of 16.8 NLG/GJ.

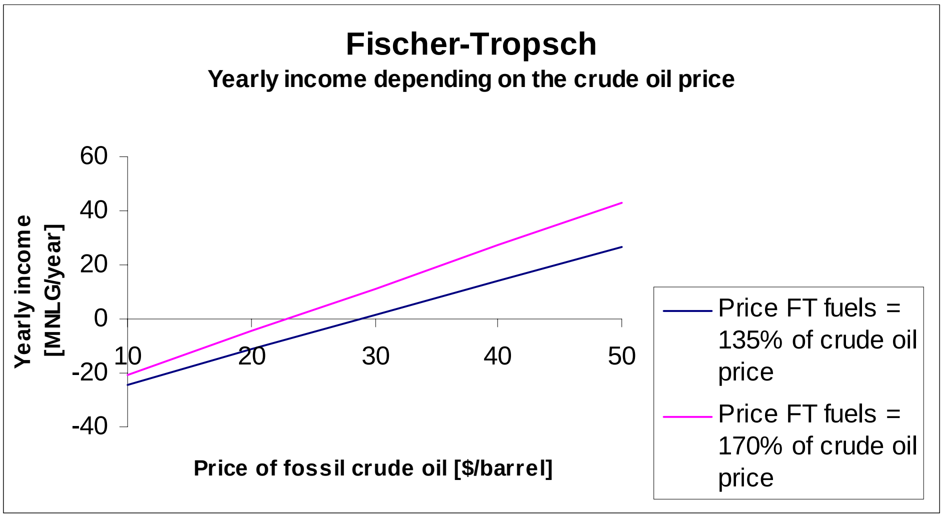

Figure 7. *Net yearly income versus fossil crude oil market price for the Fischer-Tropsch transportation fuels co-production option*

4.3.5. SNG

The net yearly income that can be generated with the co-production of SNG, as calculated using suitable input data obtained by interpreting and adapting the data from the GASTEC report (Van Rens, 2000), is presented in Figure 8. This evaluation of the SNG option shows that it is cost-effective at a SNG price of approximately 5.6 NLG/GJ. At the assumed market price for SNG of 10 NLG/GJ, the net yearly income would be 8.8 MNLG/year. The yearly income increases with 2 MNLG/year per NLG/GJ increase in product price. The built-up of the 8.8 MNLG/year net result consists of a gross income of 22.5 MNLG/year and total costs of 13.7 MNLG/year, including capital cost (21.9%), fuel cost (63.5%) and O&M cost (14.6%).

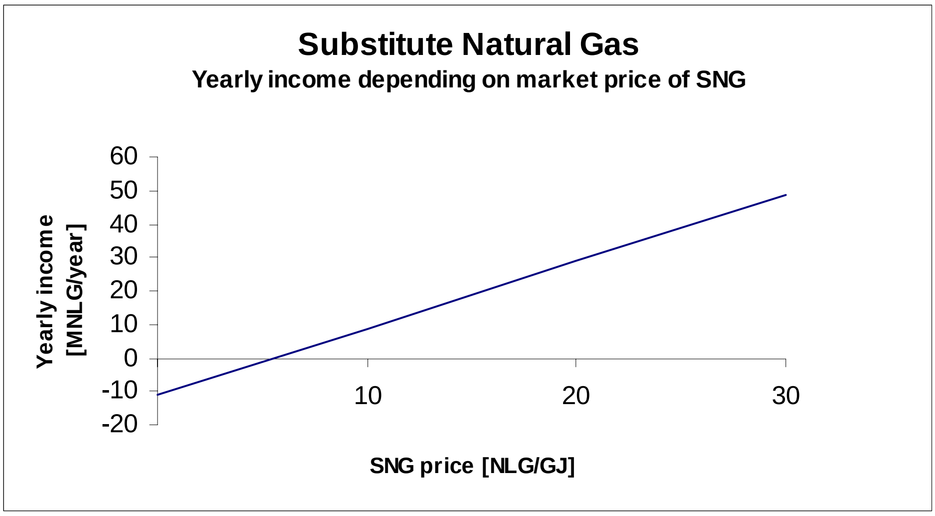

Figure 8. *Net yearly income versus SNG market price for the SNG co-production option*

4.3.6. CO_2 removal and storage

In all co-production options, except for the LPMEOH methanol option, a concentrated CO_2 process stream becomes readily available. For the Lurgi methanol and Fischer-Tropsch transportation fuels options, this is a concentrated CO_2 purge stream from the Selexol CO_2-

removal unit, whilst for the hydrogen and SNG options, this is the purge gas stream from the PSA unit. In case of CO_2 removal and storage, the thermal energy in this PSA purge gas stream could be utilised to produce steam in a separate boiler instead of feeding the purge gas into the combined cycle.

The costs for conditioning and pressurising these concentrated CO_2 streams are estimated to be 11 NLG/tonne. The costs for transportation and injection of the CO_2 in coal beds are estimated to be 44-88 NLG/tonne (Hamelink et al., 2000). Prices for transportation and injection of the CO_2 in aquifers in the province of Brabant or in empty gas fields in the province of Friesland or the North Sea are estimated to be roughly 90 NLG/tonne and >100 NLG/tonne respectively (Wildenborg, 1998).

Not only the costs show large uncertainties, also the height of the extra income that could be generated by producing CO_2-neutral products is very uncertain. Therefore, the costs and extra income for CO_2 removal and storage have not been included in the financial evaluation.

To give a rough indication, the yearly CO_2-reduction in case of CO_2-removal and injection amounts to 0.3 Mtonne/year and 0.2 Mtonne/year for the hydrogen and Lurgi methanol options respectively. Based on this amount it becomes clear that for each Dutch guilder that the income per tonne CO_2 is higher than cost per tonne CO_2, the yearly revenues increase by 0.2-0.3 MNLG. Apart from potential reductions in cost for CO_2 storage, potential subsidies might increase the financial viability.

Finally, the availability of concentrated CO_2 at the co-production plant could also offer other opportunities by looking into possibilities to sell it to industries that need it as a feedstock.

4.3.7. Sensitivity on investment costs and fuel input price

In order to get some information on the impact of higher or lower investment costs or fuel costs, a sensitivity analysis has been conducted assuming the following product prices for the secondary products (see also Paragraph 4.1.2):

• Hydrogen	10 NLG/GJ
• Fuel Grade Methanol	12.6 NLG/GJ
• AA Grade Methanol	18.9 NLG/GJ
• Fischer-Tropsch transportation fuels (at 153% of fossil crude price)	9.8 NLG/GJ
• SNG	10 NLG/GJ

Given these product prices, the net yearly results of the different options are as follows:

• Hydrogen	8.8 MNLG/year
• Lurgi Fuel Grade Methanol	6.0 MNLG/year
• LPMEOH Fuel Grade Methanol	9.3 MNLG/year
• Lurgi AA Grade Methanol	15.4 MNLG/year
• LPMEOH AA Grade Methanol	12.8 MNLG/year
• Fischer-Tropsch transportation fuels	-15.4 MNLG/year
• SNG	8.8 MNLG/year

The differences in net yearly income assuming a change in investment costs of + or - 25% are shown in Figure 9.

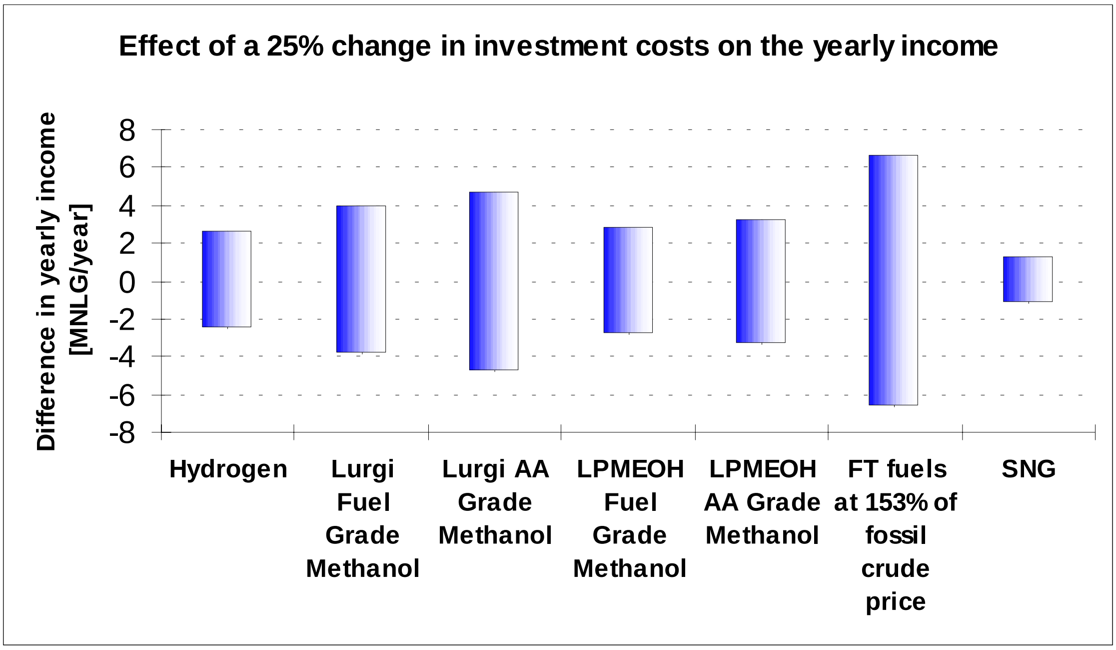

Figure 9. *Effect on the net yearly income of a 25% change in investment costs*

From Figure 9 and the yearly results given above, it appears that the relative effect of a 25% change in investment costs is the highest in case of Lurgi Fuel Grade Methanol (approx. 65%) and Fischer-Tropsch transportation fuels co-production (approx. 40%). SNG co-production shows the lowest relative sensitivity (approx. 15%) due to the low share of the investment cost in the total costs.

A similar sensitivity analysis has been conducted for a 25% change in fuel costs. The results are shown in Figure 10. In all cases, the sensitivity in terms of a change in net yearly income is the same. This is not surprising as in all cases the plant is operating at full load, thus using the same amount of fuel. Keeping in mind the yearly results as described above, the relative sensitivity ranges from approx. 13% (Fischer-Tropsch transportation fuels) to 35% (Lurgi Fuel Grade Methanol).

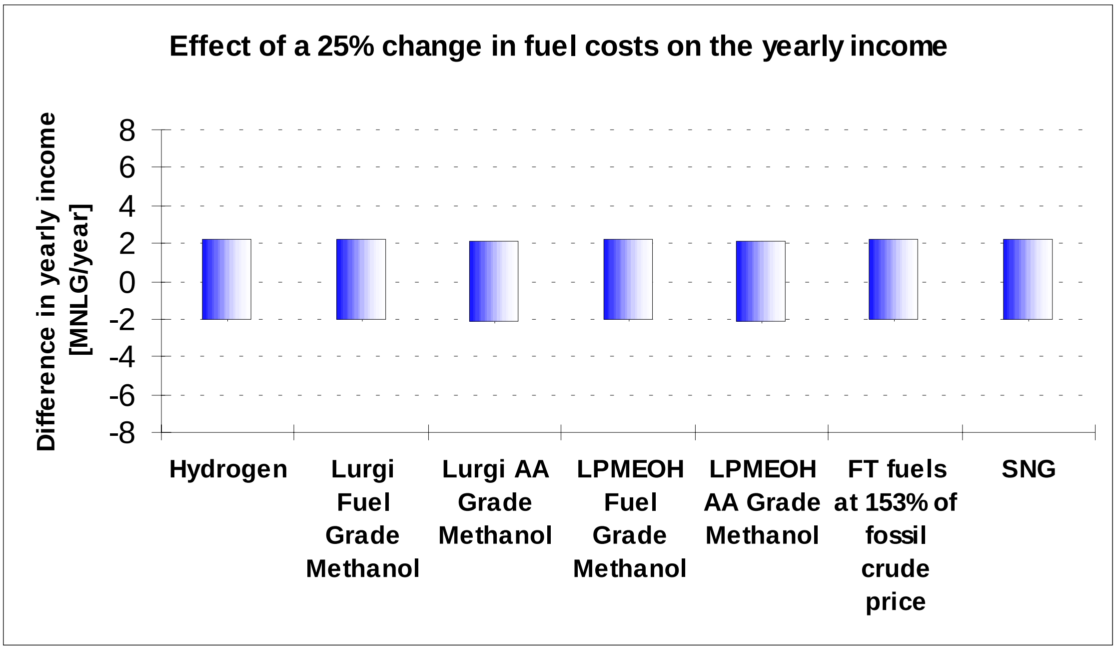

Figure 10. *Effect on the net yearly income of a 25% change in fuel costs*

4.4. Alternative modes of plant operation

In the present situation, the Demkolec power plant is running mainly on coal. Because the price for the produced electricity during off-peak hours is lower than the production costs, the output of the power plant is being decreased during off-peak hours as far as technically possible (from 460 MWth at full load down to 260 MWth). This situation has been chosen as the base case for the financial evaluation: 200 MWth of the syngas is available for the co-production process during off-peak hours.

However, at the same time it has been assumed for the base case that the fuel input changes drastically. Instead of running on coal for almost 100%, the fuel mixture is envisaged to include a substantial fraction of relatively cheap alternative fuels, of which more than 50% is green. Consequently, electricity production during off-peak hours may become cost-effective. If that is the case, it could be an attractive option to continue full-load power generation during off-peak hours. Then, the viability of the different co-production options should be related to this alternative.

Apart from this continuous full-load power generation alternative, also another alternative has been considered briefly. In the course of the study, it became clear that operating in a regulating power mode probably is the most likely operating mode for Demkolec in the liberalised future power market. In fact, the Demkolec plant is operating in this way already.

4.4.1. Off-peak fuels co-production versus continuous full-load power generation

To enable a financial-economic comparison between the different off-peak co-production options and continuous full-load power generation, the additional net yearly income has been determined as a function of the off-peak electricity price for the latter as presented in Figure 11. At the current off-peak electricity price of 0.03 NLG/kWh, the income amounts to 15 MNLG/year. This clearly indicates that electricity production during off-peak hours has become cost-effective indeed. It is interesting to notice that even when the off-peak electricity price would be 0 NLG/kWh, the off-peak electricity production would still be cost-effective due to the income from the partial "greenness" of the fuel mixture. The additional income from the "greenness" is apparently covering already the costs for fuel and other operational expenses.

Furthermore, it must be realised that Figure 11 presents the extra income by producing electricity at full load instead of the regular 57% load during off-peak hours. The actual income is even higher taking into account the benefits of the 57% that is already being produced during off-peak hours.

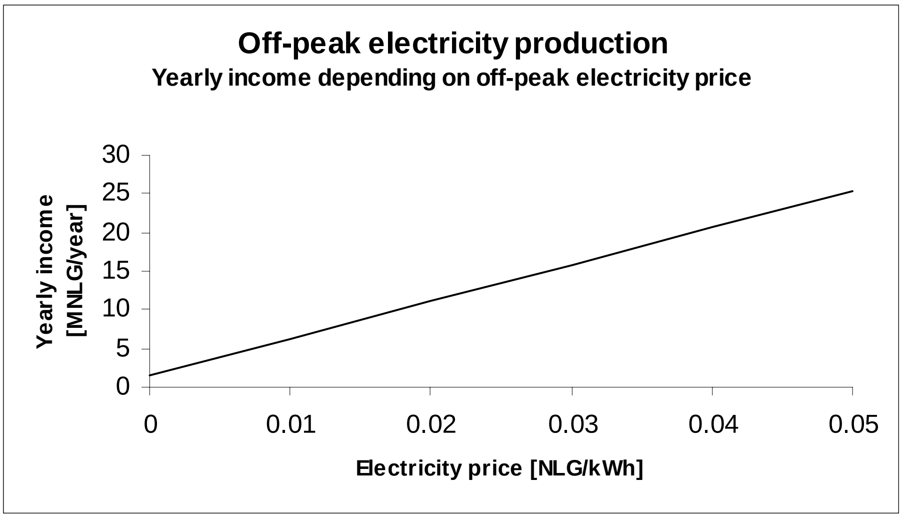

Figure 11. *Yearly additional income from continuous full-load power generation at different electricity prices*

If compared based on net yearly income at a best estimate product prices (see also Paragraph 4.3.7), it becomes clear that it is worthwhile to consider the alternative of continuous full-load power generation seriously. Apart from Lurgi AA Grade Methanol, all off-peak co-production options have a lower net yearly income. In addition, continuous full-load power generation does not involve any additional investments or risks.

However, depending on the situation on the liberalised electricity market, it will probably be more attractive to operate the Demkolec plant in a regulation power mode. In that case, which will be discussed in the next paragraph, the plant will be operated on partial-load electricity production most of the time, and the option of continuous full-load power generation will not be relevant. Instead, the partial-load electricity production during most of the time implies that gasifier capacity is available almost continuously for a co-production option.

4.4.2. Operation in a regulating power mode

Due to the flexibility of operation of the Demkolec plant, the difference between full load (100%) and minimum off-peak load (57%) can be sold as regulating power capacity. This can be done during off-peak hours as well as during peak hours. One of the characteristics of this mode of operation is a high power demand during relatively short periods. The total amount of electricity being produced is limited, but the price for it is high. The Demkolec plant extended with a co-production section could offer interesting advantages over other power plants for regulating power. A large part of the production capacity of these other, mostly gas-fired, power plants will not be used. In case of the Demkolec plant, this production capacity of the gasifier can be used for the co-production process. With a sufficient flexibility of the co-production process, load changes in the electricity production can be accommodated by the co-production section. The gasifier would be operated continuously at full load. In this way, it would be possible to get the high prices for the electricity being produced in a regulating power mode and, at the same time, it could lead to more operating hours of the co-production section, thus increasing its financial viability. Furthermore, accounting for lost income if operation of the co-production section has to be maintained at a certain minimum load during peak hours, as applied in the financial-economic evaluations, is not necessary any more.

In the figures below, the different modes of operation are visualised. Figure 12 shows the base case as used for this study. During peak hours, the gasifier as well as the combined cycle runs at full load. During off-peak hours, when electricity prices are low, the gasifier output is being reduced from 460 MWth to 260 MWth, being the minimum load for the combined cycle.

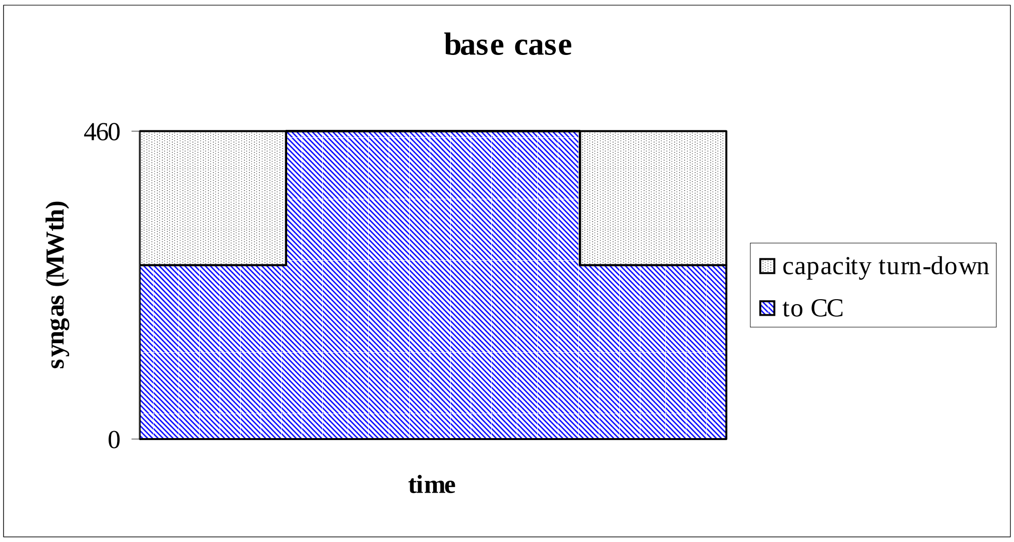

Figure 12. *Visualisation of base-load electricity production (base case)*

Figure 13 shows the case in which the gasifier is being operated continuously at full load and the surplus of syngas during off-peak hours is being used for co-production. Except for the LPMEOH process, it is necessary to maintain the co-production process also during peak hours at a certain minimum load. For SNG the minimum load requirement is 10%, for the other options (hydrogen, Lurgi methanol and Fischer-Tropsch transportation fuels), this is 20%.

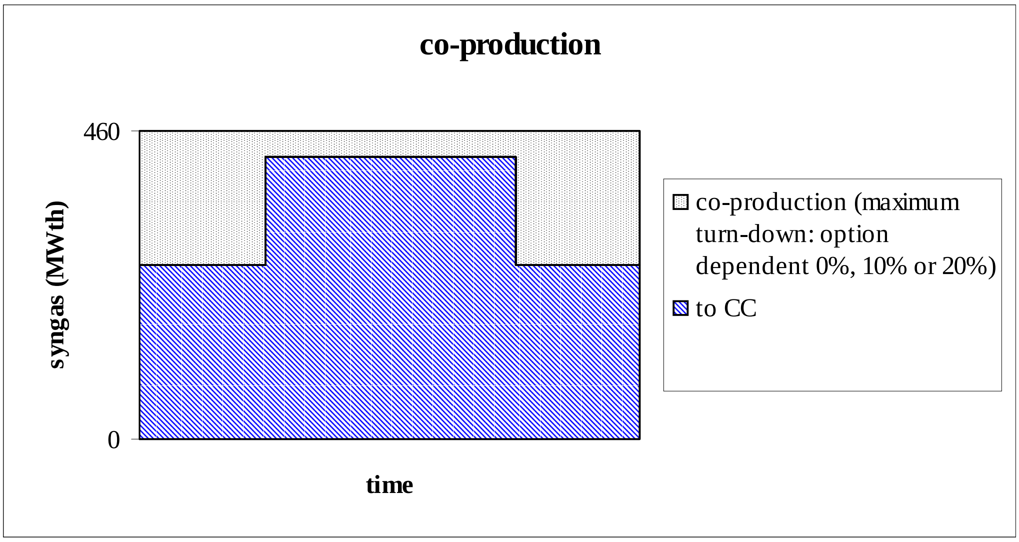

Figure 13. *Visualisation of base-load electricity production in combination with co-production during off-peak hours*

Figure 14 shows the alternative as described earlier in this paragraph. Apart from the minimum load requirement for the combined cycle (260 MWth) and the minimum load requirement for the co-production process (0 to 40 MWth), the syngas is used for either (regulating) power generation or co-production depending on the electricity demand.

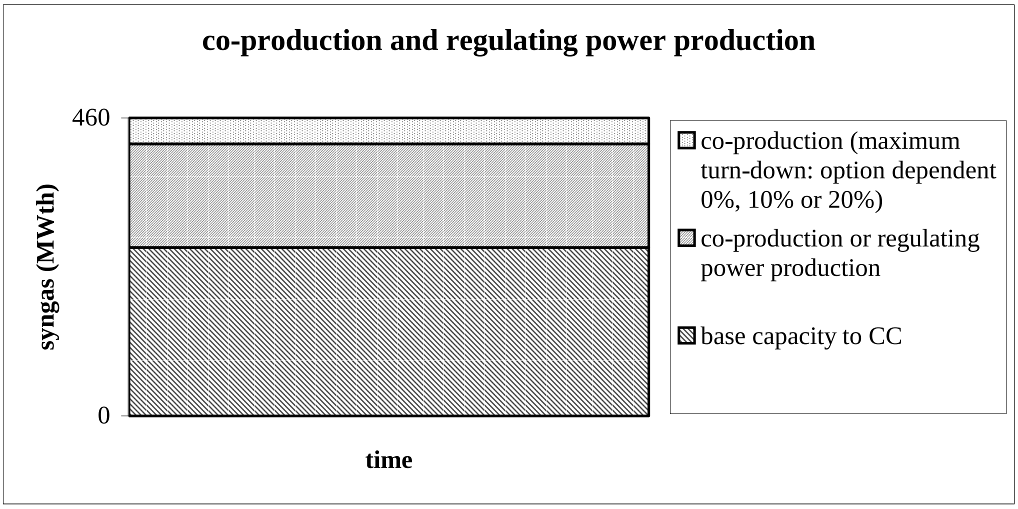

Figure 14. *Visualisation of base-load electricity production in combination with co-production and regulating power production*

Unfortunately, no further analysis could be done on the prospects of the regulating power mode operation, since it emerged in a rather late stage of the study. It is recommended, however, to evaluate this alternative more thoroughly as it represents the most likely future for the Demkolec plant and it offers important opportunities to further increase its overall financial-economic performance.

5. MAIN RESULTS AND CONCLUSIONS

In this study, four different co-production options for the Demkolec plant have been evaluated both technically and financially. These options are: hydrogen, methanol, Fischer-Tropsch transportation fuels and SNG. For the methanol option, two processes have been considered, viz. the Lurgi process including prior adjustment of the H_2/CO ratio (CO-shift) and Air Products' once-through Liquid Phase Methanol (LPMEOH^{TM}) process without this prior adjustment, and two methanol product qualities (Fuel Grade and AA Grade). The options have been compared with a base case, comprising full-load (460 MWth) power generation during peak hours and power generation at minimum load (260 MWth), as determined by the gas turbine, during off-peak hours. Therefore, it has been assumed that in principle the co-production section will be in operation during off-peak hours only and involves 200 of the 460 MW total thermal plant capacity. In addition, it has been assumed that the gasifier is operated on a mixture of 24% "green fuels" (biomass), 21% opportunity fuels and 55% coal (energy basis).

With respect to the main results and conclusions of this evaluation, the limitations of the applied evaluation approach (a preliminary review as stated in the subtitle of the report) should be noticed clearly. For each co-production option either a specific process configuration has been considered for which specifications of the process streams and costs were available from other studies (hydrogen, Lurgi and LPMEOH methanol and SNG to some extent) or only a general configuration with rough estimates for the process stream specifications and costs (Fischer-Tropsch transportation fuels). This means that for all options other process configurations are possible and that the selected ones have not been optimised for the Demkolec plant.

With this in mind, the following main results and conclusions can be drawn from the technical and financial evaluation of the different co-production options:

1. From a viewpoint of technical maturity, hydrogen and Lurgi methanol are the preferred co-production options. Both are mature technologies that are demonstrated on a wide scale. Air Products' LPMEOH process is an innovative process of which a demonstration plant is running since 1997. Also, the Fischer-Tropsch transportation fuels and SNG processes are regarded to be innovative processes, although all individual process steps are mature technology.

2. In terms of operating flexibility, the (once-through) LPMEOH process design without prior adjustment of the H_2/CO ratio in a CO-shift section is the most attractive option. The absence of a CO-shift section in combination with the robustness of the slurry bubble column reactor for methanol synthesis enable rapid ramping and extreme stop/start actions; the process can be decreased to 0% capacity during peak hours. However, the size of the once-through LPMEOH process units is rather large since the total syngas stream of 460 MWth has to be processed. The flexibility of the hydrogen, Lurgi methanol, Fischer-Tropsch transportation fuels and SNG co-production options is limited mainly by the CO-shift section, the CO-shift + methanol synthesis sections, the Fischer-Tropsch synthesis section and the methanation section respectively. These processes cannot be hold on a hot stand-by; their minimum capacity amounts to 10% (SNG) or 20% (hydrogen, Lurgi methanol and Fischer-Tropsch Transportation fuels).

3. A comparison of the financial viability of the different co-production options is given in the table below. From this table it can be seen that all co-production options, except for Fischer-Tropsch transporations fuels, are found to be financially viable, but the viability appears to be strongly dependent on the product market price. The co-production of Lurgi (AA Grade)

Methanol proves to yield the highest net yearly income for the given best estimates of the product market price. The lower financial viability of the Fuel Grade Methanol options compared to the AA Grade Methanol options is because the assumed difference in product market price is larger than the difference in production cost. The economics of hydrogen and SNG co-production are quite comparable. The 99.9% purity of the hydrogen is probably higher than required for supply into the natural gas network. The production of hydrogen with a lower purity is expected to yield a higher net yearly income. The co-production of Fischer-Tropsch transportation fuels becomes financially viable at an average product price of 16.8 NLG/GJ, which corresponds to a fossil crude oil price of 23-29 US$/barrel.

Co-production option	Product market price (best estimate, NLG/GJ)	Product price for which zero yearly income (NLG/GJ)	Net yearly income at product market price (MNLG/year)	Δ yearly income per Δ NLG/GJ product price (MNLG/year)
Hydrogen	10.0	6.8	8.8	2.8
Lurgi Fuel Grade methanol	12.6	10.2	6.0	2.5
LPMEOH Fuel Grade Methanol	12.6	6.4	9.3	1.5
Lurgi AA Grade methanol	18.9	12.8	15.4	2.5
LPMEOH AA Grade Methanol	18.9	10.1	12.8	1.5
FT transportation fuels	9.8	16.8	-15.4	2.2
SNG	10.0	5.6	8.8	2.0

4. For the given best estimate product market prices, a 25% change in investment costs leads a change in net yearly income ranging from 15% for the SNG option to 65% for the Lurgi Fuel Grade Methanol option. A 25% change in fuel costs leads to a change in net yearly income ranging from 13% for the Fischer-Tropsch transportation fuels option to 35% for the Lurgi Fuel Grade Methanol option.

In addition, the following more general conclusions can be drawn:

5. In the different cases no investment subsidies like IEA and VAMIL have been taken into account. These subsidies by the Dutch government could (with certain limitations) amount up to approximately 20% of the investment costs. They are especially applicable for investments in energy saving measures, renewable energy or environmental measures in the energy sector. Most options for co-production, but especially investments to enable the replacement of coal by biomass as a fuel for the Demkolec plant, could benefit from these subsidies.

6. Because of present energy and environmental policy, the ‘greenness’ of electricity and fuels represents an extra value for the product. For electricity, this value is rather clear and is strongly related to the regulating energy tax. For the other products, however, the value of this greenness is still less clear. To stimulate the implementation of fuels (co-)production options, it is important that this issue will be clarified. In this study, it has been assumed that the extra value for greenness for the produced fuels can be equalised to that for electricity on a primary input basis.

7. In all co-production options, except for the LPMEOH methanol option, a concentrated CO_2 process stream becomes readily available. For the Lurgi methanol and Fischer-Tropsch transportation fuels options, this is a concentrated CO_2 purge stream from the Selexol CO_2-removal unit, whilst for the hydrogen and SNG options, this is the purge gas stream from the PSA unit. The costs for conditioning, pressurising, transportation and storage in aquifers, empty gas fields or coal beds show large uncertainties, just as the height of the extra income that could be generated by producing CO_2-neutral products. Therefore, the costs and extra income for CO_2 removal and storage have not been included in the financial

evaluation. To give a rough indication, the yearly CO_2-reduction in case of CO_2-removal and injection amounts to 0.3 Mtonne/year and 0.2 Mtonne/year for the hydrogen and Lurgi methanol options respectively. The availability of concentrated CO_2 at the co-production plant could also offer other opportunities by looking into possibilities to sell it to industries that need it as a feedstock.

8. In the present situation, with the Demkolec power plant being operated mainly on coal, the revenues for the produced electricity during off-peak hours are lower than the production costs. However, if the plant switches to the new fuel mixture, then electricity production during off-peak hours becomes cost-effective. Consequently, continuous full-load power generation may become an attractive option from a financial point of view. In fact, at the current off-peak electricity price of 0.03 NLG/kWh the calculated net yearly income of 15 MNLG/year is fairly large compared to the values for the different co-production options. Even when the off-peak electricity price would be 0 NLG/kWh, the off-peak electricity production would still be cost-effective due to the income from the partial "greenness" of the fuel mixture. In addition, continuous full-load power generation does not involve any additional investments or risks. However, given the present situation on the liberalised electricity market, continuous full-load power generation is probably not the most profitable alternative.

9. Due to the flexibility of operation of the Demkolec plant, the difference between full load (100%) and minimum off-peak load (57%) can be sold as regulating power capacity (at a higher price). In fact, the Demkolec plant is operating to a considerable extent in such a mode already, both during off-peak hours as well as during peak hours. This mode of operation enforces the opportunities for fuels co-production. With a sufficient flexibility of the co-production process, load changes in the electricity production can be matched by changing the load on the co-production section. The gasifier can be operated continuously at full load. In this way, it will be possible to get the high prices for the electricity being produced in a regulating power mode and, at the same time, it can lead to more operating hours of the co-production section, thus increasing its financial viability. It is recommended to evaluate this alternative more thoroughly as it represents the most likely future for the Demkolec plant and it offers important opportunities to further increase its overall financial-economic performance.

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APPENDIX A: HYDROGEN CO-PRODUCTION

Table A1. *Process stream specifications of hydrogen co-production*

		IN	**IN**	**IN**	**IN**	**IN**	**TOTAL**
	description	syngas	make-up water	shift-steam	cooling water	electricity	
	from	desulfurization	make-up	steam/water system	CW system		
	to	saturator	saturator	shift reactor 1	desaturator		
T	°C	37,5	200,0	275,0	23,0		
p	bar	22,4	22,4	22,0	5,2		
phase	g / v / l	v	l	v	l		
H2	mol%	30,0	0,0	0,0	0,0		
CO	mol%	60,0	0,0	0,0	0,0		
CO2	mol%	2,5	0,0	0,0	0,0		
H2O	mol%	0,0	100,0	100,0	100,0		
N2	mol%	6,8	0,0	0,0	0,0		
Ar	mol%	0,7	0,0	0,0	0,0		
Total	[-]	100,0	100,0	100,0	100,0		
Flow	kmol/s	0,911	0,190	0,396	2,610		
Flow	kg/s	18,85	3,42	7,13	46,98		76,38
density	kg/m3	18,9	802,3	9,3	995,4		
LHV	MJ/kg	11,75	0,0	0,0	0,0		
Enthalpy	MWth	221,5	0,0	3,6	0,0		
Electrical Power	MWe	0,0	0,0	0,0	0,0	3,48	

		OUT	**OUT**	**OUT**			
	description	hydrogen	PSA purge	cooling water			
	from	PSA	PSA	desaturator			
	to	storage / pipeline	Combined Cycle	CW system			
T	°C	41,2	41,2	122,3			
p	bar	18,0	1,3	5,0			
phase	g / v / l	v	v	l			
H2	mol%	99,9	10,8	0,0			
CO	mol%	0,0	0,8	0,0			
CO2	mol%	0,0	73,5	0,0			
H2O	mol%	0,0	6,0	100,0			
N2	mol%	0,1	8,0	0,0			
Ar	mol%	0,0	0,8	0,0			
Total	[-]	100,0	100,0	100,0			
Flow	kmol/s	0,732	0,766	2,610			
Flow	kg/s	1,48	27,92	46,98			76,38
density	kg/m3	1,4	1,94	893,9			
LHV	MJ/kg	118,3	0,77	0,0			
Enthalpy	MWth	176,2	21,5	0,0			
Electrical Power	MWe	0,0	0	0,0			

APPENDIX B: LURGI METHANOL CO-PRODUCTION

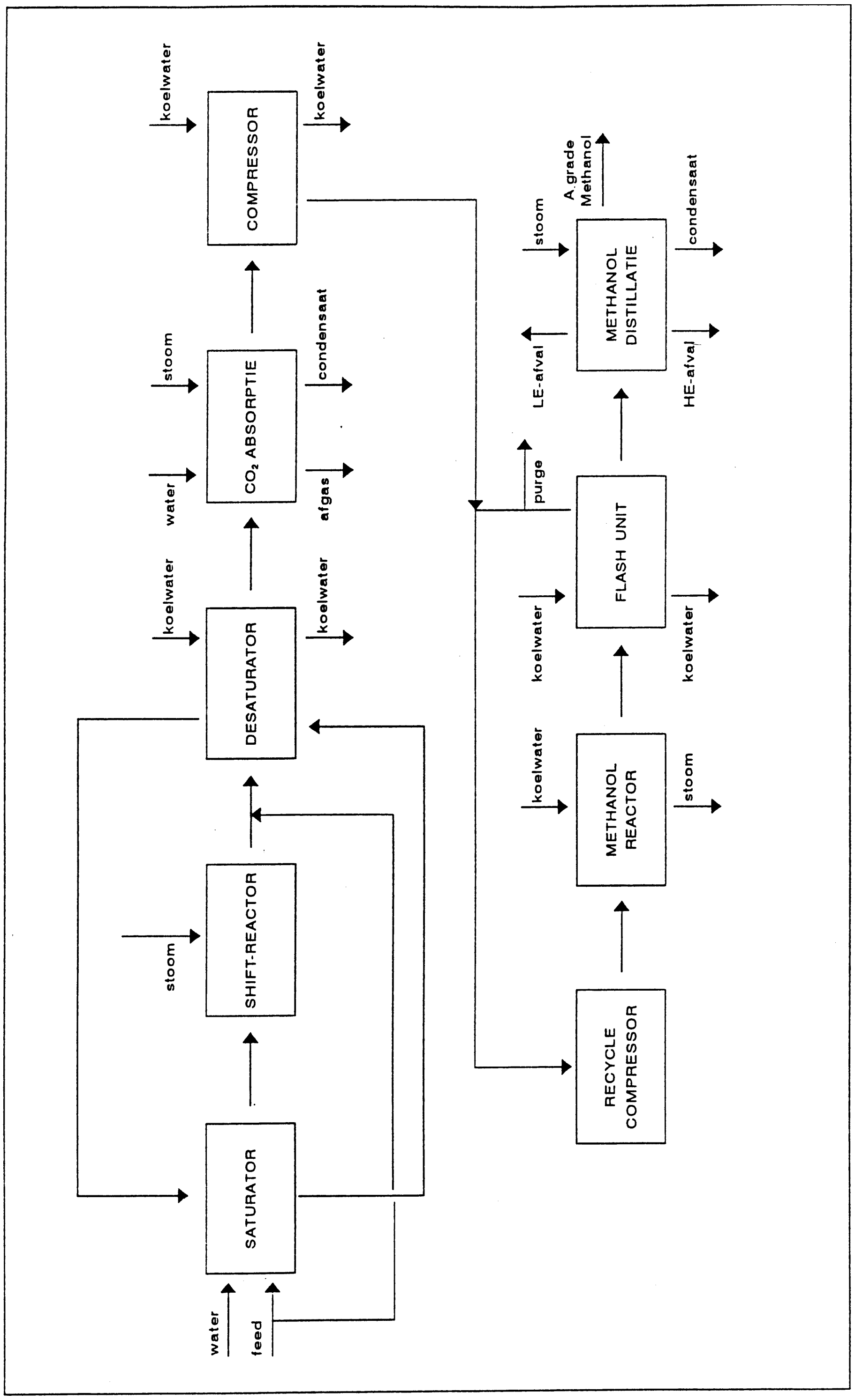


Figure B1. *Block scheme of Lurgi methanol co-production (Lensink - van der Veen, 1996a)*

Table B1. *Process stream specifications of Lurgi Fuel Grade Methanol co-production*

		IN	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**TOTAL**
	description	syngas	make-up water	shift-steam	cooling water	cooling water	cooling water	cooling water	steam	make-up water	electricity	
	from	desulfurization	make-up	steam/water system	CW system	CW system	CW system	CW system	steam/water system	make-up		
	to	saturator	saturator	shift reactor 1	cooler 1	cooler 4	compressor 1	MeOH reactor	CO2-absorption	CO2-absorption		
T	°C	37,5	200,0	350,0	23,0	23,0	23,0	117,8	152,0	128,3		
p	bar	21,4	21,4	21,0	5,2	5,2	5,2	39,9	5,0	5,0		
phase	g / v / l	v	l	v	l	l	l	l	v	l		
H2	mol%	30,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
CO	mol%	60,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
CO2	mol%	2,50	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
H2O	mol%	0,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00		
N2	mol%	6,80	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
Ar	mol%	0,70	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
MeOH	mol%	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
Total	[-]	100,00	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0		
Flow	kmol / s	0,897	0,034	0,264	1,927	1,367	0,373	0,559	0,002	1,390		
Flow	kg / s	18,56	0,62	4,75	34,68	24,60	6,71	10,05	0,03	25,02		125,01
density	kg / m3	17,1	802,3	7,6	995,4	753,8	753,8	696,8				
LHV	MJ / kg	11,75	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0		
Enthalpy	MWth	218,1	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0		
Electrical Power	MWe	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	7,22	

		OUT	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**		
	description	fuel grade MeOH	CO2	purge MeOH cycle	cooling water	cooling water	cooling water	cooling water	condensate	recycle water		
	from		CO2 absorption		cooler 1	cooler 4	compressor 1	MeOH reactor	CO2 absorption	CO2 absorption		
	to	storage	atmosphere	Combined Cycle	CW system	CW system	CW system	CW system	steam/water system	water make-up		
T	°C	40,0	40,0	40,0	117,8	128,3	128,3	250,0	105,0	105,0		
p	bar	98,7	1,4	98,7	5,0	5,0	5,0	39,7	1,2	4,8		
phase	g / v / l	l	v	v	l	l	l	v	l	l		
H2	mol%	1,01	0,54	36,78	0,00	0,00	0,00	0,00	0,00	0,00		
CO	mol%	0,35	0,25	11,04	0,00	0,00	0,00	0,00	0,00	0,00		
CO2	mol%	1,96	92,93	4,92	0,00	0,00	0,00	0,00	0,00	0,00		
H2O	mol%	1,00	6,22	0,00	100,00	100,00	100,00	100,00	100,00	100,00		
N2	mol%	0,99	0,06	42,29	0,00	0,00	0,00	0,00	0,00	0,00		
Ar	mol%	0,10	0,01	4,35	0,00	0,00	0,00	0,00	0,00	0,00		
MeOH	mol%	94,59	0,00	0,62	0,00	0,00	0,00	0,00	0,00	0,00		
Total	[-]	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00		
Flow	kmol / s	0,257	0,324	0,129	1,927	1,367	0,373	0,559	0,002	1,390		
Flow	kg / s	8,20	13,00	2,72	34,68	24,60	6,71	10,05	0,03	25,02		125,01
density	kg / m3	594,2	2,3	64,1	898,8	688,2	688,2	19,1	0,0	0,0		
LHV	MJ / kg	19,2	0,00	7,9	0,0	0,0	0,0	0,0	0,0	0,0		
Enthalpy	MWth	156,8	0,6	18,1	0,0	0,0	0,0	0,0	0,0	0,0		
Electrical Power	MWe	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0		

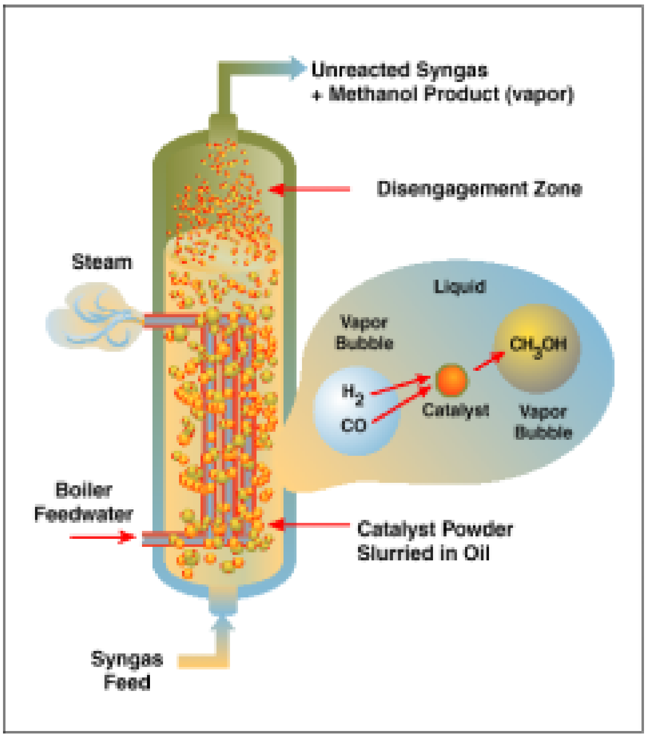
Table B2. *Process stream specifications of Lurgi AA Grade Methanol co-production*

		IN	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**IN**	**TOTAL**
	description	syngas	make-up water	shift-steam	cooling water	cooling water	cooling water	cooling water	steam	make-up water	steam	electricity	
	from	desulfurization	make-up	steam/water system	CW system	CW system	CW system	CW system	steam/water system	make-up	steam/water system		
	to	saturator	saturator	shift reactor 1	cooler 1	cooler 4	compressor 1	MeOH reactor	CO2-absorption	CO2-absorption	distillation		
T	°C	37,5	200,0	350,0	23,0	23,0	23,0	117,8	152,0	128,3	159,0		
p	bar	21,4	21,4	21,0	5,2	5,2	5,2	39,9	6,0	5,0	6,0		
phase	g / v / l	v	l	v	l	l	l	l	v	l	v		
H2	mol%	30,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
CO	mol%	60,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
CO2	mol%	2,50	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
H2O	mol%	0,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00		
N2	mol%	6,80	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
Ar	mol%	0,70	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
MeOH	mol%	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
Total	[-]	100,00	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0	100,0		
Flow	kmol / s	0,917	0,035	0,270	1,969	1,397	0,381	0,571	0,002	1,421	3,923		
Flow	kg / s	18,97	0,63	4,85	35,45	25,14	6,86	10,28	0,03	25,57	70,64		198,42
density	kg / m3	17,1	802,3	7,6	995,4	753,8	753,8	696,8					
LHV	MJ / kg	11,75	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0		
Enthalpy	MWth	222,9	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0		
Electrical Power	MWe	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	7,57	

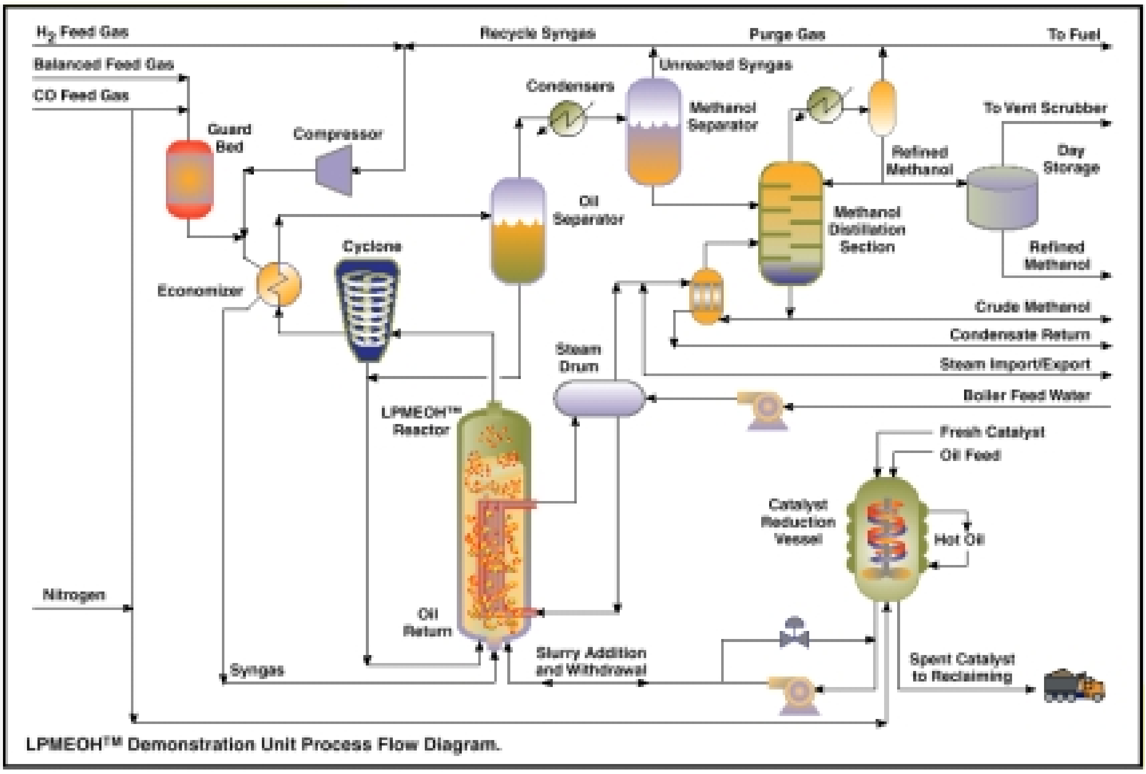
		OUT	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**OUT**	**TOTAL**
	description	AA grade MeOH	impurities	CO2	purge MeOH cycle	cooling water	cooling water	cooling water	cooling water	condensate	recycle water	condensate	
	from		distillation	CO2 absorption		cooler 1	cooler 4	compressor 1	MeOH reactor	CO2 absorption	CO2 absorption	distillation	
	to	storage	Combined Cycle	atmosphere	Combined Cycle	CW system	CW system	CW system	CW system	steam/water system	water make-up	steam/water system	
T	°C	40,0	40,0	40,0	40,0	117,8	128,3	128,3	250,0	105,0	105,0	152,0	
p	bar	98,7	4,0	1,4	98,7	5,0	5,0	5,0	39,7	1,2	4,8	5,0	
phase	g / v / l	l	l	v	v	l	l	l	v	l	l	l	
H2	mol%	0,00	13,34	0,54	36,78	0,00	0,00	0,00	0,00	0,00	0,00	0,00	
CO	mol%	0,00	4,62	0,25	11,04	0,00	0,00	0,00	0,00	0,00	0,00	0,00	
CO2	mol%	0,00	26,03	92,93	4,92	0,00	0,00	0,00	0,00	0,00	0,00	0,00	
H2O	mol%	0,15	11,43	6,22	0,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	
N2	mol%	0,00	13,12	0,06	42,29	0,00	0,00	0,00	0,00	0,00	0,00	0,00	
Ar	mol%	0,00	1,35	0,01	4,35	0,00	0,00	0,00	0,00	0,00	0,00	0,00	
MeOH	mol%	99,85	30,10	0,00	0,62	0,00	0,00	0,00	0,00	0,00	0,00	0,00	
Total	[-]	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	100,00	
Flow	kmol / s	0,263	0,020	0,332	0,132	1,969	1,397	0,381	0,571	0,002	1,421	3,923	
Flow	kg / s	8,15	0,23	13,28	2,78	35,45	25,14	6,86	10,28	0,03	25,57	70,64	198,42
density	kg / m3	594,2		2,3	64,1	898,8	688,2	688,2	19,1	0,0	0,0		
LHV	MJ / kg	19,2		0,00	7,9	0,0	0,0	0,0	0,0	0,0	0,0	0,0	
Enthalpy	MWth	155,8	4,4	0,6	18,5	0,0	0,0	0,0	0,0	0,0	0,0	0,0	
Electrical Power	MWe	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	

note: cooling water for distillation is not included

APPENDIX C: LPMEOH METHANOL CO-PRODUCTION



LPMEOH™ Slurry Bubble Column Reactor Schematic. In a SBCR, reactants from the gas phase dissolve in the liquid and diffuse to the catalyst surface, where they react. Products then diffuse through the liquid back to the gas phase. Heat is removed by generating steam in an internal tubular heat exchanger.



LPMEOH™ Demonstration Unit Process Flow Diagram.

Figure C1. *Reactor schematic and demonstration unit flow diagram (CCT, 1999)*

Table C1. *Process stream specifications of once-through LPMEOH methanol co-production*

Fuel grade Methanol

		IN		**TOTAL**
	description	syngas		
	from	desulfurization		
	to	saturator		
T	°C	37,5		
p	bar	21,4		
phase	g / v / l	v		
H2	mol%	30,00		
CO	mol%	60,00		
CO2	mol%	2,50		
H2O	mol%	0,00		
N2	mol%	6,80		
Ar	mol%	0,70		
MeOH	mol%	0,00		
Total	[-]	100,00		
Flow	kmol / s	1,893		
Flow	kg / s	39,15		39,15
density	kg / m3	17,1		
LHV	MJ / kg	11,75		
Enthalpy	MWth	460,0		460,00
Electrical Power	MWe	0,0		
		OUT	**OUT**	
	description	fuel grade MeOH	Balance	
	from			
	to	storage		
T	°C			
p	bar			
phase	g / v / l	l		
H2O	mol%	1,00		
Other	mol%	1,00		
MeOH	mol%	98,00		
Total	[-]	100,00		
Flow	kmol / s			
Flow	kg / s	5,74	33,41	39,15
density	kg / m3	594,2		
LHV	MJ / kg	19,2		
Enthalpy	MWth	110,2	349,8	460,00
Electrical Power	MWe	0,0		

AA grade Methanol

		IN		**TOTAL**
	description	syngas		
	from	desulfurization		
	to	saturator		
T	°C	37,5		
p	bar	21,4		
phase	g / v / l	v		
H2	mol%	30,00		
CO	mol%	60,00		
CO2	mol%	2,50		
H2O	mol%	0,00		
N2	mol%	6,80		
Ar	mol%	0,70		
MeOH	mol%	0,00		
Total	[-]	100,00		
Flow	kmol / s	1,893		
Flow	kg / s	39,15		39,15
density	kg / m3	17,1		
LHV	MJ / kg	11,75		
Enthalpy	MWth	460,0		460,00
Electrical Power	MWe	0,0		
		OUT	**OUT**	
	description	AA grade MeOH	Balance	
	from			
	to	storage		
T	°C			
p	bar			
phase	g / v / l	l		
H2O	mol%	0,15		
Other	mol%	0,00		
MeOH	mol%	99,85		
Total	[-]	100,00		
Flow	kmol / s			
Flow	kg / s	5,58	33,57	39,15
density	kg / m3	594,2		
LHV	MJ / kg	19,2		
Enthalpy	MWth	107,1	352,9	460,00
Electrical Power	MWe	0,0		

APPENDIX D: SUBSTITUTE NATURAL GAS CO-PRODUCTION AS EVALUATED BY GASTEC

Below, an English translation is presented of the summary of the report, prepared on the SNG co-production option by GASTEC for Demkolec (Van Rens, 2000). In addition, the process flow diagram for SNG co-production, as taken from this report, is presented.

SUMMARY

The technical and economic feasibility of SNG production from syngas has been studied by order of Demkolec. The current mode of operation consists of electricity production using a gas turbine, but the installation is turned down to minimum load during off-peak hours, since operation is not economic then. The design of the SNG installation is based on the amount of syngas, which is available when the gas turbine operates at minimum load. The other starting points for the design are mentioned in Chapter 3.

The syngas is supplied to the SNG installation at the same quality as it is currently offered to the gas turbine. This means that the syngas is cleaned and preheated to 130 ^{o}C. Producing SNG from the syngas requires an installation consisting of the following components:

- Guard bed for the removal of sulphur compounds down to ppb levels.
- Heating of the syngas and steam addition.
- Methanation reactor, in which methane and carbon dioxide are formed.
- Cooling of the gas after the methanation, which involves water condensation.
- Drying.
- PSA-system for CO_2-removal to bring the gas on the right Wobbe-index.
- Compression of the SNG to the pressure level of the gas network, into which the SNG is supplied.

Besides erection of the components mentioned above, it is recommended to modify the existing installations in the following aspects:

1. Apply carbon dioxide instead of nitrogen as transport medium for coal, since this will lead to a larger SNG yield at limited additional cost.
2. Extension of the sulphur removal unit by installing a second scrubber. The advantage compared to just a guard bed for sulphur removal is a reduction in amount of waste.

The SNG cost price consists of:

- Syngas production from coal. The fuel cost of this production amount to 0.244 NLG/m_n^3 SNG.
- Syngas conversion into SNG. The cost price of this part amounts to:
  0.091 NLG/m_n^3 for operating mode 1[1]
  0.058 NLG/m_n^3 for operating mode 2

This leads to a total cost price of 0.34 NLG/m_n^3 for operating mode 1 and 0.30 NLG/m_n^3 for operating mode 2, with an accuracy of 20%.

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[1] For the financial analysis, two modes of operation have been defined:
Operating mode 1: the SNG installation operates at full load during off-peak hours and at minimum load during peak hours
Operating mode 2: the SNG installation operates at full load continuously.

The most important price-determining factors are the supply pressure and the level of the investment costs. Since it has not been determined into which gas network the SNG will be supplied, also the supply pressure is not known yet. Because of this uncertainty, the cost price of the syngas conversion into SNG can vary from 0.035 to 0.078 NLG/m_n^3 (operating mode 2).

The revenues of the SNG are dependent on the sales potential and the added value for the gas because of Demkolec being the entry point. The commodity price for natural gas ranges currently between 0.28 and 0.44 NLG/m_n^3. Co-gasification of biomass will give an extra market value of maximal approx. 0.31 NLG/m_n^3 SNG.

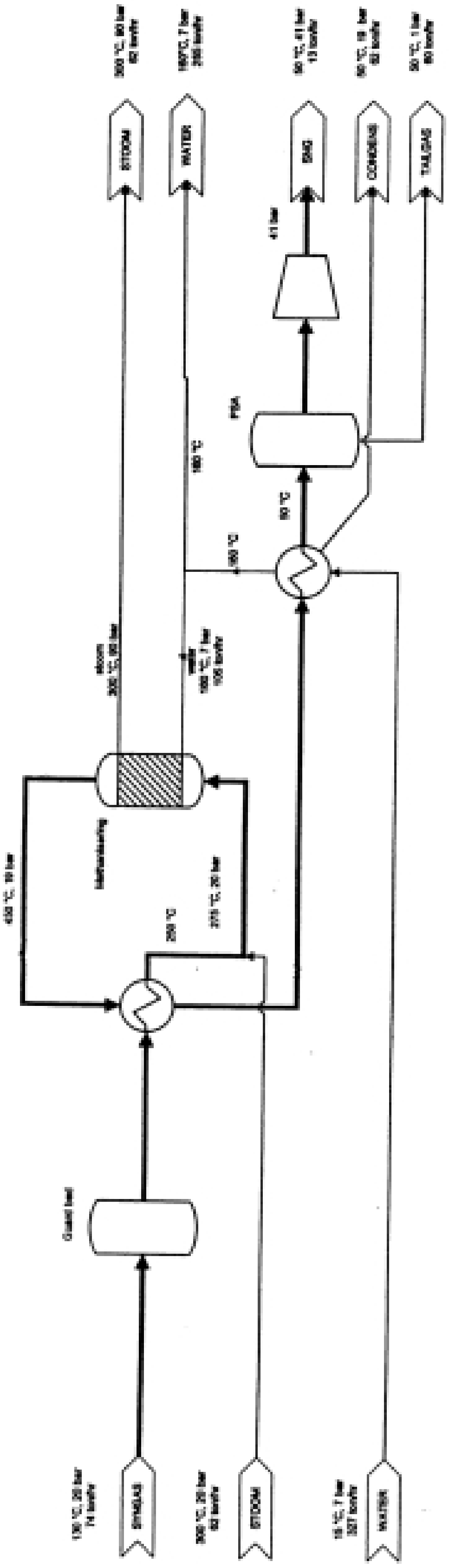

Figure D1. *Process flow diagram of SNG co-production (Van Rens, 2000)*