SKOGSKEMI – OLEFINS VALUE CHAIN

Sub-project report to the Skogskemi project

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This report should be cited as:
Olefins Project

Forest Chemical Project Vinnova

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Stenungsund
9/25/2014
Summary
Ethylene and propylene, also referred to as light olefins, are important building blocks used for producing e.g. polymers. Currently most olefins are produced via thermal cracking of naphtha or other light fractions of petroleum with steam, which is often referred to as steam cracking. During the process the hydrocarbon feedstock, usually of fossil origin, is cracked into smaller molecules. The process is very energy demanding and consumes fossil resources. The Forest Chemical project started with the aim to investigate opportunities to produce chemicals from forest material. Involved in the project is parties working with all parts along the value chain. The aim of the Olefin part of the Forest Chemical project was to investigate opportunities of producing green olefins (ethylene, propylene and butylene’s) origin from 2nd generation ethanol (ethanol produced from cellulosic raw material) and green methanol (methanol e.g. produced via synthesis of syngas derived via gasification of cellulosic material) in Stenungsund. One requirement in this project was to at least produce 200 kton of green olefins origin from forest residues utilizing ethanol and methanol.

The UOP/Hydro Methanol-to-olefins (MTO) process was chosen to be investigated in this project. UOP has provided a Technology Information package for the production of 300 kton C2, C3 and C4 olefins a year from methanol. The Chematur Ethanol to Ethylene (E2E) process was also chosen to be investigated in this project. Chematur Engineering has provided technical information for the production of 60 kton ethylene a year from ethanol. The two process routes were evaluated based on e.g. mass and energy balances, availability, synergies with existing processes, process related aspects, environmental aspects, safety aspect, storage possibilities etc. resulting in an estimated investment cost. The profitability of the processes was calculated based on projected raw material and product prices produced within the Forest Chemical project.

The project as such is a sustainability driven project, an investigation of possible transitions from non-renewable resources to renewable resources in the form of forest material. However, to be able to implement any of the more sustainable alternatives (at least seen in the resource perspective), such as E2E or MTO, they have to be economically attractive. In that perspective, the project also has to be economically justified. Today, green raw materials for the production of Olefins are less attractive then fossil. For the project to be economically attractive to implement, either the green raw materials have to be more cost competitive with fossil ones, or it has to be possible to charge a higher price for the green olefins, a so called green premium price.

The project resulted in a ±40% investment cost for the MTO process of MSEK 2 864 (MEUR 315) and ±40% investment cost for the E2E process of MSEK 636 (MEUR 70), with equal probability to overrun as underrun. Using the projected raw material and product prices it is possible to conclude that the MTO process only is economically attractive when assuming a green premium price of at least 30% and that the E2E process only is economically attractive assuming a green premium price of about 20 to 30%. The resulting IRR when assuming a green premium of 30% for the MTO process is 24% respectively 21% depending on price scenario. The resulting IRR when assuming a green premium of 30% for the E2E process is 29 respectively 28% depending on price scenario.
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1 Background

Ethylene and propylene, also referred to as light olefins, are important building blocks used for producing e.g. polymers (Zimmermann & Walzl, 2009). Ethylene is one of the largest consumed base chemicals by volume, and is mostly used as a feedstock in the manufacturing of plastics, fibres, and other organic chemicals. Ethylene is a fundamental building unit in the global petrochemical industry. Products produced from ethylene include polyethylene (PE), polyvinylchloride (PVC) and polyethylene terephthalate (PET) (Global Industry Analysts, Inc, 2008). Propylene is also an important feedstock for industrial derivatives such as polypropylene, acrylonitrile, propylene-oxide and phenol. Propylene usage spans over various industries, from automotive and construction to packaging, medical and electronics (CHEMSYSTEMS, 2012). Currently most olefins are produced via thermal cracking of naphtha or other light fractions of petroleum with steam, which is often referred to as steam cracking. During the process the hydrocarbon feedstock, usually of fossil origin, is cracked into smaller molecules. The process is very energy demanding. The resulting product mix from the cracking process, which varies depending on the process conditions, must then be separated into the desired products by using a sequence of separation and other chemical treatment steps (Zimmermann & Walzl, 2009). Olefins are currently produced via steam cracking at Borealis located in Stenungsund. Borealis is part of a chemical cluster in Stenungsund, and their steam cracker plant is the heart of this cluster (Hackl, et al., 2011), as can be seen in Figure 1.

![Diagram](image_url)

Figure 1: Illustration of the existing infrastructure of the chemical cluster in Stenungsund.

Within this chemical cluster they have a common vision that “in 2030 Stenungsund industry park will be the hub for the manufacturing of sustainable products within the Swedish chemical industry” (Business Region Göteborg, 2011). A major challenge in this vision is to shift towards renewable feedstock and energy carriers. There is also a need for energy efficiency and energy integration within the cluster (Business Region Göteborg, 2011). A first transition towards this vision at Borealis could be to replace a part of the olefins produced via steam cracking with olefins produced from forest residues or other forest products. It is also possible to increase the production capacity when designing and constructing new processes.
Assuming forest residues are the starting material for green olefins, there exist at least four different intermediate products for the production of olefins. One route is to utilize methanol produced via gasification and methanol synthesis, or extraction of methanol from the paper- and pulp industries, in the methanol-to-olefins (MTO) process or the methanol-to-propylene (MTP) process. Another route is to utilize ethanol produced via fermentation in the ethanol to ethylene (E2E) process. In the third route one can directly utilize synthesis gas for the production of olefins in the Fischer-Tropsch to olefins (FTO) process. A fourth alternative is to utilize methane and oxygen to produce ethylene via a process licensed by Siluria, oxidative coupling of methane.

UOP/Norsk Hydro developed the technology of producing olefins from methanol (Methanol-to-olefins, MTO) using a catalyst based on silicoaluminophosphate molecular sieve technology. At their pilot plant at Norsk Hydro’s Porsgrunn facility in Norway it is claimed that they achieve near perfect methanol conversion rates, i.e. almost all methanol fed to the reactor is converted into olefins and other by-products (Hamblett, 1996). The pilot plant which processed 0.5 tons/day of methanol (in 1995) showed stability of the catalyst even after more than 450 cycles of reaction and regeneration. In 1995 Norsk Hydro and UOP announced that the technology was available for license (Hamblett, 1996). However, in 2009 the MTO process still was unproven at commercial scale. An industrial scale demonstration MTO process plant has been built by the French oil and gas company Total in Feluy in Belgium with the intention to verify the performance of a full scale MTO plant (Håvard, 2011). Several projects are currently ongoing in China, where it is possible to exploit cheap and abundant coal via gasification and methanol synthesis (CHEMSYSTEMS, 2009). The first commercial scale installation of the UOP/Hydro MTO process was announced in 2011 (UOP Honeywell, 2011), and started up in September 2013 (Gregor, 2013). The referred process has a production capacity of approximately 300 kton C2, C3 and C4 olefins a year. The Honeywell UOP company announced in November 2012 that they got a third technology license for starting a new MTO project at a company in China (UOP Honeywell, 2012), a fourth license was sold in 2013 (Gregor, 2013), two additional licenses has been sold in 2014.

The MTP process is licensed by Lurgi. It is claimed (Koempel, et al., 2005) that the MTP process licensed by Lurgi is based on an efficient combination of the most suitable reactor system and a very selective and stable zeolite-based catalyst. In the MTP process, methanol is converted to propylene, LPG, Gasoline and to some extent ethylene. There are two plants in commercial operation (470 kt/a), and a third plant in construction. Successful performance test runs of the first world scale MTP plant was achieved in May 2012 (AIR LIQUIDE GLOBAL E&C Solutions, 2013).

BASF is considering constructing a world-scale methane-to-propylene plant via methanol on the US Gulf Coast. Details of the potential investment, including capacity, capital costs, and exact location, are under evaluation, says BASF, which envisions bringing the plant on stream in 2019. The company currently produces propylene in the United States at Port Arthur, TX, where BASF holds 60% of a cracker joint venture with Total. The investment “would secure needed propylene supply for BASF downstream business as less propylene is available in the United States due to the shift from naphtha to ethane feedstock to the cracker plants. BASF is likely to use its own technology for the methane-to-propylene plant using shale gas as feedstock. However, whether or not they are going to use any of the todays known methanol to olefins technologies (e.g. MTO or MTP) or develop their own technology is not yet announced.
In the E2E process, ethanol is dehydrated into ethylene and water using fixed catalytic beds. Two E2E processes, one undiluted and one diluted are licensed by Chematur Engineering AB, where the undiluted is of interest for this project. It was developed in the 1980’s by Scientific Design (SD) based on their catalyst SynDol. SD has also supplied the SynDol catalyst to other producers of ethylene using an iso-thermal process. Chematur Engineering AB (parent company of Weatherly Inc.) engineered (basic and detail engineering) an ethylene plant for the Swedish Civil Defense in the mid 1980’s based on a license from SD. During this time Chematur negotiated and purchased the ethylene technology from SD (Danielsson, 2014). A Chematur undiluted ethanol to ethylene process has been built in Egypt, Sanmar, and is currently (may 2014) in the start-up phase (Danielsson, 2014). The referred process is going to have a total ethylene capacity of 185 kton per year divided into three parallel process lines, i.e. each line having an ethylene capacity of about 60 kton per year. Chematur has in total sold four process licenses for the undiluted E2E process including the one sold to the Swedish Civil Defense and the one sold to Sanmar, Egypt.

In the Fischer-Tropsch synthesis, the reaction is between CO and H₂ in the presence of an active catalyst forming hydrocarbons and alcohols. The reaction can be seen as a surface polymerization reaction, and due to the nature of the reaction, the product stream is going to consist of a range of products instead of a single component. Depending on process conditions, type of catalyst, and chemical promoters it is possible to affect the product distribution of the Fischer-Tropsch process. Fischer-Tropsch to olefins process aims to maximize the yield of lower olefins and to minimize the production of CO₂ (Torres Galvis & de Jong, 2013). Researchers in the Netherlands have demonstrated the direct conversion of synthesis gas through a Fischer-Tropsch process to C2 through C4 light olefins with selectivity up to 60wt% (Galvis & al., 2012). However, no commercial technology is available.

The oxidative coupling of methane (OCM) reaction is a direct reaction, meaning that the desired and targeted products are formed during the reaction, and thus avoiding sequential steps to derive the desired product. In OCM, methane and oxygen react over a catalyst exothermically to form ethane and ethylene and small amounts of higher hydrocarbons (Weinberger, et al., 2012). The OCM process is currently unproven at commercial scale. Siluria and Linde (one of the leading technology and EPC service providers to the global ethylene industry) recently announced an Ethylene Technology Partnership (Market Wired, 2014-06-02). They expect to be offering the technologies to the broader ethylene industry in the second half of 2015.

Due to the setup of the Forest Chemical project, consisting of two main intermediate products in the form of ethanol and methanol, the two (ethanol and methanol) were treated as raw materials for the production of olefins in this part of the Forest Chemical project.

2 Aim and scope

The aim of the Olefin part of the Forest Chemical project was to investigate opportunities in Stenungsund for the production of green olefins (ethylene, propylene and butylenes) origin from 2nd generation ethanol (ethanol produced from cellulosic raw material) and green methanol (methanol e.g. produced via synthesis of syngas derived via gasification of cellulosic material). One requirement in this project was to at least produce 200 kton/year of green olefins origin from forest residues utilizing ethanol and methanol.
The process route for the production of green olefins from ethanol and methanol are commercially available, i.e. proven technologies. For the Ethanol route to green olefins Chematur Engineering was chosen as technology licensor, due to following reasons:

- Located in Karlskoga, Sweden, close to Stenungsund to reduce traveling costs and ease communication
- Contact already established in previous projects
- Well proven process design

For the Methanol route to green olefins the process licensed by UOP was chosen due to the following reasons:

- The flexibility of producing ethylene, propylene and butylene’s in various ratios depending on process conditions
- The only process licensor given positive response to our request
- Ineos (former Hydro and part-developer of the MTO process) is part of the Forest Chemical Project and able to contribute and to work in-kind in the olefins project
- Proven technology and carefully evaluated synthesis

The possibility of producing 60 kton of green ethylene in the Ethanol to Ethylene (E2E) process licensed by Chematur Engineering and the possibility of producing 300 kton of green olefins (C2, C3 and C4) in the UOP/Hydro Methanol-to-Olefins (MTO) process licensed by UOP at Borealis site in Stenungsund has been evaluated based on e.g. profitability, availability, synergies with existing processes (e.g. energy integration), process related aspects, environmental aspects, safety aspect, storage possibilities etc.

A 60 kton E2E process was chosen due to the following reasons:

- A Chematur E2E process with the same production capacity is currently in the start-up phase
- Match available polymer capacity at Borealis PE plant
- Start in a small scale with the intention to explore the green olefins market
- Possibility to increase the capacity by adding additional parallel process lines

A 300 kton MTO process was chosen due to the following reasons:

- A UOP/Hydro MTO process with the same production capacity was started up in China (September 2013)
- More complex and expensive process than the E2E process, i.e. require a rather high production capacity to pay back the investment
- Possible to utilize quite a large turn down ratio if wanted, i.e. if more conventional cracking is beneficial

As part of the project Life Cycle Assessments, LCA, has been made for the entire value chain going from cellulosic material via ethanol and methanol to olefins. The LCA has, however, not been based on the production of 300 kton of green olefins from the UOP/Hydro MTO process and 60 kton green ethylene from the E2E process. Implementation of the MTO process would imply that olefins
produced via steam cracking at the plant would have to be decreased. The purpose of the LCA, Olefins part of the project, was to compare the environmental effects of producing olefins via steam cracking (2011) with a fictive case including steam cracking, MTO and E2E. To be able to compare the two cases the production of ethylene was set as base, i.e. the same production of ethylene in both cases. The LCA assessment for the fictive case has therefore been based on approximately 200 kton green olefins from the MTO process, 30 kton of green ethylene from the E2E process and the conventional olefins from the cracker was adjusted so that the ethylene production would match the ethylene production of year 2011.

The project as such is a sustainability driven project, an investigation of possible transitions from non-renewable resources to renewable resources in the form of forest material. However, to be able to implement any of the more sustainable alternatives (at least seen in the resource perspective), such as E2E or MTO, they have to be economically attractive. In that perspective, the project is also economically driven. In terms of economic incentives, today, green raw materials for the production of Olefins are less attractive then fossil. For the project to be economically attractive to implement, either the green raw materials have to be more cost competitive compared with fossil ones, or it has to be possible to charge a higher price for the green olefins, a so called green premium price.

3 Studies performed within the Olefins project

3.1 MTO process

Following studies for the MTO process has been performed as a part of the Olefins project within the Forest chemical project:

- UOP (Licensor): UOP/Hydro MTO™ Process Technical information package (Confidential)
- Chalmers master thesis work: MTO process description (Public) (Johansson, 2013)
- Chalmers Geo Engineering: Storage of MeOH/EtOH in existing rock caverns; Screening perspective (Public) (Fransson, et al., 2014)
- Bilfinger Industrial Services: Storage of MeOH/EtOH in underground caverns; Engineering perspective including cost estimation (Public) (Johansson, 2014-06-26 revA)
- COWI: Constructability study (Confidential)
- Borealis: Cost estimate development
- Borealis: Mass balance study for the MTO process integrated with the cracker
- Borealis: Business case development
- Ineos (former Hydro): Impact of co-feeding methanol and ethanol in the MTO process
- Borealis: LCA-data input

3.2 E2E process

Following studies for the E2E process has been performed as a part of the Olefins project within the Forest chemical project:

- Chematur Engineering (Licensor): Engineering study for a 60 000 TPY Ethanol to Ethylene process (Confidential)
- COWI: Constructability study (Confidential)
- COWI/Borealis: Cost estimate development
- Borealis: Business case development
4 Chemical process

4.1 MTO process

In the MTO process, methanol is converted, mostly, into light olefins such as ethylene, propylene and butylenes. The MTO process is a catalytically driven process. A proposed chemical reaction for the MTO process over catalyst SAPO-34 consists of three steps seen below, and also illustrated in reaction 1 (Dahl & Kolboe, 1993):

1. Dehydration of methanol to dimethyl ether (DME)
2. Formation of intermediate species [CH₂]
3. Formation of olefins and other paraffins

\[
2\text{CH}_3\text{OH} \rightleftharpoons \text{CH}_3\text{OCH}_3 (+ \text{H}_2\text{O}) \rightarrow [\text{CH}_2] \rightarrow \text{Olefins}
\]  

Due to that the reaction mechanism of the MTO process is not fully understood, and due to that the source of the olefins are uncertain, the methanol and DME can be lumped to one single component called oxygenates (Chen, et al., 2007). A proposed reaction network for oxygenates can be seen in Figure 2. The reaction constant, \( k \), is for example dependent upon temperature, i.e. the reaction temperature will affect the outcome of the process.

![Figure 2: Proposed reaction network for the MTO process over SAPO-34, methanol and DME lumped as oxygenates.](image)

In the MTO process, methanol fed to the reactor is converted to light olefins with carbon selectivity of about 75 to 80% towards ethylene and propylene. Product ratio between ethylene and propylene (ethylene production rate divided with the propylene production rate) is in the range of 0.7 to 1.4, depending on different process parameters such as:
1. Operating temperature (where higher temperatures favouring higher ethylene to propylene ratios)
2. Operating pressure (where higher pressure leads to higher propylene ratio)
3. Water content in the feed, possible to use up to 20mole% water (where higher water content leads to higher ethylene ratio) (Vora, et al., 2001).

However, it is the temperature having most effect on the outcome of the reaction. The combined reaction network is exothermic, i.e. the process will generate heat during reaction.

The methanol/DME reacting on SAPO-34 will also to some extent generate coke. The coke has a deactivating effect on the catalyst (Vora, et al., 2001), and therefore there is a need for catalyst regeneration. The regeneration will occur continuously by burning of coke in a separate compartment called regenerator.

After reaction the gas has to be purified in a sequence of generally known separation processes such as absorption, quenching, caustic cleaning and distillation. A schematic figure of the MTO process can be seen in Figure 3. Second part of the figure, the separation processes, is similar to a separation sequence in a naphtha cracker. In order to save capital investments and to utilize available equipment at Borealis as much as possible, the idea is to integrate the MTO process with the existing cracker and utilize the existing separation processes, referred to as the light end (LE) and ethylene recovery (ER). The MTO process will generate two product streams, one vapour containing C4- and one liquid containing C4+. The vapour stream is assumed to be fed to the existing caustic wash at the cracker and the liquid stream at front-end of the cracker plant.

Figure 3: The MTO process licensed by UOP. First part (similar to Refinery/FCC Equipment): Reactor, regenerator, water quench and oxygenate recovery system included in scope. Second part: Separation sequence, assumed to utilize existing one at the cracker plant.
The MTO reactor and regenerator scheme, the quench scheme and the DME & oxygen recovery scheme is further described in Appendix 1 to Appendix 3. As mentioned, the rest of the scheme in Figure 3 is similar to the separation processes available at the cracker plant, conventional separation sequence at a naphtha cracker. The separation sequence e.g. includes the following unit of operations and main equipment, caustic wash, dryers, demethanizer, deethanizer, depropanizer, debutanizer, C2 splitter, C3 splitter etc. After the separation sequence products such as polymer grade ethylene, chemical grade propylene, C4 raff, H2 and fuel gas will be available.

As the production of paraffins (ethane and propane) are much less in a MTO process than in a conventional cracker, distillation processes will actually be off loaded when implementing a MTO process. In addition, acetylene contents from the MTO process are almost zero, reducing the loads on the acetylene reactors.

4.2 E2E process

In the E2E process, ethanol is catalytically dehydrated to produce ethylene according to the chemical reaction (2) seen below (Weatherly Inc. Chematur Engineering group, 2010)

$$C_2H_5OH \rightarrow C_2H_4 + H_2O$$

The reaction is endothermic requiring about 1630 kJ/kg ethylene produced. As recovery of unconverted ethanol for recycling is energy and capital intensive, reaction conditions enabling 99%+ conversion of ethanol are usually preferred in the process licensed by Chematur Engineering.

The conversion of ethanol is highly selective with the reaction product containing essentially only ethylene and water. Other constituents are traces of methane, carbon oxides, ethane, propylene, butanes, acetaldehydes, diethyl ether, hydrogen and any unconverted ethanol.

Processing of the reactor effluent to produce polymer grade ethylene is quite simple. It is first cooled to allow most of the water to condense. The resulting gas stream is compressed, water-scrubbed to remove oxygenated impurities, cooled to about -40°C to remove condensables, dried in a molecular sieve bed to remove residual water, distilled in a C2 splitter to remove ethane and stripped to reduce CO forming polymer grade ethylene. A process flow sheet representing the ethanol to ethylene process licensed by Chematur Engineering can be seen in Figure 4. The ethanol to ethylene process is further described in Appendix 4.
Process design

The main product at Borealis steam cracker is ethylene, the rest (e.g. propylene, butylenes, etc.) is seen as by-products. Ethylene is delivered to all industries in Stenungsund, and to cover the entire demand of the Stenungsund industries, ethylene is also being imported to Stenungsund. During 2011, the cracker produced 590 kton of ethylene, which has been defined as minimum requirement for the study. The olefins at the steam cracker are produced from various raw materials such as ethane, propane, butane and naphtha. Which raw materials used and to what extent is optimized on weekly basis based on profitability by using an off-line program titled PIMS. PIMS is optimizing the cracker operation e.g. based on current raw material and product prices and different process capacities of the cracker plant.

As mentioned earlier, the MTO process is planned to be integrated with the existing separation processes at the cracker plant resulting in less olefin production via conventional steam cracking during MTO process operation. The program PIMS has been utilized using the MTO product as a side stream to optimize the cracker plant integrated with the MTO process resulting in mass balances (maximum capacity of the plant).

The MTO process evaluated in this study has a maximum production capacity of approximately 300 kton C2, C3 and C4 olefins. The MTO process has however quite a large turndown ratio (operation ratio min/max). Standard design is based on a hydraulic turndown to 70%, but designs down to 60% can also be managed, leading to increased flexibility. Sixteen (16) different cases were modelled in PIMS, based on the MTO process combined with existing cracker set-up, using following sets of parameters:
• Two different operational base cases for the cracker (cracking of 40 t/h ethane and 68.5 t/h ethane, 68.5 t/h is the most likely future scenario)
• Two different operational base cases for the MTO process (high ethylene and high propylene)
• Four different MTO process sizes (200, 230, 270 and 300 kton C2, C3 and C4 olefins). However, further on in this report, only the 300 kton cases will be elaborated due to readily available data for such unit from UOP.

The E2E process is planned to be a “stand-alone process”, and will therefore not affect the rest of the cracker process. The E2E process evaluated in this study has a maximum production capacity of 60 kton ethylene.

The profitability of the MTO process integrated with the cracker has been based on the production of 270 kton of green olefins a year assuming maximization of propylene in the MTO process. Maximization of propylene has been assumed due to the assumption that the use of ethane will be increased in the future, resulting in decreased propylene production.

Profitability of the E2E process has been based on the production of 60 kton of green ethylene from ethanol.

The study case for the LCA has been chosen in order to, as much as possible, match the production of 2011 and been based on:

• Production of 590 kton of ethylene (same as 2011) (generated by the steam cracker plus the MTO process plus the E2E process)
• 30 kton green ethylene from the E2E process
• Approximately 200 kton green C2, C3 and C4 olefins from the MTO process
• ~68.5 t/h of ethane cracking
• Maximization of propylene in the MTO process

5.1 MTO process
The MTO process design has been based on confidential information received from UOP in the form of a Technical Information Package containing information e.g. regarding process description, feedstock specifications, mass- and energy balances (both for high ethylene and high propylene operational cases), PFD’s and equipment data sheets. However, due to the confidentiality agreement between Borealis and UOP, most of this information cannot be presented in this public report.

For the mass integration of the MTO process with the steam cracker, as mentioned, the program PIMS has been utilized using input from UOP Technical Information Package, making it possible to present an overall material balance for the MTO process combined with the steam cracker. Energy balance, steam balance, for the MTO process combined with the cracker has to some extent been based on the simulation performed in PIMS. Information regarding steam flow through turbines, compressors, dilution steam, steam production from the furnaces etc. was directly obtained from PIMS. Steam through heat exchangers was based on a study made at the cracker (Hedström & Johansson, 2008) and adjusted based on mass flows. For example the steam flow through a reboiler was adjusted using the ratio between the “old” and the new flow through the top of the column simulated in PIMS. Steam demand from the MTO process was used as an input to the combined steam balance. The boiler load was then adjusted to match the steam deficits. Required make-up of
natural gas was obtained using information from PIMS (usage of energy – bought energy = produced energy) and adjusted based on the change in boiler load.

As a part of the MTO process study, the possibilities of storing methanol in underground caverns has been investigated by Chalmers Geo Engineering and Bilfinger Industrial Services and is presented in the report.

The possibilities to actually fit a MTO process (excluding separation processes) at the existing cracker plant and to implement the integration between the MTO process and the existing cracker have been investigated by COWI.

5.1.1 Feedstock specifications
Methanol specifications required for the MTO process can be seen in Table 1 below.

Table 1: Methanol specifications for the MTO process

<table>
<thead>
<tr>
<th>Item</th>
<th>Specification</th>
<th>Test Method</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methanol (dry basis)</td>
<td>98.85 wt-% min</td>
<td>IMPCA 001-02</td>
</tr>
<tr>
<td>Water plus DME</td>
<td>5.0 wt-% max</td>
<td>ASTM E1064-12</td>
</tr>
<tr>
<td>Appearance</td>
<td>Clear and free of suspended matter</td>
<td>IMPCA 003-98</td>
</tr>
<tr>
<td>Color (Pt-Co)</td>
<td>5 max</td>
<td>ASTM D1209-05</td>
</tr>
<tr>
<td>Hydrocarbons</td>
<td>Pass Test</td>
<td>ASTM D-1722-04</td>
</tr>
<tr>
<td>Acidity (as acetic acid)</td>
<td>30 wt-ppm max</td>
<td>ASTM D-1813-06</td>
</tr>
<tr>
<td>Total Organic Nitrogen</td>
<td>1 wt-ppm max</td>
<td>ASTM D4829</td>
</tr>
<tr>
<td>Halides</td>
<td>2 wt-ppm max</td>
<td>IMPCA 002-98</td>
</tr>
<tr>
<td>Alkaline/Akline Earth Metals</td>
<td>1 wt-ppm max</td>
<td>UOP Method 389-10</td>
</tr>
<tr>
<td>Sulfur</td>
<td>1 wt-ppm max</td>
<td>ASTM D-5453-09</td>
</tr>
<tr>
<td>Nonvolatile matter</td>
<td>50 mg/1000ml max</td>
<td>ASTM 1353-09</td>
</tr>
<tr>
<td>Potassium Permanganate Test Time at 15°C</td>
<td>30 Minutes</td>
<td>ASTM D1363-06</td>
</tr>
</tbody>
</table>

Water plus DME content – This case assumes relatively high water content of 5wt%. The MTO process does not require a high purity feed such as AA-Grade methanol.

5.1.2 Product specifications (incl. by-products)
As mentioned, the MTO process is planned to be integrated with the cracker process, i.e. a raw product feed from the MTO process will be fed to light end of the cracker process, affecting the amount and composition of raw product from the cracker furnaces. There will be two hydrocarbon product streams from the MTO reactor, one liquid product and one vapor product. The vapor product is assumed to be fed to the caustic tower of the light end, and the liquid product further upstreams in the cracker plant. The MTO process is flexible when it comes to ratio between ethylene and propylene, i.e. one can either maximize the production of ethylene, high ethylene, or propylene, high propylene, by e.g. varying reaction temperature within the reactor. Due to the confidentiality agreement, the actual used product specifications cannot be included in the report.

5.1.3 Process/Systems integration
Heat integration is applied within the MTO process design, all steam that is generated within the process is also used within the process. The steam generated within the process is however not always enough (differs between the high ethylene and the high propylene cases and whether or not
one chooses to implement turbine or electric driven motors), so an input of steam produced at the cracker plant has to be included. There will however be some excess heat with an adequate temperature available within the MTO process. This intermediate stream from the MTO process that requires cooling can be used as heating media in the C3-splitter reboiler at the cracker plant. The heat available in the MTO stream exceeds the required heat in the reboiler. The remaining heat can be utilized for pre-heating the methanol feed coming from storage. Additional utilization of this low temperature heat within the cracker plant, require further evaluation. But due to that LP steam today often is available in excess this has to be investigated further. More elaborate energy integration would require an extensive steam and energy analysis/optimisation of the whole cracker, including several scenarios which is out of scope for this study.

Stripped water, containing traces of methanol, is generated in the water stripper in the MTO process. The intention is to utilize the stream as boiler feed water for generation of dilution steam that is used to dilute the conventional cracker feed to the furnaces. The stripper water flow of about 60 000 kg/h is pressurized to 9 barg and vaporized using approximately 83 000 kg/h 20 barg steam. 20 barg steam is assumed to be produced by reducing 85 barg steam available and mix it with 20 barg 130°C condensate (67.3 t/h 85 barg steam and 15.7 t/h 20 barg condensate).

The requirement of dilution steam at the Cracker plant today is estimated to 80-90 ton/h at full production and about 60 ton/h when the MTO plant is in operation. Operating the MTO process at maximum production, it should be possible to decrease the outtake of raw water for dilutions steam production with about 80 to 90 t/h i.e. operating the MTO process at maximum capacity (about 60 t/h of “waste” water production) it would be possible to decrease the demand of “conventional” dilution steam production with 100%. The contribution of dilution steam from stripped water vaporization is thereby assumed to replace a major part of the 8.8 barg steam used as dilution steam today.

To be able to implement the MTO water as dilution steam a thoroughly steam balance analysis is needed as dilution steam at 8.8 barg will be reduced, but demand for 20 barg steam increase. One possibility could be to rebuild the existing turbo alternator (that today produces electricity by expanding 85barg steam to 8.8 barg steam) to add another extract tap at 20 barg. The upside potential with the idea is that the cracker outtake of raw water can be substantially decreased.

5.1.4 Process description
The plant will be designed to produce 300 kton olefins (C2, C3 and C4) per year from bio based methanol. The plant will use approximate 100 ton dry methanol per hour as feedstock into the plant. Methanol is assumed to be supplied from offsite storage in two underground caverns (see Section 5.2 and on-wards). The methanol supplied from the caverns will be heated and vaporized utilizing internal heat exchanging to reach appropriate temperature for the MTO process. Some of the vaporized methanol is fed to an oxygenate purification sequence to purify the product gas and to recover some of the unreacted methanol and DME. The methanol stream is then fed to the MTO reactor for reaction. After the MTO reactor, the product gas is fed to a quench scheme to purify it and quench the reaction. The product gas is then directed to the oxygenate recovery scheme before it is being fed to the existing system at the cracker plant, Figure 5. For more information see Chemical process, MTO process Figure 3.
5.1.5 Overall material balance

Due to that the MTO process is assumed to be integrated with the cracker process and fed to the light end of the process, the composition and mass of the MTO product will affect amount and composition of the total feed to the cracker furnaces. Therefore, the planning optimization program PIMS has been utilized, as mentioned in Section 5, using the MTO product as a side stream to the steam cracker model. Two different cracking base modes has been used, one with ethane cracked fixed at 40 t/h whith ethane cracking fixed at 68.5 t/h (rest of the raw material is optimized by PIMS). The first case corresponds to today’s operation and the second case corresponds to an ongoing project at the cracker. In the second case, the capacity of the C2-splitters in the ethylene recovery was assumed to be increased with 10% due to planned activities.

Mass balance for the most probable operating case, cracking of 68.5 t/h ethane and maximization of propylene can be seen in Table 2 and Table 3, included is also “today” operation and “future” operation with no MTO included for comparison. To notice is the decrease in propylene production going from 40 to 68.5 t/h ethane cracking. The decrease in propylene production can be solved by introducing a MTO process that is operated to maximize the propylene production, which is seen as an upside potential. Raw material feed for the other cases can be seen in Table 30 and Table 32 in Appendix 7, included is also a case for the cracker process without integration with a MTO process. Products from the cracker process combined with the MTO process can be seen in Table 31 and Table 33 in Appendix 7.

When generating the overall material balances, especially to generate an overall energy balance to obtain the required make-up of natural gas, steam balances for each case was developed. An example of a steam balance, 40 t/h C2 no MTO, can be seen in Table 29 in Appendix 6. The main difference between the cases is the amount of HP steam generated from the furnaces, steam requirement of the MTO process, dilution steam and the calculated boiler load. In Table 28 in Appendix 5 are these differences presented.

The overall mass- and energy balances have been based on a few assumptions:
- Boiler efficiency of 85% → 3.2964 GJ of gas / ton HP steam produced
- Fuel gas sold: 333 GJ/h
- Other fuel gas uses (excluding furnace fuel and boiler fuel): 26.1 GJ/h
- Furnace fuel from PIMS
- Dilution steam from PIMS
- Steam to turbines etc. from PIMS
- “Export” of 25 t/h HP steam
- Change in original MTO process design
  - Electrical driven compressor and air blower (original design steam driven)

### Table 2: Feed to Cracker furnaces and the MTO process. Cracking of 68.5 t/h ethane and maximization of propylene (HP) in the MTO process, integrated with the cracker

<table>
<thead>
<tr>
<th>Fresh feed t/h</th>
<th>No MTO 40 t/h ethane (“today”)</th>
<th>No MTO 68.5 t/h ethane</th>
<th>HP 300 kton olefins 68.5 t/h ethane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethane</td>
<td>40</td>
<td>68.5</td>
<td>68.5</td>
</tr>
<tr>
<td>Propane</td>
<td>20.9</td>
<td>0.16</td>
<td>19.9</td>
</tr>
<tr>
<td>Butane</td>
<td>90</td>
<td>68.3</td>
<td>1.4</td>
</tr>
<tr>
<td>Naphtha</td>
<td>22.11</td>
<td>22.11</td>
<td>22.11</td>
</tr>
<tr>
<td>Total fresh feed</td>
<td>173</td>
<td>159</td>
<td>112</td>
</tr>
<tr>
<td>Methanol (100wt%) feed to MTO</td>
<td>0</td>
<td>0</td>
<td>100</td>
</tr>
<tr>
<td>Ethanol (for ETBE)</td>
<td>2.05</td>
<td>1.73</td>
<td>0.5</td>
</tr>
<tr>
<td>NG as make-up (48 MJ/kg)</td>
<td>1.5</td>
<td>3.5</td>
<td>9.2</td>
</tr>
</tbody>
</table>

1Same amount of electricity as used 2011 is assumed
2Electricity for the MTO case: same as 2011, minus 2 MW due to reduction of feed through raw gas compressor, plus electricity to drive pumps, compressors, blowers etc. in the MTO process

### Table 3: Product from Cracker furnaces combined with the MTO process. Cracking of 68.5 t/h ethane and maximization of propylene (HP) in the MTO process, integrated with the cracker

<table>
<thead>
<tr>
<th>Products t/h</th>
<th>No MTO 40 t/h ethane (“today”)</th>
<th>No MTO 68.5 t/h ethane</th>
<th>HP 300 kton olefins 68.5 t/h ethane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethylene</td>
<td>75.2</td>
<td>77.1</td>
<td>77.8</td>
</tr>
<tr>
<td>Propylene</td>
<td>27.3</td>
<td>19.5</td>
<td>28</td>
</tr>
<tr>
<td>ETBE</td>
<td>4.5</td>
<td>3.8</td>
<td>1.1</td>
</tr>
<tr>
<td>CBFS</td>
<td>2.1</td>
<td>2.3</td>
<td>1.8</td>
</tr>
<tr>
<td>Raw and treated SCN</td>
<td>15.2</td>
<td>15</td>
<td>13.6</td>
</tr>
<tr>
<td>Raff 2 (C4 olefins)</td>
<td>13.5</td>
<td>11.2</td>
<td>11.1</td>
</tr>
<tr>
<td>H₂</td>
<td>0.25</td>
<td>0.25</td>
<td>0.25</td>
</tr>
<tr>
<td>Fuel gases (57.26 MJ/kg)</td>
<td>35.7³</td>
<td>32.5³</td>
<td>24.9³</td>
</tr>
<tr>
<td>Stripped water to BL</td>
<td>-</td>
<td>-</td>
<td>60.92</td>
</tr>
<tr>
<td>Waste water to WWT</td>
<td>-</td>
<td>-</td>
<td>6.75</td>
</tr>
<tr>
<td>Coke (burned in regenerator)</td>
<td>-</td>
<td>-</td>
<td>1.16</td>
</tr>
</tbody>
</table>

³Whereof 5.8 ton sold (333 GJ/h)

Energy used (electricity, natural gas as make-up and fuel gas) in MWh per ton of ethylene and propylene produced was calculated using information from Table 2 and Table 3 above. The
calculations resulted in that MTO HP 300 kton olefins 68.5 t/h ethane amounted to 85% of the estimated energy usage “today” (No MTO 40 t/h ethane), see Table 4.

<table>
<thead>
<tr>
<th>Weighted Energy usage MWh/tethylene + propylene</th>
<th>No MTO 40 t/h ethane</th>
<th>No MTO 68.5 t/h ethane</th>
<th>HP 300 kton olefins 68.5 t/h ethane</th>
</tr>
</thead>
<tbody>
<tr>
<td>100%</td>
<td>101.4%</td>
<td>85%</td>
<td></td>
</tr>
</tbody>
</table>

300 kton of green C2, C3 and C4 olefins from the MTO process is maximum capacity. A turn-down ratio of 60% is possible, and a more realistic, profitable, starting point is to produce less than maximum capacity, and also assume a cracker furnace operating case lower than maximum. A study case has been chosen to be used for the LCA analysis, and in this study case, production of 30 kton of green ethylene from the E2E process has also been included. As mentioned, the study case has been chosen to match the production of ethylene in 2011:

- Production of 590 kton of ethylene (cracker + MTO process + E2E process)
- 30 kton green ethylene from the E2E process
- Approximately 200 kton green C2, C3 and C4 olefins from the MTO process
- ~68.5 t/h of ethane cracking
- Maximization of propylene in the MTO process

An overall material balance for the LCA study case can be seen in Table 34 and Table 35 in Appendix 8, also including some key figures for the production of 30 kton of green ethylene from the E2E process.

5.1.6 Utility requirements
Utility description for the MTO process can be seen in Appendix 9.

5.1.7 Material handling description
The MTO reactor and regeneration scheme is a continuous reaction, and some catalyst is continuously consumed in the MTO process and fresh catalyst needs to be added to the MTO process. Catalyst material will automatically be added to the reactor/regenerator during production. However, catalyst has to be loaded into a catalyst hopper a few times a week. Bulk storage of fresh catalyst is required on the Borealis site.

5.1.8 HSE
The plant shall be configured to minimize emissions and discharges during construction phase and operation.

An early safety study has been performed to verify the placement and safety distance between the equipment at the MTO-plant and possible impact on surrounding installations at Borealis. Special attention has been paid to the nearby propane loading unit and also the control room next to the MTO area. In further studies, an internal HAZOP study will be performed, with special attention paid to the MTO reactor and regenerator, since these are vulnerable equipment.
5.1.8.1 Emissions, effluents, and waste streams

Emissions
Emissions from the production plant and environmental impact are not investigated in detail in this study. However, the following sources of emissions to atmosphere are identified within the MTO plant:

- Regenerator flue gases from regeneration flue gas filter and the catalyst recovery system will be released to atmosphere through the vent stack.

- Flue gases from gas fired start-up heaters; air heater for start-up of the catalyst regenerator, reactor start-up gas preheater and reactor start-up gas superheater. Estimated total fuel gas capacity: 10 MW taken from Borealis fuel gas header. The start-up heaters Operates during a 12 to 24 h start-up period.

- Ejection steam/air/nitrogen from the fresh catalyst hopper and spent catalyst hopper steam jet ejectors. If the stream contains catalyst fines, additional filters will be required to minimize emissions of particles.

- Flow of nitrogen from start-up gas preheater and start-up gas superheater will be released to atmosphere during start-up procedure before methanol is introduced to the reactor.

Overall emissions from the cracker combined with the MTO process are included in the more comprehensive LCA study.

Effluents
Storm water from the process area will pass through oil separators and applicable treatment (Borealis existing system) before it is released to the recipient.

Water and foam used at firefighting will be collected and treated in the existing system at Borealis.

Waste streams
Waste water
The process produces a waste water stream that contains some amounts of methanol, sodium acetate and small amounts of other substances. This stream will be sent to the municipal waste water treatment plant (WWTP). Methanol and the other substances present in the waste water stream are easily degradable in a WWTP. However, the bacteria involved in the nitrogen removal step are sensitive to changes in their environment. As a further investigation regarding the waste streams, it is recommended to do some tests according the standard ISO 9509:2006, to assess the inhibition of the active organisms in the WWTP when the MTO waste stream is sent to the WWTP.

Before sending the stream, the pH also needs to be adjusted to a lower level, from 9-10 to 6-8, and the temperature needs to be lowered from 90°C to around 10°C. This can be performed in a buffer tank placed upstream of the municipal sewer.
Catalyst
Spent catalyst and catalyst fines from the MTO reactor system will be collected. To minimize environmental impact a plan for collection and destruction of the disposal needs to be developed in more detailed studies.

Area classification
The area classification of the MTO plant will be according to ATEX standard SS EN 60079-10 and the Swedish interpretation of the standard. The classification will be part of later stages of the project and will take the existing and surrounding facilities into consideration.

The overall philosophy should be to minimize the need for classification by concentrating flammable goods at certain areas.

5.1.9 Equipment description
Main equipment includes (for more comprehensive descriptions see Appendix 1 MTO reactor/regenerator description, Appendix 2 MTO Water quench description and Appendix 3 MTO deoxygenate description):

- Continuous MTO reactor regenerator scheme including catalyst coolers – Similar to a FCC type of reactor
- Stack, including flue gas filter – release of flue gases from regenerator
- Air blower – Air is blown continuously into the regenerator (electrical driven motor, in original design turbine driven)
- Catalyst storage – Feeding reactor/regenerator, storage for spent catalyst, storage for fresh catalyst
- Quench tower – Column using addition of caustic to neutralize and to separate out the most dirtiest water
- Product separator – Distillation column, mainly separation of main products and water
- Water stripper – Column to strip the most of the water, almost clean water is the result
- Raw gas compressor – Compressor to increase product pressure to enable further cleaning and separation (electrical driven motor, in original design turbine driven)
- DME stripper column – Strip the main product (liquid) form DME
- Oxygenate absorber – Absorption of DME/methanol using methanol from the main product (gas)
- Pumps
- Reboilers
- Condensers
- Heat exchangers
- Steam drums
5.1.10 Plot plan and Layout

![3D model of the MTO process, adapted to Borealis site conditions](image)

5.1.11 Operation
For the operation of the MTO process it has been assumed that 1 additional operator per shift (1*5), 1 additional day operator and 2 additional engineers, in total 8, is required for the operation of the MTO process.

5.2 Storage of MeOH/EtOH in underground caverns
Pre-studies for storage of MeOH/EtOH in underground caverns has been made by Chalmers Geo Engineering (Storage of methanol in existing rock caverns, Pre-study and solutions suggested for further assessment 2014-08-20) and Bilfinger Industrial services (Skogskemi Storage of methanol and ethanol in caverns revA 2014-06-26). Below the most important results from the studies are presented, for further and more detailed information, see the referred reports.

5.2.1 Storage of MeOH/EtOH in existing rock caverns
The pre-study performed by Chalmers Geo Engineering (Fransson, et al., 2014) consists of three main parts:

- A minor literature survey
- General description of the system
  - a. Product to be stored (MeOH/EtOH)
  - b. Rock mass
  - c. Fractures
  - d. Water
  - e. Previously stored product (light oil)
- Advantages and disadvantages of proposed solutions
5.2.1.1 Storage and offsite description

The principle of storing hydrocarbons in underground caverns is that the product confinement is secured by the groundwater pressure in rock fractures. Pressure of product must be lower than the actual groundwater pressure (both walls and roof) and the key aspect is that the product does not leak out in the groundwater (Fransson, et al., 2014).

According to Fransson, et al., (2014) there are two main alternatives for the storage of methanol or ethanol in caverns, either unlined caverns or lined caverns, both defined below.

Unlined caverns

Unlined caverns means that naked rock will be more or less visible in the cavern, the product will interact with groundwater and previous contaminants on site.

Extensive work has been carried out in with regard to storage of gas in lined or unlined caverns. Gas storage use either low temperature (liquid state storage) or high pressure for containing the gas, which introduce the possibility to simplify and use concepts for storage of temperate liquids in caverns.

For an unlined solution of MeOH/EtOH storage allowable effects on surrounding environment and on stored product is the main focus of further work.

By waste majority, unlined caverns are used for products that don’t mix with water, such as oil and naptha.

Lined caverns

Lined caverns means that the rock surfaces are covered with a lining material that essentially cancels all interaction between the stored product and the surrounding.

A cavern lining may be constructed from concrete combined with a suitable liner material. Materials covered are plastic/rubber and steel. Alternatives described in the report (Fransson, et al., 2014):

- Steel lining
- Membranes
- Spray-on liners
- Geo-membranes and rubber sheets
- Flexitanks

Conceptual description of the system rock-groundwater

The rock matrix and the nearby fracture system can be expected to be contaminated with light oil. The storage was built in an era where environmental concerns were of less importance than today. Combining the fact that the rock is generally low permeable in the area, and that the intended storage product is lighter (light oil) then the water and does not mix with water forms the basis that the stored product stays in the caverns as long as the groundwater flow is directed inwards.

The present day situation with much stricter regulations regarding describing environmental impacts and level of allowable impacts, together with a product (methanol) that mixes with water make the understanding of the system rock-groundwater-product much more important.
5.2.1.2 Comparison of possible solutions

In below table, Table 5, the main results from the study (Fransson, et al., 2014) are presented. Note that “no action” and “grouting” alternative are questionable since it is uncertain whether they fulfil dilution/contamination aspects.

Table 5: Comparison of “pros and cons” for the evaluated alternatives. Green: good, yellow: intermediate, red: bad.

<table>
<thead>
<tr>
<th>Group \ Alternative</th>
<th>Unlined</th>
<th>Lined</th>
</tr>
</thead>
<tbody>
<tr>
<td>Aspect</td>
<td>No action</td>
<td>Grouting</td>
</tr>
<tr>
<td>Time for planning/design</td>
<td>Green</td>
<td>Yellow</td>
</tr>
<tr>
<td>Times for cavern works</td>
<td>Green</td>
<td>Yellow</td>
</tr>
<tr>
<td>Contaminant transport/effect on surrounding</td>
<td>Red</td>
<td>Green</td>
</tr>
<tr>
<td>Difficulty/complexity of solution</td>
<td>Green</td>
<td>Yellow</td>
</tr>
<tr>
<td>Effect on product</td>
<td>Red</td>
<td>Green</td>
</tr>
<tr>
<td>Familiarity of solution</td>
<td>Green</td>
<td>Yellow</td>
</tr>
<tr>
<td>Price</td>
<td>Green</td>
<td>Yellow</td>
</tr>
<tr>
<td>Further work</td>
<td>-allowable mixing -Environment/ surrounding</td>
<td>-Allowable mixing -Environment/ Surrounding -Identifying leaking spots</td>
</tr>
</tbody>
</table>

5.2.1.3 Further work

An unlined approach of the storage needs an extensive environmental impact analysis containing groundwater modelling that describe flow patterns and contaminant spread in the area, with input and follow-up with data from monitoring wells.

A lined solution with drainage between liner and rock will guarantee that the diesel contamination closest to the caverns will flow toward the caverns where it can be treated. A general approach like this is anticipated to give a simpler analysis needed for the environmental impact analysis and higher probability for shorter time with the authorities and their acceptance. Focus can then be directed towards designing and optimizing the works in the caverns.

5.2.2 Skogskemi Storage of methanol and ethanol in caverns

The project scope of the pre-study performed by Bilfinger Industrial Services (Johansson, 2014-06-26 revA) is to define measures for storage of MeOH/EtOH in existing caverns at Vattenfall in Stenungsund. The measures identified and scope included in this pre-study is following, whereof some will be more elaborated than others in this summary:

- Cleaning of caverns
- Measures to prevent explosive atmosphere and to prevent dangerous gases to penetrate spaces nearby caverns
• Measures to prevent high in-leakage of water into caverns (unlined and lined solutions)
• Exchange of submersible pumps in caverns
• Piping for methanol from Vattenfall harbour to caverns, and from caverns to MTO-process located at Borealis cracker plant
• Cavern modifications and measures for access of cavern cleaning and reconstruction work.
• Loading arm and instrumentation for the system including supervision system for the caverns
• Cost estimate ±40%

5.2.2.1 Design capacity

There are totally 11 caverns at Vattenfall area in Stenungsund. For storage of MeOH/EtOH caverns OL10 and OL11 are to be used. Total storage capacity in OL10 and OL11 is approx. 100 000 m³ (60 000 + 40 000 m³) and for MeOH/EtOH this means approx. 80 000 ton. Today it is a fixed water bed level in the caverns, which is controlled by pumping out the water to waste water treatment plant located in Vattenfall harbour. Caverns OL10 and OL11 have today the following design data:

Cavern OL10
• Storage capacity: 60 000 m³
• Temperature: 14°C
• Ground water level: min 5 m above top of storage
• Distance from cavern to top of mountain: about 10 -15 m

Cavern OL11
• Storage capacity: 40 000 m³
• Temperature: 14°C
• Ground water level: min 5 m above top of storage
• Distance from cavern to top of mountain: approx. 10 -15 m

The temperature in the caverns can, when used for storage of methanol, differ slightly from the temperature data given above depending on the temperature of the incoming methanol.

Imported feed to cavern OL10, OL11:
• Mass flow methanol: 960 ton/h
• Volume flow methanol: 1200 m³/h
• Operating pressure methanol: max 9 bar(g)
• Operating temperature methanol: ambient/water temperature

Feed to MTO-process:
• Mass flow methanol: 102.5 ton/h (dry)
  o max 108 ton/h (incl. moister)
• Volume flow methanol: 129 m³/h (dry)
  o max 140 m³/h (incl. moister)
• Operating pressure methanol: pressure created by submersible pumps – pressure drop in piping system
• Operating temperature methanol: 14°C
Methanol from caverns upstream the MTO-process is at the conditions described above. Operating conditions for methanol to MTO-process are as described below.

- Operating pressure of methanol to MTO-process: 10 bar\( (g) \) (met by a booster pump)
- Operating temperature of methanol to MTO-process: 35°C (met by heat exchanging with excess heat from the MTO process)

### 5.2.2.2 Product specification

About 5% water in the methanol is acceptable for the methanol feed to the MTO-process; however the quality of the water is of high importance. The quality of the water that leak into the caverns was measured. The alkali content in the leakage water is, according to performed measurements too high to be acceptable for the MTO-process, i.e. measurements have to be taken to prevent water to mix with the methanol, e.g. utilize lined caverns.

### 5.2.2.3 Process description

Methanol is assumed to be unloaded by tankers at Vattenfall harbour in Stenungsund and pumped to Vattenfall caverns OL10 and OL11 in the existing 12” import line. A continuous flow of methanol is to be pumped to the MTO-process located at the SCN-area at the cracker plant at Borealis.

### 5.2.2.4 Overall material balance

To handle the large quantities required for the MTO-process, approx. unloading once a week or once every second week is required. Normal turnaround for tankers is approx. 18-20 hours which includes about 8 hours for unloading. Maximum capacity of the planned MTO process has a total methanol demand of approximately 820 kton per year. One tanker with a loading capacity of 25 000 ton (approximately 16 000 ton per unloading) will be needed for unloading once a week and a loading capacity of 50 000 ton (approximately 32 000 ton per unloading) will be needed for unloading once every second week for continuous transport of methanol from Iggesund to Vattenfall harbour in Stenungsund. Normally there are centrifugal pumps in the tankers. The capacity per pump is normally 200 m\(^3\)/h and there are several pumps per tanker. It is normally the capacity of the line (dimension and pressure drop in line) that limits the unloading capacity (according to discussions with Furetank shipping company).

Safety risks are associated with the unloading operation. It is preferred to have larger tankers and fewer unloading occasions per year compared to have smaller tankers with more frequent occasions for unloading. A methanol flow of 1200 m\(^3\)/h unloaded during a time period of 33 hours once every second week have therefore been assumed based on the data above. This might however be a bit too long unloading time for the boat and might result in extra shipping cost.

### 5.2.2.5 Measures to prevent an explosive atmosphere

Without any measure, the atmosphere above the MeOH/EtOH storage will be explosive (explosive limits according to Table 6) as e.g. the methanol concentration will be approx. 10% in air at the storage temperature of 14°C.
Table 6: Lower and upper explosive limits for methanol and ethanol at room temperature and atmospheric pressure, % in air (acc. to www.mathesontrigas.com).

<table>
<thead>
<tr>
<th></th>
<th>LEL</th>
<th>UEL</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methanol</td>
<td>6.7</td>
<td>36.0</td>
</tr>
<tr>
<td>Ethanol</td>
<td>3.3</td>
<td>19.0</td>
</tr>
</tbody>
</table>

Different solutions to prevent explosive atmosphere to occur have been suggested. The aim of the different solutions is to find an acceptable solution with explosive atmosphere in the caverns or to create MeOH/EtOH concentrations above UEL or below LEL in the space above the liquid level in the storage part of the caverns.

For all solutions described below shall requirements described in SÄIFS 2000:2 be fulfilled.

The most interesting alternatives and some relevant aspects are considered in Table 7 below. For a more detailed description of the different alternatives and aspects see the full report (Johansson, 2014-06-26 revA).

Table 7: Choice of system to prevent explosive atmosphere. Green: good, yellow: intermediate, red: bad.

<table>
<thead>
<tr>
<th>Alternative</th>
<th>Explosive atmosphere in caverns, all equipment and instruments placed above cavern or in existing pump rooms next to the caverns in the mountain</th>
<th>Air flow to gas phase in storage sections of caverns</th>
<th>Inert atmosphere by N₂ flow, all installations placed above cavern</th>
<th>Inert atmosphere by N₂ flow, all installations in existing pump room next to the caverns in the mountain</th>
</tr>
</thead>
<tbody>
<tr>
<td>Risk during operation</td>
<td>Red</td>
<td>Green</td>
<td>Green</td>
<td>Green</td>
</tr>
<tr>
<td>Risk during maintenance of pumps</td>
<td>Red</td>
<td>Yellow</td>
<td>Green</td>
<td>Green</td>
</tr>
<tr>
<td>Complexity</td>
<td>Green</td>
<td>Green</td>
<td>Green</td>
<td>Green</td>
</tr>
<tr>
<td>Cost</td>
<td>Green</td>
<td>Green</td>
<td>Green</td>
<td>Green</td>
</tr>
<tr>
<td>Loss of MeOH/EtOH</td>
<td>Yellow</td>
<td>Yellow</td>
<td>Yellow</td>
<td>Yellow</td>
</tr>
<tr>
<td>Other considerations and risks</td>
<td>-Allowable alternative according to Borealis requirements and safety policy?</td>
<td>-Uneven air flow distribution.</td>
<td>-Introduction of moister to methanol from air?</td>
<td>-Introduction of other unwanted substances to methanol from air?</td>
</tr>
</tbody>
</table>
An interesting alternative is to utilize a floating roof (e.g. by evaporation reducing objects, Hexa Armos Rhombo from AWTT inc. or by fiberglass reinforced polymer) in combination with all alternatives in the table above to minimize the MeOH/EtOH losses.

\( \text{N}_2 \) is in the literature described as a “best practise” method for safe storage and transportation of MeOH/EtOH. By providing a proper level of \( \text{N}_2 \) in the system, methanol can be kept below the minimum \( \text{O}_2 \) concentration (MOC) of < 10%. Air Liquide’s practice is to inert system to half this value, i.e. 5% oxygen in the case of methanol.

To be confident in the safety system of providing \( \text{N}_2 \) into the system it should be considered to monitor the \( \text{O}_2 \) concentration at various locations in the underground caverns. A challenge associated with \( \text{O}_2 \) concentration monitoring is that the liquid level will vary, making it a challenge to install monitors at appropriate locations. It might even be necessary to use monitors that both can be dry and wet to be able to rely on the system. From the table above, it was selected to proceed with alternative “Inert atmosphere by \( \text{N}_2 \) flow, installation in existing pump room”. This alternative is also the alternative that the cost estimate for the underground cavern storage is based on.

### 5.2.2.6 Afterburner system

Even though the chosen alternative (\( \text{N}_2 \) atmosphere in caverns) implies a lower product loss of methanol then when for example utilizing air flow, there will still be a significant working loss during methanol import to the caverns (filling with methanol implies that gas will be pushed out through breathing valves). According to a rough estimate, the loss will approximately be 125 ton/year. This product loss can however be significantly reduced utilizing a floating roof.

An afterburner system has been included in scope to take care of the gas from the breather valves. In the afterburner system the gases will be collected with a process gas fan which is equipped with flame arresting devices with temperature control. The afterburner chamber will be equipped with methanol burners with a duty of 500 kW including burner air fans which also provides the process gas flow with air for incineration of the methanol gases. The incineration occurs at a temperature of 850°C, with a retention time of 1 sec and “dilution” of temperature to approx. 500°C. The system will be equipped with a control system. Further investigations are needed to define the utility supply systems, localisation etc. for this afterburner system. The utility system including fuel system and electricity, localisation (need for roads etc.), instrument signal handling etc. for afterburner system is therefore not included in the cost estimate.

Alternatively, if the import of MeOH is done very frequent, it should be considered to install an off-gas condenser instead of an afterburner system to take care and recycle the methanol off-gases.

### 5.2.2.7 Lined or Unlined solution

In unlined caverns there is a risk for MeOH/EtOH liquid or gas penetration from caverns. However, as long as the MeOH/EtOH is below the ground water level and the pressure thereby is higher than the pressure inside the caverns, the MeOH/EtOH will be kept inside the caverns. The remaining possibility for MeOH/EtOH to escape from the caverns is by diffusion, e.g. diffuse out in the ground water. The diffusivity of methanol in water is approximately \( 1.5 \times 10^{-9} \text{ m}^2/\text{s} \), and the diffusivity of methanol in air is approximately \( 1.6 \times 10^{-10} \text{ m}^2/\text{s} \) at the current conditions, i.e. a significantly higher diffusivity in in gas than liquid. The amount of flow due to diffusivity has to be further investigated.
For lined cavern solution the MeOH/EtOH will also be contained by the lining, i.e. no MeOH/EtOH will diffuse out in the ground water.

As mentioned above, measures have to be taken to prevent that the leak water mix with the methanol due to the high content of alkali. This can be solved with a lined solution. Different solutions have been proposed. One proposal is to have a polymer membrane lining and a drainage carpet against the wall of the cavern, see Figure 7.

Figure 7: Polymer membrane lining alternative (Johansson, 2014-06-26 revA).
Another proposal is to install a steel lining with back filling for drainage, see Figure 8.

Figure 8: Steel plate lining alternative (Johansson, 2014-06-26 revA).
A third proposal is to have partial lining and drainage installed at locations with high in-leakage of water, see Figure 9.

![Diagram of local drainage alternative](image)

**Figure 9: Local drainage alternative (Johansson, 2014-06-26 revA).**

The lined alternatives require a quite smooth rock wall, which can be achieved with levelling and shotcrete. A cost estimate has been done for all of these solutions.
5.2.2.8 Permit plan

Environmental permit
A new permit for the new type of storage is to be enquired at Countries Agency (Länsstyrelsen) according to SFS 2013:251 (according to classification described in SFS 1998:808, Miljöbalken chapter 9). A consultation is to be done with the municipality, Countries Agency and other in the project involved parties. An environmental impact assessment (Miljökonsekvensbeskrivning, MKB) is to be done when the design is settled. The permit process takes about 6-12 month.

An unlined solution to the storage is expected to require an extensive environmental impact assessment containing groundwater modelling that describe flow patterns and contaminant spread in the area, with input and follow-up with data from monitoring wells. Also usage patterns and allowable contamination of the product is expected to be required. The analysis required for the environmental impact assessment for a lined solution is expected to be less extensive.

5.2.2.9 Alternative storage: Conventional Methanol storage tanks

To be able to determine whether storage of methanol in underground caverns is price competitive with conventional methanol storage tanks, a small study was made to determine the cost of building two conventional methanol storage tanks of 50 000 m³ each.

The methanol storage tanks are planned to be installed with inner floating roofs, outer fixed roofs with nitrogen purge to avoid an explosive atmosphere. Storing methanol in conventional storage tanks requires a dike able to contain the entire volume of the tank (excluding the volume that fits within the tank itself). It is also possible to have two storage tanks within the same dike if fulfilling some requirements. According to SÄIFS 2000:2 and changes in SÄIFS 2000:5, the dike should be able to contain 75% of the combined volume of the two tanks excluding the volume taken by tank number 2, resulting in a volume of about 87 000 m³. According to Borealis internal standards, the dike should be able to contain the larger volume of the two tanks, excluding the volume taken by tank number 2. Some design figures for the storage tanks and the tank area are presented below.

Tank dimensions:
- Diameter of about 50 m
- Height of about 25 – 30 m

Dike dimensions:
- Dike area of about 200 * 75 m
- Dike height of about 6 m
  - A high dike has been assumed
    - to minimize the required tank area
    - less safety issues in case of fire and BLEVE
  - Requirements when the dike has a height above 1.8 m (NFPA 30)
    - Provisions shall be made for normal operation of valves and for access to tank roofs without entering below to top of the dike. These provisions shall be permitted to be met through the use of remote-operated valves, elevated walkways, or similar arrangements.
    - Piping passing through dike walls shall be designed to prevent excessive stresses as a result of settlement of fire exposure.
• The minimum distance between tanks and the toe of the interior dike walls shall be 1.5 m.

Other design considerations:

• Space between the tanks according to Borealis internal standards of 50 m (one tank diameter)
• A dike or subdivision between the two tanks is required (NFPA 30)
  o Height of 1 m
  o The dike between the tanks should be able to contain at least 10% of the volume of one tank (5 000 m³)

Included in the scope is also:

• A 12” loading arm at harbour
• Piping from harbour to tank area
  o About 3 km
• 2 redundant pumps including process area
  o 140 m³/h
  o Discharge pressure at about 12 bar(g) to deliver methanol at 10 bar(g)
• 8” Piping from tank area to MTO process, about 2 km
• Fire fighting
• Instrumentation
  o At harbour
  o Tank area
  o Pump area
  o Etc.
• Electrical scope e.g. including a new switchgear and lighting
• Drainage system at new process area and tank area directed to sewage system
• Roads
• Pipe racks
• Etc.

5.3 Impact of co-feeding methanol and ethanol in the MTO-process
Within the forest chemical project, the possibility of co-feeding ethanol in the MTO-process has been studied. The idea is to use the exothermic methanol-to olefin (MTO) reaction to supply heat to the endothermic ethanol dehydration reaction. Preliminary evaluation of the heat balance suggests that about 20mole% ethanol can be added to the methanol feed. One of INEOS ChlorVinyls contributions to the project was to perform lab-scale experiments to study the impact of co-feeding 10 and 20mole% ethanol to a methanol feed on the MTO yields. This has been done by INEOS ChlorVinyls in Porsgrunn using a lab scale fluid-bed reactor with a commercial MTO catalyst at typical MTO conditions.

The basic assumption when analysing the experimental results was that the selectivities for methanol conversion was not affected by co-feeding of ethanol. The experiments performed showed that the ethanol conversion was equal to or higher than the methanol conversion, and the methanol conversion was in the range 95-99.9%. The results show that ethanol is mainly converted to ethylene. The selectivity to ethylene increases 8-17%point compared to pure MeOH feed, but there
is a drop in the selectivity to propylene and butenes so the net effect on the C₂⁻ - C₄⁻ selectivity change is only about 3%point. Mixing in ethanol in the feed also results in increased yields of acetaldehyde and acetic acid. The capacity of the equipment downstream the reactor for handling increased amount of acetic acid and acetaldehyde may restrict the amount of ethanol that can be added to the feed.

If the economics are promising, the impact of co-feeding ethanol on the process should be studied in more details:

- The impact of ethanol on the heat balance accounting for operational constraints
- Impact of increased yields of acetaldehyde and acetic acid on downstream units
- Impact of ethanol addition on the olefin production capacity

Anyhow, enabling admix of ethanol to the methanol should improve the flexibility and operating window.

5.4 The E2E process

Borealis is looking at the possibility to build a 60 kton per year polymer grade ethanol to ethylene (E2E) plant on a former process area. The reason why this area was chosen is that an old plant was closed at the end of 2013 and this area is an excellent industrial estate released for new enterprises. All necessary utilities are accessible on the site such as; steam and natural gas. The ethanol to ethylene process is based on a license from Chematur Engineering.

Chematur Engineering has contributed to the project by delivering a Technology package e.g. containing mass- and energy balances, process flow sheets, equipment data, LCA-data, typical plant layout etc. COWI has also been involved in the E2E part of the Olefins project. They have utilized the information provided by Chematur Engineering and evaluated the constructability to implement an E2E process plant at available area at Borealis site.

5.4.1 Feedstock specifications

Ethanol specifications required for the E2E process can be seen in Table 8 below.

Table 8: Ethanol specifications

<table>
<thead>
<tr>
<th>Item</th>
<th>Specification</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethanol, quality</td>
<td>95vol%</td>
</tr>
<tr>
<td>Acetaldehyde</td>
<td>100ppm (max)</td>
</tr>
<tr>
<td>Fuel oil</td>
<td>100 mg/l (max)</td>
</tr>
<tr>
<td>Acids</td>
<td>7 mg/l (max)</td>
</tr>
<tr>
<td>Methanol</td>
<td>0.3vol% (max)</td>
</tr>
<tr>
<td>Total sulphur</td>
<td>0.5wt ppm (max)</td>
</tr>
</tbody>
</table>
5.4.2 Product specifications (incl. By-products)

ARG Ethylene specifications (specifications for European pipeline) can be seen in Table 9 below.

**Table 9: Ethylene product specification**

<table>
<thead>
<tr>
<th>Item</th>
<th>Specification</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethylene</td>
<td>&gt;99.9 vol %</td>
</tr>
<tr>
<td>Methane + Ethane</td>
<td>&lt;1 000 cm³/m³</td>
</tr>
<tr>
<td>Ethane</td>
<td>&lt;500 cm³/m³</td>
</tr>
<tr>
<td>Acetylene</td>
<td>&lt;5 cm³/m³</td>
</tr>
<tr>
<td>C3 +</td>
<td>&lt;10 cm³/m³</td>
</tr>
<tr>
<td>Carbon monoxide</td>
<td>&lt;2 cm³/m³</td>
</tr>
<tr>
<td>Carbon dioxide</td>
<td>&lt;5 cm³/m³</td>
</tr>
<tr>
<td>Water</td>
<td>&lt;10 cm³/m³</td>
</tr>
<tr>
<td>Oxygen</td>
<td>&lt;5 cm³/m³</td>
</tr>
<tr>
<td>Hydrogen</td>
<td>&lt;10 cm³/m³</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>&lt;100 cm³/m³</td>
</tr>
<tr>
<td>Solvents</td>
<td>&lt;10 cm³/m³</td>
</tr>
<tr>
<td>Total sulphur</td>
<td>&lt;2 mg/kg</td>
</tr>
<tr>
<td>Ammonia</td>
<td>&lt;1 cm³/m³</td>
</tr>
<tr>
<td>Oil</td>
<td>&lt;1 mg/kg</td>
</tr>
</tbody>
</table>

By-product streams from the E2E process includes waste water to be sent to waste water treatment (for more information see section 5.4.6.1), for specifications see Table 10 and Table 11, and two hydrocarbon streams possible to utilize for energy recovery, for specifications see Table 36 and Table 37 in Appendix 10 E2E By-products.

Assuming production of 60 kton ethylene per year (8 000 h/y) assuming Chematur Engineering standard design, 7 321 kg/h waste water with a temperature of 40°C will be generated, Table 10. If the municipality waste water treatment plant has the possibility to treat dirtier water (mainly ethanol), it is possible to exclude equipment from the standard design and save capital investment. It is e.g. possible to exclude a water stripper tower with associate equipment. The waste water will then amount to 6 497 kg/h according to the specification seen in Table 11.

**Table 10: Waste water specifications (Standard design)**

<table>
<thead>
<tr>
<th>Item</th>
<th>Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>99.43wt%</td>
</tr>
<tr>
<td>Ethanol</td>
<td>0.07wt%</td>
</tr>
<tr>
<td>NaOH</td>
<td>0.10wt%</td>
</tr>
<tr>
<td>Na₂CO₃</td>
<td>0.40wt%</td>
</tr>
<tr>
<td>Calculated COD</td>
<td>1449 mg/l</td>
</tr>
<tr>
<td>Calculated BOD</td>
<td>1014 mg/l</td>
</tr>
<tr>
<td>Phenol, oil and metal</td>
<td>0</td>
</tr>
</tbody>
</table>
Table 11: Waste water specifications (no water stripper in design)

<table>
<thead>
<tr>
<th>Item</th>
<th>Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>99.75wt%</td>
</tr>
<tr>
<td>Ethanol</td>
<td>1.52wt%</td>
</tr>
<tr>
<td>NaOH</td>
<td>0.114wt%</td>
</tr>
<tr>
<td>Na2CO3</td>
<td>0.454wt%</td>
</tr>
<tr>
<td>Ethylene</td>
<td>157wtppm</td>
</tr>
<tr>
<td>Acetaldehyde</td>
<td>508wtppm</td>
</tr>
<tr>
<td>Diethyl Ether</td>
<td>954wtppm</td>
</tr>
<tr>
<td>3-Methyl-1-Butanol</td>
<td>123wtppm</td>
</tr>
</tbody>
</table>

There will be recovery of one liquid and one vapour stream from the fuel collection system (typically used as fuel, see also Section 5.4.6.1). Assuming production of 60 kton ethylene per year (8 000 h/y), the recovered liquid will amount to 370 kg/h with a temperature of 40°C and a heating value of 28 MJ/kg. Specifications of the recovered liquid stream can be seen in Table 36 in Appendix 10 E2E By-products.

The recovered vapour stream from the fuel collection system, assuming production of 60 kton ethylene per year (8 000 h/y), will amount to 259 kg/h with a temperature of 52.3°C and a heating value of 42 MJ/kg. Specifications can be seen in Table 37 in Appendix 10 E2E By-products.

5.4.3 Process Description

The plant will produce 60 kton per year ethylene from bio based ethanol. The plant will consume approximate 14 ton ethanol per hour as feedstock into the plant. Ethanol is supplied from offsite storage tankage and vaporized and heated in a gas fired furnace to reaction temperature (about 425-450°C), for a block scheme see Figure 10. The vaporized ethanol is fed to the first bed of a multi bed reactor (typically 4 beds in series) in which about 60% of the ethanol is converted to ethylene. The reaction stops as the gas cools down by about 100°C due to the endothermic nature of the reaction. The effluent from the first bed is then fed back to the gas fired furnace and reheated to reaction temperature and fed to the second bed for an additional 20-30% conversion. The gas cools down and is reheated in the furnace a third time for another 10% conversion. The fourth and final bed is to ensure a conversion of over 99% of the incoming ethanol to ethylene. The effluent from the multi bed reactor containing ethylene, water, unconverted ethanol and by products, such as acetaldehyde, C3’s and C4’s, is fed to the quench column. In the quench column, water and heavier hydrocarbons are removed. The quenched raw ethylene gas is then sent to downstream processing for purification to achieve the desired ethylene quality (crude, intermediate or polymer grade).
5.4.4 Overall material balance
Main feed of the E2E process is ethanol and the main product is ethylene. A feed of about 13.7 t/h ethanol (100wt%) is guaranteed to generate about 7 500 kg/h polymer grade ethylene. In addition to the waste streams mentioned in E2E Product specifications (incl. By-products) there will be some losses in the form of process gases sent to a flare e.g. vent vapors (CO) from the combined condenser between the ethylene column and splitter.

5.4.5 Utility requirements
Utility description and utility balance for the E2E process can be seen in Appendix 11 E2E Utility requirements.

5.4.6 HSE
The plant shall be configured to minimize emissions and discharges during construction phase and operation.

No risk analysis has been performed for the planned production plant in the scope of this study. A general, and also more detailed risk analyses and also HAZOP studies needs to be performed in the coming work. The plant can affect surrounding operations in case of an accident. The design and layout of the plant can be changed during evaluation with the aim to minimize and remove these risk sources.

5.4.6.1 Emissions, effluents, and waste streams

Emissions
Emissions from the production plant and environmental impact are included as a part in the LCA study.

Effluents
The process produces contaminated water which most likely could be sent to the municipal waste water treatment plant (WWTP). Still, the general provisions in ABVA07 (Allmänna bestämmelser för användande av Stenungsunds kommuns allmänna vatten- och avloppsanläggning) must be fullfilled.
Information regarding the content of the water should be provided to the WWTP. It should be investigated whether a buffer tank is needed in order to even out the flow to the plant. The impact in the WWTP of a malfunction in the E2E plant should be investigated as well.

Storm water from the process area should pass through oil separators and applicable treatment system before it is released to the recipient. The systems are available at Borealis plant.

Water and foam used at firefighting should be collected and treated in the existing system at Borealis.

Waste streams
The reactor produces contaminated catalyst media that needs to be taken care of. It is not yet decided if this media should be sent to destruction, taken care of by the supplier or treated in another way.

The process also generates oil and gaseous waste streams. The oily stream will be sent to a dedicated steam boiler at site and the vapour stream will be burnt in the furnace. Propylene is used as cooling media in a closed circuit. Small diffuse leakage could be expected.

Area classification
The area classification of the E2E plant will be according to ATEX standard SS EN 60079-10 and the Swedish interpretation of the standard. The classification will be part of later stages of the project and will take into consideration the existing and surrounding facilities. The overall philosophy should be to minimize the need for classification by concentrating flammable goods at certain areas.

5.4.7 Equipment description
The major equipment included in the E2E processing plant includes:

- One furnace – For heating of ethanol before and between reactors
- Four fixed bed reactors – Conversion of ethanol to ethylene
- One compressor – Increase of product gas pressure to enable separation
- Quench tower – Column to quench the reaction
- Water stripper – Stripping of water (optional – assumed not needed)
- Caustic wash column – Scrubber, removal of CO₂ using caustic
- Gas dryer – Dry the product gas from water
- Ethylene distillation column – Separation/cleaning of ethylene
- Stripper – Separation of ethylene from CO
- Refrigeration system – Enable condensation of ethylene
- Pumps
- Reboilers
- Condensers
- Heat exchangers
5.4.8 Plot plan and Layout
5.4.9 Operation
For the operation of the E2E process it is assumed that there is a need for two operators per shift (5), i.e. 10 operators in total and 2 engineers is also foreseen. Laboratory and management personnel available at the plant are foreseen to be sufficient.

The process plant is designed for minimum 8 000 h/y.

5.5 Storage of Ethanol

5.5.1 Storage and offsite description
At existing cracker plant Ethanol is already used as feedstock in a process. The existing storage capacity is however not sufficient when adding a new Ethanol to Ethylene process. However, the infrastructure for importing ethanol is already in place, loading arm at harbour, pipe between the harbour and existing storage area etc.

5.5.2 Ethanol storage
One idea is to build a new Ethanol storage tank of 10 000 ton at existing tank area utilizing existing tank dike for another storage. The dike has to be able to take the entire volume of the tank in case of leakage, but due to that the risk of having two tank leaks at the same time is so small, the requirement of the dike according to Borealis internal standards is to take the entire larger volume of the two storage tanks or according to SÄIFS 2000:5 to take 75% of the combined volume, for more information see Section 5.2.2.9. Utilizing existing tank dike it is possible to avoid most works associated with ground work and thereby saving capital investment.

A second idea is to convert and utilize existing SCN, pyrolysis gasoline, storage tanks at the cracker plant for ethanol storage. Three SCN tanks are available comprising a total storage capacity of 12 200 m³, which corresponds to approximately 28 days in full production. However, the three referred SCN storage tanks are made of carbon steel. During the past decade there has been evidence of stress corrosion cracking, SCC, of steel storage tanks and associated piping in ethanol service. COWI (have designed Agro Ethanol storage tanks) recommend that a new ethanol storage tank should be constructed out of stainless steel. Stainless steel or carbon steel with lining is the alternatives found on internet. Considering the possible problems using carbon steel in ethanol service it might not be practical to rebuild the existing storage tanks and further investigation is needed, i.e. the alternative of building a new ethanol storage is option number one.

To free the tanks, the storage of SCN has to be moved to an underground cavern, UC-902. This has been investigated by Borealis in a study from 1997 "SCN-lagring i UC-902".

Optimum alternative depend upon difference in investment cost and whether or not to convert UC-902 into SCN service.

A new pipeline is needed for transport of ethanol to the E2E production plant. The pipeline will be of material SS316L, size 3" with an approximate length of 1 800 m.

5.5.3 Ethanol buffer tank
The ethanol is pumped from existing storage area. To be able to fulfil the availability at the E2E plant and to receive a reasonable operational situation, it is recommended to install a new buffer tank for
the ethanol before feeding the process with feedstock. The buffer tank is recommended to be dimensioned for about 2 hours of full production corresponding to a storage volume of 35-40 m$^3$.

6 Business model

6.1 Logistics and raw material management
Methanol is assumed to be imported via ships with a size of 50 000 ton and ethanol is assumed to be imported via ships with a size of 8 000 ton.

6.2 Geographic localisation
Both of the new process plants, MTO and E2E, are planned to be located at Borealis facilities in Stenungsund. Borealis is the only producer of olefins in Sweden. Borealis has a good knowledge in the olefins area. In Stenungsund the market, customers and piping for olefins are available, i.e. infrastructure for olefins production is already in place.

6.3 Economy

6.3.1 Risk evaluation
According to a study performed by the nova Institute (Dammer, et al., 2013 ), the production capacity of bio-based polymers will triple from 3.5 million tonnes in 2011 to nearly 12 million tonnes by 2020. Bio-based drop-in PET and PE/PP polymers are among the polymers showing the highest growth rates on the market. The 3.5 million tonnes represents a share of 1.5% of an overall construction polymer production of 235 million tonnes in 2011. The total production capacity in 2020 is estimated to be 400 million tonnes, resulting in that the bio-based share will increase from 1.5 to a 3% share, i.e. the bio-based share will grow faster than overall production (Dammer, et al., 2013 ).

Findings made in the study performed by the nova Institute (Dammer, et al., 2013 ) is that Europe shows the strongest demand for bio-based polymers, while the production tends to take place elsewhere, namely in Asia and south America, i.e. market for production of bio-based polymers in Europe is foreseen to increase.

The nova Institute (Dammer, et al., 2013 ) have proposed a definition of Green Premium, “Green Premium is the extra price a market actor is willing to pay for the additional performance, and this is precisely the sum of additional emotional performance and strategic performance the buyer gets when choosing the bio-based alternative compared to the price for the (theoretical) conventional counterpart with the same technical performance”. The study performed by the nova Institute using the Green Premium definition clearly prove that Green Premium prices exists and are paid in value-chains of bio-based chemicals and plastics. A relevant group of market actors are already paying Green Premium process throughout examined branches and in variable levels between 10 and 100% (Dammer, et al., 2013 ).

According to the nova Institute (Dammer, et al., 2013 ) the companies do not always succeed to enforce a Green Premium price on their customers, especially in long-term existing supply chains with established products. On the other hand, there are new investments for bio-based plastics (e.g. PE, PET, PP), which only have been developing upon demand, which in turn includes necessarily a confirmed willingness to pay Green Premium prices for intermediates over a longer period. Results
from the study shows that a relevant group of companies are willing to pay around 10 to 20% Green Premium for bio-based intermediates, plastics and polymers. According to the nova Institute is also much higher Green Premium prices (50 to 100% and beyond) possible under specific market conditions under a limited timeframe.

Main reasons of achieving high Green Premium price (Dammer, et al., 2013):

- Huge image benefit (higher brand value) or specific strategic advantages for the producer
- Material costs account for a small share of overall production cost
- Niche markets are very sensitive to emotional performance
- In some instances, Green Premium expenses can be fully and directly passed on along the value-added chain

The following figure shows the results of all expert interviews and surveys performed by the nova Institute, undertaken and analysed in the context of the Green Premium study.
Figure 11: Development of Green Premium prices along the value-added chain of different bio-based chemicals, plastics and end products. Coloured lines represent one value-added chain, single dots represent single findings (Dammer, et al., 2013).

Figure 11 shows identified Green Premium levels depending on where they are paid in the value chain. For example, the polymer producer buys a building block from the chemical company and might pay a Green Premium for it or the end consumer buys the final product and might pay a Green Premium to the distributor. Some identified Green Premium levels are coherent as they belong to the same value chain – they are represented by coloured lines. Other Green Premium levels are single findings (dots) and can therefore not be allocated to specific value chains. Some of the latter represent specific materials and are coloured (e.g. PLA in blue), others represent more general findings and are marked in grey (e.g. bio-based chemicals in general).
According to ICIS Chemical Business (2012), PE e.g. from Braskem commands a price premium of around 15 to 20%, which is feasible for selected target markets and allows for the higher cost of production when compared to petrochemical-based plastics (ICIS Chemical Business, 2012).

According to Energy Technology Systems analysis programme and the International Renewable Energy Agency (IEA-ETSAP and IRENA, 2013), bio-ethylene is typically more expensive than petrochemical ethylene, and producers may be hesitant to invest in this novel production route. To overcome these barriers, producers may set a premium price on their products. In 2007, Braskem determined a premium price for bio-PE of about 15 to 30% compared to petrochemical PE. However, for widespread implementation of bio-ethylene in the long term, its prices need to be comparable to and competitive with petrochemical ethylene since there is no difference in chemical characteristics (IEA-ETSAP and IRENA, 2013).

According to Energy Technology Systems analysis programme and the International Renewable Energy Agency (IEA-ETSAP and IRENA, 2013), the lingo-cellulosic bio-ethylene production route has the potential to become far cheaper than sugar- or starch-based production because 100% of the biomass material can be used. However, there might be some years before lingo-cellulosic production reaches this stage.

According to Jeff Gotro (2013), involved in polymers innovation along with product and process development over the last 30 years, the current bio-PE price is about 50% higher than the fossil-fuel PE, but as production volume increases the premium of bio-PE should decrease (Gotro, 2013).

The Biorenewable Business Platform (2012) could in their business case calculations for the production of bio-ethylene from sugar beet conclude that this route for the production of bio-ethylene only is commercial viable with a substantial higher green premium than 20% (Biorenewable Business Platform, 2012). Green premium, that can be achieved for a limited available volume, for Bio-PE is 30 to 50% over fossil (Biorenewable Business Platform, 2012).

In the study made by the nova Institute (Dammer, et al., 2013) they also looked into policies influencing the development of the bio-based markets. They concluded that several aspects have to be kept in mind in order to be able to support the relatively young sector of bio-based chemicals and plastics. They are of the opinion that the different developments in Asia or the Americas compared to Europe should make policy makers aware of the fact that supporting demonstration and pilot plans, bans, tax incentives and public procurement can be powerful instruments to boost the bio-economy in Europe. European countries need to revise their frameworks in order to catch up on the trends and the secure investments in Europe. In their report (Dammer, et al., 2013) they also conclude that it is important to create a level playing field for bio-based materials compared to bioenergy as well as to the petro-chemical alternatives. The nova Institute are of the opinion that in order to create a fairer competition for biomass between materials and energy, it would be thinkable to modify the support systems for bioenergy in a way so that types of biomass, which can be used for high-value materials applications, are not made overly attractive to be allocated to energetic uses. Alternatively, an interesting solution could be making the support systems for bioenergy available for industrial material uses too (Dammer, et al., 2013). The nova-institute (Carus, et al., 2014) has made a comprehensive proposal to reform the Renewable Energy Directive support framework in the EU (goal is to meet 20% of the EU’s overall energy demand through renewable energy by 2020, along with a minimum of 10% of the transport demand) to a Renewable Energy and Materials Directive.
The reform proposal aims to creating a level playing field for bio-based chemicals and materials with bioenergy and biofuels in Europe. The proposal is based on the insight that the support system for bioenergy and biofuels created by the Renewable Energy Directive and the corresponding national legislations is one of the main reasons hindering the bio-based material sector from developing and therefore hindering the whole bio-based economy (Carus, et al., 2014).

The proposal (Carus, et al., 2014) calls for an opening of the support system to also make bio-based chemicals and materials accountable for the renewables quota of each member state. The basic idea is to transform the RED into a REMD, a Renewable Energy and Materials Directive. The proposal does not aim to establish a new quota for the chemical industry, instead it proposes that the material use of a bio-based building block such as bioethanol should be accounted for in the renewables quota the same way as it counts for the energy use of the same building block, e.g. fuel.

6.3.2 Business case

The business case for the MTO process has been based on the difference between “today” and “to be” when integrating a MTO process with the steam cracker plant. For example, both the raw materials for the cracker (ethane, naphtha, propane and butane) and the MTO process (methanol) will be taken into account using the difference in consumption. This method will also be applied when it comes to operational costs and incomes when selling the products (e.g. ethylene and propylene).

The E2E process is a so called stand-alone process. The business case has therefore been treated as an increase of production capacity, i.e. “today” equals zero and “to be” is the new production capacity. At Borealis site, there exists opportunities to take advantage of spare capacities when it comes to production of PE plastics. The business case for the E2E process can therefore be based on the route going from ethanol to green PE plastics if desired.

Raw material and product price projections have been made within the forest chemical project and have been used as input to the business cases. The price projections have been made using a model (ENPAC) created at Chalmers with two different scenarios from IEA’s conclusions in the World Energy Outlook (WEO) using today’s prices as input:

- Current policy: This scenario reflects the present situation on the international energy and climate arena, taking into account existing CO₂ emission reduction schemes, which on the international level are quite modest.
- 450: This scenario, as described in the IEA World Energy Outlook, reflects a future scenario where climate policy measures are implemented that would lead to a reduction in CO₂ emissions large enough that the global mean temperature with a 50% probability does not exceed pre-industrial levels by more than 2°C. The cost for emitting CO₂ implemented as a price on CO₂ emissions rights in a common EU system, is much higher in the 450 scenario compared to the Current Policy scenario.

According to the Market Scenario report, the ENPAC scenarios represent harmonized, international prices, which do not take into account local conditions (transportation costs, local supply and competition etc.), or national policy instruments (import tolls, taxes, subsidies, etc.). The focus of ENPAC is evaluation of climate change mitigation measures, and hence a carbon tax applied equally over all sectors is the main policy measure included.
The ENPAC tool was not used directly, they rather created their own market scenarios by assuming that the prices of the products in the project would be linked to the world market prices developed with the ENPAC tool.

To be mentioned as well when it comes to some of the products/raw materials studied in the project is that they also can be used as fuel, especially ethanol and methanol, replacing diesel and gasoline. The product/raw material may therefore have two markets with potentially different prices. One on the chemical market, set by the supply and demand situation for that particular chemical, and one on the transportation fuel market, set by the price of the alternative transportation fuel, e.g. diesel and gasoline. Today, the use of biomass based transportation fuels are incentivized through tax exemptions, while the biomass based products for the chemicals markets do not receive the same policy incentives, making it expensive to produce bio materials. The problem was also discussed in above section regarding the RED directive.

Currency of 1.3 $/€ and 0.11€/SEK have been assumed.

### 6.3.2.1 Investment costs

**MTO**
A ±40% cost estimate was made for the Methanol to Olefins process including methanol storage, storage in rock caverns was estimated to be the cheapest alternative. Basis for the cost estimate was the equipment cost as estimated from preliminary data sheets. Equipment based factors was applied for all other costs in the estimate. The investment is estimated to start year 2017, and the plant is estimated to be up and running year 2020. Production at full capacity (270 kton olefins per year) is estimated to start 2021.

- Investment cost ±40% of MSEK 2 864 (MEUR 315)

**E2E**
A ±40% cost estimate was made for the Ethanol to Ethylene process including ethanol storage. Basis for the cost estimate was the equipment cost as estimated from preliminary data sheets as well as some package units as priced by Chematur. Equipment based factors was applied for all other costs in the estimate. The investment is estimated to start year 2015, and the plant is estimated to be up and running year 2017. Production at full capacity (60 kton ethylene per year) is estimated to start 2018.

- Investment cost ±40% of MSEK 636 (MEUR 70)

### 6.3.2.2 Raw material and product prices

Link to the world market prices:

- Methanol: Initial price based on information from the industry. Follows the gasoline price index
- Ethanol: Initial price based on information from the industry. Follows the gasoline price index
- Ethylene: Initial price based on information from the industry. Follows a mix of the crude oil and natural gas price indices
- Propylene: Initial price based on information from the industry. Follows a mix of the crude oil and natural gas price indices
MTO

In below figure, Figure 12, are the main raw material and product prices, as chemical product or feedstock, for the different projections shown.

![Graph showing price projections for MTO raw materials/products](image)

**Figure 12: Main raw materials/products for MTO, price projections using ENPAC model.**

The methanol price has fluctuated between 146.5 EUR/ton to 525 EUR/ton between 2006 and 2014, making it really difficult to predict future methanol prices, see Figure 13 below.
Figure 13: Historical Methanol contract prices EUR/ton, Methanex.

E2E
The main raw material and product prices, as a chemical product of feedstock, from the projections can be seen in Figure 14 below.
6.3.2.3 Operational costs

For both business cases, a maintenance cost per year corresponding to 2% of the total investment cost assumed. A personnel cost of 1 049 000 SEK (150 000 $) per year and personnel has been assumed.

MTO

For the MTO process integrated with the cracker plant has the difference in operational cost been calculated, i.e. difference between “today” and “to be” when it comes to e.g. utility consumption such as steam, electricity, catalyst material and nitrogen. Standard design for the MTO process is to run a raw gas compressor and an air blower using steam turbines. Due to that electrical motors are cheaper, has the electrical driven alternative for the compressor and air blower also been included.

Due to that the MTO process is planned to be integrated with the cracker process, the most profitable operational case of the MTO process is highly dependent upon methanol price in relation to conventional cracker feeds and the possibilities to charge a green premium price. Turndown ratio of 90% (90% of total production capacity) for the MTO process has been used in the business case.

E2E

As mentioned, the implementation of the E2E process will be an increase of ethylene production and thereby is the operational cost only dependent on the cost of utilities, raw material, personnel etc. to the E2E process.
6.3.2.4 Income

MTO
See Figure 12 above for main raw material and product prices. The income is also to a great extent dependent upon the possibilities of charging a premium for green products, the so called green premium price, which is assumed to be possible. See section Result for a complete picture.

E2E
For the E2E process the income will be in the form of sold green ethylene with the potential to produce green PE.

6.3.3 Result

MTO
As the business cases for the MTO process has been based on “today” (only conventional cracking) and “to be” (conventional cracking combined with the MTO process) the business cases are influenced by the projected prices for the conventional feeds. The Naphtha price is e.g. projected using the crude oil price and the propane and butane prices is projected using both the crude oil and the natural gas prices for the two different cases. The ethane price will not affect, as the same amount of ethane is assumed to be cracked in both the “today” and the “to be” cases. Affecting the profitability is also the assumed green premium price. In the base case it was assumed that it is possible to charge a green premium of 30% during the entire lifetime of the project even though discussions in above section indicates that the possibility of charging a green premium most likely will decrease in coming years. Below, some assumptions made are presented as well as the result of the business case.

Assumptions:

- Investment cost ±40% of MSEK 2 864 (MEUR 316)
- Investment starting year 2017
- Lifetime of 20 years
- Discount rate of 10%
- Green premium of 30% for bio-ethylene, bio-propylene and bio-butylene
- Two forecasted price scenarios according to above Section, Current policies and 450

Results:

<table>
<thead>
<tr>
<th>Variable</th>
<th>Price scenario</th>
<th>Current policies</th>
<th>450</th>
</tr>
</thead>
<tbody>
<tr>
<td>NPV</td>
<td></td>
<td>3 413 MSEK</td>
<td>2 331 MSEK</td>
</tr>
<tr>
<td>IRR</td>
<td></td>
<td>24%</td>
<td>21%</td>
</tr>
<tr>
<td>Discounted PBP</td>
<td></td>
<td>8.1 years</td>
<td>8.6 years</td>
</tr>
</tbody>
</table>

The above results are judge as good though an IRR of 20% is stated as sufficient.

Below in Figure 15 and Figure 16 is the cumulated discounted cash flow presented for the MTO process integrated with the cracker process for the Current policy case and the 450 case.
Figure 15: Cumulated discounted cash flow for the MTO process, including 30% green premium on olefins, Current policy scenario.

Figure 16: Cumulated discounted cash flow for the MTO process, including 30% green premium on olefins, 450 case.

In Table 13, the revenues and costs for the first year of full production for the MTO process integrated with the cracker are presented. To notice is the importance of a green premium for the olefins.
Table 13: Delta production cost for the MTO process integrated with the cracker the first year of full operation (270 kton olefins per year).

<table>
<thead>
<tr>
<th>Revenues/Costs</th>
<th>First year of full capacity, Current policy scenario</th>
<th>First year of full capacity, 450 policy scenario</th>
</tr>
</thead>
<tbody>
<tr>
<td>Revenues</td>
<td>MSEK/y</td>
<td>MSEK/y</td>
</tr>
<tr>
<td>Delta income for products</td>
<td>768</td>
<td>733</td>
</tr>
<tr>
<td><strong>Operational Costs</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Delta cost raw materials</td>
<td>-412.5</td>
<td>-399</td>
</tr>
<tr>
<td>Delta cost utilities</td>
<td>-245</td>
<td>-228</td>
</tr>
<tr>
<td><strong>Fixed costs</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Maintenance</td>
<td>-42</td>
<td>-42</td>
</tr>
<tr>
<td>Personnel</td>
<td>-8.4</td>
<td>-8.4</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>60.1</td>
<td>55.6</td>
</tr>
<tr>
<td>Green premium (30% for green olefins)</td>
<td>866</td>
<td>823</td>
</tr>
<tr>
<td><strong>Total including green premium</strong></td>
<td>926.1</td>
<td>878.6</td>
</tr>
</tbody>
</table>

Sensitivity analysis MTO

Due to the uncertainty of the assumed variables, some sensitivity analyses has been made to display whether or not the project can be seen as a risky project or not. Below is the analysed variables presented.

- **Investment cost**
  - Project is in an early phase with an accuracy of the estimated investment cost of ±40%

- **Possibility to charge a green premium price**
  - Market driven and therefore difficult to predict the willingness of the customer to pay extra for bio-ethylene

- **Green premium for bio-methanol**
  - Forecasted methanol prices are for fossil methanol and bio-methanol producers will most likely charge a green premium as well.

Below, the sensitivity analysis for the investment cost versus green premium is presented, Current policies in Table 14 and 450 in Table 15. The estimated investment cost is stated to be a ±40% accuracy cost estimate with equal likelihood to overrun as underrun. An estimate of ±40% for MSEK 2 864, corresponds to an investment interval of MSEK 1 718 to MSEK 4 010.
Table 14: Sensitivity analysis: Investment cost versus green premium price scenario current policies MTO.

<table>
<thead>
<tr>
<th>Green premium %</th>
<th>0.00%</th>
<th>10.00%</th>
<th>20.00%</th>
<th>30.00%</th>
<th>40.00%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Investment cost</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>40%</td>
<td>-9%</td>
<td>5%</td>
<td>12%</td>
<td>17%</td>
<td>22%</td>
</tr>
<tr>
<td>20%</td>
<td>-7%</td>
<td>7%</td>
<td>14%</td>
<td>20%</td>
<td>25%</td>
</tr>
<tr>
<td>0%</td>
<td>-5%</td>
<td>9%</td>
<td>17%</td>
<td>24%</td>
<td>29%</td>
</tr>
<tr>
<td>-20%</td>
<td>-3%</td>
<td>12%</td>
<td>21%</td>
<td>28%</td>
<td>34%</td>
</tr>
<tr>
<td>-40%</td>
<td>0%</td>
<td>16%</td>
<td>27%</td>
<td>35%</td>
<td>42%</td>
</tr>
<tr>
<td>MSEK 2 864</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 15: Sensitivity analysis: Investment cost versus green premium price scenario 450 MTO.

<table>
<thead>
<tr>
<th>Green premium %</th>
<th>0.00%</th>
<th>10.00%</th>
<th>20.00%</th>
<th>30.00%</th>
<th>40.00%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Investment cost</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>40%</td>
<td>#</td>
<td>0%</td>
<td>9%</td>
<td>15%</td>
<td>20%</td>
</tr>
<tr>
<td>20%</td>
<td>#</td>
<td>2%</td>
<td>11%</td>
<td>17%</td>
<td>23%</td>
</tr>
<tr>
<td>0%</td>
<td>#</td>
<td>5%</td>
<td>14%</td>
<td>21%</td>
<td>27%</td>
</tr>
<tr>
<td>-20%</td>
<td>#</td>
<td>8%</td>
<td>18%</td>
<td>25%</td>
<td>32%</td>
</tr>
<tr>
<td>-40%</td>
<td>#</td>
<td>12%</td>
<td>24%</td>
<td>32%</td>
<td>40%</td>
</tr>
<tr>
<td>MSEK 2 864</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

A comparison between the two price scenarios (2: Current policies and 3: 450) based on the green premium is presented below for the NPV, IRR and PBP in Table 16 to Table 18. To be notice is that there are some differences in outcome between the two scenarios, i.e. depending on future political decisions, the project is more or less risky. To be notice is also the importance of the green premium.

Table 16: NPV Sensitivity analysis: Price scenario versus green premium (2: Current policies, 3: 450) MTO.

<table>
<thead>
<tr>
<th>Price scenario</th>
<th>NPV [MSEK]</th>
</tr>
</thead>
<tbody>
<tr>
<td>2</td>
<td>-2028^,,,214,,,160,,,3413,,,5227</td>
</tr>
<tr>
<td>3</td>
<td>-2393,,,818,,,756,,,2331,,,3905</td>
</tr>
</tbody>
</table>

Table 17: IRR Sensitivity analysis: Price scenario versus green premium (2: Current policies, 3: 450) MTO.

<table>
<thead>
<tr>
<th>Price scenario</th>
<th>IRR</th>
</tr>
</thead>
<tbody>
<tr>
<td>2</td>
<td>-5%</td>
</tr>
<tr>
<td>3</td>
<td>#</td>
</tr>
</tbody>
</table>
Due to the difficulty to predict future prices and due to the fact that bio-methanol most likely will be bought with a green premium, a sensitivity analysis was made on the two different price scenarios versus a green methanol premium assuming a 30% green premium on the olefins, seen in Table 19.

To notice from these results is the importance of predicting future bio-methanol prices.

Table 19: IRR Sensitivity analysis: Price scenario versus green methanol premium, assuming 30% green premium for bio-olefins (2: Current policies, 3: 450) MTO.

<table>
<thead>
<tr>
<th>IRR</th>
<th>Green premium % on methanol (assuming 30% green premium on products)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0%</td>
</tr>
<tr>
<td>Price scenario 2</td>
<td>0%</td>
</tr>
<tr>
<td>Price scenario 3</td>
<td>0%</td>
</tr>
</tbody>
</table>

E2E

In comparison to the MTO business case, the E2E business case implies an increase in production capacity. The profitability of the different scenarios is therefore only dependent upon the prices of the raw materials and products in the E2E process. The product price is to a great extent affected by implementation of the Green premium price. In the base case it was assumed that it is possible to charge a green premium of 30% during the entire lifetime of the project even though discussions in above section indicates that the possibility of charging a green premium most likely will decrease in coming years. Below, some assumptions made are presented as well as the result of the business case.

Assumptions:

- Investment cost ±40% of MSEK 636 (MEUR 70)
- Investment starting year 2015
- Lifetime of 20 years
- Discount rate of 10%
- Green premium of 30% for green ethylene
- Two forecasted price scenarios according to above Section, Current policies and 450

Results:

Table 20: E2E process business case results (green premium for ethylene of 30%).

<table>
<thead>
<tr>
<th>Variable \ Price scenario</th>
<th>Current policies</th>
<th>450</th>
</tr>
</thead>
<tbody>
<tr>
<td>NPV</td>
<td>1 239 MSEK</td>
<td>1 020 MSEK</td>
</tr>
<tr>
<td>IRR</td>
<td>29%</td>
<td>28%</td>
</tr>
<tr>
<td>Discounted PBP</td>
<td>6.6 years</td>
<td>6.7 years</td>
</tr>
</tbody>
</table>
The above results are judged as good though an IRR of 20% is stated as sufficient.

Below in Figure 17 and Figure 18 is the discounted cumulative cash flow presented for the E2E process including 30% green premium for ethylene for the current policies case respectively the 450 case.

![Figure 17: Discounted cumulative cash flow for the E2E process, 30% green premium for ethylene, Current policies case.](image1)

![Figure 18: Discounted cumulative cash flow for the E2E process, 30% green premium for ethylene, 450 policies case.](image2)

In Table 21 the revenues and costs for the first year of full production for the E2E process are presented. To notice is the importance of a green premium for ethylene.
Table 21: Delta production cost for the E2E process the first year of full operation (60 kton ethylene per year).

<table>
<thead>
<tr>
<th>Revenues/Costs</th>
<th>First year of full capacity, Current policy scenario</th>
<th>First year of full capacity, 450 policy scenario</th>
</tr>
</thead>
<tbody>
<tr>
<td>Revenues</td>
<td>MSEK/y</td>
<td>MSEK/y</td>
</tr>
<tr>
<td>Ethylene sales</td>
<td>686.5</td>
<td>674</td>
</tr>
<tr>
<td>Operational Costs</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Ethanol cost</td>
<td>-546</td>
<td>-535.5</td>
</tr>
<tr>
<td>Utilities cost</td>
<td>-46</td>
<td>-46.7</td>
</tr>
<tr>
<td>Fixed costs</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Maintenance</td>
<td>-12.6</td>
<td>-12.6</td>
</tr>
<tr>
<td>Personnel</td>
<td>-12.4</td>
<td>-12.4</td>
</tr>
<tr>
<td>Total</td>
<td>69.5</td>
<td>66.7</td>
</tr>
<tr>
<td>Green premium (30% for green ethylene)</td>
<td>206</td>
<td>202</td>
</tr>
<tr>
<td>Total including green premium</td>
<td>275.5</td>
<td>268.7</td>
</tr>
</tbody>
</table>

Sensitivity analysis E2E

Due to the uncertainty of the assumed variables, some sensitivity analyses has been made to display whether or not the project can be seen as a risky project or not. Below the analysed variables are presented.

- Investment cost
  - Project is in an early phase with an accuracy of the estimated investment cost of ±40%
- Possibility to charge a green premium price
  - Market driven and therefore difficult to predict the willingness of the customer to pay extra for bio-ethylene
- Fossil ethylene price
  - Difficult to predict future prices
- Polyethylene produced from the ethylene
  - The possibility to produce polyethylene exists and are going to have an effect on the result

No green premium price for ethanol has been assumed due to the fact that the ethanol on the market already is green today. There might be a risk however that ethanol produced via fermentation of forest residues can be more expensive than ethanol produced from e.g. sugar cane due to the complexity of the process. This has however to be concluded in the project.

Below is the sensitivity analysis for the investment cost versus green premium presented, Current policies in Table 22 and 450 in Table 23. The estimated investment cost is stated to be a ±40% accuracy cost estimate with equal likelihood to overrun as underrun. An estimated of ±40% for MSEK 636, corresponds to an investment interval of MSEK 891 to MSEK 382.
Table 22: Sensitivity analysis: Investment cost versus green premium price scenario current policies E2E.

<table>
<thead>
<tr>
<th>Green premium %</th>
<th>0,00%</th>
<th>10,00%</th>
<th>20,00%</th>
<th>30,00%</th>
<th>40,00%</th>
</tr>
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<tr>
<td>Investment cost</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>40%</td>
<td>3%</td>
<td>11%</td>
<td>17%</td>
<td>22%</td>
<td>27%</td>
</tr>
<tr>
<td>20%</td>
<td>5%</td>
<td>13%</td>
<td>20%</td>
<td>25%</td>
<td>30%</td>
</tr>
<tr>
<td>MSEK 636</td>
<td>0%</td>
<td>7%</td>
<td>16%</td>
<td>23%</td>
<td>29%</td>
</tr>
<tr>
<td>-20%</td>
<td>10%</td>
<td>19%</td>
<td>27%</td>
<td>34%</td>
<td>41%</td>
</tr>
<tr>
<td>-40%</td>
<td>13%</td>
<td>24%</td>
<td>34%</td>
<td>42%</td>
<td>49%</td>
</tr>
</tbody>
</table>

Table 23: IRR Sensitivity analysis: Investment cost versus green premium price scenario 450 E2E.

<table>
<thead>
<tr>
<th>Green premium %</th>
<th>0,00%</th>
<th>10,00%</th>
<th>20,00%</th>
<th>30,00%</th>
<th>40,00%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Investment cost</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>40%</td>
<td>0%</td>
<td>9%</td>
<td>15%</td>
<td>21%</td>
<td>25%</td>
</tr>
<tr>
<td>20%</td>
<td>2%</td>
<td>11%</td>
<td>18%</td>
<td>24%</td>
<td>29%</td>
</tr>
<tr>
<td>MSEK 636</td>
<td>0%</td>
<td>4%</td>
<td>14%</td>
<td>21%</td>
<td>28%</td>
</tr>
<tr>
<td>-20%</td>
<td>7%</td>
<td>18%</td>
<td>26%</td>
<td>33%</td>
<td>39%</td>
</tr>
<tr>
<td>-40%</td>
<td>11%</td>
<td>23%</td>
<td>32%</td>
<td>40%</td>
<td>48%</td>
</tr>
</tbody>
</table>

A comparison between the two price scenarios (2: Current policies and 3: 450) based on the green premium is presented below for the NPV, IRR and PBP in Table 24 to Table 26. To be notice is that there are some differences in outcome between the two scenarios, i.e. depending on future political decisions, the project is more or less risky. To be notice is also the importance of the green premium.

Table 24: NPV Sensitivity analysis: Price scenario versus green premium (2: Current policies, 3: 450) E2E.

<table>
<thead>
<tr>
<th>Price scenario</th>
<th>2</th>
<th>-133</th>
<th>324</th>
<th>782</th>
<th>1239</th>
<th>1696</th>
</tr>
</thead>
<tbody>
<tr>
<td>3</td>
<td>-220</td>
<td>193</td>
<td>607</td>
<td>1020</td>
<td>1433</td>
<td></td>
</tr>
</tbody>
</table>

Table 25: IRR Sensitivity analysis: Price scenario versus green premium (2: Current policies, 3: 450) E2E.

<table>
<thead>
<tr>
<th>Price scenario</th>
<th>2</th>
<th></th>
<th>7%</th>
<th>16%</th>
<th>23%</th>
<th>29%</th>
<th>35%</th>
</tr>
</thead>
<tbody>
<tr>
<td>3</td>
<td>4%</td>
<td>14%</td>
<td>21%</td>
<td>28%</td>
<td>33%</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table 26: PBP Sensitivity analysis: Price scenario versus green premium (2: Current policies, 3: 450) E2E.

<table>
<thead>
<tr>
<th>Price scenario</th>
<th>2</th>
<th>11,8</th>
<th>8,1</th>
<th>6,6</th>
<th>5,7</th>
</tr>
</thead>
<tbody>
<tr>
<td>3</td>
<td>12,8</td>
<td>8,3</td>
<td>6,7</td>
<td>5,8</td>
<td></td>
</tr>
</tbody>
</table>
Due to the difficulty of predicting future prices and the fact that a discount for the ethylene most often is applied, a sensitivity analysis was also made for the fossil ethylene price, see Table 27. In the sensitivity analysis it is still assumed that a 30% green premium can be charged. To be noticed is that a decrease of the ethylene price down to below 90% of the forecasted contract price results in an IRR of less than the “required” 20%, indicating the importance of good future price predictions. The same discussion is applicable for ethanol pricing, especially since only minor amounts of forest based/2nd generation ethanol is out on the market today.

Table 27: IRR Sensitivity analysis: Price scenario versus fossil ethylene price, assuming 30% green premium for bio-ethylene (2: Current policies, 3: 450). 100%: Forecasted contract price E2E.

<table>
<thead>
<tr>
<th>Price scenario</th>
<th>IRR</th>
<th>Fossil ethylene price (30% green premium for bioethylene)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>80%</td>
<td>85%</td>
</tr>
<tr>
<td>2</td>
<td>11%</td>
<td>16%</td>
</tr>
<tr>
<td>3</td>
<td>9%</td>
<td>14%</td>
</tr>
</tbody>
</table>

### Discussion and Conclusion

Having most effect on the outcome of the business case is the marginal between the feedstock and the raw material. It is believed to be a greater challenge to affect the cost of the feedstock (ethanol and methanol) than the price of the products (bio olefins and bio PE). It is possible to influence the price of the product through the green premium and thereby increasing the marginal between feedstock and product. Seen in the results above is that the processes only are profitable when charging a rather high green premium. However, the customers have to be willing to pay extra for the product, which is chemically identical to the fossil alternative. Another possibility is if an incentive for biomaterials from e.g. EU is implemented. However, as also discussed above, there is a need for a more thorough analysis of the methanol and ethanol market to get a better picture of what prices are realistic.

Cost of fossil methanol, as seen in Figure 13, has fluctuated a lot during the last couple of years. This fact makes it really hard to predict, or to assume, a correct enough price as base in the business case. Another dilemma regarding the methanol price is the fact that it is for fossil methanol. It is more than likely that bio methanol producers have to charge a green premium to make the business profitable. As seen in Table 19, it is hardly profitable with an MTO process with as low green premium for methanol as 10%, assuming a green premium for olefins of 30% and the predicted methanol prices. Also in discussion with the gasification and methanol synthesis project, a 10% green premium for methanol on the predicted methanol prices is too low for their projects to be profitable. Again, for the bio economy to grow there is need for incentives from EU or the Swedish government.

To remember is that the above results exclude toll, local conditions, and the competition for the feedstock as fuel and chemical. Today there is a toll for importing ethanol into EU, making it more expensive than in above result to produce bio chemicals.

An important question regarding this is whether or not there should be any import toll on ethanol for the production of bio chemicals. What should be remembered is that most of the bio chemicals, or at least bio plastics today on the market, also on the EU market, are produced from ethanol outside EU.
and then imported into EU without any toll. To notice also regarding this discussion is that ethanol produced inside EU most often are more expensive than the ethanol produced outside (excluding import toll). A bio-plastics (PE) producer inside EU has therefore difficulties to compete with the producers outside EU today, affecting the bio economics inside EU.

The competition between the feedstock (ethanol and methanol) as fuel and chemical is especially important in the 450 scenario. In the 450 case it will be really expensive to utilize fossil feedstock for transportation purposes due to the increased CO₂ emission taxes. Even though it most likely will be more expensive to utilize fossil resources as fuel, it is likely that it will be cheaper to utilize them as chemical feedstock due to the reduced total demand. The change in fossil feedstock cost will most probably also affect the price of the products (ethylene and propylene), and will probably be cheaper. Due to the increased cost of fossil fuels, the demand for bio fuels will probably be greater pushing the price for bio fuels upward. However, due to that ethanol and methanol as a chemical feedstock are linked to the gasoline price excluding the CO₂ emission taxes, the ethanol and methanol chemical prices are projected to decrease in the 450 scenario. However, since the fossil fuel prices (including CO₂ emission taxes) are projected to increase, the willingness to pay for and demand for bio fuels most likely also will increase. Due to the previous discussion, it is believed that the price for ethanol and methanol as a chemical also will increase and not be according to the projections in the 450 scenario. Why would an ethanol or methanol producer sell their products to the chemical market for a lower price than they can get for it as a fuel? For the bio economy to grow, the feedstock as a chemical and fuel has to be able to compete on the same conditions, either take away the incentives for bio fuels, or introduce incentives for bio materials, as also is discussed in above section regarding the RED/REMD.

From the conducted project it is possible to conclude:

- The need for a green premium price for the project to be profitable and for the biomaterials to be competitive with fossil alternatives
- The need for analyses of the feedstock market
- The need for a complete value chain project, forest material to bio plastics consumers
- The need for further technological studies to verify/modify investment cost
- The need for incentives from the EU or Swedish government to decrease the competition for feedstock between fuel and chemical
- The need for toll exemption for chemical feedstock, to be able to compete within the EU with producers outside EU

A fact not forget is that the biomaterials/bio polymers market are growing, and are expected to grow more and more in the future. So, it is not only the profitability of the process it selves that should be consider, but also the cost of not carry on with the project, and thereby lose costumers and market share.

8 Way forward
The Ethanol to Ethylene process is believed to be the best alternative, at least in the near future. The way forward is to perform a screen analysis of available Ethanol to Ethylene processes, and study the process believed to have the highest potential in more detail, resulting in a more precise cost
estimate. The study also aims to evaluate possible integration alternatives onsite to conclude the most optimized solution when it e.g. comes to energy consumption and resource consumption.

Due to the inexperience of the bio chemicals/materials market, the next phase of the project should also include departments/partners working explicit with the bio market, both when it comes to feedstock and products. What feedstock prices are possible and how much green premium are the customers willing to pay?

The possibilities/opportunities of receiving investment support from e.g. the EU or Swedish government should be investigated in more detail to get the project(s) profitable, or as profitable as possible,

It should also be investigated whether or not it is possible, today and in the future, to buy chemical ethanol outside the EU with toll exemption.
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11 Appendices

11.1 Appendix 1 MTO reactor/regenerator description

11.1.1 MTO Reactor and Regeneration scheme

When converting methanol into light olefins in the MTO reaction it is beneficial to use a circulating fluidized bed with catalyst regeneration, as shown in Figure 19. A fluidized bed reactor is suitable since this technology normally provides the best opportunities for catalyst circulation to a separate regenerator where the coke can be combusted. Furthermore, since both the MTO process and the combustion of coke are exothermic, this reactor type is suitable because it enables removal of heat that is generated (Vora, et al., 2001). Also at the demonstration plant in Norway they utilize a fluidized-bed reactor with a fluidized-bed regenerator due to the high heat of reaction and the frequent regeneration of catalyst (Hamblett, 1996), and it is also the technique UOP license.

![Figure 19: MTO reactor, circulating fluidized bed, B, with catalyst cooling, C, and catalyst regeneration, A, (Miller, 2011).](image)

Methanol is the preferred feedstock for the MTO process. The feed is introduced into the reactor and to the catalyst via a fluidized feed stream. The products of the MTO reaction depend upon the feed stream, catalyst and conditions within the reactor. The products will however preferably contain light olefins, ethylene and propylene.

Operating conditions of the reactor will influence the ratio of ethylene/propylene and coke formation. An increased pressure may favour the production of higher propylene ratios. Also the temperature will influence the ratio of ethylene and propylene, and a higher temperature gives a higher ratio. However, a higher temperature will also reduce the selectivity towards light olefins (Vora, et al., 2001). Temperature of the reaction is preferably between 400 and 550°C. To maximize the production of ethylene a temperature between 475 and 550°C, or more preferably between 500 and 520°C may be used. To maximize the production of propylene a temperature between 350 and 475°C, or more preferably between 400 and 430°C may be used (Miller, 2011). The pressure of the reactor is between 1 and 3 barg (Eng, et al., 1998). To maintain the selectivity towards light olefins, a...
non-reactive diluent may be used in the feed stream. The feed stream may also contain water, and may consist of 0 to 35 wt% water (Miller, 2011). It is possible to use crude methanol as feed, which nominally contains 20 wt% of water (Eng, et al., 1998).

The reaction of converting methanol into olefins is exothermic. Excess heat from the reaction will disrupt the optimal conditions of the reaction by raising the temperature to levels that are too high. Therefore it is of utmost importance to control the temperature within the reactor to be able to optimally convert the methanol into olefins and minimize the formation of by-products. The temperature can be controlled by removing heat from the reactor during the process. One way of doing this is to cool the catalyst used in the conversion reaction. Another way of controlling the heat of the reaction is to adjust the temperature of the feed stream. However, adjusting the temperature of the feed stream will not quickly change the temperature of the reaction since the catalyst within the reactor is very massive and voluminous. The catalysts will not respond to changes in temperature of the feed stream (Miller, 2011).

When the catalyst is exposed to the reacting species coke, carbonaceous material, is generated and deposited on the catalyst. Accumulation of coke on the catalyst will interfere with the catalyst ability to convert the reactants into products, and will result in a spent catalyst. As the coke deposit increases, the catalyst loses its activity and less of the feed will be converted into the desired olefins (Miller, 2011). This is also in accordance with Soundararajan, et al. (2001). The methanol conversion decreases significantly with an increase of coke on the catalyst and the coke formed during reaction causes partial or complete pore plugging, which reduces the active sites available for the reagent molecules. Coke having deactivating effect is classified as inactive coke (Soundararajan, et al., 2001). Regeneration of the catalyst, e.g. via coke combustion in air, is therefore necessary. Formation of coke is dependent on temperature and water content in the feed, and will be higher with higher temperatures and lower with water content. However, the coke does not just have deactivating effects, active coke also increases the ethylene selectivity (Vora, et al., 2001; Soundararajan, et al., 2001). An optimum catalyst operation, depending on wanted products, is found at a certain equilibrium level of coke (Vora, et al., 2001). This phenomena arises because the deposited coke increases the shape selectivity of the SAPO-34 catalyst by inducing steric hindrance (Soundararajan, et al., 2001). According to Kuechler, et al. (2005) it is important to design the reactor so that a relatively high average level of coke is maintained within the reactor, and a preferred average level of coke range from about 2 to about 20 wt%. A high average level of coke can be maintained by just regenerating a part of the catalyst (Kuechler, et al., 2005). The step of regenerating the catalyst by removing the coke via combustion with oxygen will restore the catalytic activity of the catalyst (Miller, 2011). To increase the selectivity to light olefins it is preferred to have an average level of coke between 2 and 7 wt% based on the overall weight of the catalyst (molecular sieve plus matrix and binder) (Cao, et al., 2011). Therefore only a fraction of the catalyst material will be regenerated and the remaining fraction will be cooled (Miller, 2011).

The reactor utilized in the MTO reaction, part B in Figure 19, consists of an upper disengaging chamber and a lower reaction chamber (Miller, 2011). As the gas, unreacted feedstock and products, travels through the reactor chamber to the disengaging chamber, it carries partially coked catalyst particles with it. When the mixture of catalyst particles and reaction product reaches the top of the reactor chamber, distributor arms discharge the catalyst particles and the reaction product to the bottom of the disengaging chamber by gravitational force. Catalyst particles that remain in the
reaction product stream continue upwardly to phase separators such as cyclones, which separates the catalyst from the product vapour. The product vapour is transferred to the product recovery section (Miller, 2011).

The catalyst separated from the product vapour via cyclones is transferred to the bottom of the disengaging chamber. A portion of the catalyst settled inside the disengaging chamber will be directed to a catalyst cooler, seen in part C of Figure 19, and another portion will be directed to the regenerator, seen in part A of Figure 19. The portion directed to the catalyst cooler is heat exchanged in a flow-through type of catalyst cooler. At least two catalyst cooler is preferred (Miller, 2011). The use of cooling tubes makes it possible to recover and remove excess heat from the catalysts and produce steam that can be used elsewhere. To control the amount of catalyst exiting from the reactor and thus entering the catalyst cooler a recirculation valve is utilized, and it is therefore possible to control the temperature in the reactor (Williams & Vaughn, 2003).

The catalyst material transferred to the regenerator, part A in Figure 19, is heated in the presence of oxygen to a temperature between 550 and 780˚C (Clem, et al., 2006). At these temperatures the coke deposited at the catalyst will be combusted. To achieve complete regeneration of the catalyst particles, and achieve complete combustion, a temperature inside the reactor of 730˚C is required (Sadeghbeigi, 2011). The catalyst material and the regeneration gas, often air, flows upward from the lower combustion zone towards the upper combustion zone. As the mixture of catalyst and gas flows through the regenerator and the coke is burned off, the heat generated is absorbed by the relatively carbon free catalyst. To control the temperature inside the regenerator and to return the catalyst to the reactor at a suitable temperature, a back-mixed type of catalyst cooler is used, number 102 in Figure 19 (Cetinkaya & Myers, 1986). The average level of coke on the regenerated catalyst particles should be less than 0.2wt% of the molecular sieve in the catalyst particle (Clem, et al., 2006). The temperature of the regenerator can be controlled by withdrawing regenerated catalyst and passing it through the catalyst cooler to form a cooled regenerated catalyst composition. The cooled catalyst material transferred back to the regenerator will be mixed with catalyst material containing coke and work as an inert material absorbing heat released during the combustion (Cao, et al., 2011). The regenerated catalyst material will be transferred through the catalyst cooler, and thereby it is possible to produce steam (Cetinkaya & Myers, 1986). A portion of the catalyst material transferred through the catalyst cooler will be returned to the reactor and another portion will be returned to the regenerator.

To increase the heat transfer between catalyst material and cooling tubes, a fluidizing gas is transported through the catalyst coolers. An inert gas such as nitrogen, steam or a hydrocarbon gas is used and the flow rate of the fluidizing gas is sufficiently high to accomplish fluidization. There are two ways to control the temperature of the recirculated catalyst material, one can either control the amount of catalyst travelling through the catalyst cooler or vary the fluidizing gas in the catalyst cooler. All catalyst particles that enter the catalyst cooler are transferred back to the dense phase of the reactor chamber (Miller, 2011). By adjusting the amount of fluidizing gas, more or less of the heat available can be used to produce steam, too much fluidizing gas will result in more heat leaves with the fluidizing gas, and too little will decrease the heat transfer from the catalyst particles. It is at least possible to recover 80% of the incoming heat and produce steam (Reh, et al., 1979).
11.2 Appendix 2 MTO Water quench description

11.2.1 Water quench scheme

Due to the high content of water, and the content of catalyst particles and dust in the product vapour, the vapour is quenched in a water tower to separate the catalyst dust and some of the water vapour from the product gas. An operating temperature of the water quench of 95 to 115°C is suitable (Wang, et al., 2011). Since catalyst particles and dust are present in the product vapour it is not possible to heat exchange the product gas below the dew point, the catalyst dust will then stick to the heat exchanger wall (Fuglerud, 2013).

To be able to separate the product gas from the entrained catalyst fines, to remove water and any heavy by-products such as C₆⁺ hydrocarbons, the gas coming from the reactor is quenched using a two-stage process. A two-stage process is used since the reactor effluent can contain small amounts of acetic acid, which could build up in a conventional quench process scheme (Miller & Senetar, 2002). An example of the two-stage quench process can be seen in Figure 20.

![Two-stage quench process](image)

Figure 20: Two-stage quench process (Miller & Senetar, 2002).

To recover the heat of the reactor effluent it is first heat exchanged. However, the effluent cannot be cooled below its dew point since the catalyst fines then will stick to the heat exchanger wall. The cooled reactor effluent is then passed to the first quench tower, number 42 in Figure 20, where the effluent is contacted with a relatively pure aqueous stream and a neutralizing agent, which is introduced at the top of the tower. This configuration will result in a hydrocarbon vapour stream leaving at the top of the quench tower and a waste water stream leaving at the bottom. The waste water stream will contain the heavy by-products, neutralized organic acid components, and the catalyst fines present in the reactor effluent. Part of the waste water stream leaving at the bottom may be recycled back to the quench tower at a point above where the reactor effluent is entering. The other part of the waste water stream will be withdrawn and the catalyst fines present will be recovered (Miller & Senetar, 2002).
The vapour stream leaving the first quench tower is first heat exchanged indirectly in an intercondenser, where the vapour stream is partially condensed. The partially condensed reactor effluent is passed further to the second quench tower, number 46 in Figure 20. The reactor effluent is partially condensed before the second quench tower both because one can utilize some of the heat available and because this configuration will require less quench water to reduce the amount of water in the reactor effluent. At the top of the second quench tower, the light olefin vapour is recovered and passed on to the separation processes. At the bottom of the quench tower, a relatively pure aqueous stream is recovered. A portion of the bottom product will be returned to the first quench tower, a portion will be heat exchanged and recycled back to the second quench tower, and the last portion will be passed to a water stripper zone. In the stripper zone any remaining oxygenates and small hydrocarbons, such as propane, will be removed from the water stream as a stripper overhead stream and passed to the reactor effluent leaving the first quench tower. A purified water stream will leave the stripper at the bottom (Miller & Senetar, 2002). The vapour leaving at the top of the second quench tower should contain little more water than the saturation level of the remaining dewatered effluent stream. Preferably it should contain less than 3% of the total amount of water present in the gas leaving the reactor (Beech, et al., 2006).
11.3 Appendix 3 MTO deoxygenate description

11.3.1 DME & Oxygenate recovery scheme

After the quench scheme, the product gas will be cleaned from oxygenates such as methanol and DME. Oxygenates cannot be fed to the coming separation processes. In Figure 21 (Senetar, 2004) is a MTO process, similar to the MTO process licensed by UOP, illustrated in a block flow diagram. Included in Figure 21 is an oxygenate recovery scheme similar to the one included in the MTO process licensed by UOP.

![Figure 21: The MTO process BFD. 1: MTO reaction zone 2: Quench zone 3: Oxygenate recovery zone 4: Primary product separator zone 5: DME stripping zone 6: Primary DME absorption zone 7: Light olefins stripping zone 8: Secondary DME absorption zone (optional) 9: Methanol recovery zone (Senetar, 2004)](image_url)

Block 1 in Figure 21 represents the reaction and regeneration part of the MTO process, where stream no 10 represent the methanol feed. As can be seen in the figure is part of the methanol feed (stream no 11 and 12) used within the oxygenate recovery scheme to absorb oxygenates remaining in the product gas before it is being fed to the MTO reactor (stream no 17 and 30). Block 6 is the primary DME absorption zone, where methanol is used as absorbent. The primary DME absorption zone is a conventional liquid-gas contacting zone packed with a suitable material to enhance vapour liquid interaction as the ascending vapour stream meets the descending liquid stream. The methanol is passed to the upper region of the absorption zone wherein it is passed into counter current contact with the up flowing DME-rich light hydrocarbon vapour stream charged to the absorption zone. The primary absorption zone is operated at scrubbing conditions effective to produce a liquid solvent bottom stream containing methanol, DME, water and substantial and undesired amounts of ethylene and propylene and a light olefin-rich, DME lean overhead vapour product stream. Block no 8 is a
secondary DME absorption zone similar to the primary one. The secondary absorption zone may however be excluded from the scheme (Senetar, 2004).

Block no 3 is referred to as oxygenate recovery zone. This zone operates to strip substantially all oxygenates that are dissolved in the circulating aqueous scrubbing medium, e.g. from the quench scheme, and to recover and recycle these oxygenates to the MTO reactor, block no 1 (Senetar, 2004).

Block no 4 is referred to as the primary product separation zone. A liquid hydrocarbon phase containing substantial amounts of DME, light olefins and unreacted methanol dissolved therein will form in the product separation zone. The liquid hydrocarbon stream will be fed to the DME stripping zone, block no 5. In the DME stripping zone will the liquid hydrocarbon stream contact an up flowing vapour stream under stripping conditions effective to remove DME and light olefins from the liquid hydrocarbon stream generating an overhead vapour stream containing DME, ethylene and propylene. The liquid hydrocarbon stream withdrawn from the bottom of the DME stripping zone, block no 5, will contain C4, C5 and C6 olefins in admixture with a minor amount of C4+ saturates. In a preferred MTO operation will this liquid hydrocarbon stream approximately comprise 8 to 18% of the methanol equivalent and will be transferred back to the MTO reaction zone (not shown in Figure 21) (Senetar, 2004).

At least a portion of the liquid solvent bottom stream recovered from the DME absorption zone block no 6, is passed to the light olefin stripping zone, block no 7. The light olefins stripping zone is operated at a severity level sufficient to lift a substantial portion of the ethylene and propylene contained in the liquid solvent bottom stream from the DME absorption zone without stripping any significant portion of the methanol. A stripper overhead stream containing DME, ethylene and propylene and an aqueous bottom stream containing DME, methanol, water and reduced amount of light olefins will form. The liquid stream will be recycled back to the reaction zone (Senetar, 2004).

The light olefin-containing vapour stream withdrawn from the olefin stripping zone, block no 7, will contain minor amounts of DME that co-boil with the propylene material contained therein and consequently needs to be further treated in order to remove the DME. At least a portion is sent to the primary DME absorption zone, block no 6, in order to remove the residual DME (Senetar, 2004).

At least a portion of the DME-lean and light olefin-rich overhead vapour product stream from the primary DME absorption zone, block no 6, is passed to the lower region of the methanol recovery zone, block no 9, in order to remove residual methanol vapour there from. In the methanol recovery zone is the vapour product scrubbed free of methanol with a portion of the by-product water stream recovered in the oxygenate recovery zone, block no 3. The resulting methanol-lean light olefin vapour stream is then withdrawn and constitutes the light olefin product stream (Senetar, 2004).

The resulting light olefin product stream is assumed to be sent to the existing light end separation processes available at the cracker site e.g. forming polymer grade ethylene and propylene.
11.4 Appendix 4 E2E process description

11.4.1 Process description
Due to that the reaction is endothermic, and the reactor chosen is adiabatic, one utilizes four fixed bed reactors with intermediate heating in a furnace in this process design. The catalyst used in the reactor is the Syndol catalyst and is stable for at least two years. Compared to the MTO process, no regeneration of the catalyst material is needed in the ethanol to ethylene process (Weatherly Inc. Chematur Engineering group, 2010).

The ethanol feed is first supplied to a vaporizer. The vaporized ethanol is then heated to the reaction temperature in the preheat section of the furnace and then passed to the first of the four adiabatically operated catalyst beds in the reactor. Ethanol is converted to ethylene as it passes over the catalyst and because of the endothermic reaction the reaction gases cool progressively thereby lowering the reaction rate. The effluent gases from the first bed are therefore heated in the furnace again to the desired temperature and then send to the second bed of the reactor where some more ethanol is converted. Similar operations are conducted in the third and fourth beds of the reactor (Weatherly Inc. Chematur Engineering group, 2010).

The hot ethylene-water vapour mixture leaving the last bed on the reactor enters the waste heat boiler where some of the heat is recovered as steam. The cooled gases are quenched with water in the quench system. In this system, the ethylene is cooled and freed of unreacted ethanol and some of the side products. The quench water is recirculated for heat removal and waste water is purged (Weatherly Inc. Chematur Engineering group, 2010).

Upon leaving the top of the quench tower, the ethylene is compressed and passed in series through a caustic scrubber and a fixed bed gas dryer to reduce the concentrations of oxygenated impurities and water to specification levels. The gas dryers need regeneration with nitrogen gas. One of the two dryers is therefore being regenerated whilst the other is in operation (Weatherly Inc. Chematur Engineering group, 2010).

The ethylene from the dryer is distilled in an ethylene column operating at low temperature where heavy impurities are removed. The ethylene product from the overhead of this column meets all polymer grade specifications except for carbon monoxide (Weatherly Inc. Chematur Engineering group, 2010).

To reduce the carbon monoxide to specification levels, the overhead to the ethylene column is sent to a stripper where the carbon monoxide is stripped from the ethylene. The overheads from the ethylene column and stripper use a common condenser and accumulator. The carbon monoxide is purged off the accumulator with a small percentage of the ethylene. Product ethylene obtained at the bottom of the stripper is used to cool a number of process streams and is then delivered to the battery limits as a vapour at 6.5 bar(g) (Weatherly Inc. Chematur Engineering group, 2010).

Excluded from Figure 4 above is a water stripper column which is used to strip the waste water formed during reaction. One waste water stream in the need for treatment is formed in the water stripper.
### 11.5 Appendix 5 MTO/Cracker steam balance

#### 11.5.1 Main steam balance differences

**Table 28: Main differences between the steam balances of the cases.**

<table>
<thead>
<tr>
<th>Case</th>
<th>Steam from furnaces t/h</th>
<th>Dilution steam to furnaces t/h</th>
<th>MTO steam (20 barg) requirements t/h</th>
<th>MTO steam (85 barg 485°C / 20 barg 130°C) requirements to meet 20 barg t/h</th>
<th>Calculated boiler load t/h</th>
</tr>
</thead>
<tbody>
<tr>
<td>No MTO 40 t/h ethane</td>
<td>262</td>
<td>79</td>
<td>0</td>
<td>0</td>
<td>72</td>
</tr>
<tr>
<td>HE 300 kton olefins 40 t/h ethane</td>
<td>181</td>
<td>56</td>
<td>32.4</td>
<td>26.3 / 6.1</td>
<td>113</td>
</tr>
<tr>
<td>HP 300 kton olefins 40 t/h ethane</td>
<td>182</td>
<td>57</td>
<td>42.6</td>
<td>34.5 / 8.1</td>
<td>120.5</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>No MTO 68.5 t/h ethane</td>
<td>245</td>
<td>83</td>
<td>0</td>
<td>0</td>
<td>74</td>
</tr>
<tr>
<td>HE 300 kton olefins 68.5 t/h ethane</td>
<td>187</td>
<td>60</td>
<td>32.4</td>
<td>26.3 / 6.1</td>
<td>108</td>
</tr>
<tr>
<td>HP 300 kton olefins 68.5 t/h ethane</td>
<td>189</td>
<td>58</td>
<td>42.6</td>
<td>34.5 / 8.1</td>
<td>121.5</td>
</tr>
</tbody>
</table>
### 11.6 Appendix 6 Example cracker steam balance

#### 11.6.1 Steam balance example

Table 29: Example of a steam balance (t/h). 40 t/h C2 case No MTO.

<table>
<thead>
<tr>
<th></th>
<th>Furnaces</th>
<th>Boilers</th>
<th>MTO</th>
<th>Export</th>
<th>CT-1951</th>
<th>Turbo-alternator</th>
<th>C-1701</th>
<th>Pump turbines</th>
<th>Dilution steam</th>
<th>HP Let-down</th>
<th>Import</th>
<th>Losses HP steam</th>
<th>MP let-down</th>
<th>Reboiler s etc.</th>
<th>IP let-down</th>
<th>Dearator steam</th>
<th>LP let-down</th>
</tr>
</thead>
<tbody>
<tr>
<td>HP steam, 85 barg</td>
<td>262</td>
<td>72</td>
<td>0</td>
<td>-25</td>
<td>-187</td>
<td>-108</td>
<td></td>
<td>-3</td>
<td></td>
<td>-5</td>
<td>8</td>
<td>-5</td>
<td>8</td>
<td>-1</td>
<td>7</td>
<td>-7</td>
<td>7</td>
</tr>
<tr>
<td>MP steam, 8.8 barg</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>187</td>
<td>108</td>
<td>-110</td>
<td>-60</td>
<td>-79</td>
<td>5</td>
<td>-7</td>
<td>-44</td>
<td>7</td>
<td>-7</td>
<td>7</td>
</tr>
<tr>
<td>IP steam, 2.7 barg</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>7</td>
<td>-7</td>
<td>7</td>
<td>-7</td>
<td>7</td>
</tr>
<tr>
<td>LP steam, 1.8 barg</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>7</td>
<td>-127</td>
<td>7</td>
<td>-30</td>
<td>0</td>
</tr>
<tr>
<td>CW</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>57</td>
<td>0</td>
<td>57</td>
<td>0</td>
<td>0</td>
</tr>
</tbody>
</table>
11.7 Appendix 7 MTO/Cracker Mass Balances

11.7.1 Overall Mass Balances for MTO 300 kton combined with the cracker plant

Table 30: Feed to Cracker furnaces and the MTO process. Cracking of 40 t/h ethane, maximization of ethylene (HE) and maximization of propylene (HP) in the MTO process, integrated with the cracker.

<table>
<thead>
<tr>
<th>Fresh feed t/h</th>
<th>No MTO 40 t/h ethane</th>
<th>HE 300 kton olefins 40 t/h ethane</th>
<th>HP 300 kton olefins 40 t/h ethane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethane</td>
<td>40</td>
<td>40</td>
<td>40</td>
</tr>
<tr>
<td>Propane</td>
<td>20.9</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Butane</td>
<td>90</td>
<td>46.1</td>
<td>46.9</td>
</tr>
<tr>
<td>Naphtha</td>
<td>22.11</td>
<td>22.11</td>
<td>22.11</td>
</tr>
<tr>
<td>Total fresh feed</td>
<td>173</td>
<td>108.2</td>
<td>109</td>
</tr>
<tr>
<td>Methanol (100wt%) feed to MTO</td>
<td>0</td>
<td>103</td>
<td>100</td>
</tr>
<tr>
<td>Ethanol (for ETBE)</td>
<td>2.05</td>
<td>1.26</td>
<td>1.27</td>
</tr>
<tr>
<td>NG as make-up (48 MJ/kg)</td>
<td>1.5</td>
<td>9.3</td>
<td>9.8</td>
</tr>
</tbody>
</table>

Table 31: Product from Cracker furnaces combined with the MTO process. Cracking of 40 t/h ethane, maximization of ethylene (HE) and maximization of propylene (HP) in the MTO process, integrated with the cracker.

<table>
<thead>
<tr>
<th>Products t/h</th>
<th>No MTO 40 t/h ethane</th>
<th>HE 300 kton olefins 40 t/h ethane</th>
<th>HP 300 kton olefins 40 t/h ethane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethylene</td>
<td>75.2</td>
<td>68.4(^1)</td>
<td>68.6(^1)</td>
</tr>
<tr>
<td>Propylene</td>
<td>27.3</td>
<td>30.5</td>
<td>30.5</td>
</tr>
<tr>
<td>ETBE</td>
<td>4.5</td>
<td>2.8</td>
<td>2.8</td>
</tr>
<tr>
<td>CBFS</td>
<td>2.1</td>
<td>1.6</td>
<td>1.6</td>
</tr>
<tr>
<td>Raw and treated SCN</td>
<td>15.2</td>
<td>14.4</td>
<td>14.5</td>
</tr>
<tr>
<td>Raff 2 (C4 olefins)</td>
<td>13.5</td>
<td>13.1</td>
<td>13.1</td>
</tr>
<tr>
<td>(\text{H}_2)</td>
<td>0.25</td>
<td>0.25</td>
<td>0.25</td>
</tr>
<tr>
<td>Fuel gases (57.26 MJ/kg)</td>
<td>35.7(^1)</td>
<td>22.9(^1)</td>
<td>23(^1)</td>
</tr>
<tr>
<td>Stripped water to BL</td>
<td>-</td>
<td>60.96</td>
<td>60.92</td>
</tr>
<tr>
<td>Waste water to WWT</td>
<td>-</td>
<td>6.76</td>
<td>6.75</td>
</tr>
<tr>
<td>Coke (burned in regenerator)</td>
<td>1.88</td>
<td>1.16</td>
<td></td>
</tr>
</tbody>
</table>

\(^1\) Lower than base case due to limitations that arises when adding too much MTO product into the light end. Compared to the below case, cracking of additional ethane, has modifications on the C2 splitters been included due to coming activities.
Table 32: Feed to Cracker furnaces and the MTO process. Cracking of 68.5 t/h ethane, maximization of ethylene (HE) and maximization of propylene (HP) in the MTO process, integrated with the cracker.

<table>
<thead>
<tr>
<th>Fresh feed t/h</th>
<th>No MTO 68.5 t/h ethane</th>
<th>HE 300 kton olefins 68.5 t/h ethane</th>
<th>HP 300 kton olefins 68.5 t/h ethane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethane</td>
<td>68.5</td>
<td>68.5</td>
<td>68.5</td>
</tr>
<tr>
<td>Propane</td>
<td>0.16</td>
<td>0</td>
<td>19.9</td>
</tr>
<tr>
<td>Butane</td>
<td>68.3</td>
<td>21.8</td>
<td>1.4</td>
</tr>
<tr>
<td>Naphtha</td>
<td>22.11</td>
<td>22.11</td>
<td>22.11</td>
</tr>
<tr>
<td>Total fresh feed</td>
<td>159</td>
<td>112.4</td>
<td>112</td>
</tr>
<tr>
<td>Methanol (100wt%) feed to MTO</td>
<td>0</td>
<td>103</td>
<td>100</td>
</tr>
<tr>
<td>Ethanol (for ETBE)</td>
<td>1.73</td>
<td>0.87</td>
<td>0.5</td>
</tr>
<tr>
<td>NG as make-up (48 MJ/kg)</td>
<td>3.5</td>
<td>8.3</td>
<td>9.2</td>
</tr>
</tbody>
</table>

Table 33: Product from Cracker furnaces combined with the MTO process. Cracking of 68.5t/h ethane, maximization of ethylene (HE) and maximization of propylene (HP) in the MTO process, integrated with the cracker.

<table>
<thead>
<tr>
<th>Products t/h</th>
<th>No MTO 68.5 t/h ethane</th>
<th>HE 300 kton olefins 68.5 t/h ethane</th>
<th>HP 300 kton olefins 68.5 t/h ethane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethylene</td>
<td>77.1</td>
<td>77.9</td>
<td>77.8</td>
</tr>
<tr>
<td>Propylene</td>
<td>19.5</td>
<td>26.4</td>
<td>28</td>
</tr>
<tr>
<td>ETBE</td>
<td>3.8</td>
<td>1.9</td>
<td>1.1</td>
</tr>
<tr>
<td>CBFS</td>
<td>2.3</td>
<td>1.9</td>
<td>1.8</td>
</tr>
<tr>
<td>Raw and treated SCN</td>
<td>15</td>
<td>14.8</td>
<td>13.6</td>
</tr>
<tr>
<td>Raff 2 (C4 olefins)</td>
<td>11.2</td>
<td>11.5</td>
<td>11.1</td>
</tr>
<tr>
<td>H₂</td>
<td>0.25</td>
<td>0.25</td>
<td>0.25</td>
</tr>
<tr>
<td>Fuel gases (57.26 MJ/kg)</td>
<td>32.5¹</td>
<td>24.9¹</td>
<td>24.9¹</td>
</tr>
<tr>
<td>Stripped water to BL</td>
<td>-</td>
<td>60.96</td>
<td>60.92</td>
</tr>
<tr>
<td>Waste water to WWT</td>
<td>-</td>
<td>6.76</td>
<td>6.75</td>
</tr>
<tr>
<td>Coke (burned in regenerator)</td>
<td>-</td>
<td>1.88</td>
<td>1.16</td>
</tr>
</tbody>
</table>

¹Whereof 5.8 t/h sold (333 GJ/h)
### 11.8 Appendix 8 LCA study Mass Balance

#### 11.8.1 Overall Material Balance for the LCA study case

**Table 34: Feed to Cracker furnaces and the MTO process (production of 560 kton ethylene/y).** Cracking of ~68.5 t/h ethane, maximization of propylene (HP) in the MTO process (production of ~200 kton green olefins/y), integrated with the cracker and production of 30 kton ethylene from E2E.

<table>
<thead>
<tr>
<th>Fresh feed t/h</th>
<th>HP ~200 kton olefins ~68.5 t/h ethane [t/h]</th>
<th>HP ~200 kton olefins ~68.5 t/h ethane [kton/y]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethane</td>
<td>61.65</td>
<td>493.2</td>
</tr>
<tr>
<td>Propane</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Butane</td>
<td>32.68</td>
<td>261.4</td>
</tr>
<tr>
<td>Naphtha</td>
<td>19.9</td>
<td>159.2</td>
</tr>
<tr>
<td>Methanol (100wt%) feed to MTO</td>
<td>70.2</td>
<td>552.2</td>
</tr>
<tr>
<td>Ethanol (for ETBE)</td>
<td>1.03</td>
<td>8.26</td>
</tr>
<tr>
<td>Natural gas (48 MJ/kg) (as make-up)</td>
<td>10.63</td>
<td>85</td>
</tr>
<tr>
<td>Ethanol (E2E)</td>
<td>6.8</td>
<td>54.75</td>
</tr>
<tr>
<td>NG (48 MJ/kg to E2E)</td>
<td>0.35</td>
<td>2.82</td>
</tr>
</tbody>
</table>

**Table 35: Product from Cracker furnaces combined with the MTO process (560 kton ethylene/y).** Cracking of ~68.5 t/h ethane, maximization of propylene (HP) in the MTO process (production of ~200 kton green olefins/y), integrated with the cracker and production of 30 kton ethylene from E2E.

<table>
<thead>
<tr>
<th>Products t/h</th>
<th>HP ~200 kton olefins ~68.5 t/h ethane [t/h]</th>
<th>HP ~200 kton olefins ~68.5 t/h ethane [kton/y]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethylene</td>
<td>70</td>
<td>560</td>
</tr>
<tr>
<td>Propylene</td>
<td>24.94</td>
<td>199.5</td>
</tr>
<tr>
<td>ETBE</td>
<td>2.25</td>
<td>18.02</td>
</tr>
<tr>
<td>CBFS</td>
<td>1.68</td>
<td>13.41</td>
</tr>
<tr>
<td>Raw and treated SCN</td>
<td>13.38</td>
<td>107</td>
</tr>
<tr>
<td>Raff 2 (C4 olefins)</td>
<td>11.43</td>
<td>91.4</td>
</tr>
<tr>
<td>H₂</td>
<td>0.66</td>
<td>5.3</td>
</tr>
<tr>
<td>Fuel gases (57.26 MJ/kg)</td>
<td>23.6</td>
<td>189</td>
</tr>
<tr>
<td>Stripped water to BL</td>
<td>40.61</td>
<td>324.9</td>
</tr>
<tr>
<td>Waste water to WWT</td>
<td>6.75</td>
<td>36</td>
</tr>
<tr>
<td>Coke (burned in regenerator)</td>
<td>0.773</td>
<td>6.18</td>
</tr>
<tr>
<td>Ethylene (E2E)</td>
<td>3.75</td>
<td>30</td>
</tr>
<tr>
<td>Waste fuel (33.8 MJ/kg)</td>
<td>0.31</td>
<td>2.52</td>
</tr>
<tr>
<td>Waste water to WWT</td>
<td>3.25</td>
<td>26</td>
</tr>
</tbody>
</table>
11.9 Appendix 9 MTO Utility requirements

11.9.1 MTO Utility description

The following utilities and chemicals are used within the MTO process (original design), net input to MTO ISBL from Cracker site OSBL.

Caustic
Caustic (10wt%) is used as a neutralizing agent in the quench tower scheme, see section Appendix 2, for more information. Specification of the caustic:

- 10wt%
- 6.6 barg
- 30°C

Methane
Methane purge gas is used within the MTO reactor for the various instruments. Specification of the methane purge:

- 10.1 barg
- 20°C

Steam
Steam purges and fluidizing steam is used within the MTO reactor to e.g. help fluidizing the catalyst bed. Specification for the steam:

- 20 barg
- 215°C

Nitrogen
Nitrogen gas is used for the transfer MTO catalyst material in the system. Specification for the nitrogen:

- 7 barg
- 75°C

Air
Air is fed to the MTO catalyst regenerator and used as oxygen supply. Specification for the air:

- 4.6 barg
- 27°C

Demineralised water
Net input of BFW due to BFW losses. Specification for the BFW:

- 7 barg
- 90°C

Cooling water
Cooling water is used in various heat exchangers. Cooling water specifications:

- Supply temperature 30°C
- Maximum return temperature 40°C
Electricity

Electricity is used in various units within the MTO process.

HHP stem

Net input of HHP steam to the MTO process (to drive compressor and air blower). Specification for the HHP steam:

- Pressure 100 barg
- Temperature 510°C
- Condensate pressure 3.1 barg
- Condensate temperature 90°C
- Saturated liquid 314°C

HP stem

Net input of HP steam to the MTO process (only during start-up). Specification for the HHP steam:

- Pressure 40 barg
- Temperature 420°C
- Condensate pressure 7.0 barg
- Condensate temperature 90°C
- Saturated liquid 252°C

MP steam

Net input of MP steam to the MTO process (steam level used for heat exchangers etc. within the MTO process). Specification for the MP steam:

- Pressure 20 barg
- Temperature 215°C
- Condensate pressure 7.0 barg
- Condensate temperature 90°C
- Saturated liquid 215°C

Due to that electrical driven motors often are cheaper than turbine driven ones and to avoid using high pressure steam, electrical motors for the compressor and air blower was chosen. This change in design result in that more MP steam (20barg) and electricity needs to be fed to the MTO process (see Table 28, Table 30 and Table 32 in above appendices) compared to original design, however one avoids the input of HHP steam. Another change is the exchange of steam heated start-up heaters (use HP steam) to direct fired start-up heaters to avoid the need for an extra steam pressure level.
**11.10 Appendix 10 E2E By-products**

**11.10.1 Fuel By-products from the E2E process**

Table 36: Recovered liquid stream from fuel collection system.

<table>
<thead>
<tr>
<th>Item</th>
<th>Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ethylene</td>
<td>0.55wt%</td>
</tr>
<tr>
<td>Ethane</td>
<td>0.45wt%</td>
</tr>
<tr>
<td>Propylene</td>
<td>0.34wt%</td>
</tr>
<tr>
<td>1-Butene</td>
<td>19.89wt%</td>
</tr>
<tr>
<td>Acetaldehyde</td>
<td>3.36wt%</td>
</tr>
<tr>
<td>Diethyl Ether</td>
<td>1.75wt%</td>
</tr>
<tr>
<td>Ethanol</td>
<td>63.02wt%</td>
</tr>
<tr>
<td>Water</td>
<td>10.17wt%</td>
</tr>
<tr>
<td>3-Methyl-1-Butanol</td>
<td>0.47wt%</td>
</tr>
</tbody>
</table>

Table 37: Recovered vapour stream from the fuel collection system.

<table>
<thead>
<tr>
<th>Item</th>
<th>Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrogen</td>
<td>0.14wt%</td>
</tr>
<tr>
<td>Carbon monoxide</td>
<td>1.42wt%</td>
</tr>
<tr>
<td>Methane</td>
<td>0.13wt%</td>
</tr>
<tr>
<td>Ethylene</td>
<td>21.47wt%</td>
</tr>
<tr>
<td>Ethane</td>
<td>11.14wt%</td>
</tr>
<tr>
<td>Propylene</td>
<td>2.53wt%</td>
</tr>
<tr>
<td>1-Butene</td>
<td>44.40wt%</td>
</tr>
<tr>
<td>Acetaldehyde</td>
<td>3.00wt%</td>
</tr>
<tr>
<td>Diethyl Ether</td>
<td>5.50wt%</td>
</tr>
<tr>
<td>Ethanol</td>
<td>9.23wt%</td>
</tr>
<tr>
<td>Water</td>
<td>1.03wt%</td>
</tr>
<tr>
<td>3-Methyl-1-Butanol</td>
<td>0.01wt%</td>
</tr>
</tbody>
</table>
11.11 Appendix 11 E2E Utility requirements

11.11.1 E2E Utility description

Below are the utilities and chemicals utilized within the E2E process introduced.

Caustic
Caustic is used in the caustic wash to neutralize the product gas in a mixture with water. The solution should be 25-50% NaOH in water. Specifications for the caustic:

- 3 barg
- 19°C

Caustic is not available at the existing process plant. A storage tank and facilities for import of caustic via trailer is recommended to be located close to the process area OSBL.

Demineralised water
Demineralised water is used together with the caustic in the caustic wash. Demineralized water is used instead of common potable water to prevent unwanted reactions in some parts of the process. Specifications for the demineralised water:

- 5 barg
- 19°C

The required demineralized water supply is met by taking out a partial flow from BFW supply before necessary chemicals are added.

Natural Gas
Natural gas is used in the furnace to supply the ethanol gas with heat. Assumed is also that the waste streams from the E2E process (referred to in E2E Product specifications (incl. By-products)) is used to generate heat in the furnace. Specifications for the natural gas:

- 48 MJ/kg
- 3 barg
- 25°C

At the process plant, natural gas is available in an existing pipe-rack and the capacity is assumed to be sufficient for the plant.

Nitrogen
Nitrogen is used for inerting of equipment and pipelines, regenerating the molecular sieve dryers as well as other maintenance purposes. Specifications for the nitrogen gas:

- 2.5 barg
- 25°C

At the process plant, nitrogen is available at the existing piperack and the capacity is assumed to be sufficient.

Electricity
Electricity is used for various process units within the E2E process. Power to the plant will be supplied via existing electrical sub-station. Based on a comparison of power consumers for the existing old process and the new E2E-plant the total process power supply is assumed to be sufficient.
Steam
Steam is used to supply heat to the product stream at various locations, e.g. in reboilers. Steam is also used for direct injection to an aqueous waste column. Specifications for the steam used:

- Supply pressure 9 barg
- Supply temperature 180°C
- Condensate return pressure 1 barg
- Condensate return temperature 121°C

At the process plant, steam with a pressure of 9 barg is available in the existing piperack and the capacity is assumed to be sufficient for the plant.

Boiler Feed Water
BFW is e.g. utilized within the waste heat boiler to generate steam. Specifications for the BFW:

- Supply pressure 2.8 barg
- Supply temperature 115°C
- Condensate return pressure 1 barg
- Condensate return temperature 121°C

At the process plant, BFW is available in the existing piperack and the capacity is assumed to be sufficient for the plant.

Cooling water
Cooling water is also used within the E2E process. Specifications for the cooling water:

- Supply pressure 3.5 barg
- Supply temperature 28°C
- Return pressure 1.5 barg
- Return temperature 35°C

At the process plant, cooling water will be supplied from an existing fresh water cooling system via underground piping. Connection is possible at nozzles above ground in the actual area. The capacity and the temperature are assumed to be sufficient.

Instrument air
Instrument air is required for normal instrument and valve operation purposes.

At the process plant, instrument air is available in the existing piperack and the capacity is assumed to be sufficient for the plant.

Catalyst material

Molecular sieves
Type 3A molecular sieve aluminium silicates for dryer. Expected life of 2 years.
11.11.2 E2E Utility balance

Input of utilities based on the production of 60 kton green ethylene in the E2E process can be seen in Table 38 below.

**Table 38: Input and return of utilities when producing 60 kton of ethylene within the E2E process based on 8000 h/y.**

<table>
<thead>
<tr>
<th>Item</th>
<th>Production of 60 kton ethylene</th>
</tr>
</thead>
<tbody>
<tr>
<td>Caustic (as 100% NaOH)</td>
<td>239 ton/y</td>
</tr>
<tr>
<td>Demineralised water</td>
<td>2296 ton/y</td>
</tr>
<tr>
<td>Natural Gas</td>
<td>$132.7 \times 10^6$ MJ/y</td>
</tr>
<tr>
<td>Nitrogen (regeneration of dryers)</td>
<td>160000 Nm3/y</td>
</tr>
<tr>
<td>Electricity</td>
<td>$20.4 \times 10^6$ kWh/y</td>
</tr>
<tr>
<td>Steam*</td>
<td>75.66 kton/y</td>
</tr>
<tr>
<td>Boiler Feed Water*</td>
<td>26.8 kton/y</td>
</tr>
<tr>
<td>Condensate return*</td>
<td>93.51 kton/y</td>
</tr>
<tr>
<td>Boiler feed water blow-down*</td>
<td>1.31 kton/y</td>
</tr>
<tr>
<td>Cooling water</td>
<td>11 160 000 m³/y (net input=0 m³/y)</td>
</tr>
<tr>
<td>Catalyst material</td>
<td>10 200 kg/y (start-up: 20 400kg)</td>
</tr>
<tr>
<td>Molecular sieve</td>
<td>900 kg/y (start-up: 1 800kg)</td>
</tr>
</tbody>
</table>

*Net input of water equals 6.68 kton/y