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MASTER'S THESIS

Biomass- and Coke Oven Gas Based Methanol Production

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Preface

This Master Thesis has been carried out at the division of Energy Engineering at Luleå University of Technology during the spring term 2007.

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Abstract

This study considers production of methanol as a motor fuel, generated from coke oven gas at the steel mill SSAB Tunnplåt in Luleå. In addition, different systems with a biomass gasification plant included are considered. The produced synthesis gas is mixed with the coke oven gas to enhance the methanol production. Different technologies for production of the alcohol are applied. The selected techniques are based upon their advantageous by economical and technical characteristics. The selected systems are modelled in Aspen plus, which is a widely used process simulation software. The resulting dimensions of streams, units and the energy balances are subsequently used for the economic analysis.

Global warming caused by an increased concentration of greenhouse gases is of major concern today. It is predicted that global warming will cause sea level rise and other environmental impacts that may harm humans and nature. The significant increase of greenhouse gases over the last 200 years is almost entirely due to human activities and particularly due to combustion of fossil fuels. By using biomass as fuel for production of methanol as motor fuel, it would contribute to decrease the total green house gas emission level. The coke oven gas which origins from fossil based coal, could be used in a better manner as fuel for methanol production than for increase the present electricity production at LULEKRAFT AB (LUKAB). That is because the methanol production leads to a product with lower specific CO_2 emissions due to that the conversion- and total plant efficiency are much higher.

The wood-pellets produced from sawdust at Bioenergi are to a large extent exported from the county to heat and power plants, which purchase the pellets to a low price. A better option would be to use the sawdust for methanol production, which would be more profitable and could be used within the county and thereby also minimize the transportation costs and green house gas emissions.

The conclusion is that methanol produced from 400 GWh coke oven gas would have a production cost in the range of 1,20-2,44 SEK/litre. Methanol produced from both 420 GWh sawdust and the coke oven gas would result in a production cost in the range of 1,20-2,70 SEK/litre.

The minimum methanol sale price was calculated via a fuel value analysis which analyzed the fuel input value within the MEOH plant, LUKAB and Bioenergi. The methanol produced from the coke oven gas would have a minimum sale price in the range of 1,20-2,68 SEK/litre. Methanol produced from both sawdust and coke oven gas would result in a minimum sale price in the range of 1,28-3,17 SEK/litre.

The produced methanol from sawdust and coke oven gas could potentially replace about half of the present gasoline demand in the county. This would decrease the import of fossil based motor fuel considerable.

Sammanfattning

Denna studie behandlar produktion av metanol som fordonsbränsle från koksgas producerad vid stålverket SSAB Tunnplåt i Luleå. Även system med en förgasningsanläggning inkluderad för produktion av syntesgas som mixas med koksgasen för att ökat metanolutbytet är studerad. Olika tekniker för metanolproduktion är beaktad, vilka är baserad på deras fördelar vad gäller ekonomi och teknisk karakteristik. Systemen är modellerade i Aspen plus, ett vida använt processimuleringsprogram. Simuleringsresultaten i form av flöden och energibalanser används parallellt för ekonomiska analyser.

Ökad växthuseffekt på grund av ökande koncentrationer av växthusgaser är ett stort problem idag. Det förväntas att växthuseffekten kommer ha stora inverkningar på miljön, så som stigande havs nivåer och andra miljöeffekter som kan skada människan och naturen. Den markanta ökningen av växthusgaser som skett de senaste 200 åren beror nästan enbart på människans påverkan och speciellt på grund av förbränning av fossila bränslen. Att använda biomassa som bränsle för produktion av metanol som fordonsbränsle kan bidra till minskande utsläpp av växthusgaser. Koksgasen som från början utvinns från fossilbaserat kol, kan som bränsle för metanolproduktion komma till större nytta än som bränsle för att öka dagens elproduktion vid LULEKRAFT AB (LUKAB). Detta beror på att produktion av metanol som produkt kan leda till lägre specifikt CO_2 utsläpp, eftersom konverteringsverkningsgraden är högre så väl som den totala verkningsgraden.

Pellets producerad från sågspån vid Bioenergi exporteras i stor utsträckning till värme och kraftvärmeanläggningar utanför länet som köper den till lågt pris. Ett bättre alternativ vore att använda sågspånet för metanolproduktion som skulle inbringa större vinst samt att metanolen skulle kunna användas inom länet vilket skulle minska transportkostnaderna och konsekvent utsläpp av växthusgaser.

Slutsatsen är att metanol producerad från 400 GWh koksgas har en produktionskostnad inom spannet 1,20-2,44 SEK/liter. Metanol producerad från både 420 GWh sågspån och koksgas har en produktionskostnad inom spannet 1,20-2,70 SEK/liter.

Minsta försäljningspris beräknades via en bränslevärdeanalys som analyserar värdet på använd råvara vid MEOH anläggning, LUKAB och Bioenergi. Metanol producerad från koksgas har ett minsta försäljningspris inom spannet 1,20-2,68 SEK/liter. Metanol producerad från både sågspån och koksgas har ett minsta försäljningspris inom spannet 1,28-3,17 SEK/liter.

Metanol producerad från sågspån och koksgas kan potentiellt ersätta omkring halva dagens behov av bensin i länet. Detta skulle kunna sänka importen av fossilbaserat fordonsbränsle väsentligt.

Abbreviations

ATR	Autothermal reforming
BFBG	Bubbling fluidized bed gasifier
CFBG	Circulated fluidized bed gasifier
CHP	Combined heat and power
FBG	Fluidized bed gasifier
HRSG	Heat Recovery Steam Generator
HTHP	High temperature and high pressure
LEAB	Luleå Energi AB
LTLP	Low temperature and low pressure
LTU	Luleå University of Technology
LUKAB	LULEKRAFT AB
MC	Moister content
MEOH	Methyl Alcohol
MFV	Minimum fluidizing velocity
SMR	Steam methane reforming
twe	Tonne water evaporated
WGS	Water gas shift
wt	Weight
Al_2O_3	Aluminium oxide
C	Carbon
C_2H_2	Acetylene
C_2H_4	Ethylene
C_2H_6	Ethane
C_3H_8	Propane
CH_3OH	Methanol
CO	Carbon monoxide
CO_2	Carbon dioxide
COS	Carbonyl sulphide
CH_4	Methane

Cl_2	Chlorine
Cr_2O_3	Chrome oxide
Cu	Copper
Fe	Iron
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen sulphide
HCL	Hydrogen chloride
HCN	Hydrogen cyanide
N_2	Nitrogen
NH_3	Ammonia
Ni	Nickel
NO	Nitrogen oxide
NO_2	Nitrogen dioxide
O_2	Oxygen
S	Sulphur
SO_2	Sulphur dioxide
SO_3	Sulphur trioxide
ZnO	Zink oxide

Contents

1 Introduction

1.1 Background

Global warming caused by an increased concentration of greenhouse gases in the atmosphere is of major concern today. It is predicted that the global warming will cause sea level rise and other environmental impacts that may harm humans and nature. The significant increase of greenhouse gases over the last 200 years is almost entirely due to human activities and particularly due to combustion of fossil fuels [1].

Renewable fuels origin, e.g. wood, do not contribute to global warming since the CO_2 emitted is taken up by new growth to form a closed loop. It is therefore desirable to produce liquid transportation fuels from sustainable origin instead of using fossil fuel as oil, since the transportation sector alone stands for a major part of all the greenhouse gases emitted. Also the oil supply is not infinite and sooner or later other fuels must be used [1].

Norrbotten is a county with large biomass resources and potentials to grow energy crops such as reed canary grass. The Swedish government has in accordance with the Kyoto protocol put up a goal to reduce the emissions of greenhouse gases. This is why it is of strong national and international interest that a biomass rich county as Norrbotten makes use of the potential to an as great extent as possible [2].

The integrated steelmaking industry is one of the major industries that have large CO_2 emissions, since coal is used as main reduction material when the agglomerated iron ore is reduced and melted in the blast furnace. In order to minimize the CO_2 emission from integrated steelmaking, a natural step is to minimize the energy use. This can be achieved by improving the performance of each individual process or by looking at the complete system. An analysis of the total system is preferred, as the processes in integrated steelmaking are connected through both primary and secondary products within the system. Changes within one sub-process can and will affect the total system. A key in achieving a minimised energy use for the system is to have a good utilisation of the process off-gases. Today, the process gases are used within the system as fuel in heating ovens, hot stoves etc. At SSAB in Luleå, the system is not equipped with a rolling mill at the site, instead the excess gases are burned as primary fuel in the nearby combined heat and power plant (CHP) LULEKRAFT AB (LUKAB) for production of electricity, process steam and district heating. Furthermore, flue gases from the CHP plant are delivered to a wood-pellets production plant (Bioenergi) for drying of the sawdust [3], [4].

There are no doubts that the present situation is an excellent example of process integration where industrial energy residues can be of great use for other industries as well as for the general public. However, the use of process gases to produce hot water and electricity will result in high specific CO_2 emissions as the origin is coal. One question is if there is a better and more optimal way to utilise this huge excess of energy, carried by the gases from the steel mill? Is the best product mix electricity, district heating and drying gas for wood pellets production from the viewpoint of energy, environment and economy? A more efficient use of the process gases would be to produce a product for which the alternatives have similar CO_2 emission [3], [4].

It is a relevant question to ask whether the steel work gases could be used for other and perhaps better purpose than heat production. From the viewpoint of exergy, the coke oven gas, with a heating value of 17-18 MJ/Nm^3 and the high content of H_2, should be too valuable to use for heat production. The coke oven gas could be used as fuel input for production of methanol as transportation fuel. Since the coke oven gas has high H_2 content, which is one of the key components in methanol, the production efficiency would be high [3], [4].

The sawdust used for pellets production may not be the best way to produce a renewable fuel since the sawdust could have a higher value as fuel for methanol production. The pellets to large extent are exported out from the county to heat and power plants which purchase the pellets to a low price. A better option could be to produce methanol to be used within the county and thereby perhaps minimize the transportation costs and green house gas emissions.

1.2 Project objectives

It is of great interest to make use of the generated steel work gases as efficiently as possible from the point of view of energy and consequently environment and economy. This study aims to show under which conditions production of methanol for transport purposes could be a feasible approach to utilise the generated gases in a better manner than today. Another objective is to show the possible technical- and economic gains to also include a biomass gasification plant for production of synthesis gas to enhance the methanol production and decrease the specific CO_2 emissions further [3], [4].

The most important constraint is that the district heating demand in the town of Luleå must be fulfilled as well as the electricity generation constraint. Other limitations are the availability of the steel work gases as well as the locally available biomass resources as it is crucial to minimise the feedstock transportation costs. There may also be other constraints regarding electricity production that must be considered [3], [4].

It is of great significance to have a future perspective of the study as the construction of this kind of plant may not be realised up until 15-20 years from now. Therefore, the future demands for heat and transportation fuels as well as biomass resources must be taken into account [3], [4].

More specific objectives of the project are to;

- Schematically design new possible heat, electricity, wood pellets and methanol production route(s)

- Identify and quantify all possible constraints that has to be accounted

- Identify required and suitable equipment for biomass pre-treatment, gasification, gas cleaning, purification, reforming, gas mixing and synthesis for the methanol production process.

- Find proper steel work- and biomass gas mixtures for optimal motor fuel production.

- Create a detailed process model of the production route(s) in ASPEN plus.

- Find investments and scale factors for the required process equipment and then make an economic analysis of the production cost.

1.3 The process situation between the companies

SSAB Tunnplåt in Luleå produces raw steel by different qualities. Different saleable by-products are also achieved like benzene, tar, sulphur, argon, combustible process gases etc.

The system is not equipped with a rolling mill at the site, instead the excess gases are burned as primary fuel in the nearby combined heat and power plant (LUKAB) for production of electricity, process steam and district heating. The district heating distributes by Luleå Energi (LEAB). Furthermore, flue gases from the CHP plant are delivered to a wood pellets production plant (Bioenergi) for drying of the wood. Figure 1 shows the present energy balance between the four companies [5].

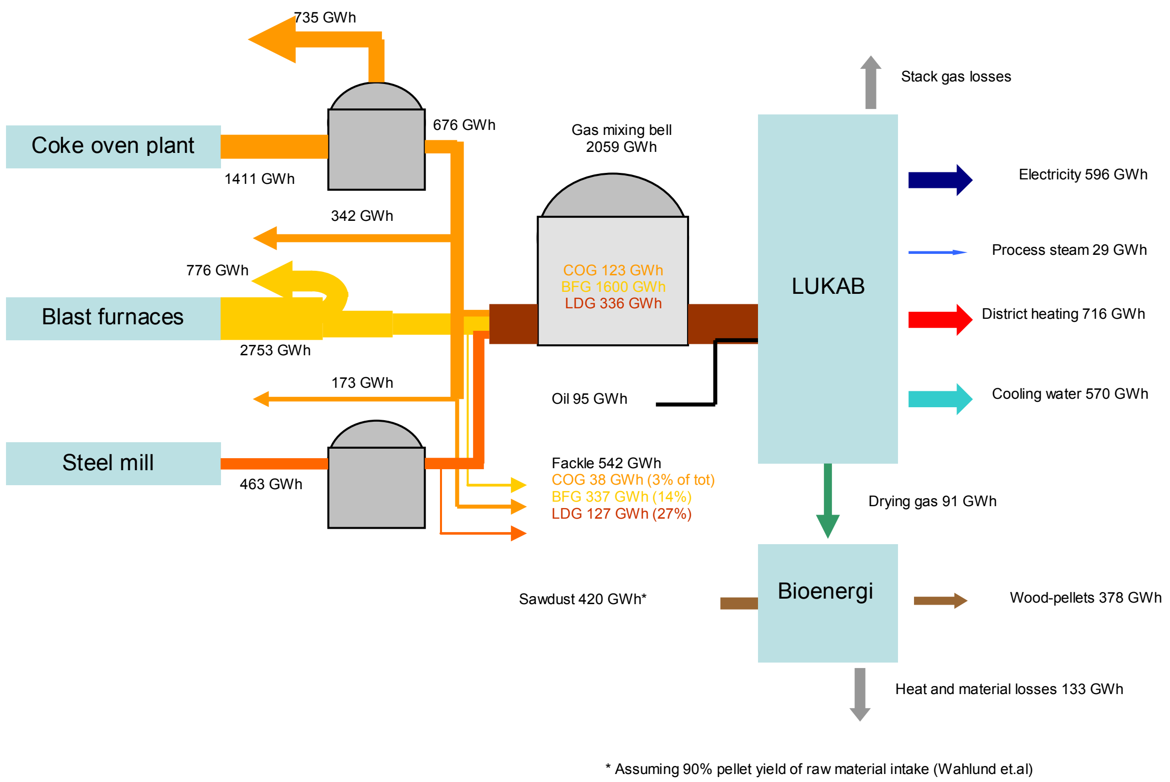

Figure 1 Present energy balance scheme between SSAB, LUKAB, Bioenergi and LEAB [5]

The main idea is to make use of the coke oven gas for MEOH production instead of burning it in the CHP plant. The MEOH plant should be integrated with the CHP plant to deliver excess steam to the existing steam turbine for electricity production. Another idea is to make use of the sawdust that Bioenergi uses for wood-pellets production to produce synthesis gas via gasification. The synthesis gas is then mixed with the coke oven gas to enhance the MEOH production. The system with coke oven gas as fuel input would to some extent be less complex than with both coke oven gas and synthesis gas, since the later system requires a gasification and pre-treatment unit. Though, usually bio refinery plants have a high investment, thus the scale has to be large to be profitable.

A study regarding the future potential energy balances between SSAB, LUKAB, Bioenergi and LEAB, shows that it could be possible to extract coke oven gas for MEOH production in the order of about 400 GWh/year in the year 2025. The extraction will not affect the production capacity at SSAB and LUKAB. Figure 2 shows an energy balance sheet for the year 2025 which this study is based on [6].

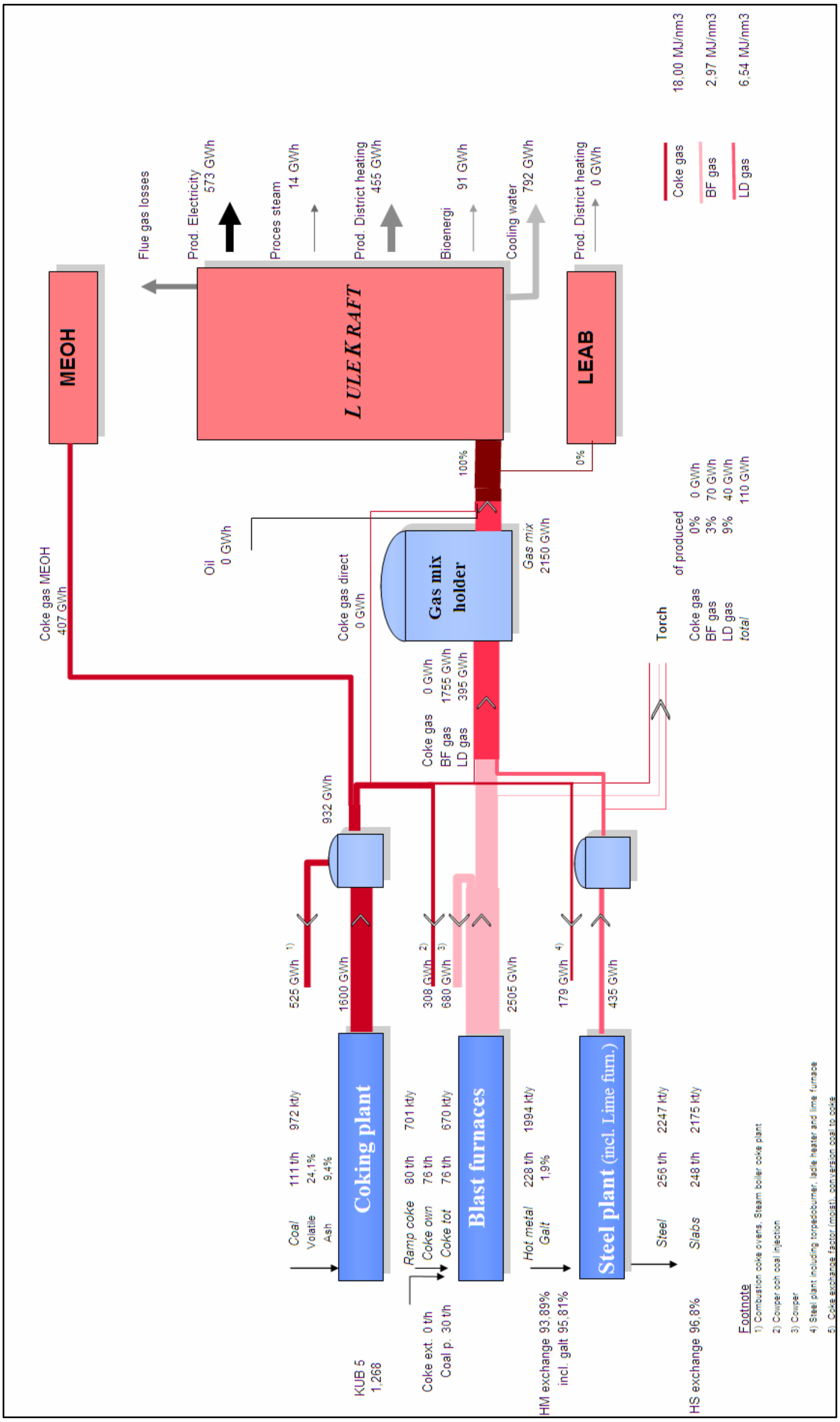

Figure 2 Energy balance scheme between SSAB, LUKAB, Bioenergi and LEAB in the year 2025 [6]

1.4 Methanol in internal combustion engines (ICE)

Gasoline and diesel have been developed together with the engines for more than hundred years. To change the fuel in an ICE may therefore cause problem. Examples of properties of the fuel that may demand a modification in the engine are:

- Octane number (RON – Research Octane Number, MON – Motor Octane Number).
- Cetane number, CN.
- Energy content and stoichiometry (SAFR – Stoichiometric Air Fuel Ratio).
- States of matter (liquid, gas).
- Physical properties such as density, viscosity, surface tension.
- Aggressiveness toward metals and plastics.
- Composition of combustion products.
- The ability to store the fuel.

Typical motor fuel characteristics can be seen in Table 1**.**

Table 1 Fuel characteristic [7]

Fuel	*RON*	*MON*	*CN*	*Density [kg/m³]*	*LHV [MJ/kg]*	*Latent Heat [kJ/kg]*	*SAFR*
Gasoline	92-98	80-90		735	44	305	14,5
Diesel			52	840	43	270	14,6
Ethanol	104	92	8	785	27	840	9,0
Methanol	105	92	3	793	20	1103	6,5

The limiting factor for how much power that can be extracted from a certain engine is how much air that can be brought into the combustion chamber. This is why turbo engines have higher specific power since they compress the air. As you can see from Table 1 methanol has a lower heating value (LHV) than gasoline and diesel while the density is roughly the same. This implies a lower power output than using the same engine running on gasoline. However the air to fuel ratio, SAFR, is lower for methanol which means that not as much air as for gasoline is needed for complete combustion. In fact a smaller increase in specific power is expected with methanol. This is due to the fact that the methanol contains more oxygen per carbon atom than gasoline and diesel [7].

1.4.1 SI-ICE fuels

Spark ignited internal combustion engines, SI-ICE, more commonly known as Otto engines has traditionally mainly been running on gasoline. This section covers how methanol affects these kinds of engines.

Methanol is a toxic alcohol with substantially lower energy content than gasoline. It takes roughly twice the volume of methanol than gasoline for a given distance. The low energy content result in that nozzles and other flow areas in the fuel system need to be increased compared to gasoline in order to maintain the energy flow constant.

Methanol is aggressive towards some metals (especially aluminium) and plastics. Because of this, many components in the fuel system need to be replaced by other materials. The exhaust from methanol combustion contains formic acid that may give rise to corrosion in the engine and exhaust system. Change of lubrication oil in the engine that suits methanol operation reduces risk of damage due to corrosion [7].

Also the steam pressure is substantially lower for methanol than for gasoline. This affects the cold starting ability of the engine and a method to go around this is to blend methanol with gasoline during start up [7].

Methanol affects the composition of the exhaust and gives lower exhaust temperature. Nitrogen oxide emission is lower due to the lower temperature. Formaldehyde emission is higher with methanol than with gasoline operation. Formaldehyde is considered to be carcinogenic. The low exhaust temperature and in combination with the low steam pressure and high latent heat results in that it is very hard to achieve low emissions of hydrocarbons during cold starts [7].

Methanol has higher octane rating than gasoline therefore it is possible to increase the compression pressure ratio of the engine and by this increase the efficiency. An engine optimized for methanol can reduce the energy use by about 10% [7].

Alcohols are unnatural fuels for compression ignited internal combustion engines, CI-ICE or diesel engines, due to their low cetane number. Methanol will simply not ignite unless the engines compression pressure ratio is highly increased or substances that improve the ignition are added. However the interest of using alcohols in diesel engines is large because of the benefits with lower emissions of nitrogen oxides and substantially lower emissions of particulates [7].

2 Technologies

2.1 Methanol production

Methanol is generally produced from coal, natural gas and biomass. This report will be focused on production of methanol from biomass and coke oven gas from a steel work. The biomass resource is sawdust from the sawmill industries around the county. The most widely used biomass to methanol process is done in the following steps: pre-treatment of the biomass, gasification, gas cleaning, methane reforming, H_2/CO shift, acid gas removal, synthesis and purification [8]. A simplified block diagram of the methanol production route can be seen in Figure 3.

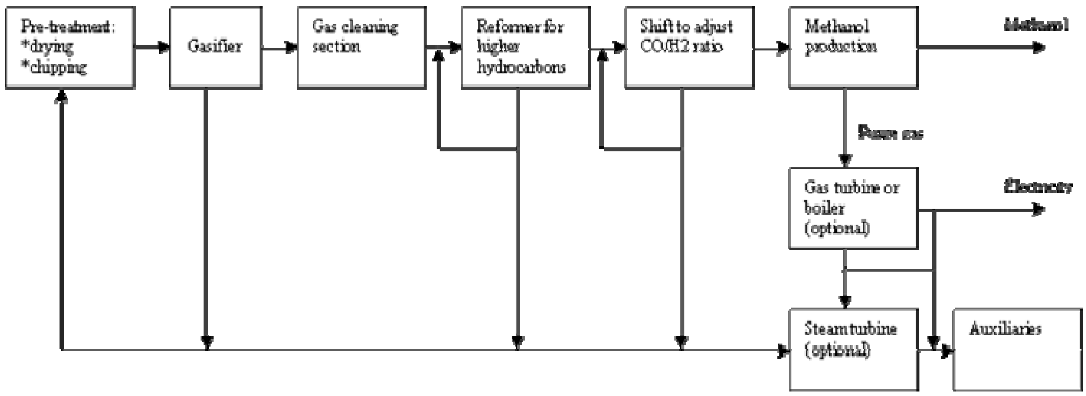

Figure 3 The methanol process [8]

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2.1.1 Pre-treatment of the biomass

The biomass has to be pre-treated before the feedstock can be entered into the gasifier. Mainly one or several of the following steps are used:

- Chipping
- Drying
- Torrefaction
- Powder

2.1.1.1 Chipping

Generally chipping is the first step in the feedstock pre-treatment procedure. The size of the biomass has to be prepared for the selected type of gasifier.

2.1.1.2 Drying

The biomass must be dried to get a moisture content in the range of 10-15% (MC) depending on the gasifier type. The drying process reduces the required heat for gasification and the moisture content of the product gas. The two most frequently ways used for drying the biomass are via steam or flue gas heat exchange. Flue gas could be passed through a HRSG to produce steam, which could be used as drying medium [8]. The specificities for steam or flue gas drying are shown in Table 2**.**

Table 2 Requirements of steam or flue gas drying for biomass feedstock with 50 % MC to 15-10 % MC [9]

	Unit	Steam	Flue gas
Temperature	°C	200	
Pressure	bar	12	
Energy use	MJ/twe	2.8	2.4-3.0
Electricity consumption	kWh/twe	40	40-100

2.1.1.3 Torrefaction

Another possibility of improving biomass properties is torrefacation. Torrefacation means that biomass is slowly heated in an inert atmosphere to a maximum temperature of 300ºC [9]. The treatment will form a product with a lower moisture content and a higher energy content compared to those in the initial biomass. The process may be called mild pyrolysis, with separation of smoke producing compounds and formation of a solid product. The final product retains approximately 70% of the initial weight and 80–90% of the original energy content. The main advantages of torrefacation are that the energy density for the biomass increases, by losing materials with low or no energy density such as water, and the biomass is no longer in danger of losing energy density due to moisture exposure [9]. Torrefied wood is easily packaged and transported, and thus constitutes an efficient fuel. The properties of torrefied biomass should lead to an improved operation in gasifiers for which the stability of the process is important. Torrefied fuel can therefore substitute charcoal and wood in a number of applications.

2.1.1.4 Powder

For the entrained flow gasifier the biomass has to be pulverized to be suitable. The production of biomass powder suitable for entrained flow gasification from different feed stocks is an extra cost, but may be reduced by an initial torrefaction process [11].

2.1.2 Gasification

Gasification is the process that converts biomass to a combustible gas called product gas or syngas. The product gas usually contains 70-80% of the energy originally present in the biomass. Biomass gasification involves two processes: pyrolysis and partial combustion of the char residue (conversion) [1].

Pyrolysis is when heat vaporizes the volatile components of the biomass in the absence of oxygen. The process occurs at a temperature in the range of 450°C to 600°C. Pyrolysis vapour mainly consists of carbon monoxide, hydrogen, methane, carbon dioxide, volatile tar and water. The synthesis gas may also contain some amounts of sulphur, nitrogen and heavier hydrocarbons depending of the fuel input and type of gasifier used. The solid residue of the process consists of charcoal and represents 10-25% wt of total fuel mass [1].

Conversion is when the charcoal residue from the pyrolysis reacts with oxygen and then converts to carbon monoxide. This occurs at temperatures in the range of 700°C to 1200°C [1].

In a combustion process, both these processes occur. When wood burns, the heat from the combustion causes pyrolysis vapour. However these pyrolytic vapours are immediately ignited and burned at temperature of 1500°C to 2000°C. Hence it could be said that gasification is a controlled combustion process [1].

The conversion of charcoal to CO requires the absence of oxygen. As explained by Barrio et al [12] "The oxidising, or gasifying, agents are air, oxygen, steam and CO_2. CO_2 is produced during the pyrolysis and early oxidation processes is generally not externally added. The most common agent is air because of its availability at zero cost. Air, though is not a perfect agent because of its nitrogen content. The product gas from air gasification has generally a low heating value of 4-7 MJ/Nm^3. Oxygen gasification produces a higher heating value (10-18 MJ/Nm^3) but has a drawback due to the high production cost of oxygen. Steam is another alternative. It also generates a medium calorific value gas (10-14 MJ/Nm^3) and moreover increases the hydrogen content of the product gas. Steam gasification is however a highly endothermic reaction and requires a temperature above 800°C to take place if no catalyst is present. The heat required for the reaction has to be transferred either by partial char combustion in the same reactor –mixing H_2O with oxygen/air- or by indirect heating".

2.1.3 Gasification techniques

There are many different gasification systems available on the market. Nevertheless the most common gasifiers for liquid fuel production with biomass as fuel are fluidized bed (FBG), indirect and entrained flow. The focus will be on the first two types due to the dependencies of industrial process integration and economic feasibility [11].

2.1.3.1 Fluidized bed gasifier

There are two common types of FBG gasifiers that could be used for liquid fuel production, Bubbling fluidized bed- and Circulated fluidized bed gasifier.

Fluidized beds normally use a quartz sand bed with a size distribution around 250 µm. Other suitable and active bed materials are dolomite or blast furnace slag which can be used to if they can be supplied in sufficient quantities. The sand bed is the medium that transfers the heat the fuel particles and also increases the mixing and kinetics, which increases the overall gasifier efficiency and fuel throughput [11].

The fluidization agent for liquid fuel production is often oxygen, steam or a mixture of them. The primary fluidization agent is added in the bottom of the bed as fluidizing medium. The velocity of the primary agent is of great importance and thus at least has to reach the minimum fluidizing velocity (MFV), otherwise the fluidization agent will not form any bubbles together with the sand. The MFV occurs when the pressure drop over the sand bed is equal to the total pressure of the bed (i.e. the pressure which the bed exerts on the bottom of the combustor). At MFV, bubbles are formed in the sand and the bed begins to float, this is the Bubbling Fluidized Bed (BFB). If the air velocity is increased above MFV, the bubbles become bigger and will finally erupt quite heavily on the surface of the bed. The bed particles will have a tendency to follow the air up through the furnace and if the air speed is increased even further, some bed material will be transported from the bed and up through the furnace. To recycle the sand a cyclone is installed at the top of the furnace which will allow the syngas to escape through the top, and the sand to be trapped in the bottom and transported back to the bottom of the bed region, this is the circulating fluidized bed (CFB) [11].

A drawback for biomass fuels used as feedstock together with fluidized bed technologies is that, depending of bed material and fuel composition, ash related problems begins to occur at fairly low temperatures. The sand grains in the bed agglomerate since the ashes begin to get sticky or melt. The sand or addition of additives must be exchanged regularly to avoid the bed agglomerate itself by the differential pressure over the bed will drop sharply and the bed defluidize. Otherwise the whole bed can collapse and form a sort of glass which often is very difficult, and takes a lot of effort, to get rid of with economic losses as a result [11].

The fuel specific agglomeration temperature depends on which fuel used. Cleaner fuels with less content of fluxing elements (mostly alkali metals) will increase the agglomeration temperature which of course closely corresponds to the melting point of the ashes. As mentioned previously, controlling these agglomeration problems in commercial facilities is done by decreasing the gasification temperature, more frequent exchange of the sand bed or addition of some proven mineral binding products, e.g. dolomite. The last two will increase the production cost slightly [11].

Bubbling fluidized bed gasifier - BFBG

The fluidisation agent is introduced at the bottom of the reactor while the fuel enters into or above the sand bed. The speed of the fluidisation agent is normally about 2-3 m/s which form bubbles up through the bed Figure 4). The syngas passes through a cyclone at the top of the reactor where the sand and fly ash are separated from the syngas. Typical gasifier data can be seen in Table 3 [11].

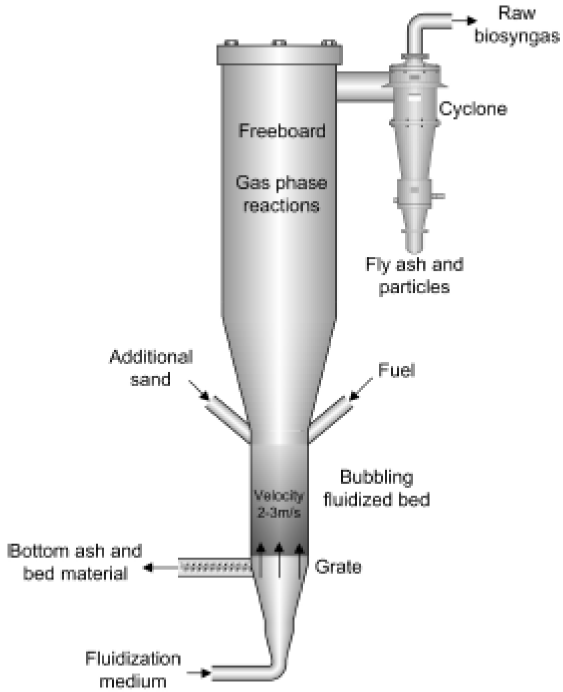

Figure 4 Typical bubbling fluidized bed gasifier [11]

These reactors allow high rates of throughput. The carbon conversion ratio is high because the mixing is good, kinetics is optimised and residence time is long. The syngas has quite low tar content since the good mixing allows gasification temperatures at about 900-950°C, although the tar content is not as low as other concepts like entrained flow gasifiers. The BFBG has a good tolerance level for uneven fuel particles distribution size and the moisture content is also allowed to be higher than for entrained flow gasifiers. There is nevertheless some disadvantage with the BFBG reactor. The technology is somewhat advanced, especially for pressurised operation. There could be a problem with bed agglomeration when biomass is used as fuel. The syngas is also rich in particulates [11].

There is no real technical scale-up limit, but size may be limited by availability of biomass [11].

Table 3 Typical gasifier data [11]

Fuel types	*Wood pellet and woodchips of different size and moisture content*
Scale up limit, dry feed (t/h)	*5 – 180t/day. No real scale up limit, mostly depending on availability of biomass.*
Heating Value (MJ/Nm^3)	*4.5-7.9 (air), 4-6 (Air and steam), 5.5-13 (O_2 and steam)*
Typical gas composition (% volume)	*5-26 H_2, 13-27 CO, 12–40 CO_2, 13-56 N_2, <18 H_2O, 3-11 CH_4*
Tar content of dry syngas (mg/Nm^3)	*13500*
Gasification agent	*Air/Oxygen/Steam/Mix*
Operating pressures (OP, bar)	*1 – 35*
Operating temperatures (°C)	*650 – 950*

Circulated fluidized bed gasifier – CFBG

The fluidisation agent is introduced at the bottom of the reactor while the fuel enters into or above the sand bed. The velocity of the fluidisation agent is normally about 5-10 m/s (Figure 5), which is enough to suspend the bed particles throughout the entire gasifier. The syngas passes through a cyclone at the top of the reactor where the sand and entrained particles are separated from the syngas and recycled back into the bed. Typical gasifier data can be seen in Table 4 [11].

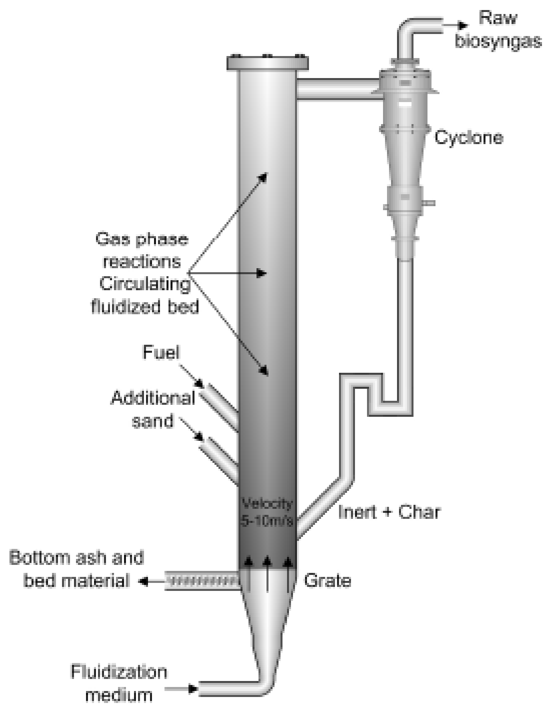

Figure 5 Typical circulated fluidized bed gasifier [11]

CFBG produces a syngas with higher quality and has also higher rates of throughput than BFBG. The CFBG has like BFBG a good mixing, optimized kinetics and long residence time and thus a high carbon conversion ratio. The tar content is also quite low but not as low as other concepts like entrained flow gasifiers. The CFBG has a good tolerance level for uneven fuel particles distribution size. The moisture is also allowed to be a bit higher. There is nevertheless some disadvantage with the CFBG reactor. The technology is somewhat advanced, especially for pressurised operation. High velocities in the reactor may result in equipment. There could be a problem with bed agglomeration when biomass is used as fuel. The syngas is also rich in particulates [11].

The CFBG reactor is for fuel capacities higher than 10 $MW_{thermal}$ and is very fuel flexible [11].

Table 4 Typical gasifier data [11]

Fuel types	*Wood pellet and woodchips of different size and moisture content*
Plant size, dry feed (t/h)	*10 – 110t/day. No real scale up limit, mostly depending on availability of biomass.*
Heating Value (MJ/Nm^3)	*4 - 7 (air)*
Typical gas composition (% volume)	*7-20 H_2, 9-22 CO, 11-16 CO_2, 46-52 N_2, 10-14 H_2O, <9 CH_4*
Tar content of dry syngas (mg/Nm^3)	*Low*
Gasification agent	*Air, oxygen, steam and mitures*
Operating pressures (OP, bar)	*1 - 19*
Operating temperatures (°C)	*800 - 1000*

2.1.3.2 Indirect gasifiers

The indirect gasifier uses gasifying agents which is indirect heated without any combustion of the feedstock. Instead steam can be used for heat transfer between an external heat source and the feedstock material. The indirect heated steam contributes to an environment with very low levels of nitrogen present. The absence of nitrogen results in a syngas with higher energy content [11].

The net reactions are highly endothermic which means that additional heat supply always has to be added to the gasification process in some way. The procedure for the heat supply can be solved with [11]:

- Increase the gasification agent temperature
- Heat the gasification bed-region by an external source; heat pipes, heat exchangers or by physically add heat to the sand-bed or ball-bed by circulating it and heat it externally.

Indirect gasification generally gives higher heating values of the dry synthesis gas than gasification with oxygen, since the methane content is higher. The methane content increases dramatically when using steam rather than air or oxygen, especially at low temperatures 800 - 900°C and high pressures 20 – 60 bar. If the syngas is intended to be used for liquid fuel production the methane content has to be decreased for which a methane reformer could be used. But if the syngas will be used in a gas turbine for electricity production, the methane content must be high to obtain high energy content [11].

Char Indirect, Two-Stage with Steam Reforming

The char indirect system consists of two separate reactors, which are of the type CFBG (Figure 6). They are called the gasifier and the combustor. The hot sand provides the heat necessary to gasify the reactor's feedstock. The sand and residual char is separated via a cyclone from the gasifier to combustor. The char is then burnt in the combustor and the sand is heated. The sand is then passed on to the gasifier again via another cyclone and the loop is closed. Typical data can be seen in Table 5 [11].

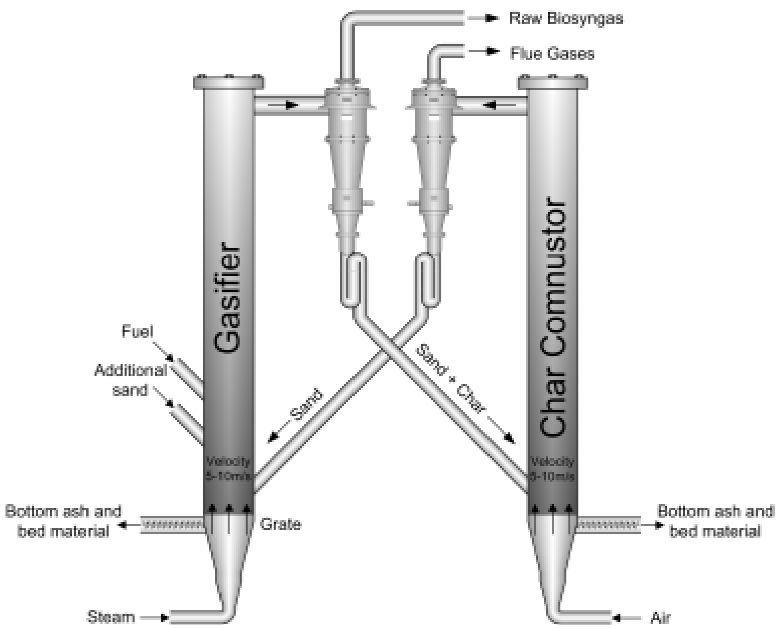

Figure 6 The Char Indirect, Two-Stage gasifier with steam reforming [11]

The advantage with this gasifier is since steam is used in the gasifier and no supply of air is needed in the gasifier the nitrogen level is very low. Air is only used in the combustor from which the flue gases are separated via the cyclone. Because the nitrogen content is low in the syngas the energy content is high. The heat exchange is good by using a twin fluidized bed, which is one of the most interesting gasification systems for large scale biomass power or liquid fuel generation. Nevertheless it has some disadvantages like the construction is complex and thus the investment cost is relatively high which makes it suitable only for larger scales [11].

Table 5 Typical gasifier data [11]

Fuel types	*Wood pellet and woodchips of different size and moisture content*
Plant size, dry feed (t/h)	*No real scale up limit, mostly depending on availability of biomass.*
Heating Value (MJ/Nm^3)	*X (air), X (O_2), X (Steam), 15 (Max H_2), 18 (normal) (Air and steam)*
Typical gas composition (% volume)	*14.9 H_2, 45.6 CO, 14.6 CO_2, 0 N_2, dry H_2O, 17.8 CH_4 (normal)* *24 H_2, 14 CO, 42.7 CO_2, 0.6 N_2, dry H_2O, 14.2 CH_4, 2 C_2H_4 (Max H_2)* *25-35 H_2, 20-30 CO, 15-25 CO_2, 3-5 N_2, dry H_2O, 8-12 CH_4 (Guessing)*
Tar content of dry syngas (mg/Nm^3)	*1500 – 4500 Medium or low (potentially, if proper bed material or fuelmix is used to increase the gasification temperature without aggl. problems)*
Gasification agent	*Air/Oxygen/Steam/Mix*
Operating pressures (OP, bar)	*Atmospheric and pressurized*
Operating temperatures (°C)	*600 - 1000*

2.1.4 Gas cleaning

The quality and pureness of the syngas always differs between different gasifiers and depending on used feedstock. The syngas contains impurities (tar, particles, halogens, alkali metals, S-compounds, N-compounds, heavy metals, calcium etc.) which some have to be reduced and other completely removed. Since these impurities can be harmful to the further refining components in the system [11].

The tar is one of the critical compounds that has to be removed before further treatment of the syngas. The tar is often summarized to all aromatic and polyaromatic compounds above benzene. The tar concentrations and compounds affect the condensation temperature. The tar compounds condensates at different temperatures on inner surfaces or forms aerosols. The aerosols are quite hard to remove by filtration or scrubbing. The tar compounds can further be subdivided into two groups; water soluble (phenolic) and non water soluble (aromatic) [11].

The crucial is not the concentration of total tar but the dew point temperature for the particular tar composition. The heaviest tar compounds will condensate at the highest temperature. Since liquid fuel catalysts generally operate in a temperature range of 220 to 275°C and at 20 to 100 bar for methanol, the dew point of tar have to be reduced to levels below the lowest expected temperature. Then problems with forming aerosol by condensation of tar are solved. The tar compounds can be divided into six five subgroups, see Table 6 [11].

Table 6 Tar classification system [11]

Class	Type	Examples
1	*GC undetectable tars*	*Biomass fragments, the heaviest tars i.e. pitch*
2	*Heterocyclic compounds that generally exhibit high water solubility*	*Phenol, cresol, quinoline, pyridine*
3	*Aromatic components. Light hydrocarbons, which are important from the point of view of tar reaction pathways, but not in particular towards condensation and solubility*	*Toluene, xylenes, ethylbenzene (excluding benzene)*
4	*Light PAHs (2-3 rings), condensate at relatively high concentrations and intermediate temperatures*	*Naphthalene, indene, biphenyl, antracene*
5	*Heavy PAHs (≥4 rings), condensate at relatively low concentrations and high temperatures*	*Fluoranthene, pyrene, crysene*
6	*GC detectable, not identified compounds*	*Unknowns*

Generally only the classes 1 and 5 have to be completely removed and the classes 2 and 4 have to be partly removed to reach a dew point of about 25°C at atmospheric pressure to fulfil the requirements when use high pressures catalyst for liquid fuel production [11].

Methane in the syngas is also of critical importance. The methane level must be held low if the syngas is intended to be used in fuel catalysts. The methane works like nitrogen as inert gas in the catalyst which affects the fuel synthesis process negative. Thus the concentrations must be kept low for obtain high fuel production efficiency. The methane chemical equilibrium concentration is favoured by high pressures (20 – 60 bar), low temperatures and in particular indirect gasification i.e. the use of steam as gasification agent [11].

There are three common cleaning methods, hot, partial hot and cold gas cleaning. In gasification systems, almost exclusively the hot gas high pressure (HTHP) route is preferred to minimise heat losses. The HTHP path is mostly carried out at temperatures above 500°C and pressures between 15-25 bar. I Figure 7 the typical gas cleaning routes for the three systems are shown [11].

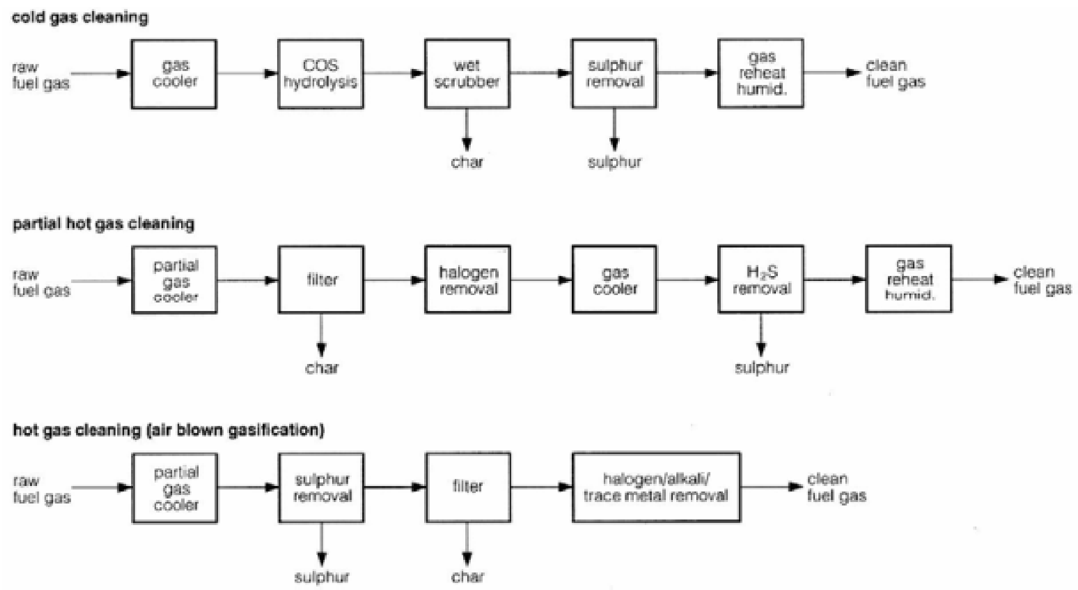

Figure 7 Typical gas cleaning routes [11]

2.1.4.1 Cold gas cleaning

The cold gas cleaning should be used if the operation temperature is below 200°C. The syngas is first cooled and then often followed by COS (carbonyl sulphide) hydrolysis, wet scrubber, sulfur removal and gas reheating. The cold gas cleaning route is the most economical preferable because expensive heat resistant materials does not have to be used. The tar, particles and char usually are removed in a wet scrubber. If the gasifier operates at atmospheric or slightly overpressure and medium to high temperature, the cold gas cleaning could be the choice for fuel synthesis processes. The make-up compressor needs anyway to be internal cooled to reach the proper pressure for methanol synthesis 20-100 bar. If the gasifier operates at high pressure in a bio fuel plant, cold gas cleaning is probably not the best way to go [11].

2.1.4.2 Partial hot gas cleaning

The partial hot gas cleaning is performed at temperatures of 260-540°C. The syngas are first partially cooled to obtain the required operation temperature. The syngas is then passed through a filter for particulate and char removal. Depending on operation temperature usually some type of high temperature fabric filter, ceramic filter or granular bed are required. The halogens (in gas phase) in the syngas are thereafter removed and the syngas is then cooled one more time to finally be cleaned from H_2S. If the gasifier operates at high pressure and medium temperature in a bio fuel plant, the partial hot gas cleaning could be an option. When the gasifier operates at low pressures and high temperature the other two gas cleanings routes are a better choice since the technique is quite expensive [11].

2.1.4.3 Hot gas cleaning

The hot gas cleaning is used for gasification processes where the pressure is high and the temperature range is medium to high (>550°C), due to reduced required pressure and temperature adjustment. The gas is first partially cooled to above 550°C and then the sulfur is removed. The syngas is then passed through ceramic filters or granular beds where the particles and char are removed. Finally gaseous halogens (Cl, F), alkali metals (K, Na) and trace metals are removed from the syngas. This gas cleaning route is preferable if the gasifier operates at high pressure and medium to high temperature because a minimum of temperature and pressure adjustments would be needed downwards in the fuel production process [11].

2.1.5 Gas processing

The syngas generally need to be processed to reach the requirements of the fuel catalysts for production of liquid fuels. The syngas is processed to minimize the presence of inert gases like CH_4, CO_2 and N_2, thereby increase the efficiencies of fuel catalysts. The right proportion of the H_2/CO-ratio also needs to be optimized before entering the fuel catalyst. Depending of used fuel input and gasifier some or all of the following methods are generally needed to reach the required syngas composition before entering the catalyst. Though, since every step also will make the system more expensive the increased efficiency is versus increased investment cost. Thus some parts may not be economical usable in the end. Some steps are only applicable in large plants [11].

2.1.5.1 Methane reforming

Methane acts as an inert in the catalyst, hence affects negative to the fuel production efficiency. The methane content in the syngas differs depending the used fuel input and gasifier. Generally the fuel catalyst can handle content up to 2% vol. in the syngas but the content can often be as high as 5-10% vol. in the raw syngas. The reforming process lets methane reacts with either steam or oxygen to form mixtures of CO_2, CO and H_2. The most frequently used routes are steam and oxygen or autothermal reforming [11].

Steam methane reforming (SMR)

The steam forming process uses steam for reaction with methane to form CO and H_2, generally over a nickel based catalyst, see Figure 8. Steam reforming is the most common method for production of syngas from natural gas or gasifier gas. The reactions that occur are [11]:

$$CH_4 + H_2O \Rightarrow CO + 3H_2 \qquad \Delta h^\circ_f = 206 \text{ MJ/kmol} \qquad \text{(R1)}$$
$$C_2H_4 + 2H_2O \Rightarrow 2CO + 4H_2 \qquad \Delta h^\circ_f = 210 \text{ MJ/kmol} \qquad \text{(R2)}$$
$$C_2H_6 + 2H_2O \Rightarrow 2CO + 5H_2 \qquad \Delta h^\circ_f = 347 \text{ MJ/kmol} \qquad \text{(R3)}$$
$$CO + H_2O \Rightarrow CO_2 + H_2 \qquad \Delta h^\circ_f = -41 \text{ MJ/kmol} \qquad \text{(R4)}$$

The reforming reaction (R1) is highly endothermic thus an external heat source is required. The inlet temperature and pressure is typically between 450-650°C and 20-30 bar respectively. The steam reformer is suitable to use in combination with pressurized gasification and hot gas cleaning. The normal outlet temperature from the reformer is approximately 700-950°C. A disadvantage with this kind of reformer can be the need of an external heat source and the investment cost [11].

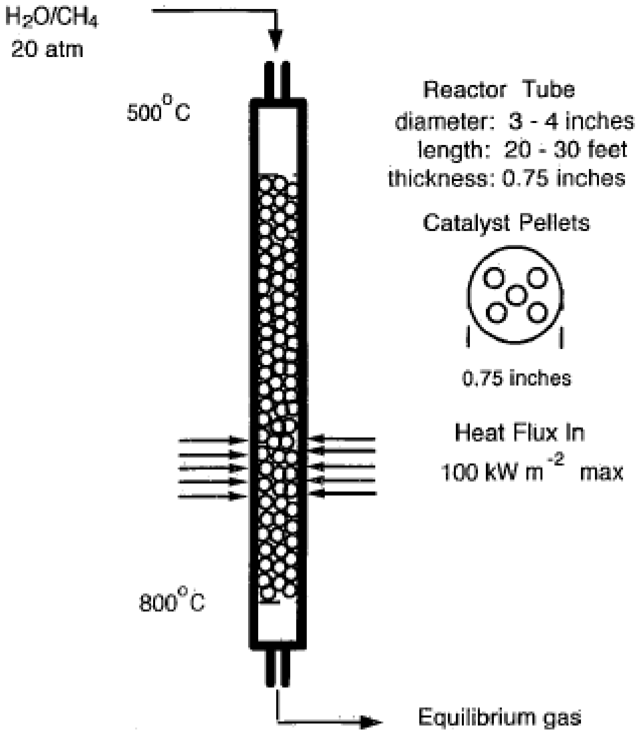

Figure 8 Steam methane reforming [11]

Oxygen- or autothermal reforming (ATR)

Autothermal reforming uses oxygen to react with methane and other hydrocarbons over a rhodium based catalyst. The methane is partly oxidised in the presence of oxygen when H_2 and CO is formed. The reactions that occur are [13]:

$$CH_4 + {}^3/_2O_2 \Rightarrow CO + 2H_2O \qquad \Delta h°_f = -519 \text{ MJ/kmol} \qquad \text{(R5)}$$
$$CH_4 + H_2O \Rightarrow CO + 3H_2 \qquad \Delta h°_f = 206 \text{ MJ/kmol} \qquad \text{(R6)}$$
$$CO + H_2O \Rightarrow CO_2 + H_2 \qquad \Delta h°_f = -41 \text{ MJ/kmol} \qquad \text{(R7)}$$

Even reactions with heavier hydro carbons occur. Since reaction R5 is highly exothermic and occur at very high temperatures (~1250°C), huge amounts of heat is produced. Thus process integration with other heat consumers would be beneficial. A disadvantage with this method is that an oxygen plant is needed which is expensive [11].

2.1.5.2 Water gas shift (WGS)

The WGS is implemented in the system to adjust the H_2/CO ratio before the syngas is passed on to the fuel catalyst step, see Figure 9. Depending of which fuel produced the required ratio differs, for methanol production the ratio should theoretically be about 2:1. The WGS separates the H_2 and CO_2 out of the syngas by a membrane and steam. Generally a by pass stream is used where the separated H_2 is fed into to get the right ratio. The reaction is exothermic and is described by R8 [11]:

$$CO + H_2O \Rightarrow CO_2 + H_2 \qquad \Delta h°_f = -41 \text{ MJ/kmol} \qquad \text{(R8)}$$

The WGS process is normally performed at 20 bar and 400-600°C or even up to 900°C depending on the membrane and reactor walls [11].

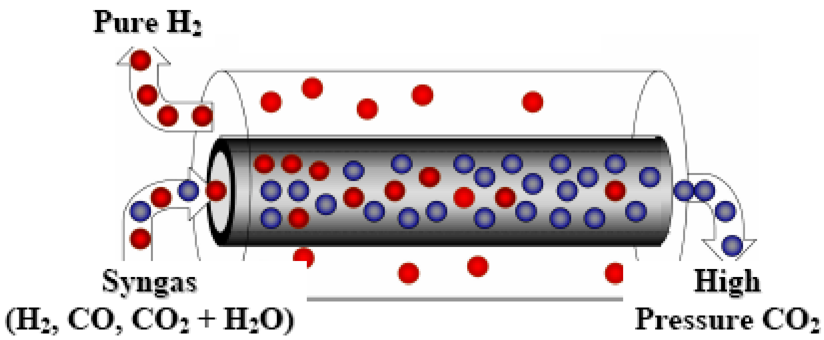

Figure 9 Water gas shift reactor [11]

2.1.5.3 CO_2 removal

Carbon dioxide is an inert gas which can be accepted in certain amount in the syngas. For methanol production with a low temperature and low pressure (LTLP) system, the catalyst should have 2-8 vol% CO_2 for best methanol activity and selectivity. If the amount increases over 8 vol% the synthesis will be inhibited. Thus lack of CO_2 makes the synthesis reaction rapidly slower. The MEOH synthesis is about 100 times slower in pure H_2/CO mix than when CO_2 is present in right proportion. The most widely used methods for CO_2 removal are: chemical and physical absorption, solid physical adsorption, cryogenic separation and the use of membrane systems. Since the chemical and physical absorption are widely applied and at the present the most suitable for application to a broad range of CO_2 containing streams the focus is applied on those [11].

Chemical absorption

Chemical absorption uses chemical solvents to let CO_2 reacts and form an intermediate compound with weak bonds. Additional heat is provided to break down the compound for production of a CO_2 stream, original solvent is regenerated as well. Typical solvents are based on amine or carbonate such as MEA, diethanolamine (DEA), ammonia and hot potassium carbonate. The most widely used absorption processes are MEA and activated potassium carbonate. The technique has though some restrictions, it can be applied at low CO_2 partial pressures but the syngas stream must be free of SO_2 and O_2. Hydrocarbons and particulates will also cause operating problems in the absorber [11].

Physical absorption

Physical absorption uses also a solvent to absorb CO_2 from the syngas and the regeneration occurs by either adding heat and/or pressure reduction. The typical solvents used are Selexol (dimethylether of polyethylene glycol) and Rectisol (cold methanol) which is frequently specified for coal gasification. Both are applied at high pressure. If the CO_2 capture process is intended to run at lower pressures, the chemical absorption processes are more economical. A drawback with the technique is that heavier hydrocarbons (C_3+) are soluble in the physical solvent and will thereby cause operating problems [11].

2.1.6 Synthesis and catalysts

In chemistry, synthesis is defined as the formation of a compound from simpler compounds or elements.

Traditionally methanol has been produced by synthesis of syngas derived from gasified coal or natural gas. Syngas can also be produced from biomass fuels via a gasification process as explained earlier [11].

Typically syngas consists of H_2 and CO which forms to methanol according to reaction (R9). The composition differs depending on the gasification process and feedstock. If more H_2 than necessary is present in the syngas it is possible to inject CO_2, which would allow methanol formation according to reaction (R10). If there is a lack of hydrogen for stoichiometric methanol formation, the CO/H_2 ratio may be adjusted in a water gas shift reactor prior to the synthesis according to (R8) [11].

$$CO + 2H_2 \Rightarrow CH_3OH \qquad \Delta h°_f = -90.6 \text{ MJ/kmol} \qquad \text{(R9)}$$
$$CO_2 + 3H_2 \Rightarrow CH_3OH + H_2O \qquad \Delta h°_f = -49.5 \text{ MJ/kmol} \qquad \text{(R10)}$$

Several catalyst have been reported and tried out, but can generally be classified as high temperature and high pressure (HTHP) or LTLP synthesis. The HTHP synthesis is at present not usually is applied, since the LTLP synthesis is more economic. Thus the focus will be on the LTLP synthesis [11].

2.1.6.1 Low temperature and Low pressure MEOH synthesis

Conventional LTLP methanol synthesis reactors can be divided into adiabatic (quench cooled) reactors and isothermal (indirectly cooled) reactors. The reactors generally use fixed beds of catalyst pellets and operate in the gas phase. Two reactor types predominate in plants built after 1970. The ICI low pressure process is an adiabatic reactor with cold unreacted gas injected between the catalyst beds. The consequent heating and cooling decreases its potential efficiency, but the reactor is very reliable and therefore still predominant. The Lurgi system have the catalyst loaded into tubes and a cooling medium circulating on the outside of the tubes, allows near-isothermal operation [8].

The methanol synthesis temperature is typically performed between 230 and 270°C and the pressure is between 50 and 150 bar. Higher pressures give economical benefit, since the equilibrium then favours methanol conversion efficiency. Only a portion of the CO in the feed gas is converted to methanol in one pass through the reactor, due to the low temperature at which the catalyst operates. The unreacted gas is recycled at a ratio typically between 2.3 and 6. The copper catalyst is poisoned by both sulphur and chlorine but the presence of free zinc oxides does help prevent poisoning [8].

ICI low pressure quench converter

This is the most widely used adiabatic methanol converter where the catalyst is contained in a single bed, see Figure 10. By adding a mixture of fresh and recycled syngas into the reactor and quenches the reaction and temperature is controlled. The syngas is injected at appropriate depths within the reactor through spargers called lozenges. There are horizontal layers of these lozenges that run across the converter from side to side and each has an outer surface covered with wire mesh and a central pipe that delivers the cold gas. The used catalyst is $Cu/ZnO/Al_2O_3$ + promoters, where the life time is 36-48 months. The operating pressure and temperature are 50-100 bar and 230-265°C respectively [11].

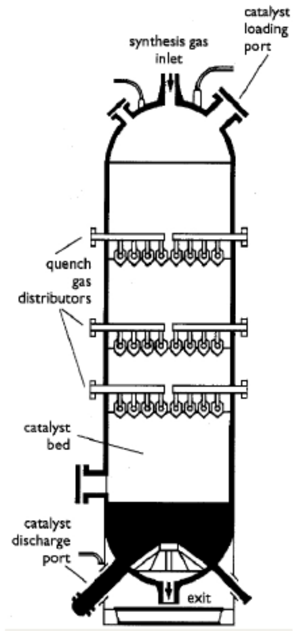

Figure 10 ICI low pressure quench converter [11]

Lurgi methanol converter

The Lurgi Methanol Converter is one of the more widely used commercial isothermal methanol converters, see Figure 11. Several solutions is also patented, e.g. the catalyst. The reactor has a shell tube design. The tubes contain the copper based catalyst and are surrounded by cold water for heat removal. The near isothermal temperature is controlled by adjusting the water pressure and thereby the boiling temperature. Steam at 40-50 bar is produced as a by-product by the cooling water which can be used to run the catalyst reactor compressor or as heat supply for the distillation process. The used catalyst is $Cu/ZnO/Cr_2O_3$ where the life time is 36-48 months. The operating pressure and temperature are 50-100 bar and 270°C respectively [11].

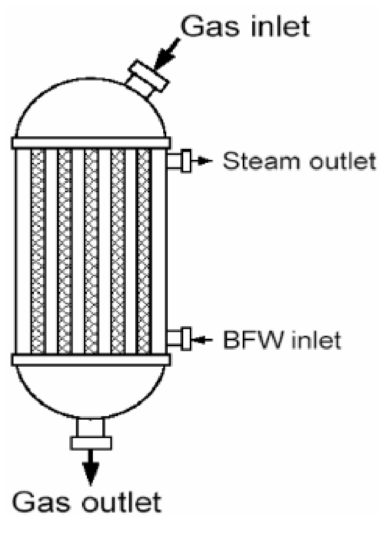

Figure 11 Lurgi shell tube methanol converter [11]

2.1.7 Purification

The last step in the methanol production process is the purification. Usually the crude methanol is first filtered to remove traces of wax and then sent to a topping column followed by a refining column. Lighter impurities are removed in the topping column such as CO, N_2 and aldehydes. Water and higher alcohols are removed in the refining column. Grade AA refined methanol has a purity of 99% [14].

The two distillation column system is recognized as using relatively much process steam. Multistage column arrangements requires less steam also slightly higher methanol yield is obtained, though it is related with higher investments cost and are only justified for very large plants [14].

Single column distillation is possible for fuel-grade methanol (as a blending component), where the objectives only are to remove dissolved gases and water [14].

3 Selected systems

The selected systems are essentially based on two types used in the Faaij et al. report (5 and 6 for MEOH production) [8]. The systems are based on equipment technologies available today and adapted to the conditions concerning the type of fuel input. Four systems are considered, two that use biomass as well as coke oven gas in a mixture and the other two use coke oven gas. The main components in the systems are:

- **IGT system:** IGT pressurised oxygen blown gasifier, hot gas cleaning, autothermal methane reforming, partial H_2/CO shift, conventional solid bed MEOH catalyst with syngas recycling and power generation via steam cycle.

- **BCL system:** BCL atmospheric air blown gasifier, cold gas cleaning, steam methane reforming, conventional solid bed MEOH catalyst with syngas recycling and power generation via steam cycle.

- **ATR system:** Hot gas cleaning, autothermal methane reforming, conventional solid bed MEOH catalyst with syngas recycling and power generation via steam cycle.

- **SMR system:** Cold gas cleaning, steam methane reforming, conventional solid bed MEOH catalyst with syngas recycling and power generation via steam cycle.

The two gasifiers selected are the IGT (Institute of Gas Technology) pressurised direct oxygen fired gasifier, and the BCL (Battelle Columbus) atmospheric indirectly fired gasifier. Both gasifiers produce a medium calorific gas, undiluted by atmospheric nitrogen [8].

The IGT gasifier produces a CO_2 rich gas. The CH_4 fraction will be reformed to hydrogen. The H_2/CO ratio (1.4:1) is attractive to produce methanol, although the large CO_2 content lowers the overall yield of methanol. The pressurised gasification allows a large throughput per reactor volume and reduces the need for pressurisation downstream, so less overall power is needed. However, the gasifier efficiency is lower and much more steam is needed. The IGT gasifier uses oxygen as gasification agent to reduce downstream equipment size [8].

The indirectly heated BCL gasifier uses air as gasification agent with no risk of nitrogen dilution. It produces a gas with a low CO_2 content, but contains heavier hydrocarbons. Therefore, reforming is a logical subsequent step in order to maximise CO and H_2 production. The tars present need to be cracked and the large CO fraction normally needs to be shifted to yield hydrogen. The reactor is fast fluidised, allowing throughputs equal to the bubbling fluidised IGT, regardless of the atmospheric operation [8].

In Figure 12 the key characteristics of the selected gasifiers are shown [8].

	IGT bubbling fluidised bed		BCL Indirectly heated fast fluidised bed	
Initial moisture content (%)	30		30	
Dry moisture content (%)	15		10	
steam (kg/kg dry feed)	0.3		0.019	
oxygen (kg/kg dry feed)	0.3		0	
air (kg/kg dry feed)	0		2.06	
Product temperature (°C)	982		863	
exit pressure (bar)	34.5		1.2	
gas yield (kmol/dry tonne)	82.0		45.8	
composition: mole fraction on wet basis (on dry basis)				
H_2O	0.318	(-)	0.199	(-)
H_2	0.208	(0.305)	0.167	(0.208)
CO	0.15	(0.22)	0.371	(0.463)
CO_2	0.239	(0.35)	0.089	(0.111)
CH_4	0.0819	(0.12)	0.126	(0.157)
C_2H_4	0.0031	(0.005)	0.042	(0.052)
C_2H_6	0		0.006	(0.0074)
O_2	0		0	
N_2	0		0	
LHV_{wet} syngas (MJ/Nm^3)	6.70		12.7	
Cold gas efficiency (%)	HHV 82.2 / LHV 78.1		HHV 80.5 / LHV 82.5	

Figure 12 Key characteristics of the IGT- and BCL gasifier [8]

4 Modelling and calculations

The selected systems are modelled in Aspen plus, a widely used process simulation software. Components as chemical reactors, pumps, turbines, heat exchanging apparatus, etc are virtually connected by pipes. Each component is specified in detail like reactions taking place, efficiencies, dimension of heating surfaces and so on. For given inputs, product streams can be calculated, or one can evaluate the influence of apparatus adjustments on electrical output. The plant efficiency is optimised by integrating the heat supply and demand. The resulting dimensions of streams, units and the energy balances are subsequently used for economical analysis.

Because biomass gasification temperatures are relatively low, significant departures from equilibrium are found in the product gas. Kinetic gasifier modelling would be complex and different for each reactor type which also was confirmed by own tryouts in Aspen plus. The syngas produced by the different gasifiers contain various contaminants: particulates, condensable tars, alkali compounds, H_2S, HCl, NH_3, HCN and COS. No full data sets of syngas compositions including these contaminants are available for the gasifiers considered [8].

The pre-treatment and gasification sections are not modelled but their energy use and conversion efficiencies are included in the energy balances. The models start with the synthesis gas composition from the gasifiers as given in Figure 12. The gas cleaning section in the models is considerable simplified since it is complex to model and no data of particulates, condensable tars and alkali compounds are available. The purge gas separated out from MEOH separator is not further used. The purge gas could be used as recycle gas into the reformer for higher methanol conversion efficiency or as fuel to LUKAB. Modelling assumptions for the process units are given in Appendix A [8].

The coke oven gas composition is showed in Table 7. The oxygen content is normally closer to zero, since there is a risk of explosion due to the high H_2 and CH_4 content [4].

Table 7 Coke oven gas composition by volume [4]

Component	Coke gas vol%
CH4	23,3
CO	5,4
CO2	1,3
C2H6	3,0
H2	56,7
N2	5,8
O2	0,3
H2O	2,8
H2S	1,2
COS	0,2
Sum	100,0

The general properties for the methanol plants are listed below.

Load hours = 8000h/a [8]
LHV for methanol = 21 GJ/tonne [7]
Density methanol = 793 kg/m^3 [7]
LHV dry sawdust = 17,1 GJ/tonne [15]
LHV wood pellets = 20 GJ/tonne [15]
LHV coke oven gas = 17,5 MJ/Nm^3 [4]
Amount of sawdust = 420 GWh/year [3]
Amount of coke oven gas = 400 GWh/year [4]

The sawdust is dried from a moisture content of 30% to 10-15% depending on the gasifier type. Drying uses roughly 10% of the energy content of the feedstock in form of steam. The drier has also a specific electricity consumption of 40 kWh/tonne water evaporated (twe) [8].

The steam delivered to the steam turbine at LUKAB has a steam/electricity factor of 0,887 at full load and back pressure production. The district heating/electricity factor is assumed to be 1,22 at full load and back pressure production. An extraction point at 12 bar is at present used to deliver process steam to SSAB. From that point also steam for the sawdust drier could be extracted [16].

The maximum methanol conversion per pass through the catalyst is 20% when the temperature and pressure is about 260°C and 100 bar respectively, see Figure 13**.** Thus a recycle stream has to be built up in the catalyst to increase the output of MEOH. Normally a recycle to feed ratio of 4-5 is used [11].

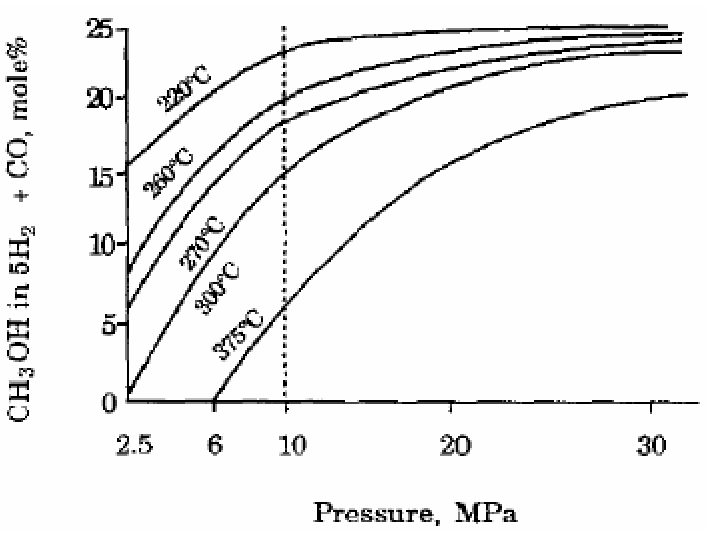

Figure 13 Equilibrium concentrations of methanol depending on pressure and temperature [11]

4.1 IGT system

The IGT system is based on mixing syngas produced from gasified sawdust with coke oven gas from the steel mill, see Figure 34 in Appendix D.1 for the process scheme. The coke oven gas is compressed and mixed with the syngas before the gas cleaning unit. After the syngas is cleaned it preheats before entering the reformer (ATR1) by passing a heat exchanger after the reformer. In ATR1 a partial combustion occur (mainly R5) to keep the temperature high for the highly endothermic CH_4 and steam reaction (R6), which take place in ATR2 [8]. The WGS reaction (R8) is also present in ATR2 [8]. When the syngas is reformed the H_2/CO ratio is still too low thus a split stream is passed a WGS reactor where H_2 and steam forms CO and CO_2 via the WGS reaction (R8) [8].

The CO_2 removal unit removes as much CO_2 necessary but no more than 98 vol% to get the R-value or $(H_2 - CO_2) / (CO + CO_2)$ ratio about 2.1 [8]. The CO_2 content is kept as low as possible to avoid that CO_2 acts as considerable inert gas later in the catalyst [8]. Still some presence of the CO_2 is wanted since CO_2 act as a promoter and the MEOH reaction with CO_2 and H_2 (R10) will increase the output of MEOH [8]. The syngas is then compressed and passed on to the catalyst unit where reaction R9 and R10 occurs [8]. The syngas is separated into two streams where one stream is preheated before entering the catalyst by passing a heat exchanger after the catalyst. Since the conversion efficiency is thermal limited to about 20% per passage, a recycle stream is built up, see Figure 13. The crude MEOH is then separated to a distillation column for purification.

The steam used by the gasifier, reformer and shift reactor is produced via a stream passing through heat exchangers after and before the shift reactor. Steam raised in the high temperature heat exchangers is passed through a steam turbine assumed to be the one at LUKAB. All steam is extracted at the pressure 12 bar to supply the sawdust drier where the steam will be completely condensed [8]. The reminding hot tempered water is summarized as potential district heating.

4.2 BCL system

The BCL system is as the IGT system based on mixing syngas produced from the gasified sawdust with the coke oven gas from the steel mill, see Figure 35 in Appendix D.2 for the process scheme. The coke oven gas is mixed with the syngas before the gas cleaning unit. After the syngas is cleaned it splits into two streams. One stream is used as fuel in SMR1 were complete combustion is achieved to supply the heat demand for SMR2 [8]. The other stream is compressed and passed on to SMR2 where primarily the reaction R1-R4 occurs [8]. Before the stream enters SMR2 it preheats by passing a heat exchanger after SMR2. When the syngas is reformed the H_2/CO ratio is still too high thus CO_2 recycling from CO_2 removal unit downstream is applied. The WGS reaction (R8) will go against left side and produce more CO and steam. Some H_2 is then consumed but still higher MEOH conversion ratio will be reached later in the catalyst. The required H_2/CO ratio is achieved thus no shift reactor unit is necessary.

The following steps are equal performed as in the IGT system [8]. The steam used by the reformer and a part of the drier is produced via a stream passing through heat exchangers after SMR2 and SMR1. The amount of steam required for the gasifier is small and therefore just heated with a heater. In reality the amount should be extracted from one of the other streams or from the steam turbine. Steam raised in the high temperature heat exchangers is passed through a steam turbine assumed to be the one at LUKAB. All steam is extracted at the pressure 12 bar to supply the sawdust drier where the steam will be completely condensed [8]. The reminding hot tempered water is summarized as potential district heating.

4.3 ATR system

The ATR system is based on using coke oven gas from the steel mill, see Figure 36 in Appendix D.3 for the process scheme. The coke oven gas is compressed and passed trough the gas cleaning unit. After the syngas is cleaned it preheats before entering the reformer (ATR1) by passing a heat exchanger after the reformer as for the IGT system [8]. When the syngas is reformed the H_2/CO ratio is still too high thus CO_2 recycling from CO_2 removal unit downstream is applied as for the BCL system.

The following steps are equal performed as in the BCL system. The amount of steam required for the reformer is produced via a stream passing through heat exchangers after the reformer. Steam raised in the high temperature heat exchangers is passed through a steam turbine assumed to be the one at LUKAB. The turbine is not modelled because the steam/electricity- and district heating/electricity factor is used [16]. The reminding hot tempered water is summarized as potential district heating.

4.4 SMR system

The SMR system is based on using coke oven gas from the steel mill as in the ATR system, see Figure 37 in Appendix D.4 for the process scheme. The following steps are equal performed as in the BCL system. The amount of steam required for the reformer is produced via a stream passing through heat exchangers after the reformer. Steam raised in the high temperature heat exchangers is passed through a steam turbine assumed to be the one at LUKAB. The turbine is not modelled because the steam/electricity- and district heating/electricity factor is used [16]. The reminding hot tempered water is summarized as potential district heating.

5 Economic calculations

The resulting dimensions of streams, units and the energy balances in the Aspen models are used for the economic analysis. The production cost for the methanol plants are calculated by hand with the software Microsoft Excel.

5.1 Methanol plant

All equipment costs and scale factors for the economic evaluations is presented in Figure 29 in Appendix B. The equipment is scaled with Equation 1 [8].

$$\frac{\mathrm{Cost}_{\mathrm{new.size}}}{\mathrm{Cost}_{\mathrm{ref.size}}} = \left(\frac{\mathrm{Size}_{\mathrm{new}}}{\mathrm{Size}_{\mathrm{ref}}}\right)^{R} \qquad (1)$$

In Equation 2 the R stands for scale factor, size is in existing unit and the specific equipment cost is in MUS$. The specific equipment cost is also adjusted with consumer price index for the year 2006 [19]. The annual capital cost for the installed units are calculated with the annuity method, Equation 2 [8].

$$I_{\mathrm{annual}} = \frac{IR}{1-\left(\frac{1}{1+IR}\right)^{t_e}} \times I_t \cdot \left(1 - \frac{1}{(1+IR)^{t_e}} \cdot \frac{t_t - t_e}{t_t}\right) \qquad (2)$$

I_{annual} = Annual investment cost
IR = Interest rate = 10 % [8]
I_t = Total initial investment (sum of unit investment)
t_e = Economical lifetime = 15 years [8]
t_t = Technical lifetime = 25 years [8]

5.1.1 Capital cost

The cost for the installed equipment includes auxiliary equipment, non-equipment, engineering and contingencies. Installation factors that are not given by literature are calculated by following numbers [8].

- 33 % investment is added to hardware (instrumentation and control 5 %, buildings 1.5 % grid connections 5%, site preparation 0.5 %, civil works 10 %, electronics 7 %, and piping 4 %)
- 40 % installation costs to investment (engineering 5 %, building interest 10 %, project contingency 10 %, fees/overheads/profits 10 %, start-up costs 5 %).

This gives a resulting overall installation factor of 1.86 which is multiplied with the investment of the specific equipment.

5.1.2 Operating and maintenance

Operator Maintenance cost = 4% of Total Plant Investment (TPI) [8]
The biomass cost is assumed to 100 SEK/MWh (2006) [17]
The coke oven gas cost is assumed to 28,4 SEK/GJ (2006) [4]
Electricity price is assumed to 500 SEK/MWh (2006) [18]

Consumer prices for $ 2006/2001 = 1,138 (2006) [19]
Currency = 1US$ = 7,38 SEK (2006) [20]
Currency = 1€ = 9,25 SEK (2006) [21]

6 Results

For given inputs, product streams are calculated. The plant efficiency is optimised by integrating the heat supply and demand. The resulting dimensions of streams, units and the energy balances are subsequently used for the economic analysis. Since the district heating demand for the community already is fulfilled by LUKAB the economic result is presented without sold district heating. The potential extraction and income for district heating will anyway be presented separately for each system.

The methanol production cost is influenced by variables such as capital, operating and maintenance cost. In order to assess the influence the different variables have on the methanol production cost, a sensitivity analysis is made. The potential maximum and minimum values of the input parameters are estimated. The variables used in the sensitivity analysis are shown in Table 8 with attached minimum and maximum values. The analysis was performed by varying one factor while keeping all other factors constant in the reference case. Then a best and worst case was created by putting all variables in min- and max position.

Table 8 Sensitivity analysis variables

Variable	Min	Ref	Max		
Biomass price	70	100	130	**SEK/MWh**	(-/+ 30%)
Coke oven gas price	24,14	28,4	32,66	**SEK/GJ**	(-/+ 15%)
Electricity price	450	500	650	**SEK/MWh**	Own choice
TOT Investment	-30	0	30	**%**	Own choice
Internal rate	6	10	12	**%**	Own choice
Economical lifetime	20	15	10	**years**	Own choice

These variables are assumed to be those that have the most influence on the final methanol production cost and therefore have to be accounted for. The maximum and minimum values chosen are explained and listed below.

- The maximum- and minimum cost for the biomass is chosen from the fact that the price for logging residue normally is in the range of 70-130 SEK/MWh, including taxes [17].
- Coke oven gas price difference by +/- 15% is an own assumption.
- Electricity price difference is based on statistics [18]
- Total installed investment differs +/- 30%, which is a normal limit to be kept within [8].
- The minimum value of 7% for the internal rate is common for industrial investment in Sweden. The higher value is chosen to reach a higher value than the reference case.
- Economic lifetime difference is an own assumption.

6.1 IGT System

The IGT system results in a MEOH conversion efficiency of 53,6% (based on LHV) and a total plant efficiency of 63,9%, see Table 9 for a short energy balance.

Table 9 Energy balance for the IGT system

Energy input		
Coke oven gas	MW	-50,0
Biomass	MW	-52,5
Internal el. use	MW	-7,7
Steam raise	MW	-1,7
Sum	MW	-111,9
Energy output		
Methanol	MW	60,0
El. production	MW	2,0
District heating	MW	5,1
Purge gas	MW	4,4
Sum	MW	71,4
MEOH conversion efficiency	%	53,6
Total plant efficiency	%	63,9

The total efficiency is the lowest of the modelled systems because the steam and electricity demand is higher and MEOH conversion efficiency is the lowest. The autothermal reformer consumes more syngas via the partial combustion than the steam reformer due to the higher operating temperature. This leads then to an increased production of high tempered steam for electricity production. But since the electrical conversion efficiency is low and the internal electricity demand is high the total plant efficiency is lower than for the other systems. Still the production cost of MEOH is comparable to the other systems since the investment is lower than for the BCL system. For a complete energy balance see Figure 30 in Appendix C.1. All stream results are shown in Table 41 in Appendix D.1. Table 10 presents a summary of the key figures for the IGT system, see also Table 45 in Appendix E.1 for detailed economical calculations.

Table 10 Key figures for the IGT system

Investments	Unit	Value
Total capital requirement	MUS$	85
Total annual cost	MUS$/year	27,8
Methanol produced	m^3/year	103707
Methanol production cost	US$/litre	0,27
Methanol production cost	SEK/litre	1,98

The sensitivity analysis by varying one factor while keeping all other factors constant in the reference case showed that the investment followed by the internal rate was the most crucial factors that influenced the MEOH production cost. In Table 11 the selected factors with respective MEOH production cost are presented.

Table 11 Sensitivity analysis results for the IGT system

Variable	Min	Ref	Max	
Biomass price	70	100	130	**SEK/MWh**
MEOH prod. cost	**1,86**	**1,98**	**2,10**	**SEK/litre**
Coke oven gas price	24,14	28,40	32,66	**SEK/GJ**
MEOH prod. cost	**1,92**	**1,98**	**2,04**	**SEK/litre**
Electricity price	450	500	650	**SEK/MWh**
MEOH prod. cost	**1,96**	**1,98**	**2,04**	**SEK/litre**
TOT Investment	-30	0	30	**%**
MEOH prod. cost	**1,69**	**1,98**	**2,27**	**SEK/litre**
Internal rate	6	10	12	**%**
MEOH prod. cost	**1,78**	**1,98**	**2,08**	**SEK/litre**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,95**	**1,98**	**2,02**	**SEK/litre**

In Figure 14 the MEOH production cost is graphically presented as a range for each factor.

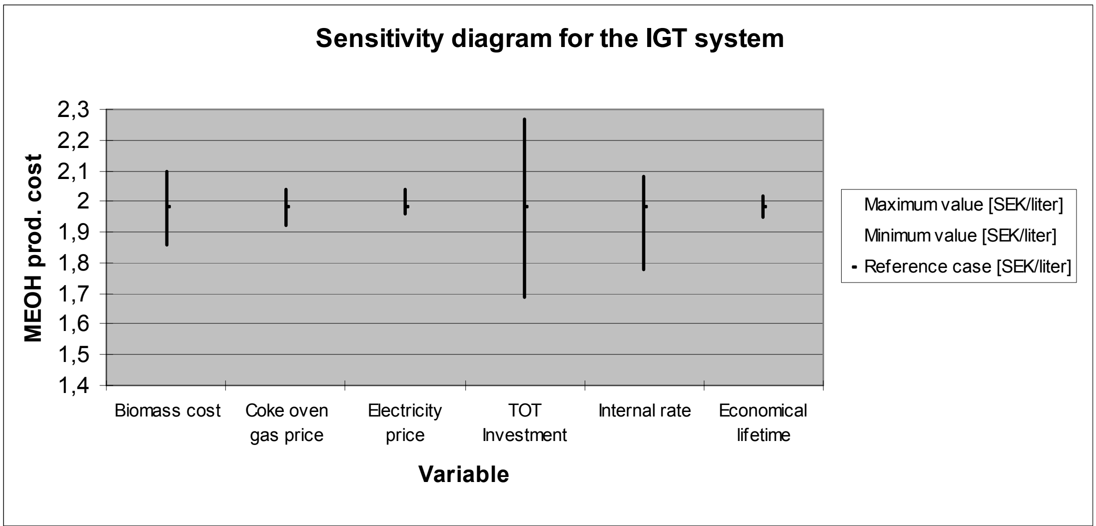

Figure 14 Sensitivity diagram for the IGT system

The analysis with the best and worst case was performed by set all variables in the minimum, maximum and reference position. In Table 12 the selected factors with respective MEOH production cost is presented.

Table 12 Best- and worst case analysis results for the IGT system

Variable	Min	Ref	Max	
Biomass price	70	100	130	**SEK/MWh**
Coke oven gas price	24,14	28,40	32,66	**SEK/GJ**
Electricity price	450	500	650	**SEK/MWh**
TOT Investment	-30	0	30	**%**
Internal rate	6	10	12	**%**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,33**	**1,98**	**2,70**	**SEK/litre**

In Figure 15 the MEOH production cost is graphically presented as a range between the cases. The MEOH production cost differs 1,37 SEK/litre between the worst- and best case.

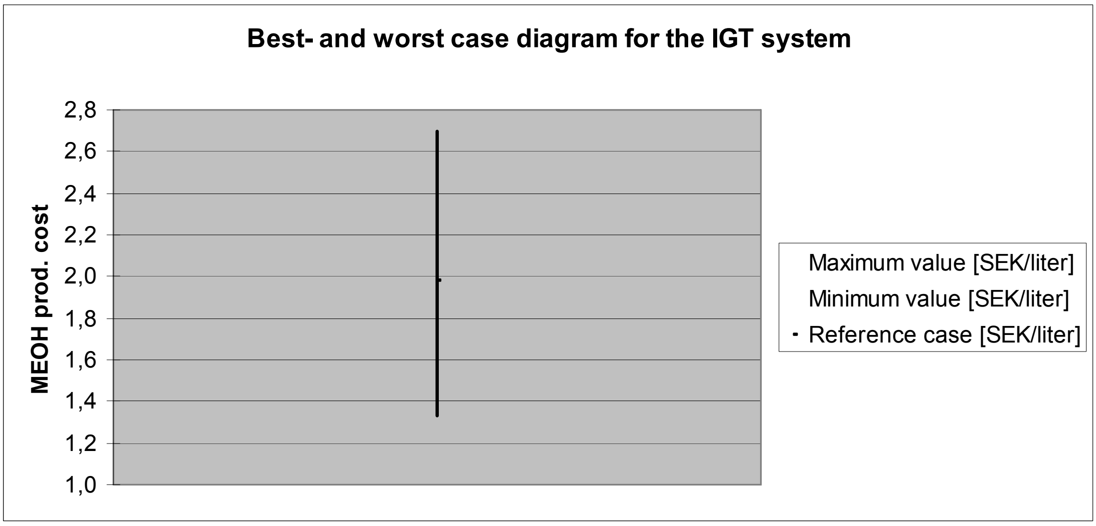

Figure 15 Best- and worst case diagram for the IGT system

6.2 BCL system

The BCL system results in a MEOH conversion efficiency of 64,8% (based on LHV) and a total plant efficiency of 75,0%, see Table 13 for a short energy balance.

Table 13 Energy balance for the BCL system

Energy input		
Coke oven gas	MW	-50,0
Biomass	MW	-52,5
Internal el. use	MW	-7,0
Steam raise	MW	-0,1
Sum	MW	-109,6
Energy output		
Methanol	MW	71,0
El. production	MW	1,5
District heating	MW	4,4
Purge gas	MW	5,2
Sum	MW	82,2
MEOH conversion efficiency	%	64,8
Total plant efficiency	%	75,0

The total efficiency is higher than the IGT system which mainly depends on the higher MEOH conversion efficiency but also the lower electricity demand. The syngas from the gasifier has a better composition than for the IGT system when mixed with the coke oven gas. The CO content is higher in the syngas from the BCL gasifier than from the IGT gasifier. This is good for optimal MEOH production since there is lack of CO in the coke oven gas but much H_2. The investment is higher than for the IGT system, but since the MEOH exchange is higher, the production cost is lower.

For a complete energy balance see Figure 31 in Appendix C.2. All stream results are shown in Table 42 in Appendix D.2. Table 14 presents a summary of the key figures for the BCL system, see also Table 46 in Appendix E.2 for detailed the economic calculations.

Table 14 Key figures for the BCL system

Investments	**Unit**	**Value**
Total capital requirement	MUS$	100,8
Total annual cost	MUS$/year	30,2
Methanol produced	m^3/year	122774
Methanol production cost	US$/litre	0,25
Methanol production cost	SEK/litre	1,82

The sensitivity analysis was performed as for the IGT system and showed that also here, the investment followed by the internal rate was the most crucial factors that influenced the MEOH production cost. In Table 15 the selected factors with respective MEOH production cost is presented.

Table 15 Sensitivity analysis results for the BCL system

Variable	**Min**	**Ref**	**Max**	
Biomass price	70	100	130	**SEK/MWh**
MEOH prod. cost	**1,71**	**1,82**	**1,92**	**SEK/litre**
Coke oven gas price	24,14	28,40	32,66	**SEK/GJ**
MEOH prod. cost	**1,77**	**1,82**	**1,87**	**SEK/litre**
Electricity price	450	500	650	**SEK/MWh**
MEOH prod. cost	**1,80**	**1,82**	**1,87**	**SEK/litre**
TOT Investment	-30	0	30	**%**
MEOH prod. cost	**1,53**	**1,82**	**2,10**	**SEK/litre**
Internal rate	6	10	12	**%**
MEOH prod. cost	**1,62**	**1,82**	**1,92**	**SEK/litre**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,79**	**1,82**	**1,85**	**SEK/litre**

In Figure 16 the MEOH production cost is graphically presented as a range for each factor.

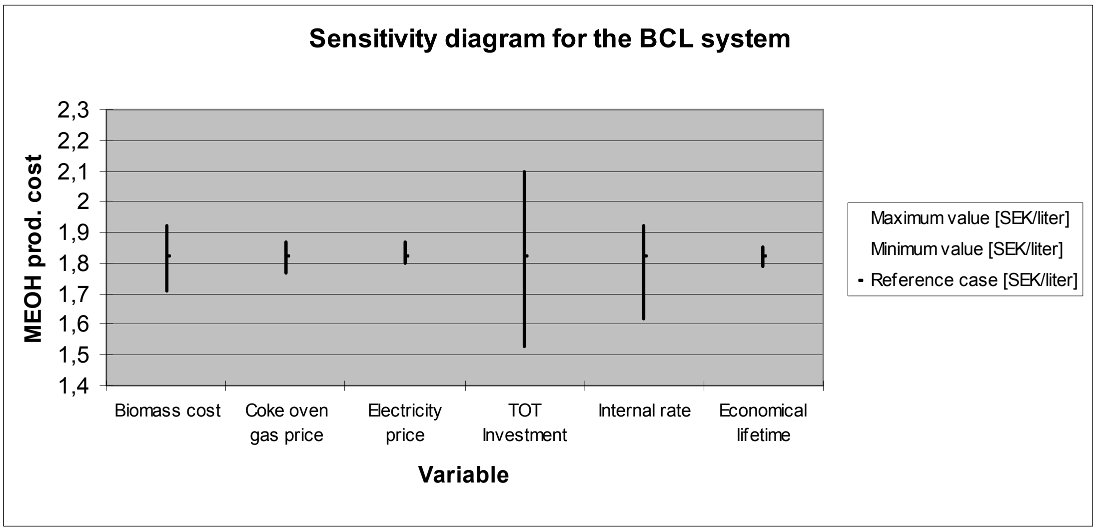

Figure 16 Sensitivity diagram for the BCL system

The analysis with the best and worst case was performed as for the IGT system. In Table 16 the selected factors with respective MEOH production cost is presented.

Table 16 Best- and worst case analysis results for the BCL system

Variable	Min	Ref	Max	
Biomass price	70	100	130	**SEK/MWh**
Coke oven gas price	24,14	28,40	32,66	**SEK/GJ**
Electricity price	450	500	650	**SEK/MWh**
TOT Investment	-30	0	30	**%**
Internal rate	6	10	12	**%**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,20**	**1,82**	**2,50**	**SEK/litre**

In Figure 17 the MEOH production cost is graphically presented as a range between the cases. The MEOH production cost differs 1,30 SEK/litre between the worst- and best case.

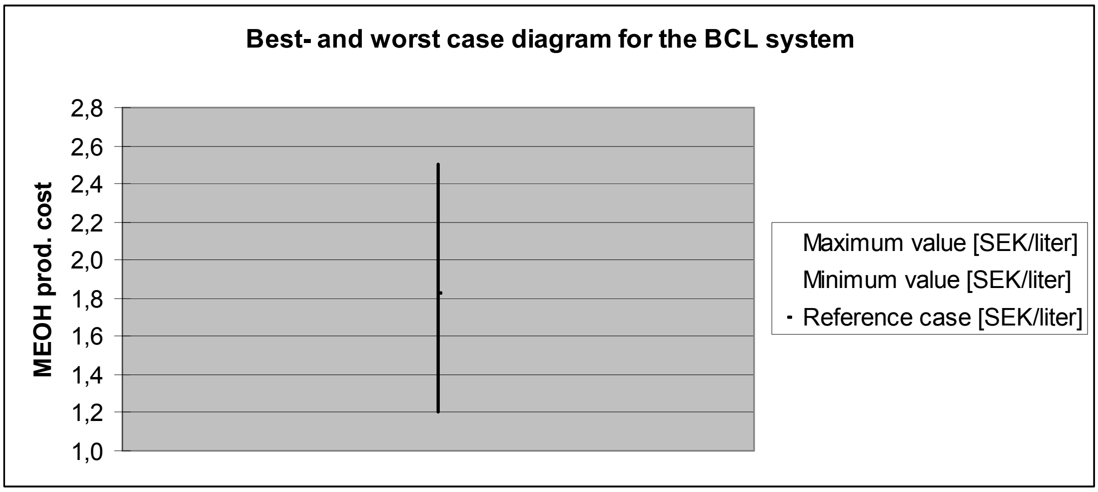

Figure 17 Best- and worst case diagram for the BCL system

6.3 ATR system

The ATR system results in a MEOH conversion efficiency of 56,2% (based on LHV) and a total plant efficiency of 70,3%, see Table 17 for a short energy balance.

Table 17 Energy balance for the ATR system

Energy input		
Coke oven gas	MW	-50,0
Biomass	MW	0,0
Internal el. use	MW	-3,8
Steam raise	MW	0,0
Sum	MW	-53,8
Energy output		
Methanol	MW	30,2
El. production	MW	1,2
District heating	MW	3,5
Purge gas	MW	2,9
Sum	MW	37,8
MEOH conversion efficiency	%	56,2
Total plant efficiency	%	70,3

The MEOH conversion efficiency is nearly the same as for IGT system but the total efficiency is higher. The higher total efficiency depends on that no gasifier or pre-treatment unit is needed. The later uses about 10% of energy input for drying the sawdust [8]. The investment cost is lower than the IGT and BCL systems because less equipment is needed. The MEOH production cost is slightly higher than for the BCL system. For a complete energy balance see
Figure *32* in Appendix C.3. All stream results are shown in Table 43 in Appendix D.3. Table 18 presents a summary of the key figures for the ATR system, see also Table 47 in Appendix E.3 for the detailed economic calculations.

Table 18 Key figures for the ATR system

Investments	Unit	Value
Total capital requirement	MUS$	35,8
Total annual cost	MUS$/year	13,1
Methanol produced	m^3/year	52257
Methanol production cost	US$/litre	0,25
Methanol production cost	SEK/litre	1,85

The sensitivity analysis was done as for the previous systems and showed that the investment followed by the internal rate and coke oven gas price was the most crucial factors that influenced the MEOH production cost. In Table 19 the selected factors with respective MEOH production cost is presented.

Table 19 Sensitivity analysis results for the ATR system

Variable	Min	Ref	Max	
Biomass price	70	100	130	**SEK/MWh**
MEOH prod. cost	**-**	**-**	**-**	**SEK/litre**
Coke oven gas price	24,14	28,4	32,66	**SEK/GJ**
MEOH prod. cost	**1,73**	**1,85**	**1,97**	**SEK/litre**
Electricity price	450	500	650	**SEK/MWh**
MEOH prod. cost	**1,82**	**1,85**	**1,93**	**SEK/litre**
TOT Investment	-30	0	30	**%**
MEOH prod. cost	**1,61**	**1,85**	**2,09**	**SEK/litre**
Internal rate	6	10	12	**%**
MEOH prod. cost	**1,68**	**1,85**	**1,94**	**SEK/litre**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,82**	**1,85**	**1,88**	**SEK/litre**

In Figure 18 the MEOH production cost is graphically presented as a range for each factor.

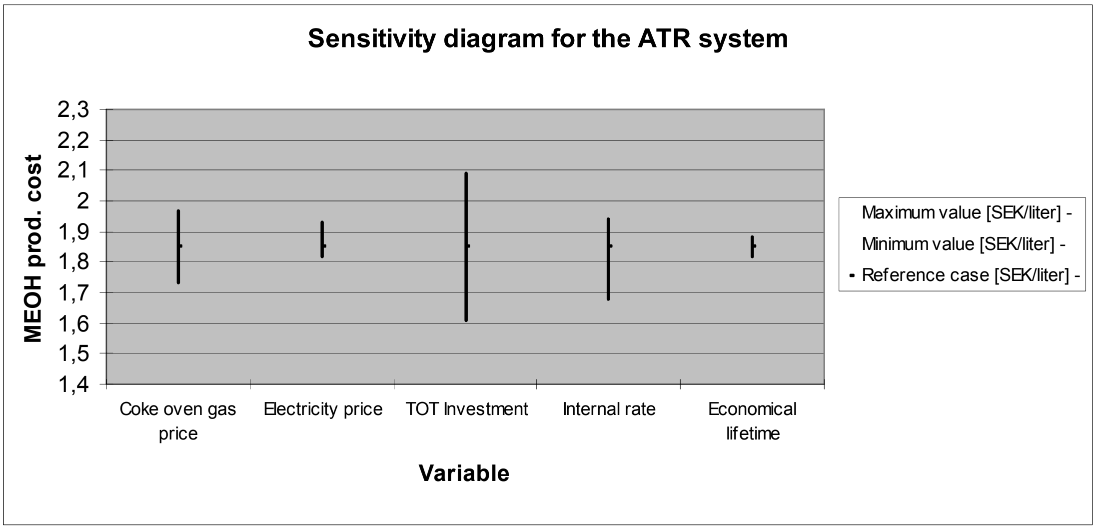

Figure 18 Sensitivity diagram for the ATR system

The best- and worst case analysis was done as for the previous systems. In Table 20 the selected factors with respective MEOH production cost is presented.

Table 20 Best- and worst case analysis results for the ATR system

Variable	Min	Ref	Max	
Biomass price	70	100	130	**SEK/MWh**
Coke oven gas price	24,14	28,40	32,66	**SEK/GJ**
Electricity price	450	500	650	**SEK/MWh**
TOT Investment	-30	0	30	**%**
Internal rate	6	10	12	**%**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,33**	**1,85**	**2,44**	**SEK/litre**

In Figure 19 the MEOH production cost is graphically presented as a range between the cases. The MEOH production cost differs 1,11 SEK/litre between the worst- and best case.

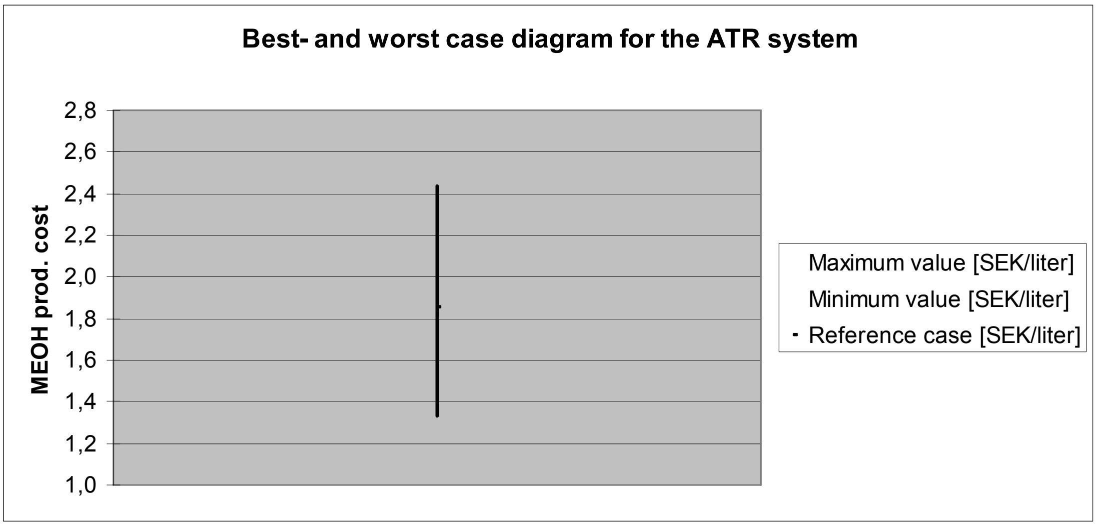

Figure 19 Best- and worst case diagram for the ATR system

6.4 SMR system

The SMR system results in a MEOH conversion efficiency of 67,2% (based on LHV) and a total plant efficiency of 86,5%, see Table 21 for a short energy balance.

Table 21 Energy balance for the SMR system

Energy input		
Coke oven gas	MW	-50,0
Biomass	MW	0,0
Internal el. use	MW	-3,5
Steam raise	MW	0,0
Sum	MW	-53,5
Energy output		
Methanol	MW	36,0
El. production	MW	1,0
District heating	MW	2,9
Purge gas	MW	6,4
Sum	MW	46,3
MEOH conversion efficiency	%	67,2
Total plant efficiency	%	86,5

The MEOH conversion efficiency is the highest of the systems which depends on a combination with good process optimisation and the fact that the coke oven gas has high H_2 content. In comparison to the ATR system, the steam reformer is a better choice for CH_4 reforming, because less syngas is needed for combustion to satisfy the heat demand for the catalytic process. The total plant efficiency is very high, mainly because the MEOH conversion efficiency is higher and the internal steam and electricity demand is lower than for the ATR system. The total investment is higher than for the ATR system, but since the MEOH extraction is higher, the production cost is lower. For a complete energy balance see
Figure *33* in Appendix C.4. All stream results are shown in Table 44 in Appendix D.4. Table 22 presents a summary of the key figures for the SMR system, see also Table 48 in Appendix E.4 for the detailed economic calculations.

Table 22 Key figures for the SMR system

Investments	Unit	Value
Total capital requirement	MUS$	48,6
Total annual cost	MUS$/year	14,6
Methanol produced	m^3/year	62245
Methanol production cost	US$/litre	0,24
Methanol production cost	SEK/litre	1,75

The sensitivity analysis was done as for the previous systems and showed that, as for the ATR system the investment followed by the internal rate and coke oven gas price was the most crucial factors that influenced the MEOH production cost. In Table 23 the selected factors with respective MEOH production cost is presented.

Table 23 Sensitivity analysis results for the SMR system

Variable	Min	Ref	Max	
Biomass price	70	100	130	**SEK/MWh**
MEOH prod. cost	**-**	**-**	**-**	**SEK/litre**
Coke gas price	24,14	28,40	32,66	**SEK/GJ**
MEOH prod. cost	**1,64**	**1,74**	**1,83**	**SEK/litre**
Electricity price	450	500	650	**SEK/MWh**
MEOH prod. cost	**1,72**	**1,74**	**1,78**	**SEK/litre**
TOT Investment	-30	0	30	**%**
MEOH prod. cost	**1,46**	**1,74**	**2,01**	**SEK/litre**
Internal rate	6	10	12	**%**
MEOH prod. cost	**1,55**	**1,74**	**1,84**	**SEK/litre**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,71**	**1,74**	**1,77**	**SEK/litre**

In Figure 20 the MEOH production cost is graphically presented as a range for each factor.

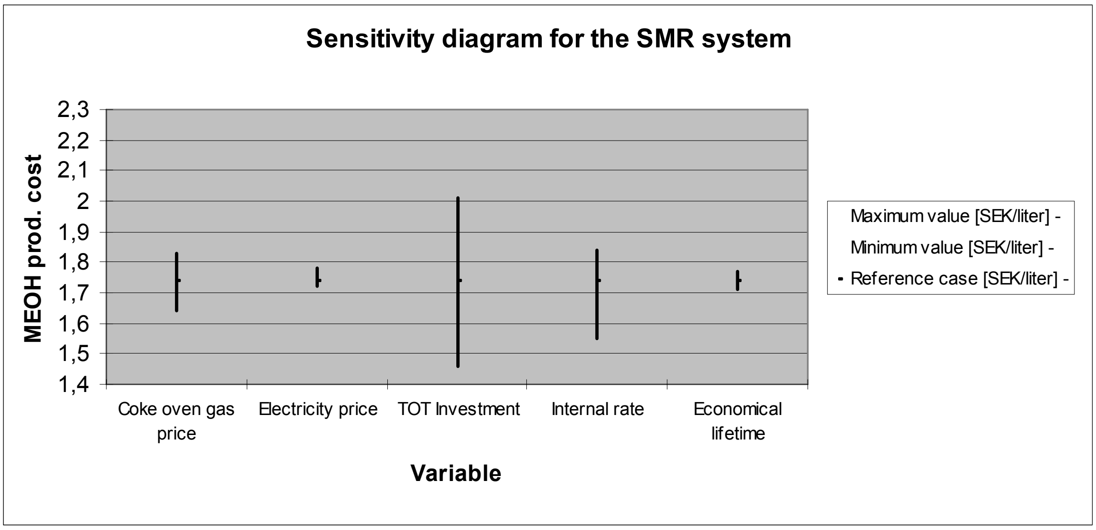

Figure 20 Sensitivity diagram for the SMR system

The best- and worst case analysis was performed as for the previous systems. In Table 24 the selected factors with respective MEOH production cost is presented.

Table 24 Best- and worst case analysis results for the SMR system

Variable	Min	Ref	Max	
Biomass price	70	100	130	**SEK/MWh**
Coke oven gas price	24,14	28,40	32,66	**SEK/GJ**
Electricity price	450	500	650	**SEK/MWh**
TOT Investment	-30	0	30	**%**
Internal rate	6	10	12	**%**
Economical lifetime	20	15	10	**years**
MEOH prod. cost	**1,20**	**1,74**	**2,34**	**SEK/litre**

In Figure 21 the MEOH production cost is graphically presented as a range between the cases. The MEOH production cost differs 1,14 SEK/litre between the worst- and best case.

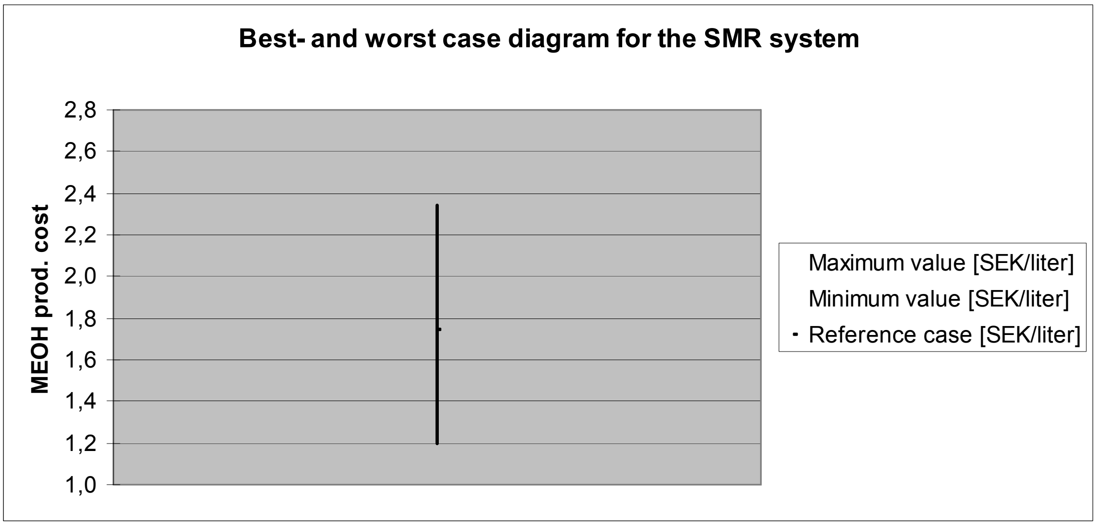

Figure 21 Best- and worst case diagram for the SMR system

6.5 Summary for the systems

A summary of the MEOH production cost range for the systems is presented in Table 25.

Table 25 Summary of production cost range for all systems

Production cost	IGT	BCL	ATR	SMR
Maximum value [SEK/liter]	2,70	2,50	2,44	2,34
Minimum value [SEK/liter]	1,33	1,20	1,33	1,20
Reference case [SEK/liter]	1,98	1,82	1,85	1,74

In Figure 22 the MEOH production cost range for the systems is graphically presented.

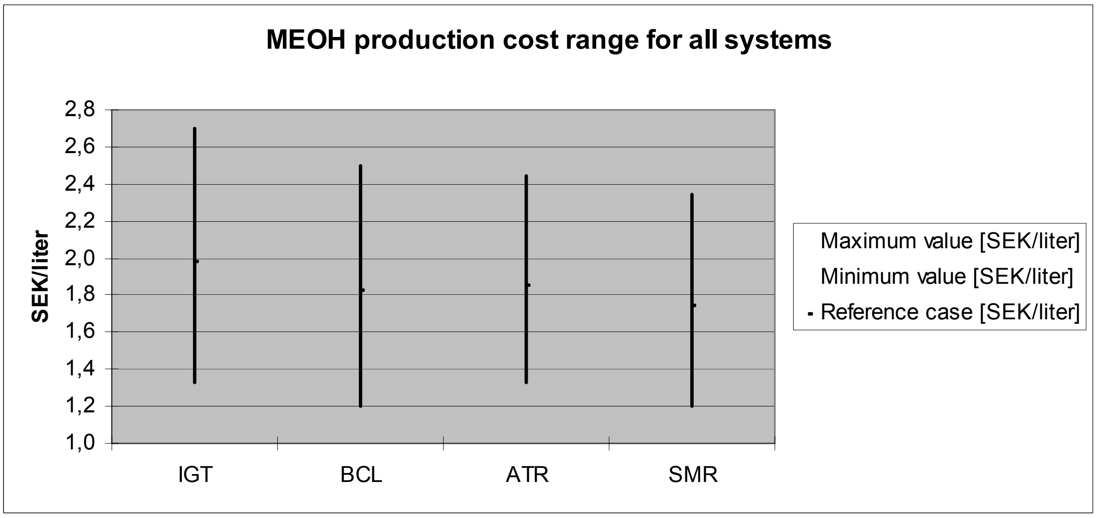

Figure 22 MEOH production cost span for all systems

The production cost for the system is nearly in the same range. In general bio refinery plants has high investments, thus the scale has to be large to be economic profitable. The remarkable conclusion here is that the small ATR- and SMR system could produce MEOH at even lower cost than the larger IGT- and BCL system.

The methanol production cost for the IGT- and BCL system in the reference case are 16,1 $\$/GJ_{MEOH}$ and 14,8 $\$/GJ_{MEOH}$ respectively. In an article made by Faaij et al. [22] methanol production cost were about 16,5 $€/GJ_{MEOH}$ for a 100 $MW_{fuel\ input}$.

In the Faaij et al. report [8] the result for system 5 and 6 with 80 $MW_{biomass\ input}$ has a production cost of 14,8 $\$/GJ_{MEOH}$ and 12,9 $\$/GJ_{MEOH}$ respectively. The used equipment in system 5 and 6 is comparable with the one used in the IGT- and BCL system respectively. This confirms that the methanol production cost in this study is in the range with previous studies. Even if the MEOH production cost in the previous studies with only biomass as fuel input seems a bit optimistic since the systems in this study has higher MEOH conversion efficiency. The higher conversion efficiency is mainly due to the high H_2 content in the coke oven gas.

6.6 Fuel value analysis

The fuel value analysis will show the minimum MEOH sale price when the coke oven gas and the sawdust has a higher economic value as fuel input for MEOH production than electricity (LUKAB) and pellets production (Bioenergi). The annual profit per GWh fuel input is calculated for electricity and pellets production. The highest annual profit per GWh fuel input is used to calculate the minimum MEOH sale price. The conditions for the analysis are:

- The cost for the increased electricity production at LUKAB is assumed to be only the cost for the coke oven gas. No increased capital, operating or maintenance costs are taken into account. Since LUKAB already has their energy need covered by the steel mill gases, the extra coke oven gas is just a bonus which will increase their electricity production. No other income than the sold electricity is assumed to be possible. In reality LUKAB probably would have to make investments to increase their production capacity since the capacity is near max today and the extra coke oven gas increases the fuel input with about 19%.

- The pellets production cost without sawdust cost is set to be 29,7 €/tonne pellets produced [23], sawdust cost is then added. The production cost has been assumed as it is not available from Bioenergi.

The analysis is made concerning the maximum, minimum and reference case for the MEOH production cost in the sensitivity analysis. The sawdust- and coke oven gas cost is the same as in the maximum, minimum and reference case in the sensitivity analysis. The sale price for pellets is assumed to change with same percentage as for the sawdust cost. The reference value for the pellets sale price is assumed on the basis of the turnover divided with amount of sold pellets at Bionergi in the year 2005 [24]. Table 26 shows the factors used.

Table 26 Fuel value analysis factors

Price		Max	Min	Ref
Coke oven gas	SEK/GWh	117576	86904	102240
Sawdust	SEK/GWh	130000	70000	100000
Electricity	SEK/GWh	550000	300000	450000
Pellets	SEK/GWh	260000	140000	200000

6.6.1 IGT system

Bioenergi has a higher profit per GWh energy input than LUKAB in all three cases. The MEOH plant must at least bring the same profit as Bioenergi to give the sawdust the same economic value as fuel input for MEOH production. The profit per GWh energy input is used to calculate the lowest MEOH sale price. The result is shown in Table 27.

Table 27 Fuel value analysis results for the IGT system

		Max	Max	Max	Min	Min	Min	Ref	Ref	Ref
Production		**LUKAB**	**Bioenergi**	**IGT-MEOH**	**LUKAB**	**Bioenergi**	**IGT-MEOH**	**LUKAB**	**Bioenergi**	**IGT-MEOH**
Coke oven gas	GWh	-400		-400	-400		-400	-400		-400
Sawdust	GWh		-420	-420		-420	-420		-420	-420
Electricity	GWh	108			108			108		
District heating	GWh									
Pellets	GWh		378			378			378	
MEOH	GWh			480			480			480
Annual cost/income										
Production cost	MSEK	-47,03	-73,29	-279,96	-34,76	-48,09	-137,90	-40,90	-60,69	-205,30
Income	MSEK	59,40	98,28	328,74	32,40	52,92	147,33	48,60	75,60	234,41
Total profit	MSEK	12,37	24,99	48,79	-2,36	4,83	9,43	7,70	14,91	29,11
Total profit/energy input	SEK/GWh	30924	59495	59495	-5904	11495	11495	19260	35495	35495

The minimum sale price for each case is shown in Table 28.

Table 28 Minimum MEOH sale price for the IGT system

Price		Max	Min	Ref
Min sale price MEOH	SEK/GWh	685391	307167	488713
Min sale price MEOH	SEK/litre	3,17	1,42	2,26
Production cost MEOH	SEK/litre	2,70	1,33	1,98
Sale price/prod. Cost	%	17,43	6,83	14,18

If the market price is higher than the minimum MEOH sale price, the coke oven gas and sawdust should be used for MEOH production to reach the best economy.

In Figure 23 the production and minimum sale price is graphically presented.

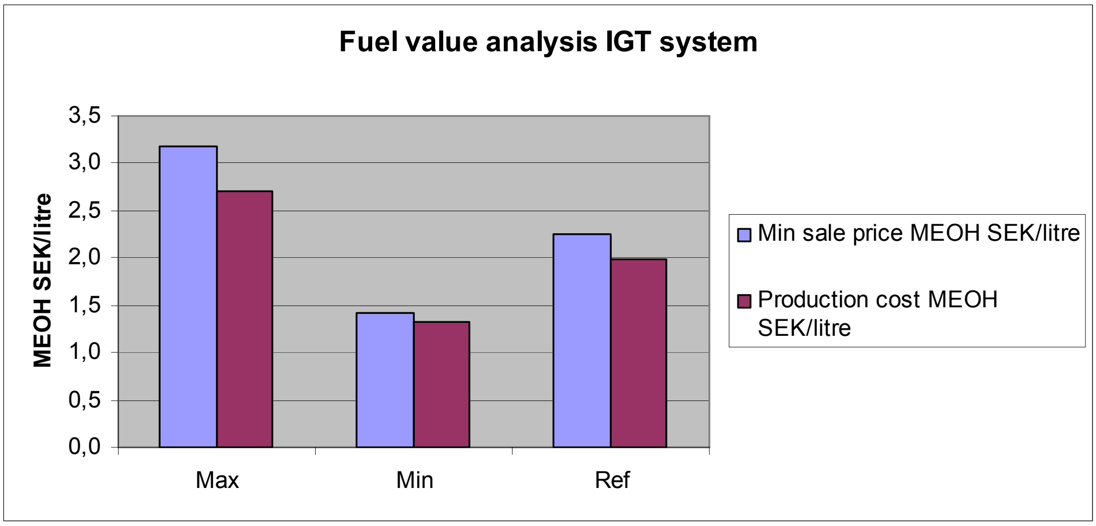

Figure 23 Production cost and minimum MEOH sale price for the IGT system

6.6.2 BCL system

Bioenergi has a higher profit per GWh energy input than LUKAB in all three cases. The MEOH plant must as for the IGT system at least bring the same profit as Bioenergi. The result is shown in Table 29.

Table 29 Fuel value analysis results for the BCL system

		Max	Max	Max	Min	Min	Min	Ref	Ref	Ref
Production		**LUKAB**	**Bioenergi**	**BCL-MEOH**	**LUKAB**	**Bioenergi**	**BCL-MEOH**	**LUKAB**	**Bioenergi**	**BCL-MEOH**
Coke oven gas	GWh	-400		-400	-400		-400	-400		-400
Sawdust	GWh		-420	-420		-420	-420		-420	-420
Electricity	GWh	108			108			108		
District heating	GWh									
Pellets	GWh		378			378			378	
MEOH	GWh			568			568			568
Annual cost/income										
Production cost	MSEK	-47,03	-73,29	-307,04	-34,76	-48,09	-147,38	-40,90	-60,69	-223,53
Income	MSEK	59,40	98,28	355,83	32,40	52,92	156,80	48,60	75,60	252,63
Total profit	MSEK	12,37	24,99	48,79	-2,36	4,83	9,43	7,70	14,91	29,11
Total profit/energy input	SEK/GWh	30924	59495	59495	-5904	11495	11495	19260	35495	35495

The minimum sale price for each case is shown in Table 30.

Table 30 Minimum MEOH sale price for the BCL system

Price		**Max**	**Min**	**Ref**
Min sale price MEOH	SEK/GWh	626314	276003	444673
Min sale price MEOH	SEK/litre	2,90	1,28	2,06
Production cost MEOH	SEK/litre	2,50	1,20	1,82
Sale price/prod. Cost	%	15,89	6,40	13,02

The coke oven gas and sawdust should be used for MEOH production to reach the best economy, if the market price is higher than the minimum MEOH sale price.

In Figure 24 the production and minimum sale price is graphically presented.

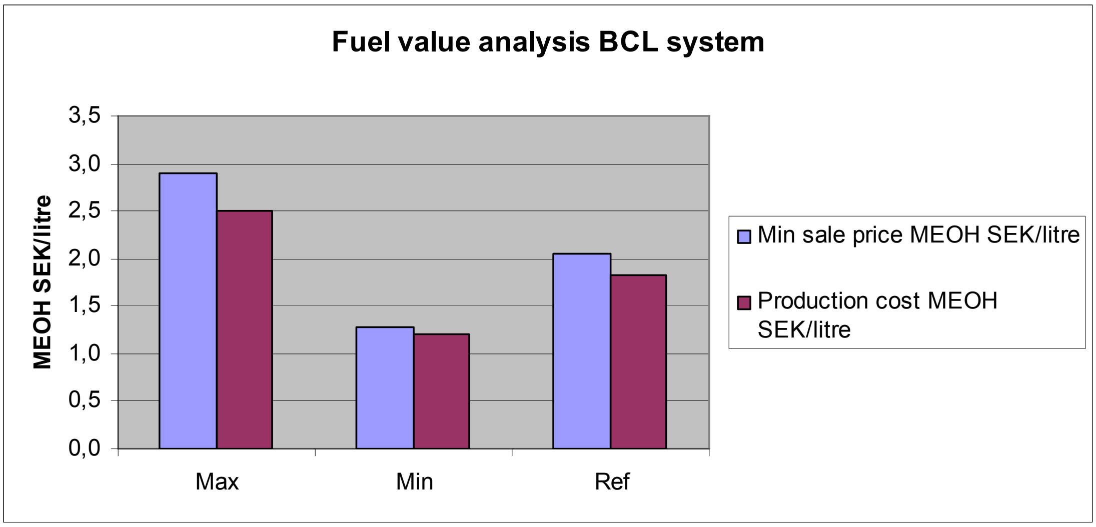

Figure 24 Production cost and minimum MEOH sale price for the BCL system

6.6.3 ATR system

The MEOH plant must at least bring the same profit as LUKAB to give the coke oven gas the same economic value as fuel for MEOH production. The profit per GWh energy input is used to calculate the lowest MEOH sale price. The result is shown in Table 31.

Table 31 Fuel value analysis results for the ATR system

		Max	Max	Min	Min	Ref	Ref
Production		**LUKAB**	**ATR-MEOH**	**LUKAB**	**ATR-MEOH**	**LUKAB**	**ATR-MEOH**
Coke oven gas	GWh	-400	-400	-400	-400	-400	-400
Sawdust	GWh						
Electricity	GWh	108		108		108	
District heating	GWh						
Pellets	GWh						
MEOH	GWh		242		242		242
Annual cost/income							
Production cost	MSEK	-47,03	-127,59	-34,76	0	-40,90	-96,74
Income	MSEK	59,40	139,96	32,40	0	48,60	104,44
Total profit	MSEK	12,37	12,37	-2,36	0	7,70	7,70
Total profit/energy input	SEK/GWh	30924	30924	-5904	0	19260	19260

In the minimum case the coke oven gas cost is higher for LUKAB than the income for the sold electricity. Thus the minimum MEOH sale price at least has to be equal to the MEOH production cost.

The minimum sale price for each case is shown in Table 32.

Table 32 Minimum MEOH sale price for the ATR system

Price		Max	Min	Ref
Min sale price MEOH	SEK/GWh	578611	287516	431778
Min sale price MEOH	SEK/liter	2,68	1,33	2,00
Production cost MEOH	SEK/liter	2,44	1,33	1,85
Sale price/prod. Cost	%	9,69	0,00	7,96

The coke oven gas should be used for MEOH production to reach the best economy, if the market price is higher than the minimum MEOH sale price.

In Figure 25 the production and minimum sale price is graphically presented.

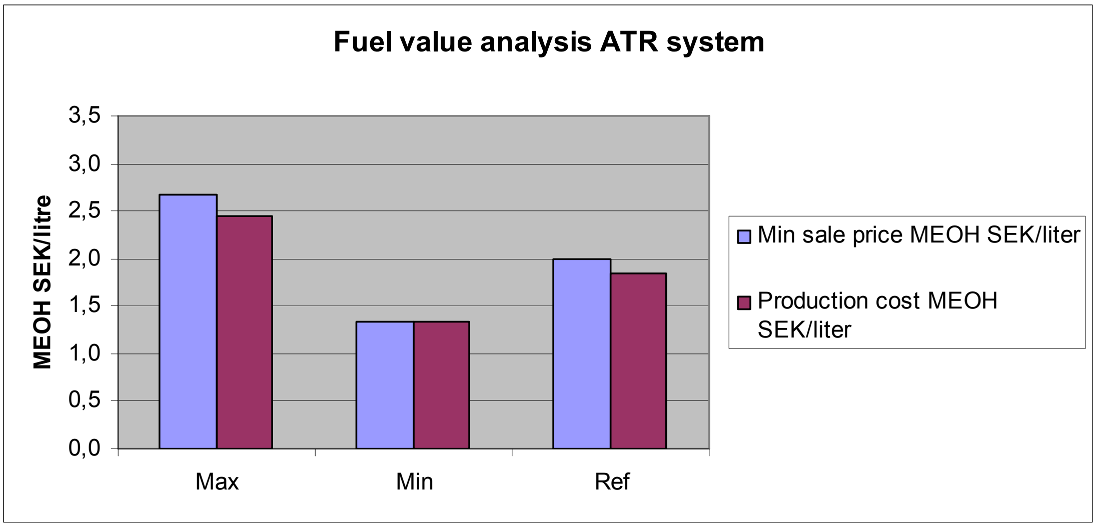

Figure 25 Production cost and minimum MEOH sale price for the ATR system

6.6.4 SMR system

The same condition as for the ATR system is applied here. The MEOH plant must at least bring the same profit as LUKAB to give the coke oven gas the same economical value as fuel for MEOH production. The profit per GWh energy input is used to calculate the lowest MEOH sale price. The result is shown in Table 33.

Table 33 Fuel value analysis results for the SMR system

		Max	Max	Min	Min	Ref	Ref
Production		**LUKAB**	**SMR-MEOH**	**LUKAB**	**SMR-MEOH**	**LUKAB**	**SMR-MEOH**
Coke oven gas	GWh	-400	-400	-400	-400	-400	-400
Sawdust	GWh						
Electricity	GWh	108		108		108	
District heating	GWh						
Pellets	GWh						
MEOH	GWh		288		288		288
Annual cost/income							
Production cost	MSEK	-47,03	-145,63	-34,76	0	-40,90	-108,29
Income	MSEK	59,40	158,00	32,40	0	48,60	116,00
Total profit	MSEK	12,37	12,37	-2,36	0	7,70	7,70
Total profit/energy input	SEK/GWh	30924	30924	-5904	0	19260	19260

In the minimum case, as for the ATR system, the coke oven gas cost is higher for LUKAB than the income for the sold electricity. Thus the minimum MEOH sale price at least has to be equal to the MEOH production cost.

The minimum sale price for each case is shown in Table 34

Table 34 Minimum MEOH sale price for the SMR system

Price		Max	Min	Ref
Min sale price MEOH	SEK/GWh	548820	259413	402908
Min sale price MEOH	SEK/litre	2,54	1,20	1,86
Production cost MEOH	SEK/litre	2,34	1,20	1,74
Sale price/prod. Cost	%	8,49	0,00	7,11

If the market price is higher than the minimum MEOH sale price the coke oven gas should be used for MEOH production to reach the best economy.

In Figure 26 the production and minimum sale price is graphically presented.

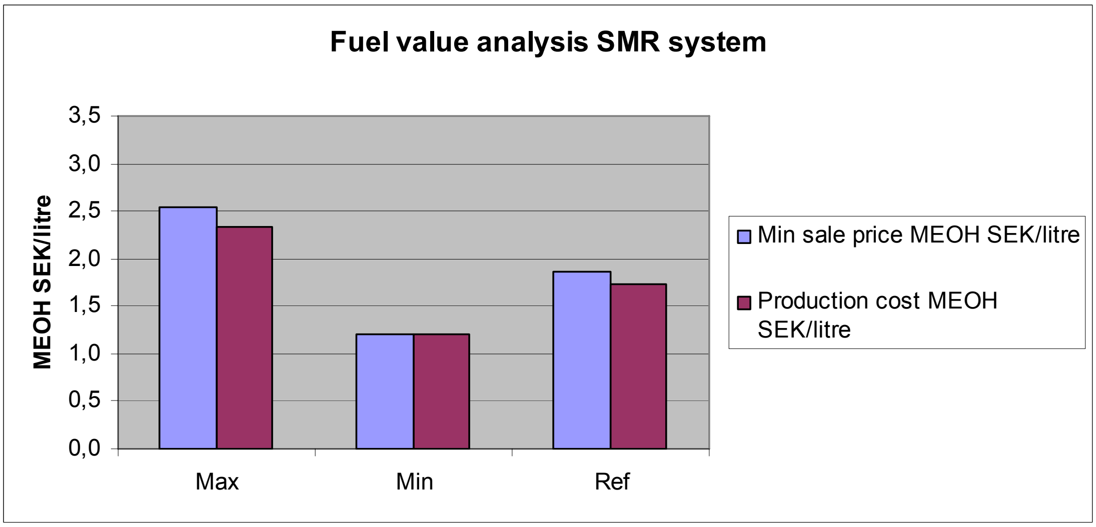

Figure 26 Production cost and minimum MEOH sale price for the SMR system

6.6.5 Summary for the fuel value analysis

A summary of the minimum MEOH sale price range for the systems is shown in Table 35.

Table 35 Summary of the minimum MEOH sale price for the systems

System		IGT	BCL	ATR	SMR
Min sale price **(MAX)**	SEK/litre	3,17	2,90	2,68	2,54
Min sale price **(MIN)**	SEK/litre	1,42	1,28	1,33	1,20
Min sale price **(Ref)**	SEK/litre	2,26	2,06	2,00	1,86

In Figure 27 the minimum MEOH sale price range for the systems is graphically presented.

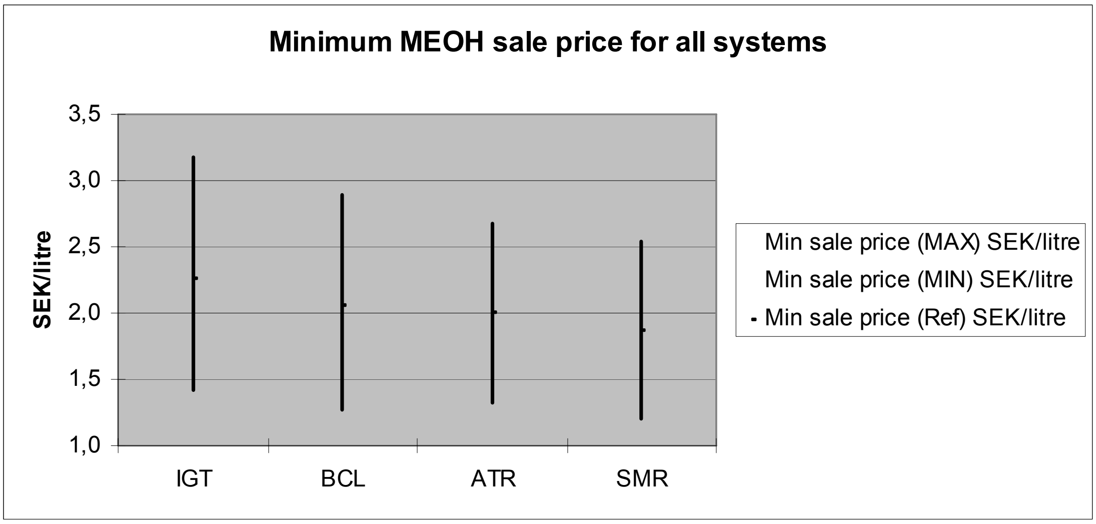

Figure 27 Minimum MEOH sale price for all systems

The minimum MEOH sale price is nearly in the same range for all of the systems.

6.7 MEOH- and Gasoline price

In the year 2006, the average 95-octane gasoline price was 11.55 SEK/litre in Sweden [25]. The price consists to a large extent by taxes, 7,05 SEK/litre. Thus the product price for gasoline is only 3,68 SEK/litre without taxes [25].

If the methanol produced from the systems considered is to be this heavy charged with taxes the price will be way too high to compete with fossil fuels. It is however reasonable to assume that the taxes will be lower for the IGT and BCL system than for gasoline since biomass is used which could be considered as renewable. Yet, there is some space for putting on taxes to methanol. It could be assumed that the cars will consume twice as much of methanol for the same driving distance, as it would be with gasoline. This due to the heating value for methanol is about the half to gasoline while the density is roughly the same. The methanol sale price can then be about 5,78 SEK/litre (gasoline equivalence). If the minimum methanol sale price for the reference case from fuel value analysis is concerned, the marginal for taxes, distributor, retailer etc. for the systems is shown in Table 36.

Table 36 Marginal for additional charge on methanol price

Price		IGT	BCL	ATR	SMR
Min sale price MEOH	SEK/litre	2,26	2,06	2,00	1,86
Max sale price MEOH	SEK/litre	5,78	5,78	5,78	5,78
Differance	SEK/litre	3,52	3,72	3,78	3,92
Differance	%	256	281	289	311

The final methanol sale price for the consumer must however be put somewhat lower than for gasoline, in order to stimulate the use of methanol. Conversely gasoline taxes could be increased further. The high taxes of transportation fuels today is officially to a large extent justified by environmental reasons. Nevertheless the state income on fuel taxes is very important for the state budget and lowering these taxes would decrease the tax income. An interesting question regarding these incomes is how much money that indirectly would be generated for the benefit of the government. That is money stays within the country, jobs are created, biomass conversions techniques could be sold and other positive effects.

6.8 District heating

To get as good total efficiency and economy as possible, useable waste heat should be sold to a district heating network. The problem in Luleå is that the municipality at present already has the district heating demand fulfilled by LUKAB. Nevertheless the potential extraction and income will be presented for the systems. The district heating is delivered from LUKAB with a temperature of 75-120°C and the return temperature is 40-70°C [16]. The temperature range depends on the municipality demand and period of the year. The calculations are based on a temperature difference of 30°C (100°C_{out}/70°C_{in}). The heat extraction taken from the Aspen models and is presented in Table 37.

Table 37 Potential extraction of district heating for all systems

District heating	IGT	BCL	ATR	SMR
Heat extracted [MW]	5,14	4,45	3,48	2,95
Heat of total fuel input [%]	5,01	4,34	6,96	5,90

The potential district heating extraction for the systems is kept low, which is a result of good process optimisation. The price for district heating in Luleå is about 200 SEK/MWh [25]. In Table 38 the MEOH production cost with accounted income for district heating is presented.

Table 38 MEOH production cost for all systems with accounted income for district heating

District heating	IGT	BCL	ATR	SMR
MEOH production cost [US$/litre]	0,27	**0,25**	**0,25**	**0,23**
MEOH production cost [SEK/litre]	1,97	1,81	1,84	1,73
Decreased production cost [%]	0,50	0,50	0,50	0,60

The loss of income for the potential district heating could in principle be neglected, which is positive for the overall system economy.

7 Motor fuel demand

The inhabitants in Norrbotten was 251 740 people in the year 2005 [27], see Table 49 in Appendix F.1. The average gasoline consumption per inhabitant in the county amounts to about 500 litres/year (2003) [28], see Table 50 in Appendix F.2. Hence the total annual gasoline consumption for the county is calculated to approximately 126 million litres. Since the heating value for methanol is about half compared to gasoline, while the density is roughly the same, the cars will use more fuel than before [7]. This results in that every person who choose methanol (M100) instead of gasoline will consume roughly twice the amount of methanol by volume. If it is assumed that the average gasoline consumption per inhabitant in the county is replaced by methanol, the demand would be about 1000 $litres_{MEOH}$/person and year. The total annual demand for the county would then be about 252 million litres of methanol.

In Table 39 the possible number of MEOH supplied inhabitants is shown for each system.

Table 39 Possible number of MEOH supplied inhabitants

System	IGT	BCL	ATR	SMR
MEOH produced [m^3/year]	103707	122774	52257	62245
No. Inhabitants supplied	103707	122774	52257	62245

The BCL system could almost supply the half of the theoretical demand in the county. The town of Luleå had 72 751 inhabitants in the year 2005 [27], which then has the potential to be self-supporting of MEOH. If the MEOH would be used as a blend like ethanol (5%) into gasoline, the remaining MEOH could be exported.

Regarding the future transportation fuel demand in the county, a study with four different scenarios describing the future transportation fuel demand (only passenger cars, no lorry, bus or cargo transports) until the year 2025 have been created. In the BAU scenarios it is assumed that the national target of 20% bio fuels of the total transportation fuel demand by the year 2020 is reached. The Green scenarios are even more optimistic assuming a biofuel share of 75% until the year 2025. Figure 28 shows the results [29].

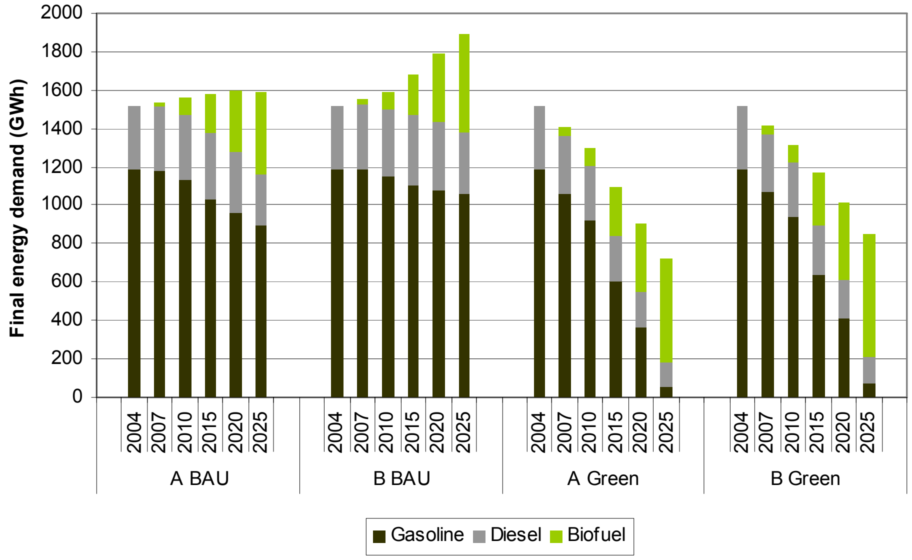

Figure 28 Total transportation fuel demand in Norrbotten according to the four scenarios [29]

The total demand amounts to 720-1900 GWh per year of which bio fuels constitute of 430-640 GWh [29].

The methanol could be included in the bio fuel part in the four scenarios for the year 2025. The annual methanol production for the four systems is used to illustrate to which extent it could supply the demand in the county for the four scenarios, see Table 40.

Table 40 Methanol produced and amount of fuel demand supplied to the county in the year 2025

MEOH Plant	IGT	BCL	ATR	SMR
Produced MEOH [GWh]	480	568	242	288
Bio fuel demand A BAU [GWh]	429	429	429	429
Fraction of MEOH supplied [%]	112	132	56	67
Bio fuel demand B BAU [GWh]	511	511	511	511
Fraction of MEOH supplied [%]	94	111	47	56
Bio fuel demand A Green [GWh]	538	538	538	538
Fraction of MEOH supplied [%]	89	106	45	54
Bio fuel demand B Green [GWh]	638	638	638	638
Fraction of MEOH supplied [%]	75	89	38	45

The BCL system itself could supply the demand in the county in all scenarios except the B green case.

8 Environment

In general, the environmental performance of biomass conversion technologies is far superior to conventional, fossil-based energy systems. But since coke oven gas also is used as fuel which originally comes from fossil based coal, the environmental impacts have to be more carefully considered. These negative impacts could occur in extraction of raw material, production or in end use of the fuel.

The integrated steelmaking industry is one of the major industries that have large CO_2 emissions, since coal is used as main reduction material when the agglomerated iron ore is reduced and melted in the blast furnace. In order to minimize the CO_2 emission from integrated steelmaking, the energy demand could be minimized or used in a more efficient way [4]. If the additional coke oven gas would be used in LUKAB to increase the present electricity production, the specific CO_2 emissions would be high since just electricity would be saleable. Then a better option would be to produce methanol instead, because the conversion- and total plant efficiency is higher, which leads to lower specific emissions. Regardless if only biomass is used there would be emissions.

8.1 Emission of GHG

Carbon dioxide (CO_2), methane, nitrous oxide and certain other gases are called greenhouse gases because they trap heat in the Earth's atmosphere. Global warming caused by an increased concentration of greenhouse gases is of major concern today. It is predicted that global warming will cause sea level rise and other environmental impacts that may harm humans and nature. The significant increase of greenhouse gases over the last 200 years is almost entirely due to human activities and particularly due to combustion of fossil fuels [30].

Trees and plants remove carbon from the atmosphere through photosynthesis, forming new biomass as they grow. This carbon is stored in the biomass. When biomass burns, carbon returns to the atmosphere in the form of CO_2. This cycle makes it possible for biomass energy to avoid increasing the net amount of CO_2 in the atmosphere [30].

There is no net increase in atmospheric CO_2 if the new growth of plants and trees fully replaces the supply of biomass consumed for energy. However, if the collection or processing of biomass consumes any fossil fuel, additional biomass would need to be grown to offset the carbon released from the fossil fuel [30].

8.2 Acid Rain

Air pollution from burning fossil fuels is the major cause of acid rain. Emissions of sulphur dioxide (SO_2) and nitrogen oxides (NO_x) react in the atmosphere with water, oxygen and oxidants to form acidic compounds (sulphuric acid and nitric acid). Some of these compounds fall to earth in the form of acid rain, snow or fog. Acid rain increases acidity of lakes and streams and damages trees at high elevations [30].

Efficient combustion of biomass results in low emissions of SO_2 and production of fewer organic compounds that cause smog compared to emissions from facilities that burn coal or oil. The level of NO_x emissions from biomass combustion facilities depends on the design of the facility and the nitrogen content of the feedstock. Pollution control equipment can further reduce NO_x and particulate emissions [30].

8.3 Local environmental impacts at plant site

When building a plant for methanol production, local environmental impacts also have to be considered. The most obvious interference for the citizens is the landscape change a new plant brings. Landscape change can be quantified in terms of area affected. Heavy trucks that transports biomass and methanol is hazardous and routes should be made so that they can take the safest way possible, avoiding schools etc. This is something that may disturb the local environment and people. For this reasons, it is desirable to place a new plant somewhat in the periphery of a city.

Thermal pollution is a temperature change in natural water bodies caused by human influence. The main cause of thermal pollution is the use of water as a coolant. A methanol plant produces heated waste water and thus is a source of thermal pollution. It is desirable to use this excess heat in a district heating network both for the economy and to reduce the thermal pollution. Thermal pollution may lead to fish death and disturbance in the ecosystem, which in turn may affect fishing, hunting, agriculture, forestry and tourism in the area. It might be important to identify the groups that are affected. The methanol plants of the sizes considered in this report should not impose that large of a risk for thermal pollution, because the amount of excess heat is low.

8.4 Emissions of carcinogenic substances

Several carcinogenic substances that are produced in combustion of gasoline and diesel are eliminated by using methanol instead. But as mentioned in section 1.4.1 formaldehydes are emitted from methanol combustion which is considered as carcinogenic. Although aldehydes are carcinogenic, it is the indoor concentration that is considered to be of importance for the risk of developing cancer. Besides the mean values of indoor concentrations is today generally ten times lower than the threshold value. Outdoor values of aldehydes are at 100 times below the threshold value. Aldehydes in exhaust from methanol are broke down when it come in contact with the surrounding air. This is why it only would affect the concentration of aldehydes in the outdoor air marginally. But it could anyway be a problem in large cities which have problem with heavy smog which then could lead to higher concentrations of aldehydes than normal [7].

9 Future improvements

9.1 Active bed materials

Fluidized beds often use active bed materials like dolomite or blast furnace slag to reduce the problem with bed agglomeration, as mentioned in chapter 2.1.3.1. The addition of the mineral binding products like dolomite will increase the production cost a bit [11]. When SSAB produces steel they deposit a large quantity of blast furnace slag which normally is used as fill up material at construction of roads. There is anyhow blast furnace slag left which possible could be used as bed material [31]. The use of the blast furnace slag could be good for the methanol production economy as well as for SSAB to have positive optional way to get rid of it. Nevertheless further studies of how the quality and composition of the blast furnace slag affects the gasifier must be done.

9.2 Fuel cell

Fuel cell powered vehicles is still under development and although some successful models exists, it will take some time before they reach the market. Fuel cell runs with hydrogen and this imposes a large problem regarding on board fuel storage and distribution to refuelling stations. Hydrogen can however be produced from methanol with an onboard reformer. A fuel cell engine with an onboard reformer that runs on methanol has a higher efficiency compared to an ICE-CI engine [1].

10 Discussion

The modelling of the systems in Aspen plus is a simplification of the reality. Components as chemical reactors specified as Gibbs free energy minimisation reactor calculates the chemical reactions against equilibrium, which does not occur completely in reality. Thus the result is somewhat better than it should be. Still the results from the models should be close to a real plant since each component is specified in detail with inputs known by experience.

The pre-treatment unit for the IGT- and BCL system could perhaps be bought from Bioenergi. Today, the drying of the sawdust at Bioenergi is done with flue gases delivered from LUKAB. The systems considered in this study uses a steam dryer, an alternative route by instead using a flue gas dryer would perhaps allow an increase of the electricity production. The steam extracted at the pressure of 12 bar for the steam dryer, could instead be further expanded in the steam turbine at LUKAB, thus increase the electricity generation.

It is desirable to examine the gas cleaning unit in more detail since that step is one of the most important steps in a methanol plant. The gas purity could in reality have significant affect on the production capacity and material lifetime which could result in higher production costs.

The purge gas was not used further from the methanol separation unit, which of course should be recycled to the reformer or used as fuel at LUKAB. The reason was that the recycle process in Aspen plus is a complex iteration process which takes long time and would anyway not increase the methanol conversion efficiency significantly.

The potential for increased district heating deliverance and revenues for the municipality should be studied more in detail. This was briefly done in this study because it was hard to find out how the future demand will develop and how exploited the market would be by LUKAB.

The part of produced steam for electricity production which originally comes from the sawdust should be able to sell like “green electricity”. That would bring an additional annual income via a green electricity certificate. This possible income was neglected in this study.

The sensitivity analysis showed that the total investment has the most influence on the methanol production cost. The investments can increase greatly in case of delayed construction time and underestimated equipment cost. Thus the investment has to be more investigated or confirmed.

The fuel value analysis was based on the assumption that no increased capital, operating and maintenance cost was concerned for LUKAB. Because the fuel input would increase about 19% from the present with the added coke oven gas, an upgrading of the system probably would be necessary. Perhaps gas turbine would be needed to be able to use the added coke oven gas. If investments are needed the minimum sale price for the methanol systems would decrease further. The assumed pellets production cost was not available from Bioenergi since that key figure normally is secret. The used figure is still probably close to the one for Bioenergi because the study was made for a Swedish pellets plant almost in the same size as Bioenergi. Nevertheless the figure should be compared with the real one at Bioenergi.

The plants considered in the study are small compared with commercial ones who use coal as fuel input. In a longer perspective more biomass like logging residuals and energy crops could be a significant contribution for larger plants. Also new vehicles with fuel cells and ordinary ICE with higher efficiencies could be possible due to the higher octane number in methanol, which then could decrease the total demand of transportation fuels.

There is however some drawbacks with methanol that makes the choice a little bit harder. Methanol is somewhat less compatible with ICE engines than for example ethanol. It should not be a problem for new engines manufactured expressly for methanol use. However more complicated and costly conversion may be expected for existing vehicles. Another issue with future methanol production is that ethanol already has gained some recognition as the future biomass based fuel among politicians and consumers in general for some reasons, probably by lobbying from the ethanol industry. It might be cumbersome to convince politicians and investors to also support the alternative fuel methanol.

11 Conclusion

From an economic perspective it is obvious that methanol production is preferable using only the coke oven gas, since the investment is lower than the sawdust- and coke oven gas systems and still have a methanol production cost that is in the same range. With a coke oven gas input of 400 GWh/year methanol is produced to a price in the range of 1,20-2,34 SEK/litre for the SMR system, while methanol produced from the ATR system would be in the range of 1,33-2,44 SEK/litre. The larger systems with an additional sawdust input of 420 GWh/year would have a methanol production cost in the range of 1,33-2,70 SEK/litre for the IGT system and 1,20-2,50 SEK/litre for the BCL system. Nevertheless the larger IGT and BCL system can produce more methanol than ATR and SMR system and on energy basis about half of the fuel input can be considered as renewable.

The fuel analysis showed that the coke oven gas could have a higher value as fuel for methanol production instead of fuel for increased electricity production at LUKAB. That because the income for LUKAB probably only would come from sold electricity, which is produced by low conversion efficiency and thus high specific CO_2. The sawdust could be used for methanol production as well because the benefits are greater with methanol as transport fuel. Since the methanol could be a more profitable product and it could be used within the county as motor fuel.

Revenues for sold district heating were not taken into account in this study because of difficulties with predict the future demand in the municipality. It is likely that the town of Luleå could not receive more district heating because of existing supplies by LUKAB. A short analysis was made to predict the influence of methanol production cost with accounted sold district heating. The result showed that the influence could almost be neglected because the optimisation of the plants is good and then the potential district heating extraction is kept low. The decrease of the methanol production cost would be about 0,5% for all systems.

The produced methanol from the IGT- and BCL system could potentially replace about half of the present gasoline demand in the county. This would decrease the import of fossil based motor fuel considerable.

The final conclusion is that all systems have an economic potential and is a good option of utilize the sawdust and the coke oven gas for methanol production. In a short term the ATR- and SMR system could be the way to go, since the technique is more close to the common methanol production via natural gas. In a long term the IGT- and BCL system could be better alternative, because the sawdust can be considered as renewable.

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Appendix A
Modelling assumptions

General	
Heat Exchangers	Counter current exchangers
Centrifugal Pump	$\eta = 0{,}9$ $\eta_{driver} = 1$
Compressor	$\eta_{polytropic} = 0{,}80$ $\eta_{mech} = 0{,}90$
Multistage Compressor	$\eta_{isentropic} = 0{,}77$ $\eta_{mech} = 1$ Compression ratio is same for each stage, maximum is 4, such that outlet temperature does not exceed 250°C. Intercooling to 25 – 130°C, last stage no duty
Gas Cleaning	
Quench Scrubber	Modelled as Two Outlet Flash drum $T_{in, gas}$ = 120°C at 1,2 bar pressure $T_{in, water}$ = 25°C T = ~60°C at 1,2 bar pressure by adjusting amount of water. TQUENCH; Minimally 1 m3 water per 1000 m^3 gas Q = 0 W
Hot Gas Cleaning	Modelled as valve $T_{in, gas}$ = ~550°C Δp = -5 bar
Reformer	
Steam Reformer	SMR1 provides heat to SMR2 by combusting flue gas. If this is not sufficient then part of gasifier product is combusted as well. SMR1: Stoichiometric Reactor $T_{flue\ gases}$ = 890°C Air is stoichiometric SMR2: Gibbs free energy minimisation Reactor $T_{in, syngas}$ = 860°C; p_{in} = 15,5 bar $T_{out, syngas}$ = 890°C 3.5 mole steam injected per mole carbon CO_2 recycling if ratio $(H_2 - CO_2) / (CO + CO_2) > 2.1$ after downstream CO_2 removal unit
Autothermal Reformer	ATR1 provides heat ATR2 requires $T_{in, syngas}$ = 700°C Overall 2 mole steam injected per mole carbon; some gas streams do not require additional steam for reforming ATR1: Stoichiometric Reactor T = 1000°C Δp = -0.5 bar Complete combustion of CH_4, C_2H_4 and C_2H_6 using stoichiometric amount of oxygen ATR2: Gibbs free energy minimisation Reactor 2% of CH_4 is inert Ar and N_2 are inert; C_2H_4 and C_2H_6 react completely CO_2 recycling if ratio $(H_2 - CO_2) / (CO + CO_2) > 2.1$ after downstream CO_2 removal unit

Shift	
Partial Shift Reactor	Part of stream splits to Shift reactor such that ratio (H_2 - CO_2) / (CO + CO_2) = ~2.1 after downstream CO_2 removal unit. Modelled as Gibbs free energy minimisation Reactor $T_{in, syngas}$ = 330°C Q = 0 W Δp = -0.5 bar Inertia: CH_4, C_2H_4, C_2H_6, Ar, N_2 Steam injected is 3 times CO – H_2O
Chemical Reactors	
Conventional Solid Bed Methanol	Modelled as Gibbs free energy minimisation Reactor p_{in} = 106 bar; Δp = -8 bar Q = 0 W $T_{in, syngas}$ = 50°C and 250°C Inertia: CH_4, C_2H_4, C_2H_6 T = ~260°C by adjusting cold / hot feed ratio Recycle to Feed ratio = ~5
Purification	
Methanol Separator	Modelled as Two Outlet Flash Drum T_{in} = 30°C Q = 0 Watt Δp = -12 bar
Methanol Distillation	Modelled as Two Outlet Separator
CO_2 Removal Unit with Selexol	Modelled as Two Outlet Separator X% of CO_2 separation depending on ratio (H_2 -CO_2) / (CO + CO_2) but max 98%. 100% of H_2O separation T_{in} = 127°C Δp = -0.5 bar
Power Generation	
Steam Turbine	Steam of 108 bar and 520°C is expanded to 12 bar for drier $\eta_{isentropic}$ = 0.89 η_{mech} = 0.99
Air Composition	Molar fraction O_2 = 0.2075 H_2O = 0.0101 CO_2 = 0.0003 N_2 = 0.7729 Ar = 0.0092 T = 15 °C, p = 1 atm

Appendix B

Equipment costs and scale factors

Unit	Base Investment Cost (fob)	Scale Factor	Base Scale	Overall installation factor	Maximum Size
Pre-treatment					
Conveyers	0.35	0.8	33.5 wet tonne/hour	1.86 (v)	110
grinding	0.41	0.6	33.5 wet tonne/hour	1.86 (v)	110
storage	1.0	0.65	33.5 wet tonne/hour	1.86 (v)	110
dryer	7.6	0.8	33.5 wet tonne/hour	1.86 (v)	110
iron removal	0.37	0.7	33.5 wet tonne/hour	1.86 (v)	110
feeding system	0.41	1	33.5 wet tonne/hour	1.86 (v)	110
Gasification System					
BCL	16.3	0.65	68.8 dry tonne/hour	1.69	83
IGT	38.1	0.7	68.8 dry tonne/hour	1.69	75
Oxygen Plant (installed)	44.2	0.85	41.7 tonne O_2/hour	1	-
Gas Cleaning					
Tar Cracker	3.1	0.7	34.2 m^3 gas/s	1.86 (v)	52
Cyclones	2.6	0.7	34.2 m^3 gas/s	1.86 (v)	180
High-temperature heat exchanger	6.99	0.6	39.2 kg steam/s	1.84 (v)	-
Baghouse filter	1.6	0.65	12.1 m^3 gas/s	1.86 (v)	64
Condensing Scrubber	2.6	0.7	12.1 m^3 gas/s	1.86 (v)	64
Hot Gas Cleaning	30	1.0	74.1 m^3 gas/s	1.72 (v)	-
Syngas Processing					
Compressor	11.1	0.85	13.2 MW_e	1.72 (v)	-
Steam Reformer	9.4	0.6	1390 kmol total/hour	2.3 (v)	-
Autothermal Reformer	4.7	0.6	1390 kmol total/hour	2.3 (v)	-
Shift Reactor (installed)	36.9	0.85	15.6 Mmol $CO+H_2$/hour	1	-
Selexol CO_2 removal (installed)	54.1	0.7	9909 kmol CO_2/hour	1	-
Methanol Production					
Gas Phase Methanol	7	0.6	87.5 tonne MeOH/hour	2.1 (v)	-
Liquid Phase Methanol	3.5	0.72	87.5 tonne MeOH/hour	2.1 (v)	-
Refining	15.1	0.7	87.5 tonne MeOH/hour	2.1 (v)	
Hydrogen Production					
PSA units A+B	28.0	0.7	9600 kmol feed/hour	1.69	-
Ceramic Membrane (installed)	21.6	0.8	17 tonne H_2/hr	1	-
Power Isle					
Gas Turbine + HRSG	18.9	0.7	26.3 MW_e	1.86 (v)	-
Steam Turbine + steam system	5.1	0.7	10.3 MW_e	1.86 (v)	-
Expansion Turbine	4.3	0.7	10.3 MW_e	1.86 (v)	-

Figure 29 Costs of system components in $MUS\$_{2001}$ (for detailed description see the reference) [8].

Appendix C
Energy balances

C.1 Energy balance for the IGT system

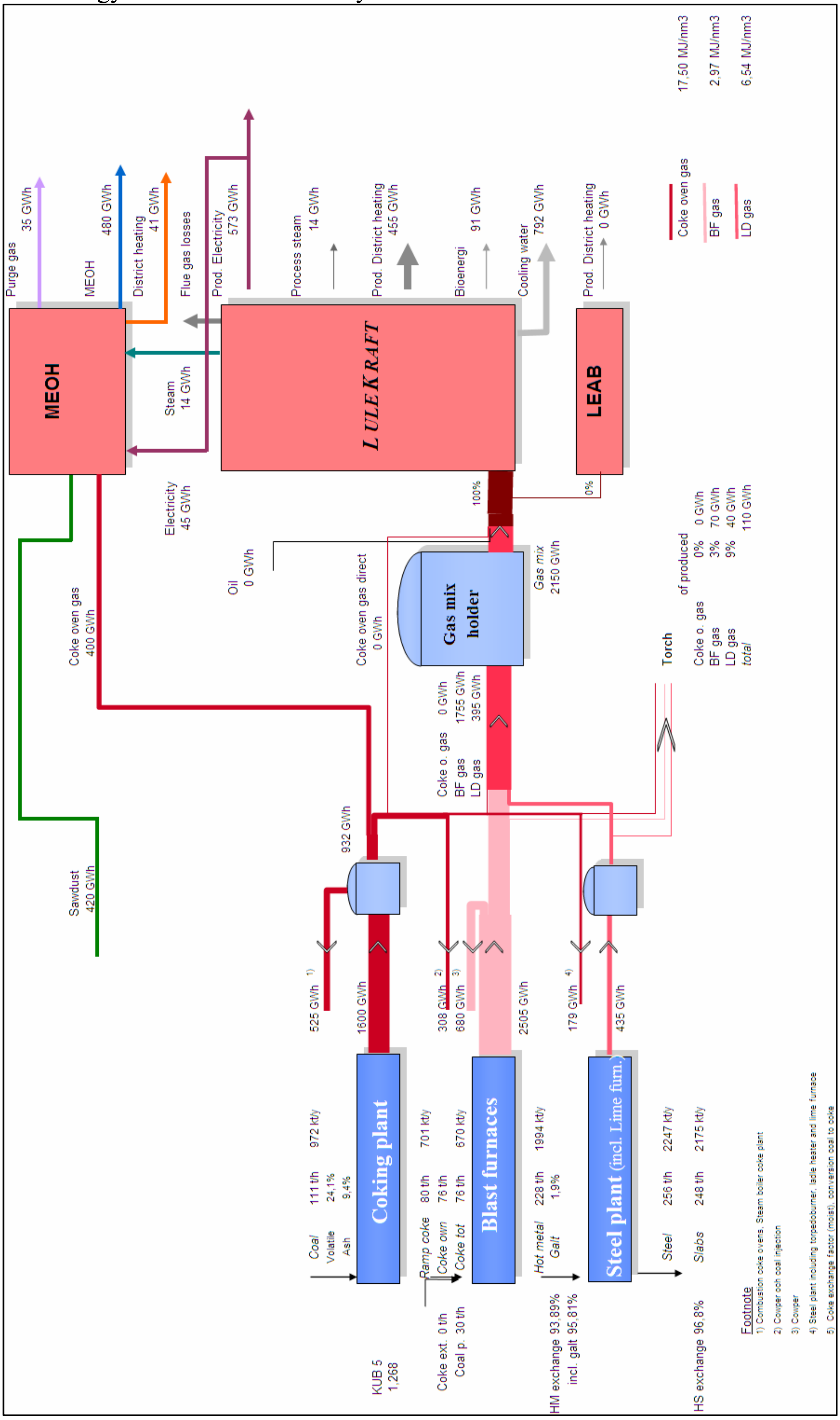

Figure 30 Energy balance for the IGT system

C.2 Energy balance for the BCL system

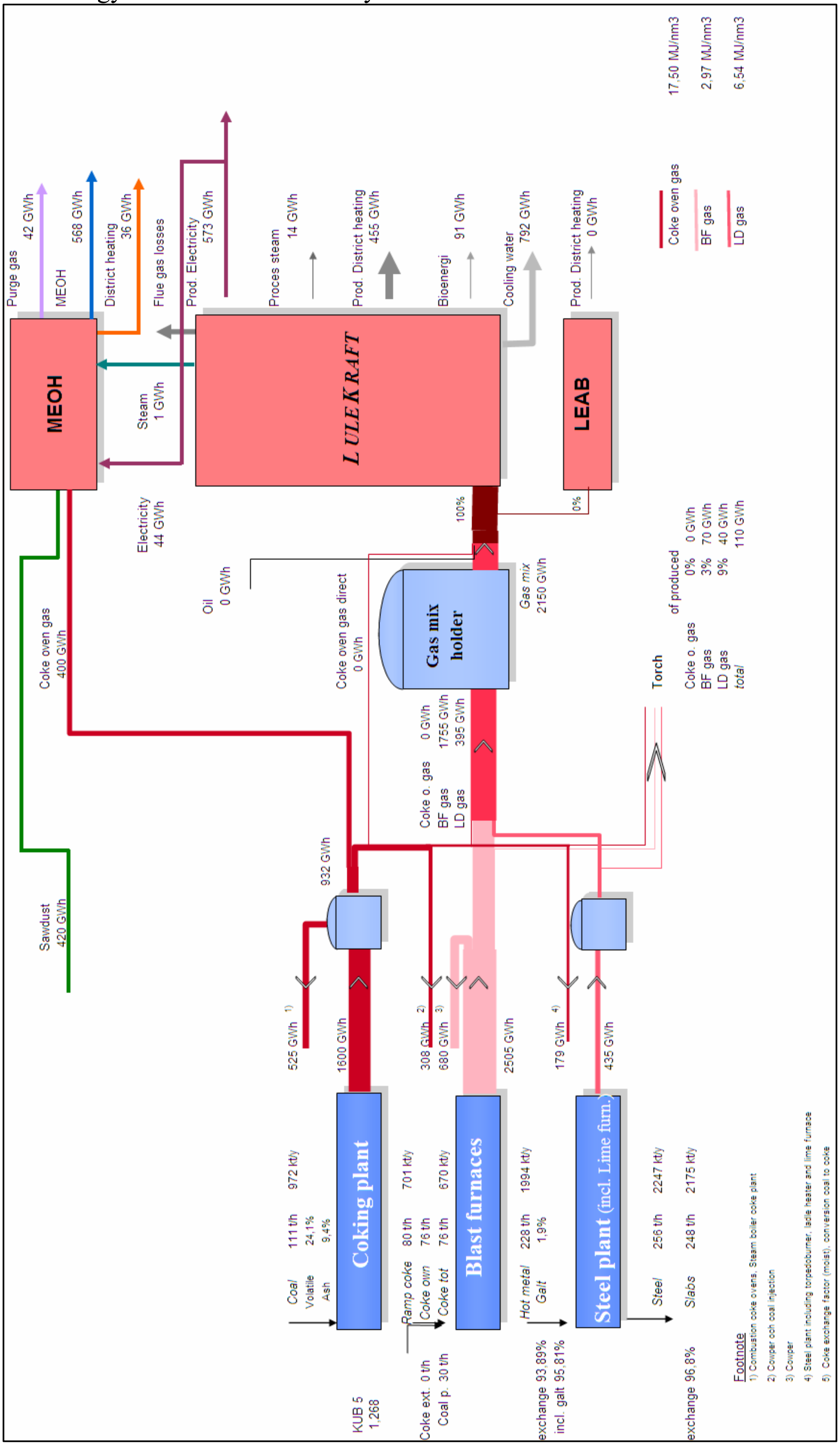

Figure 31 Energy balance for the BCL system

C.3 Energy balance for the ATR system

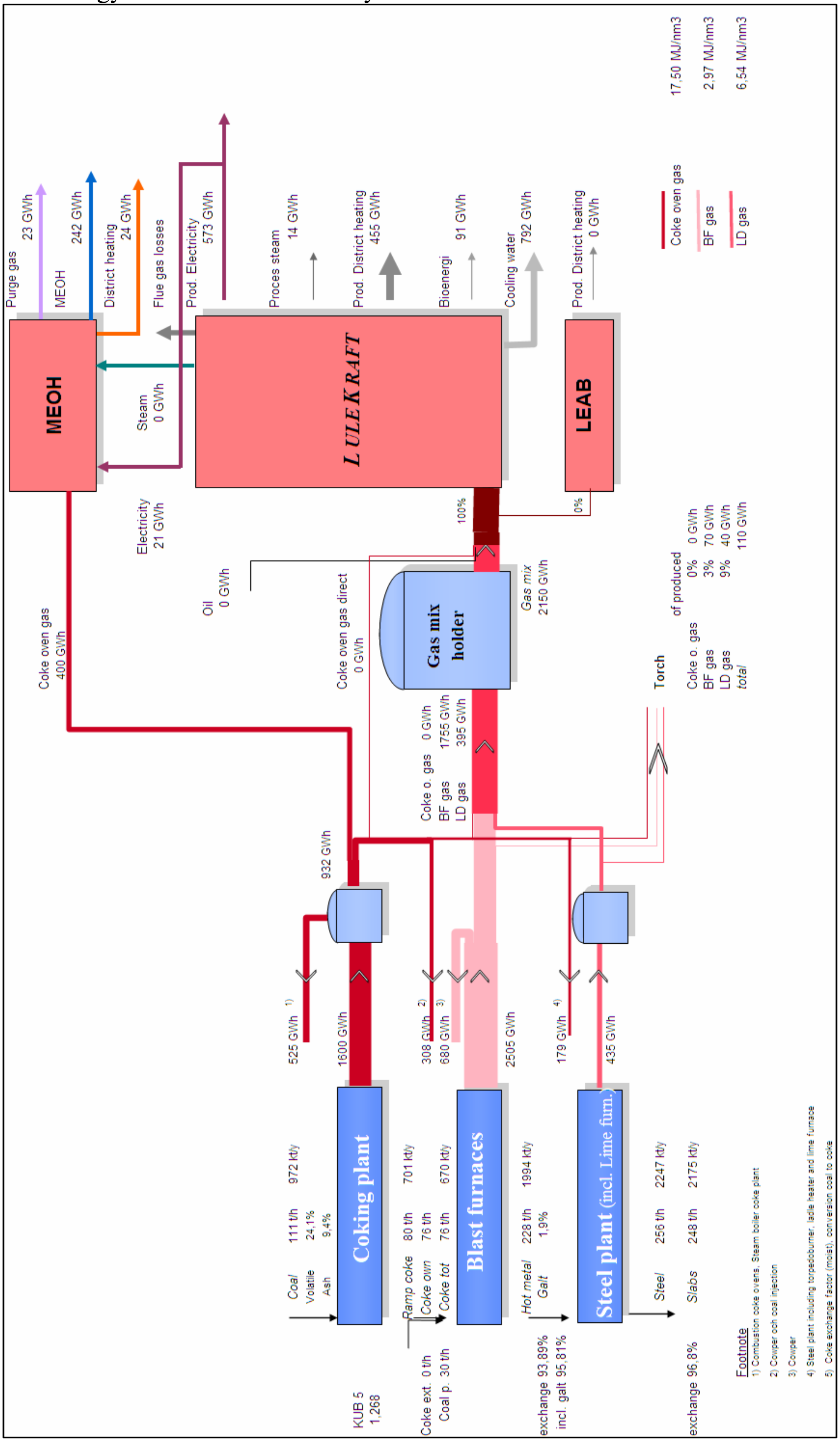

Figure 32 Energy balance for the ATR system

C.4 Energy balance for the SMR system

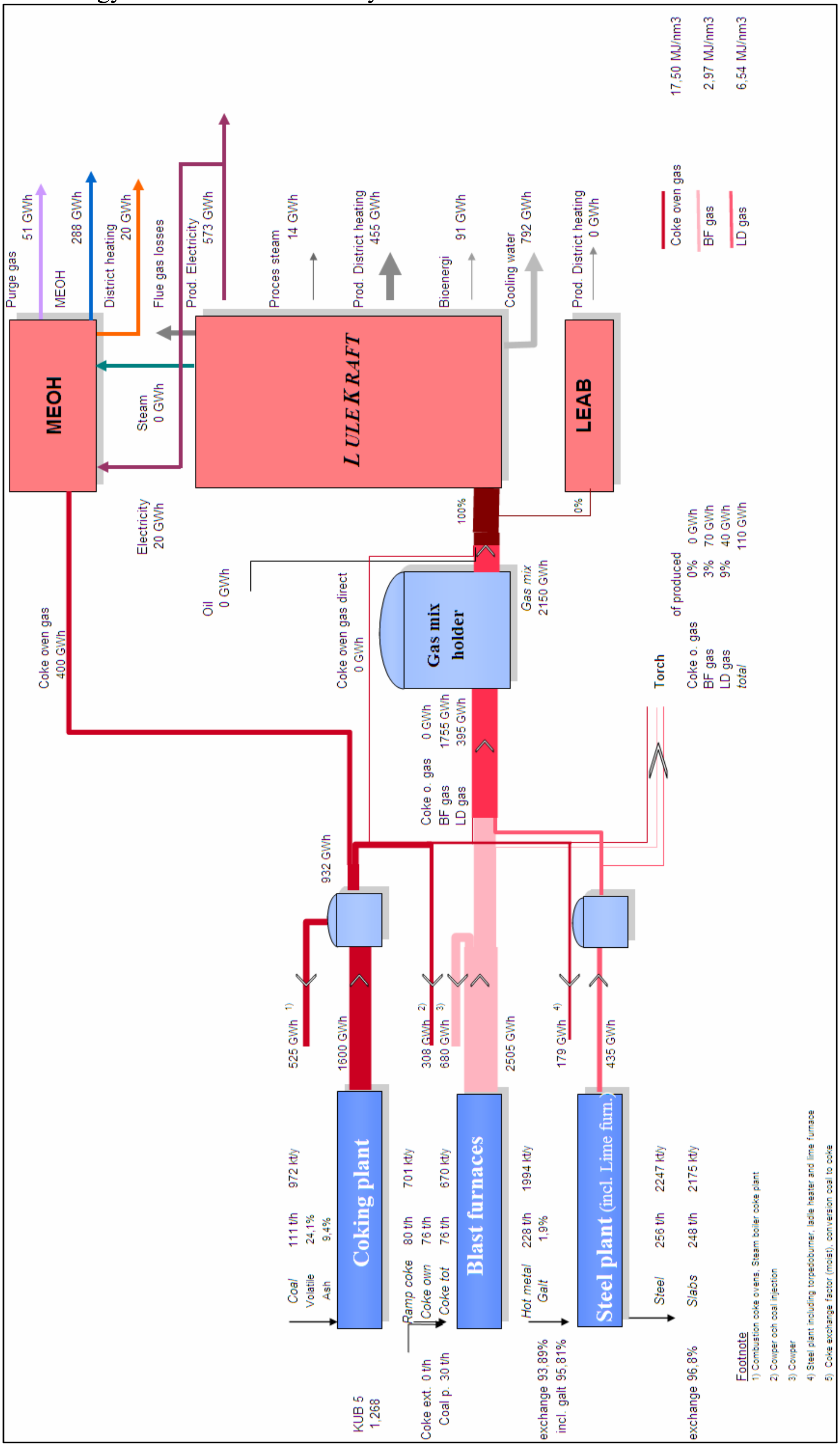

Figure 33 Energy balance for the SMR system

Appendix D

Aspen plus stream results for the systems

D.1 Stream result for the IGT methanol system

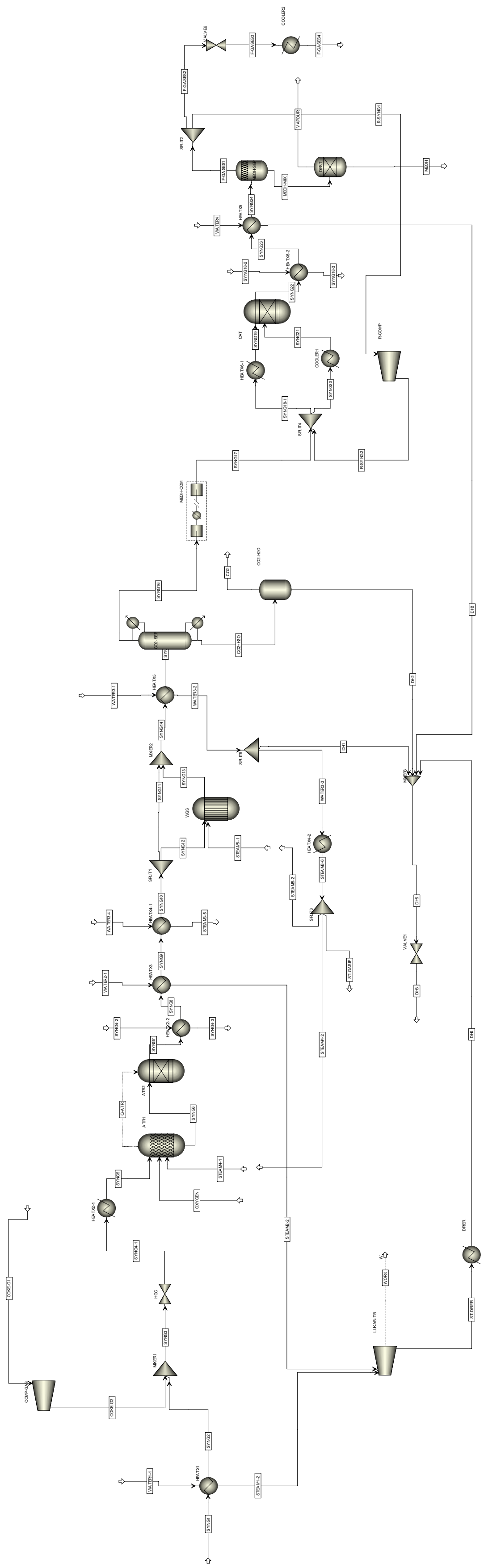

Figure 34 Aspen plus flow sheet for the IGT system

Table 41 Aspen plus stream results for the IGT system

Substream: MIXED									
Mole Flow kmol/hr	**SYNG1**	**SYNG2**	**COKE-G1**	**COKE-G2**	**SYNG3**	**SYNG4-1**	**SYNG4-2**	**SYNG4-3**	**SYNG5**
H2O	288,9698	288,9698	11,53116	11,53116	300,5009	300,5009	300,5009	300,5009	300,5009
O2	0	0	1,411156	1,411156	1,411156	1,411156	1,411156	1,411156	1,411156
N2	0	0	24,19124	24,19124	24,19124	24,19124	24,19124	24,19124	24,19124
CH4	74,42335	74,42335	96,76495	96,76495	171,1883	171,1883	171,1883	171,1883	171,1883
CO	136,3065	136,3065	22,57849	22,57849	158,885	158,885	158,885	158,885	158,885
CO2	217,1817	217,1817	5,241435	5,241435	222,4231	222,4231	222,4231	222,4231	222,4231
C2H2	2,817001	2,817001	0	0	2,817001	2,817001	2,817001	2,817001	2,817001
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	12,49881	12,49881	12,49881	12,49881	12,49881	12,49881	12,49881
C3H8	0	0	0	0	0	0	0	0	0
H2	189,0117	189,0117	235,8646	235,8646	424,8762	424,8762	424,8762	424,8762	424,8762
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	4,838248	4,838248	4,838248	4,838248	4,838248	4,838248	4,838248
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0,8063746	0,8063746	0,8063746	0,8063746	0,8063746	0,8063746	0,8063746
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	5205,872	5205,872	207,737	207,737	5413,609	5413,609	5413,609	5413,609	5413,609
O2	0	0	45,15528	45,15528	45,15528	45,15528	45,15528	45,15528	45,15528
N2	0	0	677,6808	677,6808	677,6808	677,6808	677,6808	677,6808	677,6808
CH4	1193,956	1193,956	1552,377	1552,377	2746,333	2746,333	2746,333	2746,333	2746,333
CO	3818	3818	632,4325	632,4325	4450,432	4450,432	4450,432	4450,432	4450,432
CO2	9558,123	9558,123	230,6745	230,6745	9788,797	9788,797	9788,797	9788,797	9788,797
C2H2	73,34873	73,34873	0	0	73,34873	73,34873	73,34873	73,34873	73,34873
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	375,8346	375,8346	375,8346	375,8346	375,8346	375,8346	375,8346
C3H8	0	0	0	0	0	0	0	0	0
H2	381,0249	381,0249	475,4747	475,4747	856,4995	856,4995	856,4995	856,4995	856,4995
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	164,8966	164,8966	164,8966	164,8966	164,8966	164,8966	164,8966
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	48,44408	48,44408	48,44408	48,44408	48,44408	48,44408	48,44408
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	908,71	908,71	415,7264	415,7264	1324,436	1324,436	1324,436	1324,436	1324,436
Total Flow kg/hr	20230,32	20230,32	4410,707	4410,707	24641,03	24641,03	24641,03	24641,03	24641,03
Total Flow cum/hr	2765,098	1808,668	10286	834,4774	2643,948	3088,684	3088,684	3653,083	3653,083
Temperature C	982	550	25	552,3364	550,3508	550,1643	550,1643	700	700
Pressure bar	34,5	34,5	1	34,5	34,5	29,5	29,5	29,5	29,5
Vapor Frac	1	1	0,9982924	1	1	1	1	1	1
Liquid Frac	0	0	1,71E-03	0	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-36,87039	-41,43187	-9,108616	-4,491236	-29,8366	-29,8366	-29,8366	-28,31498	-28,31498
Enthalpy kcal/kg	-1656,152	-1861,045	-858,5227	-423,3166	-1603,694	-1603,694	-1603,694	-1521,908	-1521,908
Enthalpy Gcal/hr	-33,50497	-37,65009	-3,786747	-1,867152	-39,51724	-39,51724	-39,51724	-37,50193	-37,50193
Entropy cal/mol-K	7,789305	3,351908	-2,099471	-0,4117596	2,583207	2,89582	2,89582	4,592649	4,592649
Entropy cal/gm-K	0,3498817	0,1505617	-0,1978834	-0,0388099	0,1388454	0,1556481	0,1556481	0,2468514	0,2468514
Density mol/cc	3,29E-04	5,02E-04	4,04E-05	4,98E-04	5,01E-04	4,29E-04	4,29E-04	3,63E-04	3,63E-04
Density kg/cum	7,316314	11,1852	0,4288068	5,285592	9,319788	7,977841	7,977841	6,74527	6,74527

Substream: MIXED

Mole Flow kmol/hr	**OXYGEN**	**STEAM4-1**	**SYNG6**	**SYNG7**	**SYNG8**	**SYNG9**	**SYNG10**	**SYNG11**	**SYNG12**
H2O	0	394	926,5994	797,1157	797,1157	797,1157	797,1157	717,4042	79,71157
O2	146	0	0	4,20E-13	4,20E-13	4,20E-13	4,20E-13	3,78E-13	4,20E-14
N2	0	0	24,19124	24,19124	24,19124	24,19124	24,19124	21,77211	2,419124
CH4	0	0	122,281	20,78776	20,78776	20,78776	20,78776	18,70899	2,078776
CO	0	0	214,934	329,4245	329,4245	329,4245	329,4245	296,4821	32,94245
CO2	0	0	224,2631	233,5984	233,5984	233,5984	233,5984	210,2386	23,35984
C2H2	0	0	2,012203	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	8,927982	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	303,492	662,6931	662,6931	662,6931	662,6931	596,4238	66,26931
C	0	0	0	3,13E-21	3,13E-21	3,13E-21	3,13E-21	2,81E-21	3,13E-22
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	1,61263	2,23E-05	2,23E-05	2,23E-05	2,23E-05	2,01E-05	2,23E-06
H2S	0	0	3,455993	5,52095	5,52095	5,52095	5,52095	4,968855	0,552095
CH3OH	0	0	0	5,22E-05	5,22E-05	5,22E-05	5,22E-05	4,70E-05	5,22E-06
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	1,01E-07	1,01E-07	1,01E-07	1,01E-07	9,08E-08	1,01E-08
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	2,31E-13	2,31E-13	2,31E-13	2,31E-13	2,08E-13	2,31E-14
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0,5759988	0,1236501	0,1236501	0,1236501	0,1236501	0,1112851	0,012365
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	0	7098,02	16692,95	14360,26	14360,26	14360,26	14360,26	12924,24	1436,026
O2	4671,825	0	0	1,34E-11	1,34E-11	1,34E-11	1,34E-11	1,21E-11	1,34E-12
N2	0	0	677,6808	677,6808	677,6808	677,6808	677,6808	609,9127	67,76808
CH4	0	0	1961,724	333,4931	333,4931	333,4931	333,4931	300,1438	33,34931
CO	0	0	6020,387	9227,312	9227,312	9227,312	9227,312	8304,581	922,7312
CO2	0	0	9869,774	10280,62	10280,62	10280,62	10280,62	9252,558	1028,062
C2H2	0	0	52,3935	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	268,4612	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	611,8034	1335,91	1335,91	1335,91	1335,91	1202,319	133,591
C	0	0	0	3,76E-20	3,76E-20	3,76E-20	3,76E-20	3,38E-20	3,76E-21
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	103,3128	1,43E-03	1,43E-03	1,43E-03	1,43E-03	1,29E-03	1,43E-04
H2S	0	0	117,7867	188,1643	188,1643	188,1643	188,1643	169,3479	18,81643
CH3OH	0	0	0	1,67E-03	1,67E-03	1,67E-03	1,67E-03	1,50E-03	1,67E-04
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	3,23E-06	3,23E-06	3,23E-06	3,23E-06	2,91E-06	3,23E-07
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	1,85E-11	1,85E-11	1,85E-11	1,85E-11	1,66E-11	1,85E-12
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	34,60393	7,428451	7,428451	7,428451	7,428451	6,685606	0,7428451
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	146	394	1832,345	2073,455	2073,455	2073,455	2073,455	1866,11	207,3455
Total Flow kg/hr	4671,825	7098,02	36410,88	36410,88	36410,88	36410,88	36410,88	32769,79	3641,088
Total Flow cum/hr	216,0519	519,4818	7770,191	7663,471	7060,024	5207,296	3565,858	3209,272	356,5858
Temperature C	250	250	1200	1010,182	909,095	600	330	330	330
Pressure bar	29,5	29,5	29	29	29	29	29	29	29
Vapor Frac	1	1	1	1	1	1	1	1	1
Liquid Frac	0	0	0	0	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	1,60613	-56,23615	-33,67689	-28,65894	-29,63088	-32,47998	-34,81805	-34,81805	-34,81805
Enthalpy kcal/kg	50,19346	-3121,581	-1694,76	-1632,013	-1687,362	-1849,606	-1982,75	-1982,75	-1982,75
Enthalpy Gcal/hr	0,2344984	-22,15736	-61,70859	-59,42389	-61,4392	-67,34675	-72,19471	-64,97524	-7,219471
Entropy cal/mol-K	-2,689653	-13,1068	6,786015	7,568983	6,780234	3,991441	0,7933006	0,7933006	0,7933006
Entropy cal/gm-K	-0,0840548	-0,7275378	0,3415003	0,4310237	0,3861075	0,2272968	0,0451753	0,0451753	0,0451753
Density mol/cc	6,76E-04	7,58E-04	2,36E-04	2,71E-04	2,94E-04	3,98E-04	5,81E-04	5,81E-04	5,81E-04
Density kg/cum	21,62363	13,66366	4,685969	4,751225	5,157331	6,992281	10,21097	10,21097	10,21097

Substream: MIXED

Mole Flow kmol/hr	STEAM5-1	SYNG13	SYNG14	SYNG15	SYNG16	CO2+H2O	CO2	SYNG17	SYNG18-1
H2O	17	71,34946	788,7536	788,7536	0	788,7536	25,9803	0	0,1080464
O2	0	1,31E-30	3,78E-13	3,78E-13	0	0	0	0	0
N2	0	2,419124	24,19124	24,19124	24,19124	0	0	24,19124	501,8052
CH4	0	2,078776	20,78776	20,78776	20,78776	0	0	20,78776	386,1728
CO	0	7,584764	304,0668	304,0668	304,0668	0	0	304,0668	138,2453
CO2	0	48,72196	258,9605	258,9605	23,30645	235,6541	234,8199	23,30645	36,76239
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	91,61197	688,0358	688,0358	688,0358	0	0	688,0358	805,114
C	0	0	2,81E-21	2,81E-21	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	3,88E-14	2,01E-05	2,01E-05	2,01E-05	0	0	2,01E-05	6,02E-06
H2S	0	0,5615373	5,530392	5,530392	0	5,530392	5,306697	0	0,2107137
CH3OH	0	5,01E-03	5,06E-03	5,06E-03	5,06E-03	0	0	5,06E-03	7,519984
CL2	0	0	0	0	0	0	0	0	0
S	0	9,24E-19	9,08E-08	9,08E-08	9,08E-08	0	0	9,08E-08	2,72E-08
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	3,33E-27	2,08E-13	2,08E-13	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	2,92E-03	0,11421	0,11421	0,11421	0	0	0,11421	0,0359282
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	306,2598	1285,381	14209,62	14209,62	0	14209,62	468,0424	0	1,946487
O2	0	4,21E-29	1,21E-11	1,21E-11	0	0	0	0	0
N2	0	67,76808	677,6808	677,6808	677,6808	0	0	677,6808	14057,31
CH4	0	33,34931	333,4931	333,4931	333,4931	0	0	333,4931	6195,278
CO	0	212,4523	8517,033	8517,033	8517,033	0	0	8517,033	3872,307
CO2	0	2144,244	11396,8	11396,8	1025,712	10371,09	10334,38	1025,712	1617,905
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	184,6787	1386,998	1386,998	1386,998	0	0	1386,998	1623,013
C	0	0	3,38E-20	3,38E-20	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	2,48E-12	1,29E-03	1,29E-03	1,29E-03	0	0	1,29E-03	3,86E-04
H2S	0	19,13825	188,4862	188,4862	0	188,4862	180,8622	0	7,18152
CH3OH	0	0,1606238	0,1621286	0,1621286	0,1621286	0	0	0,1621286	240,9565
CL2	0	0	0	0	0	0	0	0	0
S	0	2,96E-17	2,91E-06	2,91E-06	2,91E-06	0	0	2,91E-06	8,73E-07
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	2,67E-25	1,66E-11	1,66E-11	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0,175717	6,861323	6,861323	6,861323	0	0	6,861323	2,15844
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	17	224,3355	2090,445	2090,445	1060,507	1029,938	266,1069	1060,507	1875,974
Total Flow kg/hr	306,2598	3947,347	36717,14	36717,14	11947,94	24769,19	10983,28	11947,94	27618,06
Total Flow cum/hr	18,73865	466,8415	3735,695	1682,006	1252,178	304,8929	295,0509	315,4667	496,5933
Temperature C	250	439,2197	342,0761	127	127	127	126,8197	91,03431	56,96972
Pressure bar	34,5	28,5	28,5	28,5	28,5	28,5	28	106,3908	106
Vapor Frac	1	1	1	0,684452	1	0,2580892	1	1	1
Liquid Frac	0	0	0	0,315548	0	0,7419108	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-56,29906	-36,44713	-34,99287	-39,7871	-9,26898	-72,19705	-87,91755	-9,528862	-7,48615
Enthalpy kcal/kg	-3125,073	-2071,362	-1992,277	-2265,23	-822,7209	-3002,055	-2130,098	-845,7882	-508,5016
Enthalpy Gcal/hr	-0,9570977	-8,176504	-73,15174	-83,17396	-9,829963	-74,35956	-23,3958	-10,10557	-14,04403
Entropy cal/mol-K	-13,50207	0,383613	0,8168823	-9,684124	2,939261	-26,3658	-3,604251	-0,4210728	-8,588238
Entropy cal/gm-K	-0,749479	0,0218014	0,0465082	-0,5513538	0,2608908	-1,096327	-0,0873251	-0,0373747	-0,5833617
Density mol/cc	9,07E-04	4,81E-04	5,60E-04	1,24E-03	8,47E-04	3,38E-03	9,02E-04	3,36E-03	3,78E-03
Density kg/cum	16,34375	8,455433	9,828729	21,82937	9,541727	81,239	37,22504	37,87386	55,61504

Substream: MIXED

Mole Flow kmol/hr	**SYNG18-2**	**SYNG18-3**	**SYNG19**	**SYNG20**	**SYNG21**	**SYNG22**	**SYNG23**	**SYNG24**	**F-GASES1**
H2O	0,1080464	0,1080464	0,1080464	0,2521084	0,2521084	19,41724	19,41724	19,41724	0,3648894
O2	0	0	0	0	0	0	0	0	0
N2	501,8052	501,8052	501,8052	1170,879	1170,879	1672,684	1672,684	1672,684	1670,041
CH4	386,1728	386,1728	386,1728	901,07	901,07	1287,243	1287,243	1287,243	1283,011
CO	138,2453	138,2453	138,2453	322,5724	322,5724	158,9008	158,9008	158,9008	158,7959
CO2	36,76239	36,76239	36,76239	85,7789	85,7789	103,4842	103,4842	103,4842	100,5302
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	805,114	805,114	805,114	1878,599	1878,599	2022,367	2022,367	2022,367	2021,832
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	6,02E-06	6,02E-06	6,02E-06	1,40E-05	1,40E-05	0	0	0	0
H2S	0,2107137	0,2107137	0,2107137	0,4916654	0,4916654	0,8159784	0,8159784	0,8159784	0,7115754
CH3OH	7,519984	7,519984	7,519984	17,54663	17,54663	346,1542	346,1542	346,1542	25,38931
CL2	0	0	0	0	0	0	0	0	0
S	2,72E-08	2,72E-08	2,72E-08	6,36E-08	6,36E-08	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,0359282	0,0359282	0,0359282	0,0838325	0,0838325	6,18E-03	6,18E-03	6,18E-03	5,62E-03
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	1,946487	1,946487	1,946487	4,541803	4,541803	349,8071	349,8071	349,8071	6,573584
O2	0	0	0	0	0	0	0	0	0
N2	14057,31	14057,31	14057,31	32800,39	32800,39	46857,7	46857,7	46857,7	46783,67
CH4	6195,278	6195,278	6195,278	14455,65	14455,65	20650,93	20650,93	20650,93	20583,03
CO	3872,307	3872,307	3872,307	9035,383	9035,383	4450,874	4450,874	4450,874	4447,938
CO2	1617,905	1617,905	1617,905	3775,112	3775,112	4554,321	4554,321	4554,321	4424,313
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	1623,013	1623,013	1623,013	3787,031	3787,031	4076,85	4076,85	4076,85	4075,771
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	3,86E-04	3,86E-04	3,86E-04	9,00E-04	9,00E-04	0	0	0	0
H2S	7,18152	7,18152	7,18152	16,75688	16,75688	27,81008	27,81008	27,81008	24,25183
CH3OH	240,9565	240,9565	240,9565	562,2319	562,2319	11091,53	11091,53	11091,53	813,5282
CL2	0	0	0	0	0	0	0	0	0
S	8,73E-07	8,73E-07	8,73E-07	2,04E-06	2,04E-06	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	2,158439	2,158439	2,15844	5,036359	5,036359	0,3713739	0,3713739	0,3713739	0,3378223
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	1875,974	1875,974	1875,974	4377,274	4377,274	5611,073	5611,073	5611,073	5260,681
Total Flow kg/hr	27618,06	27618,06	27618,06	64442,13	64442,13	92060,19	92060,19	92060,19	81159,41
Total Flow cum/hr	496,5933	798,1622	798,1622	1158,718	1132,159	2605,522	2308,231	1371,626	1538,259
Temperature C	56,96972	250	250	56,96972	50	260,1489	201,4628	30	29,06707
Pressure bar	106	106	106	106	106	98	98	98	86
Vapor Frac	1	1	1	1	1	1	1	0,937377	1
Liquid Frac	0	0	0	0	0	0	0	0,0626229	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-7,48615	-5,925874	-5,925874	-7,48615	-7,542015	-7,864282	-8,385936	-10,36089	-7,232406
Enthalpy kcal/kg	-508,5016	-402,5188	-402,5188	-508,5016	-512,2963	-479,3283	-511,1232	-631,4966	-468,7982
Enthalpy Gcal/hr	-14,04403	-11,11695	-11,11695	-32,7694	-33,01394	-44,12769	-47,05477	-58,13653	-38,04793
Entropy cal/mol-K	-8,588238	-4,871521	-4,871521	-8,588238	-8,759276	-7,428877	-8,46496	-13,74822	-10,64595
Entropy cal/gm-K	-0,5833617	-0,3309012	-0,3309012	-0,5833617	-0,5949795	-0,4527904	-0,5159397	-0,8379548	-0,6900613
Density mol/cc	3,78E-03	2,35E-03	2,35E-03	3,78E-03	3,87E-03	2,15E-03	2,43E-03	4,09E-03	3,42E-03
Density kg/cum	55,61504	34,60206	34,60206	55,61504	56,9197	35,33272	39,88344	67,11755	52,76058

Substream: MIXED

Mole Flow kmol/hr	F-GASES2	F-GASES3	F-GASES4	R-SYNG1	R-SYNG2	MEOH-MIX	MEOH	VAPOUR
H2O	4,71E-03	4,71E-03	4,71E-03	0,3601548	0,3601548	19,05235	0	19,05235
O2	0	0	0	0	0	0	0	0
N2	21,53786	21,53786	21,53786	1648,493	1648,493	2,642551	0	2,642551
CH4	16,54648	16,54648	16,54648	1266,455	1266,455	4,232295	0	4,232295
CO	2,047928	2,047928	2,047928	156,7509	156,7509	0,1048278	0	0,1048278
CO2	1,296498	1,296498	1,296498	99,23484	99,23484	2,954069	0	2,954069
C2H2	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0
H2	26,07476	26,07476	26,07476	1995,677	1995,677	0,5353051	0	0,5353051
C	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
H2S	9,18E-03	9,18E-03	9,18E-03	0,7023791	0,7023791	0,104403	0	0,104403
CH3OH	0,3274358	0,3274358	0,3274358	25,06155	25,06155	320,7649	320,7649	0
CL2	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0
COS	7,25E-05	7,25E-05	7,25E-05	5,55E-03	5,55E-03	5,58E-04	0	5,58E-04
HCN	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0
Mass Flow kg/hr								
H2O	0,0847769	0,0847769	0,0847769	6,488289	6,488289	343,2335	0	343,2335
O2	0	0	0	0	0	0	0	0
N2	603,3504	603,3504	603,3504	46180,02	46180,02	74,02704	0	74,02704
CH4	265,4512	265,4512	265,4512	20317,43	20317,43	67,89769	0	67,89769
CO	57,36329	57,36329	57,36329	4390,656	4390,656	2,93627	0	2,93627
CO2	57,0586	57,0586	57,0586	4367,305	4367,305	130,008	0	130,008
C2H2	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0
H2	52,5636	52,5636	52,5636	4023,046	4023,046	1,079111	0	1,079111
C	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
H2S	0,3127662	0,3127662	0,3127662	23,9384	23,9384	3,558252	0	3,558252
CH3OH	10,49175	10,49175	10,49175	803,0263	803,0263	10278	10278	0
CL2	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0
COS	4,36E-03	4,36E-03	4,36E-03	0,3334759	0,3334759	0,0335516	0	0,0335516
HCN	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0
Total Flow kmol/hr	67,84492	67,84492	67,84492	5192,741	5192,741	350,3913	320,7649	29,62636
Total Flow kg/hr	1046,681	1046,681	1046,681	80112,24	80112,24	10900,77	10278	622,7734
Total Flow cum/hr	19,83831	1634,686	1540,145	1518,411	1339,09	13,9522	13,04205	3,006385
Temperature C	29,06707	16,7219	0	29,06707	51,10073	29,06707	29,06707	29,06707
Pressure bar	86	1	1	86	106	86	86	86
Vapor Frac	1	1	1	1	1	0	0	0,3569852
Liquid Frac	0	0	0	0	0	1	1	0,6430148
Solid Frac	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-7,232406	-7,232406	-7,354842	-7,232662	-7,06897	-57,33104	-57,45826	-56,26165
Enthalpy kcal/kg	-468,7982	-468,7982	-476,7344	-468,809	-458,1987	-1842,832	-1793,208	-2676,46
Enthalpy Gcal/hr	-0,490689	-0,490689	-0,4989959	-37,55788	-36,70785	-20,08859	-18,43086	-1,666852
Entropy cal/mol-K	-10,64595	-1,820168	-2,255208	-10,64609	-10,54183	-56,37895	-59,34761	-30,68624
Entropy cal/gm-K	-0,6900613	-0,1179817	-0,1461806	-0,6900613	-0,6833039	-1,812228	-1,852173	-1,459795
Density mol/cc	3,42E-03	4,15E-05	4,41E-05	3,42E-03	3,88E-03	0,0251136	0,0245946	9,85E-03
Density kg/cum	52,76058	0,6402947	0,6795987	52,76058	59,82587	781,2942	788,0665	207,1502

Substream: MIXED

Mole Flow kmol/hr	**WATER1-1**	**STEAM1-2**	**WATER2-1**	**STEAM2-2**	**WATER3-1**	**WATER3-2**	**WATER3-3**	**WATER3-4**	**STEAM3-5**
H2O	332,735	332,735	474,1441	474,1441	2220,337	2220,337	595,55	595,55	595,55
O2	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	5994,315	5994,315	8541,838	8541,838	40000	40000	10729	10729	10729
O2	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	332,735	332,735	474,1441	474,1441	2220,337	2220,337	595,55	595,55	595,55
Total Flow kg/hr	5994,315	5994,315	8541,838	8541,838	40000	40000	10729	10729	10729
Total Flow cum/hr	6,801528	184,7742	9,692108	263,3747	39,85842	52,91174	14,19225	14,19225	665,2634
Temperature C	134	519,989	134	520,1452	15	234,7314	234,7314	234,7314	254,733
Pressure bar	108	108	108	108	34,5	34,5	34,5	34,5	34,5
Vapor Frac	0	1	0	1	0	0	0	0	1
Liquid Frac	1	0	1	0	1	1	1	1	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-66,54437	-54,08682	-66,54437	-54,08513	-68,90585	-64,39208	-64,39208	-64,39208	-56,2519
Enthalpy kcal/kg	-3693,774	-3002,275	-3693,774	-3002,181	-3824,856	-3574,304	-3574,304	-3574,304	-3122,455
Enthalpy Gcal/hr	-22,14196	-17,99684	-31,55207	-25,64451	-152,9964	-142,9742	-38,34925	-38,34925	-33,5013
Entropy cal/mol-K	-34,03407	-12,12258	-34,03407	-12,12045	-40,77714	-29,23057	-29,23057	-29,23057	-13,41234
Entropy cal/gm-K	-1,889178	-0,6729052	-1,889178	-0,672787	-2,263475	-1,622543	-1,622543	-1,622543	-0,7444978
Density mol/cc	0,0489206	1,80E-03	0,0489206	1,80E-03	0,0557056	0,041963	0,041963	0,041963	8,95E-04
Density kg/cum	881,3189	32,4413	881,3189	32,43227	1003,552	755,9758	755,9758	755,9758	16,12745

Substream: MIXED

Mole Flow kmol/hr	**STEAM3-6**	**STEAM4-2**	**STEAM5-2**	**ST-GASIF**	**ST-DRIER**	**WATER6**
H2O	595,55	394	17	184,55	806,8791	4995,759
O2	0	0	0	0	0	0
N2	0	0	0	0	0	0
CH4	0	0	0	0	0	0
CO	0	0	0	0	0	0
CO2	0	0	0	0	0	0
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	0	0	0	0	0	0
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	0	0	0	0	0	0
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Mass Flow kg/hr						
H2O	10729	7098,02	306,2598	3324,72	14536,15	90000
O2	0	0	0	0	0	0
N2	0	0	0	0	0	0
CH4	0	0	0	0	0	0
CO	0	0	0	0	0	0
CO2	0	0	0	0	0	0
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	0	0	0	0	0	0
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	0	0	0	0	0	0
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Total Flow kmol/hr	595,55	394	17	184,55	806,8791	4995,759
Total Flow kg/hr	10729	7098,02	306,2598	3324,72	14536,15	90000
Total Flow cum/hr	656,459	434,2957	18,73865	203,4246	2630,558	89,68145
Temperature C	250	250	250	250	221,4375	15
Pressure bar	34,5	34,5	34,5	34,5	12	1
Vapor Frac	1	1	1	1	1	0
Liquid Frac	0	0	0	0	0	1
Solid Frac	0	0	0	0	0	0
Enthalpy kcal/mol	-56,29906	-56,29906	-56,29906	-56,29906	-56,28722	-68,91965
Enthalpy kcal/kg	-3125,073	-3125,073	-3125,073	-3125,073	-3124,415	-3825,622
Enthalpy Gcal/hr	-33,52939	-22,18215	-0,9570977	-10,39014	-45,41763	-344,3109
Entropy cal/mol-K	-13,50207	-13,50207	-13,50207	-13,50207	-11,55341	-40,76664
Entropy cal/gm-K	-0,749479	-0,749479	-0,749479	-0,749479	-0,641312	-2,262893
Density mol/cc	9,07E-04	9,07E-04	9,07E-04	9,07E-04	3,07E-04	0,0557056
Density kg/cum	16,34375	16,34375	16,34375	16,34375	5,525881	1003,552

Substream: MIXED

Mole Flow kmol/hr	**DH1**	**DH2**	**DH3**	**DH4**	**DH5**	**DH6**
H2O	1624,787	762,7733	4995,759	806,8791	8190,199	8190,199
O2	0	0	0	0	0	0
N2	0	0	0	0	0	0
CH4	0	0	0	0	0	0
CO	0	0	0	0	0	0
CO2	0	0,834185	0	0	0,834185	0,834185
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	0	0	0	0	0	0
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0	0,2236954	0	0	0,2236954	0,2236954
CH3OH	0	0	0	0	0	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	0	0	0	0	0	0
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Mass Flow kg/hr						
H2O	29271	13741,58	90000	14536,15	1,48E+05	1,48E+05
O2	0	0	0	0	0	0
N2	0	0	0	0	0	0
CH4	0	0	0	0	0	0
CO	0	0	0	0	0	0
CO2	0	36,71231	0	0	36,71231	36,71231
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	0	0	0	0	0	0
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0	7,62396	0	0	7,62396	7,62396
CH3OH	0	0	0	0	0	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	0	0	0	0	0	0
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Total Flow kmol/hr	1624,787	763,8312	4995,759	806,8791	8191,257	8191,257
Total Flow kg/hr	29271	13785,91	90000	14536,15	1,48E+05	1,48E+05
Total Flow cum/hr	38,71949	15,49832	8153,327	17,78478	28011,42	28011,48
Temperature C	234,7314	126,8197	101,5389	188,3597	101,5047	101,5046
Pressure bar	34,5	28	1	12	1	1
Vapor Frac	0	0	0,0522339	0	0,1101537	0,110154
Liquid Frac	1	1	0,947766	1	0,8898463	0,889846
Solid Frac	0	0	0	0	0	0
Enthalpy kcal/mol	-64,39208	-66,72027	-66,70147	-65,44081	-66,12096	-66,12096
Enthalpy kcal/kg	-3574,304	-3696,747	-3702,494	-3632,517	-3669,642	-3669,642
Enthalpy Gcal/hr	-104,6249	-50,96375	-333,2292	-52,80358	-541,6215	-541,6215
Entropy cal/mol-K	-29,23057	-34,28345	-34,22944	-31,36569	-32,67029	-32,67029
Entropy cal/gm-K	-1,622543	-1,899531	-1,900023	-1,74106	-1,813166	-1,813166
Density mol/cc	0,041963	0,0492847	6,13E-04	0,045369	2,92E-04	2,92E-04
Density kg/cum	755,9758	889,5098	11,03844	817,3367	5,269031	5,26902

	LUKAB-TB	**COMP-GAS**	**MEOH-COM**	**R-COMP**	**TOTAL WORK**
Power MW	-2,045126	2,48050673	1,53889486	1,09840358	3,07267917

D.2 Stream result for the BCL methanol system

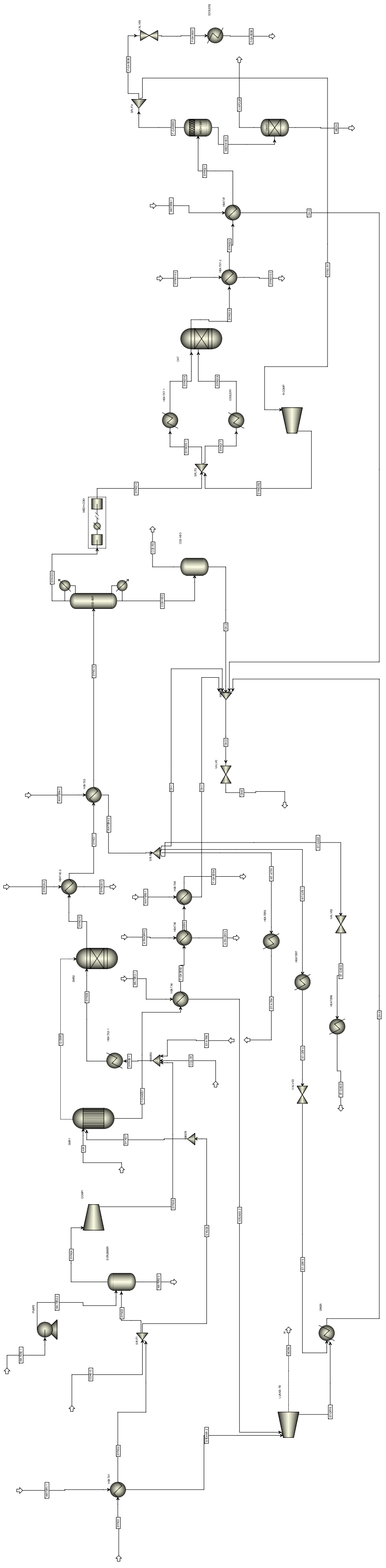

Figure 35 Aspen plus flow sheet for the BCL system

Table 42 Aspen plus stream results for the BCL system

Substream: MIXED									
Mole Flow kmol/hr	**SYNG1**	**SYNG2**	**COKE-G**	**SYNG5**	**SYNG7**	**AIR**	**F-GASES1**	**F-GASES2**	**F-GASES3**
H2O	99,89462	99,89462	13,89875	20,14143	20,14143	6,868	173,9404	173,9404	173,9404
O2	0	0	1,700896	0,3010586	0,3010586	141,1	2,053034	2,053034	2,053034
N2	0	0	29,15821	5,161004	5,161004	525,572	530,733	530,733	530,733
CH4	63,24986	63,24986	116,6329	31,83924	31,83924	0	0	0	0
CO	186,2357	186,2357	27,21433	37,78065	37,78065	0	0	0	0
CO2	44,67649	44,67649	6,317613	9,025956	9,025956	0,204	92,88462	92,88462	92,88462
C2H2	0	0	0	0	0	0	0	0	0
C2H4	21,08329	21,08329	0	3,731742	3,731742	0	0	0	0
C2H6	3,011898	3,011898	15,06508	3,199625	3,199625	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	83,83116	83,83116	284,2926	65,15791	65,15791	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	1,204234	1,204234	1,204234
H2S	0	0	5,831643	1,032201	1,032201	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0,9719405	0,1720335	0,1720335	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	6,256	6,256	6,256	6,256
Mass Flow kg/hr									
H2O	1799,63	1799,63	250,3899	362,8534	362,8534	123,7289	3133,584	3133,584	3133,584
O2	0	0	54,42663	9,633513	9,633513	4515,031	65,69461	65,69461	65,69461
N2	0	0	816,8231	144,5777	144,5777	14723,1	14867,68	14867,68	14867,68
CH4	1014,702	1014,702	1871,113	510,7893	510,7893	0	0	0	0
CO	5216,536	5216,536	762,2844	1058,251	1058,251	0	0	0	0
CO2	1966,203	1966,203	278,0369	397,2305	397,2305	8,977999	4087,833	4087,833	4087,833
C2H2	0	0	0	0	0	0	0	0	0
C2H4	591,4654	591,4654	0	104,6894	104,6894	0	0	0	0
C2H6	90,56669	90,56669	453,0015	96,21156	96,21156	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	168,9936	168,9936	573,0998	131,3505	131,3505	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	77,14903	77,14903	77,14903
H2S	0	0	198,7534	35,17934	35,17934	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	58,39069	10,33515	10,33515	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	249,9147	249,9147	249,9147	249,9147
Total Flow kmol/hr	501,983	501,983	501,0839	177,5428	177,5428	680	807,0713	807,0713	807,0713
Total Flow kg/hr	10848,1	10848,1	5316,319	2861,102	2861,102	19620,75	22481,85	22481,85	22481,85
Total Flow cum/hr	39525,88	17843,59	10286	4991,476	4991,476	16066,28	1,11E+05	53499,39	47229,52
Temperature C	863	240	25	132,7233	132,7233	15	890	287,5747	221,9375
Pressure bar	1,2	1,2	1,2	1,2	1,2	1,01325	0,70325	0,70325	0,70325
Vapor Frac	1	1	0,9938864	1	1	1	1	1	1
Liquid Frac	0	0	6,11E-03	0	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-23,56706	-29,73722	-9,157125	-19,4564	-19,4564	-0,6833247	-16,22645	-21,3673	-21,88091
Enthalpy kcal/kg	-1090,538	-1376,055	-863,0949	-1207,347	-1207,347	-23,68211	-582,5099	-767,0602	-785,498
Enthalpy Gcal/hr	-11,83043	-14,92779	-4,588554	-3,454393	-3,454393	-0,4646675	-13,09609	-17,24518	-17,65971
Entropy cal/mol-K	17,69998	9,953246	-2,622728	4,446706	4,446706	0,8757114	11,32264	5,151327	4,17733
Entropy cal/gm-K	0,8190458	0,4605748	-0,2472024	0,275936	0,275936	0,0303496	0,4064692	0,1849264	0,1499611
Density mol/cc	1,27E-05	2,81E-05	4,87E-05	3,56E-05	3,56E-05	4,23E-05	7,27E-06	1,51E-05	1,71E-05
Density kg/cum	0,2744556	0,607955	0,51685	0,5731976	0,5731976	1,221238	0,2025361	0,4202264	0,4760128

Substream: MIXED

Mole Flow kmol/hr	F-GASES4	SYNG3	SYNG4	SYNG6	CO2-R1	ST-ATR5	SYNG8-1	SYNG8-2	SYNG8-3
H2O	173,9404	93,65194	75,62385	75,62385	0	572	647,6238	647,6238	647,6238
O2	2,053034	1,399837	1,399823	1,399823	0	0	1,399823	1,399823	1,399823
N2	530,733	23,99721	23,9972	23,9972	0	0	23,9972	23,9972	23,9972
CH4	0	148,0435	148,0428	148,0428	0	0	148,0428	148,0428	148,0428
CO	0	175,6694	175,6693	175,6693	0	0	175,6693	175,6693	175,6693
CO2	92,88462	41,96814	41,96649	41,96649	110	0	151,9665	151,9665	151,9665
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	17,35154	17,35142	17,35142	0	0	17,35142	17,35142	17,35142
C2H6	0	14,87735	14,87728	14,87728	0	0	14,87728	14,87728	14,87728
C3H8	0	0	0	0	0	0	0	0	0
H2	0	302,9659	302,9656	302,9656	0	0	302,9656	302,9656	302,9656
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	1,204234	0	0	0	0	0	0	0	0
H2S	0	4,799442	4,794624	4,794624	0	0	4,794624	4,794624	4,794624
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0,799907	0,799829	0,799829	0	0	0,799829	0,799829	0,799829
HCN	0	0	0	0	0	0	0	0	0
ARGON	6,256	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	3133,584	1687,166	1362,385	1362,385	0	10304,74	11667,13	11667,12	11667,12
O2	65,69461	44,79311	44,79266	44,79266	0	0	44,79266	44,79266	44,79266
N2	14867,68	672,2454	672,245	672,245	0	0	672,245	672,245	672,245
CH4	0	2375,026	2375,016	2375,016	0	0	2375,016	2375,016	2375,016
CO	0	4920,569	4920,567	4920,567	0	0	4920,567	4920,567	4920,567
CO2	4087,833	1847,01	1846,937	1846,937	4841,078	0	6688,015	6688,015	6688,015
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	486,7761	486,7727	486,7727	0	0	486,7727	486,7727	486,7727
C2H6	0	447,3566	447,3544	447,3544	0	0	447,3544	447,3544	447,3544
C3H8	0	0	0	0	0	0	0	0	0
H2	0	610,7428	610,7423	610,7423	0	0	610,7423	610,7423	610,7423
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	77,14903	0	0	0	0	0	0	0	0
H2S	0	163,574	163,4098	163,4098	0	0	163,4098	163,4098	163,4098
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	48,05553	48,05085	48,05085	0	0	48,05085	48,05085	48,05085
HCN	0	0	0	0	0	0	0	0	0
ARGON	249,9147	0	0	0	0	0	0	0	0
Total Flow kmol/hr	807,0713	825,5241	807,4882	807,4882	110	572	1489,488	1489,488	1489,488
Total Flow kg/hr	22481,85	13303,31	12978,27	12978,27	4841,078	10304,74	28124,09	28124,09	28124,09
Total Flow cum/hr	35572,99	23208,95	18112,28	2933,156	228,6123	1519,441	4696,9	4696,9	9072,623
Temperature C	100	132,7233	50,85773	401,4812	127	250	318,905	318,905	860
Pressure bar	0,70325	1,2	1,2	15,5	15,5	15,5	15,5	15,5	15,5
Vapor Frac	1	1	1	1	1	1	1	1	1
Liquid Frac	0	0	0	0	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-22,81334	-19,4564	-19,26725	-16,28769	-93,11807	-56,07231	-37,23995	-37,23995	-31,80504
Enthalpy kcal/kg	-818,9712	-1207,347	-1198,779	-1013,395	-2115,849	-3112,486	-1972,276	-1972,276	-1684,436
Enthalpy Gcal/hr	-18,41226	-16,06195	-15,5583	-13,15231	-10,24313	-32,07382	-55,46926	-55,46926	-47,37392
Entropy cal/mol-K	2,016624	4,446706	2,855828	3,932146	-2,12549	-11,6133	-1,031281	-1,031281	5,425924
Entropy cal/gm-K	0,0723943	0,275936	0,1776853	0,2446521	-0,0482958	-0,6446361	-0,0546179	-0,0546179	0,2873639
Density mol/cc	2,27E-05	3,56E-05	4,46E-05	2,75E-04	4,81E-04	3,76E-04	3,17E-04	3,17E-04	1,64E-04
Density kg/cum	0,6319922	0,5731976	0,7165454	4,424679	21,17593	6,781927	5,987799	5,987799	3,099885

Substream: MIXED

Mole Flow kmol/hr	SYNG9	SYNG10	SYNG11	SYNG12	SYNG13	CO2+H2O	CO2-R2	SYNG14	SYNG15-1
H2O	647,6238	454,4497	454,4497	454,4497	0	454,4497	27,3566	0	0,1488937
O2	1,399823	1,10E-15	1,10E-15	1,10E-15	0	0	0	0	0
N2	23,9972	23,9972	23,9972	23,9972	23,9972	0	0	23,9972	512,9039
CH4	148,0428	25,16728	25,16728	25,16728	25,16728	0	0	25,16728	480,3509
CO	175,6693	355,0519	355,0519	355,0519	355,0519	0	0	355,0519	154,5317
CO2	151,9665	160,6073	160,6073	160,6073	32,12147	128,4859	128,2516	32,12147	44,16714
C2H2	0	0	0	0	0	0	0	0	0
C2H4	17,35142	0	0	0	0	0	0	0	0
C2H6	14,87728	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	302,9656	820,5348	820,5348	820,5348	820,5348	0	0	820,5348	1001,817
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	6,92E-07	6,92E-07	6,92E-07	6,92E-07	0	0	6,92E-07	2,08E-07
H2S	4,794624	5,485177	5,485177	5,485177	0	5,485177	5,367856	0	0,2018361
CH3OH	0	8,22E-05	8,22E-05	8,22E-05	8,22E-05	0	0	8,22E-05	8,554723
CL2	0	0	0	0	0	0	0	0	0
S	0	6,61E-09	6,61E-09	6,61E-09	6,61E-09	0	0	6,61E-09	1,98E-09
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	7,29E-16	7,29E-16	7,29E-16	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,799829	0,109276	0,109276	0,109276	0,109276	0	0	0,109276	0,0340734
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	11667,13	8187,039	8187,039	8187,039	0	8187,039	492,8367	0	2,682361
O2	44,79266	3,52E-14	3,52E-14	3,52E-14	0	0	0	0	0
N2	672,245	672,245	672,245	672,245	672,245	0	0	672,245	14368,22
CH4	2375,016	403,7526	403,7526	403,7526	403,7526	0	0	403,7526	7706,153
CO	4920,567	9945,145	9945,145	9945,145	9945,145	0	0	9945,145	4328,494
CO2	6688,015	7068,297	7068,297	7068,297	1413,659	5654,638	5644,328	1413,659	1943,787
C2H2	0	0	0	0	0	0	0	0	0
C2H4	486,7727	0	0	0	0	0	0	0	0
C2H6	447,3544	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	610,7423	1654,1	1654,1	1654,1	1654,1	0	0	1654,1	2019,543
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	4,44E-05	4,44E-05	4,44E-05	4,44E-05	0	0	4,44E-05	1,33E-05
H2S	163,4098	186,9451	186,9451	186,9451	0	186,9451	182,9466	0	6,878953
CH3OH	0	2,63E-03	2,63E-03	2,63E-03	2,63E-03	0	0	2,63E-03	274,1118
CL2	0	0	0	0	0	0	0	0	0
S	0	2,12E-07	2,12E-07	2,12E-07	2,12E-07	0	0	2,12E-07	6,36E-08
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	5,84E-14	5,84E-14	5,84E-14	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	48,05085	6,56491	6,56491	6,56491	6,56491	0	0	6,56491	2,047009
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	1489,488	1845,403	1845,403	1845,403	1256,982	588,4207	160,9761	1256,982	2202,71
Total Flow kg/hr	28124,09	28124,09	28124,09	28124,09	14095,47	14028,62	6320,111	14095,47	30651,92
Total Flow cum/hr	9072,623	11902,79	6603,546	3685,824	2804,554	351,5504	342,8811	373,0257	586,8253
Temperature C	860	887,3626	371,2634	127	127	127	127	101,5016	58,68603
Pressure bar	15,5	15	15	15	15	15	15	109,8386	106
Vapor Frac	1	1	1	0,901923	1	0,273573	1	1	1
Liquid Frac	0	0	0	0,098077	0	0,726427	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-31,80504	-20,69826	-25,08496	-27,95667	-9,498687	-71,44843	-84,03153	-9,682161	-7,629834
Enthalpy kcal/kg	-1684,436	-1358,146	-1645,986	-1834,417	-847,0579	-2996,855	-2140,321	-863,4194	-548,2956
Enthalpy Gcal/hr	-47,37392	-38,19718	-46,29252	-51,59205	-11,93985	-42,04234	-13,52726	-12,17048	-16,80655
Entropy cal/mol-K	5,425924	9,548194	4,574315	-1,498241	4,123906	-25,61278	-2,571236	-0,3807307	-8,876892
Entropy cal/gm-K	0,2873639	0,6265185	0,3001503	-0,0983092	0,3677548	-1,07431	-0,0654905	-0,0339521	-0,6379118
Density mol/cc	1,64E-04	1,55E-04	2,79E-04	5,01E-04	4,48E-04	1,67E-03	4,69E-04	3,37E-03	3,75E-03
Density kg/cum	3,099885	2,362815	4,258938	7,63034	5,025922	39,90501	18,43237	37,78685	52,23347

Substream: MIXED

Mole Flow kmol/hr	SYNG15-2	SYNG15-3	SYNG16	SYNG17	SYNG18	SYNG19	SYNG20	SYNG21	F-GASES5
H2O	0,1488937	0,1488937	0,1488937	0,3474186	0,3474186	27,75894	27,75894	27,75894	0,5026013
O2	0	0	0	0	0	0	0	0	0
N2	512,9039	512,9039	512,9039	1196,776	1196,776	1709,68	1709,68	1709,68	1707,03
CH4	480,3509	480,3509	480,3509	1120,819	1120,819	1601,17	1601,17	1601,17	1595,958
CO	154,5317	154,5317	154,5317	360,5739	360,5739	162,1735	162,1735	162,1735	162,0683
CO2	44,16714	44,16714	44,16714	103,0567	103,0567	119,9612	119,9612	119,9612	116,5528
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	1001,817	1001,817	1001,817	2337,573	2337,573	2551,412	2551,412	2551,412	2550,745
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	2,08E-07	2,08E-07	2,08E-07	4,85E-07	4,85E-07	0	0	0	0
H2S	0,2018361	0,2018361	0,2018361	0,4709509	0,4709509	0,7815782	0,7815782	0,7815782	0,6813157
CH3OH	8,554723	8,554723	8,554723	19,96102	19,96102	408,8192	408,8192	408,8192	28,87804
CL2	0	0	0	0	0	0	0	0	0
S	1,98E-09	1,98E-09	1,98E-09	4,63E-09	4,63E-09	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,0340734	0,0340734	0,0340734	0,0795046	0,0795046	4,79E-03	4,79E-03	4,79E-03	4,36E-03
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	2,682361	2,682361	2,682361	6,258843	6,258843	500,085	500,085	500,085	9,054504
O2	0	0	0	0	0	0	0	0	0
N2	14368,22	14368,22	14368,22	33525,85	33525,85	47894,07	47894,07	47894,07	47819,85
CH4	7706,153	7706,153	7706,153	17981,02	17981,02	25687,18	25687,18	25687,18	25603,57
CO	4328,494	4328,494	4328,494	10099,82	10099,82	4542,546	4542,546	4542,546	4539,598
CO2	1943,787	1943,787	1943,787	4535,503	4535,503	5279,468	5279,468	5279,468	5129,463
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	2019,543	2019,543	2019,543	4712,267	4712,267	5143,34	5143,34	5143,34	5141,996
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	1,33E-05	1,33E-05	1,33E-05	3,10E-05	3,10E-05	0	0	0	0
H2S	6,878953	6,878953	6,878953	16,05089	16,05089	26,63765	26,63765	26,63765	23,22052
CH3OH	274,1118	274,1118	274,1118	639,5942	639,5942	13099,45	13099,45	13099,45	925,3149
CL2	0	0	0	0	0	0	0	0	0
S	6,36E-08	6,36E-08	6,36E-08	1,48E-07	1,48E-07	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	2,047009	2,047009	2,047009	4,776354	4,776354	0,2876181	0,2876181	0,2876181	0,2616916
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	2202,71	2202,71	2202,71	5139,657	5139,657	6581,76	6581,76	6581,76	6162,42
Total Flow kg/hr	30651,92	30651,92	30651,92	71521,15	71521,15	1,02E+05	1,02E+05	1,02E+05	89192,32
Total Flow cum/hr	586,8253	937,0531	937,0531	1369,259	1330,528	3055,457	2710,718	1609,482	1804,835
Temperature C	58,68603	250	250	58,68603	50	260,0474	201,9565	30	29,10174
Pressure bar	106	106	106	106	106	98	98	98	86
Vapor Frac	1	1	1	1	1	1	1	0,9360961	1
Liquid Frac	0	0	0	0	0	0	0	0,0639039	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-7,629834	-6,07814	-6,07814	-7,629834	-7,699471	-8,046984	-8,566287	-10,56286	-7,369472
Enthalpy kcal/kg	-548,2956	-436,7877	-436,7877	-548,2956	-553,2999	-518,3687	-551,8211	-680,4359	-509,1669
Enthalpy Gcal/hr	-16,80655	-13,38857	-13,38857	-39,21529	-39,57321	-52,96408	-56,38206	-69,52322	-45,41443
Entropy cal/mol-K	-8,876892	-5,190125	-5,190125	-8,876892	-9,089542	-7,793827	-8,824808	-14,16064	-11,03746
Entropy cal/gm-K	-0,6379118	-0,3729731	-0,3729731	-0,6379118	-0,6531933	-0,5020609	-0,5684744	-0,9121965	-0,7625933
Density mol/cc	3,75E-03	2,35E-03	2,35E-03	3,75E-03	3,86E-03	2,15E-03	2,43E-03	4,09E-03	3,41E-03
Density kg/cum	52,23347	32,71097	32,71097	52,23347	53,75397	33,43953	37,69225	63,48197	49,41856

Substream: MIXED

Mole Flow kmol/hr	F-GASES6	F-GASES7	F-GASES8	SYNG-R1	SYNG-R2	MEOH+H2O	MEOH	VAPOUR
H2O	6,28E-03	6,28E-03	6,28E-03	0,4963123	0,4963123	27,25634	0	27,25634
O2	0	0	0	0	0	0	0	0
N2	21,3297	21,3297	21,3297	1685,682	1685,682	2,64969	0	2,64969
CH4	19,94184	19,94184	19,94184	1576,002	1576,002	5,211724	0	5,211724
CO	2,025078	2,025078	2,025078	160,0538	160,0538	0,1052378	0	0,1052378
CO2	1,456352	1,456352	1,456352	115,1023	115,1023	3,408436	0	3,408436
C2H2	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0
H2	31,87211	31,87211	31,87211	2518,856	2518,856	0,6671076	0	0,6671076
C	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
H2S	8,51E-03	8,51E-03	8,51E-03	0,6727869	0,6727869	0,1002625	0	0,1002625
CH3OH	0,3608374	0,3608374	0,3608374	28,51566	28,51566	379,9412	379,9412	0
CL2	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0
COS	5,44E-05	5,44E-05	5,44E-05	4,30E-03	4,30E-03	4,32E-04	0	4,32E-04
HCN	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0
Mass Flow kg/hr								
H2O	0,113138	0,113138	0,113138	8,941205	8,941205	491,0305	0	491,0305
O2	0	0	0	0	0	0	0	0
N2	597,5192	597,5192	597,5192	47221,83	47221,83	74,22703	0	74,22703
CH4	319,9221	319,9221	319,9221	25283,43	25283,43	83,61043	0	83,61043
CO	56,72325	56,72325	56,72325	4483,17	4483,17	2,947752	0	2,947752
CO2	64,09375	64,09375	64,09375	5065,631	5065,631	150,0046	0	150,0046
C2H2	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0
H2	64,25034	64,25034	64,25034	5077,711	5077,711	1,344809	0	1,344809
C	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
H2S	0,2901454	0,2901454	0,2901454	22,92984	22,92984	3,417133	0	3,417133
CH3OH	11,56201	11,56201	11,56201	913,7034	913,7034	12174,14	12174,14	0
CL2	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0
COS	3,27E-03	3,27E-03	3,27E-03	0,2584529	0,2584529	0,0259265	0	0,0259265
HCN	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0
Total Flow kmol/hr	77,00076	77,00076	77,00076	6085,385	6085,385	419,3404	379,9412	39,39923
Total Flow kg/hr	1114,477	1114,477	1114,477	88077,6	88077,6	12980,74	12174,14	806,6082
Total Flow cum/hr	22,5518	1859,294	1748,053	1782,278	1571,022	16,59534	15,44892	3,537866
Temperature C	29,10174	17,33759	0	29,10174	51,06692	29,10174	29,10174	29,10174
Pressure bar	86	1	1	86	106	86	86	86
Vapor Frac	1	1	1	1	1	0	0	0,3082322
Liquid Frac	0	0	0	0	0	1	1	0,6917678
Solid Frac	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-7,369472	-7,369472	-7,49673	-7,36982	-7,205911	-57,49135	-57,45734	-58,13242
Enthalpy kcal/kg	-509,1669	-509,1669	-517,9594	-509,1896	-497,8649	-1857,247	-1793,179	-2839,511
Enthalpy Gcal/hr	-0,5674631	-0,5674631	-0,5772622	-44,84884	-43,85137	-24,10879	-21,83073	-2,290405
Entropy cal/mol-K	-11,03746	-2,207534	-2,65922	-11,03749	-10,93345	-56,18868	-59,34457	-32,13722
Entropy cal/gm-K	-0,7625933	-0,1525216	-0,1837292	-0,7625933	-0,7554049	-1,815164	-1,852078	-1,56976
Density mol/cc	3,41E-03	4,14E-05	4,40E-05	3,41E-03	3,87E-03	0,0252685	0,0245933	0,0111364
Density kg/cum	49,41856	0,5994089	0,6375534	49,41856	56,06389	782,1921	788,0249	227,9929

Substream: MIXED

Mole Flow kmol/hr	WATER1-1	STEAM1-2	WATER2-1	WATER2-2	WATER2-3	WATER3-1	STEAM3-2	WATER4-1	WATER4-2
H2O	248,618	248,618	1443,219	1443,219	1461,247	333,0506	333,0506	1387,711	1387,711
O2	0	0	0	0	1,43E-05	0	0	0	0
N2	0	0	0	0	1,35E-05	0	0	0	0
CH4	0	0	0	0	6,49E-04	0	0	0	0
CO	0	0	0	0	7,89E-05	0	0	0	0
CO2	0	0	0	0	1,65E-03	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	1,20E-04	0	0	0	0
C2H6	0	0	0	0	7,36E-05	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	2,46E-04	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	4,82E-03	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	7,80E-05	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	4478,922	4478,922	26000	26000	26324,78	6000	6000	25000	25000
O2	0	0	0	0	4,56E-04	0	0	0	0
N2	0	0	0	0	3,79E-04	0	0	0	0
CH4	0	0	0	0	0,010405	0	0	0	0
CO	0	0	0	0	2,21E-03	0	0	0	0
CO2	0	0	0	0	0,0726413	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	3,37E-03	0	0	0	0
C2H6	0	0	0	0	2,21E-03	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	4,96E-04	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0,1642004	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	4,69E-03	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	248,618	248,618	1443,219	1443,219	1461,255	333,0506	333,0506	1387,711	1387,711
Total Flow kg/hr	4478,922	4478,922	26000	26000	26325,04	6000	6000	25000	25000
Total Flow cum/hr	5,082068	138,0757	26,15807	26,15807	27,177	6,807979	184,9531	24,91151	64,64837
Temperature C	134	520,0433	25	25	50,85773	134	520	15	200,0773
Pressure bar	108	108	1,2	1,2	1,2	108	108	15,5	15,5
Vapor Frac	0	1	0	0	0	0	1	0	0,0103544
Liquid Frac	1	0	1	1	1	1	0	1	0,9896455
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-66,54437	-54,08624	-68,72482	-68,72482	-68,22124	-66,54437	-54,0867	-68,91368	-65,09483
Enthalpy kcal/kg	-3693,774	-3002,242	-3814,807	-3814,807	-3786,837	-3693,774	-3002,268	-3825,29	-3613,312
Enthalpy Gcal/hr	-16,54436	-13,447	-99,18641	-99,18641	-99,69006	-22,16296	-18,01387	-95,63363	-90,33411
Entropy cal/mol-K	-34,03407	-12,12184	-40,1023	-40,1023	-38,483	-34,03407	-12,12243	-40,77119	-30,63229
Entropy cal/gm-K	-1,889178	-0,6728641	-2,226016	-2,226016	-2,136121	-1,889178	-0,6728969	-2,263145	-1,70035
Density mol/cc	0,0489206	1,80E-03	0,055173	0,055173	0,053768	0,0489206	1,80E-03	0,0557056	0,0214655
Density kg/cum	881,3189	32,43816	993,957	993,957	968,6514	881,3189	32,44066	1003,552	386,7073

Substream: MIXED

Mole Flow kmol/hr	WT-ATR1	WT-DRY1	ATR+DRY2	ATR+DRY3	ST-ATR4	ST-DRY4	ST-DRY5	ST-DRY6	ST-GASI1
H2O	572,0144	346,9277	918,9421	918,9421	572,0144	346,9277	346,9277	581,6686	11,54575
O2	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	10305	6250	16555	16555	10305	6250	6250	10478,92	208
O2	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	572,0144	346,9277	918,9421	918,9421	572,0144	346,9277	346,9277	581,6686	11,54575
Total Flow kg/hr	10305	6250	16555	16555	10305	6250	6250	10478,92	208
Total Flow cum/hr	26,64806	16,16209	2162,882	2441,046	1519,48	921,5669	1194,652	1896,123	0,5378745
Temperature C	200,0773	200,0773	200,0773	250	250	250	245,6562	221,3896	200,0773
Pressure bar	15,5	15,5	15,5	15,5	15,5	15,5	12	12	15,5
Vapor Frac	0,0103544	0,0103544	1	1	1	1	1	1	0,0103544
Liquid Frac	0,9896455	0,9896455	0	0	0	0	0	0	0,9896455
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-65,09483	-65,09483	-56,52339	-56,07231	-56,07231	-56,07231	-56,07231	-56,28764	-65,09483
Enthalpy kcal/kg	-3613,312	-3613,312	-3137,525	-3112,486	-3112,486	-3112,486	-3112,486	-3124,439	-3613,312
Enthalpy Gcal/hr	-37,23572	-22,58353	-51,94247	-51,52795	-32,07463	-19,45332	-19,45332	-32,74122	-0,7515798
Entropy cal/mol-K	-30,63229	-30,63229	-12,51954	-11,6133	-11,6133	-11,6133	-11,1292	-11,55427	-30,63229
Entropy cal/gm-K	-1,70035	-1,70035	-0,6949401	-0,6446361	-0,6446361	-0,6446361	-0,6177645	-0,6413596	-1,70035
Density mol/cc	0,0214655	0,0214655	4,25E-04	3,76E-04	3,76E-04	3,76E-04	2,90E-04	3,07E-04	0,0214655
Density kg/cum	386,7073	386,7073	7,654141	6,781927	6,781927	6,781927	5,231647	5,526499	386,7073

Substream: MIXED

Mole Flow kmol/hr	**ST-GASI2**	**ST-GASI3**	**WATER5-1**	**WATER6-1**
H2O	11,54575	11,54575	277,5422	4995,759
O2	0	0	0	0
N2	0	0	0	0
CH4	0	0	0	0
CO	0	0	0	0
CO2	0	0	0	0
C2H2	0	0	0	0
C2H4	0	0	0	0
C2H6	0	0	0	0
C3H8	0	0	0	0
H2	0	0	0	0
C	0	0	0	0
NH3	0	0	0	0
SO2	0	0	0	0
H2S	0	0	0	0
CH3OH	0	0	0	0
CL2	0	0	0	0
S	0	0	0	0
NO2	0	0	0	0
NO	0	0	0	0
SO3	0	0	0	0
HCL	0	0	0	0
COS	0	0	0	0
HCN	0	0	0	0
ARGON	0	0	0	0
Mass Flow kg/hr				
H2O	208	208	5000	90000
O2	0	0	0	0
N2	0	0	0	0
CH4	0	0	0	0
CO	0	0	0	0
CO2	0	0	0	0
C2H2	0	0	0	0
C2H4	0	0	0	0
C2H6	0	0	0	0
C3H8	0	0	0	0
H2	0	0	0	0
C	0	0	0	0
NH3	0	0	0	0
SO2	0	0	0	0
H2S	0	0	0	0
CH3OH	0	0	0	0
CL2	0	0	0	0
S	0	0	0	0
NO2	0	0	0	0
NO	0	0	0	0
SO3	0	0	0	0
HCL	0	0	0	0
COS	0	0	0	0
HCN	0	0	0	0
ARGON	0	0	0	0
Total Flow kmol/hr	11,54575	11,54575	277,5422	4995,759
Total Flow kg/hr	208	208	5000	90000
Total Flow cum/hr	61,24519	311,687	4,982303	89,68145
Temperature C	106,6173	120	15	15
Pressure bar	1,2	1,2	1	1
Vapor Frac	0,2029964	1	0	0
Liquid Frac	0,7970036	0	1	1
Solid Frac	0	0	0	0
Enthalpy kcal/mol	-65,09484	-57,00708	-68,91965	-68,91965
Enthalpy kcal/kg	-3613,312	-3164,374	-3825,622	-3825,622
Enthalpy Gcal/hr	-0,7515798	-0,6581992	-19,12839	-344,3109
Entropy cal/mol-K	-30,01615	-8,724509	-40,76664	-40,76664
Entropy cal/gm-K	-1,66615	-0,4842839	-2,262893	-2,262893
Density mol/cc	1,89E-04	3,70E-05	0,0557056	0,0557056
Density kg/cum	3,396185	0,6673362	1003,552	1003,552

Substream: MIXED

Mole Flow kmol/hr	DH-1	DH-2	DH-3	DH-4	DH-5	DH-6	DH-7
H2O	457,223	427,0931	4995,759	928,5963	7086,214	7086,214	277,5422
O2	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0
CO2	0	0,234265	0	0	0,234265	0,234265	0
C2H2	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
H2S	0	0,1173211	0	0	0,1173211	0,1173211	0
CH3OH	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0
Mass Flow kg/hr							
H2O	8237	7694,202	90000	16728,92	1,28E+05	1,28E+05	5000
O2	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0
CO2	0	10,30996	0	0	10,30996	10,30996	0
C2H2	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
H2S	0	3,998522	0	0	3,998522	3,998522	0
CH3OH	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0
Total Flow kmol/hr	457,223	427,4447	4995,759	928,5963	7086,565	7086,565	277,5422
Total Flow kg/hr	8237	7708,51	90000	16728,92	1,28E+05	1,28E+05	5000
Total Flow cum/hr	21,30035	8,669272	14464,84	20,4676	24136,88	24136,9	872,5129
Temperature C	200,0773	127	101,5389	188,3597	101,5247	101,5246	101,5389
Pressure bar	15,5	15	1	12	1	1	1
Vapor Frac	0,0103544	0	0,0931612	0	0,1097056	0,1097057	0,1012047
Liquid Frac	0,9896455	1	0,9068387	1	0,8902944	0,8902943	0,8987953
Solid Frac	0	0	0	0	0	0	0
Enthalpy kcal/mol	-65,09483	-66,70962	-66,28924	-65,44081	-66,12319	-66,12319	-66,20823
Enthalpy kcal/kg	-3613,312	-3699,116	-3679,612	-3632,517	-3670,165	-3670,165	-3675,115
Enthalpy Gcal/hr	-29,76328	-28,51508	-331,1698	-60,76897	-468,593	-468,593	-18,37584
Entropy cal/mol-K	-30,63229	-34,29025	-33,12926	-31,36569	-32,68358	-32,68358	-32,91304
Entropy cal/gm-K	-1,70035	-1,901429	-1,838953	-1,74106	-1,814101	-1,814101	-1,826952
Density mol/cc	0,0214655	0,0493057	3,45E-04	0,045369	2,94E-04	2,94E-04	3,18E-04
Density kg/cum	386,7073	889,1762	6,221983	817,3367	5,2896	5,289595	5,730574

	LUKAB-TB	COMP-GAS	MEOH-COM	R-COMP	TOTAL WORK
Power MW	-1,4741379	-	2,74878984	1,28894201	2,56359395

D.3 Stream result for the ATR methanol system

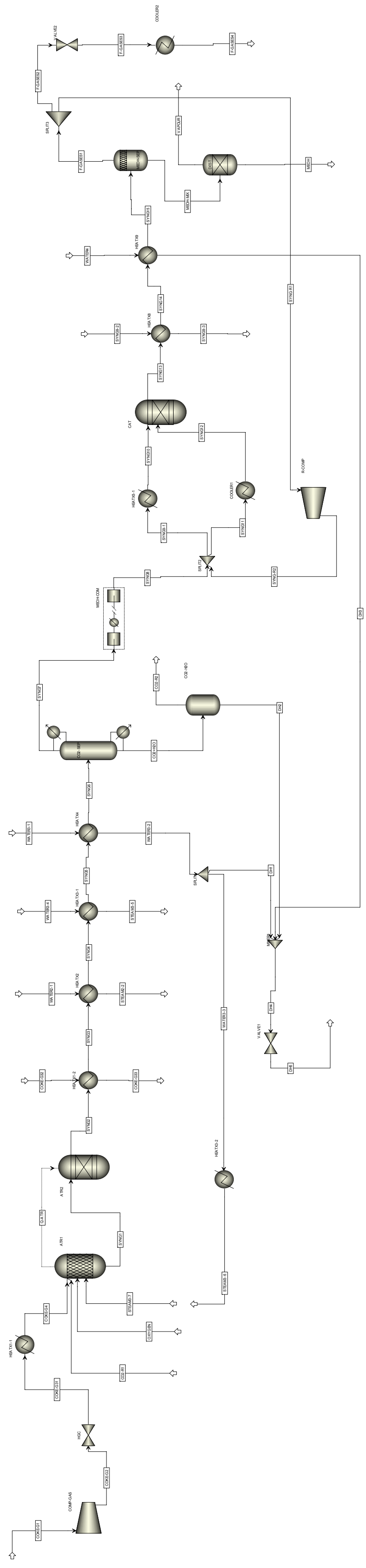

Figure 36 Aspen plus flow sheet for the ATR system

Table 43 Aspen plus stream results for the ATR system

Substream: MIXED									
Mole Flow kmol/hr	**COKE-G1**	**COKE-G2**	**COKE-G31**	**COKE-G32**	**COKE-G33**	**COKE-G4**	**CO2-R1**	**OXYGEN**	**STEAM3-7**
H2O	11,53116	11,53116	11,53116	11,53116	11,53116	11,53116	0	0	253
O2	1,411156	1,411156	1,411156	1,411156	1,411156	1,411156	0	89	0
N2	24,19124	24,19124	24,19124	24,19124	24,19124	24,19124	0	0	0
CH4	96,76495	96,76495	96,76495	96,76495	96,76495	96,76495	0	0	0
CO	22,57849	22,57849	22,57849	22,57849	22,57849	22,57849	0	0	0
CO2	5,241435	5,241435	5,241435	5,241435	5,241435	5,241435	100	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	12,49881	12,49881	12,49881	12,49881	12,49881	12,49881	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	235,8646	235,8646	235,8646	235,8646	235,8646	235,8646	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	4,838248	4,838248	4,838248	4,838248	4,838248	4,838248	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,8063746	0,8063746	0,8063746	0,8063746	0,8063746	0,8063746	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	207,737	207,737	207,737	207,737	207,737	207,737	0	0	4557,866
O2	45,15528	45,15528	45,15528	45,15528	45,15528	45,15528	0	2847,893	0
N2	677,6808	677,6808	677,6808	677,6808	677,6808	677,6808	0	0	0
CH4	1552,377	1552,377	1552,377	1552,377	1552,377	1552,377	0	0	0
CO	632,4325	632,4325	632,4325	632,4325	632,4325	632,4325	0	0	0
CO2	230,6745	230,6745	230,6745	230,6745	230,6745	230,6745	4400,98	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	375,8346	375,8346	375,8346	375,8346	375,8346	375,8346	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	475,4747	475,4747	475,4747	475,4747	475,4747	475,4747	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	164,8966	164,8966	164,8966	164,8966	164,8966	164,8966	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	48,44408	48,44408	48,44408	48,44408	48,44408	48,44408	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	415,7264	415,7264	415,7264	415,7264	415,7264	415,7264	100	89	253
Total Flow kg/hr	4410,707	4410,707	4410,707	4410,707	4410,707	4410,707	4400,98	2847,893	4557,866
Total Flow cum/hr	10286	1076,111	1350,393	1350,393	1732,749	1732,749	163,816	198,9797	526,0134
Temperature C	25	484,5285	484,6406	484,6406	700	700	127	250	250
Pressure bar	1	24,5	19,5	19,5	19,5	19,5	19,5	19,5	19,5
Vapor Frac	0,9982924	1	1	1	1	1	1	1	1
Liquid Frac	1,71E-03	0	0	0	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-9,108616	-5,153576	-5,153576	-5,153576	-2,995969	-2,995969	-93,14277	1,612143	-56,11748
Enthalpy kcal/kg	-858,5227	-485,7448	-485,7448	-485,7448	-282,3818	-282,3818	-2116,41	50,38135	-3114,993
Enthalpy Gcal/hr	-3,786747	-2,142508	-2,142508	-2,142508	-1,245521	-1,245521	-9,314411	0,1434828	-14,19793
Entropy cal/mol-K	-2,099471	-0,5637858	-0,1077826	-0,1077825	2,393742	2,393742	-2,626877	-1,853673	-12,12802
Entropy cal/gm-K	-0,1978834	-0,053139	-0,0101589	-0,0101589	0,2256196	0,2256196	-0,0596884	-0,0579294	-0,6732075
Density mol/cc	4,04E-05	3,86E-04	3,08E-04	3,08E-04	2,40E-04	2,40E-04	6,10E-04	4,47E-04	4,81E-04
Density kg/cum	0,4288068	4,098748	3,266241	3,266241	2,545496	2,545496	26,86539	14,31248	8,664923

Substream: MIXED

	SYNG1	SYNG2	SYNG3	SYNG4	SYNG5	SYNG6	SYNG7	CO2+H2O	CO2-R2
Mole Flow kmol/hr									
H2O	405,3848	350,8804	350,8804	350,8804	350,8804	350,8804	0	350,8804	12,34314
O2	0	1,51E-13	1,51E-13	1,51E-13	1,51E-13	1,51E-13	0	0	0
N2	24,19124	24,19124	24,19124	24,19124	24,19124	24,19124	24,19124	0	0
CH4	67,87191	10,8595	10,8595	10,8595	10,8595	10,8595	10,8595	0	0
CO	58,93557	150,6129	150,6129	150,6129	150,6129	150,6129	150,6129	0	0
CO2	105,4822	88,81138	88,81138	88,81138	88,81138	88,81138	17,76228	71,0491	70,82109
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	8,766788	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	165,4378	358,1213	358,1213	358,1213	358,1213	358,1213	358,1213	0	0
C	0	1,22E-21	1,22E-21	1,22E-21	1,22E-21	1,22E-21	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	1,685428	1,75E-05	1,75E-05	1,75E-05	1,75E-05	1,75E-05	1,75E-05	0	0
H2S	3,393595	5,539538	5,539538	5,539538	5,539538	5,539538	0	5,539538	5,334404
CH3OH	0	1,39E-05	1,39E-05	1,39E-05	1,39E-05	1,39E-05	1,39E-05	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	1,13E-07	1,13E-07	1,13E-07	1,13E-07	1,13E-07	1,13E-07	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	1,35E-13	1,35E-13	1,35E-13	1,35E-13	1,35E-13	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,5655992	0,1050663	0,1050663	0,1050663	0,1050663	0,1050663	0,1050663	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	7303,12	6321,209	6321,209	6321,209	6321,209	6321,209	0	6321,209	222,3651
O2	0	4,82E-12	4,82E-12	4,82E-12	4,82E-12	4,82E-12	0	0	0
N2	677,6808	677,6808	677,6808	677,6808	677,6808	677,6808	677,6808	0	0
CH4	1088,853	174,2164	174,2164	174,2164	174,2164	174,2164	174,2164	0	0
CO	1650,809	4218,727	4218,727	4218,727	4218,727	4218,727	4218,727	0	0
CO2	4642,251	3908,571	3908,571	3908,571	3908,571	3908,571	781,7142	3126,857	3116,822
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	263,6142	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	333,5027	721,9295	721,9295	721,9295	721,9295	721,9295	721,9295	0	0
C	0	1,46E-20	1,46E-20	1,46E-20	1,46E-20	1,46E-20	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	107,9766	1,12E-03	1,12E-03	1,12E-03	1,12E-03	1,12E-03	1,12E-03	0	0
H2S	115,6601	188,7979	188,7979	188,7979	188,7979	188,7979	0	188,7979	181,8065
CH3OH	0	4,47E-04	4,47E-04	4,47E-04	4,47E-04	4,47E-04	4,47E-04	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	3,63E-06	3,63E-06	3,63E-06	3,63E-06	3,63E-06	3,63E-06	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	1,08E-11	1,08E-11	1,08E-11	1,08E-11	1,08E-11	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	33,97916	6,312007	6,312007	6,312007	6,312007	6,312007	6,312007	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	841,7149	989,1214	989,1214	989,1214	989,1214	989,1214	561,6523	427,4691	88,49863
Total Flow kg/hr	16217,45	16217,45	16217,45	16217,45	16217,45	16217,45	6580,582	9636,864	3520,994
Total Flow cum/hr	5440,507	5535,81	5116,646	3483,207	2241,294	1289,276	990,7897	153,9562	151,4861
Temperature C	1200	1001,963	905,3488	530	248,1018	127	127	127	126,7072
Pressure bar	19	19	19	19	19	19	19	19	18,5
Vapor Frac	1	1	1	1	1	0,7428192	1	0,2064999	1
Liquid Frac	0	0	0	0	0	0,2571808	0	0,7935001	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-31,63488	-24,88272	-25,78956	-29,1431	-31,48889	-34,9304	-9,681824	-70,02212	-82,76125
Enthalpy kcal/kg	-1641,908	-1517,627	-1572,936	-1777,473	-1920,545	-2130,447	-826,3431	-3106,02	-2080,168
Enthalpy Gcal/hr	-26,62793	-24,61238	-25,50937	-28,82648	-31,14679	-34,5509	-5,437897	-29,93272	-7,324363
Entropy cal/mol-K	6,753966	8,367066	7,627587	4,210621	0,6203258	-7,362748	3,525145	-27,69047	-2,378026
Entropy cal/gm-K	0,3505431	0,5103174	0,4652156	0,2568108	0,0378344	-0,4490628	0,300871	-1,228285	-0,0597706
Density mol/cc	1,55E-04	1,79E-04	1,93E-04	2,84E-04	4,41E-04	7,67E-04	5,67E-04	2,78E-03	5,84E-04
Density kg/cum	2,98087	2,929553	3,169546	4,655894	7,23575	12,57872	6,641755	62,59484	23,24302

Substream: MIXED									
Mole Flow kmol/hr	**SYNG8**	**SYNG9-1**	**SYNG9-2**	**SYNG9-3**	**SYNG10**	**SYNG11**	**SYNG12**	**SYNG13**	**SYNG14**
H2O	0	0,0731848	0,0731848	0,0731848	0,0731848	0,1707647	0,1707647	13,63489	13,63489
O2	0	0	0	0	0	0	0	0	0
N2	24,19124	320,5086	320,5086	320,5086	320,5086	747,8534	747,8534	1068,362	1068,362
CH4	10,8595	134,1445	134,1445	134,1445	134,1445	313,0037	313,0037	447,1482	447,1482
CO	150,6129	73,03599	73,03599	73,03599	73,03599	170,4173	170,4173	94,9171	94,9171
CO2	17,76228	31,24625	31,24625	31,24625	31,24625	72,90791	72,90791	90,76326	90,76326
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	358,1213	388,8508	388,8508	388,8508	388,8508	907,3185	907,3185	958,6111	958,6111
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	1,75E-05	5,26E-06	5,26E-06	5,26E-06	5,26E-06	1,23E-05	1,23E-05	0	0
H2S	0	0,1837143	0,1837143	0,1837143	0,1837143	0,4286668	0,4286668	0,7167324	0,7167324
CH3OH	1,39E-05	3,721841	3,721841	3,721841	3,721841	8,684295	8,684295	174,4376	174,4376
CL2	0	0	0	0	0	0	0	0	0
S	1,13E-07	3,39E-08	3,39E-08	3,39E-08	3,39E-08	7,92E-08	7,92E-08	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,1050663	0,0333526	0,0333526	0,0333526	0,0333526	0,0778229	0,0778229	6,84E-03	6,84E-03
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	0	1,318446	1,318446	1,318446	1,318446	3,076373	3,076373	245,6364	245,6364
O2	0	0	0	0	0	0	0	0	0
N2	677,6808	8978,561	8978,561	8978,561	8978,561	20949,98	20949,98	29928,54	29928,54
CH4	174,2164	2152,047	2152,047	2152,047	2152,047	5021,444	5021,444	7173,491	7173,491
CO	4218,727	2045,767	2045,767	2045,767	2045,767	4773,457	4773,457	2658,666	2658,666
CO2	781,7142	1375,141	1375,141	1375,141	1375,141	3208,663	3208,663	3994,473	3994,473
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	721,9295	783,8765	783,8765	783,8765	783,8765	1829,045	1829,045	1932,445	1932,445
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	1,12E-03	3,37E-04	3,37E-04	3,37E-04	3,37E-04	7,86E-04	7,86E-04	0	0
H2S	0	6,26133	6,26133	6,26133	6,26133	14,60977	14,60977	24,42759	24,42759
CH3OH	4,47E-04	119,2558	119,2558	119,2558	119,2558	278,2636	278,2636	5589,356	5589,356
CL2	0	0	0	0	0	0	0	0	0
S	3,63E-06	1,09E-06	1,09E-06	1,09E-06	1,09E-06	2,54E-06	2,54E-06	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	6,312007	2,00371	2,00371	2,00371	2,00371	4,675323	4,675323	0,4110428	0,4110428
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	561,6523	951,7981	951,7981	951,7981	951,7981	2220,862	2220,862	2848,598	2848,598
Total Flow kg/hr	6580,582	15464,23	15464,23	15464,23	15464,23	36083,21	36083,21	51547,44	51547,44
Total Flow cum/hr	172,281	254,5243	254,5243	405,1964	405,1964	593,8901	574,9407	1325,447	1175,624
Temperature C	104,3415	59,80594	59,80594	250	250	59,80594	50	260,9077	202,6503
Pressure bar	106,8652	106	106	106	106	106	106	98	98
Vapor Frac	1	1	1	1	1	1	1	1	1
Liquid Frac	0	0	0	0	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-9,847077	-7,632051	-7,632051	-6,124147	-6,124147	-7,632051	-7,709825	-8,057343	-8,561177
Enthalpy kcal/kg	-840,4475	-469,7402	-469,7402	-376,9312	-376,9312	-469,7402	-474,5271	-445,2622	-473,1049
Enthalpy Gcal/hr	-5,530713	-7,264276	-7,264276	-5,829035	-5,829035	-16,94998	-17,12271	-22,95246	-24,3877
Entropy cal/mol-K	-0,400175	-7,187297	-7,187297	-3,60706	-3,60706	-7,187297	-7,424394	-5,974745	-6,973541
Entropy cal/gm-K	-0,0341549	-0,4423663	-0,4423663	-0,2220086	-0,2220086	-0,4423663	-0,4569593	-0,3301744	-0,3853695
Density mol/cc	3,26E-03	3,74E-03	3,74E-03	2,35E-03	2,35E-03	3,74E-03	3,86E-03	2,15E-03	2,42E-03
Density kg/cum	38,1968	60,75738	60,75738	38,16479	38,16479	60,75738	62,75988	38,89061	43,84689

Substream: MIXED

	SYNG15	F-GASES1	F-GASES2	F-GASES3	F-GASES4	SYNG-R1	SYNG-R2	MEOH-MIX	MEOH	VAPOUR
Mole Flow kmol/hr										
H2O	13,63489	0,2492072	5,28E-03	5,28E-03	5,28E-03	0,2439495	0,2439495	13,38569	0	13,38569
O2	0	0	0	0	0	0	0	0	0	0
N2	1068,362	1066,762	22,60612	22,60612	22,60612	1044,171	1044,171	1,600374	0	1,600374
CH4	447,1482	445,7289	9,445596	9,445596	9,445596	436,2887	436,2887	1,419264	0	1,419264
CO	94,9171	94,8572	2,010152	2,010152	2,010152	92,84039	92,84039	0,0599019	0	0,0599019
CO2	90,76326	88,26189	1,870389	1,870389	1,870389	86,39189	86,39189	2,501361	0	2,501361
C2H2	0	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0	0
H2	958,6111	958,3656	20,30906	20,30906	20,30906	938,048	938,048	0,2455128	0	0,2455128
C	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
H2S	0,7167324	0,6256284	0,0132579	0,0132579	0,0132579	0,6123811	0,6123811	0,091104	0	0,091104
CH3OH	174,4376	12,67407	0,2685806	0,2685806	0,2685806	12,40612	12,40612	161,7635	161,7635	0
CL2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0
COS	6,84E-03	6,24E-03	1,32E-04	1,32E-04	1,32E-04	6,11E-03	6,11E-03	6,00E-04	0	6,00E-04
HCN	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr										
H2O	245,6364	4,489538	0,0951393	0,0951393	0,0951393	4,394819	4,394819	241,1469	0	241,1469
O2	0	0	0	0	0	0	0	0	0	0
N2	29928,54	29883,7	633,276	633,276	633,276	29250,86	29250,86	44,83205	0	44,83205
CH4	7173,491	7150,722	151,5334	151,5334	151,5334	6999,274	6999,274	22,76892	0	22,76892
CO	2658,666	2656,988	56,30516	56,30516	56,30516	2600,496	2600,496	1,677877	0	1,677877
CO2	3994,473	3884,388	82,31542	82,31542	82,31542	3802,09	3802,09	110,0844	0	110,0844
C2H2	0	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0	0
H2	1932,445	1931,95	40,94063	40,94063	40,94063	1890,992	1890,992	0,4949243	0	0,4949243
C	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
H2S	24,42759	21,32259	0,4518545	0,4518545	0,4518545	20,8711	20,8711	3,104996	0	3,104996
CH3OH	5589,356	406,1046	8,605903	8,605903	8,605903	397,5189	397,5189	5183,252	5183,252	0
CL2	0	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0	0
COS	0,4110428	0,3749909	7,95E-03	7,95E-03	7,95E-03	0,3670252	0,3670252	0,0360519	0	0,0360519
HCN	0	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	2848,598	2667,53	56,52856	56,52856	56,52856	2611,008	2611,008	181,0673	161,7635	19,3038
Total Flow kg/hr	51547,44	45940,04	973,5315	973,5315	973,5315	44966,86	44966,86	5607,398	5183,252	424,1461
Total Flow cum/hr	696,5119	780,9579	16,54955	1361,537	1283,272	764,4142	674,9278	7,127498	6,577142	1,662245
Temperature C	30	29,0637	29,0637	16,61397	0	29,0637	51,38747	29,0637	29,0637	29,0637
Pressure bar	98	86	86	1	1	86	106	86	86	86
Vapor Frac	0,9362501	1	1	1	1	1	1	0	0	0,3064642
Liquid Frac	0,0637499	0	0	0	0	0	0	1	1	0,6935358
Solid Frac	0	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-10,53123	-7,319591	-7,319591	-7,319591	-7,43986	-7,319572	-7,155578	-57,84591	-57,45835	-61,29173
Enthalpy kcal/kg	-581,9735	-425,0155	-425,0155	-425,0155	-431,999	-425,0121	-415,4898	-1867,89	-1793,211	-2789,519
Enthalpy Gcal/hr	-29,99968	-19,52551	-0,4137719	-0,4137719	-0,4205706	-19,11174	-18,68354	-10,47415	-9,294797	-1,18318
Entropy cal/mol-K	-12,23998	-9,026828	-9,026828	-0,1985856	-0,6260064	-9,026876	-8,922258	-55,69967	-59,34791	-31,04715
Entropy cal/gm-K	-0,6764017	-0,524147	-0,524147	-0,0115309	-0,0363493	-0,524147	-0,5180724	-1,798586	-1,852182	-1,413023
Density mol/cc	4,09E-03	3,42E-03	3,42E-03	4,15E-05	4,41E-05	3,42E-03	3,87E-03	0,025404	0,0245948	0,011613
Density kg/cum	74,00798	58,82525	58,82525	0,715024	0,7586319	58,82525	66,6247	786,7274	788,0705	255,1646

Substream: MIXED

	WATER2-1	STEAM2-2	WATER3-1	WATER3-2	WATER3-3	WATER3-4	STEAM3-5	STEAM3-6	WATER4
Mole Flow kmol/hr									
H2O	266,419	266,419	943,6434	943,6434	253	253	253	253	2220,337
O2	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	4799,614	4799,614	17000	17000	4557,866	4557,866	4557,866	4557,866	40000
O2	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	0
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	266,419	266,419	943,6434	943,6434	253	253	253	253	2220,337
Total Flow kg/hr	4799,614	4799,614	17000	17000	4557,866	4557,866	4557,866	4557,866	40000
Total Flow cum/hr	5,445945	147,7769	16,93983	20,99945	5,630158	5,636669	526,0134	526,0134	39,85842
Temperature C	134	519,3415	15	194,5713	194,5713	195,3106	250	250	15
Pressure bar	108	108	19,5	19,5	19,5	19,5	19,5	19,5	1
Vapor Frac	0	1	0	0	0	0	1	1	0
Liquid Frac	1	0	1	1	1	1	0	0	1
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-66,54437	-54,09383	-68,91203	-65,30466	-65,30466	-65,28852	-56,11748	-56,11748	-68,91965
Enthalpy kcal/kg	-3693,774	-3002,664	-3825,199	-3624,96	-3624,96	-3624,064	-3114,993	-3114,993	-3825,622
Enthalpy Gcal/hr	-17,72894	-14,41183	-65,02932	-61,6252	-16,52232	-16,51823	-14,19793	-14,19793	-153,0271
Entropy cal/mol-K	-34,03407	-12,13141	-40,77245	-31,08229	-31,08229	-31,04781	-12,12802	-12,12802	-40,76664
Entropy cal/gm-K	-1,889178	-0,6733955	-2,263215	-1,725329	-1,725329	-1,723416	-0,6732075	-0,6732075	-2,262893
Density mol/cc	0,0489206	1,80E-03	0,0557056	0,0449365	0,0449365	0,0448846	4,81E-04	4,81E-04	0,0557056
Density kg/cum	881,3189	32,47879	1003,552	809,5449	809,5449	808,6098	8,664923	8,664923	1003,552

Substream: MIXED

Mole Flow kmol/hr	**DH1**	**DH2**	**DH3**	**DH4**	**DH5**
H2O	690,6434	338,5373	2220,337	3249,518	3249,518
O2	0	0	0	0	0
N2	0	0	0	0	0
CH4	0	0	0	0	0
CO	0	0	0	0	0
CO2	0	0,2280125	0	0,2280125	0,2280125
C2H2	0	0	0	0	0
C2H4	0	0	0	0	0
C2H6	0	0	0	0	0
C3H8	0	0	0	0	0
H2	0	0	0	0	0
C	0	0	0	0	0
NH3	0	0	0	0	0
SO2	0	0	0	0	0
H2S	0	0,2051342	0	0,2051342	0,2051342
CH3OH	0	0	0	0	0
CL2	0	0	0	0	0
S	0	0	0	0	0
NO2	0	0	0	0	0
NO	0	0	0	0	0
SO3	0	0	0	0	0
HCL	0	0	0	0	0
COS	0	0	0	0	0
HCN	0	0	0	0	0
ARGON	0	0	0	0	0
Mass Flow kg/hr					
H2O	12442,13	6098,844	40000	58540,98	58540,98
O2	0	0	0	0	0
N2	0	0	0	0	0
CH4	0	0	0	0	0
CO	0	0	0	0	0
CO2	0	10,03478	0	10,03478	10,03478
C2H2	0	0	0	0	0
C2H4	0	0	0	0	0
C2H6	0	0	0	0	0
C3H8	0	0	0	0	0
H2	0	0	0	0	0
C	0	0	0	0	0
NH3	0	0	0	0	0
SO2	0	0	0	0	0
H2S	0	6,99136	0	6,99136	6,99136
CH3OH	0	0	0	0	0
CL2	0	0	0	0	0
S	0	0	0	0	0
NO2	0	0	0	0	0
NO	0	0	0	0	0
SO3	0	0	0	0	0
HCL	0	0	0	0	0
COS	0	0	0	0	0
HCN	0	0	0	0	0
ARGON	0	0	0	0	0
Total Flow kmol/hr	690,6434	338,9704	2220,337	3249,951	3249,951
Total Flow kg/hr	12442,13	6115,87	40000	58558	58558
Total Flow cum/hr	15,36929	6,875936	5728,414	10368,82	10368,87
Temperature C	194,5713	126,7072	101,5389	101,5013	101,501
Pressure bar	19,5	18,5	1	1	1
Vapor Frac	0	0	0,0829422	0,1027284	0,1027291
Liquid Frac	1	1	0,9170578	0,8972716	0,8972709
Solid Frac	0	0	0	0	0
Enthalpy kcal/mol	-65,30466	-66,69618	-66,39217	-66,19277	-66,19277
Enthalpy kcal/kg	-3624,96	-3696,617	-3685,325	-3673,679	-3673,679
Enthalpy Gcal/hr	-45,10288	-22,60836	-147,4151	-215,1264	-215,1264
Entropy cal/mol-K	-31,08229	-34,28522	-33,40396	-32,8696	-32,8696
Entropy cal/gm-K	-1,725329	-1,900249	-1,854202	-1,824253	-1,824253
Density mol/cc	0,0449365	0,049298	3,88E-04	3,13E-04	3,13E-04
Density kg/cum	809,5449	889,46	6,982736	5,647509	5,647481

	LUKAB-TB	**COMP-GAS**	**MEOH-COM**	**R-COMP**	**TOTAL WORK**
Power MW	-	2,12469058	1,07999584	0,55331279	3,75799921

D.4 Stream result for the SMR methanol system

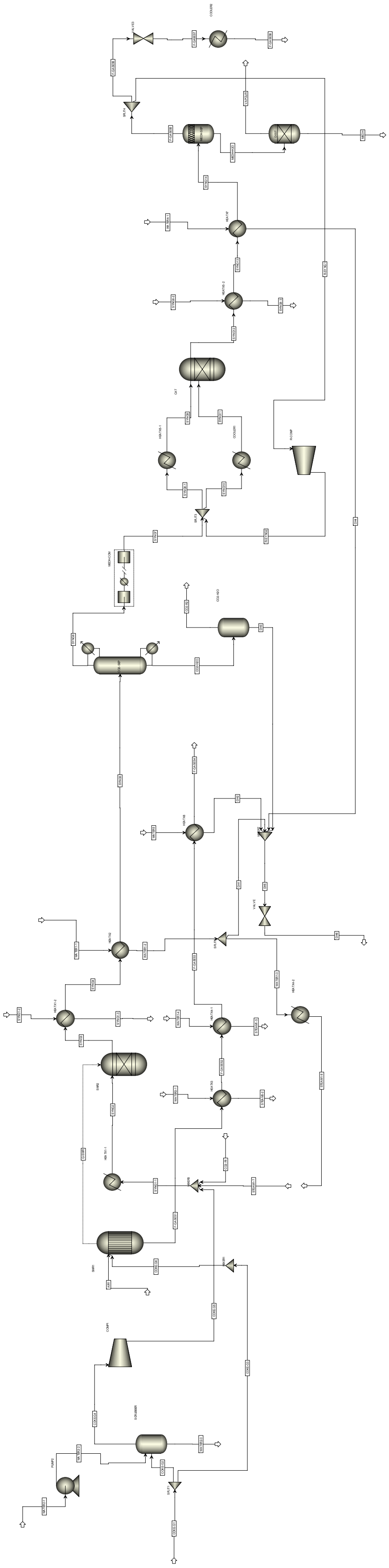

Figure 37 Aspen plus flow sheet for the SMR system

Table 44 Aspen plus stream results for the SMR system

Substream: MIXED									
Mole Flow kmol/hr	**COKE-G1**	**COKE-G3**	**COKE-G6**	**AIR1**	**F-GASES1**	**F-GASES2**	**F-GASES3**	**F-GASES4**	**COKE-G2**
H_2O	13,89875	2,849244	2,849244	4,545	123,9542	123,9542	123,9542	123,9542	11,04951
O_2	1,700896	0,3486837	0,3486837	93,375	1,073455	1,073455	1,073455	1,073455	1,352212
N_2	29,15821	5,977434	5,977434	347,805	353,7824	353,7824	353,7824	353,7824	23,18078
CH_4	116,6329	23,90974	23,90974	0	0	0	0	0	92,72312
CO	27,21433	5,578938	5,578938	0	0	0	0	0	21,6354
CO_2	6,317613	1,295111	1,295111	0,135	37,29472	37,29472	37,29472	37,29472	5,022503
C_2H_2	0	0	0	0	0	0	0	0	0
C_2H_4	0	0	0	0	0	0	0	0	0
C_2H_6	15,06508	3,088341	3,088341	0	0	0	0	0	11,97674
C_3H_8	0	0	0	0	0	0	0	0	0
H_2	284,2926	58,27998	58,27998	0	0	0	0	0	226,0126
C	0	0	0	0	0	0	0	0	0
NH_3	0	0	0	0	0	0	0	0	0
SO_2	0	0	0	0	1,394735	1,394735	1,394735	1,394735	0
H_2S	5,831643	1,195487	1,195487	0	0	0	0	0	4,636156
CH_3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO_2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO_3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,9719405	0,1992478	0,1992478	0	0	0	0	0	0,7726927
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	4,14	4,14	4,14	4,14	4,14	0
Mass Flow kg/hr									
H_2O	250,3899	51,32992	51,32992	81,87945	2233,07	2233,07	2233,07	2233,07	199,0599
O_2	54,42663	11,15746	11,15746	2987,888	34,34928	34,34928	34,34928	34,34928	43,26917
N_2	816,8231	167,4487	167,4487	9743,228	9910,677	9910,677	9910,677	9910,677	649,3743
CH_4	1871,113	383,5782	383,5782	0	0	0	0	0	1487,535
CO	762,2844	156,2683	156,2683	0	0	0	0	0	606,0161
CO_2	278,0369	56,99756	56,99756	5,941323	1641,333	1641,333	1641,333	1641,333	221,0393
C_2H_2	0	0	0	0	0	0	0	0	0
C_2H_4	0	0	0	0	0	0	0	0	0
C_2H_6	453,0015	92,8653	92,8653	0	0	0	0	0	360,1362
C_3H_8	0	0	0	0	0	0	0	0	0
H_2	573,0998	117,4855	117,4855	0	0	0	0	0	455,6143
C	0	0	0	0	0	0	0	0	0
NH_3	0	0	0	0	0	0	0	0	0
SO_2	0	0	0	0	89,35339	89,35339	89,35339	89,35339	0
H_2S	198,7534	40,74444	40,74444	0	0	0	0	0	158,0089
CH_3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO_2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO_3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	58,39069	11,97009	11,97009	0	0	0	0	0	46,4206
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	165,3847	165,3847	165,3847	165,3847	165,3847	0
Total Flow kmol/hr	501,0839	102,7222	102,7222	450	521,6395	521,6395	521,6395	521,6395	398,3617
Total Flow kg/hr	5316,319	1089,845	1089,845	12984,32	14074,17	14074,17	14074,17	14074,17	4226,474
Total Flow cum/hr	10286	2108,63	2108,631	10632,1	71744,24	32502,76	30712,92	22991,6	8177,37
Temperature C	25	25	25,00015	15	890	253,9562	224,9689	100	25
Pressure bar	1,2	1,2	1,2	1,01325	0,70325	0,70325	0,70325	0,70325	1,2
Vapor Frac	0,9938864	0,9938864	0,9938864	1	1	1	1	1	0,9938864
Liquid Frac	6,11E-03	6,11E-03	6,11E-03	0	0	0	0	0	6,11E-03
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-9,157125	-9,157125	-9,157125	-0,6833247	-13,61491	-18,91746	-19,13974	-20,08217	-9,157125
Enthalpy kcal/kg	-863,0949	-863,0949	-863,0949	-23,68211	-504,6178	-701,1497	-709,3881	-744,3178	-863,0949
Enthalpy Gcal/hr	-4,588554	-0,9406535	-0,9406535	-0,3075005	-7,102178	-9,868239	-9,98419	-10,4758	-3,6479
Entropy cal/mol-K	-2,622728	-2,622728	-2,622723	0,8757114	10,76763	4,235484	3,801764	1,624737	-2,622728
Entropy cal/gm-K	-0,2472024	-0,2472024	-0,2472019	0,0303496	0,3990875	0,1569823	0,1409071	0,0602186	-0,2472024
Density mol/cc	4,87E-05	4,87E-05	4,87E-05	4,23E-05	7,27E-06	1,60E-05	1,70E-05	2,27E-05	4,87E-05
Density kg/cum	0,51685	0,51685	0,5168497	1,221238	0,1961714	0,4330145	0,4582491	0,6121439	0,51685

Substream: MIXED

Mole Flow kmol/hr	COKE-G4	COKE-G5	STEAM1-7	CO2-R1	SYNG1-1	SYNG1-2	SYNG1-3	SYNG2	SYNG3
H2O	8,613909	8,613909	250	0	258,6139	258,6139	258,6139	258,6139	203,3433
O2	1,352206	1,352206	0	0	1,352206	1,352206	1,352206	1,352206	3,97E-16
N2	23,18078	23,18078	0	0	23,18078	23,18078	23,18078	23,18078	23,18078
CH4	92,72298	92,72298	0	0	92,72298	92,72298	92,72298	92,72298	12,98122
CO	21,63539	21,63539	0	0	21,63539	21,63539	21,63539	21,63539	171,7351
CO2	5,022398	5,022398	0	100	105,0224	105,0224	105,0224	105,0224	59,30225
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	11,97672	11,97672	0	0	11,97672	11,97672	11,97672	11,97672	0
C3H8	0	0	0	0	0	0	0	0	0
H2	226,0126	226,0126	0	0	226,0126	226,0126	226,0126	226,0126	476,0124
C	0	0	0	0	0	0	0	0	1,66E-23
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	3,82E-07
H2S	4,632774	4,632774	0	0	4,632774	4,632774	4,632774	4,632774	5,317192
CH3OH	0	0	0	0	0	0	0	0	4,86E-05
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	6,28E-09
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	3,26E-16
HCL	0	0	0	0	0	0	0	0	0
COS	0,7726597	0,7726597	0	0	0,7726597	0,7726597	0,7726597	0,7726597	0,0882412
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	155,182	155,182	4503,82	0	4659,002	4659,002	4659,002	4659,002	3663,286
O2	43,26898	43,26898	0	0	43,26898	43,26898	43,26898	43,26898	1,27E-14
N2	649,3742	649,3742	0	0	649,3742	649,3742	649,3742	649,3742	649,3742
CH4	1487,533	1487,533	0	0	1487,533	1487,533	1487,533	1487,533	208,2546
CO	606,016	606,016	0	0	606,016	606,016	606,016	606,016	4810,369
CO2	221,0348	221,0348	0	4400,98	4622,015	4622,015	4622,015	4622,015	2609,88
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	360,1356	360,1356	0	0	360,1356	360,1356	360,1356	360,1356	0
C3H8	0	0	0	0	0	0	0	0	0
H2	455,6142	455,6142	0	0	455,6142	455,6142	455,6142	455,6142	959,5838
C	0	0	0	0	0	0	0	0	1,99E-22
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	2,45E-05
H2S	157,8937	157,8937	0	0	157,8937	157,8937	157,8937	157,8937	181,2199
CH3OH	0	0	0	0	0	0	0	0	1,56E-03
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	2,01E-07
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	2,61E-14
HCL	0	0	0	0	0	0	0	0	0
COS	46,41861	46,41861	0	0	46,41861	46,41861	46,41861	46,41861	5,301218
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	395,9224	395,9224	250	100	745,9224	745,9224	745,9224	745,9224	951,9605
Total Flow kg/hr	4182,471	4182,471	4503,82	4400,98	13087,27	13087,27	13087,27	13087,27	13087,27
Total Flow cum/hr	8177,244	1349,794	664,0915	207,8293	2218,648	2218,648	4544,911	4544,911	6159,284
Temperature C	24,99995	359,5146	250	127	284,5219	284,5219	860	860	890,9838
Pressure bar	1,2	15,5	15,5	15,5	15,5	15,5	15,5	15,5	15
Vapor Frac	1	1	1	1	1	1	1	1	1
Liquid Frac	0	0	0	0	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-8,790699	-6,020812	-56,07231	-93,11807	-34,4723	-34,4723	-28,65171	-28,65171	-16,30133
Enthalpy kcal/kg	-832,1479	-569,944	-3112,486	-2115,849	-1964,784	-1964,784	-1633,034	-1633,034	-1185,749
Enthalpy Gcal/hr	-3,480484	-2,383808	-14,01828	-9,311941	-25,71403	-25,71403	-21,37226	-21,37226	-15,51844
Entropy cal/mol-K	-2,392367	-1,324089	-11,6133	-2,12549	-3,002048	-3,002048	4,085303	4,085303	9,404097
Entropy cal/gm-K	-0,226467	-0,1253413	-0,6446361	-0,0482958	-0,1711048	-0,1711048	0,232846	0,232846	0,6840486
Density mol/cc	4,84E-05	2,93E-04	3,76E-04	4,81E-04	3,36E-04	3,36E-04	1,64E-04	1,64E-04	1,55E-04
Density kg/cum	0,5114768	3,098598	6,781927	21,17593	5,89876	5,89876	2,879544	2,879544	2,124804

Substream: MIXED

	SYNG4	SYNG5	SYNG6	CO2+H2O	CO2-R2	SYNG7	SYNG8-1	SYNG8-2	SYNG8-3
Mole Flow kmol/hr									
H2O	203,3433	203,3433	0	203,3433	8,332408	0	0,1010457	0,1010476	0,1010476
O2	3,97E-16	3,97E-16	0	0	0	0	0	0	0
N2	23,18078	23,18078	23,18078	0	0	23,18078	212,6678	212,6664	212,6664
CH4	12,98122	12,98122	12,98122	0	0	12,98122	113,7654	113,765	113,765
CO	171,7351	171,7351	171,7351	0	0	171,7351	61,78879	61,78974	61,78974
CO2	59,30225	59,30225	23,7209	35,58135	35,48339	23,7209	14,55163	14,55163	14,55163
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	476,0124	476,0124	476,0124	0	0	476,0124	765,6573	765,631	765,631
C	1,66E-23	1,66E-23	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	3,82E-07	3,82E-07	3,82E-07	0	0	3,82E-07	1,15E-07	1,15E-07	1,15E-07
H2S	5,317192	5,317192	0	5,317192	5,147622	0	0,1454775	0,1454695	0,1454695
CH3OH	4,86E-05	4,86E-05	4,86E-05	0	0	4,86E-05	3,559766	3,559599	3,559599
CL2	0	0	0	0	0	0	0	0	0
S	6,28E-09	6,28E-09	6,28E-09	0	0	6,28E-09	1,88E-09	1,88E-09	1,88E-09
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	3,26E-16	3,26E-16	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,0882412	0,0882412	0,0882412	0	0	0,0882412	0,0267142	0,0267142	0,0267142
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	3663,286	3663,286	0	3663,286	150,1107	0	1,820366	1,8204	1,8204
O2	1,27E-14	1,27E-14	0	0	0	0	0	0	0
N2	649,3742	649,3742	649,3742	0	0	649,3742	5957,566	5957,526	5957,526
CH4	208,2546	208,2546	208,2546	0	0	208,2546	1825,112	1825,105	1825,105
CO	4810,369	4810,369	4810,369	0	0	4810,369	1730,729	1730,755	1730,755
CO2	2609,88	2609,88	1043,952	1565,928	1561,617	1043,952	640,4145	640,4143	640,4143
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	959,5838	959,5838	959,5838	0	0	959,5838	1543,473	1543,42	1543,42
C	1,99E-22	1,99E-22	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	2,45E-05	2,45E-05	2,45E-05	0	0	2,45E-05	7,35E-06	7,35E-06	7,35E-06
H2S	181,2199	181,2199	0	181,2199	175,4406	0	4,958147	4,957872	4,957872
CH3OH	1,56E-03	1,56E-03	1,56E-03	0	0	1,56E-03	114,0626	114,0572	114,0572
CL2	0	0	0	0	0	0	0	0	0
S	2,01E-07	2,01E-07	2,01E-07	0	0	2,01E-07	6,04E-08	6,04E-08	6,04E-08
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	2,61E-14	2,61E-14	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	5,301218	5,301218	5,301218	0	0	5,301218	1,604893	1,604897	1,604897
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	951,9605	951,9605	707,7187	244,2418	48,96342	707,7187	1172,264	1172,237	1172,237
Total Flow kg/hr	13087,27	13087,27	7676,837	5410,434	1887,168	7676,837	11819,74	11819,66	11819,66
Total Flow cum/hr	3239,151	1983,875	1578,98	108,1241	104,161	209,8739	320,1667	320,1654	499,4296
Temperature C	339,3451	127	127	127	127	101,3675	61,21266	61,21912	250
Pressure bar	15	15	15	15	15	109,8386	106	106	106
Vapor Frac	1	0,9405465	1	0,200471	1	1	1	1	1
Liquid Frac	0	0,0594534	0	0,799529	0	0	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-20,86213	-23,07332	-9,166068	-68,86971	-77,63958	-9,350585	-4,198891	-4,198955	-2,772089
Enthalpy kcal/kg	-1517,499	-1678,34	-845,0091	-3108,967	-2014,394	-862,0196	-416,4397	-416,439	-274,927
Enthalpy Gcal/hr	-19,86021	-21,96521	-6,48709	-16,8211	-3,801554	-6,617679	-4,922279	-4,922239	-3,249591
Entropy cal/mol-K	4,125455	-0,5714747	3,371526	-27,70729	-1,562583	-1,135297	-7,307904	-7,307725	-3,926828
Entropy cal/gm-K	0,3000832	-0,0415687	0,3108171	-1,250783	-0,0405419	-0,1046617	-0,7247869	-0,724757	-0,3894504
Density mol/cc	2,94E-04	4,80E-04	4,48E-04	2,26E-03	4,70E-04	3,37E-03	3,66E-03	3,66E-03	2,35E-03
Density kg/cum	4,04034	6,596821	4,861896	50,03913	18,11779	36,57833	36,91746	36,91736	23,66632

Substream: MIXED

Mole Flow kmol/hr	SYNG9	SYNG10	SYNG11	SYNG12	SYNG13	SYNG14	F-GASES5	F-GASES6	F-GASES7
H2O	0,1010457	0,2357733	0,2357733	22,56669	22,56669	22,56669	0,3477627	0,0109368	0,0109368
O2	0	0	0	0	0	0	0	0	0
N2	212,6678	496,2249	496,2249	708,8928	708,8928	708,8928	707,9765	22,26523	22,26523
CH4	113,7654	265,4527	265,4527	379,2181	379,2181	379,2181	378,1282	11,89179	11,89179
CO	61,78879	144,1738	144,1738	35,36065	35,36065	35,36065	35,34143	1,111456	1,111456
CO2	14,55163	33,95381	33,95381	26,27558	26,27558	26,27558	25,59042	0,8047958	0,8047958
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	765,6573	1786,534	1786,534	2144,033	2144,033	2144,033	2143,57	67,41336	67,41336
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	1,15E-07	2,68E-07	2,68E-07	0	0	0	0	0	0
H2S	0,1454775	0,3394475	0,3394475	0,5730607	0,5730607	0,5730607	0,5006639	0,0157454	0,0157454
CH3OH	3,559766	8,306121	8,306121	204,7859	204,7859	204,7859	12,25052	0,3852677	0,3852677
CL2	0	0	0	0	0	0	0	0	0
S	1,88E-09	4,40E-09	4,40E-09	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0,0267142	0,0623331	0,0623331	9,12E-04	9,12E-04	9,12E-04	8,32E-04	2,62E-05	2,62E-05
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	1,820366	4,247521	4,247521	406,5452	406,5452	406,5452	6,265043	0,19703	0,19703
O2	0	0	0	0	0	0	0	0	0
N2	5957,566	13900,99	13900,99	19858,55	19858,55	19858,55	19832,89	623,7265	623,7265
CH4	1825,112	4258,593	4258,593	6083,705	6083,705	6083,705	6066,221	190,7772	190,7772
CO	1730,729	4038,367	4038,367	990,466	990,466	990,466	989,9275	31,13233	31,13233
CO2	640,4145	1494,3	1494,3	1156,383	1156,383	1156,383	1126,229	35,4189	35,4189
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	0
C3H8	0	0	0	0	0	0	0	0	0
H2	1543,473	3601,438	3601,438	4322,113	4322,113	4322,113	4321,18	135,8972	135,8972
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	7,35E-06	1,71E-05	1,71E-05	0	0	0	0	0	0
H2S	4,958147	11,56901	11,56901	19,53099	19,53099	19,53099	17,06357	0,5366339	0,5366339
CH3OH	114,0626	266,1461	266,1461	6561,781	6561,781	6561,781	392,533	12,34481	12,34481
CL2	0	0	0	0	0	0	0	0	0
S	6,04E-08	1,41E-07	1,41E-07	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	1,604893	3,744751	3,744751	0,054795	0,054795	0,054795	0,050006	1,57E-03	1,57E-03
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	1172,264	2735,283	2735,283	3521,707	3521,707	3521,707	3303,706	103,8986	103,8986
Total Flow kg/hr	11819,74	27579,39	27579,39	39399,13	39399,13	39399,13	32752,36	1030,032	1030,032
Total Flow cum/hr	499,4412	747,0556	721,5453	1642,624	1462,316	887,3116	995,4873	31,30718	2581,85
Temperature C	250	61,21266	50	259,9407	201,7584	30	29,61412	29,61412	25,65213
Pressure bar	106	106	106	98	98	98	86	86	1
Vapor Frac	1	1	1	1	1	0,9378239	1	1	1
Liquid Frac	0	0	0	0	0	0,062176	0	0	0
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-2,771979	-4,198891	-4,283136	-4,249255	-4,724202	-6,623554	-3,228227	-3,228227	-3,228227
Enthalpy kcal/kg	-274,9207	-416,4397	-424,795	-379,8213	-422,2746	-592,0489	-325,6289	-325,6289	-325,6289
Enthalpy Gcal/hr	-3,249537	-11,48532	-11,71576	-14,96485	-16,63749	-23,32655	-10,66527	-0,3354131	-0,3354131
Entropy cal/mol-K	-3,926867	-7,307904	-7,564184	-6,157818	-7,101104	-12,17357	-9,018383	-9,018383	-0,1219286
Entropy cal/gm-K	-0,3894607	-0,7247869	-0,7502044	-0,550419	-0,6347349	-1,08814	-0,9096778	-0,9096778	-0,0122988
Density mol/cc	2,35E-03	3,66E-03	3,79E-03	2,14E-03	2,41E-03	3,97E-03	3,32E-03	3,32E-03	4,02E-05
Density kg/cum	23,66593	36,91746	38,22268	23,98549	26,94298	44,40281	32,90083	32,90083	0,3989513

Substream: MIXED

	F-GASES8	R-SYNG1	R-SYNG2	MEOH+H2O	MEOH	VAPOUR
Mole Flow kmol/hr						
H2O	0,0109368	0,3368189	0,3368189	22,21893	0	22,21893
O2	0	0	0	0	0	0
N2	22,26523	685,712	685,712	0,9162368	0	0,9162368
CH4	11,89179	366,2369	366,2369	1,089866	0	1,089866
CO	1,111456	34,22751	34,22751	0,019223	0	0,019223
CO2	0,8047958	24,78455	24,78455	0,6851565	0	0,6851565
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	67,41336	2076,179	2076,179	0,4630691	0	0,4630691
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0,0157454	0,484925	0,484925	0,0723968	0	0,0723968
CH3OH	0,3852677	11,86584	11,86584	192,5353	192,5353	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	2,62E-05	8,06E-04	8,06E-04	7,97E-05	0	7,97E-05
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Mass Flow kg/hr						
H2O	0,19703	6,067888	6,067888	400,2802	0	400,2802
O2	0	0	0	0	0	0
N2	623,7265	19209,18	19209,18	25,66698	0	25,66698
CH4	190,7772	5875,45	5875,45	17,48445	0	17,48445
CO	31,13233	958,7264	958,7264	0,5384455	0	0,5384455
CO2	35,4189	1090,763	1090,763	30,1536	0	30,1536
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	135,8972	4185,327	4185,327	0,9334918	0	0,9334918
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0,5366339	16,52716	16,52716	2,467419	0	2,467419
CH3OH	12,34481	380,2071	380,2071	6169,248	6169,248	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	1,57E-03	0,048427	0,048427	4,79E-03	0	4,79E-03
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Total Flow kmol/hr	103,8986	3199,828	3199,828	218,0003	192,5353	25,46495
Total Flow kg/hr	1030,032	31722,3	31722,3	6646,778	6169,248	477,5294
Total Flow cum/hr	2360,004	964,1793	850,2799	8,426793	7,834855	1,265968
Temperature C	0	29,61412	52,56067	29,61412	29,61412	29,61412
Pressure bar	1	86	106	86	86	86
Vapor Frac	1	1	1	0	0	0,1270462
Liquid Frac	0	0	0	1	1	0,8729538
Solid Frac	0	0	0	0	0	0
Enthalpy kcal/mol	-3,410082	-3,228017	-3,05947	-58,07835	-57,44374	-63,20818
Enthalpy kcal/kg	-343,9725	-325,6101	-308,6087	-1904,847	-1792,755	-3370,669
Enthalpy Gcal/hr	-0,3543079	-10,32925	-9,789919	-12,66128	-11,06011	-1,609616
Entropy cal/mol-K	-0,75823	-9,018318	-8,910998	-55,89079	-59,29961	-36,33504
Entropy cal/gm-K	-0,0764821	-0,9096778	-0,8988524	-1,8331	-1,850675	-1,93762
Density mol/cc	4,40E-05	3,32E-03	3,76E-03	0,0258699	0,0245742	0,020115
Density kg/cum	0,4364536	32,90083	37,30806	788,7672	787,4107	377,2048

Substream: MIXED

Mole Flow kmol/hr	WATER1-1	WATER1-2	WATER1-3	WATER1-4	STEAM1-5	STEAM1-6	WATER2-1	WATER2-2	WATER2-3
H2O	444,0675	444,0675	250	250	250	250	554,0166	554,0166	556,4522
O2	0	0	0	0	0	0	0	0	5,85E-06
N2	0	0	0	0	0	0	0	0	3,73E-06
CH4	0	0	0	0	0	0	0	0	1,40E-04
CO	0	0	0	0	0	0	0	0	2,80E-06
CO2	0	0	0	0	0	0	0	0	1,04E-04
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	1,88E-05
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	6,05E-05
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	3,38E-03
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	3,30E-05
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Mass Flow kg/hr									
H2O	8000	8000	4503,82	4503,82	4503,82	4503,82	9980,765	9980,765	10024,64
O2	0	0	0	0	0	0	0	0	1,87E-04
N2	0	0	0	0	0	0	0	0	1,05E-04
CH4	0	0	0	0	0	0	0	0	2,24E-03
CO	0	0	0	0	0	0	0	0	7,85E-05
CO2	0	0	0	0	0	0	0	0	4,58E-03
C2H2	0	0	0	0	0	0	0	0	0
C2H4	0	0	0	0	0	0	0	0	0
C2H6	0	0	0	0	0	0	0	0	5,64E-04
C3H8	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	1,22E-04
C	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0,1152681
CH3OH	0	0	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0	0	0
NO	0	0	0	0	0	0	0	0	0
SO3	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0	0	1,98E-03
HCN	0	0	0	0	0	0	0	0	0
ARGON	0	0	0	0	0	0	0	0	0
Total Flow kmol/hr	444,0675	444,0675	250	250	250	250	554,0166	554,0166	556,456
Total Flow kg/hr	8000	8000	4503,82	4503,82	4503,82	4503,82	9980,765	9980,765	10024,77
Total Flow cum/hr	7,971684	138,4551	77,94708	607,2214	687,5259	687,5259	10,04145	10,04145	10,08572
Temperature C	15	198,5387	198,5387	198,5387	250	250	25	25	24,99995
Pressure bar	15	15	15	15	15	15	1,2	1,2	1,2
Vapor Frac	0	0,1202561	0,1202561	1	1	1	0	0	0
Liquid Frac	1	0,8797439	0,8797439	0	0	0	1	1	1
Solid Frac	0	0	0	0	0	0	0	0	0
Enthalpy kcal/mol	-68,91388	-64,17369	-64,17369	-56,53054	-56,06675	-56,06675	-68,72482	-68,72482	-68,7244
Enthalpy kcal/kg	-3825,302	-3562,181	-3562,181	-3137,922	-3112,177	-3112,177	-3814,807	-3814,807	-3814,762
Enthalpy Gcal/hr	-30,60285	-28,49786	-16,04365	-14,13284	-14,01689	-14,01689	-38,07524	-38,07524	-38,24265
Entropy cal/mol-K	-40,77104	-28,67804	-28,67804	-12,47425	-11,54097	-11,54097	-40,1023	-40,1023	-40,10188
Entropy cal/gm-K	-2,263137	-1,591873	-1,591873	-0,6924259	-0,6406212	-0,6406212	-2,226016	-2,226016	-2,22598
Density mol/cc	0,0557056	3,21E-03	3,21E-03	4,12E-04	3,64E-04	3,64E-04	0,055173	0,055173	0,0551726
Density kg/cum	1003,552	57,78048	57,78048	7,417097	6,550764	6,550764	993,957	993,957	993,9562

Substream: MIXED

	WATER3-1	STEAM3-2	WATER4-1	WATER5
Mole Flow kmol/hr				
H2O	222,0337	222,0337	2220,337	166,5253
O2	0	0	0	0
N2	0	0	0	0
CH4	0	0	0	0
CO	0	0	0	0
CO2	0	0	0	0
C2H2	0	0	0	0
C2H4	0	0	0	0
C2H6	0	0	0	0
C3H8	0	0	0	0
H2	0	0	0	0
C	0	0	0	0
NH3	0	0	0	0
SO2	0	0	0	0
H2S	0	0	0	0
CH3OH	0	0	0	0
CL2	0	0	0	0
S	0	0	0	0
NO2	0	0	0	0
NO	0	0	0	0
SO3	0	0	0	0
HCL	0	0	0	0
COS	0	0	0	0
HCN	0	0	0	0
ARGON	0	0	0	0
Mass Flow kg/hr				
H2O	4000	4000	40000	3000
O2	0	0	0	0
N2	0	0	0	0
CH4	0	0	0	0
CO	0	0	0	0
CO2	0	0	0	0
C2H2	0	0	0	0
C2H4	0	0	0	0
C2H6	0	0	0	0
C3H8	0	0	0	0
H2	0	0	0	0
C	0	0	0	0
NH3	0	0	0	0
SO2	0	0	0	0
H2S	0	0	0	0
CH3OH	0	0	0	0
CL2	0	0	0	0
S	0	0	0	0
NO2	0	0	0	0
NO	0	0	0	0
SO3	0	0	0	0
HCL	0	0	0	0
COS	0	0	0	0
HCN	0	0	0	0
ARGON	0	0	0	0
Total Flow kmol/hr	222,0337	222,0337	2220,337	166,5253
Total Flow kg/hr	4000	4000	40000	3000
Total Flow cum/hr	4,538652	123,3021	39,85842	2,989382
Temperature C	134	520	15	15
Pressure bar	108	108	1	1
Vapor Frac	0	1	0	0
Liquid Frac	1	0	1	1
Solid Frac	0	0	0	0
Enthalpy kcal/mol	-66,54437	-54,0867	-68,91965	-68,91965
Enthalpy kcal/kg	-3693,774	-3002,268	-3825,622	-3825,622
Enthalpy Gcal/hr	-14,77531	-12,00925	-153,0271	-11,47703
Entropy cal/mol-K	-34,03407	-12,12243	-40,76664	-40,76664
Entropy cal/gm-K	-1,889178	-0,6728969	-2,262893	-2,262893
Density mol/cc	0,0489206	1,80E-03	0,0557056	0,0557056
Density kg/cum	881,3189	32,44066	1003,552	1003,552

Substream: MIXED

Mole Flow kmol/hr	DH1	DH2	DH3	DH4	DH5	DH6
H2O	194,0675	166,5253	195,0109	2220,337	2775,941	2775,941
O2	0	0	0	0	0	0
N2	0	0	0	0	0	0
CH4	0	0	0	0	0	0
CO	0	0	0	0	0	0
CO2	0	0	0,0979653	0	0,0979653	0,0979653
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	0	0	0	0	0	0
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0	0	0,1695706	0	0,1695706	0,1695706
CH3OH	0	0	0	0	0	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	0	0	0	0	0	0
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Mass Flow kg/hr						
H2O	3496,18	3000	3513,175	40000	50009,36	50009,36
O2	0	0	0	0	0	0
N2	0	0	0	0	0	0
CH4	0	0	0	0	0	0
CO	0	0	0	0	0	0
CO2	0	0	4,311433	0	4,311433	4,311433
C2H2	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C2H6	0	0	0	0	0	0
C3H8	0	0	0	0	0	0
H2	0	0	0	0	0	0
C	0	0	0	0	0	0
NH3	0	0	0	0	0	0
SO2	0	0	0	0	0	0
H2S	0	0	5,779285	0	5,779285	5,779285
CH3OH	0	0	0	0	0	0
CL2	0	0	0	0	0	0
S	0	0	0	0	0	0
NO2	0	0	0	0	0	0
NO	0	0	0	0	0	0
SO3	0	0	0	0	0	0
HCL	0	0	0	0	0	0
COS	0	0	0	0	0	0
HCN	0	0	0	0	0	0
ARGON	0	0	0	0	0	0
Total Flow kmol/hr	194,0675	166,5253	195,2784	2220,337	2776,209	2776,209
Total Flow kg/hr	3496,18	3000	3523,266	40000	50019,45	50019,45
Total Flow cum/hr	60,50798	646,3544	3,963022	9029,369	11818,97	11818,99
Temperature C	198,5387	101,5389	127	101,5389	101,5178	101,5176
Pressure bar	15	1	15	1	1	1
Vapor Frac	0,1202561	0,1251028	0	0,1311039	0,1372847	0,137285
Liquid Frac	0,8797439	0,8748972	1	0,8688961	0,8627153	0,862715
Solid Frac	0	0	0	0	0	0
Enthalpy kcal/mol	-64,17369	-65,96752	-66,67078	-65,90708	-65,84325	-65,84325
Enthalpy kcal/kg	-3562,181	-3661,754	-3695,254	-3658,399	-3654,471	-3654,471
Enthalpy Gcal/hr	-12,4542	-10,98542	-13,01955	-146,3381	-182,7972	-182,7972
Entropy cal/mol-K	-28,67804	-32,27063	-34,26272	-32,10931	-31,94079	-31,94079
Entropy cal/gm-K	-1,591873	-1,791292	-1,899024	-1,782338	-1,772796	-1,772796
Density mol/cc	3,21E-03	2,58E-04	0,0492751	2,46E-04	2,35E-04	2,35E-04
Density kg/cum	57,78048	4,641417	889,0353	4,429989	4,232132	4,232125

	LUKAB-TB	COMP-GAS	MEOH-COM	R-COMP	TOTAL WORK
Power MW	-	-	1,54723349	0,69751317	2,24474666

Appendix E
Economical calculations

E.1 Economical calculations for the IGT methanol system

Table 45 Economical calculations for the IGT system

Investments IGT-Methanol

Installed Unit	Base investment cost MUS$-2001	Scale factor	Base scale	Actual scale	Maximum Size		Overall installation factor	Actual TOT investment cost MUS$
Pre-treatment								
Conveyers	0,35	0,80	33,5	13,46	110	tonne/hour wet biomass	1,86	0,36
Grinding	0,41	0,60	33,5	13,46	110	tonne/hour wet biomass	1,86	0,50
Storage	1,00	0,65	33,5	13,46	110	tonne/hour wet biomass	1,86	1,17
Dryer	7,60	0,80	33,5	13,46	110	tonne/hour wet biomass	1,86	7,76
Iron Removal	0,37	0,70	33,5	13,46	110	tonne/hour wet biomass	1,86	0,41
Feeding System	0,41	1,00	33,5	13,46	110	tonne/hour wet biomass	1,86	0,35
Gasification System								
BCL	16,30	0,65	68,8	0	83	tonne/hour dry biomass	1,69	0,00
IGT	38,10	0,70	68,8	11,08	75	tonne/hour dry biomass	1,69	20,42
Oxygen Plant (installed)	44,20	0,85	41,7	8,00	-	tonne O2/hour	1,00	12,36
Gas Cleaning								
Tar Cracker	3,10	0,70	34,2	0	52	m3 gas/s	1,86	0,00
Cyclones	2,60	0,70	34,2	0,77	180	m3 gas/s	1,86	0,39
High-temperature Heat Exchanger	6,99	0,60	39,2	5,99	-	kg steam/s	1,84	4,74
Baghouse Filter	1,60	0,65	12,1	0	64	m3 gas/s	1,86	0,00
Condensing Scrubber	2,60	0,70	12,1	0	64	m3 gas/s	1,86	0,00
Hot Gas Cleaning	30,00	1,00	74,1	0,73	-	m3 gas/s	1,72	0,58
Syngas Processing								
Compressor	11,10	0,85	13,2	2,48	-	MWe	1,72	5,25
Steam Reformer	9,40	0,60	1390,0	0	-	kmol total/hour	2,30	0,00
Autothermal Reformer	4,70	0,60	1390,0	1864,44	-	kmol total/hour	2,30	14,68
Shift Reactor (installed)	36,90	0,85	15,6	0,10	-	Mmol CO+H2/hour	1,00	0,57
Selexol CO2 removal (installed)	54,10	0,70	9909,0	235,65	-	kmol CO2/hour	1,00	4,50
Methanol Production					-			
Make Up Compressor	11,10	0,85	13,2	1,54	-	MWe	1,72	3,50
Gas Phase Methanol Reactor	7,00	0,60	87,5	11,09	-	tonne MeOH/hour	2,10	4,85
Recycle Compressor	11,10	0,85	13,2	1,10	-	MWe	1,72	2,63
Refining	15,10	0,70	87,5	10,28	-	tonne MeOH/hour	2,10	
Power Isle					-			
Steam Turbine + Steam System	5,10	0,70	10,3	0	-	MWe	1,86	0,00
Expansion Turbine	4,30	0,70	10,3	0	-	MWe	1,86	0,00
Total Installed Investment								85,00
Total Installed Investment corrected for lifetime								76,86
LHV dry biomass	GJ/tonne,dry	19,28						17,07
Biomass input	MWth	428,4						52,5
Biomass input	tonnedry/h	80,00						11,08
Load hours	h/a	8000						8000
Biomass input	GJ/a LHV	12339200						1513357,92
LHV Coke oven gas	MJ/Nm3	17,5						17,5
Coke oven gas input	Nm3/h							10286
Annual Cost/income								
Capital								10,1
Operating and Maintenance								3,40
Biomass								5,70
Coke oven gas								5,54
Auxiliaries (el. compressors, pumps and drier)								2,83
Oxygen (el. oxygen separator)								1,33
Income Electricity								-1,08
Income Disctrict Heating								0
Total Annual Cost								27,8
Production								
Methanol produced	tonne/h							10,28
Fuel output	MW LHV							59,96
Efficiency fuel	% LHV							53,60
Cost of fuel produced	US$/GJ LHV							16,1
Cost of fuel produced	US$/liter							0,27
Cost of fuel produced	SEK/liter							1,98

Internal rate	0,1
Economical lifetime	15
Technical lifetime	25
Currency 1$=xSEK	7,38
Consumer prices 2006/2001	1,138
Biomass price SEK/MWh	100
Electricity price SEK/MWh	500
Coke oven gas price SEK/GJ	28,4

E.2 Economical calculations for the BCL methanol system

Table 46 Economical calculations for the BCL system

Investments BCL-Methanol

Installed Unit	Base investment cost MUS$-2001	Scale factor	Base scale	Actual scale	Maximum Size		Overall installation factor	Actual TOT investment cost MUS$-2006
Pre-treatment								
Conveyers	0,35	0,80	33,5	14,09	110	tonne/hour wet biomass	1,86	0,37
Grinding	0,41	0,60	33,5	14,09	110	tonne/hour wet biomass	1,86	0,52
Storage	1,00	0,65	33,5	14,09	110	tonne/hour wet biomass	1,86	1,21
Dryer	7,60	0,80	33,5	14,09	110	tonne/hour wet biomass	1,86	8,05
Iron Removal	0,37	0,70	33,5	14,09	110	tonne/hour wet biomass	1,86	0,43
Feeding System	0,41	1,00	33,5	14,09	110	tonne/hour wet biomass	1,86	0,37
Gasification System								
BCL	16,30	0,65	68,8	10,96	83	tonne/hour dry biomass	1,69	9,50
IGT	38,10	0,70	68,8	0	75	tonne/hour dry biomass	1,69	0,00
Oxygen Plant (installed)	44,20	0,85	41,7	0	-	tonne O2/hour	1,00	0,00
Gas Cleaning								
Tar Cracker	3,10	0,70	34,2	10,98	52	m3 gas/s	1,86	2,96
Cyclones	2,60	0,70	34,2	10,98	180	m3 gas/s	1,86	2,49
High-temperature Heat Exchanger	6,99	0,60	39,2	3,05	-	kg steam/s	1,84	3,16
Baghouse Filter	1,60	0,65	12,1	7,81	64	m3 gas/s	1,86	2,55
Condensing Scrubber	2,60	0,70	12,1	6,45	64	m3 gas/s	1,86	3,54
Hot Gas Cleaning	30,00	1,00	74,1	0	-	m3 gas/s	1,72	0,00
Syngas Processing								
Compressor	11,10	0,85	13,2	2,80	-	MWe	1,72	5,81
Steam Reformer	9,40	0,60	1390,0	2347,03	-	kmol total/hour	2,30	33,70
Autothermal Reformer	4,70	0,60	1390,0	0	-	kmol total/hour	2,30	0,00
Shift Reactor (installed)	36,90	0,85	15,6	0	-	Mmol CO+H2/hour	1,00	0,00
Selexol CO2 removal (installed)	54,10	0,70	9909,0	128,25	-	kmol CO2/hour	1,00	2,94
Methanol Production					-			
Make Up Compressor	11,10	0,85	13,2	2,75	-	MWe	1,72	5,73
Gas Phase Methanol Reactor	7,00	0,60	87,5	13,10	-	tonne MeOH/hour	2,10	5,35
Recycle Compressor	11,10	0,85	13,2	1,29	-	MWe	1,72	3,01
Refining	15,10	0,70	87,5	12,17	-	tonne MeOH/hour	2,10	9,08
Power Isle					-			
Steam Turbine + Steam System	5,10	0,70	10,3	0	-	MWe	1,86	0,00
Expansion Turbine	4,30	0,70	10,3	0	-	MWe	1,86	0,00
Total Installed Investment								100,76
Total Installed Investment corrected for lifetime								91,11
LHV dry biomass	GJ/tonne,dry	19,28						17,25
Biomass input	MWth	428,4						52,5
Biomass input	tonnedry/h	80,00						10,96
Load hours	h/a	8000						8000
Biomass input	GJ/a LHV	12339200						1512480
LHV Coke oven gas	MJ/Nm3	17,5						17,5
Coke oven gas input	Nm3/h							10286
Annual Cost/income								
Capital								12,0
Operating and Maintenance								4,03
Biomass								5,69
Coke oven gas								5,54
Auxiliaries (el. compressors, pumps and drier)								3,77
Oxygen (el. oxygen separator)								0,00
Income Electricity								-0,80
Income Disctrict Heating								0
Total Annual Cost								30,2
Production								
Methanol produced	tonne/h							12,17
Fuel output	MW LHV							71,02
Efficiency fuel	% LHV							64,82
Cost of fuel produced	US$/GJ LHV							14,8
Cost of fuel produced	US$/liter							0,25
Cost of fuel produced	SEK/liter							1,82

Internal rate	0,1
Economical lifetime	15
Technical lifetime	25
Currency 1$=xSEK	7,38
Consumer prices 2006/2001	1,138
Biomass price SEK/MWh	100
Electricity price SEK/MWh	500
Coke oven gas price SEK/GJ	28,4

E.3 Economical calculations for the ATR methanol system

Table 47 Economical calculations for the ATR system

Investments ATR-Methanol

Installed Unit	Base investment cost MUS$-2001	Scale factor	Base scale	Actual scale	Maximum Size		Overall installation factor	Actual TOT investment cost MUS$-2006
Pre-treatment								
Conveyers	0,35	0,80	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Grinding	0,41	0,60	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Storage	1,00	0,65	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Dryer	7,60	0,80	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Iron Removal	0,37	0,70	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Feeding System	0,41	1,00	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Gasification System								
BCL	16,30	0,65	68,8	0	83	tonne/hour dry biomass	1,69	0,00
IGT	38,10	0,70	68,8	0	75	tonne/hour dry biomass	1,69	0,00
Oxygen Plant (installed)	44,20	0,85	41,7	2,85	-	tonne O2/hour	1,00	5,14
Gas Cleaning								
Tar Cracker	3,10	0,70	34,2	0	52	m3 gas/s	1,86	0,00
Cyclones	2,60	0,70	34,2	0	180	m3 gas/s	1,86	0,00
High-temperature Heat Exchanger	6,99	0,60	39,2	1,33	-	kg steam/s	1,84	1,93
Baghouse Filter	1,60	0,65	12,1	0	64	m3 gas/s	1,86	0,00
Condensing Scrubber	2,60	0,70	12,1	0	64	m3 gas/s	1,86	0,00
Hot Gas Cleaning	30,00	1,00	74,1	0,30	-	m3 gas/s	1,72	0,24
Syngas Processing								
Compressor	11,10	0,85	13,2	2,12	-	MWe	1,72	4,60
Steam Reformer	9,40	0,60	1390,0	0	-	kmol total/hour	2,30	0,00
Autothermal Reformer	4,70	0,60	1390,0	857,73	-	kmol total/hour	2,30	9,21
Shift Reactor (installed)	36,90	0,85	15,6	0	-	Mmol CO+H2/hour	1,00	0,00
Selexol CO2 removal (installed)	54,10	0,70	9909,0	100	-	kmol CO2/hour	1,00	2,47
Methanol Production					-			
Make Up Compressor	11,10	0,85	13,2	1,08	-	MWe	1,72	2,59
Gas Phase Methanol Reactor	7,00	0,60	87,5	5,59	-	tonne MeOH/hour	2,10	3,21
Recycle Compressor	11,10	0,85	13,2	0,55	-	MWe	1,72	1,47
Refining	15,10	0,70	87,5	5,18	-	tonne MeOH/hour	2,10	4,99
Power Isle					-			
Steam Turbine + Steam System	5,10	0,70	10,3	0	-	MWe	1,86	0,00
Expansion Turbine	4,30	0,70	10,3	0	-	MWe	1,86	0,00
Total Installed Investment								35,84
Total Installed Investment corrected for lifetime								32,41
LHV dry biomass	GJ/tonne,dry	19,28						17,07
Biomass input	MWth	428,4						0,0
Biomass input	tonnedry/h	80,00						0,00
Load hours	h/a	8000						8000
Biomass input	GJ/a LHV	12339200						0
LHV Coke oven gas	MJ/Nm3	17,5						17,5
Coke oven gas input	Nm3/h							10286
Annual Cost/income								
Capital								4,3
Operating and Maintenance								1,43
Biomass								0,00
Coke oven gas								5,54
Auxiliaries (el. compressors and pumps)								2,04
Oxygen (el. oxygen separator)								0,47
Income Electricity								-0,64
Income Disctrict Heating								0
Total Annual Cost								13,1
Production								
Methanol produced	tonne/h							5,18
Fuel output	MW LHV							30,24
Efficiency fuel	% LHV							56,24
Cost of fuel produced	US$/GJ LHV							15,0
Cost of fuel produced	US$/liter							0,25
Cost of fuel produced	SEK/liter							1,85

Internal rate	0,1
Economical lifetime	15
Technical lifetime	25
Currency 1$=xSEK	7,38
Consumer prices 2006/2001	1,138
Biomass price SEK/MWh	100
Electricity price SEK/MWh	500
Coke oven gas price SEK/GJ	28,4

E.4 Economical calculations for the SMR methanol system

Table 48 Economical calculations for the SMR system

Investments SMR-Methanol

Installed Unit	Base investment cost MUS$-2001	Scale factor	Base scale	Actual scale	Maximum Size		Overall installation factor	Actual TOT investment cost MUS$-2006
Pre-treatment								
Conveyers	0,35	0,80	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Grinding	0,41	0,60	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Storage	1,00	0,65	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Dryer	7,60	0,80	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Iron Removal	0,37	0,70	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Feeding System	0,41	1,00	33,5	0	110	tonne/hour wet biomass	1,86	0,00
Gasification System								
BCL	16,30	0,65	68,8	0	83	tonne/hour dry biomass	1,69	0,00
IGT	38,10	0,70	68,8	0	75	tonne/hour dry biomass	1,69	0,00
Oxygen Plant (installed)	44,20	0,85	41,7	0	-	tonne O2/hour	1,00	0,00
Gas Cleaning								
Tar Cracker	3,10	0,70	34,2	0	52	m3 gas/s	1,86	0,00
Cyclones	2,60	0,70	34,2	0	180	m3 gas/s	1,86	0,00
High-temperature Heat Exchanger	6,99	0,60	39,2	1,11	-	kg steam/s	1,84	1,73
Baghouse Filter	1,60	0,65	12,1	2,86	64	m3 gas/s	1,86	1,33
Condensing Scrubber	2,60	0,70	12,1	2,86	64	m3 gas/s	1,86	2,00
Hot Gas Cleaning	30,00	1,00	74,1	0	-	m3 gas/s	1,72	0,00
Syngas Processing								
Compressor	11,10	0,85	13,2	1,28	-	MWe	1,72	2,98
Steam Reformer	9,40	0,60	1390,0	1298,64	-	kmol total/hour	2,30	23,63
Autothermal Reformer	4,70	0,60	1390,0	0	-	kmol total/hour	2,30	0,00
Shift Reactor (installed)	36,90	0,85	15,6	0	-	Mmol CO+H2/hour	1,00	0,00
Selexol CO2 removal (installed)	54,10	0,70	9909,0	100	-	kmol CO2/hour	1,00	2,47
Methanol Production					-			
Make Up Compressor	11,10	0,85	13,2	1,55	-	MWe	1,72	3,51
Gas Phase Methanol Reactor	7,00	0,60	87,5	6,56	-	tonne MeOH/hour	2,10	3,54
Recycle Compressor	11,10	0,85	13,2	0,70	-	MWe	1,72	1,79
Refining	15,10	0,70	87,5	6,17	-	tonne MeOH/hour	2,10	5,64
Power Isle					-			
Steam Turbine + Steam System	5,10	0,70	10,3	0	-	MWe	1,86	0,00
Expansion Turbine	4,30	0,70	10,3	0	-	MWe	1,86	0,00
Total Installed Investment								48,61
Total Installed Investment corrected for lifetime								43,95
LHV dry biomass	GJ/tonne,dry	19,28						17,07
Biomass input	MWth	428,4						0,0
Biomass input	tonnedry/h	80,00						0,00
Load hours	h/a	8000						8000
Biomass input	GJ/a LHV	12339200						0
LHV Coke oven gas	MJ/Nm3	17,5						17,5
Coke oven gas input	Nm3/h							10286
Annual Cost/income								
Capital								5,8
Operating and Maintenance								1,94
Biomass								0,00
Coke oven gas								5,54
Auxiliaries (el. compressors and pumps)								1,91
Oxygen (el. oxygen separator)								0,00
Incom Electricity								-0,53
Income Disctrict Heating								0
Total Annual Cost								14,6
Production								
Methanol produced	tonne/h							6,17
Fuel output	MW LHV							35,99
Efficiency fuel	% LHV							67,24
Cost of fuel produced	US$/GJ LHV							14,1
Cost of fuel produced	US$/liter							0,24
Cost of fuel produced	SEK/liter							1,74

Internal rate	0,1
Economical lifetime	15
Technical lifetime	25
Currency 1$=xSEK	7,38
Consumer prices 2006/2001	1,138
Biomass price SEK/MWh	100
Electricity price SEK/MWh	500
Coke oven gas price SEK/GJ	28,4

Appendix F

Statistics

F.1 Population in Norrbotten.

Table 49 Population in the county of Norrbotten in the year 2005 [27]

Code	County and communities	Population	Growth	Live offspring	Dead	Birth surplus	Move into	Move out
25	**Norrbottens county**	**251 740**	**-845**	**2 291**	**2 802**	**-511**	**6 215**	**6 562**
2506	Arjeplog	3 159	-65	18	49	-31	104	134
2505	Arvidsjaur	6 814	-80	54	100	-46	250	283
2582	Boden	28 176	-101	252	313	-61	1 217	1 255
2523	Gällivare	19 077	-127	136	243	-107	577	598
2583	Haparanda	10 184	-24	111	137	-26	621	619
2510	Jokkmokk	5 534	-65	33	77	-44	220	241
2514	Kalix	17 483	-170	148	215	-67	555	659
2584	Kiruna	23 135	-119	222	247	-25	741	834
2580	Luleå	72 751	186	752	631	121	3 460	3 411
2521	Pajala	6 798	-127	44	103	-59	184	254
2581	Piteå	40 873	43	389	429	-40	1 481	1 396
2560	Älvsbyn	8 655	-120	67	117	-50	297	367
2513	Överkalix	3 872	-61	27	67	-40	115	137
2518	Övertorneå	5 229	-15	38	74	-36	254	235

F.2 Gasoline and diesel consumption per inhabitant and mean car.

Table 50 Gasoline and diesel consumption per inhabitant and mean car in the county of Norrbotten [28]

NR. 6. Gasoline and diesel consumption per inhabitant and mean car
(with respect to car possession divided among car models, the cars fuel consumption per 10 km and driving distances per car model.)

Code	County and communities	Year 2003 Diesel		Year 2003 Gasoline		Year 2003 Fuel
		Liter/inhabitant.	Liter/mean car	Liter/inhabitant.	Liter/mean car	Liter/inhabitant.
25	Norrbottens community	107	1,495	500	1,048	607
2505	Arvidsjaur	165	1,416	446	1,006	611
2506	Arjeplog	125	1,475	524	1,089	649
2510	Jokkmokk	109	1,479	515	1,127	624
2513	Överkalix	214	1,380	402	961	615
2514	Kalix	135	1,449	469	1,028	604
2518	Övertorneå	239	1,485	336	984	575
2521	Pajala	161	1,389	409	1,008	570
2523	Gällivare	100	1,554	525	1,069	625
2560	Älvsbyn	152	1,378	557	945	709
2580	Luleå	82	1,680	507	1,127	590
2581	Piteå	88	1,626	548	1,093	636
2582	Boden	98	1,552	497	1,089	595
2583	Haparanda	164	1,392	425	1,030	589
2584	Kiruna	80	1,602	502	1,080	582