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Techno-Economic Evaluation of HYCO Plant Integrated to Ammonia / Urea or Methanol Production with CCS

IEA GREENHOUSE GAS R&D PROGRAMME

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Techno-Economics of Deploying CCS in a Natural Gas Based Production of Methanol and Ammonia/Urea

Key Messages

- The IEA Greenhouse Gas R&D Programme (IEAGHG) have systematically evaluated the performance and cost of integrating CO_2 Capture and Storage (CCS) in various energy intensive industries. To date, the programme has assessed the economics of deploying CCS in the cement, iron and steel, hydrogen production from NG, pulp and paper industry, and with study in the oil refining industry underway.

- Ammonia/urea and methanol production are the pillars of the basic chemicals industry worldwide. Ammonia/urea is an important commodity used in the agriculture (fertiliser) and food industry. Whilst methanol is an important feedstock in production of various chemicals and fuel used in our daily lives. Globally (except for China), these commodities are mainly produced from NG or light hydrocarbons.

- The study presented a detailed baseline information of the performance and cost of deploying CO_2 capture in a SMR Based HyCO plant using natural gas as feedstock / fuel and operating as a captive plant (i.e. integrated within an industrial complex). This study presents the levelised cost of methanol and urea production and its corresponding CO_2 avoidance cost.

- The addition of capture of CO_2 from the flue gas of SMR increases the energy demand of the plant.
    - For the ammonia/urea production – an additional 8.6MWe of electricity is imported from the grid.
    - For the methanol production – an additional 17.9 MWe of electricity is imported from the grid.

- In general, the addition of CCS increases the levelised cost of production.
    - For the ammonia/urea production – this increases by 23 €/t urea.
    - For the methanol production – this increases by 24 €/t methanol

- The CO_2 avoided cost of capturing additional CO_2 from the SMR plant is in the range of €80 to 100 per tonne CO_2 for both cases.

- The use of CO_2 in the production of urea or methanol is considered a mature technology. However, the economics of having or integrating both industrial CCS and CCU is not yet well established in the open literature. This study should provide a good basis for future study in understanding the performance and cost of implementing both industrial CCS and CCU.

Background to the Study

The IEA Greenhouse Gas R&D Programme (IEAGHG) has undertaken a series of studies evaluating the performance and cost of capturing CO_2 from different energy intensive industries.

The ammonia and methanol industry is an allied industry very related to the production of hydrogen or HyCO gas. Globally, around 60% of the produced hydrogen is consumed by these industries. Outside China, production of these chemicals from natural gas is predominant. In fact, the production of ammonia and methanol is always an important strategy on how natural gas assets are monetised.

An important aspects of this study is to demonstrate how an SMR based hydrogen/HyCO production is integrated to an industrial complex. Furthermore, it is essential to understand the different aspects of the production process and how will it be affected when additional CO_2 is captured from the SMR's flue gas.

In the case of ammonia production, it is essential to remove the bulk of the CO_2 from the syngas produced from the reforming process. In other words, CO_2 capture is an inherent part of the production process. Normally, the captured CO_2 is used to produce urea in the downstream process. But not all plants would be able to consume all the captured CO_2 from the process as most of the fertiliser complex also produces other types of ammonia products. Thus, any surplus CO_2 is normally vented to the atmosphere. For this reason, this type of process is considered a low hanging fruit for demonstrating industrial CCS (for example: the ACTL project in Canada and the Yara project in Norway).

In the case of methanol production, the CO_2 is part of the methanol synthesis reaction. As such, the bulk of the carbon in the natural gas should end up in the methanol product. Nonetheless, it is important to understand the implication of capturing CO_2 from the methanol production as it is related to how this industry could become an important player in the industrial CCU context.

IEAGHG has commissioned this study to evaluate the performance and cost of deploying CO_2 capture and storage in mega-plants producing urea and methanol from natural gas as feedstock. The results presented in this study should form the basis of future studies in industrial CCS and CCU.

Scope of Work

Study Cases

The study assessed the performance and cost of a new build plant producing ammonia/urea and methanol without and with CCS. Specifically, this study examines how an SMR based hydrogen/HyCO/syngas production are integrated into an industrial complex.

The following cases were evaluated:

- Ammonia with Urea Production. In this case, the syngas generation unit or HyCO plant, that uses natural gas as feedstock/fuel, is integrated to a nominal 1350 t/d ammonia plant. Around 95% of the ammonia produced is further converted in a downstream urea plant producing 2260 t/d urea using the CO_2 captured from the syngas (as part of the process). For the capture case, the same ammonia and urea complex as described above but would have additional CO_2 to be captured from the flue gas of the SMR. The plant will maximise the urea production producing 2380 t/d urea.

- Methanol Production. In this case, the syngas generation unit or HyCO plant, that uses natural gas as feedstock/fuel, is integrated into a nominal 5000 t/d methanol plant. For the capture case, the same methanol plant would capture the CO_2 from the flue gas of the SMR.

Both cases assessed in this study represent the average plant size that will be built in the short to medium term.

Figures 1 and 2 present the simplified block flow diagram of the ammonia/urea production without and with CO_2 capture. Figures 3 and 4 present the block flow diagram of the methanol production without and with CO_2 capture.

Basis of Design

a.) HyCO plant / Syngas Generation

Both cases examined involve a syngas generation unit based on two-step reforming (also known as tandem reforming).

For the ammonia production case, the syngas is produced from steam methane reforming (SMR) in tandem with an air-blown auto thermal reforming (ATR). The syngas produced then passes through a series of processes which include the high temperature and low temperature shift, the bulk CO_2 removal unit, the methanation process before being fed into the ammonia/urea synthesis plant.

For the methanol production case, the syngas is produced from the SMR in tandem with an oxygen blown ATR. The syngas (also called make-up gas or MUG) produced is cooled and compressed. This is then mixed with the recycled gas from the methanol synthesis reactor before being fed into the methanol synthesis plant.

b.) Ammonia / Urea Synthesis

The ammonia synthesis takes place in the Ammonia Synthesis Converter (normally cooled by steam generation or quenched). The product gas from the converter is then refrigerated to $0^{o}C$ to let the ammonia condensed. The ammonia is then sent to the urea synthesis plant.

The Urea plant primarily consists of the (a.) synthesis reactor, (b.) stripper, (c.) carbamate condenser, (d.) decomposer, (e.) vacuum concentrator, (f.) evaporator; and (g.) granulation plant.

c.) Methanol Synthesis

The methanol synthesis takes place in a methanol synthesis reactor (normally cooled by steam generation or quenched). The product gas is then cooled and sent to a methanol catchpot where the crude methanol is separated and sent to the purification section which consist of 3 column distillation unit producing grade AA methanol.

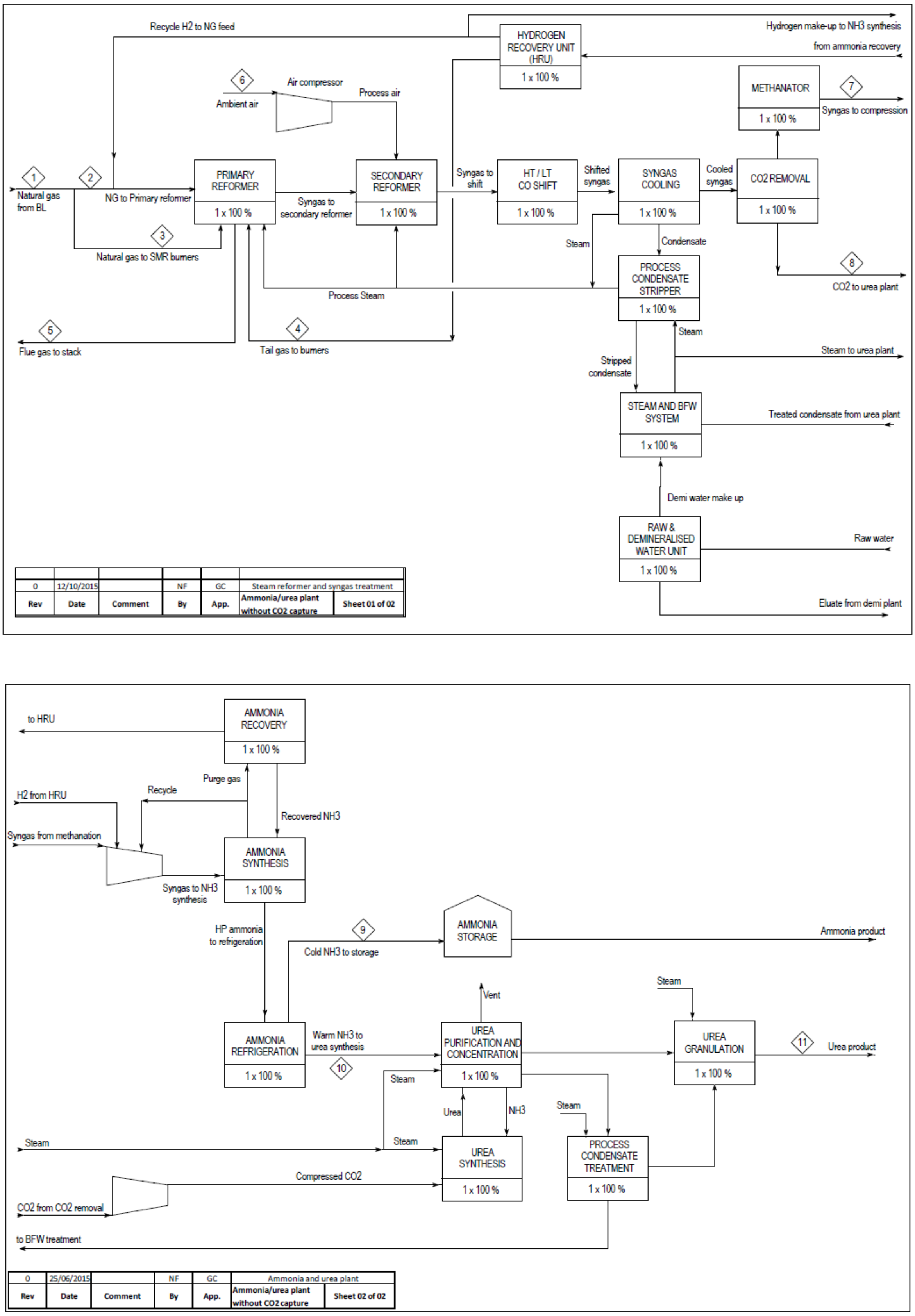


Figure 1: Simplified block flow diagram of the ammonia/urea production without CCS

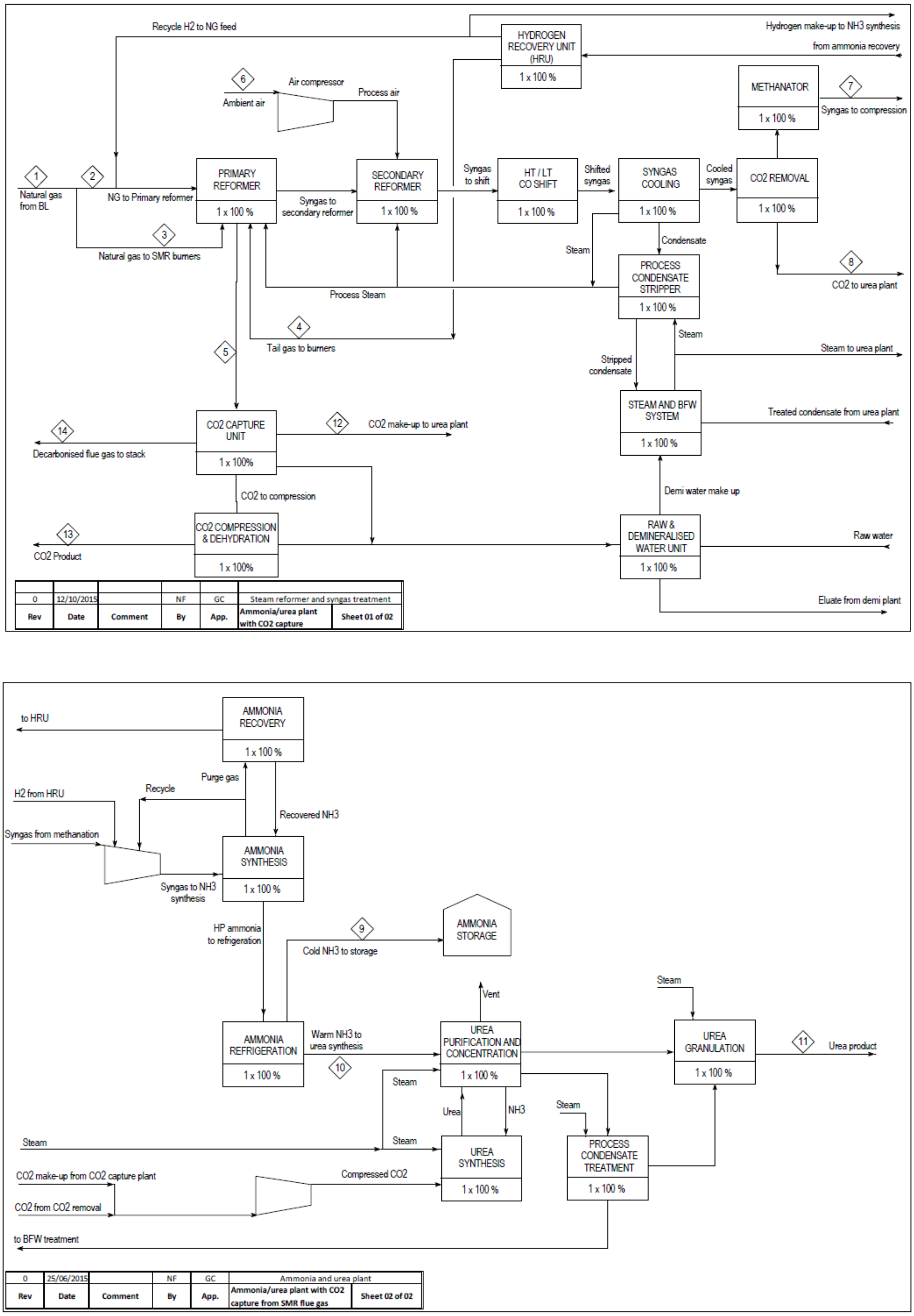


Figure 2: Simplified block flow diagram of the ammonia/urea production with CCS

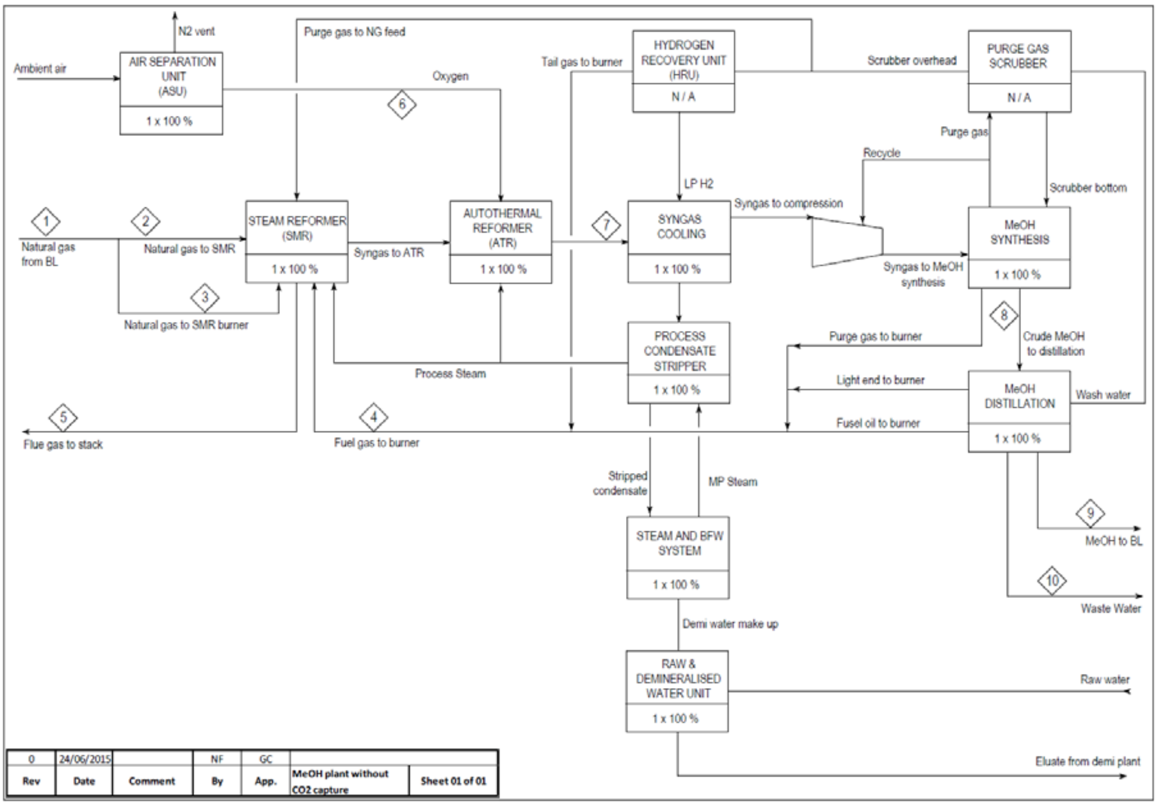


Figure 3: Simplified block flow diagram of the methanol plant without CCS

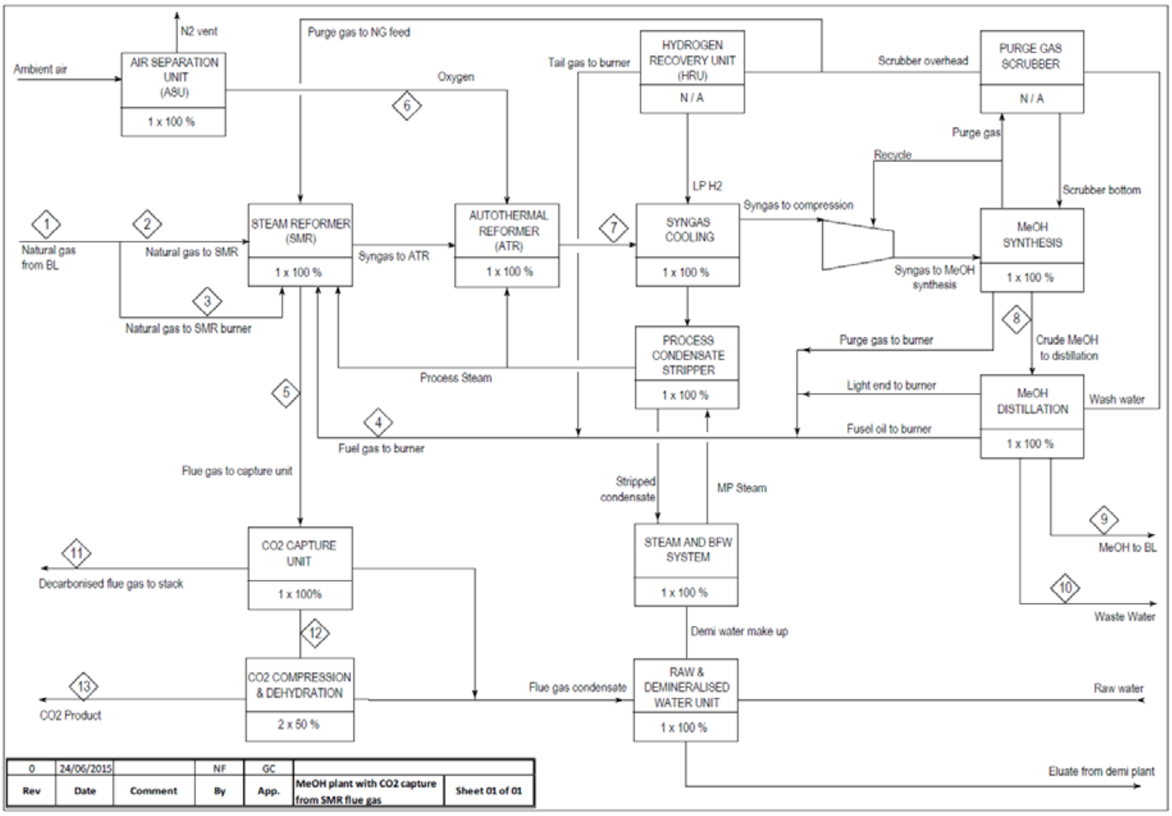


Figure 4: Simplified block flow diagram of the methanol plant with CCS

Economic Assessment Criteria and Key Assumptions

Where applicable, the criteria for the economic evaluation used in this study is based on the information retrieved from IEAGHG document “Criteria for Technical and Economic Assessment of Plants with Low CO_2 Emissions” Version C-6, March 2014. Other key criteria and assumptions relevant to the operation of the methanol plant are based on the information provided by Amec Foster Wheeler.

The criteria used in the evaluation of the cost of hydrogen production and CO_2 avoided cost are summarized below:

- Plant Life: The plant is design for an economic life of 25 years.
- Financial leverage (debt / invested capital): All capital requirements are treated as debt, i.e. financial leverage equal to 100%.
- Discount Rate: Discounted cash flow analysis is used to evaluate the levelised cost of H_2 production and CO_2 avoided cost. The discount rate of 8% is assumed.
- Inflation Rate: Not considered in the discounted cash flow analysis.
- Depreciation: Not considered in the discounted cash flow analysis. The results presented in this study is reported on Earnings Before Interest, Taxes, Depreciation and Amortization (EBITDA) basis.
- Design and Construction Period

  The design and construction period and the curve of capital expenditure assumed in this study is presented below:

  ⇒ Construction period: 3 years

  ⇒ Curve of capital expenditure

Year	Investment cost %
1	20
2	45
3	35

- Decommissioning Cost: This is not included in the discounted cash flow analysis. The salvage value of equipment and materials is normally assumed to be equal to the costs of dismantling and site restoration, resulting to a zero net cost for decommissioning.
- Estimate accuracy: The estimate is based on AACE Class 4 estimate (with accuracy in the range of +35%/-15%), using 4Q2014 price level, in euro (€).

The cost of the plant is estimated based on the capital expenditure and annual operating cost. These are defined as follows:

- Capital Cost

  The capital cost is presented as the Total Plant Cost (TPC) and the Total Capital Requirement (TCR).

  TPC is defined as the installed cost of the plant, including project contingency. This is broken down into:
  ⇒ Direct materials
  ⇒ Construction
  ⇒ EPC services
  ⇒ Other costs
  ⇒ Contingency

  TCR is defined as the sum of:
  ⇒ Total plant cost (TPC)
  ⇒ Interest during construction
  ⇒ Owner's costs
  ⇒ Spare parts cost
  ⇒ Working capital
  ⇒ Start-up costs

  For each of the cases the TPC has been determined through a combination of licensor/vendor quotes, the use of Amec Foster Wheeler's in-house database and the development of conceptual estimating models, based on the specific characteristics, materials and design conditions of each item of equipment in the plant. The other components of the TCR have been estimated mainly as percentages of the TPC of the plant.

- Fixed Operating Cost

  The fixed operating cost includes the following:
  ⇒ direct labour cost
  ⇒ administrative and general overhead cost
  ⇒ annual operating and maintenance cost
  ⇒ insurance
  ⇒ local taxes and fees

- Variable Operating Cost

  The variable operating cost include the consumptions of the following key items:
  ⇒ Feedstock (natural gas)
  ⇒ Raw water make-up
  ⇒ Catalysts
  ⇒ Chemicals

The assumptions used to calculate of the total capital requirements:

- Spare parts cost: 0.5% of the TPC is assumed to cover spare part costs. It is also assumed that spare parts have no value at the end of the plant life due to obsolescence.

- Start-up costs consist of:
  - ⇒ 2% of TPC, to cover modifications to equipment that needed to bring the unit up to full capacity.
  - ⇒ 25% of the full capacity feedstock and fuel cost for one month, to cover inefficient operation that occurs during the start-up period.
  - ⇒ Three months of operating and maintenance labour costs, to include training.
  - ⇒ One month of chemicals, catalyst and waste disposal costs and maintenance materials costs.

- Owner's cost: This is assumed to be incurred in the first year of construction, allowing for the fact that some of the costs would be incurred before the start of construction. 7% of the TPC is assumed to cover the Owner's cost and fees.

- Interest during construction: This is calculated from the plant construction schedule and the interest rate is assumed to be the same as the discount rate.

- Working capital: The working capital includes inventories of fuel and chemicals (materials held in storage outside of the process plants). Storage for 30 days at full load is considered for chemicals and consumables. It is assumed that the cost of these materials are recovered at the end of the plant life.

The assumptions used to calculate of the fixed operating cost are as follows:

- Direct labour cost

  This study assumes fixed number personnel needed for the Base Case and 5 more additional personnel for all cases with CO_2 capture. The yearly cost of the direct labour is calculated assuming an average salary of 60,000 €/y.

- Admin and general overhead cost (indirect labour cost)

  This study assumes that the indirect labour cost is equal to 30% of the direct labour and maintenance labour cost. Generally, admin and general overhead cost is dependent on the company's organization structure and complexity of its operation. This normally covers functions which are not directly involved in the daily operation of the plant.

- Annual operating and maintenance cost

  A precise evaluation of the cost of maintenance would require a breakdown of the costs amongst the numerous components and packages of the plant. Since these costs are all strongly dependent on the type of equipment selected and their statistical maintenance data provided by the selected vendors. This kind of evaluation of the maintenance cost is premature for this type of study.

  For this reason the annual maintenance cost of the plant is estimated as a percentage of the TPC. 1.5% is assumed for all cases. This generally applies to all of the major processes, utilities and off-sites.

Additionally, estimates can be separately expressed as maintenance labour and maintenance materials. A maintenance labour to materials ratio of 40:60 can be statistically considered for this study.

- Insurance cost: 0.5% of the TPC is assumed to cover the insurance cost.

- Local taxes and fees: 0.5% of the TPC is also assumed to cover the local taxes and fees.

Key assumptions used in estimating the variable operating cost are as follows:

- The assumed prices of the consumables and miscellaneous items are presented in Table 1.

Table 1: Assumed prices for consumables and other miscellaneous items

Item	Unit	Cost	Sensitivity Range
Natural gas	€/GJ (LHV)	6	4 to 16
Raw process water	€/m^3	0.2	-
Electricity	€/MWh	80	40 to 120
Ammonia	€/t NH3	340	-
CO_2 transport and storage	€/t CO_2 stored	10	0 to 20
CO_2 emission cost	€/t CO_2 emitted	0	0 to 100

- Chemical and Catalyst Cost

  The cost of chemicals is assumed fixed at an annual cost for all cases. This generally accounts for the cost of chemicals used in the treatment of demi-water, process water, boiler feed water, cooling water, and others.

  The catalyst cost which covers the catalyst used in the syngas production, and ammonia, urea and methanol synthesis. Fixed cost is assumed for all cases.

Definition of Levelised Cost and Cost of CO_2 Avoidance

a.) Levelized Cost of Production

The levelised cost of production defined in this report include:

- levelised cost of urea production (LCOU)
- levelised cost of methanol production (LCOMeOH)

The Levelized Cost of Production is used to calculate the selling price of main product which enables the present value from all sales of the product(s) over the economic lifetime of the plant to equal the present value of all costs of building, maintaining and operating the plant over its lifetime. In other word, the selling price of the product is calculated based on the assumption that NPV = 0 (over the whole life time of the plant).

The method of calculation is based on a discounted cash flow analysis. This is similar to how the Levelized Cost of Electricity (LCOE) are calculated in other IEAGHG studies, except that it is necessary to take into account the revenues from the sale of electricity as co-product.

The Levelised Cost of Production in this study is calculated assuming constant prices (in real terms) for fuel and other costs; and constant operating capacity factors throughout the plant lifetime, apart from lower capacity factors in the first year of operation.

b.) Cost of CO_2 avoidance

Costs of CO_2 avoidance were calculated by comparing the CO_2 emissions per tonne product and the Levelised Cost of Production of the plants with capture and a reference plant without capture.

$$\text{CO}_2 \text{ Avoidance Cost (CAC)} = \frac{\text{LCO(Product)}_{\text{CCS}} - \text{LCO(Product)}_{\text{Reference}}}{\text{CO}_2\text{Emissions}_{\text{ Reference}} - \text{CO}_2\text{Emissions}_{\text{ CCS}}}$$

where:

- CAC is expressed in € per tonne of CO_2
- LCO(Product) is expressed in Euro per tonne of product
- CO_2 emission is expressed in tonnes of CO_2 per tonne of product

Key Findings of the Study

Plant Performance

a.) Ammonia/Urea Production

Table 2 presents the summary of the energy balance of the ammonia/urea plant without and with CCS.

For both cases the total natural gas consumption are fixed. For the base case, only 95% of the ammonia is converted to urea (producing 2260 t/d granulated urea). For the case with CCS, all the ammonia is converted to urea (producing 2380 t/d granulated urea).

As it could be noted that the main difference in the plant performance could be manifested in the amount of electricity imported by the plant. This has increased by 8.6 MWe for the ammonia plant with CCS (i.e. +97% as compared to the Base Case). This is mainly due to the reduced amount of steam available to the ammonia plant to drive various machineries (compressors, pumps, etc…) as most of the excess steam are diverted to the CO_2 capture plant (+4.9 MWe) and the additional electricity demand of the CO_2 capture and compression plant (+2.6 MWe).

Table 3 presents the carbon balance of the plant without and with CCS.

Table 2: Energy Balance of the Ammonia/Urea Plant

Performance Data			
		Base case	CO_2 capture case
INLET STREAMS			
Natural Gas Feedstock	t/h	27.021	27.021
Natural Gas Fuel	t/h	10.630	10.630
Natural Gas LHV	MJ/kg	46.50	46.50
Total Energy Input	**MW**	**486.3**	**486.3**
OUTLET STREAMS			
Urea Product to BL	TPD	2,260	2,380
	t/h	94.17	99.17
Urea energy content	MJ/kg	32.7	32.7
Total Energy in product	**MW**	**855.4**	**900.8**
Ammonia Product to BL	t/d	68.4	NA
	t/h	2.85	NA
Ammonia LHV	MJ/kg	18.60	NA
Total Energy in the Ammonia Product	**MW**	**14.73**	**NA**
POWER BALANCE			
Ammonia/Urea Plant Power Consumption	MWe	4.490	9.415
Steam and BFW Consumption	MWe	1.100	1.100
Utilities + BoP Consumption	MWe	3.300	4.400
CO2 Capture Plant Consumption	MWe	-	0.680
CO2 Compressor Consumption	MWe	-	1.880
Power import from the grid	MWe	8.890	17.475
SPECIFIC CONSUMPTIONS			
Natural Gas Feedstock	GJ/t urea	13.343	12.670
Natural Gas Fuel	GJ/t urea	5.249	4.984
Feed + Fuel	**GJ/t urea**	**18.592**	**17.654**
SPECIFIC EMISSIONS			
Specific CO2 emission	t/t urea	0.325	0.033
Equivalent CO2 in Urea product		69.3%	73.0%
Captured CO2 to storage		NA	23.8%
Overall CO2 not emitted		69.3%	96.7%

Table 3: Overall Carbon Balance of the Plant

CO_2 removal efficiency	**Equivalent flow of CO_2**	
	kmol/h	
	Ammonia/Urea plant w/o capture	**Ammonia/Urea with capture from SMR flue gas**
Natural gas feedstock	1626	1626
Natural gas fuel	637	637
TOTAL IN	**2263**	**2263**
OUTPUT		
Urea carbon content	1568	1651
CO2 to storage	-	538
Total not emitted (C)	**1568**	**2189**
Flue gas to stack	695	74
Vents	-	-
Emission	**695**	**74**
TOTAL OUT	**2263**	**2263**
Equivalent CO_2 in Urea product, %	**69.3%**	**73.0%**
Captured CO_2 to storage, %	**-**	**23.8%**
AMOUNT OF CARBON NOT EMITTED	**69.3%**	**96.7%**

b.) Methanol Production

Table 4 presents the summary of the energy balance of the methanol plant without and with CCS.

For both cases (i.e. methanol plant without and with capture) the total natural gas consumption and the amount of the methanol produced are fixed.

As it could be noted that the main difference in the plant performance could be manifested in the amount of electricity imported by the plant. This has increased by 17.9 MWe for the methanol plant with CCS (i.e. +96% as compared to the Base Case). This is mainly due to the reduced amount of steam available to the methanol plant to drive various machineries (compressors, pumps, etc…) as most of the excess steam are diverted to the CO_2 capture plant (+11 MWe) and the electricity demand of the CO_2 capture and compression plant (+6.9 MWe).

Table 5 presents the carbon balance of the plant without and with CCS.

Table 4: Energy Balance of the Methanol Plant

Performance Data			
		Base case	CO_2 capture case
INLET STREAMS			
Natural Gas Feedstock	t/h	119.098	119.098
Natural Gas Fuel	t/h	17.119	17.119
Natural Gas LHV	MJ/kg	46.50	46.50
Total Energy Input	**MW**	**1,760**	**1,760**
OUTLET STREAMS			
Methanol Product to BL	TPD	5,000	5,000
	t/h	208.36	208.36
Methanol LHV	MJ/kg	20.094	20.094
Total Energy in product	**MW**	**1,163**	**1,163**
POWER BALANCE			
Methanol Plant Power Consumption	MWe	11.150	20.295
Steam and BFW Consumption	MWe	2.920	2.920
Utilities + BoP Consumption	MWe	4.400	6.250
CO2 Capture Plant Consumption	MWe	-	1.655
CO2 Compressor Consumption	MWe	-	5.200
Power import from the grid	MWe	18.470	36.320
SPECIFIC CONSUMPTIONS			
Natural Gas Feedstock	GJ/t MeOH	26.579	26.579
Natural Gas Fuel	GJ/t MeOH	3.820	3.820
Feed + Fuel	**GJ/t MeOH**	**30.399**	**30.399**
SPECIFIC EMISSIONS			
Specific CO_2 emission	t/t MeOH	0.3533	0.0353
Equivalent CO_2 in MeOH product		79.3%	79.3%
Captured CO_2 to storage		NA	18.4%
Overall CO_2 not emitted		79.3%	97.7%

Table 5: Overall Carbon Balance of the Plant

CO_2 removal efficiency	**Equivalent flow of CO_2**	
	kmol/h	
	Methanol plant w/o capture	**Methanol plant w/ capture from SMR flue gas**
Natural gas feedstock	7168	7168
Natural gas fuel	1030	1030
TOTAL IN	**8199**	8199
Methanol carbon content	6503	6503
CO_2 to storage	-	1505
Total not emitted	**6503**	**8008**
Flue gas to stack	1673	167
Vents	23	23
Emission	**1696**	**190**
TOTAL OUT	**8199**	**8199**
Equivalent CO_2 in MeOH product, %	**79.3%**	**79.3%**
Captured CO_2 to storage, %		**18.4%**
AMOUNT OF CARBON NOT EMITTED	**79.3%**	**97.7%**

Economic Evaluation

a.) Ammonia/Urea Production

Tables 6 and 7 present the summary of the total capital cost and annual operating cost of the ammonia/urea plant without and with CCS.

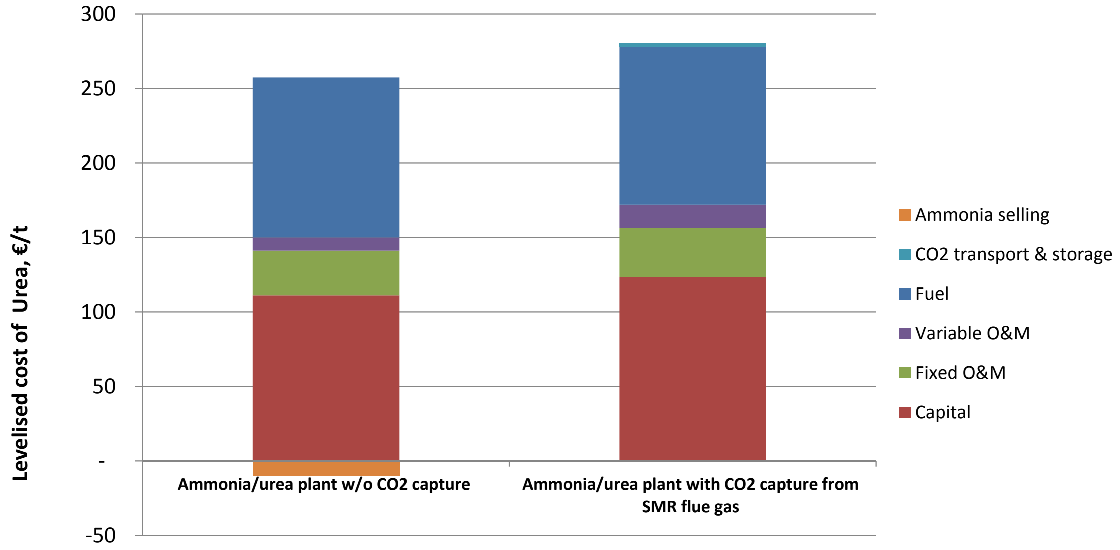
Figure 5 presents the levelised cost of urea production. Table 8 summarised the results for the levelised cost of urea production (LCOU) and CO_2 avoided cost (CAC). The CAC is calculated based on fixed electricity price of €80/MWh and indirect CO_2 emissions of the electricity production based on gas and coal (at 345 kg/MWh and 746 kg/MWh respectively).

Table 6: TPC and TCR of the ammonia/urea plant

		Ammonia/urea plant w/o CO_2 capture	**Ammonia/urea plant with CO_2 capture**
Ammonia/urea plant	M€	493	500
Utilities and BoP	M€	53.6	57.4
CO_2 capture unit	M€	-	46.5
CO_2 compression unit	M€	-	10.0
Total Installed Cost (TIC)	**M€**	**546.7**	**613.9**
Contingency	-	20%	20%
Total Plant Cost (TPC)	**M€**	**656.7**	**733.9**
Total Capital Requirement (TCR)	**M€**	**849.2**	**954.4**

Table 7: Annual Operating Cost of Ammonia/Urea Plant

ANNUAL O&M COST		
	Ammonia/Urea Plant w/o CCS €/year	Ammonia/Urea Plant w/ CCS €/year
Fixed Costs		
Direct labour	3,960,000	4,260,000
Adm./gen. overheads	2,368,656	2,604,024
Insurance & local taxes	6,559,200	7,366,800
Maintenance	9,838,800	11,050,200
Sub-total	22,726,656	25,281,024
Variable Costs (Availability - 90%)		
Feedstock & fuel	82,821,922	82,821,922
Electricity (Import from Grid)	5,607,120	11,021,840
Raw water (make-up)	34,847	6,465
Chemicals & catalysts	1,200,000	1,200,000
Sub-total	89,663,889	95,050,227
Total Fixed & Variable Cost	**112,390,545**	**120,331,251**
Other Revenues		
Ammonia	-7,640,000	-
Other Cost		
CO2 Transport & Storage	-	1,866,300
Annual O&M Cost	**104,750,545**	**122,197,551**



Natural gas: 6 €/GJ (LHV); Discount rate: 8%

CO_2 transport & storage: 10 €/t; 90% capacity factor; Constant €, 2014.

Figure 5: Levelized cost of Urea (LCOU)

Table 8. LCOU and CAC of the ammonia/urea plant study cases

Description	Levelized Cost of Urea	CO_2 emission avoidance cost
	€/t	€/t
Ammonia/urea plant w/o CO_2 capture	257.3	-
Ammonia/urea plant with CO_2 capture from SMR flue gas (Indirect emission from NG plant w/o CCS)	280.3	87.4
Ammonia/urea plant with CO_2 capture from SMR flue gas (Indirect emission from coal plant w/o CCS)	280.3	99.8

b.) Methanol Production

Tables 9 and 10 present the summary of the total capital cost and annual operating cost of the plant without and with CCS.

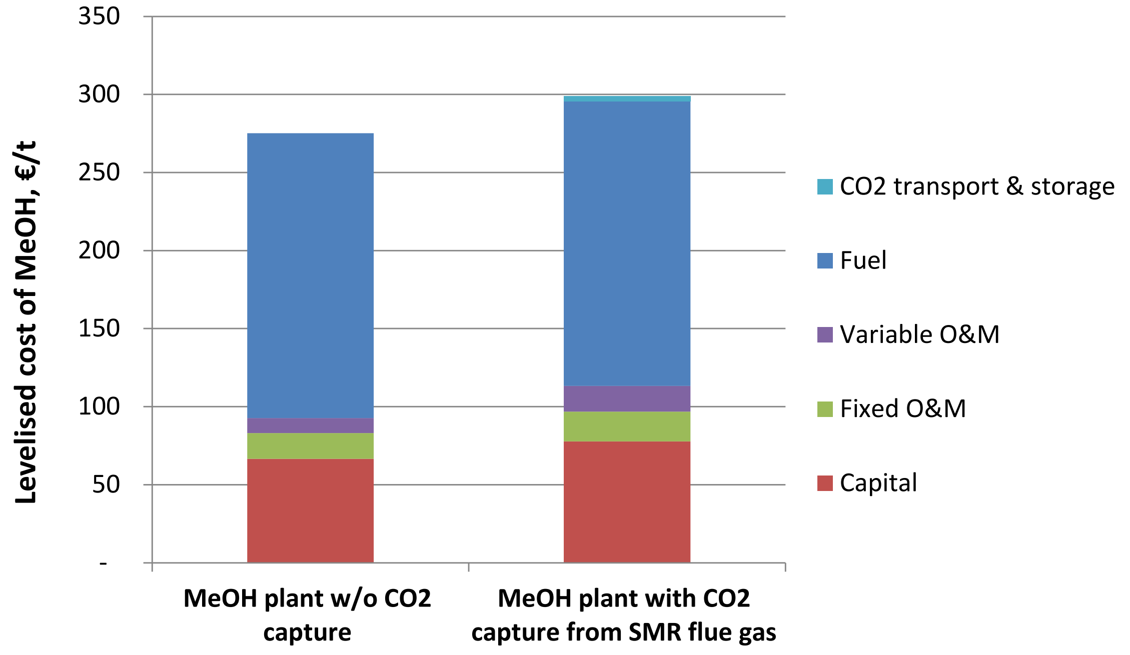
Figure 6 presents the levelised cost of methanol production. Table 11 summarised the results for the levelised cost of methanol production (LCOMeOH) and CO_2 avoided cost (CAC). The CAC is calculated based on fixed electricity price of €80/MWh and indirect CO_2 emissions of the electricity production based on gas and coal (at 345 kg/MWh and 746 kg/MWh respectively).

Table 9: TPC and TCR of the methanol plant

		Methanol plant w/o CO_2 capture	**Methanol plant with CO_2 capture**
Methanol plant (including ASU and steam/BFW system)	M€	600	600
Methanol storage	M€	8.7	8.7
Utilities and BoP	M€	85.0	92.8
CO_2 capture unit	M€	-	88.1
CO_2 compression unit	M€	-	20.5
Total Installed Cost (TIC)	**M€**	**693.7**	**810.1**
Contingency	-	20%	20%
Total Plant Cost (TPC)	**M€**	**832.4**	**972.1**
Total Capital Requirement (TCR)	**M€**	**1082.3**	**1264.6**

Table 10: Annual operating cost of methanol plant

ANNUAL O&M COST		
	Methanol Plant w/o CCS €/year	Methanol Plant w/ CCS €/year
Fixed Costs		
Direct labour	3,360,000	3,660,000
Adm./gen. overheads	2,506,392	2,847,816
Insurance & local taxes	8,324,400	9,721,200
Maintenance	12,486,600	14,581,800
Sub-total	26,677,392	30,810,816
Variable Costs (Availability - 90%)		
Feedstock & fuel	299,638,487	299,638,487
Electricity (Import from Grid)	11,649,360	22,907,760
Raw water (make-up)	89,720	54,715
Chemicals & catalysts	4,000,000	4,000,000
Sub-total	315,377,567	326,600,962
Total Fixed & Variable Cost	**342,054,959**	**357,411,778**
Other Cost		
CO2 Transport & Storage	-	5,222,519
Annual O&M Cost	**342,054,959**	**362,634,297**



Natural gas: 6 €/GJ (LHV); Discount rate: 8%
CO_2 transport & storage: 10 €/t; 90% capacity factor; Constant €, 2014.

Figure 6. Levelized cost of Methanol (LCOMeOH)

Table 11. LCOMeOH and CAC of the methanol plant study cases

Description	Levelised Cost of Methanol €/t	CO_2 emission avoidance cost €/t
MeOH plant w/o CO_2 capture	275.1	-
MeOH plant with CO_2 capture from SMR flue gas (Indirect emission from NG plant w/o CCS)	298.9	82.6
MeOH plant with CO_2 capture from SMR flue gas (Indirect emission from coal plant w/o CCS)	298.9	93.7

Conclusions

- The study presented a detailed baseline information of the performance and cost of deploying CO_2 capture in a SMR Based HyCO plant using natural gas as feedstock / fuel and operating as a captive plant (i.e. integrated with other industrial complex).

- An SMR could be integrated into an industrial complex such as an ammonia or methanol plant.
    - In general, the syngas generation used in ammonia plant is based on SMR in tandem with an air blown ATR. This is the typical conventional ammonia production configuration (i.e. no need of air separation unit).
    - Normally, with large methanol plants as presented in this study, the SMR is always in tandem with an oxygen blown ATR.

- The addition of capture of CO_2 from the flue gas of SMR increases the energy demand of the plant.
    - For the ammonia/urea production – an additional 8.6MWe of electricity is imported from the grid.
    - For the methanol production – an additional 17.9 MWe of electricity is imported from the grid.

- In general, the addition of CCS increases the levelised cost of production.
    - For the ammonia/urea production – this increases by 23 €/t urea.
    - For the methanol production – this increases by 24 €/t methanol

- The CO_2 avoided cost of capturing additional CO_2 from the SMR plant is in the range of €80 to 100 per tonne CO_2 for both cases.

- It is well established that CO_2 is used as reactant to both urea and methanol production. This is considered as mature technology. This study should provide a good basis for understanding the performance and cost of implementing both industrial CCS and CCU.

Recommendations

For future studies, it is highly recommended to evaluate the following case scenarios:

- Evaluate cases for methanol plant (mainly in revamp scenario) to determine the performance and cost of industrial CCS together with CCU.

- Evaluate cases for ammonia plant (mainly in fertilizer production complex) where the capture of CO_2 is not entirely used in the urea production (i.e. a scenario with multi-product configuration).

As an example to how CO_2 could be used in the production, Figure 7 illustrates how a smaller methanol plant could increase its production capacity by up to 20% using CO_2 injection. Preliminary assessment indicated that 50% of the CO_2 in the flue gas of the SMR could be used as feedstock to the methanol plant.

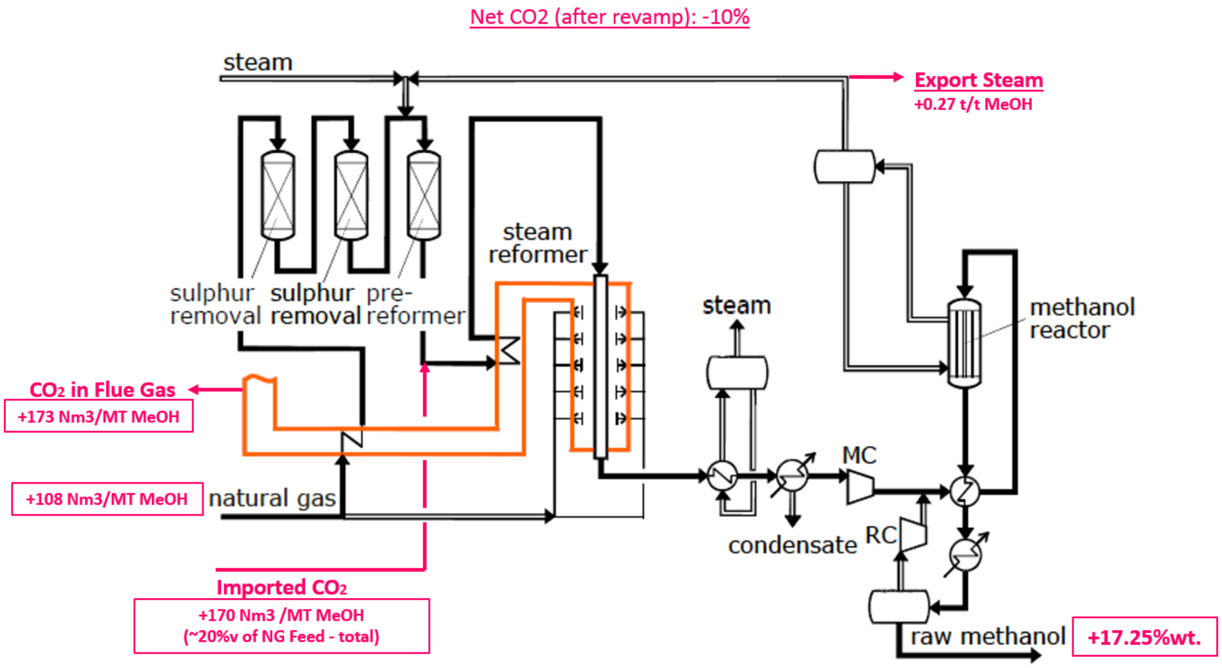


Figure 7: Impact of boosting methanol production by addition of CO_2 in a one-step reforming methanol plant (revamp option)

IEAGHG	Revision No.:	FINAL
Techno-Economic Evaluation of H_2 or HYCO Plant Integrated to an Industrial Complex (Task 4)	Date:	February 2017
		Sheet: 1 of 80

CLIENT : IEA Greenhouse Gas R&D Programme (IEAGHG)
PROJECT NAME : Techno-Economic Evaluation of H_2 Production with CO_2 Capture
DOCUMENT NAME : Techno-Economic Evaluation of H_2 or HYCO Plant Integrated to an Industrial Complex
FWI Contract : 1BD0840A

ISSUED BY : G. AZZARO / N. FERRARI
CHECKED BY : G. COLLODI
APPROVED BY : G. COLLODI

DATE	REVISED PAGES	ISSUED BY	CHECKED BY	APPROVED BY

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List of Abbreviations

AACE	Association for Advancement of Cost Engineering
ACTL	Alberta Carbon Trunk Line
ASTM	American Society for Testing and Materials
ASU	air separation unit
ATR	autothermal reformer
B.L.	battery limit
BFD	block flow diagram
BFW	boiler feed water
BHP	boiler horsepower
BoP	balance of plant
CAC	CO_2 avoidance cost
CCS	CO_2 capture and storage
CCU	CO_2 capture and utilisation
CWR	cooling water return
CWS	cooling water system
DRI	direct iron reduction
EBITDA	earnings before interest, taxes, depreciation and amortisation
H&MB	heat and mass balance
HC	hydrocarbon
HRU	hydrogen recovery unit
HTS	high temperature shift
HYCO	hydrogen and carbon monoxide (gas mixture)
LCOE	levelised cost of electricity
LCOH	levelised cost of hydrogen
LCOMeOH	levelised cost of methanol
LCOU	levelised cost of urea
LHV	low heating value
LTS	low temperature shift
MAC	main air compressor
MTS	medium temperature shift
MUG	make-up gas
NG	natural gas
NGCC	natural gas combined cycle
POX	partial oxidation
PSA	pressure swing adsorption
SMR	steam methane reformer
TCR	total capital requirement
TIC	total installed cost
TPC	total plant cost
USC-PC	ultra-supercritical pulverised coal fired boiler
VSA	vacuum swing adsorption
WHB	waste heat boiler
WWT	waste water treatment plant

1. Introduction

The objective of this document is to define and evaluate the techno-economics of deploying CO_2 capture in a H_2 or HYCO Plant integrated to the following industrial complex:

- Case 4A: Ammonia with Urea Production

  For the base case (Case 4A-1), the syngas generation unit or HYCO plant is integrated to a nominal 1350 t/d ammonia plant using natural gas as feedstock and fuel. Around 95% of the ammonia produced is further converted in a downstream urea plant producing 2260 t/d urea using the CO_2 captured from the syngas (as part of the process).

  For the capture case (Case 4A-2), the same ammonia and urea complex as described above would have additional CO_2 to be captured from the flue gas of the SMR. The plant would maximise the urea production by producing 2380 t/d urea.

  For both cases, the plant battery limit would include all the utilities required for the plant operation.

  These cases are described in Section 2.

- Case 4B: Methanol Production

  For the base case (Case 4B-1), the syngas generation unit or HYCO plant is integrated into a nominal 5000 t/d methanol plant using natural gas as feedstock and fuel.

  For the capture case (Case 4B-2), the same methanol plant as described above would capture the CO_2 from the flue gas of the SMR.

  For both cases, the plant battery limit would include all the utilities required for the plant operation.

  These cases are described in Section 3.

2. Case 4A: SMR Plant for Ammonia - Urea Production

2.1. Basis of Design

The general plant design data and assumptions used in these cases should be referred to Annex I - Reference Document (Task 2). Whilst, the specific information relevant to the ammonia-urea production are reported in this section.

2.1.1. *Capacity*

The ammonia plant is designed to produce 1,350 t/d of liquid ammonia.

For the plant without CCS (Case 4A-1), the urea plant is designed to produce of 2,260 t/d of granulated urea, corresponding to about 95% ammonia conversion.

For the plant with CCS (Case 4A-2) where additional CO_2 is captured from the flue gas of the SMR, all the ammonia are converted into granulated urea; thus increasing the capacity to 2,380 t/d.

2.1.2. *Product Specifications*

Typical commercial specifications of ammonia and granulated urea are listed in the following table.

Ammonia		Value	Unit
Purity	min	99.5	wt. %
Water content	max	0.5	wt. %
Oil	max	5	ppm wt.

Urea (Granulated)		Value	Unit
Total Nitrogen	min	46	wt. %
Free Ammonia	max	100	ppm wt.
Biuret	max	0.9	wt. %
Moisture	max	0.3	wt. %
Size distribution	between 1&4 mm in diameter	95	wt. %

2.1.3. *Syngas Specifications*

Below summarised the list of requirements regarding the specifications of the syngas or HYCO gas used by the ammonia synthesis plant.

- Process Feed Gas to the Primary Reformer (%v dry basis)

Sulphur	< 0.05 ppm	poison to the catalyst
Chloride	< 0.05 ppm	poison to the catalyst
As, V, Pb, Hg	< 5 ppb	poison to the catalyst
Olefin	< 1-2%	carbon formation

- Syngas to the Low Temperature Shift reactor (%v dry basis)

Sulphur	< 0.1 ppm	poison to the catalyst
Chloride	< 5 ppb	poison to the catalyst

- Syngas to the Methanator

K_2CO_3	could cause pore blockage to the methanator catalyst (1)
As (as As_2O_3)	< 5 ppb, poison to the methanator catalyst (2)
Sulpholane	could cause sulphur poisoning to the methanator catalyst (3)

- Syngas to Ammonia Synthesis (%v dry basis)

CO_2 + CO	< 5 ppm	poison to the NH_3 synthesis catalyst

2.1.4. *Plant Battery Limits*

Main plant battery limits are listed below:

- Natural gas (in)
- Waste water streams (out)
- Raw water (in)
- Sea water (in/out)
- Electric power (in)
- Compressed CO_2 rich stream (out) applicable to the CO_2 capture case only
- Ammonia (out) - applicable to the base case only
- Urea (out)

---

[1] This is an important considerations if Benfield, Vetrocoke or Catacarb process are used in the bulk CO_2 removal unit.

[2] This is an important considerations if Vetrocoke process is used in the bulk CO_2 removal unit.

[3] This is an important considerations if Sulfinol process is used in the bulk CO_2 removal unit. Sulpholane could decompose and sulphur could act as poison to the methanator catalyst.

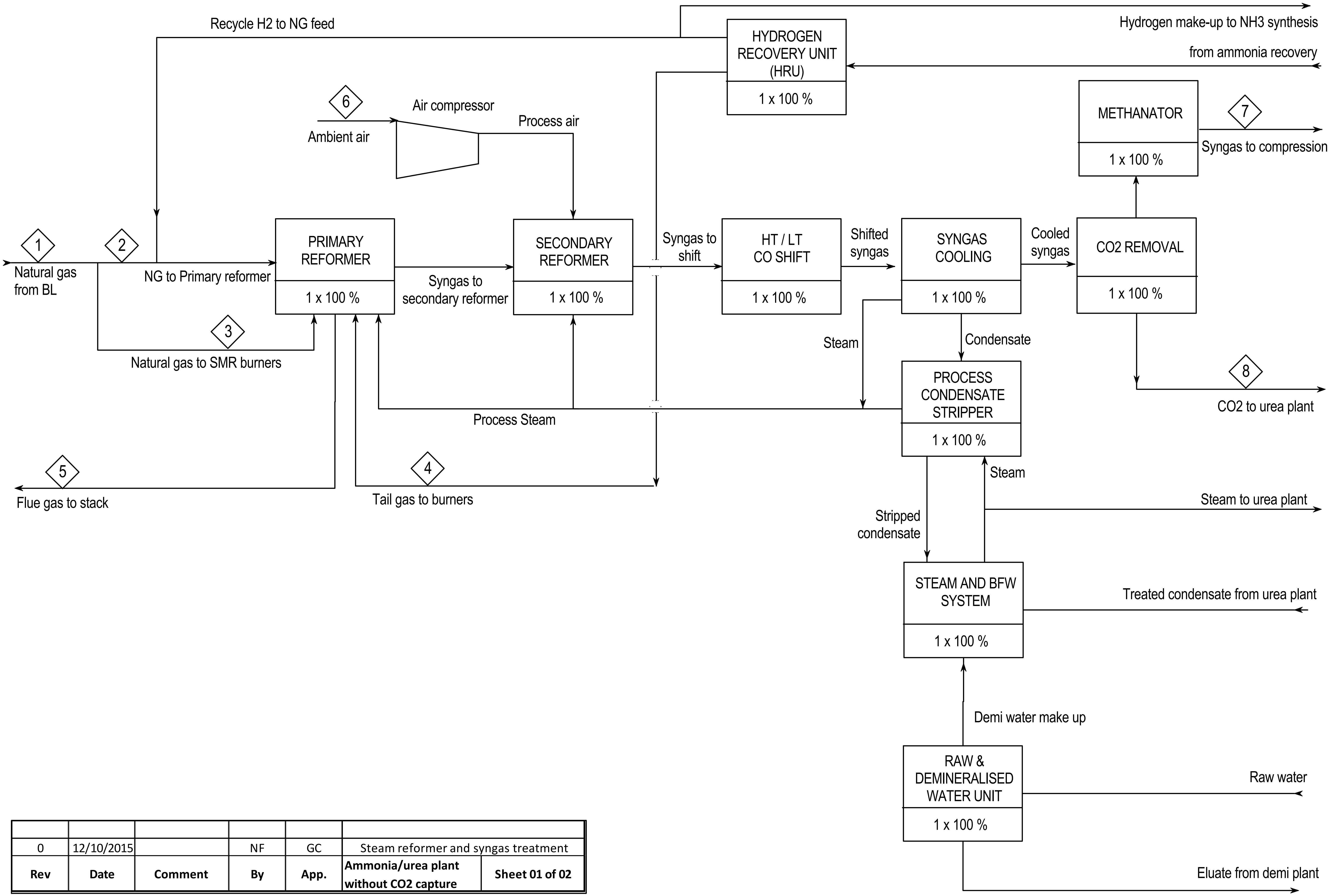
2.2. Units Arrangement

Ammonia-Urea Plant (Case 4A-1 and Case 4A-2) consists of the following units:

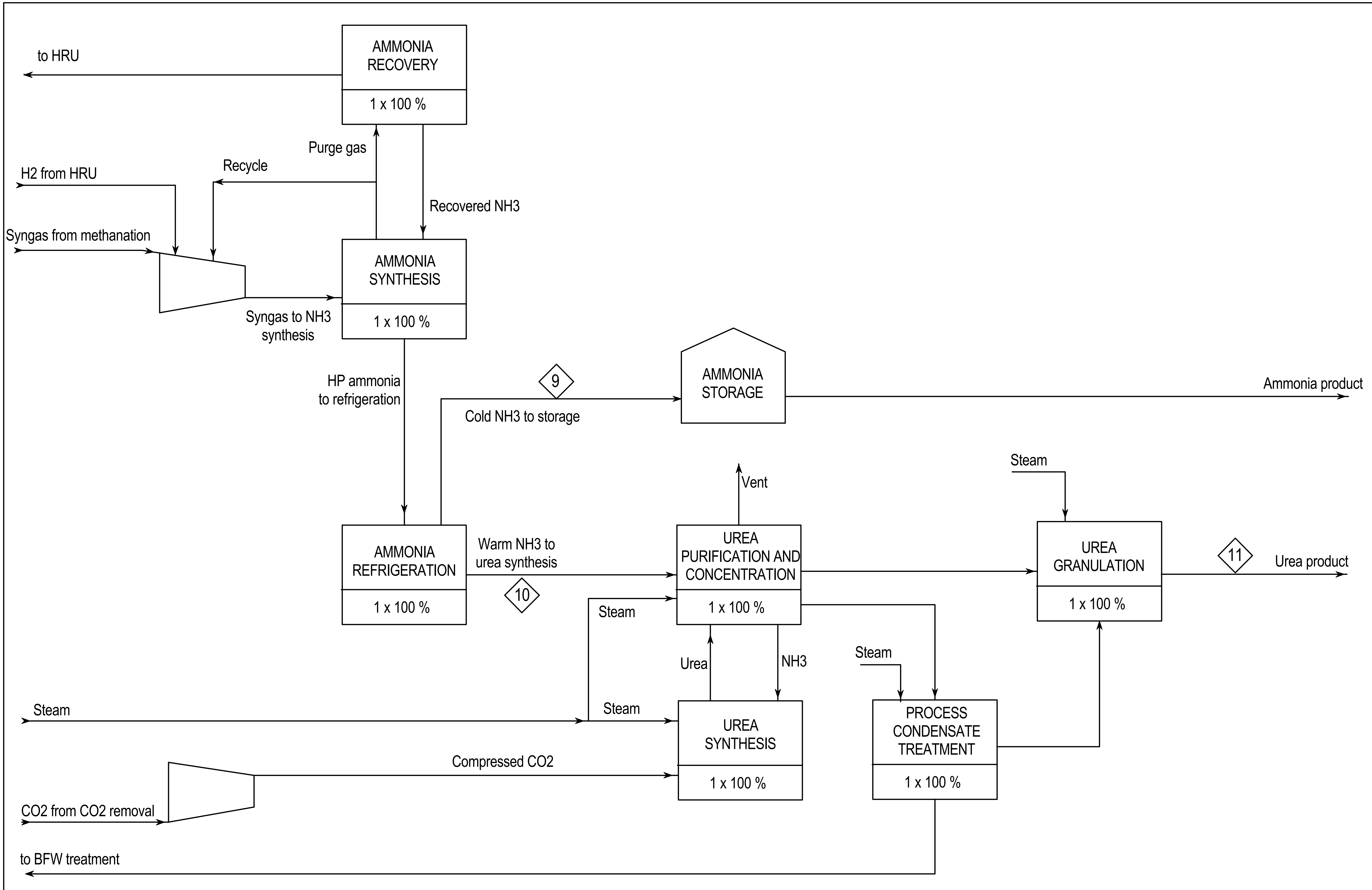
- Ammonia plant include:
  - Primary Reformer (Feedstock Pre-treatment and SMR)
  - Secondary Reformer (ATR)
  - High and Low Temperature CO Shift Conversion Section
  - Bulk CO_2 Removal Section
  - Methanation
  - Ammonia Synthesis and Refrigeration
  - Ammonia Storage
- Urea plant include:
  - Urea Synthesis
  - Urea Purification
  - Granulation
  - Waste Water Treatment
- Steam and BFW plant
- Demi-Water Plant
- Utilities and Balance of Plant (BoP), consisting of:
  - Cooling Water System
  - Flare System
  - Interconnecting
  - Drain System
  - Buildings (Control Room, Laboratories, Electrical Sub-station).

2.3. Overall Block Flow Diagram

The BFDs presented in the succeeding pages shows the different unit processes included in the Ammonia and Urea Production Complex (without and with CCS).



0	12/10/2015		NF	GC	Steam reformer and syngas treatment	
Rev	**Date**	**Comment**	**By**	**App.**	**Ammonia/urea plant without CO2 capture**	**Sheet 01 of 02**



0	25/06/2015		NF	GC	Ammonia and urea plant	
Rev	**Date**	**Comment**	**By**	**App.**	**Ammonia/urea plant without CO2 capture**	**Sheet 02 of 02**

HEAT AND MATERIAL BALANCE
Ammonia-Urea plant without CO2 capture

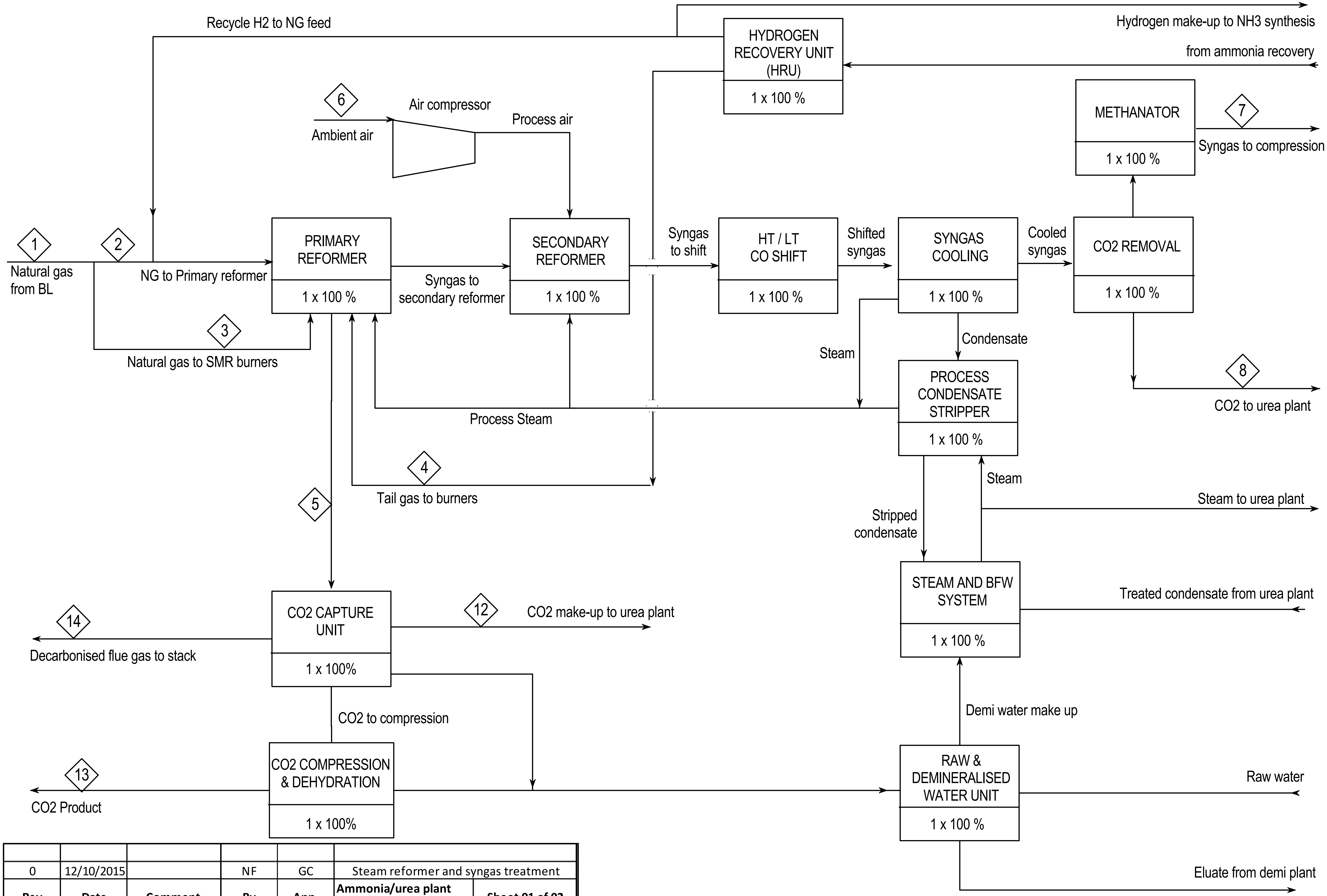
		REV	**DATE**	**BY**	**CHKD**	**APP**
CLIENT:	IEAGHG					
PROJECT NAME:	TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE FOR INDUSTRY	0	October 2015	NF	GC	GC
FWI CONTRACT:	1BD0840A					
LOCATION:	THE NETHERLAND					

Stream		**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Description		Natural Gas From B.L.	Natural Gas feedstock to Ammonia Plant	Natural Gas fuel to burners	Tail gas to burners	Flue gas to stack	Process Air	Syngas to compression	CO2 to Urea Plant	Ammonia to storage	Ammonia to Urea Plant	Urea product	
Temperature	℃	9	15	15	50	155	Ambient	43	43	-33	10	Ambient	
Pressure	*MPa*	7,00	4,20	0,15	0,50	atm	atm	3,00	0,15	atm	2,10	atm	
Molar Flow	*kmol/h*	2090	1500	590	282	7969	2286	7007	1668	167,0	3136	1568	
Mass Flow	*kg/h*	37651	27021	10630	5583	221304	66079	62853	70593	2850	53400	94170	
Composition													
CO2	*mol/mol*	0,0200	0,0200	0,0200	0,0000	0,0870	0,0000	0,0000	0,9400	0,0000	0,0000	0,0000	
CO	*mol/mol*	0,0000	0,0000	0,0000	0,0000	(2)	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	
Hydrogen	*mol/mol*	0,0000	0,0000	0,0000	0,2690	0,0000	0,0000	0,7300	0,0080	0,0000	0,0000	0,0000	
Nitrogen	*mol/mol*	0,0089	0,0089	0,0089	0,4770	0,7070	0,7800	0,2580	0,0010	0,0000	0,0000	0,0000	
Ar	*mol/mol*	0,0000	0,0000	0,0000	0,0760	0,0120	0,0100	0,0030	0,0000	0,0000	0,0000	0,0000	
Oxygen	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0140	0,2100	0,0000	0,0000	0,0000	0,0000	0,0000	
Methane	*mol/mol*	0,8900	0,8900	0,8900	0,1780	0,0000	0,0000	0,0070	0,0000	0,0000	0,0000	0,0000	
Ethane	*mol/mol*	0,0700	0,0700	0,0700	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	
Propane	*mol/mol*	0,0100	0,0100	0,0100	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	
n-Butane	*mol/mol*	0,0010	0,0010	0,0010	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	
n-Pentane	*mol/mol*	0,0001	0,0001	0,0001	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	
H2O	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,1800	0,0000	0,0020	0,0510	0,0000	0,0000	0,0000	
NH3	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	1,0000	1,0000	0,0000	
Urea	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	1,0000	
Contaminants:													
H2S	*ppm v*	(1)											
NOx	*mg/Nm3*												

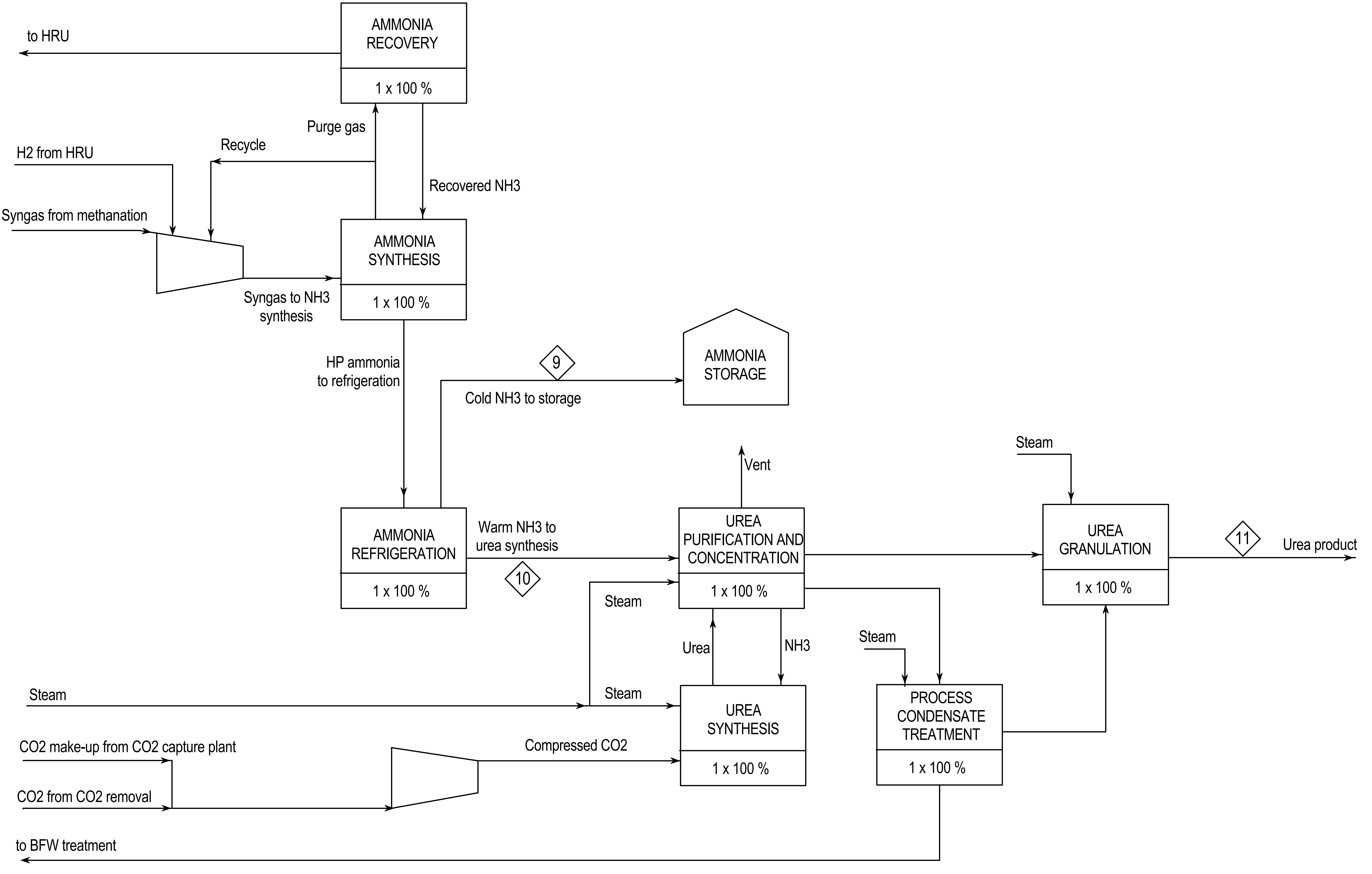
Notes:

(1) For feedstock purification section design purposes 5 ppmv of H2S have been assumed in NG to plant

(2) 30 mg/Nm3 max



0	12/10/2015		NF	GC	Steam reformer and syngas treatment	
Rev	**Date**	**Comment**	**By**	**App.**	**Ammonia/urea plant with CO2 capture**	**Sheet 01 of 02**



0	25/06/2015		NF	GC	Ammonia and urea plant	
Rev	**Date**	**Comment**	**By**	**App.**	**Ammonia/urea plant with CO2 capture from SMR flue gas**	**Sheet 02 of 02**

FOSTER WHEELER

HEAT AND MATERIAL BALANCE
Ammonia-Urea plant with CO2 capture

CLIENT:	IEAGHG	**REV**	**DATE**	**BY**	**CHKD**	**APP**
PROJECT NAME:	TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE FOR INDUSTRY	0	October 2015	NF	GC	GC
FWI CONTRACT:	1BD0840A					
LOCATION:	THE NETHERLAND					

Stream		1	2	3	4	5	6	7	8	9	10	11	12	13	14
Description		Natural Gas From B.L.	Natural Gas feedstock to Ammonia Plant	Natural Gas fuel to burners	Tail gas to burners	Flue gas to capture unit	Process Air	Syngas to compression	CO2 to Urea Plant from CO2 removal	Ammonia to storage	Ammonia to Urea Plant	Urea product	CO2 to Urea Plant from flue gas	CO2 product	Decarbonized Flue gas to Stack
Temperature	℃	9	15	15	50	155	Ambient	43	43	-33	10	Ambient	43	24	43
Pressure	*MPa*	7,00	4,20	0,15	0,50	atm	atm	3,00	0,15	atm	2,10	atm	0,15	11,00	0,10
Molar Flow	*kmol/h*	2090	1500	590	282	7969	2286	7007	1668	0,0	3303	1651	83,2	538,0	6475,4
Mass Flow	*kg/h*	37651	27021	10630	5583	221304	66079	62853	70593	0	56227	99167	3662	23677	178493
Composition															
CO2	*mol/mol*	0,0200	0,0200	0,0200	0,0000	0,0870	0,0000	0,0000	0,9400	0,0000	0,0000	0,0000	0.9999+	0.9999+	0,0114
CO	*mol/mol*	0,0000	0,0000	0,0000	0,0000	(2)	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	(2)
Hydrogen	*mol/mol*	0,0000	0,0000	0,0000	0,2690	0,0000	0,0000	0,7300	0,0080	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Nitrogen	*mol/mol*	0,0089	0,0089	0,0089	0,4770	0,7070	0,7800	0,2580	0,0010	0,0000	0,0000	0,0000	0,0000	0,0000	0,8701
Ar	*mol/mol*	0,0000	0,0000	0,0000	0,0760	0,0120	0,0100	0,0030	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0148
Oxygen	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0140	0,2100	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0172
Methane	*mol/mol*	0,8900	0,8900	0,8900	0,1780	0,0000	0,0000	0,0070	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Ethane	*mol/mol*	0,0700	0,0700	0,0700	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Propane	*mol/mol*	0,0100	0,0100	0,0100	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
n-Butane	*mol/mol*	0,0010	0,0010	0,0010	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
n-Pentane	*mol/mol*	0,0001	0,0001	0,0001	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2O	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,1800	0,0000	0,0020	0,0510	0,0000	0,0000	0,0000	0,0000	0,0000	0,0865
NH3	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	1,0000	1,0000	0,0000	0,0000	0,0000	0,0000
Urea	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	1,0000	0,0000	0,0000	0,0000
Contaminants:															
H2S	*ppm v*	(1)													
NOx	*mg/Nm3*														

Notes:

(1) For feedstock purification section design purposes 5 ppmv of H2S have been assumed in NG to plant

(2) 30 mg/Nm3 max

2.4. Process Description

There are various technologies used in the production of ammonia and urea that are commercially available. The process scheme selected for this study is generic, with no reference to specific licensor and equipment suppliers.

This section should be referred to the Block Flow Diagram presented in Section 2.3.

2.4.1. Ammonia Plant

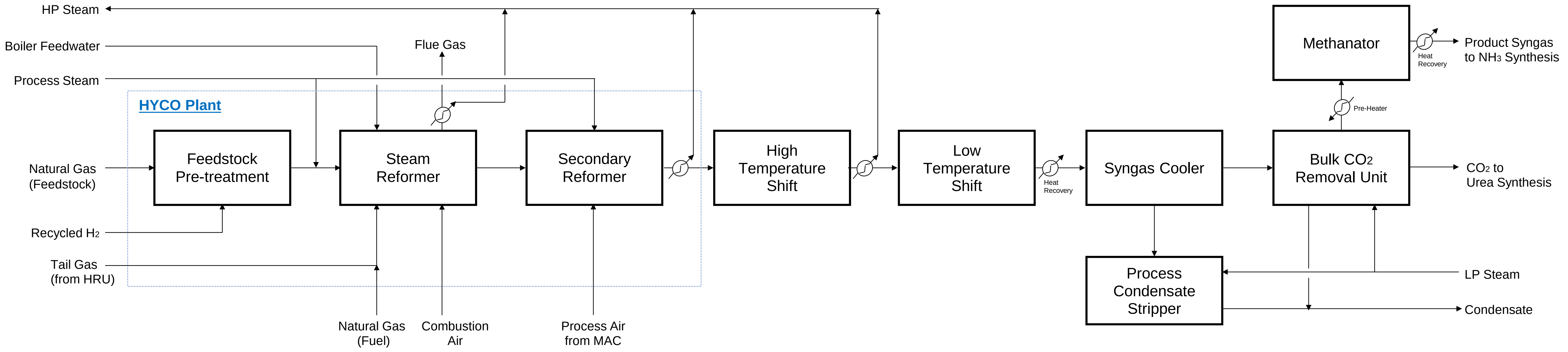
Ammonia is produced by conversion of the hydrogen (H_2) and nitrogen (N_2) in “synthesis gas” or “syngas” with a typical molar ratio of approximately 3:1.

Hydrogen is supplied by reforming the hydrocarbon feedstock; whilst the nitrogen is supplied by introducing the process air into the secondary reformer. For this study, natural gas is the main hydrocarbon feedstock.

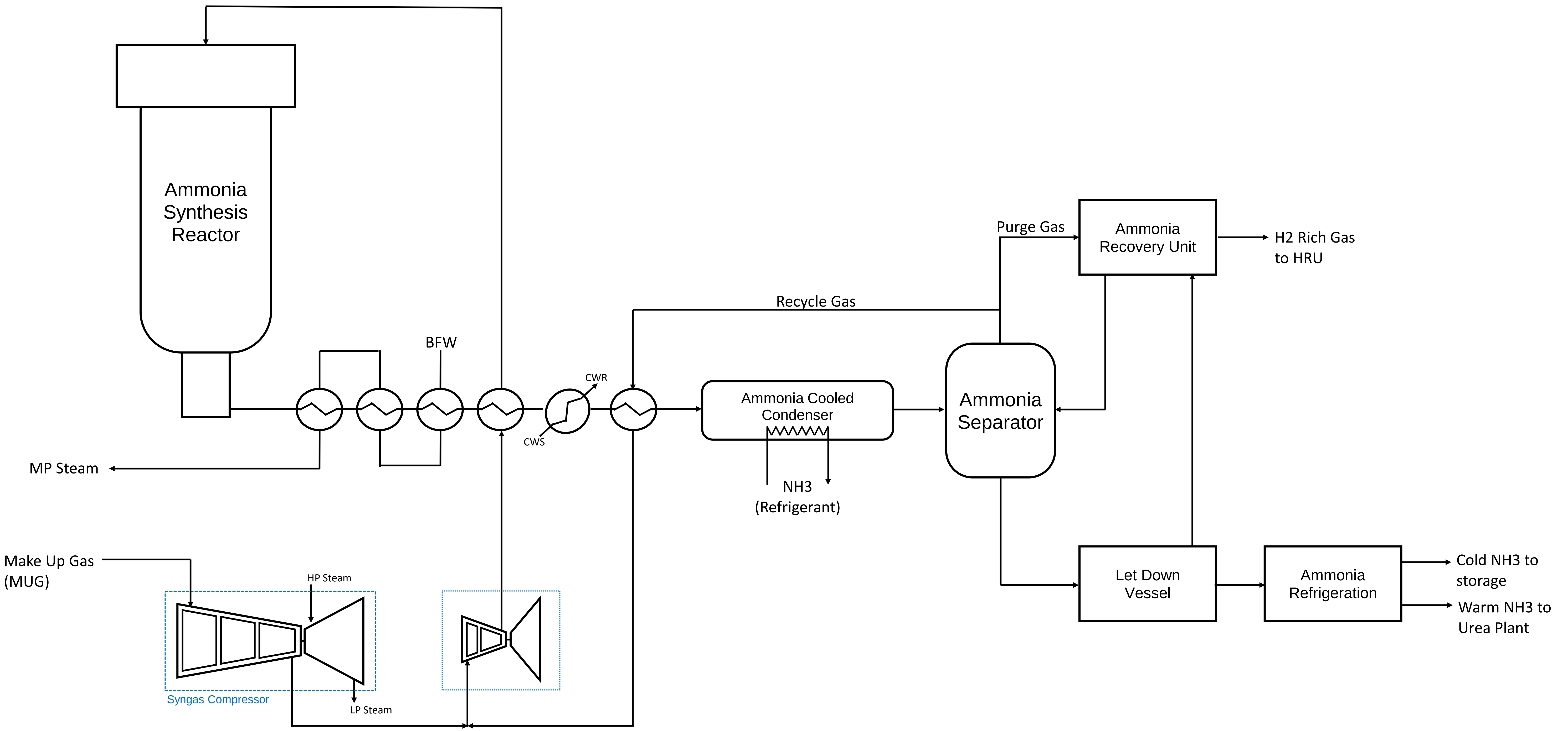
The steps to produce ammonia are as follows:

- The hydrocarbon feed is pre-treated by removing any sulphur and chloride.
- The treated hydrocarbon feed is reformed in a two-steps reformer (which consists of the SMR unit as the primary reformer and the air blown ATR as the secondary reformer) to produce the raw syngas which mainly consists of H_2, N_2, CO, CO_2, CH_4, Ar and steam.
- The raw syngas is further purified to remove any CO and CO_2 (as this poisons the ammonia synthesis catalyst). This involves (a.) the conversion of CO in the high and low temperature shift reactors; (b.) the bulk removal of CO_2 using chemical absorption; and (c.) the polishing step to remove any remaining CO and CO_2 in the methanation reactor
- The purified syngas is compressed and then sent to the ammonia synthesis loop where it is converted to liquid ammonia.

Figures shown in the succeeding pages present the generic block flow diagram of the syngas generation and the ammonia synthesis section.



SCHEMATIC BLOCK FLOW DIAGRAM	REVISION	DATE	BY	CHKD	APP.	SHEET
SYNGAS GENERATION & TREATMENT SECTION (AMMONIA / UREA PLANT)	v 1.0	Feb-17	SS			1 of 4



SCHEMATIC BLOCK FLOW DIAGRAM	REVISION	DATE	BY	CHKD	APP.	SHEET
Ammonia Synthesis Plant	v 1.0	Feb-17	SS			2 of 4

Desulphurization Section

The natural gas feedstock, which contains up to 5 ppm (v/v) of sulphur compounds, must be desulphurized, as the primary reformer catalyst and the low temperature CO shift catalysts are poisoned by any sulphur compounds.

The process description of the Desulphurisation Section is described in the Task 1 report.

The desulphurisation takes place in two stages – (1.) the hydrogenation of any organic sulphur and the saturation of any olefins; and (2.) removal of H_2S.

ZnO absorber is used to remove any H_2S in the feedstock; and should bring this down to less than 0.05 ppm (v/v) H_2S.

Reforming Section

The desulphurised feedstock is reformed in the primary reformer and secondary reformer to produce the H_2 needed for the ammonia synthesis.

The Primary Reformer is based on the principle of steam methane reforming (SMR). This is described in the reports of Task 1 and Task 3. Natural gas (consists mainly of methane) is reformed with steam in the bank of reformer tubes containing the catalyst. The syngas produced mainly consists of H_2, CO_2, CO, CH_4 and steam. The main heat required by the reaction is supplied indirectly by burning the tail gas and supplementary fuel in the SMR furnace.

As compared to the conventional SMR (normally used in H_2 production), the operation of the primary reformer is less intensive which allows higher methane slip (up to 10%v).

The Secondary Reformer is based on the principle of air blown auto-thermal reforming (ATR). The introduction of air in the secondary reformer provides the nitrogen required for the synthesis of ammonia.

In the ATR, the main heat is supplied by the combustion (in sub-stoichiometric condition) of the gas mixture obtained from the primary reformer with air. The combustion occurs in the upper section (combustion chamber or mixing volume) of the ATR. The lower section of the ATR consists of the catalyst bed which reforms the partially combusted gas mixture to produce the desired amount of hydrogen needed for the ammonia synthesis. Additionally, due to higher operating temperature, methane slip is reduce to less than 0.8%v.

Since the molar ratio (H_2/N_2) are fixed to maintain a value as close to 3, the amount of process air introduced into the ATR is therefore fixed. To control the methane slip from the ATR, the firing of the primary reformer (i.e. temperature of the SMR) is adjusted accordingly.

The gas leaving the ATR is around 1000°C. Cooling of the syngas is carried out in the Waste Heat Boiler, to produce high pressure (HP) steam.

Shift Section

The syngas leaving the reforming section contains substantial amount of CO. This is converted to CO_2 and H_2 via shift reaction:

$$CO + H_2O \Leftrightarrow CO_2 + H_2 + heat$$

The shift reaction takes place in two different adiabatic reactors in series. The first reactor is the high temperature shift (HTS) reactor using iron based catalyst promoted by chromium oxides or copper oxides. This operates at around 350-360°C. The residual CO of around 2-3% (dry basis) is expected.

The gas leaving from the first reactor is then cooled to around 180-200°C before being introduced into the low temperature shift reactor (LTS) using copper catalyst. The residual CO of around 0.1-0.3% (dry basis) should be expected. The gas leaving the low temperature shift reactor is then cooled to around 50°C before being delivered to the bulk CO_2 removal section.

Overall, ~95% of the CO that is fed into the two shift reactors should be converted into CO_2.

Bulk CO_2 Removal Section

The bulk of the CO_2 in the syngas is removed by using absorption process (based on aMDEA solvent)[4].

The process mainly consist of the absorber, flash column, rich-lean HX, and regeneration columns. The CO_2 is removed from the syngas by contacting with the solvent at high pressure. The rich solvent is then delivered to the flash column at lower pressure therefore releasing the volatiles into the gas phase. This is sent to the burners of the SMR. The remaining CO_2 rich solution is then pre-heated by the lean solution before being fed into the upper section of the regeneration column where CO_2 is released into the vapour phase by steam stripping.

The CO_2 released from the solvent in the regeneration column is then sent to the Urea Plant as feedstock to the Urea production (see Section 2.4.2).

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[4] It should be noted that there are several other technologies that could be used in the bulk removal of CO_2 from the shifted syngas. These are described in Task 1. Some of the notable examples include chemical absorption – i.e. aMDEA (BASF), Hot Potassium Carbonate (Giammarco-Vetrocoke and UOP); physical absorption – i.e. Rectisol (Linde, Lurgi), Selexol (UOP), etc…

Methanation Section

Any oxygen containing compounds such as CO and CO_2 are severe poison to the ammonia synthesis catalyst.

To remove any residual CO and CO_2 in the syngas, the methanation process is used as a polishing step. This takes place in the Methanator that involves the following reactions:

$$CO + 3H_2 \Leftrightarrow CH_4 + H_2O + \text{heat}$$
$$CO_2 + 4H_2 \Leftrightarrow CH_4 + 2H_2O + \text{heat}$$

Besides the activity of the catalyst, the temperature, pressure, and moisture content of the syngas determines the conversion efficiency of the methanation reactions. Lower temperature, higher pressure and lower water vapour content favours the methane formation.

The methanation reactions are exothermic. In normal operation, temperature rise of around 20^oC is generally expected.

After the methanation step, the purified syngas that is fed into the ammonia synthesis loop is mostly N_2 and H_2 with approximately 1% (mol) of Ar and CH_4 and less than 5 ppm (mol) of CO and CO_2.

Ammonia Synthesis Section

The ammonia synthesis takes place in the Ammonia Synthesis Converter according to the following reaction:

$$3H_2 + N_2 \Leftrightarrow 2NH_3 + \text{heat}$$

The reaction is reversible. In the ammonia synthesis converter, about 25 - 30% of the N_2 and the H_2 are converted into NH_3. The unconverted part is recycled to the converter after separation of the liquid ammonia product.

Higher pressure and lower temperature favours the formation of ammonia. As the reaction is exothermic, the choice of the operating temperature is based on a compromise between the theoretical conversion and approach to equilibrium.

The normal operating pressure of modern ammonia synthesis may vary between 130 and 220 Bar(g) (inlet of the ammonia converter), depending on load and catalyst activity. The normal operating temperatures are in the range of approx. 370-500°C.

The heat liberated by the reaction (about 750 kcal/kg produced ammonia) is typically utilised to generate high pressure steam and to pre-heat the boiler feed water and the converter feed gas.

After the synthesis gas passing through the converter, the effluent gas will be cooled to about 0°C in the ammonia chillers where most of the ammonia is condensed. The condensed ammonia is then separated from the effluent gas in the ammonia separator.

The recycling of the effluent gas from the converter is generally carried out in the recirculator which is an integral part of the syngas compressor (where the syngas from the methanator is compressed to the synthesis loop pressure).

From the separator, the effluent gas is recycled back to the ammonia synthesis converter through the cold heat exchanger, the recirculating stage of the compressor, and then finally through the hot heat exchanger.

Purge Gas System – Inert Gases Removal

The purified synthesis gas from the methanator contains a small amount of inert gases, mainly Ar and CH_4. These inerts could accumulate due to the recycling of the effluent gas to the ammonia synthesis converter.

A high level of inert gases tends to inhibit the formation of ammonia (thus requiring large catalyst volume or very high operating pressure). To avoid the accumulation of the inert gases, a continuous purge from the synthesis loop is required.

A large portion of the inert are removed through the purge gas system. The purge gas is then sent to the ammonia and hydrogen recovery unit to recover any ammonia (which is mainly recycled back to the ammonia synthesis loop) or hydrogen (which is used as feedstock to the hydrogenation unit); whilst the tail gas is then sent to the SMR as fuel.

A small part of the inert could also be dissolved in the liquid product; and these are normally released during the let-down of the liquid product in the NH_3 separator. The quantity of inert gas leaving the loop this way is proportional to the partial pressure of the inerts (inert level).

Refrigeration

The purpose of the refrigeration circuit is to carry out the various cooling duty within the ammonia synthesis loop. The primary task is to condense the ammonia, which is produced in the converter. Other cooling duties include the cooling of the purge gas, let-down gas, and inert gas.

The refrigeration circuit typically includes the following main equipment: four chillers operating at two different pressure levels, the refrigeration compressor, the ammonia booster compressor, the ammonia condenser and finally the ammonia accumulator.

Liquid ammonia can be produced at different pressure/temperature levels depending on its final use. If ammonia is converted into urea, then ammonia is sent to the Urea Plant at 10-15°C and 10-20 Bar(g); otherwise ammonia is stored in cryogenic atmospheric storage tanks at -33°C.

2.4.2. Urea Plant

Urea is produced from the synthesis of liquid NH_3 and gaseous CO_2, both produced in the Ammonia Plant (see Section 2.4.1.).

In the urea reactor the ammonia and carbon dioxide react to form ammonium carbamate, and then a portion of which are dehydrated to form the urea and water. The reactions are as follows:

$$2NH_3 + CO_2 \leftrightarrow NH_2\text{-}COO\text{-}NH_4$$
$$+ 32560 \text{ kcal/kmol of carbamate (at 1.033 kg/cm}^2\text{; 25°C);}$$

$$NH_2\text{-}COO\text{-}NH_4 \leftrightarrow NH_2\text{-}CO\text{-}NH_2 + H_2O$$
$$- 4200 \text{ kcal/kmol of urea (at 1.033 kg/cm}^2\text{; 25°C)}$$

The first reaction occurs rapidly to completion; whilst the second reaction occurs slowly and this determines the reactor volume.

The Urea plant primarily consists of the (a.) synthesis reactor, (b.) stripper, (c.) carbamate condenser, (d.) decomposer, (e.) vacuum concentrator, (f.) evaporator; and (g.) granulation plant. Figures shown in the succeeding pages present the generic block flow diagram of the urea synthesis and the urea granulation plant.

Depending on the licensor, the urea synthesis reactor is typically operated at around 190°C and 150-160 Bar(g).

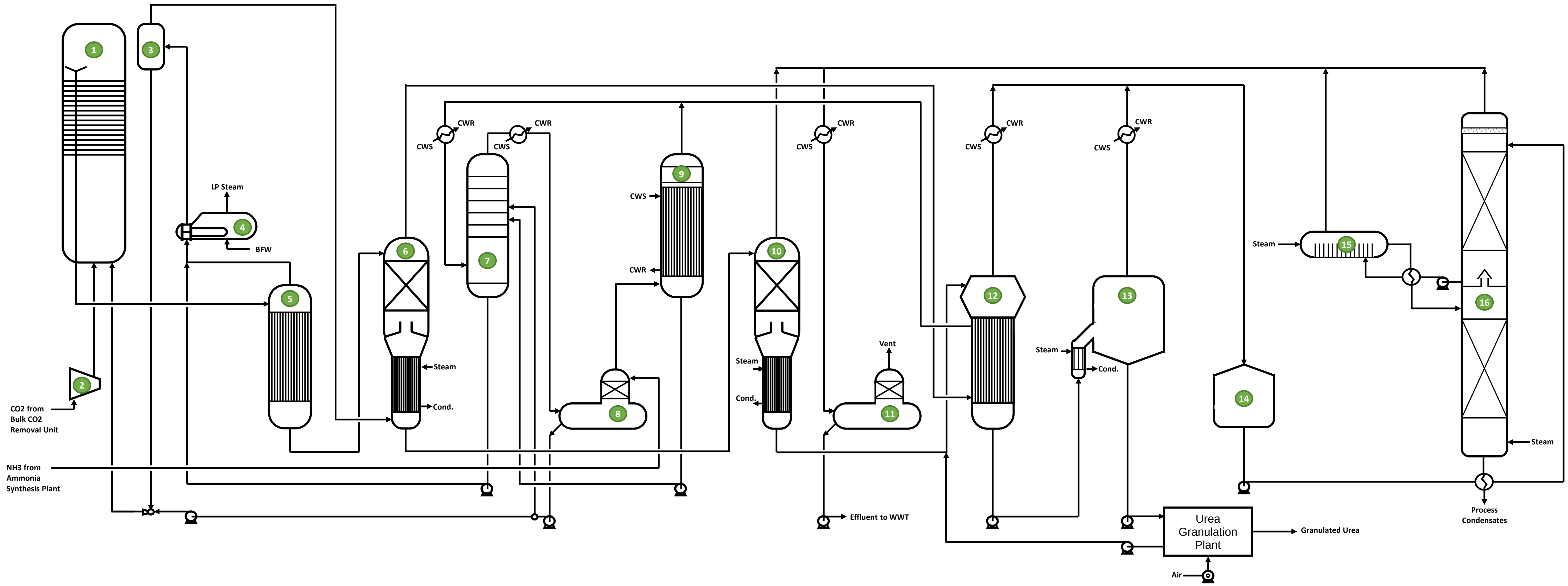
The fraction of ammonium carbamate that dehydrates is determined by the ratios of the various reagents, operating temperature, pressure and their residence time in the reactor.

The CO_2 (with a purity of at least 98.5% vol. and at 0.50 to 0.75 Bar(g)), coming mainly from the Bulk CO_2 Removal Section, is compressed to about 160 Bar(g).

Downstream the urea synthesis the decomposition (and relevant recovery) of unconverted chemical reagents is carried out in several subsequent steps at lower pressures. The decomposition reaction is the reverse reaction of the first one above showed:

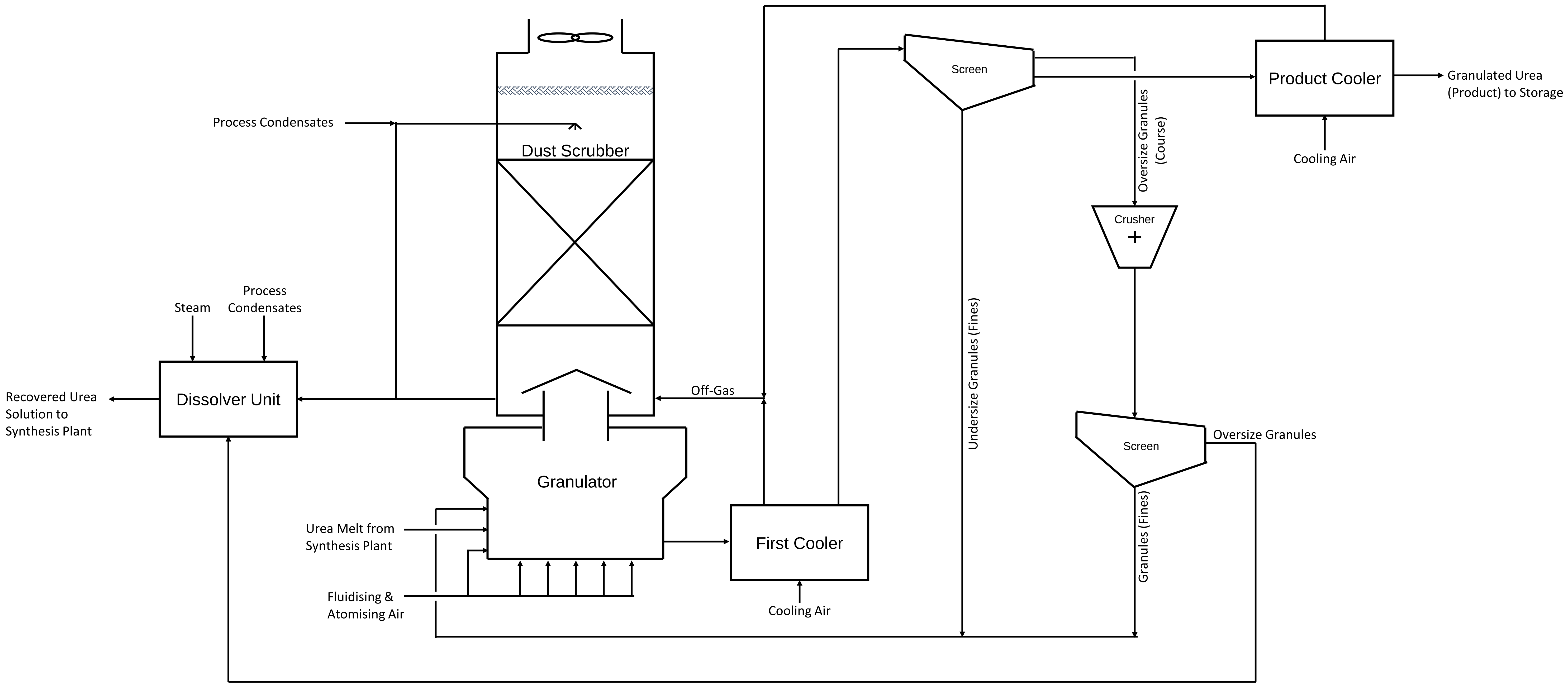
$$NH_2\text{-}COO\text{-}NH_4 \leftrightarrow 2NH_3 + CO_2 \text{ (- Heat)}$$

and, as can be inferred from the equation, it is promoted by reducing pressure and/or adding heat.



1	Urea synthesis reactor	9	Flash tank condenser
2	CO2 compressor	10	LP decomposer
3	Separator	11	Blowdown tank
4	Carbamate condenser	12	Concentrator / pre-evaporator
5	Stripper	13	Vacuum evaporator
6	MP decomposer	14	Ammonia water storage tank
7	Carbamate absorber condenser	15	Hydrolyser
8	Ammonia receiver condenser	16	Process condensate stripper

SCHEMATIC BLOCK FLOW DIAGRAM	REVISION	DATE	BY	CHKD	APP.	SHEET
Urea Synthesis Plant	v 1.0	Feb-17	SS			3 of 4



SCHEMATIC BLOCK FLOW DIAGRAM	REVISION	DATE	BY	CHKD	APP.	SHEET
Urea Granulation Plant	v 1.0	Feb-17	SS			4 of 4

As it is necessary, in order to granulate urea, to concentrate the urea solution up to 96 % by wt., one vacuum concentration stage is provided.

This section is fed by the solution leaving the decomposer bottom at low pressure with about 70 % wt. urea, and is sent first to the vacuum pre-concentrator. The urea solution leaving the vacuum pre-concentrator is sent by urea solution pumps, to the vacuum concentrator. Low pressure saturated steam is supplied concentrate the urea solution.

The mixed phase coming out from the concentrator enters the gas-liquid vacuum separator, from where vapours are extracted by the vacuum system, while the urea solution (~ 96 % by wt.), is sent to the granulation unit.

The process water containing NH_3, CO_2 and urea, coming from the vacuum system, is collected and sent to a urea hydrolyser, where process conditions are suitable to decompose urea into CO_2 and NH_3 so as to have an almost NH_3-CO_2 urea free process condensate to be recovered as boiler feed water.

The concentrated urea solution (~ 96 % by wt.) is fed to the injection heads of the granulation unit where it is finely atomized, assisted by air. Fluidization air flows through the product layer to create a fluid bed and is discharged at the granulator top. Granulated urea flows from the granulator to a cooler and then to the screening section. The fine fraction is recycled directly to the granulator while the coarse material is crushed before being sent to the granulator. The final urea product is then sent to warehouse after final cooling.

2.4.3. *Steam and BFW system*

The steam and BFW system mainly includes the following sub-systems.

- Condensate Polishing Unit receives the process condensate from the ammonia and urea plants, after being stripped in their dedicated sections.

- Deaeration System receives the condensate from the Polishing Unit and the demineralized make-up water from the demi plant. LP steam from the main LP steam header is used as degassing agent. The Deaerator vent, consisting mainly of steam, is discharged to the atmosphere.

- HP Steam System includes the HP BFW pumps, HP steam drum and superheated steam header. The equipment required for BFW pre-heating and steam generation are included in the process unit of the syngas generation and ammonia synthesis. The HP steam from the steam header, at about 12.0 MPa and 510°C, is typically fed to the steam turbine driver of the ammonia syngas compressor. To balance the MP steam requirements of the plant, some of the MP steam is extracted from the steam turbine and is sent to the MP header.

- The MP steam from the steam header, typically at about 4.2 MPa and 375°C, is used as process steam in the ammonia plant and urea plant (i.e. hydrolyser, granulator, stripper). MP steam is also fed to the various steam turbine drivers of the process air compressor, ammonia refrigeration compressor, CO_2 compressor of the urea plant, SMR air and flue gas fan and BFW pumps. Exhaust LP steam from the various steam turbine drivers (back-pressure type) are collected and sent to the LP steam header at 0.6 MPa. This is to be used by the different LP consumers within the process sections.

- Steam Condensate System, which includes the MP and LP headers and drums, recovers the condensate from the different steam users within the plant. Steam released in the MP condensate and LP condensate flash drums are recovered and sent to the LP steam header and to the dearator. The liquid condensates collected from the flash drums of steam headers; and the condensate collected from the condensers of the different steam turbine drivers are sent to the condensate polishing unit.

- It should be noted that, for the case with CO_2 capture (Case 4A-2), where additional LP steam is required for the stripper reboiler, some of the steam drivers driving the compressor (in this case the booster ammonia compressor) has to be modified into electrical driven compressor, therefore increasing the plant electricity demand.

- Blowdown System, which includes the blowdown drum, collects all the blow down steam from the MP and LP steam headers. The LP steam recovered from the blowdown drum is sent to the deaerator as part of the degassing agent; whilst the liquid effluent collected is sent to the waste water treatment plant as effluent.

- Chemical Packages, include chemicals for pH control, oxygen scavenger used in conditioning of the BFW in the Deaerator System, and phosphate injection package used in all the steam systems.

2.4.4. *Demi-Water Plant/Cooling Water System*

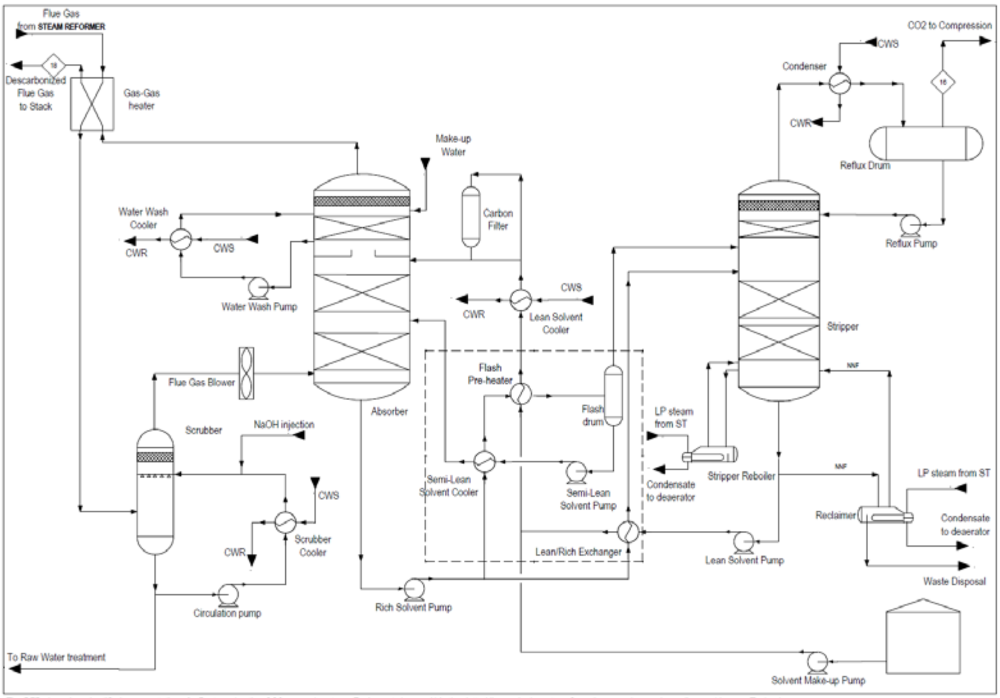
The demi-water required for the steam production is produced by processing raw water using reverse-osmosis system followed by an electro-deionization system. The plant includes a raw water tank, a demi water tank, relevant pumps, and a potable water package and storage.

Chemically treated demi-water is also used as cooling water in a close circuit system (secondary system). This is mainly used for process coolers and for machinery cooling. The secondary cooling system is indirectly cooled by the sea water using plate heat exchangers.

Sea water in once through system (primary system) is used directly for the different steam turbine condensers and for the CO_2 compressor intercooler (for the CO_2 capture case).

2.4.5. *CO_2 Removal from the SMR Flue Gas using MEA Solvent (for Case 4A-2 only)*

The typical CO_2 Removal System consists of the flue gas quench cooler, CO_2 absorption section, heat exchanger network and the CO_2 stripper section. The simplified block flow diagram is presented below.



The BFD above is a simplified representation of a Post combustion CO2 removal system. Equipment shown within the dotted line and relevant configuration may change depending on Licensor Technology

Figure 1. Typical process configuration for a CO_2 absorption and desorption unit

The flue gas of the primary SMR exiting the Combustion Air/Flue Gas Heat Exchanger with a temperature of about 135°C is initially cooled in a gas-gas heat exchanger by the CO_2 lean (decarbonised) flue gas coming from the top of the absorber column before leaving the flue gas stack.

The cooled flue gas is then fed into a quench scrubber using sodium hydroxide or sodium bicarbonate solution to reduce the SOx down to around 1 ppm thus minimising the solvent degradation. The flue gas is further cooled in a direct contact cooler using water wash.

The cooled treated flue gas is then fed to the bottom of the absorber. The column consists of two different level of pack beds and a water wash section.

The flue gas is contacted with a Semi-lean MEA solvent in the first packed bed and then with a Lean MEA solvent in the second pack bed where the CO_2 in the flue gas are absorbed. To improve the efficiency of the absorption process, some of the heat of absorption is removed by using a water cooled pump around cooler situated in the middle section of the absorber.

The flue gas leaving the second pack bed is then scrubbed in the water wash section and passes through a demister section to remove any MEA and/or degradation by-products such as ammonia and any entrained mist. The CO_2 lean flue gas leaves the stack at around 90°C after being heated by the hot flue gas from the SMR.

The Rich MEA solvent leaves the bottom of the absorbers. This is divided into two different streams. The first Rich Amine stream is heated by the partially cooled Lean MEA coming from the stripper in a Rich-Lean Amine Heat Exchanger before being fed to the stripper. The second Rich Amine stream is sent to the flash drum to generate the Semi-lean MEA solvent.

The Rich Amine stream going to the flash drum is first pre-heated by the Semi-Lean MEA coming from the flash drum and further heated by the Lean MEA solvent coming from the reboiler section of the stripper column.

Rich MEA solvent is regenerated in the stripping column. This consists of a stripping and a rectification section.

The vapour from the amine flash drum which mainly consists of the desorbed CO_2 and steam is fed to the top of the rectification section. The Rich MEA is fed into the lower end of the stripping section. This is heated by a vertical thermosyphon reboiler situated at the base of the stripping column. The reboiler is heated by the LP steam. The steam condensate collected is sent to the Condensate Polishing Unit to be processed and recycled back to the Steam and BFW System of the plant.

Periodically some of the circulating amine is sent to the reclaimer unit to remove any heat stable salts which are formed from the reaction of the trace impurities with the MEA. The heavy residues produced after every batch regeneration process are collected and then sent outside the B.L. for disposal. Fresh MEA from the amine storage tanks is added to replenish the lost solvent.

The overhead vapour from the column passes through a demister and sent to the column's condenser that is cooled by the sea water. The wet CO_2 is separated from the two phase mixture in a reflux drum or overhead accumulator. Some of the liquid (mainly water) is recycled back to the column as reflux and the balance are pumped to a water storage tank.

The wet CO_2 at 0.16 MPa is compressed and dried to 11.0 MPa in the compression and dehydration unit as defined by the B.L. definition.

2.4.6. *Balance of Plant (BoP)*

The operation of the whole unit is supported by additional utilities and facilities such as:

- Instrument/Plant Air System
- Flare System
- Drain System
- Interconnecting
- Buildings (Control Room, Electrical Sub-station, Laboratories).

2.4.7. *CO_2 Compression and Dehydration (for Case 4A-2 only)*

The compression and dehydration unit includes the compressor, inter-stage coolers, knockout drums, dehydration package and liquid CO_2 pump.

Wet CO_2 from the stripper's condenser is compressed to 8.0 MPa by using a single train eight-stage centrifugal compressor and then finally pumped to 11.0 MPa ready for transport.

The CO_2 compressor is an integrally geared and electrically driven machine. The compression system includes the associated equipment for anti-surge control, vent, inter-stage coolers, knockout drums and condensate draining facilities.

At the discharge of each compressor stage, CO_2 is cooled by seawater in an inter-stage cooler. The condensate are separated from the compressed gas and collected in the knockout drum. This is then sent to the waste water treatment plant.

After the sixth compression stage, the compressed gas is sent to the dehydration package unit. Drying is achieved by using solid desiccants. For this case, molecular sieve is considered. Other solid desiccants such as Activated Alumina or Silica Gel could also be used.

The dehydration unit consists of two parallel trains of 2-Bed Adsorbers producing the dried CO_2 product with a dew point temperature of -40°C. In normal operation, one bed is used for drying, while the other water-saturated bed is regenerated by recycling a small part (ca.10%) of the dried product gas.

Final compression stages downstream of the dehydration unit increase the product CO_2 pressure up to 8.0 MPa. After being cooled, dried CO_2 product is in dense phase. This is then pumped and delivered the to the battery limits at the specified pipeline pressure of 11.0 MPa.

2.5. Heat and Mass Balance

The Heat and Mass Balances of the ammonia/urea production without and with CCS are presented in the succeeding pages. The information should be referred to the Block Flow Diagram presented in Section 2.3.

The table below shows the breakdown of the CO_2 balance of the ammonia/urea production plant.

	EQUIVALENT FLOW OF CO_2 kmol/h	
	Ammonia/Urea Plant w/o CCS	Ammonia/Urea Plant w/ CCS
INPUT STREAMS		
Natural gas feedstock	1626	1626
Natural gas fuel	637	637
TOTAL (IN)	**2263**	**2263**
OUTPUT STREAMS		
Total carbon in urea product	1568	1651
Captured CO_2 to storage	-	538
SUB-TOTAL (Carbon Not Emitted)	**1568**	**2189**
Flue gas to stack (A)	695	74
Vents	-	-
Emission	695	74
TOTAL (OUT)	**2263**	**2263**
Equivalent CO_2 in Urea product (%)	69.3%	73.0%
Capture Rate - Capture from Flue Gas (%)	-	90.0%
Captured CO_2 to Storage (%)	-	23.8%
AMOUNT OF CARBON NOT EMITTED (%)	**69.3%**	**96.7%**

Indirect CO_2 Emission* (B)	70 – 151	138 - 296
TOTAL CO_2 EMISSION - (A) + (B)	**765 - 846**	**212 - 370**

(*) "Indirect CO_2 emission" is the specific CO_2 emission related to the power imported (from the grid) to the ammonia/urea production complex. For this study, the specific emissions of 348 kg/MWh and 746 kg/MWh from NG and coal fired power plant are used respectively.

2.6. Plant Performance Data

The table below summarizes the productions/ consumptions and CO_2 emissions relevant to the ammonia/urea plant (with and without capture).

Plant Performance Data		Base Case	CCS Case
INLET STREAMS			
Natural Gas (as Feedstock)	t/h	27.021	27.021
Natural Gas (as Fuel)	t/h	10.630	10.630
Natural Gas (Total Consumption)	**t/h**	**37.651**	**37.651**
Natural Gas LHV	MJ/kg	46.50	46.50
Total Energy Input	MW	486.33	486.33
OUTLET STREAMS			
Urea Product to BL	t/d	2,260	2,380
	t/h	94.17	99.17
Urea LHV	MJ/kg	32.70	32.70
Total Energy in the Urea Product	**MW**	**855.4**	**900.8**
Ammonia Product to BL	t/d	68.4	NA
	t/h	2.85	NA
Ammonia LHV	MJ/kg	18.60	NA
Total Energy in the Ammonia Product	**MW**	**14.73**	**NA**
POWER BALANCE			
Ammonia / Urea Production	MWe	-4.490	-9.415
Steam + BFW System	MWe	-1.100	-1.100
Utilities + BoP	MWe	-3.300	-4.400
CO2 Capture Plant	MWe	NA	-0.680
CO2 Compression and Dehydration Unit	MWe	NA	-1.880
Imported Power from the Grid	MWe	8.890	17.475
SPECIFIC CONSUMPTIONS			
Natural Gas (as Feedstock)	GJ/t urea	13.343	12.670
Natural Gas (as Fuel)	GJ/t urea	5.249	4.984
Feed + Fuel	**GJ/t urea**	**18.592**	**17.654**
SPECIFIC EMISSIONS			
Specific CO_2 Emission (Direct)	t/t urea	0.3250	0.0328
Specific CO_2 Captured	t/t urea	NA	0.2387
Equivalent CO_2 in the Urea Product		69.3%	73.0%
Equivalent CO_2 Captured to Storage		NA	23.8%

2.7. Preliminary Utilities Consumption

This section presents the main utilities consumption of the different processes for the ammonia/urea plant without and with CCS.

ESTIMATED UTILITY CONSUMPTIONS

CUSTOMER NAME: IEAGHG
PROJECT NAME: TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE
FWI CONTRACT: 1BD0840 A
LOCATION: THE NETHERLAND

Ammonia-Urea Plant without CO2 Capture

REV.	REV. 0	REV. 1	REV. 2	SHEET
BY	GC			1
CHKD	GC			OF
DATE	October 15			1

	ELECTRIC POWER		STEAM t / h			BFW	EFFLUENT (3)	LOSSES (2)	DMW (1)	RAW WATER	COOLING WATER		SEA WATER		FUEL	INSTR. AIR	Nitrogen
	LOAD BHP	kW	LP	MP	HP	t/h	t/h	t/h	t/h	t/h	ΔT (°C)	m³/hr	ΔT (°C)	m³/hr	MMKcal/h	Nm^3/h	Nm^3/h
AMMONIA and UREA PLANT		4.490		175,0		0,0		28,0			11	4.871	7	1.510	118,1	200	(250)
								-38,7	-164,3								
STEAM AND BFW PLANT		1.100			0,0				284,3				7	7.600			
			0,0	-175,0		0,0		-6,3	-103,0								
UTILITIES / BoP		3.300								22,1	11	-4.871	7	7.661		100	(250)
							-5,1		-17,0 (1)							-300	(-500)
TOTAL		8.890	0	0	0	0	-5,1	-17,0	0	22,1	-	0	-	16.771	118,1	0	0

NOTES:

(1) DMW is the sum of DMW plus condensate from the process unit

(2) Losses includes net water consumptions (-) and productions (+) in the reactions and steam/BFW system vent and steam losses

(3) Water effluent (to be sent to WWT) includes demi plant eluate, process condensate from NH3-Urea plant and steam system blowdown

ESTIMATED UTILITY CONSUMPTIONS

CUSTOMER NAME: IEAGHG		**Ammonia-Urea Plant with CO2 Capture from SMR flue gas**	REV.	REV. 0	REV. 1	REV. 2	SHEET
PROJECT NAME:	TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE		BY	GC			1
FWI CONTRACT:	1BD0840 A		CHKD	GC			OF
LOCATION:	THE NETHERLAND		DATE	October 15			1

	ELECTRIC POWER		STEAM t / h			BFW	EFFLUENT (3)	LOSSES (2)	DMW (1)	RAW WATER	COOLING WATER		SEA WATER		FUEL	INSTR. AIR	Nitrogen
	LOAD BHP	kW	LP	MP	HP	t/h	t/h	t/h	t/h	t/h	ΔT (°C)	m³/hr	ΔT (°C)	m³/hr	MMKcal/h	Nm^3/h	Nm^3/h
AMMONIA and UREA PLANT		9.415	0,0	179,0	0,0	0,0		29,5 (2)			11	4.875	7	1.600	118,1	200,0	(250)
				0,0	0,0		0,0	-38,7	-169,8							-300	(-500)
STEAM AND BFW PLANT		1.100			0,0				288,3				7	7.600		100	(250)
			-31,0	-179,0		0,0	0,0	-6,3	-72,0								
CO2 CAPTURE		680	31,0								11	1.902					
									-31,0	-16,1							
CO2 COMPRESSION in CO2 CAPTURE PLANT		1.880									11	23	7	488			
UTILITIES / BoP		4.400								20,2	11,00	-6.800	7	10.695			
							-4,7		-15,5 (1)								
TOTAL		17.475	0	0	0	0	-4,7	-15,5	0	4,1	-	0	-	20.383	118,1	0	0

NOTES:

(1) DMW is the sum of DMW plus condensate from the process unit

(2) Losses includes net water consumptions (-) and productions (+) in the reactions and steam/BFW system vent and steam losses

(3) Water effluent (to be sent to WWT) includes demi plant eluate, process condensate from NH3-Urea plant and steam system blowdown

2.8. Economic Evaluation

The purpose of this section is to present the economic analysis carried out to evaluate the Levelized Cost of Urea (LCOU) and the CO_2 Avoidance Cost (CAC) for the study cases of ammonia/urea plant without and with CCS.

The capital cost and annual operating & maintenance (O&M) costs for the different cases without and with CCS have been evaluated. These are presented in this section of the report, along with the results of the financial model.

All the general assumptions used in performing the analysis are described in the techno-economic criteria and methodology reported in the Reference Document (Annex I). Only the information and assumptions specific to the ammonia/urea plants are presented in this report.

2.8.1. Investment Cost Estimate

The estimates for the Total Plant Cost (TPC) and Total Capital Requirement (TCR) for the ammonia/urea plant without and with CCS are summarised in the table shown in the succeeding pages. The methodologies used in estimating capital cost are outlined and briefly described in the Reference Document (Annex I).

The **Total Capital Requirement** (**TCR**) is defined as the sum of:

- Total Plant Cost (TPC)
- Spare parts cost
- Start-up costs
- Owner's costs
- Interest during construction
- Working capital

The **Total Plant Cost** (**TPC**) is estimated using the cost database of the Amec Foster Wheeler based on previous projects undertaken and pro-rated to the capacity the different cases evaluated. The TPC of the different study cases are further broken down into the cost estimates of the different main process units:

- Base Case (Case 4A-1): Ammonia/urea w/o CCS
    - Ammonia and Urea Plant
    - Other Utilities and Balance of Plant (BoP)

- CCS Case (Case 4A-2): Ammonia/urea w/additional CO_2 capture from SMR's flue gas
    - Ammonia and Urea Plant
    - CO_2 Capture Plant
    - CO_2 Compression and Dehydration Unit
    - Other Utilities and Balance of Plant (BoP)

In summary, the assumptions used in estimating the other components of the TCR are as follows:

- Spare parts cost

  0.5% of the TPC is assumed to cover the spare part costs. It is also assumed that spare parts have no value at the end of the plant life due to obsolescence.

- Start-up costs consist of:

  ⇒ 2% of TPC, to cover any modifications to equipment that needed to bring the unit up to full capacity.
  ⇒ 25% of the full capacity feedstock and fuel cost for one month, to cover inefficient operation that occurs during the start-up period.
  ⇒ Three months of operating labour and maintenance labour costs, to include training.
  ⇒ One month of chemicals, catalyst and waste disposal costs and maintenance materials costs.

- Owner's cost

  7% of the TPC is assumed to cover the Owner's cost and fees. This is assumed to be incurred in the first year of construction, allowing for the fact that some of the costs would be incurred before the start of construction.

- Interests during construction

  This is calculated from the plant construction schedule and the interest rate is assumed to be the same as the discount rate.

- Working capital:

  The working capital includes inventories of fuel and chemicals (materials held in storage outside of the process plants). Storage for 30 days at full load is considered for chemicals and consumables. It is assumed that the cost of these materials are recovered at the end of the plant life.

Table 1. TPC and TCR of the ammonia/urea plant without and with CCS

		Ammonia/urea plant w/o CCS	Ammonia/urea plant with CCS
Ammonia/urea plant	M€	493	500
Utilities and Balance of Plants (BOP)	M€	53.6	57.4
CO_2 capture unit	M€	-	46.5
CO_2 compression unit	M€	-	10.0
Total Installed Cost (TIC)	**M€**	**546.7**	**613.9**
Contingency	-	20%	20%
Total Plant Cost (TPC)	**M€**	**656.7**	**733.9**
Total Capital Requirement (TCR)	**M€**	**849.2**	**954.4**

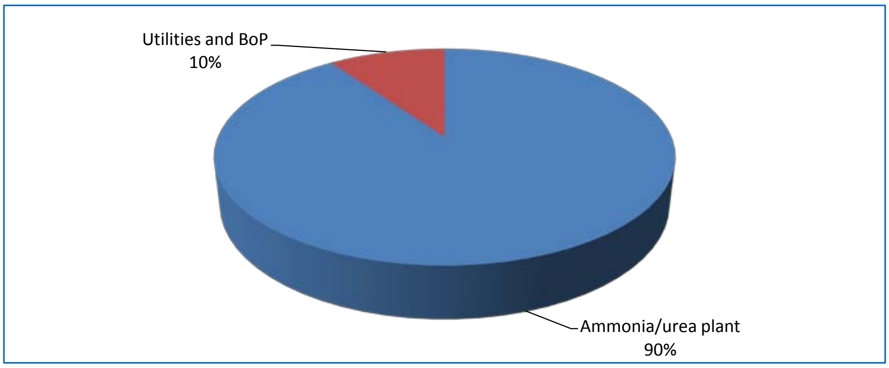


Figure 2. Cost breakdown of the ammonia/urea plant without capture

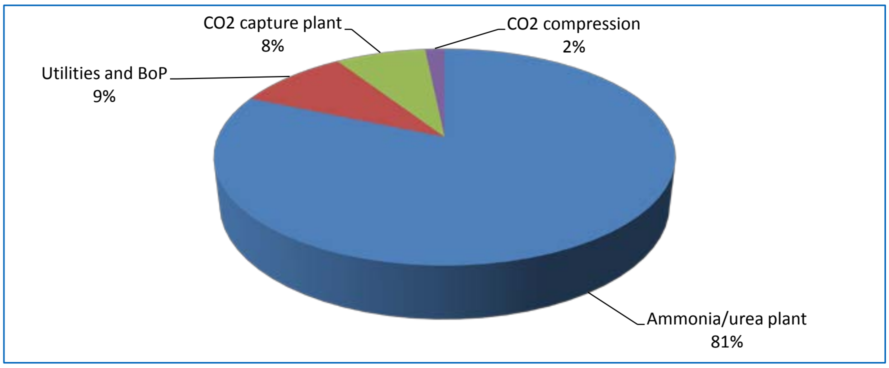


Figure 3. Cost breakdown of ammonia/urea plant with additional CO_2 capture from the SMR's flue gas.

2.8.2. *Annual Operating and Maintenance Cost*

The Annual Operating and Maintenance (O&M) costs consists of the variable cost and the fixed cost. The assumptions used in these cost estimates are briefly described in Annex I – Reference Document.

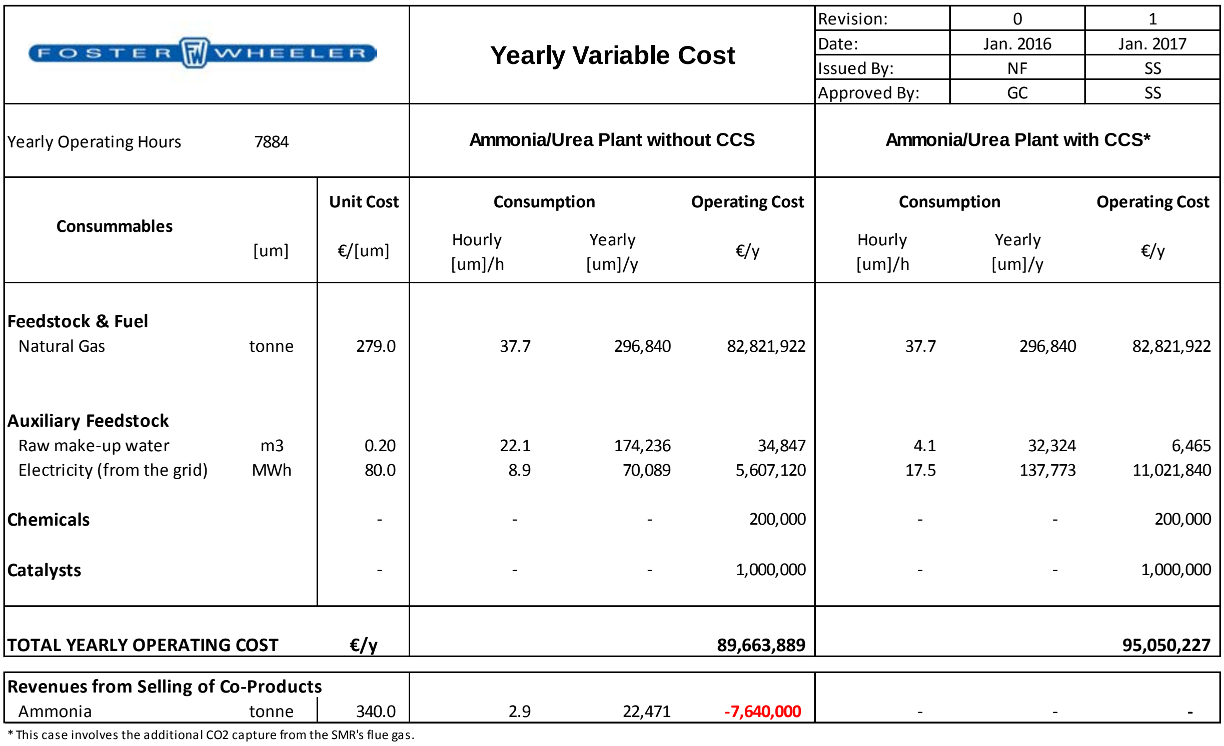
Variable Cost

Table 2 presents the variable cost related to the production of ammonia/urea which mainly consist of the following items:

- Feedstock and fuel (natural gas)
- Raw water make-up
- Electricity
- Catalysts and chemicals
- Additional revenues from the selling of ammonia (base case only).

The annual consumption of the different items included in the variable cost are calculated using the overall mass and energy balances reported in Section 2.5. The cost are estimated based on their annual consumptions using the expected availability of the plant (i.e. 70% and 90% capacity factor for year 1 and year 2 to 25 respectively).

Table 2. Yearly variable costs



				Revision:	0	1
FOSTER WHEELER		Yearly Variable Cost		Date:	Jan. 2016	Jan. 2017
				Issued By:	NF	SS
				Approved By:	GC	SS

Yearly Operating Hours	7884		Ammonia/Urea Plant without CCS			Ammonia/Urea Plant with CCS*		
Consummables	[um]	Unit Cost €/[um]	Consumption Hourly [um]/h	 Yearly [um]/y	Operating Cost €/y	Consumption Hourly [um]/h	 Yearly [um]/y	Operating Cost €/y
Feedstock & Fuel								
Natural Gas	tonne	279.0	37.7	296,840	82,821,922	37.7	296,840	82,821,922
Auxiliary Feedstock								
Raw make-up water	m3	0.20	22.1	174,236	34,847	4.1	32,324	6,465
Electricity (from the grid)	MWh	80.0	8.9	70,089	5,607,120	17.5	137,773	11,021,840
Chemicals		-	-	-	200,000	-	-	200,000
Catalysts		-	-	-	1,000,000	-	-	1,000,000
TOTAL YEARLY OPERATING COST		**€/y**			**89,663,889**			**95,050,227**
Revenues from Selling of Co-Products								
Ammonia	tonne	340.0	2.9	22,471	-7,640,000	-	-	-

* This case involves the additional CO2 capture from the SMR's flue gas.

The reference values for the prices of the natural gas and other consumables are shown in the table below.

Item	Unit	Cost
Natural gas	€/GJ (LHV)	6
Raw water	€/m^3	0.2
Electricity	€/MWh	80
Ammonia (*)	€/t	340
CO_2 transport and storage	€/t CO_2 stored	10
CO_2 emission cost	€/t CO_2 emitted	0

(*) Selling price of the ammonia

Fixed Cost

The fixed cost mainly include:

- Operating Labour Cost
- Overhead Charges
- Maintenance Cost
- Other Fixed Cost.

The yearly cost of the direct labour is calculated assuming for each individual an average cost equal to 60,000 Euro/year, referred to year 2014. The tables presented in the succeeding page presents the line-up of the labour force for the ammonia/urea plant without and with CCS.

The overhead charges are equal to 30% of the operating labour and maintenance labour cost (as described below).

The annual maintenance cost of the plant is estimated as 1.5% of the Total Plant Cost for each case; and these could be sub-divided as maintenance labour and maintenance materials. A maintenance labour to materials ratio of 40:60 can be statistically considered for this breakdown.

The other fixed cost includes local taxes and fees, and insurance cost. This study assumed that the other fixed cost could be covered by 1% of the TPC.

Ammonia/urea plant w/o capture

	SR + utilities	Ammonia/ urea plant	TOTAL	Notes
OPERATION				
Area Responsible	1		1	daily position
Assistant Area Responsible	1		1	daily position
Shift Superintendent	5		5	1 position per shift
Electrical Assistant	5		5	1 position per shift
Shift Supervisor	5	5	10	2 positions per shift
Control Room Operator	5	10	15	3 positions per shift
Field Operator	5	10	15	3 positions per shift
Subtotal			**52**	
MAINTENANCE				
Mechanical group	3	1	4	daily position
Instrument group	3	1	4	daily position
Electrical group	3	1	4	daily position
Subtotal			**12**	
LABORATORY				
Superintendent+Analysts	2		2	daily position
Subtotal			**2**	
TOTAL			**66**	

Cost for personnel

Yearly individual average cost =	60,000	Euro/year
Total cost =	3,960,000	Euro/year

Ammonia/urea plant with capture

	SR + utilities	CO2 capture	Ammonia/ urea plant	TOTAL	Notes
OPERATION					
Area Responsible	1			1	daily position
Assistant Area Responsible	1			1	daily position
Shift Superintendent	5			5	1 position per shift
Electrical Assistant	5			5	1 position per shift
Shift Supervisor	5		5	10	2 position per shift
Control Room Operator	5	5	10	20	4 positions per shift
Field Operator	5		10	15	3 positions per shift
Subtotal				**57**	
MAINTENANCE					
Mechanical group	3		1	4	daily position
Instrument group	3		1	4	daily position
Electrical group	3		1	4	daily position
Subtotal				**12**	
LABORATORY					
Superintendent+Analysts	2			2	daily position
Subtotal				**2**	
TOTAL				**71**	

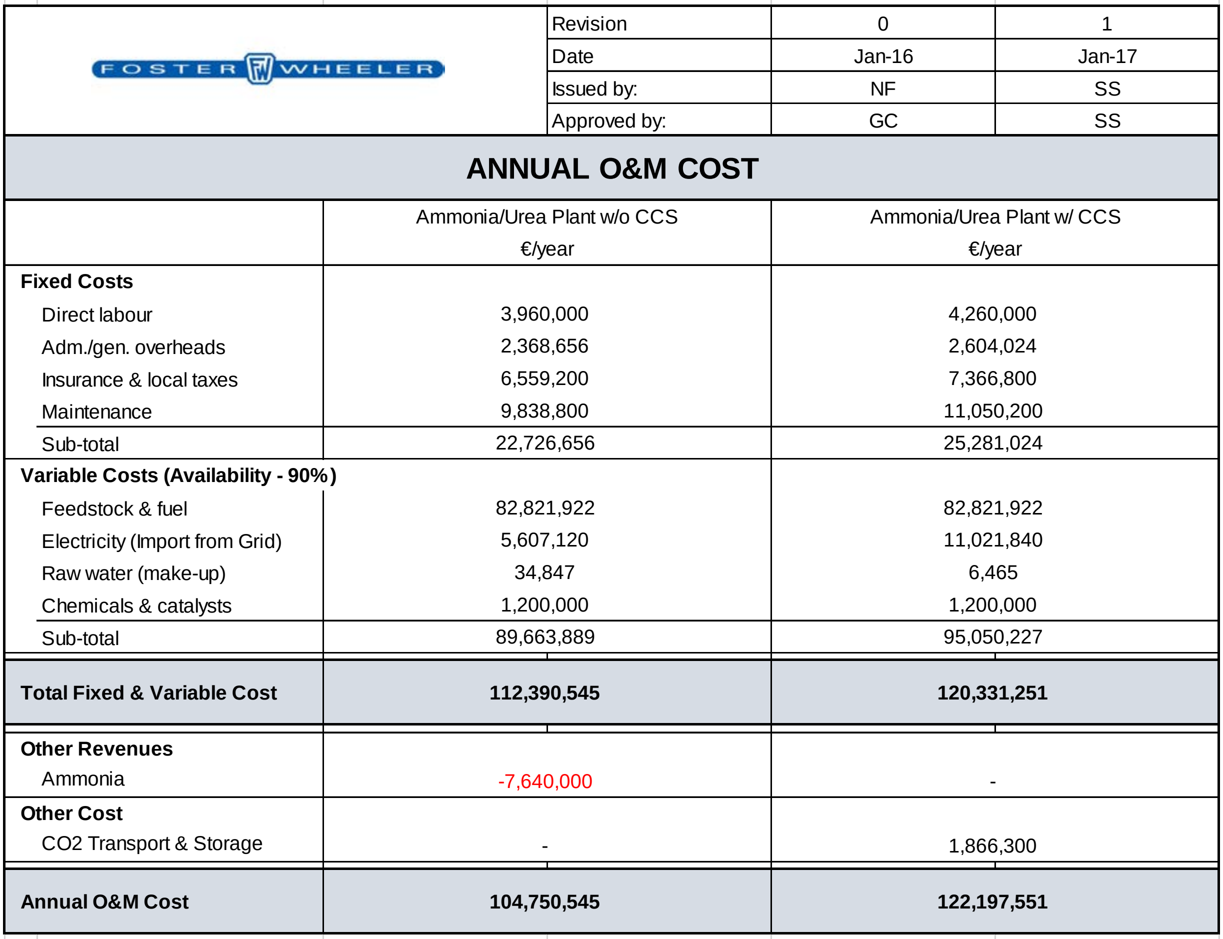
Cost for personnel

Yearly individual average cost =	60,000	Euro/year
Total cost =	4,260,000	Euro/year

Annual Maintenance Cost (2014)

	Maintenance % of TPC	TPC €	Annual Cost €/y
Ammonia/Urea Production w/o CCS	1.5%	655,920,000	9,838,800
Ammonia/Urea Production w/ CCS	1.5%	736,680,000	11,050,200

The following table summarised the annual O&M cost for the two cases.



Revision	0	1
Date	Jan-16	Jan-17
Issued by:	NF	SS
Approved by:	GC	SS

ANNUAL O&M COST

	Ammonia/Urea Plant w/o CCS €/year	Ammonia/Urea Plant w/ CCS €/year
Fixed Costs		
Direct labour	3,960,000	4,260,000
Adm./gen. overheads	2,368,656	2,604,024
Insurance & local taxes	6,559,200	7,366,800
Maintenance	9,838,800	11,050,200
Sub-total	22,726,656	25,281,024
Variable Costs (Availability - 90%)		
Feedstock & fuel	82,821,922	82,821,922
Electricity (Import from Grid)	5,607,120	11,021,840
Raw water (make-up)	34,847	6,465
Chemicals & catalysts	1,200,000	1,200,000
Sub-total	89,663,889	95,050,227
Total Fixed & Variable Cost	**112,390,545**	**120,331,251**
Other Revenues		
Ammonia	-7,640,000	-
Other Cost		
CO2 Transport & Storage	-	1,866,300
Annual O&M Cost	**104,750,545**	**122,197,551**

2.8.3. *Levelized Cost of Urea (LCOU)*

The Levelised Cost of Urea (LCOU) is defined as the selling price at which urea must be produced to achieve break even by the end of its economic lifetime.

In other word, the present value of all the revenues obtained from selling urea and other co-products should be equal to the present value of all costs of building, maintaining and operating the plant over its lifetime.

The selling price of the urea is calculated based on the assumption that NPV = 0 (over the whole life time of the plant).

Also, it should be noted that in this type of analysis, the assumptions for the long-term inflation and the price/cost variations throughout the project life-time are not considered.

2.8.4. *CO_2 Avoidance Cost (CAC)*

The CO_2 Avoidance Cost (CAC) is calculated by comparing the costs and specific emissions of the ammonia/urea plant with CCS with those of the reference case without CCS, based on the following definition:

$$\text{CO}_2\text{ Avoidance Cost (CAC)} = \frac{\text{LCOU}_{\text{CCS}} - \text{LCOU}_{\text{Reference}}}{\text{CO}_2\text{Emissions}_{\text{Reference}} - \text{CO}_2\text{Emissions}_{\text{CCS}}}$$

where:

- LCOU is expressed in € per tonne of urea
- CO_2 emissions is expressed in tonne of CO_2 per tonne of urea
- Cost of CO_2 avoidance is expressed in € per tonne of CO_2

Given that the ammonia/urea production evaluated in this study imports some amount of electricity from the grid as part of their production process, it is therefore essential to take into account the CO_2 emissions associated to the imported electricity.

For simplification, this study assumes the following specific emissions (i.e. from NG and coal fired power plant) associated to the imported electricity consumed by the plant. It should be emphasised that this only provides an indicative range of values related to the CO_2 avoidance cost.

Gas fired power plant without CO_2 capture	348	kg CO_2 / MWh
Coal fired power plant without CO_2 capture	746	kg CO_2 / MWh

For the reference case without CCS, the indirect CO_2 emission corresponding to the electricity imported from a power plant without CCS (coming from either natural gas or coal fired plant), are summarised below:

Ammonia/Urea Production w/o CCS		
Direct specific CO_2 emission	t/t urea	0.3250

Indirect CO_2 emission based on a NGCC power plant		
Electric power import	MWh	8.89
Specific CO_2 emission	kg/MWh	348
Indirect CO_2 emission	t/h	3.09
Indirect specific CO_2 emission	t/t urea	0.0329

Indirect CO_2 emission based on a coal fired power plant		
Electric power import	MWh	8.89
Specific CO_2 emission	kg/MWh	746
Indirect CO_2 emission	t/h	6.63
Indirect specific CO_2 emission	t/t urea	0.0704

For the case with CCS, the indirect CO_2 emission corresponding to the electricity imported from a power plant without CCS (coming from either natural gas or coal fired power plant) are also considered and these are summarised below:

Ammonia/Urea Production w/ CCS		
Direct specific CO_2 emission	t/t urea	0.0328

Indirect CO_2 emission based on a NGCC power plant		
Electric power import	MWh	17.48
Specific CO_2 emission	kg/MWh	348
Indirect CO_2 emission	t/h	6.08
Indirect specific CO_2 emission	t/t urea	0.0613

Indirect CO_2 emission based on a coal fired power plant		
Electric power import	MWh	17.48
Specific CO_2 emission	kg/MWh	746
Indirect CO_2 emission	t/h	13.04
Indirect specific CO_2 emission	t/t urea	0.1315

2.8.5. *Macroeconomic Bases*

The economic assumptions and macroeconomic bases are reported in Annex I - Reference Document. For reader's convenience, the key financial bases assumed in the financial modelling are summarised below:

ITEM	DATA
Discount Rate	8%
Capacity factor	90%
Inflation Rate	Constant Euro
Currency	Euro reported in 4Q2014

2.8.6. *Financial Analysis Results*

This section summarizes the results of the financial analysis performed for ammonia/urea plant study cases, based on the input data reported above.

Table 3 presents the Levelised Cost of Urea (LCOU) and the range of CO_2 avoidance cost (when indirect CO_2 emission from NG or coal fired power plant is included). It should be noted that CAC has been calculated based on a constant electricity price of €80/MWh (i.e. without any distinction between cost of electricity from NG or coal fired power plant – typical for wholesale market price).

Table 3. **LCOU and CAC of the ammonia/urea plant study cases**

Description	Levelized Cost of Urea €/t	CO_2 emission avoidance cost €/t
Ammonia/urea plant w/o CCS	257.3	-
Ammonia/urea plant with CCS* (Indirect emission from NG power plant w/o CCS)	280.3	87.4
Ammonia/urea plant with CCS* (Indirect emission from coal power plant w/o CCS)	280.3	99.8

* This case involves the additional CO_2 capture from SMR's flue gas.

Figure 4 presents the LCOU for the ammonia/urea plant without and with CCS cases and the relative weight of the capital investment, fixed O&M, variable O&M, feedstock and fuel, and CO_2 transport storage cost to the LCOU.

Additionally, for the case without CCS, the revenues from the selling of ammonia should off-set the cost of urea production. This results could be illustrated by having a starting point with a negative value.

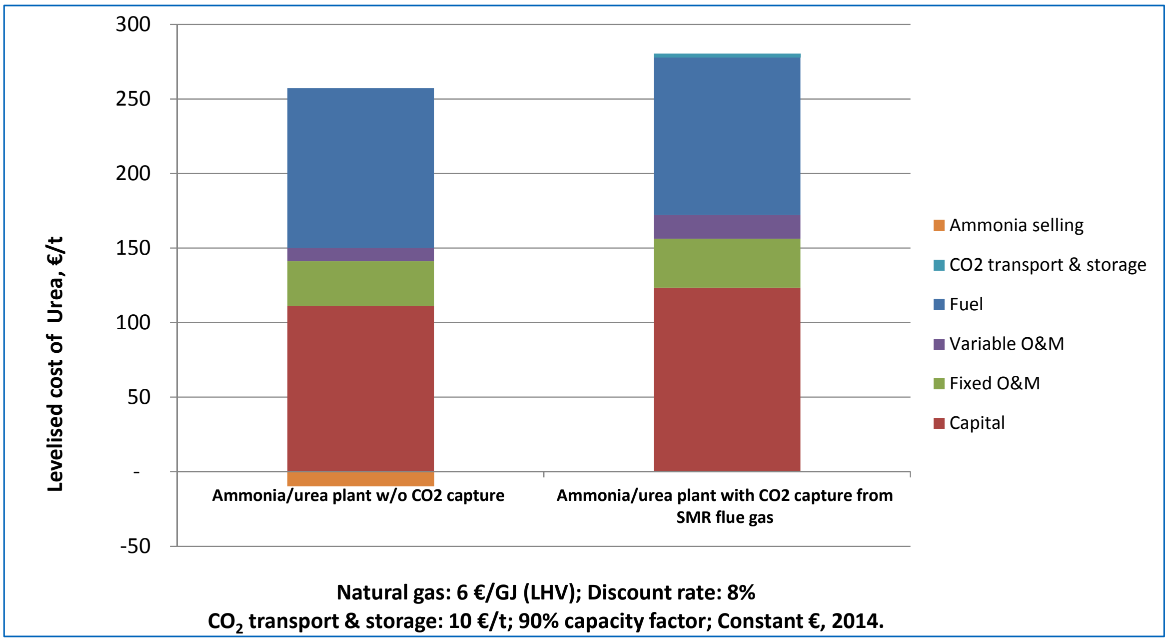


Figure 4. **Levelized Cost of Urea (without and with CCS)**

2.8.7. *Sensitivity Analysis*

Due to the possible floating of some economic input data, a sensitivity analysis is also performed on key parameters such as:

- Natural gas price,
- Electricity price,
- Discount rate,
- CO_2 emission cost,
- Costs related to CO_2 emission or transport & storage

The sensitivity range is summarised in the following table.

Sensitivity Relevant to Ammonia/Urea Study Cases			
Criteria	Unit	Base Case	Sensitivity Range
Natural gas price	€/GJ (LHV)	6	2 to 18
Electricity price	€/MWh	80	20 to 100
Discount rate	%	8	4 to 12
CO_2 transport & storage	€/t stored	10	-20 to 40
CO_2 emission costs	€/t emitted	0	0 to 100

Sensitivity to Natural Gas Price (with Constant Electricity Price)

Figures 5 and 6 present the sensitivity of LCOU and CAC to the NG price. In these analysis, the electricity price is assumed constant at 80€/MWh (without any distinction between cost of electricity produced from NG or coal fired power plant).

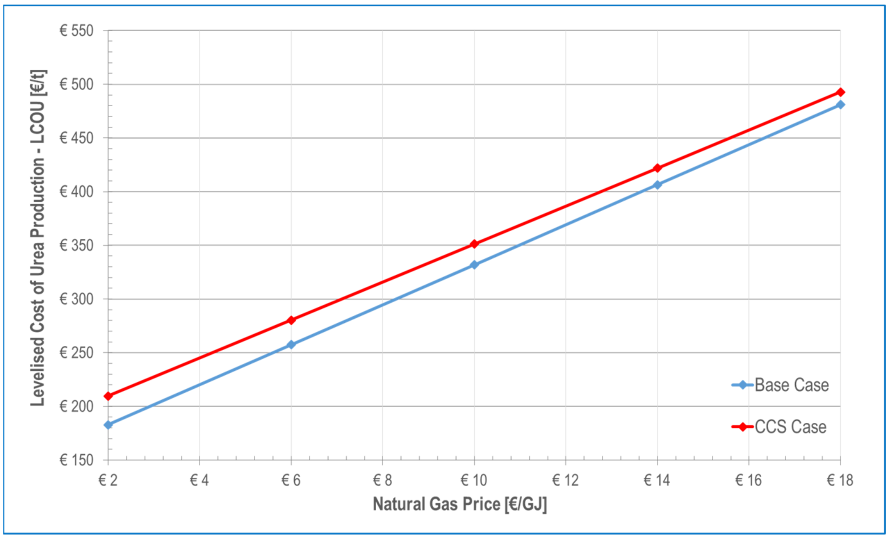


Figure 5. **Sensitivity of Levelised Cost of Urea to the Natural Gas Price**
(at constant electricity price of 80 €/MWh)

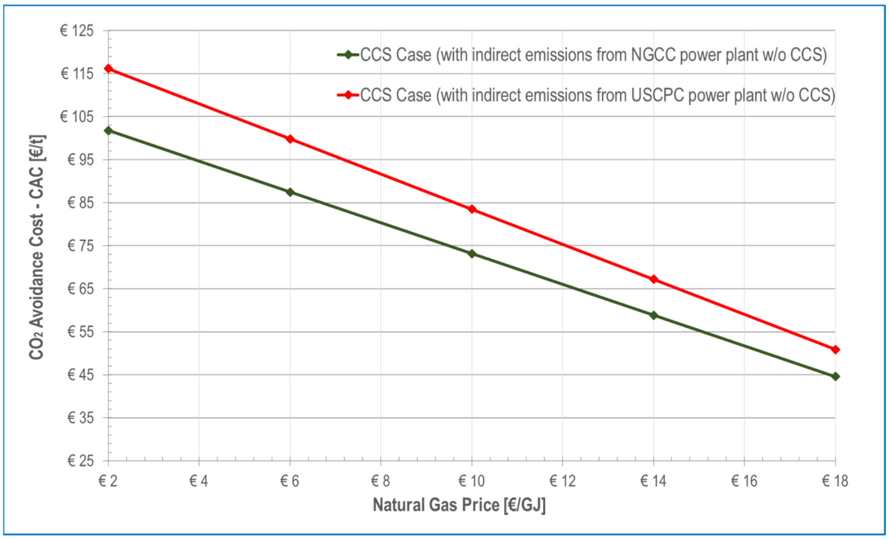


Figure 6. **Sensitivity of CO_2 Avoidance Cost to the Natural Gas Price**
(at constant electricity price of 80 €/MWh)

Sensitivity to Electricity Price (with Constant NG Price)

Figures 7 and 8 present the sensitivity of LCOU and CAC to the electricity price. In these analysis, the natural gas price is assumed constant at 6€/GJ (i.e. electricity price is not indexed to the NG price).

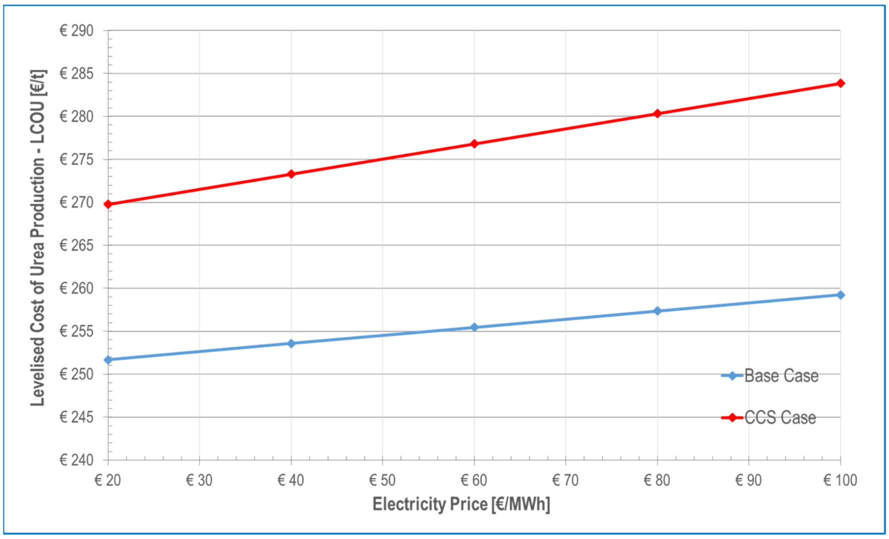


Figure 7. **Sensitivity of Levelised Cost of Urea to the Electricity Price**
(at constant natural gas price of 6 €/GJ)

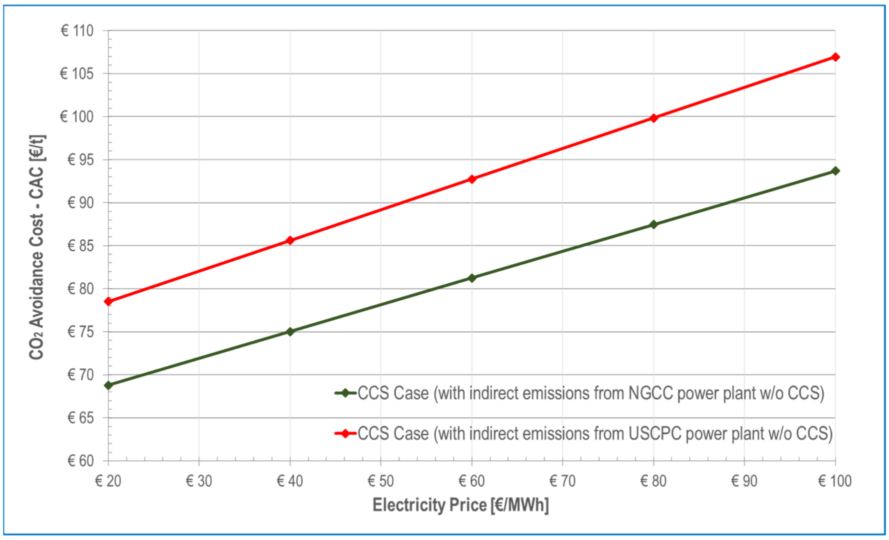


Figure 8. **Sensitivity of CO_2 Avoidance Cost to the Electricity Price**
(at constant natural gas price of 6 €/GJ)

Sensitivity to Natural Gas and Electricity Price Based on the LCOE of the Power Plant

In these analysis, the electricity price is assumed to be equal to the LCOE of a NG or coal fired power plant (i.e. over the fence contract scenario) instead of assuming a constant electricity price (i.e. wholesale market scenario) as presented previously.

Results from the previous IEAGHG reports[5] were used as the basis to estimate the LCOE of NG or coal fired power plants without CCS. Figures 9 present the sensitivity of the LCOE from NG and coal fired power plants without CCS with respect to the price of natural gas.

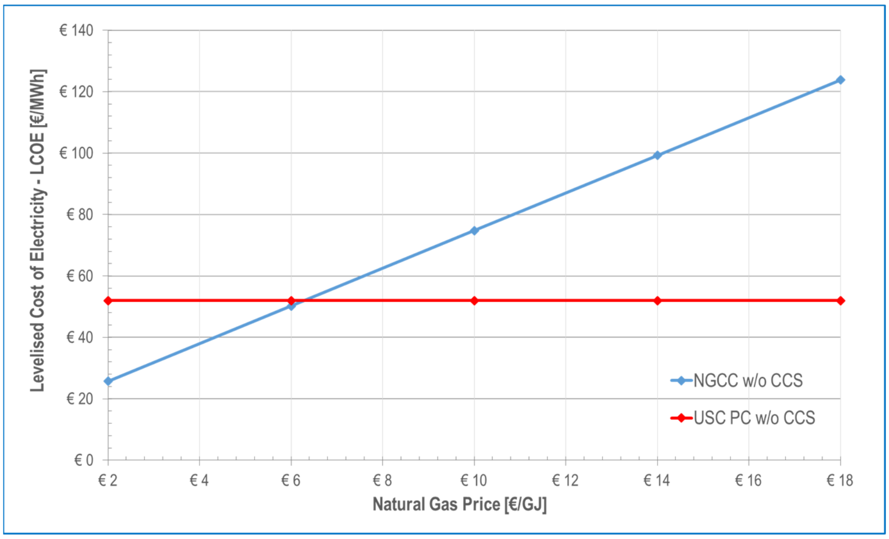


Figure 9. Sensitivity of LCOE to the Price of the Natural Gas
(Note: coal price used by the USC PC w/o CCS is assumed at €2.5/GJ)

Key Performance Data – NGCC or USC PC Power Plant without CCS [5]		
	NGCC w/o CCS	USC PC w/o CCS
Type of Power Plant	Base Load	Base Load
Configuration	2 x GT + 1 x ST	-
Gross Power Output (MWe)	927.5	1076.7
Net Power Output (MWe)	903.8	1029.6
Net Efficiency – LHV basis (%)	58.84%	44.10%
CO_2 Emission Factor (kg/MWh)	348	746

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[5] Previous IEAGHG report in determining the LCOE from the NG or coal fired power plants without CCS.

- IEAGHG Report No. 2014-03 “CO_2 Capture at Coal Based Power and Hydrogen Plant”
- IEAGHG Report No. 2015-05 “Oxy-Combustion Turbine Power Plants”

Figures 10 and 11 present the corresponding sensitivity of LCOU and CAC to the natural gas price using the LCOE from NGCC or USC PC without CCS. These results should illustrate the effect to the LCOU and CAC if electricity price are indexed to the natural gas price (should be referred to when the ammonia/urea plant is buying electricity directly from the NGCC power plant without CCS).

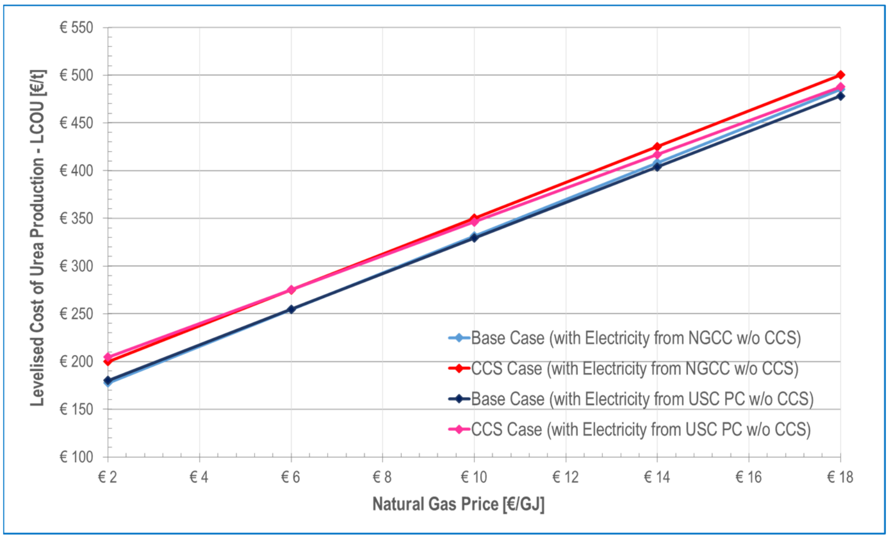


Figure 10. Sensitivity of Levelised Cost of Urea to the Natural Gas Price
(based on electricity price presented in Figure 8)

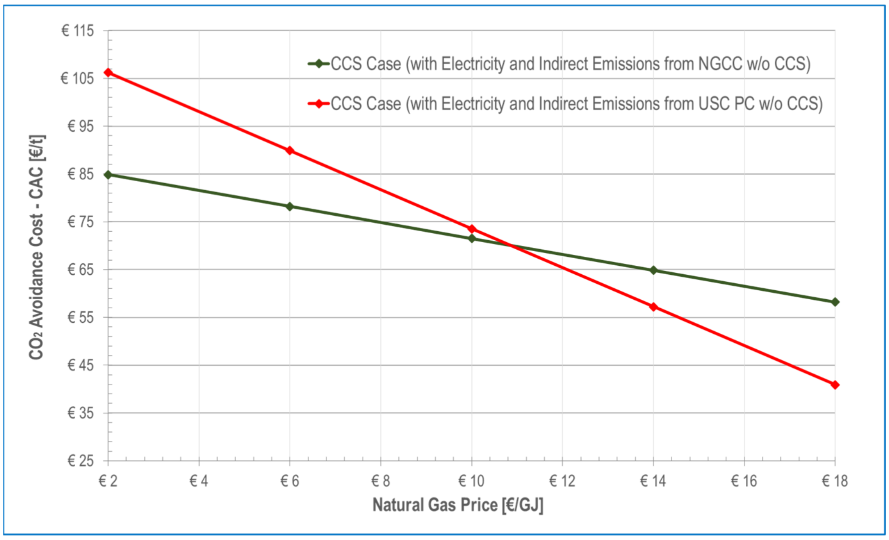


Figure 11. Sensitivity of CO_2 Avoidance Cost to the Natural Gas Price
(based on electricity price presented in Figure 8)

Sensitivity to Discount Rate

Figures 12 and 13 present the sensitivity of LCOU and CAC to the discount rate. In these analysis, the natural gas and electricity price are assumed constant at 6 €/GJ and 80 €/MWh respectively.

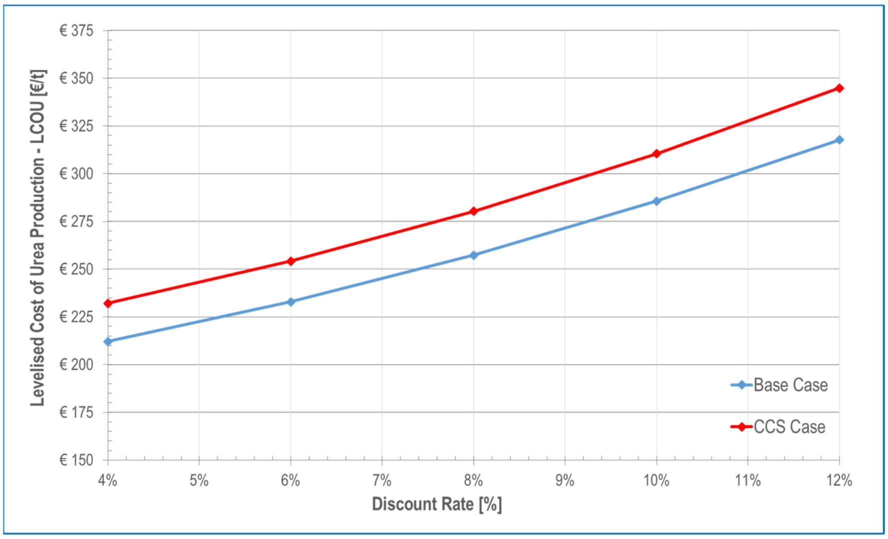


Figure 12. Sensitivity of Levelised Cost of Urea to the Discount Rate

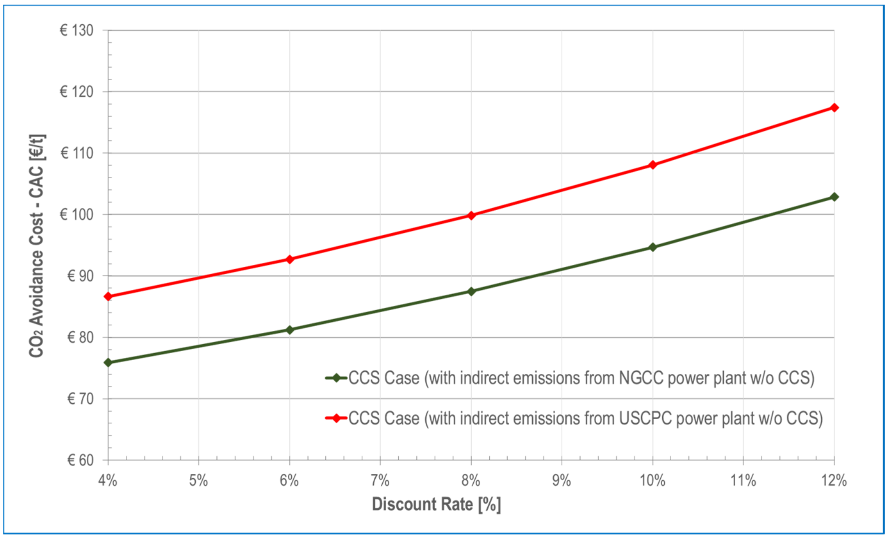


Figure 13. Sensitivity of CO_2 Avoidance Cost to the Discount Rate

Sensitivity to CO_2 Transport & Storage Cost

Figures 14 and 15 present the sensitivity of LCOU and CAC to the cost of CO_2 transport and storage. In these analysis, a negative value for CO_2 transport and storage cost could represent cash credit (which could be applicable to scenarios for EOR operation where CO_2 is bought as a commodity or any CO_2 stored is given tax credits).

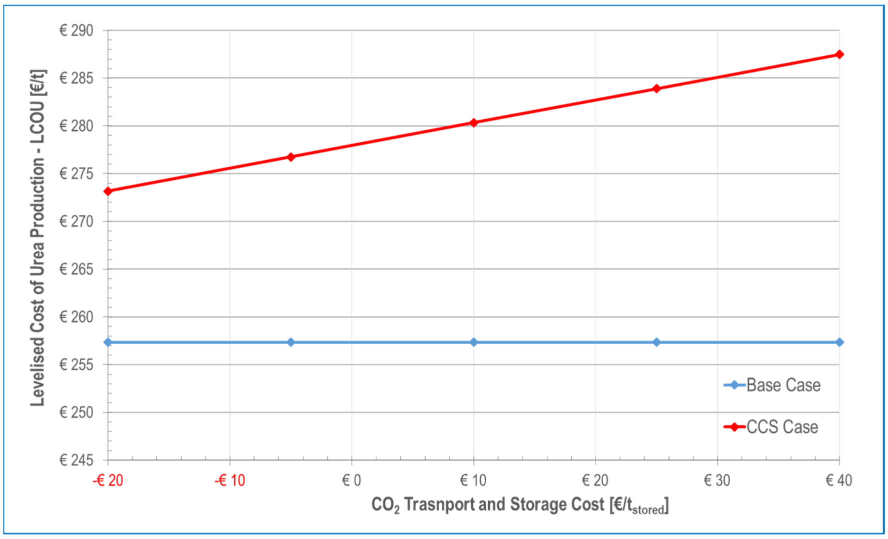


Figure 14. Sensitivity of Levelised Cost of Urea to the CO_2 Transport & Storage Cost

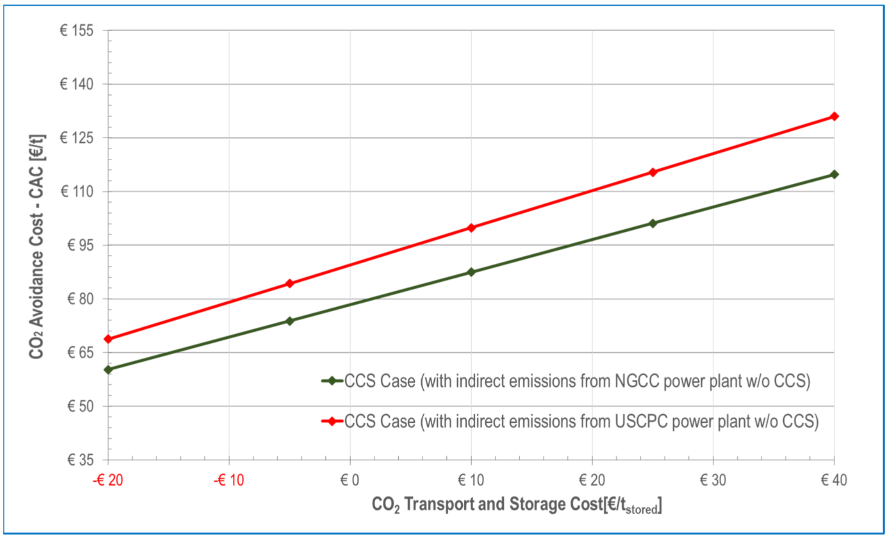


Figure 15. Sensitivity of CO_2 Avoidance Cost to the CO_2 Transport & Storage Cost

Sensitivity to CO_2 Emission Cost

Figures 16 and 17 present the sensitivity of LCOU to the cost of CO_2 emissions with the indirect CO_2 emissions accounted for in the cost of electricity bought from NGCC or Coal fired power plants respectively. In these analysis, the calculation of the LCOU is based on electricity price of the NGCC or USCPC power plant (as shown in Figure 18) where the LCOE accounts for cost of the indirect CO_2 emission.

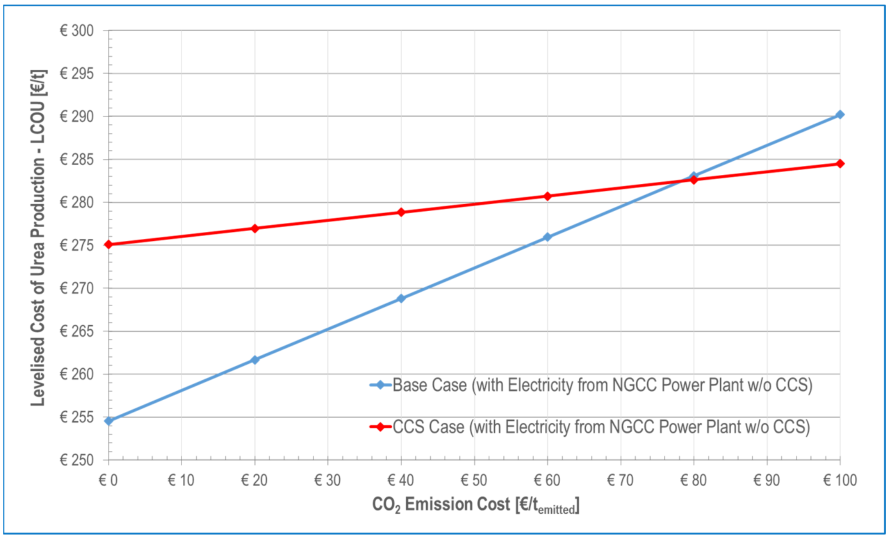


Figure 16. **Sensitivity of Levelised Cost of Urea for the Ammonia/Urea Plant (with Electricity from NGCC Power Plant w/o CCS) to the CO_2 Emission Cost**
(calculation is based on electricity price presented in Figure 18)

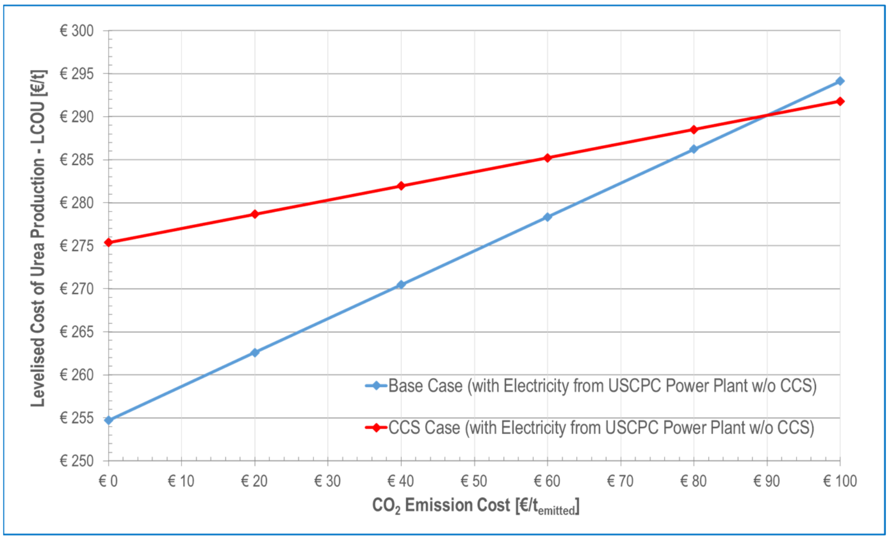


Figure 17. **Sensitivity of Levelised Cost of Urea for the Ammonia/Urea Plant (with Electricity from USCPC Power Plant w/o CCS) to the CO_2 Emission Cost**
(calculation is based on electricity price presented in Figure 18)

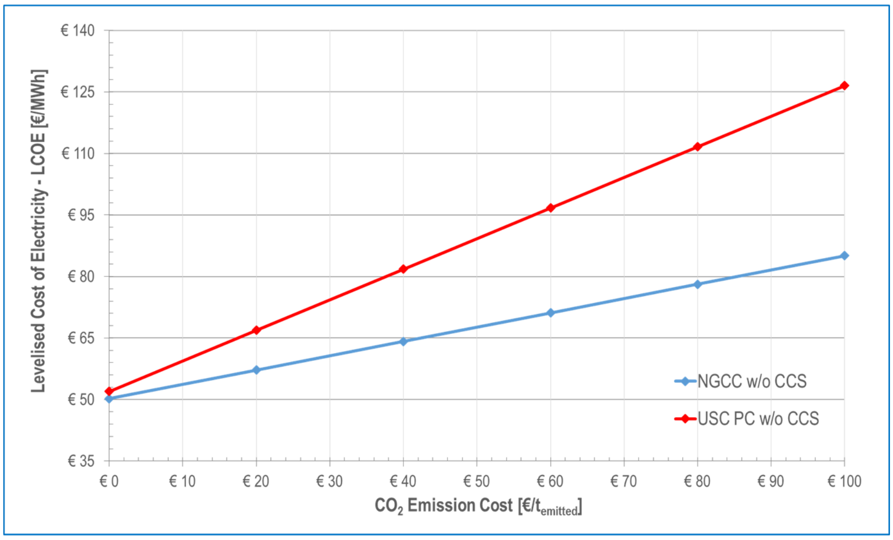


Figure 18. Sensitivity of Levelised Cost of Electricity (from NGCC or USCPC Power Plant) to the CO_2 Emission Cost

(Calculation is based on a constant natural gas price of 6 €/GJ)

3. Case 4B: Methanol Production from Natural Gas

3.1. Basis of Design

The general plant design data and assumptions used in these cases shall be referred to the Reference Document (Task 2). Whilst, the specific information relevant to the methanol production are reported in this section.

3.1.1. *Capacity*

The plant is designed to produce 5,000 t/d of Grade AA methanol.

3.1.2. *Product Specification*

Properties of the AA grade methanol produced at the battery limit are listed below.

Parameter		Value	Unit
Purity, methanol	min	99.85	wt %
Water content	max	0.1	wt %
Ethanol content	max	10	ppm
Acetone	max	20	ppm
Acetone & Aldehyde	less than	30	ppm
Non volatile content	max	0.8	mg/100 ml
Iron	max	0.1	ppm
Specific gravity	max	0.7928 at 20°C	g/cm^3
Appearance	-	Clear, free from suspended matter	
Aromatics		1	ppm

3.1.3. *Syngas Specifications*

The main considerations in the specification of the syngas used in the methanol synthesis is the level of sulphur and chloride compounds which could as poison to both primary reformer’s and methanol synthesis catalyst.

As such the following specification are expected:

- Process Feed Gas to the Primary Reformer (%v dry basis)

Sulphur	< 0.05 ppm	poison to the catalyst
Chloride	< 0.05 ppm	poison to the catalyst
As, V, Pb, Hg	< 5 ppb	poison to the catalyst
Olefin	< 1-2%	carbon formation

- Syngas to the Low Temperature Shift reactor (%v dry basis)
    - Sulphur < 0.1 ppm poison to the catalyst
    - Chloride < 5 ppb poison to the catalyst

Additionally, based on conventional methanol synthesis catalyst (i.e. all copper based catalyst), CO is more preferred than CO_2 as the later tends to become the limiting step in achieving the equilibrium during the methanol synthesis. As such module number of around 2 normally maintained.

3.1.4. *Plant Battery Limits*

Main plant battery limits are listed below:

- Natural gas (in)
- Waste Water streams (out)
- Raw water (in)
- Sea water (in/out)
- Electric power (in)
- Compressed CO_2 rich stream (out) applicable to the CO_2 capture case
- Methanol (out)

3.1.5. *Steam*

Conditions of HP, MP and LP steam used in the methanol plant are summarised in Table 4:

Table 4. Steam Conditions

Steam (at Process Unit BL)	**Pressure (MPag)**				**Temperature (°C)**			
	min	**normal**	**max**	**design**	**min**	**normal**	**max**	**design**
HP Steam	11.4	12.0	12.22	13.4	500	510	520	530
MP Steam	3.92	4.13	4.21	4.68/FV	375	395	405	425
LP Steam	0.31	0.34	0.37	0.63/FV	150	177	180	210

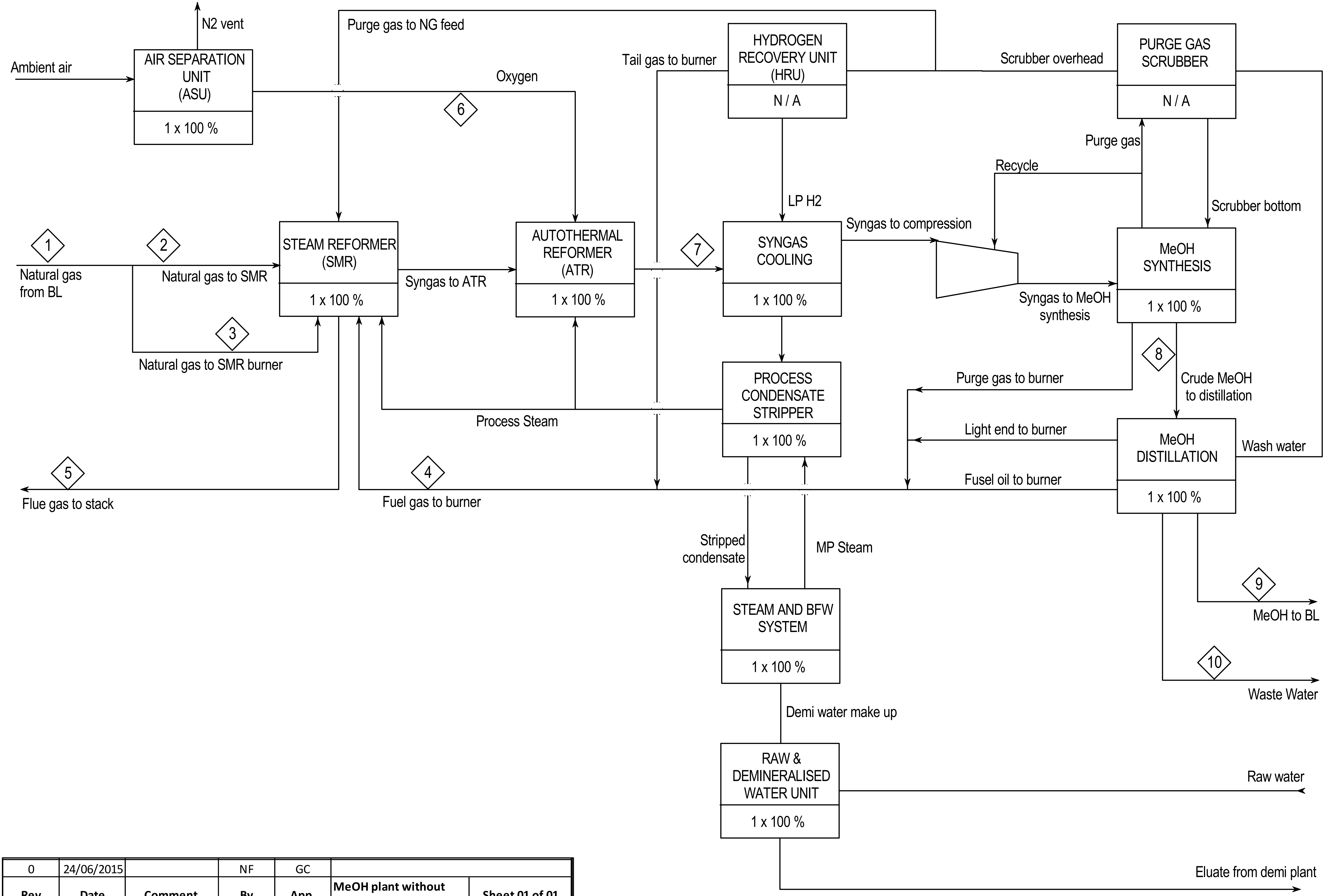
3.2. Units Arrangement

Methanol Plant (Case 4B-1 and Case 4B-2) consists of the following units:

- Syngas Generation Unit
  - Feedstock Pre-treatment
  - Pre-Reformer
  - Primary Steam Reformer (SMR)
  - Autothermal Reformer (ATR)
  - Syngas Cooler
- Methanol Production Unit
  - Syngas Compression
  - Methanol Synthesis Loop
  - Methanol Distillation Unit
  - Purge Gas Scrubber
  - Hydrogen Recovery Unit
  - Methanol Storage
- Steam and BFW System
- Air Separation Unit
- Demi Water Plant
- Utilities and Balance of Plant (BoP), consisting of:
  - Cooling Water System
  - Flare System
  - Interconnecting
  - Drain System
  - Buildings (Control Room, Laboratories, Electrical Sub-Station).
- CO_2 Capture Plant (only for Case 4B-2)
- CO_2 Compression and Drying (only for Case 4B-2).

3.3. Overall Block Flow Diagram

The BFDs and associated H&MB shown in the succeeding pages present the different unit processes included in the Methanol Plant (without and with CCS).



0	24/06/2015		NF	GC		
Rev	**Date**	**Comment**	**By**	**App.**	**MeOH plant without CO2 capture**	**Sheet 01 of 01**

HEAT AND MATERIAL BALANCE
MeOH plant without CO2 capture

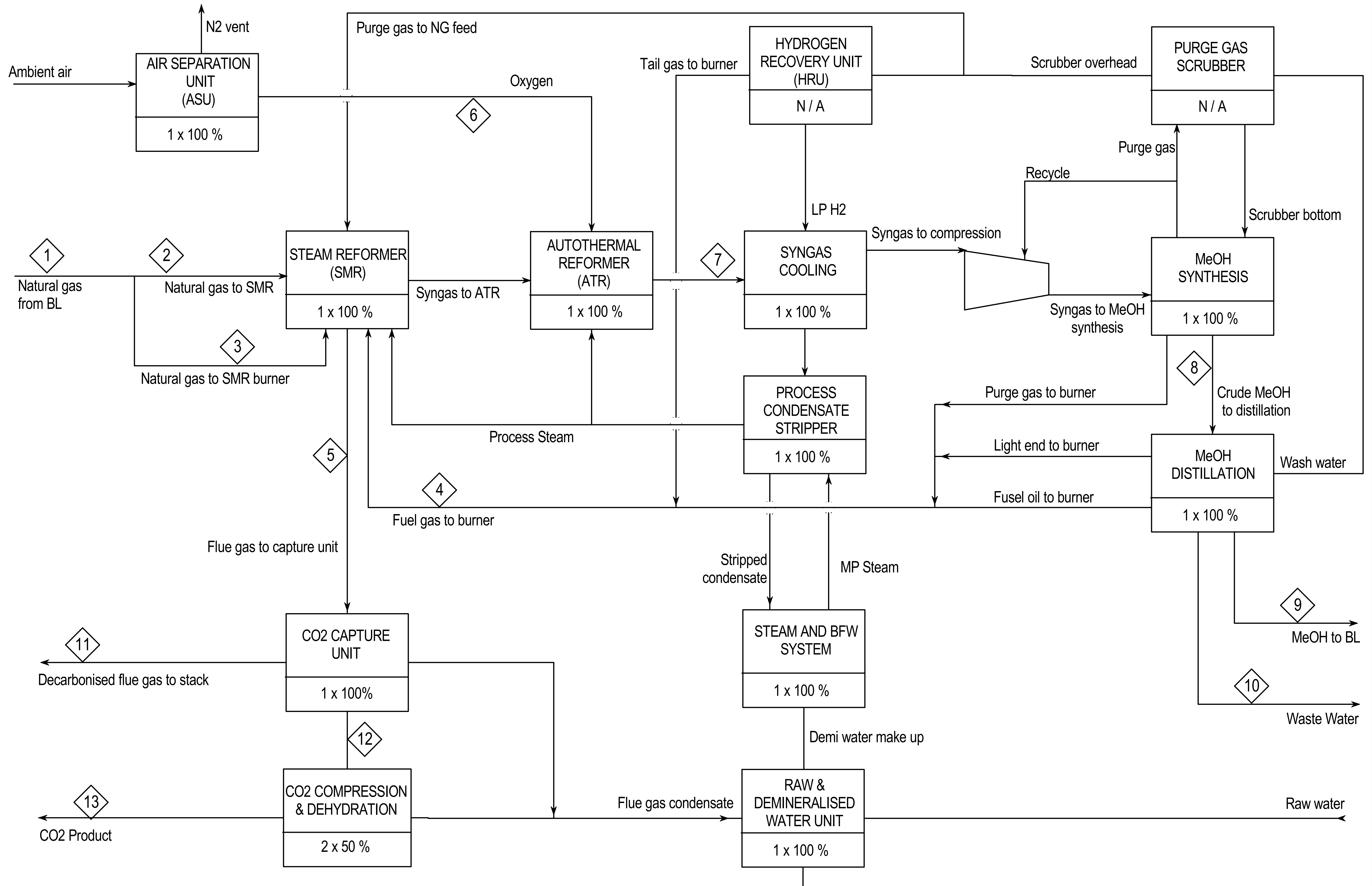
		REV	DATE	BY	CHKD	APP
CLIENT:	IEAGHG					
PROJECT NAME:	TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE FOR INDUSTRY	0	June 2015	NF	GC	GC
FWI CONTRACT:	1BD0840A					
LOCATION:	THE NETHERLAND					

Stream		1	2	3	4	5	6	7	8	9	10	11	12
Description		Natural Gas From B.L.	Natural Gas feedstock to Methanol Plant	Natural Gas fuel to burners	Tail gas + Purge gas + fusel oil + light end	Flue gas to stack	Oxygen to ATR	ATR outlet	Crude MeOH to distillation	MeOH to storage	Waste water effluent		
Temperature	*°C*	9	15	15	50	155	225	1029	50	45	45		
Pressure	*MPa*	7,00	3,50	0,15	0,13	atm	3,16	2,56	0,62	atm	atm		
Molar Flow	*kmol/h*	7560	6610	950	1632	16296	3441	37131	9415	6502,8	2042		
Mass Flow	*kg/h*	136216	119098	17119	30918	453441	110238	523662	264894	208364	36780		
Composition													
CO2	*mol/mol*	0,0200	0,0200	0,0200	0,2594	0,1026	0,0000	0,0620	0,0111	0,0000	0,0000		
CO	*mol/mol*	0,0000	0,0000	0,0000	0,0284	(2)	0,0000	0,1289	0,0000	0,0000	0,0000		
Hydrogen	*mol/mol*	0,0000	0,0000	0,0000	0,5059	0,0000	0,0000	0,4292	0,0001	0,0000	0,0000		
Nitrogen	*mol/mol*	0,0089	0,0089	0,0089	0,0564	0,6727	0,0100	0,0035	0,0002	0,0000	0,0000		
Ar	*mol/mol*	0,0000	0,0000	0,0000	0,0232	0,0109	0,0100	0,0010	0,0000	0,0000	0,0000		
Oxygen	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0134	0,9800	0,0000	0,0000	0,0000	0,0000		
Methane	*mol/mol*	0,8900	0,8900	0,8900	0,0323	0,0000	0,0000	0,0014	0,0000	0,0000	0,0000		
Ethane	*mol/mol*	0,0700	0,0700	0,0700	0,0030	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000		
Propane	*mol/mol*	0,0100	0,0100	0,0100	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000		
n-Butane	*mol/mol*	0,0010	0,0010	0,0010	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000		
n-Pentane	*mol/mol*	0,0001	0,0001	0,0001	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000		
H2O	*mol/mol*	0,0000	0,0000	0,0000	0,0240	0,2004	0,0000	0,3740	0,2878	0,0000	1,0000		
MeOH	*mol/mol*	0,0000	0,0000	0,0000	0,0673	0,0000	0,0000	0,0000	0,7007	1,0000	0,0000		
Contaminants:													
H2S	*ppm v*	(1)											
NOx	*mg/Nm3*												

Notes:

(1) For feedstock purification section design purposes 5 ppmv of H2S have been assumed in NG to Methanol Plant

(2) 30 mg/Nm3 max



0	24/06/2015		NF	GC		
Rev	**Date**	**Comment**	**By**	**App.**	**MeOH plant with CO2 capture from SMR flue gas**	**Sheet 01 of 01**

HEAT AND MATERIAL BALANCE
MeOH plant with CO2 capture from SMR flue gas

CLIENT:	IEAGHG	**REV**	**DATE**	**BY**	**CHKD**	**APP**
PROJECT NAME:	TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE FOR INDUSTRY	0	June 2015	NF	GC	GC
FWI CONTRACT:	1BD0840A					
LOCATION:	THE NETHERLAND					

Stream		**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
Description		Natural Gas From B.L.	Natural Gas feedstock to Methanol Plant	Natural Gas fuel to burners	Tail gas + Purge gas + fusel oil + light end	Flue gas to capture unit	Oxygen to ATR	ATR outlet	Crude MeOH to distillation	MeOH to storage	Waste water effluent	Decarbonized Flue gas to Stack	CO2 to compressor	CO2 to Pipeline
Temperature	*°C*	9	15	15	50	155	225	1029	50	45	45	43	43	24
Pressure	*MPa*	7,00	3,50	0,15	0,13	atm	3,16	2,56	0,62	atm	atm	0,10	0,16	11,00
Molar Flow	*kmol/h*	7560	6610	950	1632	16296	3441	37131	9415	6502,8	2041,5	12616,0	1593,1	1505,5
Mass Flow	*kg/h*	136216	119098	17119	30918	453441	110238	523662	264894	208364	36780	349132	67835	66256
Composition														
CO2	*mol/mol*	0,0200	0,0200	0,0200	0,2594	0,1026	0,0000	0,0620	0,0111	0,0000	0,0000	0,0133	0,9450	0.9999+
CO	*mol/mol*	0,0000	0,0000	0,0000	0,0284	0,0000	0,0000	0,1289	0,0000	0,0000	0,0000	(2)	0,0000	0,0000
Hydrogen	*mol/mol*	0,0000	0,0000	0,0000	0,5059	0,0000	0,0000	0,4292	0,0001	0,0000	0,0000	0,0000	0,0000	0,0000
Nitrogen	*mol/mol*	0,0089	0,0089	0,0089	0,0564	0,6727	0,0100	0,0035	0,0002	0,0000	0,0000	0,8689	0,0000	0,0000
Ar	*mol/mol*	0,0000	0,0000	0,0000	0,0232	0,0109	0,0100	0,0010	0,0000	0,0000	0,0000	0,0141	0,0000	0,0000
Oxygen	*mol/mol*	0,0000	0,0000	0,0000	0,0000	0,0134	0,9800	0,0000	0,0000	0,0000	0,0000	0,0173	0,0000	0,0000
Methane	*mol/mol*	0,8900	0,8900	0,8900	0,0323	0,0000	0,0000	0,0014	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Ethane	*mol/mol*	0,0700	0,0700	0,0700	0,0030	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
Propane	*mol/mol*	0,0100	0,0100	0,0100	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
n-Butane	*mol/mol*	0,0010	0,0010	0,0010	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
n-Pentane	*mol/mol*	0,0001	0,0001	0,0001	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000	0,0000
H2O	*mol/mol*	0,0000	0,0000	0,0000	0,0240	0,2004	0,0000	0,3740	0,2878	0,0000	1,0000	0,0865	0,0550	0,0000
MeOH	*mol/mol*	0,0000	0,0000	0,0000	0,0673	0,0000	0,0000	0,0000	0,7007	1,0000	0,0000	0,0000	0,0000	0,0000
Contaminants:														
H2S	*ppm v*	(1)												
NOx	*mg/Nm3*													

Notes:

(1) For feedstock purification section design purposes 5 ppmv of H2S have been assumed in NG to Methanol Plant

(2) 30 mg/Nm3 max

3.4. Process Description

The study evaluates the techno-economic performance of a nominal 5000 t/d methanol plant without and with CO_2 capture using natural gas as feedstock.

There are several technologies available for producing ASTM AA grade methanol. The process scheme selected for this study is generic, with no reference to specific licensor and equipment suppliers.

This section refers to the overall Block Flow Diagram presented in Section 3.3.

The methanol production process selected for this study is based on a typical process flow scheme that consists of a single train combined reformers which consist of conventional steam methane reformer (SMR) in series with an oxygen blown autothermal reformer (ATR). The syngas produced is then fed into the methanol synthesis loop (reactor); and the AA grade methanol is recovered from the methanol distillation unit.

3.4.1. Syngas Generation Unit

The figure in the succeeding page presents a simplified schematic diagram of the front-end section of the methanol plant or syngas generation unit.

The syngas generation unit mainly includes: (a.) feedstock pre-treatment, (b.) pre-reformer, (c.) primary steam reformer, (d.) auto-thermal reformer and (e.) final syngas cooler.

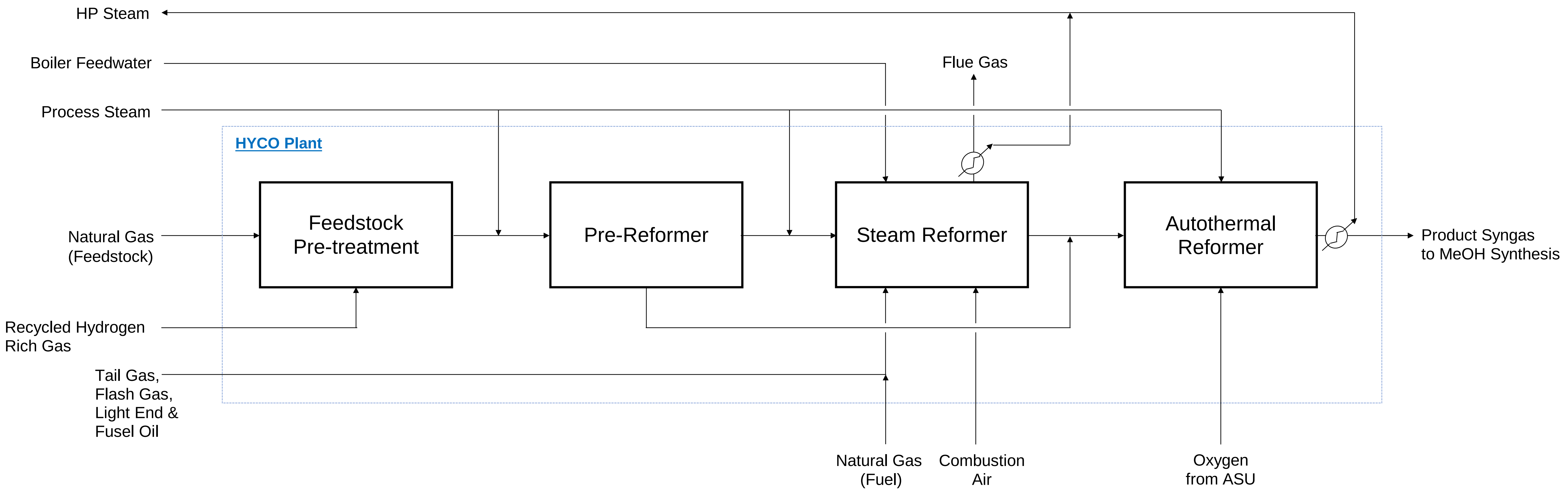
Natural gas from B.L. is used as feedstock and supplemental fuel to the SMR.

The NG is fed into the feedstock pre-treatment section to remove any sulphur compounds that could poison the reformer catalyst. From the B.L., the NG is let-down to 3.50 MPa and preheated in feed preheater. This is then mixed with a slipstream of hydrogen rich purge gas obtained from the downstream Hydrogen Recovery Unit (HRU); and the gas mixture is preheated in the "Feed Preheater Coil” to around 380°C.

The pre-heated feed is sent to the pre-treatment section which includes the hydrogenation reactor (converting any organic sulphur compounds in the feedstock into H_2S); and the sulphur removal unit (absorbing the H_2S using ZnO).

The treated feedstock is then mixed with a stream of MP steam (from the MP steam header) in order to maintain a fixed steam to carbon molar ratio. The resulting gas mixture is further heated in the “Pre-Reformer Feed Preheat Coil” to around 500^oC before entering the Pre-Reformer.

The Pre-Reformer mainly converts any heavy hydrocarbons (i.e. C2 and above) in the feed to CH_4, CO_2, CO or H_2. The outlet gas from the Pre-Reformer is then divided into two different streams: (a.) about 45% of the pre-reformed gas is mixed with second stream of superheated



SCHEMATIC BLOCK FLOW DIAGRAM	REVISION	DATE	BY	CHKD	APP.	SHEET
SYNGAS GENERATION SECTION (METHANOL PLANT)	FINAL	Feb-17	SS	GC	GC	1 of 3

MP steam to adjust the steam to carbon ratio (of the syngas produced by the primary SMR) and then further preheated in the "Reformer Preheat Coil" to around 600-650^oC, before being fed into the primary reformer; and (b.) the remaining portion of the pre-reformed gas is further pre-heated in the "ATR Heater Coil" and mixed with the syngas from the SMR before being fed into the ATR at around 670^oC.

The primary Steam Reformer provides most of the hydrogen in the product syngas. The main fuel for the SMR Furnace consists of the tail gas from the HRU, the flash gas from the methanol synthesis loop, the light end and fusel oil from the methanol distillation unit. This is supplemented by the NG from the B.L. The combustion air used by the SMR furnace is preheated by the flue gas in the "Combustion Air Preheater". HP steam, generated in the waste heat boiler downstream the ATR, is superheated by recovering heat in the convective section of the furnace. This is exported to the HP steam header at 12.0 MPa and 515^oC.

One of the main purpose of the ATR is to adjust the CO_2, CO and H_2 content of the product syngas by maintaining a methanol module number "$M = (H_2 – CO_2)/(CO_2 + CO)$" of about 2 in mole basis. The ATR also ensures that the methane slip is at minimum.

The oxygen required by the ATR is supplied by the air separation unit (ASU). This is preheated to around 225^oC and mixed with small amount of the MP steam (for cooling purpose) before being fed into the ATR burner. The final syngas product leaves the ATR at around 1030°C with a methane slip of around 0.2% mol (dry basis).

The product syngas is then cooled in a series of heat exchangers that includes:

- Waste Heater Boiler
- BFW Preheating
- Reboilers of the Methanol Distillation Unit

This is then mixed with the hydrogen recovered from the HRU and finally cooled by the cooling water to around 30^oC before being fed into the syngas compressor.

During the cooling of the syngas, the condensate is collected in separators. This is then pumped to the process condensate stripper. Process condensate recovered from the stripper is then fed into the condensate polishing unit.

3.4.2. Syngas Compressor

The Make-Up Gas or MUG, which consists of the cooled product syngas coming from ATR plus the hydrogen recovered from the HRU, is compressed to about 9.3 MPa in the syngas compressors.

A three stage centrifugal compressor with inter-cooling between the first and second stage is used, driven by a condensation steam turbine, with MP steam extraction, in single shaft arrangements using HP steam at 12.0 MPa and 515°C.

The last compressor stage acts as recirculator and receives the recycled gas coming from the methanol synthesis reactor. Typical flow rate of the recycled gas is about 4 to 5 times the flow rate of the MUG (in mole basis).

3.4.3. Methanol Synthesis Loop (Converter)

Figure presented in the succeeding page shows the simplified schematic block flow diagram of the production of crude methanol. This mainly consists of the (a.) methanol synthesis reactor or converter, (b.) feed-product heat exchanger, (c.) crude methanol cooler, (d.) crude methanol separator, and (e.) flash drum.

Majority of the feed gas (consisting of the recycled gas and MUG) that is fed into the converter is preheated in the Feed-Product Heat Exchanger. Whilst some of the feed gas is by-passed from the Feed-Product Heat Exchanger to control the reactor inlet temperature.

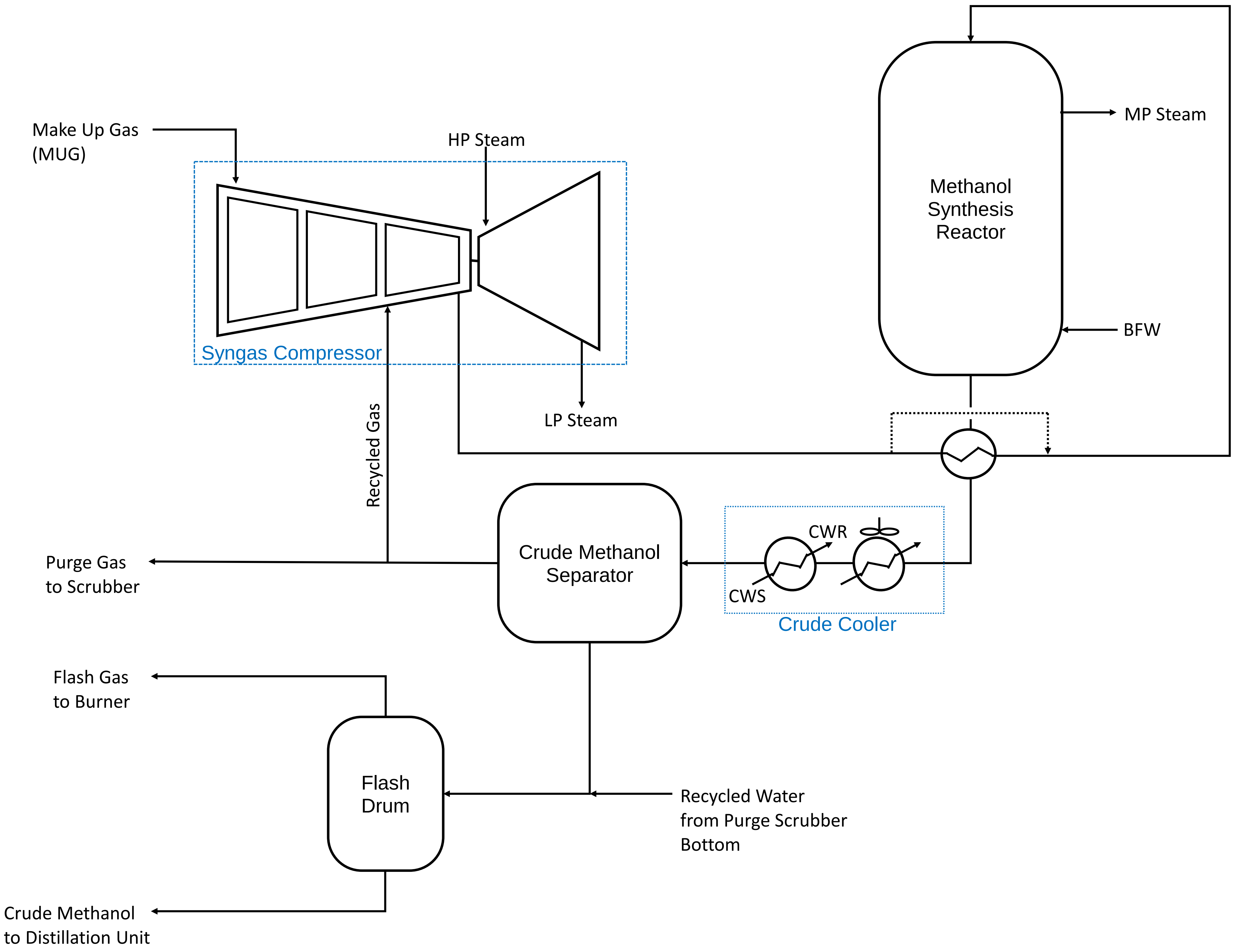
The methanol synthesis reactor (converter) considered is a typical steam raising pseudo-isothermal converter which consists of a pressure vessel with a heat exchanger (coil, plates or tube bundle) within the converter shell. The methanol synthesis catalyst is in this case contained in the shell side of the converter. The feed gas enters the converter and directly flows into these catalyst zones; where part of the syngas is converted to methanol (N.B. methanol production from the synthesis reaction is limited by its equilibrium reaction rates).

The methanol synthesis is exothermic. The core temperature of the reactor is kept fairly constant by steam generation producing saturated MP steam at 3.8 MPa. This is sent to the steam header to be used as process steam within the plant.

The product gas (containing the methanol, water, unreacted syngas and other by-products) exits the converter; and this is cooled in the Feed-Product Heat Exchanger and the Crude Methanol Cooler - consisting of an air cooler and a water cooled trim cooler, where the methanol and other condensable components in the product gas are condensed.

The gas and liquid mixture is then fed into a high pressure crude methanol separator.

The condensed liquid is then separated and collected from the bottom of the separator. This liquid is then fed into the Flash Drum (Crude Methanol Let-Down Separator) where the pressure is let down to about 0.6 MPa to release the dissolved gases in the liquid. The crude methanol collected from the flash drum is then sent to the methanol distillation unit via level control valve. The flash gas collected from the flash drum is released via pressure control valve and sent to the Reformer Fuel Gas Header to be used as fuel to the SMR burners.



SCHEMATIC BLOCK FLOW DIAGRAM	REVISION	DATE	BY	CHKD	APP.	SHEET
METHANOL SYNTHESIS LOOP	FINAL	Feb-17	SS	GC	GC	2 of 3

Most of the gas recovered from the crude methanol separator is recycled back to the Methanol Synthesis Loop. This is recirculated by sending it to the last stage syngas compressor. Whilst, a small part of the gas recovered from the separator is purged by using a suitable release valve; and this is sent to the purge gas scrubber to be washed. The water wash taken from scrubber bottom is collected and fed into the distillation section to recover any methanol in the wash liquid.

The purge gas coming from the scrubber is then split into two streams. The first stream (a small amount of the total purge gas flow) is sent to the Feedstock Pre-treatment of the Syngas Generation Unit to supply the required hydrogen for the hydrogenation reactor. The second stream (the bulk amount of the total purge gas flow) is sent to the Hydrogen Recovery Unit (HRU) where the hydrogen is recovered by using a membrane. The low pressure permeate (recovered hydrogen) is mixed with the cooled syngas from the ATR. The tail gas of the membrane is sent to the Reformer Fuel Gas Header to be used as fuel to the SMR burners.

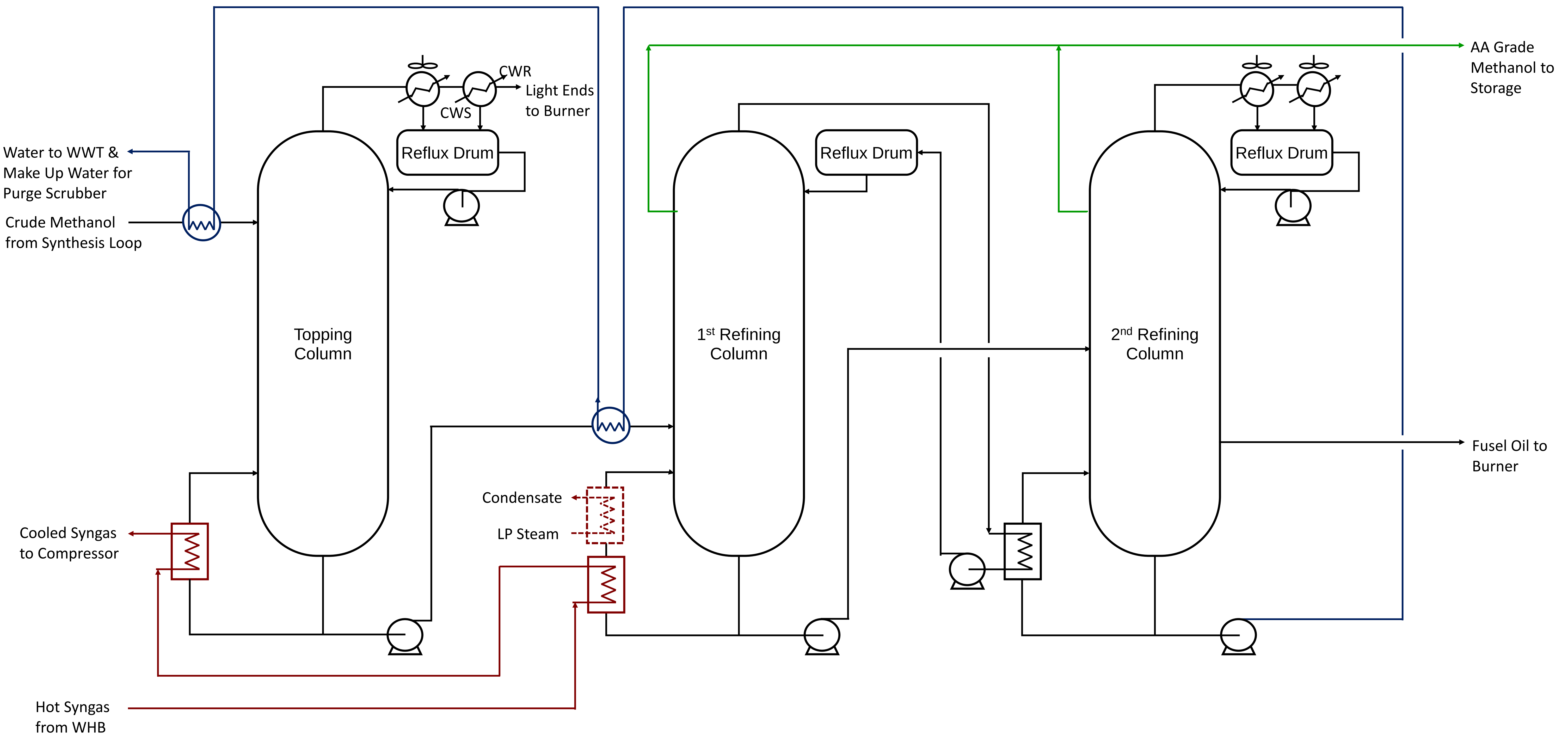
3.4.4. Methanol Distillation Unit

The crude methanol (mainly consists of 70-80% methanol, 20-30% water and other impurities) is purified in the distillation section to produce ASTM “AA grade” methanol. For this section, a three-column scheme is considered. This includes a Topping Column followed by two Refining Columns. A simplified schematic flow diagram is illustrated in the succeeding page

The crude methanol coming from the synthesis loop is pre-heated by the hot water recovered from the second refining column before being fed into the feed tray of the Topping Column. The column operates at slightly above atmospheric pressure with a bottom temperature of around 90°C. Heat is supplied to the column by a reboiler, using the hot syngas from the final syngas cooler (i.e. syngas cooled by the waste heat boiler and BFW preheater).

The lighter components (mainly dissolved gases - such as CO_2, CH_4; and other traces of by-products such as dimethyl ether, methyl formate, acetone) are separated as vapour that leaves the top of the column. The collected vapours is then cooled and partially condensed in an air cooled primary condenser and a water cooled secondary condenser. The condensed liquid collected are recycled back to the Topping Column as reflux. The gas (light end) collected from the condenser are sent to the Reformer Fuel Gas Header to be used as fuel for the SMR burners. The bottom product collected from the Topping Column is preheated and then fed into the first refining column.

The first refining column operates at around 0.8 MPa. Heat required by the distillation process is supplied mainly by the reboiler using hot syngas from the final syngas cooler as heating medium. However, to ensure full flexibility during operation, a reboiler using LP steam is also used as back up.



SCHEMATIC BLOCK FLOW DIAGRAM	REVISION	DATE	BY	CHKD	APP.	SHEET
METHANOL DISTILLATION UNIT	FINAL	Feb-17	SS	GC	GC	3 of 3

The second refining column operates at nearly atmospheric pressure. Heat required by the distillation process is supplied by the reboiler heated by condensing all the vapours collected from the top of the first refining column with the condensed liquid returned as reflux.

Pure liquid methanol is withdrawn from the top trays of both first and second refining columns. This is delivered to a dedicated storage tank, ready for sale as “AA Grade Methanol”.

Vapour leaving the top of the second refining column is fully condensed by air cooled condensers and then returned as reflux to the second refining column.

The heavier components (which mainly consist of ethanol and other alcohols produced as by-products during methanol synthesis) are withdrawn from the lower section (near the bottom) of the second refining column as Fusel Oil. This is cooled and then sent to the Reformer Fuel Gas Header to be used as fuel to the SMR burners.

Bottom product of the second refining column is mainly water with traces of methanol and other alcohols (in ppm level). This is cooled by using this stream as pre-heating medium to the feed streams of the first refining column and the topping column. The water recovered is then sent to the waste water treatment plant or recycled as make up water to the purge gas scrubber.

3.4.5. Steam and BFW System

The Steam and BFW System mainly includes the following sub-systems.

- Condensate Polishing Unit receives the process condensate collected from the syngas during cooling (taken from the process condensate stripper) and the steam condensates collected from the different steam headers, condensers of steam turbines (drivers to the different compressors, pumps, fans, et. al.) and reboiler of the stripping column (for CO_2 capture case).

- Deaeration System receives the condensates from the polishing unit and the make-up demineralized water from the demi plant. LP steam from the main LP steam header and the blowdown system is used as degassing agent. The deaerator vent is discharged to the atmosphere.

- HP Steam System includes the high pressure BFW pumps, HP steam drum, and superheated steam header. The pre-heating of the BFW and the HP steam generation and superheating are included in the syngas generation section (i.e. steam generation coil situated in the convective section of the SMR) and the syngas cooling section (i.e. waste heat boiler and BFW pre-heaters). HP steam from the steam header, at 12.0 MPa and 515°C, is fed to the steam turbine drivers of the syngas compressor and to the main air compressor/booster air compressor of the ASU. To balance the MP steam requirements of the plant, some of the MP steam is extracted from the steam turbine of the syngas compressor and is sent to the MP header.

- MP Steam System includes medium pressure BFW pumps, MP steam drum and superheated steam header. Saturated MP steam is generated from the heat exchanger coils within the methanol reactor. This is sent to a condensate stripper before being used as process steam to the syngas generation section. The balance of the MP steam required by the process is taken from the MP steam header. A small part of the MP steam is slightly superheated to 4.2MPa and 375°C. This is fed to various minor users such as steam turbine drivers for the combustion air blower, flue gas FD fan and different BFW pumps.

- LP Steam System includes the LP steam drum. Exhaust LP steam from the back pressure steam turbines (used as drivers to different compressors) are collected and sent to the LP steam header at 0.6 MPa. This is then distributed to the different LP consumers within the process sections.

For the case with CO_2 capture, it should be noted that additional LP steam is required for the regeneration of the solvent (i.e. used as heating medium for the stripper reboiler). Thus, some of the steam driven compressor should be changed to electrically driven compressor consequently increasing the plant electricity demand.

- Steam Condensate System, which includes the MP and LP headers and drums, recovers the condensate from the different steam users within the plant. The liquid condensates collected from the flash drums of steam headers; and the condensate collected from the condensers of the different steam turbine drivers are sent to the condensate polishing unit.

- Blowdown System, which includes the blowdown drum, collects all the blow down steam from the MP and LP steam headers. The LP steam recovered from the blowdown drum is sent to the deaerator as part of the degassing agent; whilst the liquid effluent collected is sent to the waste water treatment plant as effluent.

- Chemical Packages, include chemicals for pH control, oxygen scavenger used in conditioning of the BFW in the Deaerator System, and phosphate injection package used in all the steam systems.

3.4.6. *Demi Water Plant / Cooling Water System*

The demi-water required for the steam production is produced by processing raw water using reverse-osmosis system followed by an electro-deionization system. The plant includes a raw water tank, a demi water tank, relevant pumps, and a potable water package and storage.

Chemically treated demi water is also used as cooling water in a close circuit system (secondary system). This is mainly used for process coolers and for machinery cooling. The secondary cooling system is indirectly cooled by the sea water using plate heat exchangers.

Sea water in once through system (primary system) is used directly for the different steam turbine condensers and for the CO_2 compressor intercooler (for the CO_2 capture case).

3.4.7. *CO_2 Removal from the SMR Flue Gas using MEA Solvent (for Case 4B-2 only)*

The typical CO_2 Removal System consists of the flue gas quench cooler, CO_2 absorption section, heat exchanger network and the CO_2 stripper section. The simplified block flow diagram presented in Figure 1 (See Section 2.4.5) could be referred.

The flue gas of the primary SMR exiting the Combustion Air/Flue Gas Heat Exchanger with a temperature of about 135°C is initially cooled in a gas-gas heat exchanger by the CO_2 lean (decarbonised) flue gas coming from the top of the absorber column before leaving the flue gas stack.

The cooled flue gas is then fed into a quench scrubber using sodium hydroxide or sodium bicarbonate solution to reduce the SOx down to around 1 ppm thus minimising the solvent degradation. The flue gas is further cooled in a direct contact cooler using water wash.

The cooled treated flue gas is then fed to the bottom of the absorber. The column consists of two different level of pack beds and a water wash section.

The flue gas is contacted with a Semi-Lean MEA solvent in the first packed bed and then with a Lean MEA solvent in the second pack bed where the CO_2 in the flue gas are absorbed. To improve the efficiency of the absorption process, some of the heat of absorption is removed by using a water cooled pump around cooler situated in the middle section of the absorber.

The flue gas leaving the second pack bed is then scrubbed in the water wash section and passes through a demister section to remove any MEA and/or degradation by-products such as ammonia and any entrained mist. The CO_2 lean flue gas leaves the stack at around 90°C after being heated by the hot flue gas from the SMR.

The Rich MEA solvent leaves the bottom of the absorbers. This is divided into two different streams. The first Rich Amine stream is heated by the partially cooled Lean MEA coming from the stripper in a Rich-Lean Amine Heat Exchanger before being fed to the stripper. The second Rich Amine stream is sent to the flash drum to generate the Semi-Lean MEA solvent.

The Rich Amine stream going to the flash drum is first pre-heated by the Semi-Lean MEA coming from the flash drum and further heated by the Lean MEA solvent coming from the reboiler section of the stripper column.

Rich MEA solvent is regenerated in the stripping column. This consists of a stripping and a rectification section.

The vapour from the amine flash drum which mainly consists of the desorbed CO_2 and steam is fed to the top of the rectification section. The Rich MEA is fed into the lower end of the stripping section. This is heated by a vertical thermosyphon reboiler situated at the base of the stripping column. The reboiler is heated by the LP steam extracted from the Back Pressure Steam Turbine that drives the air compressor in ASU. The steam condensate collected is sent to the Condensate Polishing Unit to be processed and recycled back to the Steam and BFW System of the plant.

Periodically some of the circulating amine is sent to the reclaimer unit to remove any heat stable salts which are formed from the reaction of the trace impurities with the MEA. The heavy residues produced after every batch regeneration process are collected and then sent outside the B.L. for disposal. Fresh MEA from the amine storage tanks is added to replenish the lost solvent.

The overhead vapour from the column passes through a demister and sent to the column's condenser that is cooled by the sea water. The wet CO_2 is separated from the two phase mixture in a reflux drum or overhead accumulator. Some of the liquid (mainly water) is recycled back to the column as reflux and the balance are pumped to a water storage tank.

The wet CO_2 at 0.16MPa is compressed and dried to 11.0 MPa in the compression and dehydration unit as defined by the B.L. definition.

3.4.8. CO_2 Compression and Dehydration (for Case 4B-2 only)

The compression and dehydration unit includes the compressor, inter-stage coolers, knockout drums, dehydration package and liquid CO_2 pump.

Wet CO_2 from the stripper's condenser is compressed to 8.0 MPa by using a single train eight-stage centrifugal compressor and then finally pumped to 11.0 MPa ready for transport.

The CO_2 compressor is an integrally geared and electrically driven machine. The compression system includes the associated equipment for anti-surge control, vent, inter-stage coolers, knockout drums and condensate draining facilities.

At the discharge of each compressor stage, CO_2 is cooled by seawater in an inter-stage cooler. The condensate are separated from the compressed gas and collected in the knockout drum. This is then sent to the waste water treatment plant.

After the sixth compression stage, the compressed gas is sent to the dehydration package unit. Drying is achieved by using solid desiccants. For this case, molecular sieve is considered. Other solid desiccants such as Activated Alumina or Silica Gel could also be used.

The dehydration unit consists of two parallel trains of 2-Bed Adsorbers producing the dried CO_2 product with a dew point temperature of -40°C. In normal operation, one bed is used for drying, while the other water-saturated bed is regenerated by recycling a small part (ca.10%) of the dried product gas.

Final compression stages downstream of the dehydration unit increase the product CO_2 pressure up to 8.0 MPa. After being cooled, dried CO_2 product is in dense phase. This is then pumped and delivered the to the battery limits at the specified pipeline pressure of 11.0 MPa.

3.4.9. Air Separation Unit

The Air Separation Unit or ASU is treated in this study as a package unit supplied by specialised vendors to be included as part of the methanol plant package.

The ASU produces a low purity (98%) and medium pressure (2.56 MPa) oxygen to be used by the ATR. High purity nitrogen is also produced to provide inert stream for plant start-up and as nitrogen blanket for the liquid methanol storage.

The ASU is based on cryogenic distillation of atmospheric air. This primarily consists of the main air compressor (MAC), pre-cooling and purification of ambient air, booster compressor (BAC) and the cold box which include the cold production section and the distillation section based on a low and high pressure columns configuration.

A steam turbine is used to drive the main air compressor and booster compressor in single shaft arrangement. HP steam at 12.0 MPa and 515°C is used and obtained from the HP header as supplied by the syngas generation unit and syngas cooling.

The liquid oxygen produced from the cold box is pumped to the required delivery pressure (2.56MPa) before being evaporated. This is then preheated by a steam coil to 225°C, and mixed with small amount of MP steam as required by the ATR burner.

3.4.10. Balance of Plant (BOP)

The operation of the whole unit is supported by additional utilities and facilities such as:

- Instrument/Plant Air System
- Flare System
- Drain System
- Interconnecting
- Buildings (Control Room, Electrical Sub-Station, Laboratories).

3.5. Heat and Mass Balance

The Heat and Mass Balances of the methanol production without and with CCS are presented in the succeeding pages. The information should be referred to the Block Flow Diagram presented in Section 3.3.

The table below shows the breakdown of the CO_2 balance of the methanol production plant.

	EQUIVALENT FLOW OF CO_2 kmol/h	
	Ammonia/Urea Plant w/o CCS	Ammonia/Urea Plant w/ CCS
INPUT STREAMS		
Natural gas feedstock	7168	7168
Natural gas fuel	1030	1030
TOTAL (IN)	**8199**	**8199**
OUTPUT STREAMS		
Total carbon in methanol product	6503	6503
Captured CO_2 to storage	-	1505
SUB-TOTAL (Carbon Not Emitted)	**6503**	**8008**
Flue gas to stack (A)	1673	167
Vents	23	23
Emission	1696	190
TOTAL (OUT)	**8199**	**8199**
Equivalent CO_2 in Urea product (%)	79.3%	79.3%
Capture Rate - Capture from Flue Gas (%)	-	90.0%
Captured CO_2 to Storage (%)	-	18.4%
AMOUNT OF CARBON NOT EMITTED (%)	**79.3%**	**97.7%**

Indirect CO_2 Emission* (B)	146 - 313	287 - 616
TOTAL CO_2 EMISSION - (A) + (B)	**1842 - 2009**	**477 - 806**

(*) "Indirect CO_2 emission" is the specific CO_2 emission related to the power imported (from the grid) to the methanol production complex. For this study, the specific emissions of 348 kg/MWh and 746 kg/MWh from NG and coal fired power plant are used respectively.

3.6. Plant Performance Data

The table below summarizes the productions/consumptions and CO_2 emissions relevant to the methanol plant (with and without capture).

Plant Performance Data		**Base Case**	**CCS Case**
INLET STREAMS			
Natural Gas (as Feedstock)	t/h	119.097	119.097
Natural Gas (as Fuel)	t/h	17.119	17.119
Natural Gas (Total Consumption)	**t/h**	**136.216**	**136.216**
Natural Gas LHV	MJ/kg	46.50	46.50
Total Energy Input	MW	1,759.53	1,759.53
OUTLET STREAMS			
Methanol Product to BL	t/d	5,000	5,000
	t/h	208.36	208.36
Methanol LHV	MJ/kg	20.094	20.094
Total Energy in the Methanol Product	**MW**	**855.4**	**900.8**
POWER BALANCE			
Methanol Production	MWe	-11.150	-20.295
Steam + BFW System	MWe	-2.920	-2.920
Utilities + BoP	MWe	-4.400	-6.250
CO2 Capture Plant	MWe	NA	-1.655
CO2 Compression and Dehydration Unit	MWe	NA	-5.200
Imported Power from the Grid	MWe	18.470	36.320
SPECIFIC CONSUMPTIONS			
Natural Gas (as Feedstock)	GJ/t MeOH	26.579	26.579
Natural Gas (as Fuel)	GJ/t MeOH	3.820	3.820
Feed + Fuel	**GJ/t MeOH**	**30.399**	**30.399**
SPECIFIC EMISSIONS			
Specific CO_2 Emission (Direct)	t/t MeOH	0.3531	0.0354
Specific CO_2 Captured	t/t MeOH	NA	0.3179
Equivalent CO_2 in the Methanol Product		79.3%	79.3%
Equivalent CO_2 Captured to Storage		NA	18.4%

3.7. Preliminary Utility Consumption

This section presents the main utilities consumption of the different processes for the methanol plant without and with CCS.

FOSTER WHEELER

ESTIMATED UTILITY CONSUMPTIONS

CUSTOMER NAME:	IEAGHG	**MeOH Plant without CO2 Capture**	REV.	REV. 0	REV. 1	REV. 2	SHEET
PROJECT NAME:	TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE		BY	NF			1
FWI CONTRACT:	1BD0840 A		CHKD	GC			OF
LOCATION:	THE NETHERLAND		DATE	June 15			1

	ELECTRIC POWER		STEAM t / h			BFW	EFFLUENT (3)	LOSSES (2)	DMW (1)	RAW WATER	COOLING WATER		SEA WATER		FUEL	INSTR. AIR	Nitrogen
	LOAD BHP	kW	LP	MP	HP	t/h	t/h	t/h	t/h	t/h	ΔT (°C)	m^3/hr	ΔT (°C)	m^3/hr	MMKcal/h	Nm^3/h	Nm^3/h
MeOH PLANT		11.150	22,2	301,7	175,9	635,8					11	6.633	7	20.121	190,2	200	(250)
				-147,1	-488,7		-36,8	-7,7	-455,3								
STEAM AND BFW PLANT		2.920			312,8				645,8				7	4.441			
			-22,2	-154,7		-635,8	-7,7	-4,7	-133,6								
UTILITIES / BoP		4.400								73,9	11	-6.633	7	10.432	0,5	100	(250)
							-17,1		-56,9 (1)							-300	(-500)
TOTAL		18.470	0	0	0	0	-61,6	-12,4	0	73,9	-	0	-	34.994	190,7	0	0

NOTES:

(1) DMW is the sum of DMW plus condensate from the process unit

(2) Losses includes net water consumptions in the reactions and steam/BFW system vent and steam losses

(3) Water effluent (to be sent to WWT) includes demi plant eluate, waste water from MeOH distillation section and steam system blowdown

ESTIMATED UTILITY CONSUMPTIONS

CUSTOMER NAME:	IEAGHG	**MeOH Plant with CO2 Capture from SMR flue gas**	REV.	REV. 0	REV. 1	REV. 2	SHEET
PROJECT NAME:	TECHNO-ECONOMIC EVALUATION OF H2 PRODUCTION WITH CO2 CAPTURE		BY	NF			1
FWI CONTRACT:	1BD0840 A		CHKD	GC			OF
LOCATION:	THE NETHERLAND		DATE	June 15			1

	ELECTRIC POWER		STEAM t / h			BFW	EFFLUENT (3)	LOSSES (2)	DMW (1)	RAW WATER	COOLING WATER		SEA WATER		FUEL	INSTR. AIR	Nitrogen
	LOAD BHP	kW	LP	MP	HP	t/h	t/h	t/h	t/h	t/h	ΔT (°C)	m^3/hr	ΔT (°C)	m^3/hr	MMKcal/h	Nm^3/h	Nm^3/h
MeOH PLANT		20.295	22,2	301,7	175,9	635,8					11	6.633	7	20.121	190,2	200,0	(250)
				-147,1	-488,7		-36,8	-7,7	-455,3							-300	(-500)
STEAM AND BFW PLANT		2.920			312,8				645,8				7	4.441		100	(250)
			-97,5	-154,7		-635,8	-7,7	-4,7	-58,3								
CO2 CAPTURE		1.655	75,3								11	4.638					
									-75,3	-39,2							
CO2 COMPRESSION		5.200									11	63	7	1.356			
UTILITIES / BoP		6.250								73,9	11,00	-11.334	7	17.826	0,5		
							-17,1		-56,9 (1)								
TOTAL		36.320	0	0	0	0	-61,6	-12,4	0	34,7	-	0	-	43.744	190,7	0	0

NOTES:

(1) DMW is the sum of DMW plus condensate from the process unit

(2) Losses includes net water consumptions in the reactions and steam/BFW system vent and steam losses

(3) Water effluent (to be sent to WWT) includes demi plant eluate, waste water from MeOH distillation section and steam system blowdown

3.8. Economic Evaluation

The purpose of this section is to present the results of the economic analysis carried out to evaluate the Levelized Cost of Methanol (LCOMeOH) and the CO_2 Avoidance Cost (CAC) for the study cases of methanol production without and with CCS.

The capital cost and annual operating & maintenance (O&M) costs for the different cases without and with CCS have been evaluated. These are presented in this section of the report, along with the results of the financial model.

All the general assumptions used in performing the analysis are described in the techno-economic criteria and methodology reported in the Reference Document (Annex I). Only the information and assumptions specific to the methanol plant are presented in this report.

3.8.1. *Investment Cost Estimate*

The estimates for the Total Plant Cost (TPC) and Total Capital Requirement (TCR) for the methanol plant without and with CCS are summarised in Table 5. The methodologies used in estimating capital cost are outlined and briefly described in the Reference Document (Annex I).

The **Total Capital Requirement** (**TCR**) as defined in Section 2.8 is the sum of: Total Plant Cost (TPC), spare parts cost, start-up costs, owner's costs, interest during construction, and working capital.

The **Total Plant Cost** (**TPC**) is estimated using the cost database of the Amec Foster Wheeler based on previous projects undertaken and pro-rated to the capacity the different cases studies evaluated. The TPC of the different study cases are further broken down into the cost estimates of the different main process units:

- Base Case (Case 4B-1): Methanol Plant w/o CO_2 Capture
    - Methanol Plant (including ASU and steam/BFW system)
    - Methanol Storage
    - Other Utilities and Balance of Plant (BoP)

- CCS Case (Case 4B-2): Methanol Plant w/ CO_2 Capture from SMR's flue gas
    - Methanol Plant (including ASU and steam/BFW system)
    - Methanol Storage
    - CO_2 Capture Plant
    - CO_2 Compression and Dehydration Unit
    - Other Utilities and Balance of Plant (BoP)

In summary, the assumptions used in estimating the other components of the TCR are as follows:

- Spare parts cost

  0.5% of the TPC is assumed to cover the spare part costs. It is also assumed that spare parts have no value at the end of the plant life due to obsolescence.

- Start-up costs consist of:

  ⇒ 2% of TPC, to cover any modifications to equipment that needed to bring the unit up to full capacity.
  ⇒ 25% of the full capacity feedstock and fuel cost for one month, to cover inefficient operation that occurs during the start-up period.
  ⇒ Three months of operating labour and maintenance labour costs, to include training.
  ⇒ One month of chemicals, catalyst and waste disposal costs and maintenance materials costs.

- Owner's cost

  7% of the TPC is assumed to cover the Owner's cost and fees. This is assumed to be incurred in the first year of construction, allowing for the fact that some of the costs would be incurred before the start of construction.

- Interests during construction

  This is calculated from the plant construction schedule and the interest rate is assumed to be the same as the discount rate.

- Working capital:

  The working capital includes inventories of fuel and chemicals (materials held in storage outside of the process plants). Storage for 30 days at full load is considered for chemicals and consumables. It is assumed that the cost of these materials are recovered at the end of the plant life.

Table 5. **TPC and TCR of the methanol plant study cases**

		Methanol Plant w/o CCS	Methanol Plant w/ CCS
Methanol plant (incl. ASU & steam/BFW system)	M€	600	600
Methanol storage	M€	8.7	8.7
Utilities and BoP	M€	85.0	92.8
CO_2 capture unit	M€	-	88.1
CO_2 compression and dehydration unit	M€	-	20.5
Total Installed Cost (TIC)	**M€**	**693.7**	**810.1**
Contingency	-	20%	20%
Total Plant Cost (TPC)	**M€**	**832.4**	**972.1**
Total Capital Requirement (TCR)	**M€**	**1082.3**	**1264.6**

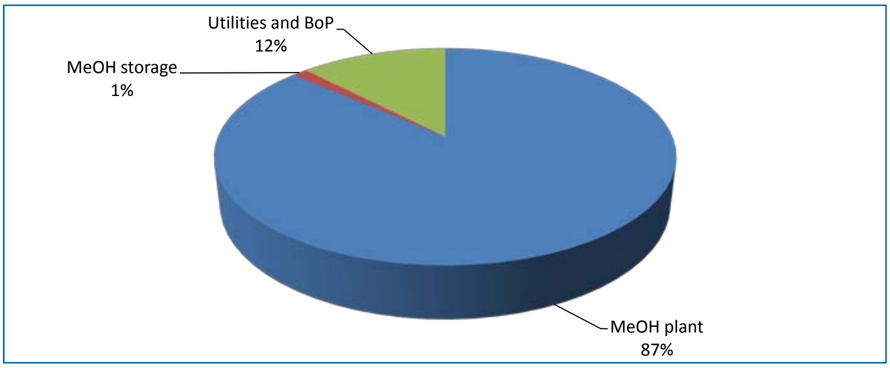


Figure 19. **Methanol plant without CCS: process units cost breakdown**

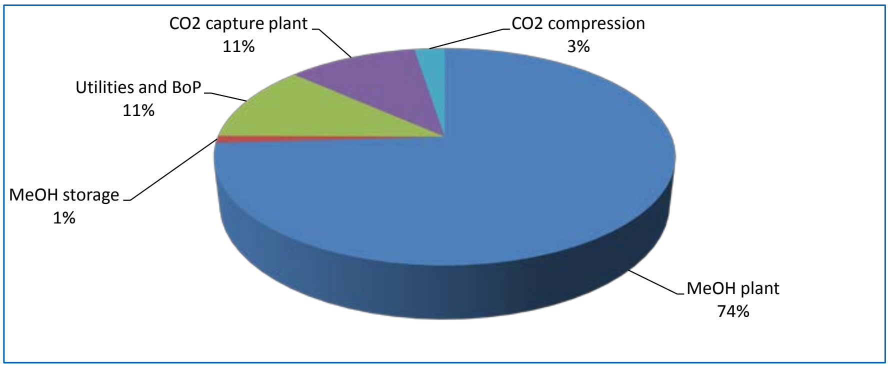


Figure 20. **Methanol plant with CCS: process units cost breakdown**

3.8.2. *Annual Operating and Maintenance Cost*

The Annual Operating and Maintenance (O&M) costs consists of the variable cost and the fixed cost. The assumptions used in these cost estimates are briefly described in Annex I – Reference Document.

Variable Cost

Table 6 summarised the elements of the variable cost for the methanol plants which consists of the following items:

- Fuel (natural gas)
- Raw water make-up
- Electricity (imported from grid)
- Catalysts and chemicals.

The consumption of the various items and the corresponding costs are estimated on yearly basis using the heat and mass balance reported in Section 3.5 using an expected availability of 90% of the methanol plant for year 2 to 25, and 70% of availability during year 1.

Table 6. Yearly variable costs

	Yearly Variable Cost			
		Revision:	0	
		Date:	Jan. 2016	
		Issued By:	NF	
		Approved By:	GC	

Yearly Operating Hours	7884		Methanol Plant without CCS			Methanol Plant with CCS*		
Consummables		**Unit Cost**	**Consumption**		**Operating Cost**	**Consumption**		**Operating Cost**
	[um]	€/[um]	Hourly [um]/h	Yearly [um]/y	€/y	Hourly [um]/h	Yearly [um]/y	€/y
Feedstock & Fuel								
Natural Gas	tonne	279.0	136.2	1,073,927	299,638,487	136.2	1,073,927	299,638,487
Auxiliary Feedstock								
Raw make-up water	m3	0.20	56.9	448,600	89,720	34.7	273,575	54,715
Electricity (from the grid)	MWh	80.0	18.5	145,617	11,649,360	36.3	286,347	22,907,760
Chemicals		-	-	-	200,000	-	-	200,000
Catalysts		-	-	-	3,800,000	-	-	3,800,000
TOTAL YEARLY OPERATING COST	**€/y**				**315,377,567**			**326,600,962**

* This case involves the capture of CO2 from the SMR's flue gas.

The reference values for the price of the natural gas and other main consumables are summarized in the table below.

Item	Unit	Cost
Natural gas	€/GJ (LHV)	6
Raw water	€/m^3	0.2
Electricity	€/MWh	80
CO_2 transport and storage	€/t CO_2 stored	10
CO_2 emission cost	€/t CO_2 emitted	0

Fixed Cost

The fixed cost mainly include:

- Operating Labour Cost
- Overhead Charges
- Maintenance Cost
- Other Fixed Cost.

The yearly cost of the direct labour is calculated assuming for each individual an average cost equal to 60,000 Euro/year, referred to year 2014. The tables presented in the succeeding page presents the line-up of the labour force for the ammonia/urea plant without and with CCS.

The overhead charges are equal to 30% of the operating labour and maintenance labour cost (as described below).

The annual maintenance cost of the plant is estimated as 1.5% of the Total Plant Cost for each case; and these could be sub-divided as maintenance labour and maintenance materials. A maintenance labour to materials ratio of 40:60 can be statistically considered for this breakdown.

The other fixed cost includes local taxes and fees, and insurance cost. This study assumed that the other fixed cost could be covered by 1% of the TPC.

Methanol plant w/o capture				
	SR + utilities	**Methanol plant**	**TOTAL**	**Notes**
OPERATION				
Area Responsible	1		1	daily position
Assistant Area Responsible	1		1	daily position
Shift Superintendent	5		5	1 position per shift
Electrical Assistant	5		5	1 position per shift
Shift Supervisor	5	5	10	2 positions per shift
Control Room Operator	5	5	10	2 positions per shift
Field Operator	5	5	10	2 positions per shift
Subtotal			**42**	
MAINTENANCE				
Mechanical group	3	1	4	daily position
Instrument group	3	1	4	daily position
Electrical group	3	1	4	daily position
Subtotal			**12**	
LABORATORY				
Superintendent+Analysts	2		2	daily position
Subtotal			**2**	
TOTAL			**56**	

Cost for personnel

Yearly individual average cost =	60,000	Euro/year
Total cost =	3,360,000	Euro/year

Methanol plant with capture					
	SR + utilities	**CO2 capture**	**Methanol plant**	**TOTAL**	**Notes**
OPERATION					
Area Responsible	1			1	daily position
Assistant Area Responsible	1			1	daily position
Shift Superintendent	5			5	1 position per shift
Electrical Assistant	5			5	1 position per shift
Shift Supervisor	5		5	10	2 position per shift
Control Room Operator	5	5	5	15	3 positions per shift
Field Operator	5		5	10	2 positions per shift
Subtotal				**47**	
MAINTENANCE					
Mechanical group	3		1	4	daily position
Instrument group	3		1	4	daily position
Electrical group	3		1	4	daily position
Subtotal				**12**	
LABORATORY					
Superintendent+Analysts	2			2	daily position
Subtotal				**2**	
TOTAL				**61**	

Cost for personnel

Yearly individual average cost =	60,000	Euro/year
Total cost =	3,660,000	Euro/year

Annual Maintenance Cost (2014)			
	Maintenance % of TPC	TPC €	Annual Cost €/y
Methanol Production w/o CCS	1.5%	832,440,000	12,486,600
Methanol Production w/ CCS	1.5%	972,120,000	14,581,800

The following table summarised the annual O&M cost for the two cases.

FOSTER WHEELER	Revision	0	1
	Date	Jan-16	Jan-17
	Issued by:	NF	SS
	Approved by:	GC	SS

ANNUAL O&M COST

	Methanol Plant w/o CCS €/year	Methanol Plant w/ CCS €/year
Fixed Costs		
Direct labour	3,360,000	3,660,000
Adm./gen. overheads	2,506,392	2,847,816
Insurance & local taxes	8,324,400	9,721,200
Maintenance	12,486,600	14,581,800
Sub-total	26,677,392	30,810,816
Variable Costs (Availability - 90%)		
Feedstock & fuel	299,638,487	299,638,487
Electricity (Import from Grid)	11,649,360	22,907,760
Raw water (make-up)	89,720	54,715
Chemicals & catalysts	4,000,000	4,000,000
Sub-total	315,377,567	326,600,962
Total Fixed & Variable Cost	**342,054,959**	**357,411,778**
Other Cost		
CO2 Transport & Storage	-	5,222,519
Annual O&M Cost	**342,054,959**	**362,634,297**

3.8.3. *Levelized Cost of Methanol and CO_2 emissions avoidance cost*

The Levelised Cost of Methanol (LCOMeOH) is defined as the selling price at which methanol must be produced to achieve break even by the end of its economic lifetime.

In other word, the present value of all the revenues obtained from selling methanol should be equal to the present value of all costs of building, maintaining and operating the plant over its lifetime.

The selling price of the methanol is calculated based on the assumption that NPV = 0 (over the whole life time of the plant).

Also, it should be noted that in this type of analysis, the assumptions for the long-term inflation and the price/cost variations throughout the project life-time are not considered.

3.8.4. *CO_2 Avoidance Cost (CAC)*

The CO_2 Avoidance Cost (CAC) is calculated by comparing the costs and specific emissions of the methanol plant with CCS with those of the reference case without CCS based on the following definition:

$$CO_2 \text{ Avoidance Cost (CAC)} = \frac{\text{LCOMeOH}_{\text{CCS}} - \text{LCOMeOH}_{\text{Reference}}}{CO_2\text{Emissions}_{\text{Reference}} - CO_2\text{Emissions}_{\text{CCS}}}$$

where:

- LCOMeOH is expressed in € per tonne of MeOH
- CO_2 emissions is expressed in tonne of CO_2 per tonne of MeOH
- Cost of CO_2 avoidance is expressed in € per tonne of CO_2

Given that the methanol production evaluated in this study imports some amount of electricity from the grid as part of their production process, it is therefore essential to take into account the CO_2 emissions associated to the imported electricity.

For simplification, this study assumes the following specific emissions (i.e. from NG and coal fired power plant) associated to the imported electricity consumed by the plant. It should be emphasised that this only provides an indicative range of values related to the CO_2 avoidance cost.

Gas fired power plant without CO_2 capture	348	kg CO_2 / MWh
Coal fired power plant without CO_2 capture	746	kg CO_2 / MWh

For the reference case without CCS, the indirect CO_2 emission corresponding to the electricity imported from a power plant without CCS (coming from either natural gas or coal fired plant), are summarised below:

Methanol Production w/o CCS		
Direct specific CO_2 emission	t/t MeOH	0.3531

Indirect CO_2 emission based on a NGCC power plant		
Electric power import	MWh	18.47
Specific CO_2 emission	kg/MWh	348
Indirect CO_2 emission	t/h	6.43
Indirect specific CO_2 emission	t/t MeOH	0.0308

Indirect CO_2 emission based on a coal fired power plant		
Electric power import	MWh	18.47
Specific CO_2 emission	kg/MWh	746
Indirect CO_2 emission	t/h	13.78
Indirect specific CO_2 emission	t/t MeOH	0.0661

For the case with CCS, the indirect CO_2 emission corresponding to the electricity imported from a power plant without CCS (coming from either natural gas or coal fired power plant) are also considered and these are summarised below:

Methanol Production w/ CCS		
Direct specific CO_2 emission	t/t MeOH	0.0354

Indirect CO_2 emission based on a NGCC power plant		
Electric power import	MWh	36.32
Specific CO_2 emission	kg/MWh	348
Indirect CO_2 emission	t/h	12.64
Indirect specific CO_2 emission	t/t MeOH	0.0607

Indirect CO_2 emission based on a coal fired power plant		
Electric power import	MWh	36.32
Specific CO_2 emission	kg/MWh	746
Indirect CO_2 emission	t/h	27.10
Indirect specific CO_2 emission	t/t MeOH	0.1300

3.8.5. *Macroeconomic Bases*

The economic assumptions and macroeconomic bases are reported in the reference document of the report. The principal financial bases assumed for the financial modelling are reported also hereafter for reader's convenience:

ITEM	DATA
Discount Rate	8%
Capacity factor	90%
Inflation Rate	Constant Euro
Currency	Euro reported in 4Q2014

3.8.6. *Financial analysis results*

This section summarizes the results of the financial analysis performed for methanol plant study cases, based on the input data reported above.

Table 7 present the Levelised Cost of Methanol (LCOMeOH) and CO_2 avoidance cost (CAC) for the methanol plant without and with CCS.

Figure 21 presents the breakdown of the LCOMeOH with respect to the relative weight of the following cost elements:

- Capital investment,
- Fixed O&M,
- Variable O&M,
- Fuel,
- CO_2 transportation & storage.

Table 7. LCOMeOH and CAC of the methanol plant study cases

Description	Levelized Cost of Urea €/t	CO_2 emission avoidance cost €/t
Methanol plant w/o CCS	275.1	-
Methanol plant with CCS* (Indirect emission from NG power plant w/o CCS)	298.9	82.6
Methanol plant with CCS* (Indirect emission from coal power plant w/o CCS)	298.9	93.7

* This case involves the additional CO_2 capture from SMR's flue gas.

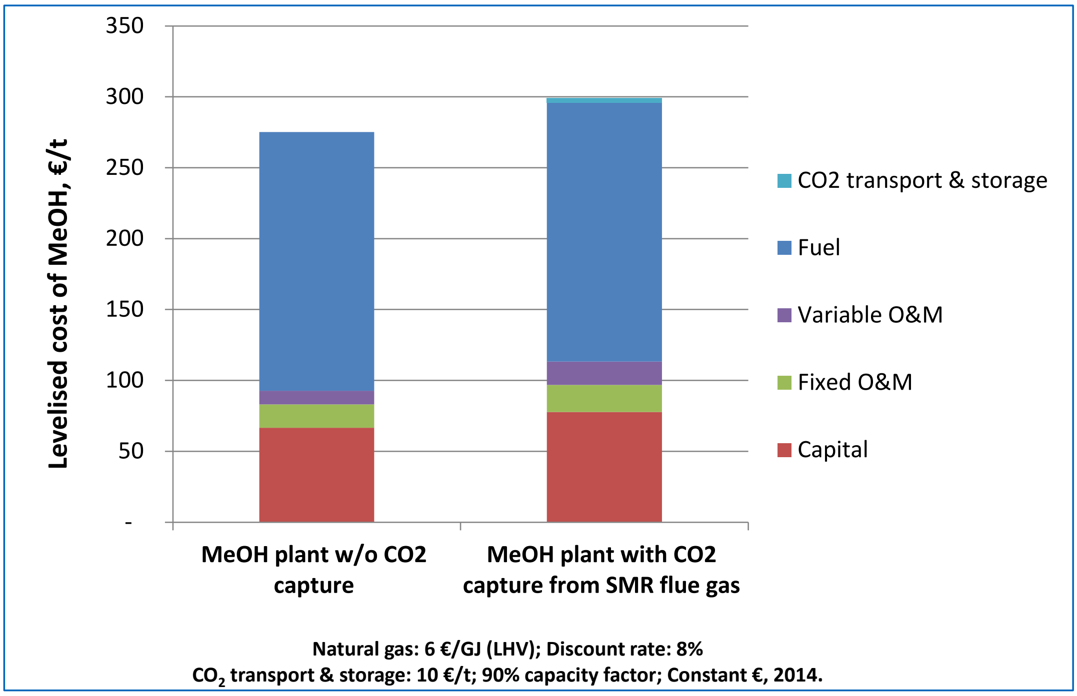


Figure 21. **Levelized Cost of Methanol**

3.8.7. Sensitivity Analysis

Due to the possible variability of some economic input data, sensitivity analysis is also performed on key parameters such as:

- Natural gas price,
- Electricity price,
- Discount rate,
- CO_2 emission cost
- CO_2 transport & storage cost

The sensitivity range is summarised in the following table.

Sensitivity Relevant to Methanol Study Cases			
Criteria	Unit	Base Case	Sensitivity Range
Natural gas price	€/GJ (LHV)	6	2 to 18
Electricity price	€/MWh	80	20 to 100
Discount rate	%	8	4 to 12
CO_2 transport & storage	€/t stored	10	-20 to 40
CO_2 emission costs	€/t emitted	0	0 to 100

Sensitivity to Natural Gas Price (with Constant Electricity Price)

Figures 22 and 23 present the sensitivity of LCOMeOH and CAC to the NG price. In these analysis, the electricity price is assumed constant at 80€/MWh (without any distinction between cost of electricity produced from NG or coal fired power plant – i.e. scenario using wholesale price electricity).

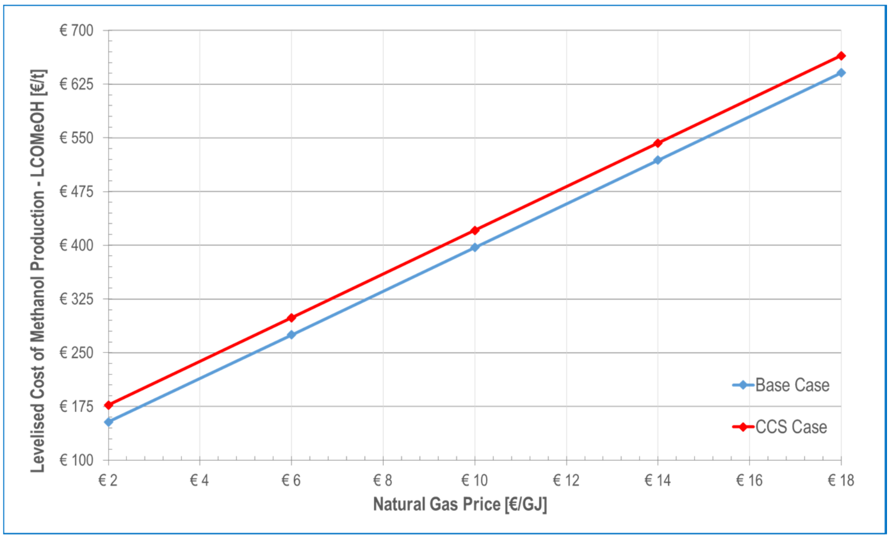


Figure 22. Sensitivity of Levelised Cost of MeOH to Natural Gas Price
(at constant electricity price of 80 €/MWh)

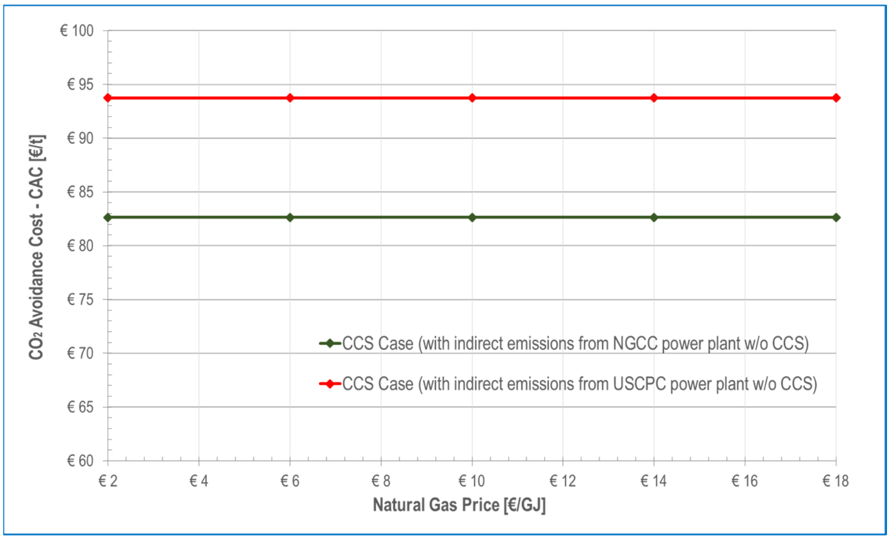


Figure 23. Sensitivity of CO_2 Avoidance Cost to Natural Gas Price
(at constant electricity price of 80 €/MWh)

Sensitivity to Electricity Price (with Constant NG Price)

Figures 24 and 25 present the sensitivity of LCOMeOH and CAC to the electricity price. In these analysis, the natural gas price is assumed constant at 6€/GJ (i.e. electricity price is not indexed to the NG price).

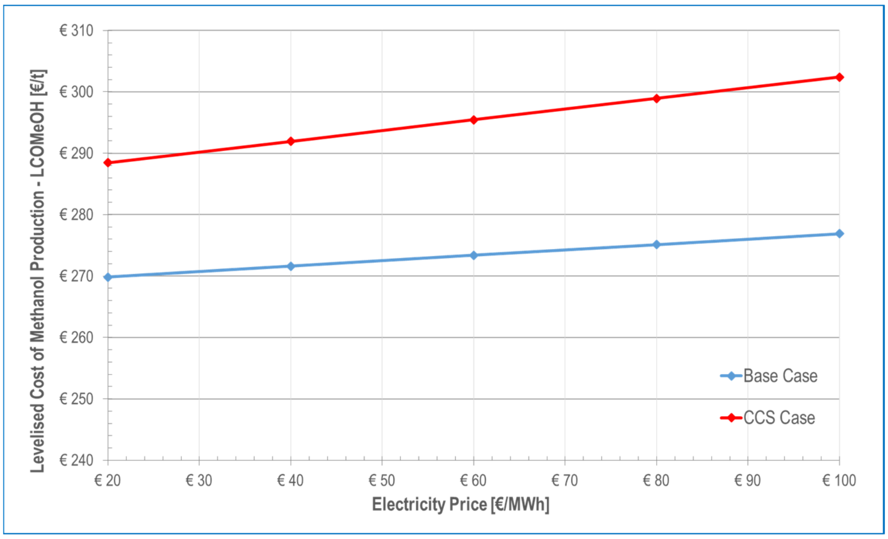


Figure 24. **Sensitivity of Levelised Cost of MeOH to Electricity Price**
(at constant natural gas price of 6 €/GJ)

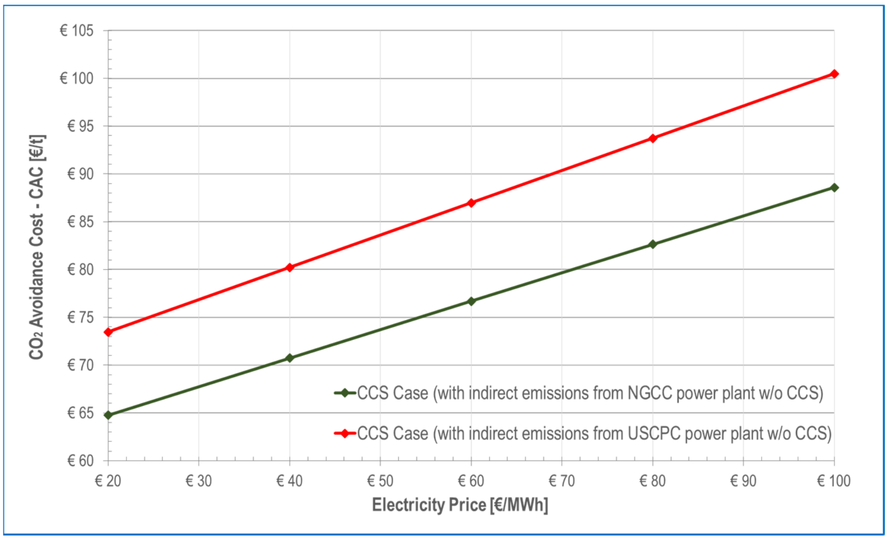


Figure 25. **Sensitivity of CO_2 Avoidance Cost to Electricity Price**
(at constant natural gas price of 6 €/GJ)

Sensitivity to Natural Gas and Electricity Price Based on the LCOE of the Power Plant

Figures 26 and 27 present the sensitivity of LCOMeOH and CAC to the natural gas and electricity price. In these analysis, the electricity price is assumed to be equal to the LCOE of a NG or coal fired power plant (i.e. over the fence contract scenario) instead of assuming a constant electricity price (i.e. wholesale market scenario) as presented previously. The LCOE from the NG and coal fired power plants without CCS with respect to the price of natural gas are presented in Figure 9 (See Section 2.8.7).

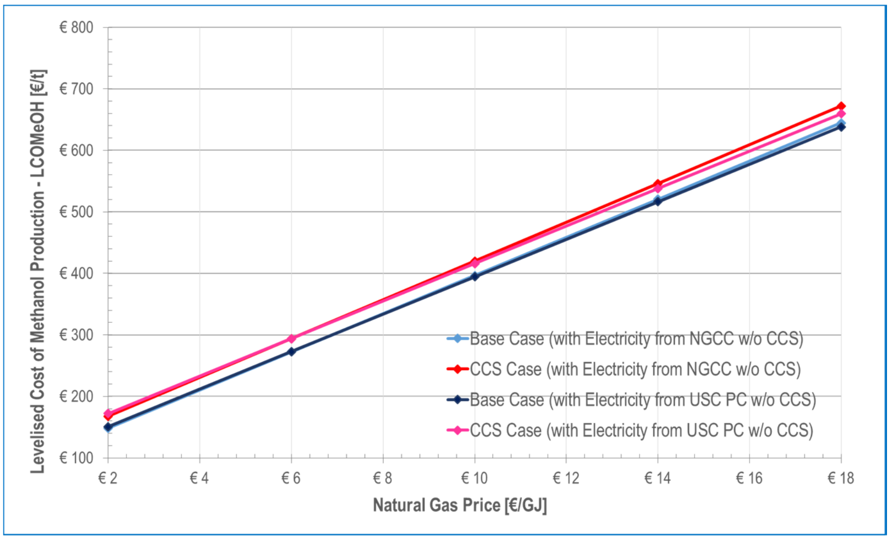


Figure 26. Sensitivity of Levelised Cost of MeOH to Natural Gas Price
(with electricity price based on LCOE presented in Figure 8)

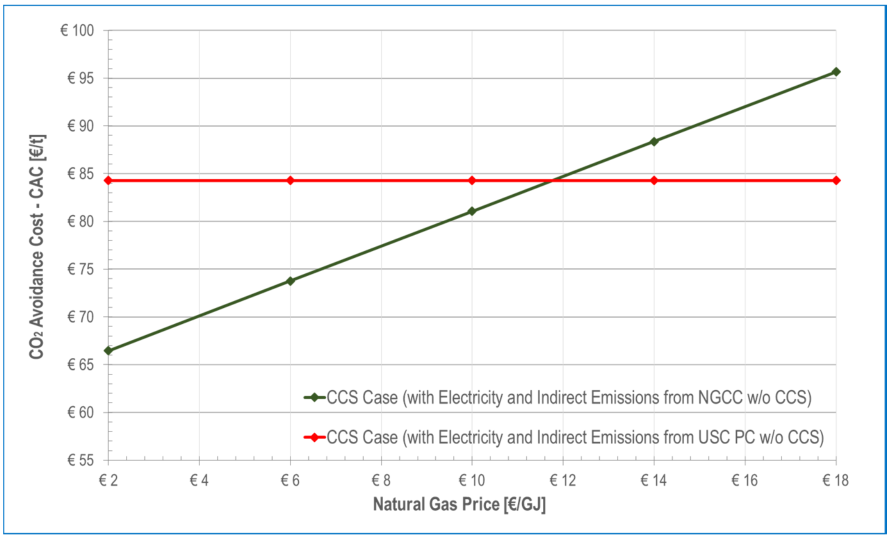


Figure 27. Sensitivity of CO_2 Avoidance Cost to Natural Gas Price
(with electricity price based on LCOE presented in Figure 8)

Sensitivity to Discount Rate

Figures 28 and 29 present the sensitivity of LCOMeOH and CAC to the discount rate. In these analysis, the natural gas and electricity price are assumed constant at 6 €/GJ and 80 €/MWh respectively.

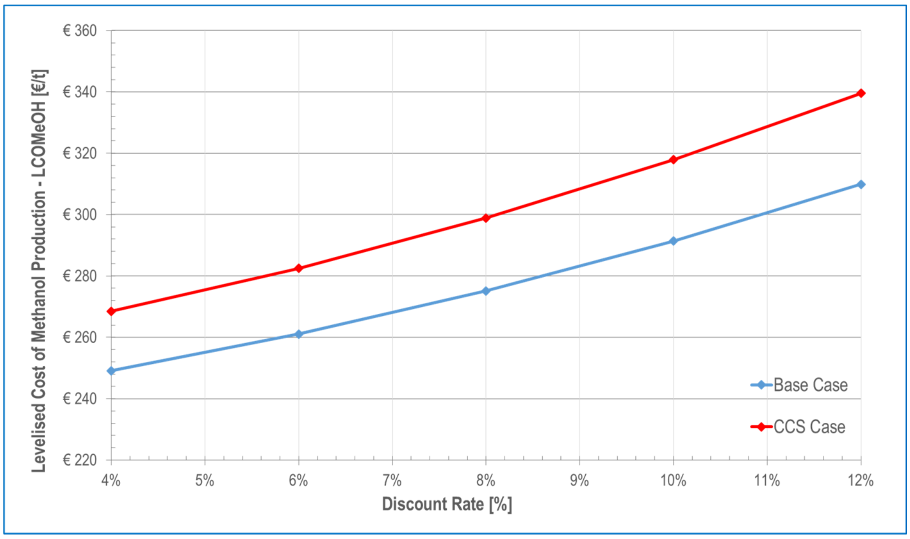


Figure 28. **Sensitivity of Levelised Cost of Methanol to the Discount Rate**

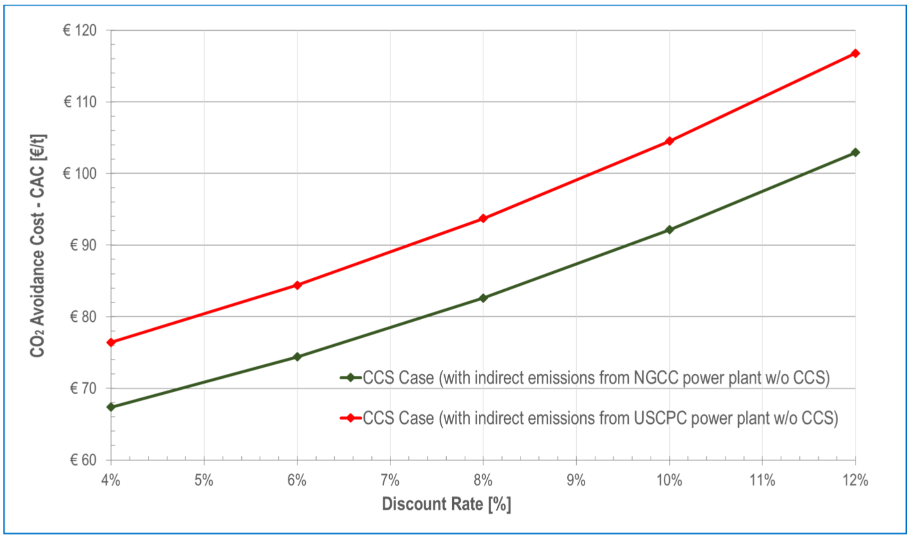


Figure 29. **Sensitivity of CO_2 Avoidance Cost to Discount Rate**

Sensitivity to CO_2 Transport & Storage Cost

Figures 30 and 31 present the sensitivity of LCOMeOH and CAC to the cost of CO_2 transport and storage. In these analysis, a negative value for CO_2 transport and storage cost could represent cash credit (which could be applicable to scenarios for EOR operation where CO_2 is bought as a commodity or any CO_2 stored is given tax credits).

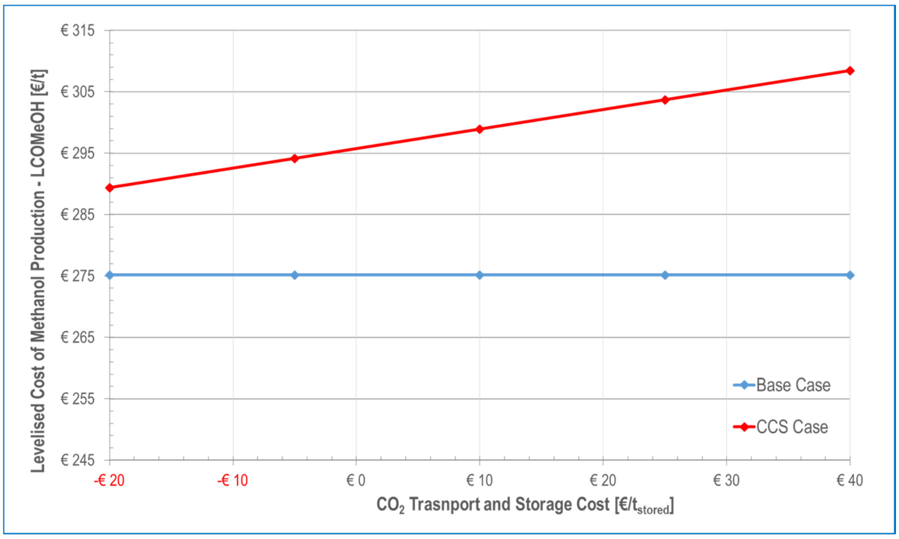


Figure 30. **Sensitivity of Levelised Cost of Methanol to CO_2 Transport and Storage Cost**

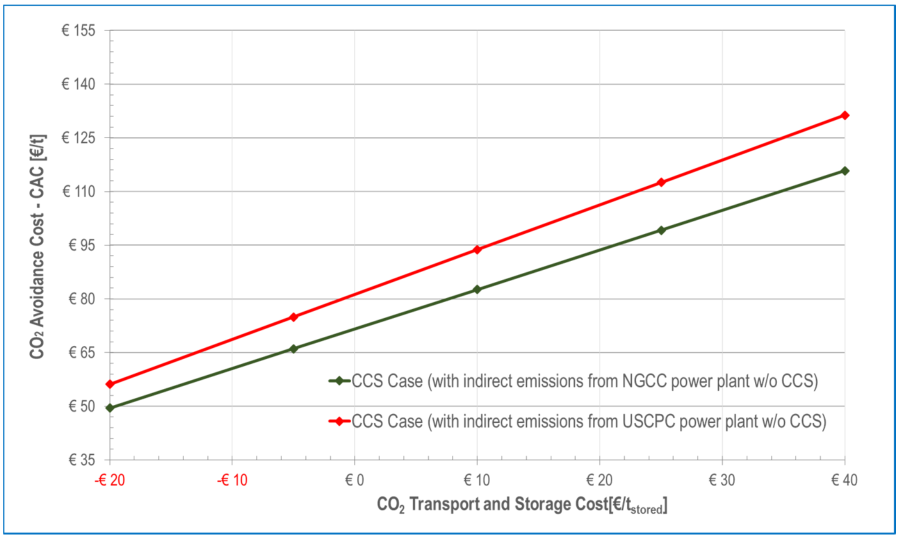


Figure 31. **Sensitivity of CO_2 Avoidance Cost to CO_2 Transport and Storage Cost**

Sensitivity to CO_2 Emission Cost

Figure 32 and 33 present the sensitivity of LCOMeOH to the cost of CO_2 emissions with the indirect CO_2 emissions accounted for in the cost of electricity bought from NGCC or Coal fired power plants respectively. In these analysis, the calculation of the LCOU is based on electricity price of the NGCC or USCPC power plant (as shown in Figure 18 – See Section 2.8.7) where the LCOE accounts for cost of the indirect CO_2 emission.

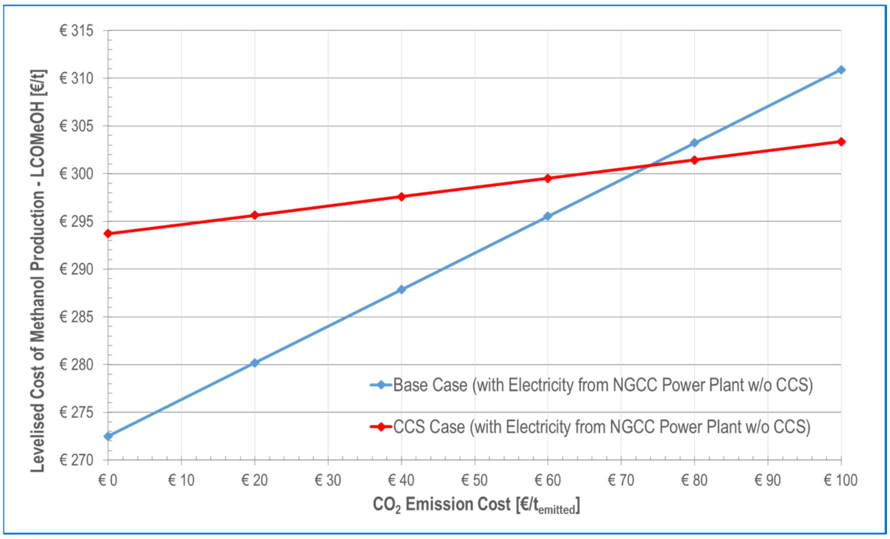


Figure 32. **Sensitivity of Levelised Cost of Methanol for the Methanol Plant (with Electricity from NGCC Power Plant w/o CCS) to the CO_2 Emission Cost**
(calculation is based on electricity price presented in Figure 18)

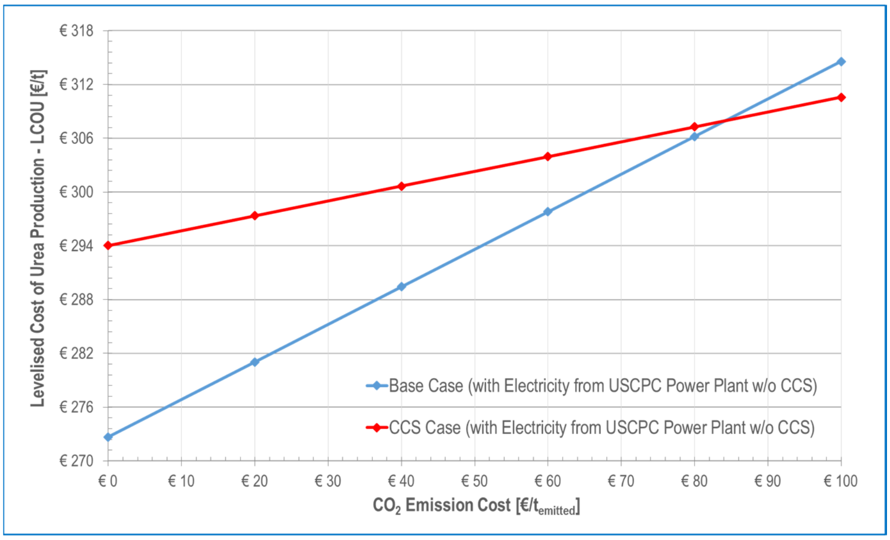


Figure 33. **Sensitivity of Levelised Cost of Methanol for the Methanol Plant (with Electricity from USCPC Power Plant w/o CCS) to the CO_2 Emission Cost**
(calculation is based on electricity price presented in Figure 18)

Annex I:

Reference Document (Task 2): Criteria for Assessing the Techno-Economic Evaluation of H_2 or HYCO Plant

CLIENT	:	IEA Greenhouse Gas R&D Programme (IEAGHG)
PROJECT NAME	:	Techno-Economic Evaluation of H_2 Production with CO_2 Capture
DOCUMENT NAME	:	Reference Document
FWI Contract	:	1BD0840A

ISSUED BY	:	G. AZZARO
CHECKED BY	:	G. COLLODI
APPROVED BY	:	G. COLLODI

DATE	REVISED PAGES	ISSUED BY	CHECKED BY	APPROVED BY

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1. Introduction

In the past years, IEA Greenhouse Gas R&D Programme (IEAGHG) has issued a series of reports presenting the performance and cost of CO_2 capture technologies when applied to energy intensive industries and these include cement, iron and steel, pulp and paper, oil refining, co-production of power and hydrogen from coal.

In line with the commitment to support and understand any development and deployment of low carbon energy technologies, IEAGHG has contracted Amec Foster Wheeler (Amec FW) to perform a study aiming to investigate the deployment of CO_2 capture technologies in a hydrogen production unit operating as standalone (merchant) or integrated to an industrial complex.

Hydrogen is used as feedstock to various industries. This could be delivered as nearly pure H_2 or as HYCO gas. Currently, nearly 95% of hydrogen generated industrially is consumed by ammonia production, oil refineries and other chemical industries (i.e basic chemicals, petrochemicals, oleo-chemicals etc...), and metal industries (i.e. direct iron reduction or DRI production).

Nearly 90% of the Hydrogen production from NG or other light hydrocarbon (HC) feedstock is produced from steam methane reforming (SMR). Other production routes include autothermal reforming (ATR) and partial oxidation (POX).

The leading technologies available for capturing CO_2 from H_2 plants include the use of chemical absorption technology (in post-combustion or pre-combustion options), cryogenic or low temperature separation technology, membrane, PSA and others.

This study mainly aims to evaluate performance and cost of capturing CO_2 from the shifted syngas, PSA tail gas or SMR’s flue gas using chemical absorption technology (as applied to all the possible CO_2 capture schemes) or cryogenic and membrane separation technology (as applied to capturing of CO_2 from the PSA tail gas only).

One of the objectives of this study is aimed to develop a common methodology of assessing the techno-economics of H_2 production. As such, the scope of TASK 2 has been set to provide a “Reference Document” to serve as a good basis to develop the different assumptions (engineering and techno-economic parameters) that could be used in evaluating the levelised cost of H_2 production without and with CO_2 capture (for TASK 3); and the levelised cost of ammonia and methanol production without and with CO_2 capture (for TASK 4).

2. Definition of the Reference Document

The scope of the techno-economic assessment consists of the following:

- TASK 3 – evaluating the performance and cost of deploying CCS in a standalone (merchant) H_2 plant.
- TASK 4 – evaluating the performance and cost of deploying CCS in an SMR based HYCO plant integrated with ammonia/urea or methanol production complex.

The "Reference Document" describes the general plant design basis and cost estimating criteria which will be used as a common basis for the techno-economic evaluation of H_2 plant without and with CO_2 capture.

The design and economic criteria outlined in the following sections will be mainly used as a reference for developing the H_2 plant configurations (scope of TASK 3) to be analysed as part of the study.

Specific criteria for TASK 4 are presented in the final report of TASK 4, as applicable.

It should be noted that in TASK 4, additional CO_2 is specifically captured from the SMR's flue gas. For the ammonia case, part of the CO_2 captured from this source will be used to maximised the production of the urea.

Where relevant, information retrieved from IEAGHG document "Criteria for Technical and Economic Assessment of Plants with Low CO_2 Emissions" Version C-6, March 2014, are included.

2.1. Merchant Hydrogen Plant

For the scope of TASK 3, the plant scheme analysed in the study includes:

- Hydrogen Plant (with or without CO_2 capture) via SMR
- Cogen Plant
- Utility Plant

The hydrogen plant will produce H_2 as the main product and HP steam as co-product.

The HP steam will be used to fulfil power requirements of the H_2 Plant by driving a steam turbine (COGEN Plant).

Should the electricity produced within the plant not be sufficient to fulfil the H_2 plant needs, the following two options will be evaluated for supplying the additional power to the H_2 plant: (a.) power import from the grid, or (b.) power production inside the battery limit via gas fired boiler (to be confirmed over the course of the work).

2.2. HYCO Plant or Syngas Generation Unit

For the scope of TASK 4, the plant scheme analysed in the study includes:

- Primary reformer
- Secondary reformer
- Associated syngas processing units
- Ammonia/urea or methanol synthesis plant

For the plants without CO_2 capture, the HP and MP steam produced in the syngas generation section and the synthesis section are mainly used to drive various turbo-machineries.

The electricity required by the plants without and with CCS are imported from the grid. The indirect CO_2 emission from the electricity will be accounted for by assuming that electricity is coming from an NGCC or USCPC power plants (without CCS).[1]

2.2.1. Ammonia/Urea Production

The HYCO plant for the ammonia/urea production consists of the SMR based primary reformer in tandem with the air blown autothermal reformer to produce a raw syngas (containing mainly CO_2, CO, CH_4, H_2 and N_2).

The raw syngas is then processed in the following units:

- Syngas cooling
- High and low temperature shift reactors
- Bulk CO_2 removal unit
- Methanation reactor
- Syngas compressor

This produces the product syngas (mainly containing H_2, N_2 with small amount of CH_4) used in the ammonia synthesis plant. The ammonia is then fed into the urea synthesis plant together with the CO_2 from the bulk removal unit.

The ammonia/urea plant produces granulated urea as the main product with chilled ammonia as co-product (for the Base Case only).

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[1] CO_2 emissions from the NGCC or USCPC will be based on the information from previous IEAGHG studies:

- IEAGHG Report No. 2014-03 "CO_2 Capture at Coal Based Power and Hydrogen Plant"
- IEAGHG Report No. 2015-05 "Oxy-Combustion Turbine Power Plants"

2.2.2. Methanol Production

The HYCO plant for the methanol production consists of the SMR based primary reformer in tandem with the oxygen blown autothermal reformer to produce a raw syngas (containing mainly CO_2, CO, CH_4, and H_2).

The raw syngas is then processed in the following units:
- Syngas cooling
- Syngas compressor

This product syngas (mainly containing CO_2, CO, H_2 with small amount of CH_4) is used as the make-up gas or MUG of the methanol synthesis plant.

The methanol plant only produces AA grade methanol as the main product.

3. General Data and Technical Assumptions

This section summarizes the general plant design criteria and assumptions used as a common basis for the design of the SMR based H_2 or HYCO plant with and without CO_2 capture.

3.1. Plant Location

The plant is situated at a greenfield site located at the North East coast of The Netherlands, with no major site preparation required. There will be no restrictions on plant area and no special civil works or constraints on the delivery of equipment are assumed. Rail lines, roads, fresh water supply and high voltage electricity transmission lines, high pressure natural gas pipeline are considered available at plant battery limits.

3.2. Climatic and Meteorological Data

Main climatic and meteorological data are listed below. Conditions marked (*) are considered reference conditions for the plant performance evaluation.

• Atmospheric pressure	101.3 kPa (*)
• Relative humidity	
o Average	80% (*)
o Maximum	95%
o Minimum	40%
• Ambient temperatures	
o minimum air temperature	-10°C
o maximum air temperature	30°C
o average air temperature	9°C (*)

3.3. Feedstock Specification

3.3.1. Natural Gas

Natural gas is used as the main feedstock and fuel to the H_2 or HYCO plant and delivered to the plant battery limits from a high pressure pipeline.

The specifications of the natural gas are shown in the table below.

Table 1. Natural Gas Specifications

Natural Gas Analysis (vol%)	
Methane	89.0
Ethane	7.0
Propane	1.0
Butane	0.1
Pentane	0.01
Carbon Dioxide	2.0
Nitrogen	0.89
Sulphur (as H_2S)	5 ppmv*
Total	100.00

HHV (MJ/kg)	51.473
LHV (MJ/kg)	46.502

Conditions at plant B.L.	
Pressure, MPa	7.0

*5 ppm_v of H_2S are assumed to be present in the natural gas for design purposes

3.4. Products and Co-Products

The main products and co-products of the Hydrogen Plant analysed in TASK 3 of this study are listed in this section, together with their main specifications.

The main products of the Ammonia/Urea and Methanol Plants analysed in this study are presented in the TASK 4 final report.

3.4.1. Hydrogen

The specification for the hydrogen used in the analysis are presented in Table 2.

The syngas or HYCO gas used in the ammonia or methanol synthesis are specified in the TASK 4 report. The pressure at the B.L. of the Syngas Generation Section is dependent on the requirement of the industrial complex.

Table 2. Hydrogen Specifications

H_2	99.5%v (min.)
$CO + CO_2$	10 ppm (max.)
CO	10 ppm (max.)
H_2S, HCl, COS, HCN, NH_3	Free
N_2 + Ar	Balance

LHV	119.96 MJ/kg
HHV	141.88 MJ/kg

Pressure at B.L.*	2.5 MPa
Temperature at B.L.	40ºC

* This is the pressure at the B.L. (to be assumed for TASK 3 evaluation).
It should be noted that this is the pressure of the H_2 product from the PSA (i.e. without any H_2 compressor)

3.4.2. *Carbon Dioxide*

The specifications of the CO_2 as delivered from the plant's B.L. to the pipeline are presented Table 3.

Table 3. Product CO_2 Specifications

Maximum allowable impurities in the product CO_2 (1)	
H_2	4% (2,4)
N_2 + Ar	4% (3,4)
O_2 (5)	100 ppm(4,6)
CO	0.2% (7)
H_2S	20 ppm(8)
H_2O	50 ppm (9)

Pressure at B.L.*	11 MPa
Temperature at B.L.	30°C

(1) Based on information available in 2012 on the requirements for CO_2 transportation and storage in saline aquifers

(2) Hydrogen concentration to be normally lower to limit loss of energy and economic value. Further investigation is required to understand hydrogen impact on supercritical CO_2 behaviour.

(3) The limit on concentrations of inerts are to reduce the volume for compression, transport and storage and limit the increase in Minimum Miscibility Pressure (MMP) in Enhanced Oil Recovery (EOR).

(4) Total non-condensable content (N_2 + O_2 + H_2 + CH_4 + Ar): maximum 4% vol. basis. This is based on the recommendations reported in the ENCAP Project (http://www.encapCO2.org)

(5) Oxygen content should be specified in conjunction with water content to limit corrosion in the downstream infrastructure.

(6) Oxygen limit is considered tentative due to the lack of practical experience on the operation of the CO_2 storage infrastructure. It is expected that stringent limit will be in place for EOR operation.

(7) CO limits are set from a health and safety perspective.

(8) H_2S specification should be specified in conjunction with water content to limit corrosion in the downstream infrastructure.

(9) Water specification is to ensure there is no free water and hydrate formation.

3.4.3. HP Steam

Refer to Section 3.11.4

3.4.4. Electric Power

High voltage grid connection:	380 kV
Frequency:	50 Hz

3.5. Environmental Limits

The environmental limits set up for each cases are outlined in this section.

3.5.1. Gaseous Emissions

The overall gaseous emissions from the plant should not exceed the following limits:

NO_x (as NO_2)*	≤ 120 mg/Nm^3
SO_x (as SO_2)*	N.A.**
CO	≤ 30 mg/Nm^3

* Emission expressed in mg/Nm^3 @ 3% O_2, dry basis.

** SOx will be very minimal – given that the PSA tail gas is expected to be sulphur free and NG as supplementary fuel contains only less than 5 ppm$_v$.

3.5.2. Liquid Effluent Discharge

Characteristics of waste water discharged from the plant should comply with the standard limits required by the EU directives currently in force.

The main liquid effluent that continuously flows out of the B.L. is coming from the blow-down of the steam drum (in the Deaerator section of the BFW system).

Sea water used in the primary cooling system is returned to the sea with allowable maximum temperature increase of 7°C.

3.5.3. Solid Wastes Disposal

Solid wastes from the Hydrogen Plant consists of the spent catalysts. All solid wastes will be handled in accordance to the instruction and guidelines provided by the catalyst vendors and the plant owner's established procedure.

The spent catalysts collected from the plant are in their oxidized/inert state; as such, these are considered non-hazardous.

The reformer's and pre-reformer's catalyst contains nickel, which can often be recovered. The other spent catalyst would normally be disposed of in the landfill.

3.5.4. Noise Pollution

All the equipment of the plant are designed to obtain a sound pressure level of 85 dB(A) at 1 meter from the equipment.

3.6. Key Features of the Hydrogen Production Plant

This key features of the SMR based Hydrogen Plant considered for TASK 3 are presented in this section of the report.

The key features of the Syngas Generation or HYCO plant considered for the ammonia and methanol production are specified in the Final Report of TASK 4.

3.6.1. Capacity

The plant capacity is assumed contact for all cases producing 100,000 Nm^3/h of high purity Hydrogen.

3.6.2. Configuration

The hydrogen production plant consists of one train and integrates the following sections:

- Feed Pre-treatment
- Pre-reforming
- Primary Reforming
- Water Gas Shift Conversion
- Final Hydrogen Purification (based on PSA)
- Steam and BFW System
- CO_2 Capture System (only for CO_2 capture cases)
- CO2 Compression and Dehydration (only for CO_2 capture cases)

3.6.3. Plant Turndown

The minimum turndown of the hydrogen plant considered in this study is assumed at 40%.

For TASK 4, the minimum turndown is dependent on the chemical complex it is integrated with.. However, it is expected that high availability is required for the syngas or HYCO production to meet the demand of the chemical production operation. Nonetheless, in typical normal operation, 40% turn down for the SMR should be necessary during start-up or upset within the chemical complex (hence reducing natural gas consumption during these events).

3.7. Capacity Factor

The table below presents the expected capacity factor (average yearly capacity factor) of the hydrogen plant evaluated in TASK 3.

Plant	Year	Average capacity factor
H_2 Production	1st year of operation	70%
	2nd – 25th year of operation	95%

The capacity factor of 90% will be assumed for the ammonia/urea and methanol production complex evaluated in TASK 4.

3.8. Process and Utility Units

This section summarised the different unit processes and utilities included in the B.L. of the hydrogen plant considered for TASK 3.

Process Units

- Feed Pre-treatment
- Pre-reforming
- Primary Reforming
- Water Gas Shift conversion
- Final Hydrogen purification (PSA)
- Steam and BFW system
- CO_2 capture system (only for CO_2 capture cases)
- CO_2 compression (only for CO_2 capture cases)

Utilities and Offsite Units

- Cooling water
- Demineralised, Condensate recovery Water Systems
- Plant/Instrument Air Systems
- Inert gas System
- Fire Fighting System
- COGEN Plant (import/export, depending on cogeneration option considered)
- Chemicals
- Flare system
- Interconnecting

The different unit processes and utilities of the ammonia/urea and methanol production complex are presented in the Final Report of TASK 4.

3.9. Units of Measurement

The units of measurement used in this study are in SI units.

3.10. Plant's Battery Limits

The plant's battery limits are defined in the Final Report of TASK 3 and TASK 4.

3.11. Utility and Service Fluids Characteristics/Conditions

The following sections present the main utilities and service fluids used within the hydrogen or HYCO plant.

3.11.1. Cooling Water

The cooling water system is based on once through seawater cooling for the primary system and close circuit demi-water cooling for the secondary system.

Primary System – Seawater Cooling Specifications

Source : sea water in once through system
Service : for steam turbine condenser and CO2 compression unit.
Type : clear filtered and chlorinated, without suspended solids and organic matter.
Salinity : 22 g/l

Supply temperature:

- average supply temperature (on yearly basis): 12°C
- max supply temperature (average summer): 14°C
- min. supply temperature (average winter): 9°C
- max. allowable seawater temperature increase: 7°C

Return temperature:

- average return temperature: 19°C
- max return temperature: 21°C

Design temperature: 50°C
Operating pressure at condenser inlet: 0.05 MPa(g)
Design pressure: 0.4 MPa(g)
Max allowable ΔP for Users: 0.05MPa(g)
Turbine condenser minimum ΔT: 5°C*
Turbine condenser conditions

- Temperature 28°C*
- Pressure 0.0038 MPa*

Secondary System: Closed Circuit Demineralised Water (Demi-Water) Cooling

Source	:	demineralised water stabilized & conditioned – seawater cooled
Service	:	for machinery cooling and for all plant users other than steam turbine condenser and CO_2 compression exchangers

Supply temperature:

- average supply temperature	19°C
- max. supply temperature:	21°C
- max. allowable temperature increase:	11°C

Design temperature:	50 °C
Operating pressure at Users:	0.3 MPa(g)
Design pressure:	0.7 MPa(g)
Max. allowable ΔP for Users:	0.15 MPa(g)

3.11.2. *Air Cooling System*

Air temperature to be considered for the air cooler design is set at 25 ^{o}C.

3.11.3. *Demineralised Water (Demi-Water)*

Type:	Treated raw water
Operating pressure at grade (min):	0.5 MPa(g)
Design pressure:	0.95 MPa(g)
Operating temperature:	Ambient
Design temperature:	38°C

Specifications:

- pH		6.5÷7.0	
- Total dissolved solids	mg/kg	0.1	max
- Conductance at 25°C	µS	0.15	max
- Iron	mg/kg as Fe	0.01	max
- Free CO_2	mg/kg as CO_2	0.01	max
- Silica	mg/kg as SiO_2	0.015	max

3.11.4. Steam Conditions

The conditions for the HP and LP steam used in the evaluation for the Hydrogen Plant in TASK 3 are summarised below.

Table 4. Steam Conditions used by the Process Units

Steam (at Process Unit's B.L.)	**Pressure - MPa(g)**				**Temperature - °C**			
	min	**normal**	**max**	**design**	**min**	**normal**	**max**	**design**
HP Steam	3.92	4.13	4.21	4.68/FV	375	395	405	425
LP Steam	0.31	0.34	0.37	0.63/FV	150	177	180	210

The conditions for the HP, MP and LP steam used in the evaluation for the Ammonia/Urea and the Methanol Production Complex are presented in the Final Report of TASK 4.

3.11.5. Instrument and Plant Air Specifications

Instrument Air

Operating pressure	
- normal:	0.7 MPa(g)
- minimum:	0.5 MPa(g)
Design pressure:	1 MPa(g)
Operating temperature (max):	40°C
Design temperature:	60°C
Dew point @ 0.7 MPa(g):	-30°C

Plant Air

Operating pressure:	0.7 MPa(g)
Design pressure:	1MPa(g)
Operating temperature (max):	40°C
Design temperature:	60°C

3.11.6. Nitrogen

Low Pressure	Nitrogen
Supply pressure:	0.65 MPa(g)
Design pressure:	1.15 MPa(g)
Supply temperature (min):	15°C
Design temperature:	70°C
Min Nitrogen Purity:	99.9 % vol. (instrument grade)

3.11.7. Chemicals

The chemicals used in the Hydrogen (TASK 3) or HYCO (TASK 4) generally consists of the additives used in treating boiler feed water and condensates. For example:

Oxygen scavenger:	Nalco Elimin-OX 100%, or equivalent,
Phosphate injection:	Water solution with 50% Na_2HPO_4 and
pH control injection:	Morpholine (100%)

Design pressure:	atmospheric pressure plus full tank of liquid solution
Design temperature:	80°C

3.12. Codes and Standards

The design of the process and utility units are in general accordance with the main International and EU Standard Codes.

3.13. Software Codes

For the design of the plant, three software codes have been mainly used to evaluate the heat and mass balances of the different study cases:

- PROMAX v3.2 (by Bryan Research & Engineering Inc.): Simulation of the CO_2 capture from the shifted syngas, PSA tail gas or SMR's flue gas using amine sweetening process.
- Aspen HYSYS v7.3 (by AspenTech): Simulation of the SMR based hydrogen or HYCO plant and the CO_2 compression and dehydration unit.
- Gate Cycle v6.1 (by General Electric): Simulation of the Power Island used by the Steam Turbine and the Preheating Line of the condensate and BFW.

4. Criteria for Economic Evaluation

The following sections describe the main bases in estimating the cost or criteria used in the economic assessment of the H_2 or HYCO plants (without and with CO_2 capture).

4.1. Economic Criteria

4.1.1. Plant Economic Life

The plant is designed for 25 years life.

4.1.2. Project Schedule

Project start		2016
Plant operation start		2019
Investment phase duration	(years)	3
Plant operating life	(years)	25
Plant operation end		2043

4.1.3. Total Capital Requirement

The Total Capital Requirement (TCR) includes:
- Total Plant Cost (TPC)
- Spare parts cost
- Start-up costs
- Owner's costs.
- Interest during construction
- Working capital

The estimates are quoted in euros (€), based on 4Q 2014 price level.

4.1.4. Total Plant Cost

The Total Plant Cost (TPC) is the installed cost of the plant including contingencies. The estimates are broken down into the main process units and, for each cases and further split into the following items:
- Direct materials
- Construction
- EPC services
- Other costs
- Contingency

4.1.5. Estimate Accuracy

The accuracy of the estimates are in the range of +35%/-15% (which in accordance to AACE Class 4 specification).

4.1.6. Contingency

The project contingency is added to the capital cost to give a 50% probability of a cost over-run or under-run.

For the accuracy considered in this study, 20% of the TPC is assumed for all the different units of the plant.

4.1.7. Design and Construction Period

Plant design and construction period and the curve of capital expenditure during construction are defined as follows:

- Construction period(1) 3 years
- Capital Expenditure Curve

Year	Investment Cost %
1	20
2	45
3	35

(1) Note: Starting from issuance of the “Notice to Proceed” to the EPC contractor

4.1.8. Financial Leverage (Ration of Debt / Invested Capital)

All the capital requirements are treated as debt, i.e. the financial leverage equal to 100%.

4.1.9. Interests during Construction

Interest during construction is calculated from the plant construction schedule and interest rate is assumed same as the discount rate.

Expenditure is assumed to take place at the end of each year and interest during construction payable in a year is calculated based on money owed at the end of the previous year.

4.1.10. Discount Rate

The analysis is based on Discounted Cash Flow analysis. The discount rate of 8% is assumed.

4.1.11. Inflation Rate

Not considered. Real constant money is assumed in all the calculation.

4.1.12. Depreciation

Not considered. The results are reported on the Earnings Before Interest, Taxes, Depreciation and Amortisaton (EBITDA) basis.

4.1.13. Spare Parts Cost

0.5% of the TPC is assumed to cover spare part costs. It is assumed that spare parts have no value at the end of the plant life due to obsolescence.

4.1.14. Start-Up Cost

Start-Up cost consists of:

- 2% of the TPC, to cover modifications to equipment that needed to bring the unit up to full capacity.
- 25% of the full capacity feedstock and fuel cost for one month, to cover any inefficient operation that occurs during the start-up period.
- Three months of operating labour and indirect labour cost and the maintenance labour cost, to include training.
- One month of chemicals (including solvent for CO_2 capture if applicable), catalysts, and waste disposal costs and the maintenance materials cost.

4.1.15. Owner's cost

7% of the TPC is assumed to cover the Owner's cost and fees.

The Owner's cost covers the expenditure related to the feasibility studies, land surveys, land purchase, construction or improvement to roads and railways, water supply, other infrastructures, etc… beyond the site boundary, owner's engineering staff costs, permitting and legal fees, arranging financing and other miscellaneous costs.

The Owner's costs are assumed to incur in the first year of construction, allowing for the fact that some of the costs would be incurred before the start of construction.

4.1.16. Working Capital

Working capital includes inventories of fuel and chemicals (materials held in storage outside of the process plants). Storage for 30 days at full load is considered for chemicals and consumables.

It is assumed that cost of these materials is recovered at the end of the plant life.

The study also assumed a zero balance for both Trade Debtors and Trade Creditors.

4.1.17. Insurance Cost

0.5% of the TPC is assumed to cover the plant's insurance cost.

4.1.18. Local Taxes and Fees

Another 0.5% of the TPC is assumed to cover the local taxes and fees.

4.1.19. Decommissioning Cost

The salvage value of equipment and materials is normally assumed to be equal to the costs of dismantling and site restoration, resulting in a zero net cost of decommissioning.

4.2. Annual Operating and Maintenance Cost

Operating and Maintenance (O&M) costs include:
- Feedstock
- Fuel
- Chemicals
- Catalysts
- Solvents
- Raw water make-up
- Direct operating labour
- Maintenance
- Overhead Charges.

The annual O&M costs are generally classified as variable and fixed costs.

Variable cost depends on the annual operating hour of the plant; and the fixed operating costs are essentially independent from the plant operating load. They can be expressed as €/y.

4.2.1. Variable Cost

Consumables are the principal components of variable O&M cost. These include feedstock, water, catalysts (Feedstock Purification catalyst, Pre-reforming catalyst, Steam Reformer Catalyst and Shift Catalyst), chemicals, solid waste disposal and others.

Reference values for Natural gas and main consumables prices are summarised in the table below.

Item	**Cost**
Natural Gas €/GJ (LHV)	6
Raw process water, €/m^3	0.2
Electric power, €/MWh	80
CO_2 transport and storage, €/t CO_2 stored (1)	10
CO_2 emission cost, €/t CO_2 emitted	0

(1) Transport and storage cost as specified by IEAGHG, in accordance with the range of costs information in the European Zero Emissions platform's report "The costs of CO_2 capture, transport and storage", published in 2009. Sensitivity to transport and storage costs are assessed to cover lower or negative cost for EOR, due to the revenue for sale of CO_2, or higher cost, in case of off shore storage with long transport distances.

4.2.2. Fixed Cost

The fixed cost of the different plants include the following items:

Direct Labour

The yearly cost of the direct labour is calculated assuming, for each individual, an average cost equal to 60,000 €/y. The number of personnel engaged is estimated for each plant type, considering a 5 shift working pattern.

Administrative and Support Labour

All other company services not directly involved in the operation of the plant fall in this category, such as:

- Management
- Administration
- Personnel services
- Technical services
- Clerical staff.

These services vary widely from company to company and are also dependent on the type and complexity of the operation.

Administrative and support labour is assumed to be 30% of the direct labour and the maintenance labour cost (see below).

Annual Maintenance Cost

A precise evaluation of the cost of maintenance would require a breakdown of the costs amongst the numerous components and packages of the plant. Since these costs are all strongly dependent on the type of equipment selected and statistical maintenance data provided by the selected supplier, this type of evaluation of the maintenance cost is premature at study level.

For this reason the annual maintenance cost of the plant is normally estimated as a percentage of the total plant cost of the facilities, as shown in the following:

- Whole Plant 1.5% of TPC

Maintenance labour is assumed to be 40% of the overall maintenance cost.

Annex II:

Annual Cash Flow

- Case 4A-1: Ammonia/Urea Production without CCS (Base Case)
- Case 4A-2: Ammonia/Urea Production with Additional CO_2 Capture from SMR's Flue Gas using MEA (CCS Case)
- Case 4B-1: Methanol Production without CCS (Base Case)
- Case 4B-2: Methanol Production with CO_2 Capture from SMR's Flue Gas using MEA (CCS Case)

Ammonia / Urea Production without and with CCS

Cash Flow (€ Million) - EBITDA Basis

Techno-Econoic Evaluation of Deploying CCS in an Ammonia/Urea Plant using NG as Feedstock & Fuel
Base Case

Period	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Sales				154.54	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	
Selling and Distribution Expenses																													
Net Sales Revenue				154.54	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	198.70	
Variable Costs				-65.64	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	-84.06	
Electricity Import Cost				-4.36	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	-5.61	
CO2 Emission Tax																													
CO2 Transport & Storage																													
Gross Margin				84.54	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	109.03	
Fixed Costs				-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	-22.73	
Net Margin				61.81	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	
Residual Value - Capital Expenditure																													
Capital Expenditure	-177.75	-312.09	-291.29	-67.89																									
Working Capital				-0.11	0.03																								0.11
Recurring Capital																													
EBITDA (€ Million)	-177.75	-312.09	-291.29	-6.19	86.34	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	86.30	0.11

Internal Rate of Return (IRR)	**8.00%**
Discount Rate	**8.00%**
Net Present Value (NPV) - € Million	**€ 0.00**
Levelised Cost of Urea	**€ 257.34**

Cash Flow (€ Million) - EBITDA Basis

Techno-Econoic Evaluation of Deploying CCS in an Ammonia/Urea Plant using NG as Feedstock & Fuel
Case 4B - CCS Case (Additional Capture from SMR Flue Gas)

Period	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Sales				170.46	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	219.17	
Selling and Distribution Expenses																													
Net Sales Revenue				***170.46***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	***219.17***	
Variable Costs				-65.62	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	-84.03	
Electricity Import Cost				-8.57	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	-11.02	
CO2 Emission Tax																													
CO2 Transport & Storage				-1.45	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	-1.87	
Gross Margin				***94.82***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	***122.25***	
Fixed Costs				-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	-25.28	
Net Margin				***69.54***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	
Residual Value - Capital Expenditure																													
Capital Expenditure	-199.64	-350.51	-326.87	-76.22																									
Working Capital				-1.11	0.03																								1.11
Recurring Capital																													
EBITDA (€ Million)	***-199.64***	***-350.51***	***-326.87***	***-7.80***	***97.00***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***96.97***	***1.11***

Internal Rate of Return (IRR)	**8.00%**
Discount Rate	**8.00%**
Net Present Value (NPV) - € Million	**(€ 0.00)**
Levelised Cost of Urea	**€ 280.32**

Methanol Production without and with CCS

Cash Flow (€ Million) - EBITDA Basis

Techno-Econoic Evaluation of Deploying CCS in an Methanol Plant using NG as Feedstock & Fuel
Base Case

Period	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11
Sales				351.55	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99
Selling and Distribution Expenses														
Net Sales Revenue				***351.55***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***
Variable Costs				-237.12	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73
Electricity Import Cost				-9.06	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65
CO2 Emission Tax														
CO2 Transport & Storage														
Gross Margin				***105.36***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***
Fixed Costs				-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68
Net Margin				***78.69***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***
Residual Value - Capital Expenditure														
Capital Expenditure	-225.59	-396.08	-373.41	-86.48										
Working Capital				-0.37	0.10									
Recurring Capital														
EBITDA (€ Million)	***-225.59***	***-396.08***	***-373.41***	***-8.16***	***110.04***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***

Period	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Sales	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	451.99	
Selling and Distribution Expenses															
Net Sales Revenue	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	***451.99***	
Variable Costs	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	-303.73	
Electricity Import Cost	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	-11.65	
CO2 Emission Tax															
CO2 Transport & Storage															
Gross Margin	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	***136.61***	
Fixed Costs	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	-26.68	
Net Margin	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	
Residual Value - Capital Expenditure															
Capital Expenditure															
Working Capital															0.37
Recurring Capital															
EBITDA (€ Million)	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***109.93***	***0.37***

Internal Rate of Return (IRR)	**8.00%**
Discount Rate	**8.00%**
Net Present Value (NPV) - € Million	**(€ 0.00)**
Levelised Cost of Hydrogen	**€ 275.14**

Cash Flow (€ Million) - EBITDA Basis

Techno-Econoic Evaluation of Deploying CCS in an Methanol Plant using NG as Feedstock & Fuel
Base Case

Period	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Sales				381.93	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	491.06	
Selling and Distribution Expenses																													
Net Sales Revenue				***381.93***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	***491.06***	
Variable Costs				-237.09	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	-303.69	
Electricity Import Cost				-17.82	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	-22.91	
CO2 Emission Tax																													
CO2 Transport & Storage				-4.06	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	-5.22	
Gross Margin				***122.96***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	***159.23***	
Fixed Costs				-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	-30.81	
Net Margin				***92.15***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	
Residual Value - Capital Expenditure																													
Capital Expenditure	-263.44	-462.54	-434.87	-100.89																									
Working Capital				-2.87	0.10																								2.87
Recurring Capital																													
EBITDA (€ Million)	***-263.44***	***-462.54***	***-434.87***	***-11.61***	***128.53***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***128.42***	***2.87***

Internal Rate of Return (IRR)	**8.00%**
Discount Rate	**8.00%**
Net Present Value (NPV) - € Million	**(€ 0.00)**
Levelised Cost of Hydrogen	**€ 298.92**

Annex III:

Breakdown of Total Capital Requirements

Total Capital Requirements

	NH3 Base Case Euro (€)	NH3 CCS Case Euro (€)	MeOH Base Case Euro (€)	MeOH CCS Case Euro (€)
Total Plant Cost (TPC)				
Total Plant Cost	546,600,000	613,900,000	693,700,000	810,100,000
Contingencies	109,320,000	122,780,000	138,740,000	162,020,000
Sub-Total	655,920,000	736,680,000	832,440,000	972,120,000
Spare Parts	3,279,600	3,683,400	4,162,200	4,860,600
Start-Up & Commissioning Cost				
Start Up CAPEX	13,118,400	14,733,600	16,648,800	19,442,400
Additional Fuel Cost	1,725,457	1,725,457	6,242,468	6,242,468
O&M	2,566,044	2,821,026	2,715,258	3,085,134
Catalyst & Chemicals	591,940	652,510	957,663	1,062,423
Owner's Cost	45,914,400	51,567,600	58,270,800	68,048,400
Interest during Construction	125,912,501	141,390,616	160,120,960	186,885,247
Working Capital	109,589	1,109,589	365,297	2,865,297
Sub-Total	193,217,931	217,683,798	249,483,446	292,491,970
Total Capital Requirements (TCR)	**849,137,931**	**954,363,798**	**1,081,923,446**	**1,264,611,970**

Annex IV:

Annual Operating Expenditure

- Case 4A-1: Ammonia/Urea Production without CCS (Base Case)
- Case 4A-2: Ammonia/Urea Production with Additional CO_2 Capture from SMR's Flue Gas using MEA (CCS Case)
- Case 4B-1: Methanol Production without CCS (Base Case)
- Case 4B-2: Methanol Production with CO_2 Capture from SMR's Flue Gas using MEA (CCS Case)

Ammonia / Urea Production without and with CCS

Annual Operating Expenditure

Techno-Econoic Evaluation of Deploying CCS in an Ammonia/Urea Plant using NG as Feedstock & Fuel
Base Case

	Base Number €'000/y	*Period*	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11
1. Fixed Cost ('000)																
1a. Direct Labour	€ *3,960.00*					3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00
1b. Admin. & General Overhead	€ *2,368.66*					2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66
1c. Insurance & Local Taxes	€ *6,559.20*					6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20
1d. Maintenance	€ *9,838.80*					9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80
		Total (Fixed O&M Cost)	**€ -**	**€ -**	**€ -**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**
2. Variable Cost ('000)																
2a. Feedstock Cost	€ *59,438.82*					46,230.20	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82
2b. Fuel Cost	€ *23,383.10*					18,186.85	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10
2c. Make Up Water	€ *34.85*					27.10	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85
2d. Chemicals	€ *200.00*					200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00
2e. Catalyst	€ *1,000.00*					1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00
		Total (Variable O&M Cost)	**€ -**	**€ -**	**€ -**	**€ 65,644.15**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**
3. Revenues ('000)																
3a. Electrcity Import from the Grid	€ *5,607.12*					4,361.09	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12
3b. Ammonia (Sold to the Market)	-€ *7,640.00*					-5,942.22	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00
		Total (Additional Revenues)	**€ -**	**€ -**	**€ -**	**-€ 1,581.13**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**
4. Cost of Emissions ('000)																
4a. CO2 Emissions Tax	€ -															
		Total (Cost of CO2 Emissions)	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**
5. Cost of CO2 Transport & Storage ('000)																
5a. CO2 Transport & Storage	€ -															
		Total (Cost of CO2 Storage)	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**

	12	13	14	15	16	17	18	19	20	21	22	23	24	25
1. Fixed Cost ('000)														
1a. Direct Labour	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00	3,960.00
1b. Admin. & General Overhead	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66	2,368.66
1c. Insurance & Local Taxes	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20	6,559.20
1d. Maintenance	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80	9,838.80
Total (Fixed O&M Cost)	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**	**€ 22,726.66**
2. Variable Cost ('000)														
2a. Feedstock Cost	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82
2b. Fuel Cost	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10
2c. Make Up Water	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85	34.85
2d. Chemicals	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00
2e. Catalyst	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00
Total (Variable O&M Cost)	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**	**€ 84,056.77**
3. Revenues ('000)														
3a. Electrcity Import from the Grid	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12	5,607.12
3b. Ammonia (Sold to the Market)	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00	-7,640.00
Total (Additional Revenues)	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**	**-€ 2,032.88**
4. Cost of Emissions ('000)														
4a. CO2 Emissions Tax														
Total (Cost of CO2 Emissions)	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**
5. Cost of CO2 Transport & Storage ('000)														
5a. CO2 Transport & Storage														
Total (Cost of CO2 Storage)	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**

Annual Operating Expenditure

Techno-Econoic Evaluation of Deploying CCS in an Ammonia/Urea Plant using NG as Feedstock & Fuel
Case 4B - CCS Case (Additional Capture from SMR Flue Gas)

	Base Number €'000/y	Period	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
1. Fixed Cost ('000)																														
1a. Direct Labour	€ *4,260.00*					4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00	4,260.00
1b. Admin. & General Overhead	€ *2,604.02*					2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02	2,604.02
1c. Insurance & Local Taxes	€ *7,366.80*					7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80	7,366.80
1d. Maintenance	€ *11,050.20*					11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20	11,050.20
Total (Fixed O&M Cost)			**€ -**	**€ -**	**€ -**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**	**€ 25,281.02**
2. Variable Cost ('000)																														
2a. Feedstock Cost	€ *59,438.82*					46,230.20	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82	59,438.82
2b. Fuel Cost	€ *23,383.10*					18,186.85	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10	23,383.10
2c. Make Up Water	€ *6.46*					5.03	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46	6.46
2d. Chemicals	€ *200.00*					200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00
2e. Catalyst	€ *1,000.00*					1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00	1,000.00
Total (Variable O&M Cost)			**€ -**	**€ -**	**€ -**	**€ 65,622.08**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**	**€ 84,028.39**
3. Revenues ('000)																														
3a. Electrcity Import from the Grid	€ *11,021.84*					8,572.54	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84	11,021.84
3b. Ammonia (Sold to the Market)	€ -																													
Total (Additional Revenues)			**€ -**	**€ -**	**€ -**	**€ 8,572.54**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**	**€ 11,021.84**
4. Cost of Emissions ('000)																														
4a. CO2 Emissions Tax	€ -																													
Total (Cost of CO2 Emissions)			**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**
5. Cost of CO2 Transport & Storage ('000)																														
5a. CO2 Transport & Storage	€ *1,866.30*					1,451.57	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30	1,866.30
Total (Cost of CO2 Storage)			**€ -**	**€ -**	**€ -**	**€ 1,451.57**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**	**€ 1,866.30**

Methanol Production without and with CCS

Annual Operating Expenditure

Techno-Econoic Evaluation of Deploying CCS in an Methanol Plant using NG as Feedstock & Fuel
Base Case

	Base Number €'000/y	Period	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
1. Fixed Cost ('000)																														
1a. Direct Labour	€ *3,360.00*					3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00	3,360.00
1b. Admin. & General Overhead	€ *2,506.39*					2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39	2,506.39
1c. Insurance & Local Taxes	€ *8,324.40*					8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40	8,324.40
1d. Maintenance	€ *12,486.60*					12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60	12,486.60
Total (Fixed O&M Cost)			**€ -**	**€ -**	**€ -**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**	**€ 26,677.39**
2. Variable Cost ('000)																														
2a. Feedstock Cost	€ *261,981.58*					203,763.45	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58
2b. Fuel Cost	€ *37,656.91*					29,288.71	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91
2c. Make Up Water	€ *89.72*					69.78	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72	89.72
2d. Chemicals	€ *200.00*					200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00
2e. Catalyst	€ *3,800.00*					3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00
Total (Variable O&M Cost)			**€ -**	**€ -**	**€ -**	**€ 237,121.94**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**	**€ 303,728.21**
3. Revenues ('000)																														
3a. Electrcity Export (Import) to the Grid	€ *11,649.36*					9,060.61	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36	11,649.36
Total (Additional Revenues)			**€ -**	**€ -**	**€ -**	**€ 9,060.61**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**	**€ 11,649.36**
4. Cost of Emissions ('000)																														
4a. CO2 Emissions Tax	€ -																													
Total (Cost of CO2 Emissions)			**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**
5. Cost of CO2 Transport & Storage ('000)																														
5a. CO2 Transport & Storage	€ -																													
Total (Cost of CO2 Storage)			**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**

Annual Operating Expenditure

Techno-Econoic Evaluation of Deploying CCS in an Methanol Plant using NG as Feedstock & Fuel
Base Case

	Base Number €'000/y	*Period*	-3	-2	-1	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
1. Fixed Cost ('000)																														
1a. Direct Labour	€ *3,660.00*					3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00	3,660.00
1b. Admin. & General Overhead	€ *2,847.82*					2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82	2,847.82
1c. Insurance & Local Taxes	€ *9,721.20*					9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20	9,721.20
1d. Maintenance	€ *14,581.80*					14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80	14,581.80
Total (Fixed O&M Cost)			**€ -**	**€ -**	**€ -**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**	**€ 30,810.82**
2. Variable Cost ('000)																														
2a. Feedstock Cost	€ *261,981.58*					203,763.45	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58	261,981.58
2b. Fuel Cost	€ *37,656.91*					29,288.71	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91	37,656.91
2c. Make Up Water	€ *54.72*					42.56	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72	54.72
2d. Chemicals	€ *200.00*					200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00	200.00
2e. Catalyst	€ *3,800.00*					3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00	3,800.00
Total (Variable O&M Cost)			**€ -**	**€ -**	**€ -**	**€ 237,094.71**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**	**€ 303,693.20**
3. Revenues ('000)																														
3a. Electrcity Export (Import) to the Grid	€ *22,907.76*					17,817.15	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76	22,907.76
Total (Additional Revenues)			**€ -**	**€ -**	**€ -**	**€ 17,817.15**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**	**€ 22,907.76**
4. Cost of Emissions ('000)																														
4a. CO2 Emissions Tax	€ -																													
Total (Cost of CO2 Emissions)			**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**	**€ -**
5. Cost of CO2 Transport & Storage ('000)																														
5a. CO2 Transport & Storage	€ *5,222.52*					4,061.96	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52	5,222.52
Total (Cost of CO2 Storage)			**€ -**	**€ -**	**€ -**	**€ 4,061.96**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**	**€ 5,222.52**

Evaluating technology options to mitigate greenhouse gas emissions

IEA Greenhouse Gas R&D Programme

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