Techno-Economic Evaluation of Deploying CCS in SMR Based Merchant H₂ Production with NG as Feedstock and Fuel

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Abstract

Hydrogen is a crucial raw material to other industries. Globally, nearly 90% of the hydrogen or HyCO gas produced is consumed by the ammonia, methanol and oil refining industries. In the future, hydrogen could play an important role in the decarbonisation of transport fuel (i.e. use of fuel cell vehicles) and space heating (i.e. industrial, commercial, building and residential heating).

This paper summarizes the results of the feasibility study carried out by Amec Foster Wheeler for the IEA Greenhouse Gas R&D Programme (IEA GHG) with the purpose of evaluating the performance and costs of a modern steam methane reforming without and with CCS producing 100,000 Nm³/h H₂ and operating as a merchant plant. This study focuses on the economic evaluation of five different alternatives to capture CO₂ from SMR. This paper provides an up-to-date assessment of the performance and cost of producing hydrogen without and with CCS based on technologies that could be erected today.

This study demonstrates that CO₂ could be captured from an SMR plant with an overall capture rate ranging between 53 to 90%. The integration of CO₂ capture plant could increase the NG consumption by -0.03 to 1.41 GJ per Nm³/h of H₂. The amount of electricity exported to the grid by the SMR plant is reduced. The levelised cost of H₂ production could increase by 2.1 to 5.1 € cent per Nm³ H₂ (depending on capture rate and technology selected). This translates to a CO₂ avoidance cost of 47 to 70 €/t.

Keywords: industrial CCS, hydrogen production, chemical absorption, MDEA, MEA, low temperature CO₂ capture, cryogenic, membrane

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1. Introduction

Hydrogen is a key raw material to other industries. Globally, nearly 90% of the hydrogen or HyCO gas produced is used by the ammonia, methanol and oil refining industries. In the future, hydrogen could play an important role in the decarbonisation of transport fuel (i.e. use of fuel cell vehicles) and space heating (i.e. industrial, commercial, building and residential heating).

The economics of hydrogen production are determined by several factors such as cost and quality of the feedstock, and utilities. Around 90% of the feedstock used in the production of hydrogen are from fossil fuels i.e. natural gas, fuel oil and coal. Other feedstock could include other hydrocarbons such as refinery off-gases, LPG, naphtha, petcoke, asphalts, vacuum tars, and others.

The conversion of fossil fuels to hydrogen also produces significant amount of CO\(_2\) as by-product. Environmental concerns regarding the reduction of CO\(_2\) emissions from energy intensive industries (including hydrogen production) should be expected in the future.

Currently, the steam methane reformer (SMR) is the leading technology for H\(_2\) production from natural gas or light hydrocarbons. Most of the modern SMR based hydrogen production facilities have achieved efficiency that could reduce CO\(_2\) emissions down to nearly 10% above its theoretical minimum. Further reduction of CO\(_2\) emissions from hydrogen production could only be achieved by the integration of CCS.

To understand the cost of deploying CO\(_2\) capture system in a hydrogen production plant, IEA Greenhouse Gas R&D Programme commissioned Amec Foster Wheeler to undertake the “Techno-Economic Evaluation of Hydrogen Production with CO\(_2\) Capture”.

This study is mainly aimed to provide a baseline information presenting the performance and costs of incorporating the CO\(_2\) capture technologies to a SMR based hydrogen plant operating as merchant plant (as a standalone plant). It should be noted that the study does not aim to provide a definitive comparison of different technologies or technology suppliers because such comparisons are strongly influenced by specific local constraints and market factors, which can be subject to rapid changes.

1.1 Background – Capture of CO\(_2\) from an SMR Plant

A modern SMR plant mainly consists of four different sections: (1.) feedstock pre-treatment, (2.) steam reformer, (3.) shift reactor, and (4.) PSA unit. These processes are described in the main report (IEAGHG, 2016). Figure 1 presents a simplified block flow diagram of a modern SMR plant.

![Figure 1: Simplified block flow diagram – SMR plant](image-url)
In this type of SMR plant, all of the CO\textsubscript{2} is emitted from the flue gas of the steam reformer. However, it should be noted that the CO\textsubscript{2} are produced from the following processes:

- CO\textsubscript{2} produced during the reforming and water-gas shift reaction;
- CO\textsubscript{2} produced during the combustion of the residual CO in the PSA tail gas and the natural gas (as supplementary fuel) in the SMR furnace.

Based on these processes, it could be gathered that the CO\textsubscript{2} could be captured from three possible locations:

- shifted syngas (Option 1)
- PSA tail gas (Option 2)
- SMR flue gas (Option 3).

1.2. **Background – Large Scale Pilot and Demonstration of SMR Plant with CCS**

The capture of CO\textsubscript{2} from an SMR plant is not a new technology. This has been done in various plants worldwide. The capture of CO\textsubscript{2} from the syngas of the SMR is a commercial technology. The current state of the art is based on chemical absorption technology. However, what’s new is the integration of capture technologies with CO\textsubscript{2} transport and storage. Additionally, new and novel CO\textsubscript{2} capture technologies are also being developed and demonstrated.

In recent years, several large scale pilot and demonstration projects have started their operation. This includes:

- Large scale demonstration projects (capturing > 1 million t/y CO\textsubscript{2})
  - Air Product’s Port Arthur Project (Texas, USA) - capture of CO\textsubscript{2} using PSA/VSA technology with the captured CO\textsubscript{2} being used for EOR operation.
  - Shell’s Quest Project (Alberta, Canada) – capture of CO\textsubscript{2} using MDEA (based on Adip-X technology) with the captured CO\textsubscript{2} being stored in saline aquifer.

- Large scale pilot projects (capturing > 100,000 t/y CO\textsubscript{2})
  - Air Liquide’s Port Jerome Project (Normandy, France) – capture of CO\textsubscript{2} using low temperature and membrane separation technology. The captured CO\textsubscript{2} is being sold as food grade CO\textsubscript{2}.
  - Japan CCS Company’s Tohokamai Project (Hokkaido, Japan) – capture of CO\textsubscript{2} using BASF’s MDEA technology with the captured CO\textsubscript{2} being sent to an off-shore CO\textsubscript{2} storage pilot facility.

In the coming years, several other demonstration projects will start their operation. These projects which are now under construction include:

- ACTL project in Alberta, Canada - which will transport the CO\textsubscript{2} obtained from a fertilizer complex (coming from an SMR plant producing syngas for ammonia / urea production) and the captured CO\textsubscript{2} from the H\textsubscript{2} production unit of the Sturgeon refinery (based on Lurgi Gasifier and Rectisol unit). The captured CO\textsubscript{2} (around 1.7 million t/y) will be transported via ACTL pipeline for EOR operation

- ADNOC Project - which will capture CO\textsubscript{2} from the SMR and DRI plant (based ENERGIRON III technology) of Emirate Steel. The captured CO\textsubscript{2} (around 800,000 t/y) will be used for EOR operation.

2. **Scope of the Study**

The primary purpose of this study is to evaluate the performance and cost of a green field modern SMR plant producing 100,000 Nm\textsuperscript{3}/h of H\textsubscript{2} from natural gas as feedstock/fuel operating in merchant plant mode.
This study focuses on the economic evaluation of five different alternatives to capture CO₂ from SMR and these include the following cases:

- **Base Case**: Modern SMR Plant equipped with Feedstock Pre-treatment, Pre-reforming, HT shift and PSA
- **Case 1A**: SMR with capture of CO₂ from the shifted syngas using MDEA
- **Case 1B**: SMR with burners firing H₂ rich fuel and capture of CO₂ from the shifted syngas using MDEA
- **Case 2A**: SMR with capture of CO₂ from the PSA tail gas using MDEA
- **Case 2B**: SMR with capture of CO₂ from the PSA tail gas using cryogenic and membrane separation
- **Case 03**: SMR with capture of CO₂ from the flue gas using MEA

All of these cases covers the different options where CO₂ could be captured within the H₂ plant (i.e. capture of CO₂ from the shifted syngas, PSA tail gas or SMR flue gas), demonstrating an overall CO₂ capture rate ranging between 50 and 90%.

3. **Basis of Design and Criteria for Economic Assessment**

3.1. **SMR Based H₂ Plant**

The primary reformer of the H₂ plant (without capture) is based on a single train Terrace Wall Fired SMR. The natural gas is fed into the feedstock pre-treatment unit to reduce any sulphur and chlorine present. This prevents any poisoning of the catalysts downstream. This is then pre-reformed in an adiabatic reactor to convert any C2+ and olefins before being fed into the primary reformer. The syngas produced from the reformer is fed into the high temperature shift (HTS) reactor to convert the CO to H₂ thus producing a syngas with residual CO of around 2.5-3%. This is then fed into the PSA to recover around 85-90% of H₂ with a purity of 99.9+% and producing the PSA Tail Gas which served as the primary fuel of the SMR.

HP Steam is generated by recovering heat from (1.) the convective section of the flue gas and (2.) the cooling of the syngas (before and after the shift reactor). The plant is optimized to minimise the amount of excess steam generated by the plant. Any excess steam is delivered to the power island, which consists of a condensing steam turbine, to generate electricity that is exported to the grid.

3.2. **Technical Design Basis**

The key assumptions used in the evaluation of the performance of the plant are as follows:

- **Plant Location:**

The site is a Greenfield location on the North East coast of The Netherlands, with no major site preparation required. No restrictions on plant area and no special civil works or constraints on 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.

- **Plant Capacity**

The plant is designed to produce 100,000 Nm³/h of high purity H₂. Any excess HP steam produced and not used by the plant are converted to electricity. This is exported to the grid.
• Capacity Factor

The study assumes a capacity factor of 70% for its first year of operation and 95% for the rest of the life of the plant. This translates to annual operating hours of 8322 h/y.

• Ambient Conditions:

Main climatic and meteorological data are listed below.

⇒ Atmospheric pressure 101.3 kPa
⇒ Average ambient temperature 9 °C
⇒ Average relative humidity 80 %

• Natural Gas (NG) specification

The natural gas (used as feedstock and fuel) is delivered to the battery limit from a high pressure pipeline. The main specification used in this study is summarized in Table 1.

Table 1: Natural Gas Specification

<table>
<thead>
<tr>
<th>Natural Gas Analysis (vol.%)</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane</td>
<td>89.0</td>
</tr>
<tr>
<td>Ethane</td>
<td>7.0</td>
</tr>
<tr>
<td>Propane</td>
<td>1.0</td>
</tr>
<tr>
<td>Butane</td>
<td>0.1</td>
</tr>
<tr>
<td>Pentane</td>
<td>0.01</td>
</tr>
<tr>
<td>CO₂</td>
<td>2.0</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>0.89</td>
</tr>
<tr>
<td>Sulphur (as H₂S)</td>
<td>5 ppmv*</td>
</tr>
<tr>
<td><strong>Total</strong></td>
<td>100.00</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Conditions at plant B.L.</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure, MPa</td>
<td>7.0</td>
</tr>
</tbody>
</table>

• Cooling Water

The primary cooling water system used by the plant is based on once through seawater cooling system. The average supply temperature is set at 12°C. The average return temperature is set at 19°C.

• Product Specifications

The H₂ produced by the plant has a purity of at least 99.5% (min). The CO and CO₂ content should be limited to 10 ppm max. It should be free of other impurities such as H₂S, HCl, COS, HCN and NH₃. The H₂ is sold and delivered to industrial consumers. The pressure and temperature at the battery limit is at 2.5MPa and 40°C respectively.

The CO₂ produced by the CO₂ capture plant has a purity of at least 99% (min). The moisture content is less than 10 ppmv. The CO₂ product is delivered to a pipeline and the conditions is suitable for use in EOR. The pressure and temperature at the battery limit is at 11.0 MPa and 30°C (max).
3.3. **General Criteria for Economic Evaluation**

Where applicable, the criteria for 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 CO₂ Emissions” Version C-6, March 2014. Other key criteria and assumptions relevant to the operation of the hydrogen plant are based on the information provided by Amec Foster Wheeler.

The criteria used in the evaluation of the cost of hydrogen production and CO₂ avoided cost are summarized below:

- **Plant Life**
  
The plant is designed 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 levelized cost of H₂ production and CO₂ 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 (€).

3.4. Definition of Cost

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. These are summarized in Section 3.5.

Fixed Operating Cost

The fixed operating cost includes the following:
- direct labor 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
Levelised Cost of Hydrogen

The Levelized Cost of Hydrogen (LCOH) is used to calculate the unit cost of producing hydrogen over their economic lifetime. This is defined as the price of hydrogen which enables the present value from all sales of hydrogen (including the additional revenue from the sale of electricity) 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.

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 LCOH in this study is calculated assuming constant (in real terms) prices 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.

Cost of CO\textsubscript{2} avoidance

Costs of CO\textsubscript{2} avoidance were calculated by comparing the CO\textsubscript{2} emissions per Nm\textsuperscript{3} H\textsubscript{2} and the LCOH of plants with capture and a reference plant without capture.

\[
\text{CO}_2 \text{ Avoidance Cost (CAC)} = \frac{\text{LCOH}_{\text{CCS}} - \text{LCOH}_{\text{Reference}}}{\text{CO}_2 \text{Emissions}_{\text{Reference}} - \text{CO}_2 \text{Emissions}_{\text{CCS}}}
\]

where:
- CAC is expressed in € per tonne of CO\textsubscript{2}
- LCOH is expressed in Euro per Nm\textsuperscript{3}/h H\textsubscript{2}
- CO\textsubscript{2} emission is expressed in tonnes of CO\textsubscript{2} per Nm\textsuperscript{3}/h H\textsubscript{2}

3.5. Key Assumptions

The assumptions used to calculate of the total capital requirements:

- Interest during construction
  Interest during construction is calculated from the plant construction schedule and the interest rate is assumed to be the 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.

- Owner’s cost
  Owner’s costs cover the costs of feasibility studies, surveys, land purchase, construction or improvement to the infrastructure beyond the site boundary (i.e. road, port, railway, water supply, etc.), owner’s engineering staff costs, permitting and legal fees, arranging financing and other miscellaneous costs.

  Owner’s costs are assumed to be all 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.
• **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.

• **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.

• **Start-up cost**

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 fuel cost for one month, to cover inefficient operation that occurs during the start-up period.
⇒ Three months of operating and maintenance labor costs, to include training.
⇒ One month of chemicals, catalyst and waste disposal costs and maintenance materials costs.

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

• **Direct labor cost**

The yearly cost of the direct labor is calculated assuming an average salary of 60,000 €/y. The number of personnel engaged in the plant’s operation is estimated for each plant type where 5 shift working pattern is considered.

This study assumes 38 personnel needed for the Base Case and 5 more additional personnel for all cases with CO₂ capture.

• **Admin and general overhead cost (indirect labor 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. These functions could include:

⇒ Management;
⇒ Administration;
⇒ Personnel services;
⇒ Clerical staff
⇒ Technical services;
⇒ R&D staff

This study assumes that the indirect labor cost is equal to 30% of the direct labor and maintenance labor cost.

• **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 labor and maintenance materials. A maintenance labor 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 2.

<table>
<thead>
<tr>
<th>Item</th>
<th>Unit</th>
<th>Cost</th>
<th>Sensitivity Range</th>
</tr>
</thead>
<tbody>
<tr>
<td>Natural gas</td>
<td>€/GJ (LHV)</td>
<td>6</td>
<td>4 to 16</td>
</tr>
<tr>
<td>Raw process water</td>
<td>€/m³</td>
<td>0.2</td>
<td></td>
</tr>
<tr>
<td>Electricity</td>
<td>€/MWh</td>
<td>80</td>
<td>40 to 120</td>
</tr>
<tr>
<td>CO₂ transport and storage</td>
<td>€/t CO₂ stored</td>
<td>10</td>
<td>0 to 20</td>
</tr>
<tr>
<td>CO₂ emission cost</td>
<td>€/t CO₂ emitted</td>
<td>0</td>
<td>0 to 100</td>
</tr>
</tbody>
</table>

The Transport and storage cost is specified in accordance to the range of costs reported by the European Zero Emissions platform’s report [ZEP 2009]. The sensitivity to the transport and storage costs are assessed to cover the lower or negative cost for EOR application (due to revenues for sale of CO₂) or higher cost – in case of off-shore CO₂ storage or with long distance CO₂ transport requirements.

- Chemical and Catalyst Cost

The cost of chemicals is assumed fixed at an annual cost of €100,000 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 feedstock pre-treatment, pre-reformer, reformer and shift reactor is also fixed at €320,000 per year (except for case 2B which is set at €405,000 per year).

4. **Case Study - Description**

Figures 2 to 7 present the simplified block flow diagram of the different cases evaluated in this study.
Figure 2: Base Case - SMR plant without CO₂ capture producing 100,000 Nm³/h H₂.

Figure 3: Case 1A - SMR Plant with capture of CO₂ from shifted syngas using MDEA

Figure 4: Case 1B - SMR Plant with H₂ rich burners and capture of CO₂ from shifted syngas using MDEA
Figure 5: Case 2A - SMR plant with capture of CO₂ from PSA tail gas using MDEA

Figure 6: Case 2B - SMR Plant with capture of CO₂ from PSA tail gas using low temperature and membrane separation

Figure 7: Case 3 - SMR Plant with capture of CO₂ from SMR flue gas using MEA
4.1. **Base Case: SMR Plant without CO₂ Capture**

The SMR based H₂ production unit consists of the following units:

- Feedstock Pretreatment
- Pre-Reformer
- Primary SMR Reformer
- High Temperature Shift
- PSA

The plant has the following key features:

- **Capacity:** 100,000 Nm³/h H₂ production
- **Syngas Capacity:** ~188,000 Nm³/h syngas (wet)
- **Residual CH₄ (after SMR):** 3.36%mole (dry basis)
- **Residual CO (after HTS):** 3.67%mole (dry basis)
- **PSA Efficiency (H₂ Recovery):** ~89%

The feedstock pre-treatment unit consists of the hydrogenation unit and a desulphurization unit involving two vessels in a lead and lag arrangement. The hydrogenation unit converts any mercaptans, other organic sulphur and organic chloride compounds in the feed to H₂S and HCl in a bed of CoMo/NiMo catalyst in the presence of H₂. The desulphurization unit consists of a bed of ZnO that absorbs the H₂S and converts it to ZnS. If any chloride compounds are present, an additional bed of sodium-based alumina is added on top of the ZnO bed. This reduces the H₂S in the feed down to less than 0.1 ppmv (dry). The removal of H₂S is to prevent any poisoning of the catalyst downstream. Normally, the feedstock is pre-heated to 370°C.

The pre-reformer is an adiabatic reactor with a bed of Ni based catalyst. The purpose of the Pre-reformer is to convert any heavy hydrocarbon in the feed to methane and to perform a portion of the overall reforming duty. In transferring some of the reformer duty of the main Reformer to the Pre-reformer, the efficiency of the process is improved. The residual C₂+ in the pre-reformer effluent is 500 ppmv (max). Typical operating temperature of the pre-reformer is around 500°C. The feed to the pre-reformer is mixed with superheated steam. The amount of process steam added to the feed is in excess of the stoichiometric quantity, in order to prevent any carbon formation on the catalyst of the primary reformer.

The primary SMR reformer is based on the Amec Foster Wheeler’s Terrace Wall Reformer design. This features a radiant section consisting of a fire-box(es), containing a single row of catalyst tubes and discharging hot flue gas into the convective section.

In the radiant section of the Steam Reformer, the mixture of feed and steam flows into the catalyst filled tubes where it reacts to produce an equilibrium mixture of H₂, CO, CO₂, CH₄ and H₂O. The residual methane is in the range of 3.3-4% (dry basis). The catalyst used in the primary reformer is a Ni based catalyst.

The convection section has several coils which recover heat from the flue gas leaving the radiant section for various process and utility duties: generation and superheating of steam, preheating of the feed to pre-treatment section and of the pre-reformer/reformer feed gases. The Reformer is designed to recover as much heat from the flue gas as it is economic, whilst avoiding dew point problems.

The syngas leaving the Steam Reformer is cooled in the Reformer Waste Heat Boiler to 320°C and the recovered heat is used to generate high pressure saturated steam in a natural circulation steam generator.
The main fuel for the Steam Reformer Furnace is the Tail Gas from the PSA Unit and supplemented with natural gas coming from unit’s B.L.

The flue gas leaving the convective section of the reformer is further cooled to around 140°C by preheating the combustion air. The flue gas has ~21%mole (wet) of CO₂.

The syngas from the primary reformer (at around 320°C) enters the high temperature shift reactor (HT Shift). This converts most of the CO to H₂ and CO₂. The residual CO in the syngas is around 3 to 4%mole (dry). The catalyst used in the HT Shift is iron based catalyst promoted by copper. This type of catalyst would require minimum amount of steam to dry gas ratio to reduce the level of unwanted side reactions.

The syngas leaving the HT shift is cooled in series of heat exchangers to produce saturated steam and preheating of the feed gas, boiler feed water, demi-water and process water. The cooled shifted syngas (raw H₂) enters the PSA unit at around 35°C. The raw H₂ is at pressure of 25 Bar(a) and consist of ~16%mole (wet) CO₂.

The PSA unit recovers the H₂ from the shifted syngas. The recovery rate is around 89%. The H₂ has a purity of 99.99+%. The tail gas leaving the PSA unit consists of the residual CH₄, H₂, CO and CO₂. This is sent to the primary reformer as fuel. The CO₂ content of the tail gas is ~51%mole (wet) at ~1.3 Bar(a).

The H₂ plant has a power island (COGEN unit) which receives all the surplus HP steam exported by the SMR plant. This is based on a condensing steam turbine that produces surplus electricity of ~9.9MWe and is exported to the grid.

4.2. Case 1A: SMR with CO₂ Capture from Shifted Syngas using MDEA

The hydrogen production plant is identical to the Base Case. The hydrogen and syngas production capacity of the plant are kept constant at 100,000 Nm³/h H₂ and ~188,000 Nm³/h wet syngas respectively. The PSA handles a smaller volume of syngas due to the removal of the CO₂ from the shifted syngas.

The changes to the hydrogen plant due to the addition of the CO₂ capture and compression plant is indicated in red box (as shown in Figure 3).

The capture plant is designed based on MDEA solvent. A more detailed block flow diagram of the CO₂ capture plant and the CO₂ compression plant are presented in the IEAGHG main report.

The addition of CO₂ capture facility resulted to some changes in the hydrogen plant and the power island. These mainly include:

- Primary reformer burners would require burner tuning adjustment to accommodate the PSA tail gas with higher LHV, and to ensure that NOx emissions regulation are met.
- Some of the heat exchangers and steam generation coils are slightly enlarged/modified to accommodate the production of steam required by the CO₂ capture plant.
- The power island is now based on a Back Pressure steam turbine.

The steam demand to regenerate the solvent of the CO₂ capture plant is about 64.4 t/h of LP steam. The additional steam needed is supplied from the reformer by increasing its steam generation capacity. This therefore increases the consumption of natural gas (as supplementary fuel).

The extraction of the LP steam from the power island and to account for the additional electricity required by the CO₂ capture and compression plant, the amount of electricity exported to the grid has been reduced to 1.49 MWe (as compared to the amount of electricity exported by the Base Case at 9.92 MWe).
4.3. **Case 1B: SMR with burners using H₂ rich fuel and CO₂ Capture from Shifted Syngas using MDEA**

The hydrogen production capacity has been kept constant at 100,000 Nm³/h H₂. However, the syngas production of the SMR plant has been increased to ~238,500 Nm³/h to deliver a “CO₂ lean” supplementary fuel (instead of NG) to the SMR (~27% increase in syngas production capacity).

In this case, the syngas production unit has been increased to accommodate the extra syngas produced to supply the SMR’s supplementary fuel (as indicated in brown boxes as shown in Figure 4). The CO₂ capture plant has been installed to capture CO₂ from the shifted syngas. These changes are indicated in red boxes (as shown in Figure 4).

The CO₂ capture plant is also designed based on MDEA solvent. This plant is similar but larger in capacity as compared to Case 1A. Around 21% of the sweet syngas (or “raw hydrogen” gas) from the CO₂ capture plant is used as supplementary fuel to the SMR.

The steam demand to regenerate the solvent of the CO₂ capture plant is about 77.8 t/h of LP steam.

Like in Case 1A, the power island is based on a back pressure steam turbine to produce electricity. Similarly due to the steam demand and the associated electricity demand of the CO₂ capture and compression plant, the amount of surplus electricity to be exported to the grid has been reduced to 1.54 MWe (as compared to the Base Case at 9.92 MWe).

4.4. **Case 2A: SMR with CO₂ Capture from PSA tail gas using MDEA**

The hydrogen production unit is identical to the Base Case. The hydrogen and syngas production capacity of the plant are kept constant at 100,000 Nm³/h H₂ and ~188,000 Nm³/h wet syngas respectively. CO₂ is captured from the PSA tail gas using MDEA solvent. Figure 5 presents the simplified schematic BFD.

In this case, the PSA tail gas is re-compressed to around 10 bar(a). The CO₂ capture plant is designed based on MDEA solvent. A detailed block flow diagram of the CO₂ capture plant is presented in the main IEAGHG report.

In addition to the electricity demand of the CO₂ capture and compression plant, the recompression of the PSA tail gas has resulted to a deficit to its electricity production. Therefore, this case would require an import of electricity from the grid of about 1.1MWe.

The total steam demand to regenerate the solvent of the CO₂ capture plant is 66.9 t/h of LP steam.

4.5. **Case 2B: SMR with CO₂ Capture from PSA tail gas using low temperature and membrane separation**

The hydrogen production unit is identical to the Base Case. The hydrogen and syngas production capacity of the plant are kept constant at 100,000 Nm³/h H₂ and ~188,000 Nm³/h wet syngas respectively. CO₂ is captured from the PSA tail gas using Cryogenic (Low Temperature) and Membrane Separation. Figure 6 presents a simplified schematic BFD. Detailed process flow diagram are presented in the main IEAGHG report.

In this case, the CO₂ capture plant is based on cryogenic separation and with additional CO₂ being captured from the vent stream of the cold box using commercially available membrane. Detailed block flow diagram is presented in the IEAGHG main report.

The CO₂ removal system has the following features:

- CO₂ compression is integrated into the CO₂ removal system
- Cold box is based on 3 flash-columns arrangement
Refrigeration needed by the cold box is supplied by an auto-refrigeration cycle using impure CO\(_2\) as refrigerant.

- Purity of the product CO\(_2\) is about 99.6%
- Around -55°C is the lowest temperature achieved within the cold box. This is the temperature of the feed gas to the third flash column (after the expansion valve).
- Vent of the cold box (vent of the 2\(^{nd}\) flash column) is sent to the membrane. The CO\(_2\) rich permeate is recycled back to the cold box. Whilst the Non-permeate CO\(_2\) lean gas from the membrane is pre-heated using the heat recovered from the CO\(_2\) compressors before being expanded and sent to the SMR as fuel.

With the tail gas from the CO\(_2\) capture plant having a better LHV (i.e. resulting to slightly better combustion efficiency) and with no additional steam needed the capture plant, the overall natural gas consumption of the SMR plant is around 0.032 MJ/Nm\(^3\) H\(_2\) less as compared to the Base Case.

The power island for this case is the same as with the Base Case (i.e. using condensing steam turbine). However, because of the additional electricity required by the CO\(_2\) capture plant, the amount of electricity exported is reduced to around 0.29 MWe (as compared to the Base Case at 9.92 MWe).

4.6. Case 3: SMR with CO\(_2\) Capture from flue gas using MEA

The hydrogen production plant is identical to the Base Case. The hydrogen and syngas production capacity of the plant are kept constant at 100,000 Nm\(^3\)/h H\(_2\) and ~188,000 Nm\(^3\)/h wet syngas respectively. CO\(_2\) is captured from the flue gas using MEA solvent. Figure 7 presents a simplified schematic BFD.

In this case, the changes to the H\(_2\) plant are indicated in red boxes (as shown in Figure 7). CO\(_2\) capture plant is designed based on a proprietary MEA solvent using split flow configuration. A more detailed block flow diagram of the CO\(_2\) capture plant and the CO\(_2\) compression plant are presented in the IEAGHG main report.

To meet the steam demand of the CO\(_2\) capture plant, natural gas consumption has been increased and several of the heat exchangers and steam generating coils of the reformer are enlarged.

Due to the requirement to extract LP steam, a Back Pressure steam turbine is installed.

The total steam demand to regenerate the solvent is about 96.4 t/h. The additional electricity demand of the CO\(_2\) capture and compression plant results to a small amount of electricity exported to the grid at 0.43 MWe (as compared to the Base Case at 9.92 MWe).

5. Summary of Results

5.1. Plant Performance

A summary of the performance of the SMR based H\(_2\) plants with and without capture is given in Table 3.

For Cases 1A, 2A and 3, the capture of CO\(_2\) are based on chemical absorption technology; demonstrating an overall CO\(_2\) capture rate from 55% to 90%. This results to an increase in the natural gas consumption from 0.45 to 1.40 MJ/Nm\(^3\) H\(_2\) to generate the steam required for the solvent regeneration. The steam are obtained from a back-pressure steam turbine. Additional electricity required by the CO\(_2\) capture system results to a significant reduction of electricity that could be exported (with Case 2A requiring additional import of ~1.1MWe).

For Case 1B, the overall CO\(_2\) capture rate is increased to about 64% (as compared to Case 1A at 54%). This is achieved by burning H\(_2\) rich fuel instead of natural gas as supplementary fuel. The H\(_2\) rich fuel is obtained from the
sweet syngas coming from the MDEA CO$_2$ Capture Plant. For this case, the raw syngas capacity of the SMR and associated equipment are enlarged by around 27% to maintain a fixed production capacity of 100,000 Nm$^3$/h H$_2$.

For Case 2B, the CO$_2$ is captured from the PSA Tail Gas using Low Temperature CO$_2$ Separation and Membrane technology. In this case, the natural gas consumption has been reduced by 0.03 MJ/Nm$^3$ H$_2$ as compared to the Base Case. The electricity required by the CO$_2$ capture plant also reduced the net amount of electricity that could be exported.

It should be noted that the technology used in Case 2B could be re-configured to increase the hydrogen production by around 5%. However, for this study, by fixing the production rate at 100,000 Nm$^3$/h and maximising the CO$_2$ capture rate, this option was not considered.

5.2. Economic Evaluation

Capital Cost

The capital cost of the H$_2$ plants without and with CCS is summarized in Table 4 and the breakdowns of the total plant cost are given in Figure 8.

To include the partial capture of CO$_2$ from an SMR plant (i.e. CO$_2$ capture rate of 53-65%), the increase in the specific capital cost per Nm$^3$/h H$_2$ are in the range of 18-42% as compared to the Base Case. On the other hand, to capture about 90% of CO$_2$ from an SMR plant, the increase in the specific capital cost is about 78% as compared to the Base Case.

Operating Cost

The OPEX of the plant include the cost for labor, O&M, feedstock, fuel, catalyst, and chemicals. Table 5 summarized the OPEX of all the different cases.

It could be noted that the biggest factors that could increase the operating cost of the plant with CO$_2$ capture are the cost of Feedstock/Fuel, Maintenance, and the loss of Revenues from the sale of electricity.

Levelized Cost of Hydrogen and CO$_2$ Avoidance Cost

The levelized costs of hydrogen (LCOH) and CO$_2$ avoidance cost (CAC) are shown in Table 6 and Figure 9.

Sensitivity Analysis

Figure 10 presents the sensitivity of the CO$_2$ avoidance cost to the price of natural gas. It could be demonstrated that with a lower natural gas price, Case 1A has the lowest CAC. On the other end, higher price for natural gas could favor Case 2B.

Figure 11 presents the sensitivity of the cost of CO$_2$ emissions (i.e. CO$_2$ tax) to the LCOH. It could be demonstrated that it would need a cost of € 75-100/t on the CO$_2$ emissions to make higher capture rate option (Case 3) more attractive than partial CO$_2$ capture rate (as compared to Case 1A to Case 2B).
### Table 3: Plant Performance Summary

<table>
<thead>
<tr>
<th>Inlet Stream</th>
<th>Base Case</th>
<th>Case 1A</th>
<th>Case 1B</th>
<th>Case 2A</th>
<th>Case 2B</th>
<th>Case 3</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>NG to Fuel (t/h)</strong></td>
<td>4.332</td>
<td>5.300</td>
<td>0.000</td>
<td>5.597</td>
<td>4.264</td>
<td>7.347</td>
</tr>
<tr>
<td><strong>LHV (MJ/kg)</strong></td>
<td>46.50</td>
<td>46.50</td>
<td>46.50</td>
<td>46.50</td>
<td>46.50</td>
<td>46.50</td>
</tr>
<tr>
<td><strong>Total Energy Input (A) (MW)</strong></td>
<td>394.77</td>
<td>407.68</td>
<td>430.55</td>
<td>411.11</td>
<td>393.89</td>
<td>433.72</td>
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<table>
<thead>
<tr>
<th>Outlet Stream</th>
<th>Base Case</th>
<th>Case 1A</th>
<th>Case 1B</th>
<th>Case 2A</th>
<th>Case 2B</th>
<th>Case 3</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>H2 to B.L. (t/h)</strong></td>
<td>8.994</td>
<td>8.994</td>
<td>8.994</td>
<td>8.994</td>
<td>8.994</td>
<td>8.994</td>
</tr>
<tr>
<td><strong>LHV (Nm^3/h)</strong></td>
<td>100000</td>
<td>100000</td>
<td>100000</td>
<td>100000</td>
<td>100000</td>
<td>100000</td>
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<tr>
<td><strong>Total Energy in Product (B) (MW)</strong></td>
<td>300.02</td>
<td>300.02</td>
<td>300.02</td>
<td>300.02</td>
<td>300.02</td>
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<thead>
<tr>
<th>Electricity Balance</th>
<th>Base Case</th>
<th>Case 1A</th>
<th>Case 1B</th>
<th>Case 2A</th>
<th>Case 2B</th>
<th>Case 3</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Gross Power Output from COGEN Plant (MWe)</strong></td>
<td>11.500</td>
<td>6.700</td>
<td>8.000</td>
<td>6.900</td>
<td>11.000</td>
<td>11.700</td>
</tr>
<tr>
<td><strong>H2 Plant (MWe)</strong></td>
<td>-1.216</td>
<td>-1.257</td>
<td>-1.582</td>
<td>-1.264</td>
<td>-1.216</td>
<td>-1.314</td>
</tr>
<tr>
<td><strong>COGEN Plant Auxiliaries, Utilities, BOP (MWe)</strong></td>
<td>-0.366</td>
<td>-0.377</td>
<td>-0.440</td>
<td>-0.397</td>
<td>-0.511</td>
<td>-1.677</td>
</tr>
<tr>
<td><strong>CO2 Capture Plant (MWe)</strong></td>
<td>NA</td>
<td>-0.569</td>
<td>-0.717</td>
<td>-3.435</td>
<td>-8.989</td>
<td>-2.001</td>
</tr>
<tr>
<td><strong>CO2 Compressor (MWe)</strong></td>
<td>NA</td>
<td>-3.005</td>
<td>-3.719</td>
<td>-2.874 incl. in CO2 capture plant</td>
<td>-6.282</td>
<td></td>
</tr>
<tr>
<td><strong>Export Power to the Grid (C) (MWe)</strong></td>
<td>9.918</td>
<td>1.492</td>
<td>1.542</td>
<td>-1.070</td>
<td>0.284</td>
<td>0.426</td>
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<table>
<thead>
<tr>
<th>Specific Consumption</th>
<th>Base Case</th>
<th>Case 1A</th>
<th>Case 1B</th>
<th>Case 2A</th>
<th>Case 2B</th>
<th>Case 3</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>NG to Feedstock (MJ/Nm^3 H2)</strong></td>
<td>12.197</td>
<td>12.212</td>
<td>15.500</td>
<td>12.197</td>
<td>12.197</td>
<td>12.197</td>
</tr>
<tr>
<td><strong>NG to Fuel (MJ/Nm^3 H2)</strong></td>
<td>2.014</td>
<td>2.465</td>
<td>0.000</td>
<td>2.603</td>
<td>1.983</td>
<td>3.416</td>
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</table>

<table>
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<tr>
<th>Plant Performance</th>
<th>Base Case</th>
<th>Case 1A</th>
<th>Case 1B</th>
<th>Case 2A</th>
<th>Case 2B</th>
<th>Case 3</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Specific CO2 Emissions (kg/Nm^3 H2)</strong></td>
<td>0.8091</td>
<td>0.3704</td>
<td>0.2918</td>
<td>0.3870</td>
<td>0.3772</td>
<td>0.0888</td>
</tr>
<tr>
<td><strong>Specific CO2 Captured (kg/Nm^3 H2)</strong></td>
<td>NA</td>
<td>0.4660</td>
<td>0.5899</td>
<td>0.4556</td>
<td>0.4289</td>
<td>0.8004</td>
</tr>
<tr>
<td><strong>Overall CO2 Capture Rate (Case Specific)</strong></td>
<td>NA</td>
<td>55.7%</td>
<td>66.9%</td>
<td>54.1%</td>
<td>53.2%</td>
<td>90.0%</td>
</tr>
<tr>
<td><strong>Overall CO2 Avoided (as compared to Base Case)</strong></td>
<td>NA</td>
<td>54.2%</td>
<td>63.9%</td>
<td>52.2%</td>
<td>53.4%</td>
<td>89.0%</td>
</tr>
</tbody>
</table>
Table 4: Capital Cost of SMR Based H₂ Plant

<table>
<thead>
<tr>
<th></th>
<th>Total Plant Cost (TPC) (million €)</th>
<th>Total Capital Requirement (TCR) (million €)</th>
<th>% Increase to the TPC as compared to Base Case</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Base Case</strong></td>
<td>170.95</td>
<td>222.9</td>
<td></td>
</tr>
<tr>
<td><strong>CO₂ Capture from Shifted Syngas</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Case 1A</td>
<td>201.8</td>
<td>263.91</td>
<td>18.0%</td>
</tr>
<tr>
<td>Case 1B</td>
<td>228.48</td>
<td>298.70</td>
<td>33.7%</td>
</tr>
<tr>
<td><strong>CO₂ Capture from PSA Tail Gas</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Case 2A</td>
<td>226.07</td>
<td>295.20</td>
<td>32.2%</td>
</tr>
<tr>
<td>Case 2B</td>
<td>241.44</td>
<td>313.85</td>
<td>41.2%</td>
</tr>
<tr>
<td><strong>CO₂ Capture from Flue Gas</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Case 3</td>
<td>305.33</td>
<td>398.54</td>
<td>78.6%</td>
</tr>
</tbody>
</table>

Figure 8: Specific Total Plant Cost - SMR Based H₂ Plant
Table 5: Operating Cost of SMR Based H₂ Plant

<table>
<thead>
<tr>
<th>Case</th>
<th>LCOH</th>
<th>CO₂ Emission Avoidance Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>€/year</td>
<td>€/t</td>
</tr>
<tr>
<td>Base Case</td>
<td>11.4</td>
<td>-</td>
</tr>
<tr>
<td>Case 1A</td>
<td>13.5</td>
<td>47.1</td>
</tr>
<tr>
<td>Case 1B</td>
<td>14.6</td>
<td>62.0</td>
</tr>
<tr>
<td>Case 2A</td>
<td>14.2</td>
<td>66.3</td>
</tr>
<tr>
<td>Case 2B</td>
<td>14.0</td>
<td>59.5</td>
</tr>
<tr>
<td>Case 3</td>
<td>16.5</td>
<td>69.8</td>
</tr>
</tbody>
</table>

Figure 9: Levelized Cost of Hydrogen
Figure 10: Sensitivity of CAC to the price of natural gas

Figure 11: Sensitivity of LCOH to the CO₂ Emissions Cost (i.e. CO₂ Tax)
6. Key Findings and Concluding Remarks

- This study provides an up-to-date assessment of the performance and costs of a modern SMR based H₂ plant without and with CCS producing 100,000 Nm³/h H₂ and operating as a merchant plant (i.e. without any integration to an industrial complex).

- CO₂ from an SMR based H₂ plant could be captured from (1.) shifted syngas, (2.) PSA tail gas, or (3.) SMR flue gas.

- The LCOH of the SMR based H₂ plant with CCS as compared to the Base Case (without CCS) could increase by 2.1 to 5.2 c€/Nm³. This corresponds to an overall capture rate from 53% to 90%.

- The current state-of-the-art technology for capturing CO₂ from an SMR Based H₂ plant is presented in Case 1A where CO₂ is captured from the shifted syngas using MDEA solvent. This represents the SMR based H₂ plants with CO₂ capture that are commercially deployed today.
  - This plant consumes about 14.67 MJ/Nm³ H₂ and captures around 56% of the total CO₂ emitted (avoiding around 54% CO₂ as compared to the Base Case).
  - The LCOH for Case 1A is 13.5 c€/Nm³ which is about 2.1 c€/Nm³ higher as compared to the Base Case. The increase in the LCOH is predominantly contributed by the increase in CAPEX and cost of NG consumption; and the loss of revenues from sale of electricity.

- Case 1B presents a scenario where H₂ rich fuel could be used as supplementary fuel (instead of natural gas). As compared to Case 1A, the capture rate has been increased to 65%. This case would require the scaling up of the capacity of the SMR and associated equipment by 27% to produce enough syngas to maintain the fix production rate of 100,000 Nm³/h H₂ (as part of the sweet syngas is used as supplementary fuel). This results to an increase in the natural gas consumption by 1.3 MJ/Nm³ H₂ as compared to the Base Case. Consequently, this increases the LCOH by 3.2 c€/Nm³ (at 14.6 c€/Nm³).

- Case 2A presents one of the conventional way of capturing CO₂ from a PSA tail gas using chemical absorption (using MDEA). This achieve a capture rate of 54%. The plant’s natural gas consumption increases to 14.8 MJ/Nm³ (this is an increase of 0.59 MJ/Nm³ as compared to Base Case). The tail gas is re-compressed to 10 bar; thus resulting to higher electricity consumption. Therefore would require to buy electricity from the grid of around 1.1MWe to cover the deficit.

- Case 2B presents the capture of CO₂ from the PSA tail using low temp erature and membrane separation technology. This case demonstrated that the NG consumptions could be slightly reduced by 0.03 MJ/Nm³ as compared to the Base Case. This results to an increase to the LCOH of around 2.6 c€/Nm³ (at 14.0 c€/Nm³).
  - Case 2B is considered a high CAPEX and low OPEX plant option. If price of NG is high, Case 2B would become favorable in comparison to the other cases with CO₂ capture (i.e. CO₂ capture using Chemical Absorption Technology).
  - Furthermore, this option has the potential to improve its economics by recovering additional hydrogen from the PSA tail gas during CO₂ capture. Nonetheless, with this option, CO₂ emissions may increase given that more supplementary fuel are necessary to cover the fuel value of the “CO₂ and H₂ lean” tail gas (i.e. after additional H₂ recovery).

- Case 3 presents one of the options to capture around 90% CO₂ emitted by the SMR (high capture rate scenario). In this case, CO₂ is captured from the SMR flue gas. It could illustrated that the NG consumption of SMR plant increases by 1.6 MJ/Nm³ H₂ as compared to the Base Case. This results to an increase to the LCOH of around 5.2 c€/Nm³ (at 16.5 c€/Nm³).
7. Acknowledgement

The authors would like to acknowledge the support provided by IEA Greenhouse Gas R&D Programme in the completion of this study.

<table>
<thead>
<tr>
<th>Nomenclature</th>
<th>Definition</th>
</tr>
</thead>
<tbody>
<tr>
<td>CAC</td>
<td>cost of CO$_2$ avoided</td>
</tr>
<tr>
<td>CCS</td>
<td>CO$_2$ capture and storage</td>
</tr>
<tr>
<td>EE</td>
<td>electrical energy / electricity</td>
</tr>
<tr>
<td>HT shift</td>
<td>high temperature shift</td>
</tr>
<tr>
<td>HyCO</td>
<td>hydrogen &amp; carbon monoxide (syngas)</td>
</tr>
<tr>
<td>LCOH</td>
<td>levelized cost of hydrogen production</td>
</tr>
<tr>
<td>LCOE</td>
<td>levelized cost of electricity</td>
</tr>
<tr>
<td>LHV</td>
<td>low heating value</td>
</tr>
<tr>
<td>MDEA</td>
<td>methyl-di-ethanolamine (also known as N-methyl di-ethanolamine)</td>
</tr>
<tr>
<td>MEA</td>
<td>mono-ethanolamine</td>
</tr>
<tr>
<td>NG</td>
<td>natural gas</td>
</tr>
<tr>
<td>PSA</td>
<td>pressure swing adsorption</td>
</tr>
<tr>
<td>SMR</td>
<td>steam methane reforming</td>
</tr>
<tr>
<td>TPC</td>
<td>total plant cost</td>
</tr>
<tr>
<td>TCR</td>
<td>total capital requirement</td>
</tr>
<tr>
<td>VSA</td>
<td>vacuum swing adsorption</td>
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</tbody>
</table>