Investigating Cost Effective Pathways for Blue Hydrogen Production in Alberta

Elnigoumi, Abdalla

doctoral thesis

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Investigating Cost Effective Pathways for Blue Hydrogen Production in Alberta

by

Abdalla Elnigoumi

A THESIS

SUBMITTED TO THE FACULTY OF GRADUATE STUDIES
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Abstract

Large amounts of hydrogen are used in many industries in Alberta, the most significant of which is upgrading of bitumen into synthetic crude oil. Hydrogen also plays a large role in the recent Natural Gas Strategy, which aims to support diversification of the Alberta economy by growing the low carbon intensity energy sector. Because current hydrogen (H₂) production is responsible for a substantial share of the greenhouse emissions from oil sands operations in Alberta, applying carbon capture and storage (CCS) to H₂ production to make what is becoming known as “blue H₂” would both significantly reduce the province’s emissions and unlock new opportunities for the current industry.

In this thesis, I evaluate the trade-offs between location and scale of blue H₂ production to decarbonize Oil Sands mining operations under different future emissions prices. I develop new cost models for steam-methane reforming (SMR) and autothermal reforming (ATR) units incorporating CCS that allow estimation of production cost and life-cycle emissions for facilities in Alberta. I also expand an existing CO₂ pipeline model to be usable for H₂ pipelines, and use this to estimate the cost and environmental trade-offs between moving CO₂ and H₂. I apply these models to compare the cost of production of blue H₂ near Edmonton (with transport of H₂ north and local CO₂ storage) to production in Fort MacMurray (and transport of CO₂ south for storage) for two different demand scenarios.

Results from these models show that SMR and ATR plants capturing upwards of 90% of total direct emissions have a comparable production cost of $1.6/kgH₂ (USD) at a scale of 350 tH₂/day. This represents a 60% increase in cost compared to an SMR without capture. However, considering the full life cycle, the estimated cost of CO₂ avoided for the SMR plant ($80/tCO₂eq) are lower than for the ATR (90 $/tCO₂eq) due to the Alberta electricity grid. The unit cost of moving H₂ is several times that of CO₂, owing to the lower (gaseous) density of H₂. However, normalized to production of one kilogram of hydrogen, the costs of transport are comparable for an optimized system. The scenario analysis suggests constructing SMR with post-combustion capture in Fort MacMurray is the lowest cost option when a 170 $/tCO₂ (CAD) carbon tax is applied to the full chain life cycle emissions.
Acknowledgements

All praise and thanks are due to Allah the most merciful the most gracious. I would like to dedicate this thesis to my mom and dad, without their support I wouldn’t have been able to push through this journey.

I would like to extend my deepest thanks to my supervisor Dr. Sean McCoy for his guidance, encouragements, and for always making sure that I had all required resources and the conducive environment to complete my research. Most of all I would like to thank him for always being on my side. I would also like to thank Dr. David Layzell for creating the opportunity to research the topic of Blue Hydrogen and I would like to thank him and Dr. Md. Kibria for their valuable feedback and comments as my committee members.

I’m grateful for my colleagues and friends from the McCoy research group, our discussions have broadened my understanding in new fields of knowledge and for that and for our tea meetings, both in-person and virtual, I thank you guys for being there.

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# List of Symbols, Abbreviations and Nomenclature

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>AACE</td>
<td>Association for the Advancement of Cost Engineering</td>
</tr>
<tr>
<td>ACTL</td>
<td>Alberta Carbon Trunk Line</td>
</tr>
<tr>
<td>AER</td>
<td>Alberta Energy Regulator</td>
</tr>
<tr>
<td>ATR</td>
<td>Auto Thermal Reforming reaction/reactor</td>
</tr>
<tr>
<td>ASU</td>
<td>Air Separation Unit</td>
</tr>
<tr>
<td>ASME</td>
<td>American Society for Mechanical Engineers</td>
</tr>
<tr>
<td>BEC</td>
<td>Bare Erected Cost</td>
</tr>
<tr>
<td>bbl</td>
<td>Barrel</td>
</tr>
<tr>
<td>CAPEX</td>
<td>Capital Expenditure</td>
</tr>
<tr>
<td>C₁</td>
<td>Known cost for Plant 1 (or Equipment 1) with capacity Q₁</td>
</tr>
<tr>
<td>C₂</td>
<td>Estimated cost for Plant 2 (or Equipment 2) with capacity Q₂</td>
</tr>
<tr>
<td>CCS</td>
<td>Carbon Capture and Storage</td>
</tr>
<tr>
<td>CCUS</td>
<td>Carbon Capture Utilization and Storage</td>
</tr>
<tr>
<td>CEPCI</td>
<td>Chemical Engineering Plant Cost Index</td>
</tr>
<tr>
<td>CH₄</td>
<td>Methane</td>
</tr>
<tr>
<td>CO</td>
<td>Carbon Monoxide</td>
</tr>
<tr>
<td>CO₂</td>
<td>Carbon Dioxide</td>
</tr>
<tr>
<td>EPC</td>
<td>Engineering, Procurement and Construction Cost</td>
</tr>
<tr>
<td>EOR</td>
<td>Enhanced Oil Recovery</td>
</tr>
<tr>
<td>fF</td>
<td>Fanning friction factor</td>
</tr>
<tr>
<td>GCCSI</td>
<td>Global CCS Institute</td>
</tr>
<tr>
<td>GHG</td>
<td>Green House Gas</td>
</tr>
<tr>
<td>GJ</td>
<td>Gigajoule</td>
</tr>
<tr>
<td>GWP</td>
<td>Global Warming Potential</td>
</tr>
<tr>
<td>H</td>
<td>Pipeline elevation (m)</td>
</tr>
<tr>
<td>H₂O</td>
<td>Water</td>
</tr>
<tr>
<td>HDSAM</td>
<td>Hydrogen Delivery Scenario Analysis Model</td>
</tr>
<tr>
<td>ID</td>
<td>Internal Diameter</td>
</tr>
<tr>
<td>IECM</td>
<td>Integrated Environmental Control Model</td>
</tr>
<tr>
<td>IEAGHG</td>
<td>IEA Greenhouse Gas R&amp;D Programme</td>
</tr>
<tr>
<td>LCOH</td>
<td>Levelized Cost of Hydrogen ($/kgH₂)</td>
</tr>
<tr>
<td>ṗm</td>
<td>Design mass flow rate (kg/s)</td>
</tr>
<tr>
<td>MDEA</td>
<td>Methydiethanolamine</td>
</tr>
<tr>
<td>MEA</td>
<td>Monoethanolamine</td>
</tr>
<tr>
<td>M_w</td>
<td>Molecular weight of the fluid (kg/kmol)</td>
</tr>
<tr>
<td>NETL</td>
<td>National Energy Technology Laboratory</td>
</tr>
<tr>
<td>NREL</td>
<td>National Renewable Energy Laboratory</td>
</tr>
<tr>
<td>NEW</td>
<td>Northwest Europe</td>
</tr>
<tr>
<td>NG</td>
<td>Natural Gas</td>
</tr>
<tr>
<td>Acronym</td>
<td>Definition</td>
</tr>
<tr>
<td>---------</td>
<td>------------</td>
</tr>
<tr>
<td>NPS</td>
<td>Nominal Pipe Size</td>
</tr>
<tr>
<td>OPEX</td>
<td>Operational Expenditure</td>
</tr>
<tr>
<td>p₁</td>
<td>Segment upstream pressure (Pa)</td>
</tr>
<tr>
<td>p₂</td>
<td>Segment downstream pressure (Pa)</td>
</tr>
<tr>
<td>Pave</td>
<td>Average pressure (Pa)</td>
</tr>
<tr>
<td>POX</td>
<td>Partial Oxidation</td>
</tr>
<tr>
<td>PSA</td>
<td>Pressure Swing Adsorption</td>
</tr>
<tr>
<td>Q₁</td>
<td>Known capacity for Plant 1(or Equipment 1)</td>
</tr>
<tr>
<td>Q₂</td>
<td>Known capacity for Plant 2(or Equipment 2)</td>
</tr>
<tr>
<td>USGC</td>
<td>United States Gulf Coast</td>
</tr>
<tr>
<td>R</td>
<td>Universal gas constant (Pa.m³/mol.K)</td>
</tr>
<tr>
<td>ROA</td>
<td>Right of Way</td>
</tr>
<tr>
<td>S/C</td>
<td>Steam to Carbon Ratio</td>
</tr>
<tr>
<td>SAGD</td>
<td>Steam Assisted Gravity Drainage</td>
</tr>
<tr>
<td>SMR</td>
<td>Steam Methane Reforming reaction/reactor</td>
</tr>
<tr>
<td>Tave</td>
<td>Average fluid temperature (K)</td>
</tr>
<tr>
<td>TASC</td>
<td>Total-As Spent- Cost</td>
</tr>
<tr>
<td>TPC</td>
<td>Total Plant Cost</td>
</tr>
<tr>
<td>TOC</td>
<td>Total Overnight Cost</td>
</tr>
<tr>
<td>UCCI</td>
<td>Upstream Capital Cost Index</td>
</tr>
<tr>
<td>W</td>
<td>Work (J/kg)</td>
</tr>
<tr>
<td>WGS</td>
<td>Water Gas Shift reaction/reactor</td>
</tr>
<tr>
<td>x</td>
<td>Scale factor representing technology of Plant 1 and 2 (or Equipment 1 and 2)</td>
</tr>
<tr>
<td>Zave</td>
<td>Average fluid compressibility over the segment</td>
</tr>
<tr>
<td>ΔHᵣ</td>
<td>Heat of reaction (kJ/mol)</td>
</tr>
<tr>
<td>ηᵣₐₖₜ</td>
<td>Isentropic efficiency</td>
</tr>
<tr>
<td>ηₘₑᶜʰ</td>
<td>Mechanical efficiency</td>
</tr>
</tbody>
</table>
1. Introduction and Literature Review

1.1 Global Momentum for Hydrogen

As the world energy system begin a radical transformation to avoid climate change, hydrogen, produced through low-emission processes, has become the focus of attention as a clean energy carrier. World governments have developed and started pursuing hydrogen strategies to grow hydrogen economies and diversify their energy mix.

Prior to the recent focus on hydrogen as an energy carrier, demand for the lightest element increased 3-fold between 1975 and 2018, reaching about 70 million tonnes per year. This hydrogen is produced predominantly from fossil resources and, hence, is responsible for 830 million tonnes of CO₂ emissions every year [1]. In order to achieve the commitments made in the Paris agreement which aims to limit global warming below 2°C, countries aim to reach a peak of greenhouse gas emissions as soon as possible. Therefore, the future growth in supply must be met through hydrogen production from fossil fuels with carbon capture and storage and water electrolysis using renewable electricity as an input.

![Global annual demand for hydrogen since 1975](image)

Figure 1: Global annual demand for hydrogen since 1975[1]

Recently colours have been used to reference different production sources of hydrogen. Typically, “black”, “grey” and “brown” refer to hydrogen produced from coal, natural gas and
lignite respectively. Hydrogen produced from fossil fuels with reduced carbon intensity through carbon capture, utilization and storage (CCUS) is typically referred to as “blue”, and “green” refers to hydrogen produced through electrolysis of water using power from renewables.

In this work the focus will be on investigating the potential for blue hydrogen generation particularly from natural gas using reforming technologies and incorporating CCUS.

1.2 Alberta’s Opportunity

Canada is well poised for a hydrogen economy being one of the top ten Hydrogen producers with an estimated annual production rate of 3 million tonnes from natural gas alone [2]. An independent study of potential hydrogen markets, delivery infrastructure, and supply options for Canada concluded these goals strategy should be supported by focusing on development of hydrogen “nodes” [3]. In this concept, similar to that proposed for other regions, each node would be based on an “advantaged” hydrogen supply – whether fossil, renewable, or even waste hydrogen – and develop a market that could be activated and grown at relatively low cost [1].

The Government of Alberta aims to use hydrogen production from fossil fuels as a means to help diversify Alberta’s natural gas industry [4]. In the provincial Natural Gas Vision and Strategy [5], the government aims to utilize and expand on existing carbon capture and storage facilities and infrastructure with the goal of becoming a global exporter of hydrogen by 2040. The initial phases of production will be dominated by “blue” hydrogen particularly from natural gas with carbon capture.

1.2.1 Current Hydrogen Production and Future Projects

Alberta produces over 5000 tons of H₂ per day, primarily for upgrading of bitumen to produce synthetic crude oil, as industrial feedstock for fertilizer production, and for other petroleum refining facilities [6]. This production is of relatively low cost owing to the provinces abundance of natural gas resource and is predominately from steam methane reforming (SMR); although
low-cost, it does have a high carbon intensity where the produced CO₂ is vented into the atmosphere [7].

But Alberta also has one of the biggest demonstrated projects for blue hydrogen production, the Quest carbon capture and storage facility near Edmonton [8]. In this project, CO₂ is captured from an SMR facility supplying hydrogen to the Scotford Upgrader, the captured CO₂ is then compressed and piped 65 km north to be stored 2 km underground in a saline aquifer. The project has captured and stored over 5 million tonnes of CO₂ since its launch in 2015.

Recently, more projects have been planned for construction, Air Products and Chemicals announced a multi-billion dollar net-zero hydrogen complex in Edmonton[9], and Suncor and ATCO are collaborating on a world scale clean hydrogen project which will produce over 300,000 tons of hydrogen annually while capturing 90% of the emissions from the production process[10]. Although not much detail is released on the specifics of the technologies that will be used in these projects, it is clear that the plants will have both reforming and capture facilities for their hydrogen production.

1.2.2 GHG Analysis and Hydrogen Contribution in the Oil Sands

Alberta is one the largest emitters of greenhouse gases (GHGs) in Canada, with an estimated emission of 272.6 MtCO₂e in 2018 [11]. The largest contributor to these emissions is the oil sands sector with about 31% as shown in Figure 2, comprising of oil and gas, surface mining, in-situ bitumen extraction and upgrading into synthetic crude oil activities.
A lot of the emissions from the oil sands sector can be traced back to hydrogen production that is required for upgrading primarily from SMRs, where upgraders typically consume around 0.33-0.44 GJ of hydrogen per barrel of synthetic crude produced [12], the carbon intensity of hydrogen from the SMR process can be 9-11 kgCO$_2$eq/kgH$_2$ [7], [13]. Using these estimates about 20-40 kgCO$_2$eq for every barrel of synthetic crude oil produced can be attributed to hydrogen production process. Comparing these numbers to the GHG emission intensities for some of the upgrading projects as shown in Table 1 they represent about half the overall project emissions.

This confirms the notion that an opportunity to significantly reduce the provincial emissions can be achieved through decarbonizing hydrogen production meeting the current upgrading demand as well as for future hydrogen markets, both domestic and foreign.

Figure 2: Alberta’s GHG emission breakdown in 2018 of total 273 MtCO$_2$e (data adopted from National Inventory Report to produce figure)
### Table 1: Alberta Oil Sands upgrading projects synthetic crude production capacities and GHG emission intensities for 2018 as recorded by the AER [14] (capacities were rounded to the nearest thousand bbl/day)

<table>
<thead>
<tr>
<th>Company</th>
<th>Facility/Project</th>
<th>Synthetic crude capacity (bbl/ day)</th>
<th>Carbon Intensity (kg CO$_{2\text{eq}}$/bbl)</th>
<th>Estimate H$_2$ Contribution</th>
</tr>
</thead>
<tbody>
<tr>
<td>Canadian Natural Resources Ltd.</td>
<td>Horizon Oil Sands Processing Plant and Mine</td>
<td>231,000</td>
<td>58</td>
<td>35-68 %</td>
</tr>
<tr>
<td>Canadian Upgrading Ltd.</td>
<td>Muskeg River, Jackpine Mine Scotford Upgrader</td>
<td>286,000</td>
<td>49</td>
<td>40-80%</td>
</tr>
<tr>
<td>Syncrude Canada Ltd.</td>
<td>Mildred Lake and Aurora Plant Sites</td>
<td>252,000</td>
<td>121</td>
<td>16-33%</td>
</tr>
<tr>
<td>Suncor Energy Oil Sands Limited Partnership</td>
<td>Suncor Energy Inc. Oil Sands</td>
<td>290,000</td>
<td>69</td>
<td>29-58%</td>
</tr>
</tbody>
</table>

#### 1.2.3 CO$_2$ Storage Potential in Alberta

Large CO$_2$ storage capacities would be needed to decarbonize hydrogen production. For example, from the analysis above, if 90% of the produced CO$_2$ is to be captured that would require a minimum of 8 Mt/year of CO$_2$ storage. Recognizing that the storage element is a key component of the CSS chain the suitability of geological storage and utilization of CO$_2$ has been extensively studied [15]–[19] providing a solid understanding of geologic formations suitable for storage in the province and their characteristics; feasible storage options in deep saline aquifers, depleted oil & gas reservoirs, and utilization in enhanced oil recovery (EOR); and, in some cases, the storage capacity, injectivity and likelihood of leakage for specific locations [20].

The consensus is that the South-Eastern and Central regions of the province are the most suited for CO$_2$ storage. Aquifers have the largest capacity for CO$_2$ in Alberta but provide no economic benefit in comparison to enhanced oil reservoirs (EOR) and may be more costly relative to depleted reservoirs.

While more research is required to gain higher resolutions of the capacities, point locations and cost of implementation for storage, there is good assurance in Alberta’s storage resource where both experience and expertise are available in the province demonstrated in the proven
storage capacity in Quest project and the established Alberta Carbon Trunk Line (ACTL) which represents the world’s largest capacity pipeline for CO₂ transmission, able to transport up to 14.6 MtCO₂/year into EOR or saline aquifer storage.

1.3 Hydrogen Production Through Reforming Technology- A Literature Review

1.3.1 Reforming Technology

Generally SMRs have been branded as the lowest cost and most mature pathway for hydrogen generation [21]. However, there are other reforming technology options and the selection of the type of technology depends on several factors including the type of feedstock, required capacity for hydrogen generation and a particularly interesting factor which is the economics of generated by-products (steam and/or electricity). This work compares the two dominant reforming technologies which are Steam Methane Reforming and Auto Thermal Reforming.

**Steam Methane Reforming (SMR)**

In this process, natural gas feedstock initially passes through a pre-treatment unit where traces of sulphur contaminants are removed to prevent catalyst poisoning in the main reactor. A pre-reformer is included after pre-treatment when multiple feedstocks or a feedstock heavier than natural is anticipated or for large capacity SMR (above 100 ton/day) to transfer some of the reformer duty from the primary reformer to the pre-reformer. The same endothermic reaction occurs in pre-reformer and reformer where steam reacts with hydrocarbons using a nickel catalyst to produce syngas (a mixture of carbon monoxide and hydrogen) following the simple reaction path shown in Equations 1 and 2 [22]. Increasing the steam-to-carbon ratio (S/C) generally increases the overall conversion and minimize the risk of carbon deposition on reactor catalyst. The suitable S/C ratio is usually a trade-off question between conversion, process efficiency and economic value, as steam is energetically costly to produce.

\[
C_nH_{2n+2} + nH_2O \leftrightarrow nCO + (2n + 1)H_2 \quad \text{(for saturated hydrocarbons)} \quad [1]
\]

\[
CH_4 + H_2O \leftrightarrow CO + 3H_2 \quad \Delta H_r = +206\text{kJ/mol} \quad \text{(for methane)} \quad [2]
\]
\[ CO + H_2O \leftrightarrow CO_2 + H_2 \quad \Delta H_r = -41 \text{ kJ/mol} \] [3]

After exiting the reformer, syngas passes to the water gas shift reactor (WGS) where a moderately exothermic reaction takes place producing more H\(_2\) and CO\(_2\) following Equation 3. The reaction equilibrium favors lower temperature while reaching practical reaction rates requires high temperature [13]. Thus, typically two reactors are used in series: one at high-temperature (315-439 °C) followed by one at low-temperature (190-230 °C) [23]. Other configurations may include a single high or medium temperature reactor. For the same H\(_2\) production rate, SMR units with both high- and low-temperature shift reactors have less potential to generate steam (or electricity) than those with only a high-temperature shift [24].

In a modern SMR, the H\(_2\) rich stream then goes through a pressure swing adsorption (PSA) unit where high purity hydrogen is separated from CO\(_2\) and unreacted CO and CH\(_4\). These components along with unrecovered H\(_2\) constitute the tail gas stream, which is burnt in the reformer furnace to supply heat for the reaction and recover the surplus for steam and power generation in the power island of the plant.

*Autothermal Reforming (ATR)*

Autothermal reforming and partial oxidation (POX) are commonly used for the production of syngas for further conversion processes (e.g., methanol, ammonia) and are typically require additional power generation or steam export.

Autothermal reforming is the combination of steam methane reforming and partial oxidation, where the heat released from the exothermic POX reaction is used to heat requirements for the endothermic steam reforming reaction. As a result, an ATR plant does not require an external supply of heat, as is the case with SMRs. Otherwise, ATR has a similar process flow to that of SMR including the feedstock purification, pre-reforming, shift reaction and hydrogen recovery units. A notable addition to the process, however, is air separation unit needed to supply oxygen for the partial oxidation reaction. In the ATR reactor, natural gas reacts with the methane similar to Equation 1 and oxygen to produce the syngas as shown in Equation 4.
\[ CH_4 + \frac{1}{2}O_2 \leftrightarrow CO_2 + H_2 \ \Delta H_r = -41 \text{ kJ/mol} \]  

Due to the chemistry involved in ATR, the amount of CO produced in the syngas is higher than that produced in the normal reforming reaction. The lower H₂/CO ratio, in comparison to the SMR, results in increased load on the shift reactor. Therefore, a two shift reactor configuration is typically used in the ATR plants [24]. The tail gas exiting the PSA system in an ATR (without CO₂ capture) would generally be burnt in a fired heater to pre-heat the feed streams and provide additional heat to the co-generation unit in the plant to produce steam/power.

1.3.2 CO₂ Capture Options and Locations

Collodi [25] and Meerman et al. [26] studied the capture unit configurations SMR plants for hydrogen production, reviewing the possible technologies and assessing the level of CO₂ extraction, performance and cost implications. The three capture options for an SMR are capturing from the syngas (location 1), tail gas (location 2) or flue gas (location 3) streams or a combination thereof as indicated in Figure 3. Location 1 has the highest CO₂ concentration (15-35 mol%) and partial pressure, while location 3 has the highest flow rate and lowest concentration (5-10 mol%) and partial pressure. The overall CO₂ removal potential is around 60%, 55% and over 90% for locations 1, 2 and 3, respectively, as shown in Table 2. The general view preference in the technical literature is for pre-combustion capture in location 1 since, even though it has a reduced emissions reduction potential relative to location 3, it is a lower-cost alternative with less capital investment and lower energy consumption.

In the case of an ATR plant, the only option assessed in literature is downstream of the shift reactor before entering the PSA using the same capture technology as in the pre-combustion configuration in SMR. Because the ATR doesn't have an externally heated reforming furnace, the only source of emission within such a plant would be emissions from the fired heater. This aspect has lead to many championing the ATR for blue hydrogen production because 90% of the direct plant emissions can be captured at a similar cost to capturing 55% in the case of the SMR [24], [27].
However this overlooks an important aspect that has only been recently highlighted [24], [28].

In the case of ATR, addition of CO₂ capture and compression systems increases the thermal and electric energy requirement for the facility, which causes the demand for energy to exceed that available from the ATR facility. This shortfall must be met by importing power from the grid or burning additional natural gas in the fired heater. This is a key point and the implication of this aspect will be further demonstrated in this work.
Table 2: Overview of capture location characteristics and achievable overall CO₂ removal for the SMR process [25], [26]

<table>
<thead>
<tr>
<th>Stream</th>
<th>CO₂ Concentration (mol %)</th>
<th>Total Pressure (bar)</th>
<th>CO₂ Partial Pressure (bar)</th>
<th>Typical % overall CO₂ removal</th>
</tr>
</thead>
<tbody>
<tr>
<td>Location 1</td>
<td>15-35</td>
<td>20 - 31</td>
<td>3 - 11</td>
<td>60</td>
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<tr>
<td>Location 2</td>
<td>30-60</td>
<td>0.14 - 2</td>
<td>0.04 - 1.2</td>
<td>60</td>
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<tr>
<td>Location 3</td>
<td>5-19</td>
<td>~1</td>
<td>~0.05 - 0.19</td>
<td>90</td>
</tr>
</tbody>
</table>

1.3.3 TEA and LCA studies on Blue Hydrogen from Reforming Technologies

The life cycle emissions impact of H₂ production from SMR technology has been estimated to be 11.9 to 15 and 3.9 to 1 kgCO₂eq/kgH₂ without and with capture respectively [7], [29], [30]. These are cradle-to-gate estimates, and the variation is primarily due to different upstream emission rate assumption, different locations for capture, and performance of the capture process. Antonnini et al. [24] conducted a harmonized life cycle assessment of SMR and ATR with and without CO₂ capture using natural gas and biogas as feedstock. Their study indicates that ATR has an advantage over SMR with regards to overall CO₂ capture rates and shows that the incorporation of vacuum pressure swing adsorption (VPSA) could reach almost 100% removal of CO₂, however, this comes at the expense of an increased electricity requirement for ATR. The total GHG emission from natural gas configurations ranged from 2.6 kgCO₂eq/kgH₂ for ATR with VPSA (98% removal) to 5.8 kgCO₂eq/kgH₂ for SMR with pre-combustion capture (position 1 in Figure 3) with 90% removal from the stream. Without CO₂ capture, the emissions rates were around 10.8 kgCO₂eq/kgH₂ for both plants, with ATR being slightly worse than SMR. At high levels of capture, they concluded the feedstock (gas or biogas) supply chain dominates the climate impact.

Parkinson et al. [31] conducted a comparative review of the cost of carbon mitigation and other environmental metrics for 12 hydrogen production routes benchmarked against the SMR process. They highlight a trade-off between mitigation cost and the extent of decarbonization where fossil feedstock processes are considered the most cost effective with moderate levels of
Decarbonization when full supply chain emissions are considered. Their cost for SMR was adopted from an IEAGHG report [32], which compared 5 configurations for CO₂ capture in SMR. The IEAGHG report represents the most comprehensive and up to date techno-economic study conducted on SMRs and their possible capture options but does not look at life cycle emissions. Khojasteh et al. [27] compared a variation of SMR and ATR with solvent-based CO₂ capture along with chemical looping technologies noting a potential for the latter to be more attractive financially than SMR, however acknowledging scale-up and commercialization barrier.

There are no studies available that describe a framework which analyzes the trade-offs between different production and delivery options for hydrogen. Previous studies focused on comparing technologies within the plant fence and/or upstream emissions for these technologies. However, integrating all the pieces of a hydrogen supply chain provides a wholesome view of the available routes for hydrogen delivery. A gap which this study seeks to fill.

1.4 Research Questions and Objectives

The aim of this study is to provide context for discussion of future hydrogen production in Alberta by answering the following questions:

- What are the estimated costs of Blue Hydrogen production through reforming in Alberta?
- Does location and scale of production impact cost and technology preference?
- How much does it cost to move hydrogen in comparison to CO₂ in Alberta?
- What are the economic, technical and environmental trade-offs between the different routes?

The above questions will be addressed by carrying out the objectives below:

- Developing performance and cost models for commercially available technologies as a function of scale and estimating production costs at different locations within Alberta.
• Producing comparable hydrogen and CO₂ pipeline engineering and economic models to identify the baseline costs for these two fluids.
• Conducting a scenario analysis using the case study of Canadian oil sands projects.
2. Estimating Techno-Economic Performance and Environmental impact of Blue Hydrogen Production in Alberta

In this chapter the goal is to review previous studies made on cost analysis for steam methane reformers and develop a normalized cost model for an Albertan context that calculates the levelized cost of hydrogen production (LCOH) $/kgH₂. As such the model produces capital (CAPEX) and operational (OPEX) cost estimates as a function of the produced hydrogen set by the user and other economic inputs as illustrated in Figure 4.

![Flowchart description of cost and performance model inputs and outputs](image)

Figure 4: Flowchart description of cost and performance model inputs and outputs

There are generally five classifications for capital cost estimates used in process industries, which are class 1 and class 2 also called definitive estimates used for bidding and as final control against which actual cost and resource will be measured for budget control; class 3 which are budgetary estimates used to support funding request and as an initial budget until replaced by a more detailed estimate; lastly class 4 and 5 which fall under order of magnitude estimates [33]. The estimate employed in this study use the order-of-magnitude approach which utilize
cost information from previously built plants and provides estimate for a complete process, which are then adjusted using appropriate scaling factors and cost indices to account for capacity and inflation. According to the Association for the Advancement of Cost Engineering (AACE) class 4 to 5 estimates are considered preliminary estimates as they are based on limited information [34]. However, this approach can be an effective method when conducting comparative analyses between technologies or production options in order to contrast the costs and benefits of different technologies at various scales. It is therefore the method of preference found in most academic literature when conducting techno-economic technology assessments.

The following sections will describe how the CAPEX and OPEX estimates are produced from the model.

2.1 CAPEX Estimation Using Cost-To-Capacity Equations

This section describes the approach followed to develop a cost-to-scale CAPEX model and a scaling factor for steam methane reforming and auto thermal reforming technology specifically operated for hydrogen production.

2.1.1 Method Review and Consideration

This method is appropriate to produce order of magnitude estimates for entire plants and/or individual equipment, typically applied during the initial stages of a project to assess its economic viability before further planning is made. The working concept of this method is based on economies of scale where plant/equipment costs change with size raised to a certain exponent called a scale factor. It is governed by the concept that plants or equipment having similar technologies and different capacities or sizes have costs that vary nonlinearly [35]. Put in an equation form, the cost is a function of size raised to an exponent[36]. The general equation is [37]:
\[
\frac{C_2}{C_1} = \left( \frac{Q_2}{Q_1} \right)^x\] [5]

Where,

- \( C_1 = \text{Known cost for Plant 1 (or Equipment 1) with capacity } Q_1 \)
- \( C_2 = \text{Estimated cost for Plant 2 (or Equipment 2) with capacity } Q_2 \)
- \( Q_1 = \text{Known capacity for Plant 1 (or Equipment 1)} \)
- \( Q_2 = \text{Known capacity for Plant 2 (or Equipment 2)} \)
- \( x = \text{Scale factor representing technology of Plant 1 and 2 (or Equipment 1 and 2)} \)

For this method to give meaningful cost estimate results at different scales/capacities several requirements must be met [36]:

- The plant/equipment for which estimates will be scaled should have the same technology or be very close in nature with the facility/equipment having the known cost.
- Plant layout, configuration, site characteristics should be similar, cost adjustments should be made if such differences exist.

Other considerations include:

- Location differences, which should be accounted for using cost factors to include differences in regional labor rates, material, transport, site conditions costs, etc.
- Adding cost adjustments for unique process designs and uncommon site requirements different from scaled for technology.
- Cost inflation adjustments, where the known historical cost with a specified reference year is updated using cost indices applicable to the technology in question to develop correct cost estimates for the required year.

If the actual process or plant conditions are not consistent with the underlying data and assumptions, as listed above, the results may not be reflective of real-world costs. These considerations should also be satisfied when using cost and capacity data to develop appropriate scale factors as will be discussed later in this chapter.
Hence, these considerations limited the number of sources that can be cited to those specifically addressing reforming processes that use natural gas as feedstock to produce Hydrogen as a final product.

### 2.1.2 Literature Review for Cost Data and Scale Factor Development

Over 25 papers with cost data on SMRs were reviewed, most of these sources used cost information that were based on either data from previous literature, updated historical models, or industry information which were cited in the respective sources but not described in detail. Only the original sources of cost information were investigated and adopted to reduce the possibility of compounded deviations. Most of these studies were from the 2000s, as such the influence of cost escalation and the discrepancy caused by sources adopting different inflation indices on the final cost estimates should not be significant.

The costing index used in this work was the Chemical Engineering Plant Cost Index (CEPCI) as it is a suitable index for chemical plant processes. Initial costs used to develop cost to capacity relation were updated to 2018 and the costing equation was then updated to 2020.

The variations present in costs presented in the reviewed literature were normalized. Some of these variations pertained to the aforementioned considerations of technology, configuration, location and cost dates. Others were related to capital cost aggregation methods and nomenclature of cost items. For ease of comparison, a consistent cost accounting method was used that follows the quality guidelines for energy system studies report by the US Department of Energy National Energy and Technology Lab [38]. A description of the cost elements and the aggregation method is provided in Appendix [A], and is shown in Figure 5. The cost elements in reviewed literature were fitted as best as possible in this framework, recognizing and adjusting for instances where elements were summed up in the cited source (e.g., EPC and start-up summed up in NREL 2002) or when low-level elements substituted higher ones due to lack of information provided in the source (e.g., direct equipment capital was used in place of bare erected cost, dropping the installation cost in Argonne 2003).
Variations that could not be adjusted for (e.g., literature having process flows deviating from the standard) were highlighted. Differences in operating parameters such as unit efficiencies, conversion rates, recovery rates and battery limit specifications were recorded when possible, this exercise lacked consistency as not all sources delivered sufficient level of detail for their operating parameters.

Efficiencies were conveyed when mentioned in the cited reference, otherwise, conversion rates and recovery factors were calculated from stream information when provided.

The process layout, operating parameters, and cost aggregation methods followed in each source are provided in Appendix [B], where the original nomenclature used in these sources is provided. The level of detail in the reported information varied within a reasonable range. Of the literature survey conducted seven studies contained sufficient detail to re-analyze their cost estimates.

Foster Wheeler (1996) [39] and IEAGHG (2017) [32] cost estimates were based on a self-power generation design. These estimates would be reduced by 10% and 20% for Foster Wheeler and
IEAGHG respectively on the assumption that all power is to be imported. Table 3 summarizes the breakdown of SMR plant cost estimates obtained from the literature.

For Foster Wheeler (1996) and IEAGHG (2017) the as reported costs were on the original study’s location (Netherlands) basis, the 2018 costs were converted to the US Gulf Coast (USGC) basis using an updated location factor [40]. Several sources did not provide the plant location and the cost estimates of these sources were assumed to be on the Gulf Coast basis. In Argonne (2003) the Direct Capital Cost (cost of equipment) was reported as BEC neglecting the installation cost, the TPC included inventory costs which are typically under owner’s cost and should be categorized as TOC however there was no mention of working capital or start-up costs and hence the estimate was kept as TPC. In all the NREL (2002) reports and Foster Wheeler (1996), the BEC included H₂ compressor costs. NREL (2006) and NETL (2013) gave no mention of pre-reformers in their configurations.

Economies of scale are observed in Table 4 and Figure 6 where the BEC of equipment per unit of production reduces with the increase in production capacity.

To derive a representative scale factor using the obtained data a regression analysis was performed on the natural logarithms of the cost and capacity data where the slope of the regression line represented the scale factor for the SMR technology [36].

The linear model was a good fit with an $R^2$ value of 0.99, where the initial scale factor produced when all studies were included was 0.74 with a 95% confidence interval of (0.70-0.78) as illustrated in Figure 7. As the known plant costs should represent similar configurations to those to be predicted, the scale factor should not be derived using the entire capacity range as smaller SMR facilities have significantly different layouts than the larger ones. For example, for a small-scale plant, the reformer is based on a heat exchanger design having a hot stream from a burner as the heat source for the reaction, while the mid- and large- capacity plants use a configuration of tubular furnaces, also, cooling is typically performed by air coolers and condensers in small scale plants while that is achieved primarily by heat exchange with other plant streams in the mid-large scale plants [41]. It worth mentioning that both NREL (2006) (point N) and the largest scale for NREL (2002) (point M) studies are larger than the biggest
SMR plants currently in operation, therefore they were also excluded from the regression. Adjusting the range to account for this aspect produces a scale factor of 0.79 (with a model R² of 0.97) and a 95% confidence interval of 0.68-0.91 as shown in Figure 8. The intercept in the regression model represents the fixed cost associated with a plant of zero capacity.

Table 3: Summary of cost aggregation for SMR plant as reported in literature and adjusted to reflect 2018 costs in USD

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<tbody>
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<td>Capacity (kgH₂/d)</td>
<td>480</td>
<td>1,000</td>
<td>24,000</td>
<td>100,000</td>
<td>150,000</td>
<td>500,000</td>
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<td>Gulf Coast</td>
<td>Gulf Coast</td>
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<td>15,960,000</td>
<td>44,531,275</td>
<td>56,760,000</td>
<td>185,135,530</td>
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<td>182,214</td>
<td>1,330,000</td>
<td>3,710,940</td>
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<td>3,710,940</td>
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<td>68,585,000</td>
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<td>54,550,812</td>
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<tr>
<td>Capacity (kgH₂/d)</td>
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<td>115</td>
<td>120,500</td>
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<td>Capacity (kgH₂/d)</td>
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<td>325,803</td>
<td>45,041,075</td>
<td>165,150,610</td>
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<td>N/A</td>
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<td>325,803</td>
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| IDC               | 16,515,061  |
| TCR               | 196,514,212 |
Table 4: BEC of Equipment (2018) and production capacities for the reviewed literature

<table>
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<tr>
<th>Label</th>
<th>Reference</th>
<th>Capacity (kg H2/day)</th>
<th>BEC USD$</th>
<th>BEC/Capacity</th>
<th>Ln (Capacity)</th>
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<td>325,803</td>
<td>2,833</td>
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<td>F</td>
<td>NREL 2002</td>
<td>100,000</td>
<td>67,888,807</td>
<td>679</td>
<td>11.51</td>
<td>18.03</td>
</tr>
<tr>
<td>G</td>
<td>Argonne 2003</td>
<td>120,500</td>
<td>45,041,075</td>
<td>374</td>
<td>11.70</td>
<td>17.62</td>
</tr>
<tr>
<td>H</td>
<td>NREL 2002</td>
<td>150,000</td>
<td>86,531,739</td>
<td>577</td>
<td>11.92</td>
<td>18.28</td>
</tr>
<tr>
<td>I</td>
<td>IEAGHG 2017</td>
<td>215,856</td>
<td>135,321,019</td>
<td>627</td>
<td>12.28</td>
<td>18.72</td>
</tr>
<tr>
<td>J</td>
<td>Basye &amp; Swam. 1997</td>
<td>254,360</td>
<td>165,150,610</td>
<td>649</td>
<td>12.45</td>
<td>18.92</td>
</tr>
<tr>
<td>K</td>
<td>NREL 2002</td>
<td>500,000</td>
<td>282,242,765</td>
<td>564</td>
<td>13.12</td>
<td>19.46</td>
</tr>
<tr>
<td>L</td>
<td>Foster Wheeler 1996</td>
<td>606,690</td>
<td>250,056,673</td>
<td>412</td>
<td>13.32</td>
<td>19.34</td>
</tr>
<tr>
<td>M</td>
<td>NREL 2002</td>
<td>1,102,041</td>
<td>514,068,049</td>
<td>466</td>
<td>13.91</td>
<td>20.06</td>
</tr>
<tr>
<td>N</td>
<td>NREL 2006</td>
<td>1,109,160</td>
<td>314,302,435</td>
<td>283</td>
<td>13.92</td>
<td>19.57</td>
</tr>
</tbody>
</table>

Figure 6: Economies of scale represented by the Bare Erected Cost per unit of production
Acknowledging the above-mentioned variation in plant configuration and cost aggregation, the produced economies of scale trend gave a consistent trend as can be noticed from Figures 6, 7 and 8 affirming the validity of the normalization approach that was followed in this study.

Figure 7: Regression line for the natural logs of cost versus capacity using entire capacity range

Figure 8: Regression line for the natural logs of cost versus capacity using mid- and large-scale capacities
2.1.3 Scaling Factor Comparison Against Available Literature Data

Several literature sources have provided scaling factors for the SMR process, it was noticed that scaling was made on varying bases. Yang and Ogden [42] used a scale factor of 0.70 for “capital cost of SMR plants” scaled on H2 production capacity (tons H2/day), no detail was given about the specifics of this scale factor however a note that start-up costs (e.g., land, engineering) represented 51% of the capital was mentioned. Using this scale factor gave costs close to the BECs documented in this study for certain capacities. Jiang [43] gave scale factors for BEC of individual units, steam methane reformer and pre-reformer both had a scale of 0.67 scaled on feed rather than produced Hydrogen capacity and the PSA a scale of 0.55 on production capacity. Hamelinck [44] derived scale estimates from other literature and gave individual unit cost scales of 0.60 for reformer units, 0.85 for shift and 0.70 for PSA. Towler and Sinnott [40] referenced an Amec Foster Wheeler process giving a scale of 0.79 for ISBL (costs) costs which is very close to the scale produced in this study’s regression.

A capital cost function that can be generated from the findings of this section:

\[
BEC \text{ Cost of SMR plant} = 135,321,019 \times \left( \frac{\text{Capacity} \left( \frac{kgH_2}{\text{day}} \right)}{215856} \right)^{0.792} [6]
\]

The reference cost and scale used represented a middle in the range used to produce the regression equation.

The above equation scales the bare erect cost of the SMR unit in 2018 US dollars on a USGC basis taking the produced hydrogen in (kg/day) as a scaling parameter. The equation would be used for a specific scaling range of 24,000 to 607,000 kg/day.

As there is no sufficient cost data on solvent capture units at different scales [45] a similar procedure to the one followed with the case of SMR was not feasible, hence the scaling exponent was set to 0.70 and the parameter as the inlet gas flow to the capture unit in kg/hr while CO\textsubscript{2} compression and drying were scaled on a 0.61 exponent [46].
Although using a similar technology to that of the SMR, the ATR has significant configuration differences as discussed earlier. In particular, an air separation unit is included as well as main reformer that has no external heat input and two water gas shift (WGS) reactors instead of one in the case of the SMR. The SMR cost model was adjusted to account for these differences to create an ATR capital cost model. Four scaling relationships calculate the installed cost of the individual reformer and WGS reactors for both the ATR and SMR, a fifth scaling relation is used to estimate the cost of the air separation unit. Adding the sum of the cost differentials and the ASU unit to the bare erect cost equation in the model gives an estimate of the ATR plant cost. Scaling factors, cost information and referenced sources are provided in Appendix C.

All reference base costs were normalized and updated to Gulf Coast 2020 cost basis and an overall weighted average location factor is used to adjust capital costs to Canada using data sourced from a previous report which will be discussed in the following section.

The Total Plant Cost (TPC) is obtained by summing the Bare Erect Cost, EPC and contingency costs. The process contingency was taken as 10 and 35% of the BEC for reforming units and capture units respectively, while the project contingency was taken as 40% of the sum of BEC EPC and process contingency. The Total As-Spent Cost (TASC) was then evaluated and following the NETL methodology [38]. More discussions on the values chosen for the cost items are provided in Appendix A.

### 2.1.4 Location Factor

Multiple parameters influence the cost of building plants in different location such as material, fabrication and infrastructure construction costs, labor cost and availability, feedstock and other fuels and consumable expenses, the ambient condition, contingency costs and other economic parameters [47]. Location factors are used to account for these differences in cost. The majority of cost data for equipment is given on USGC or Northwest Europe (NWE) as they are the hubs of the chemical industry for which most data is available. Hence the location factor is applied to estimate the cost in a specific location using Equation 7 [40].
Cost of plant in location $A = \text{cost of plant on USGC} \times LF_A$ \[7\]

Where $LF_A =$ Location factor for $A$ relative to USGC basis. These factors are strongly influenced by currency exchange rates therefore changing with time.

Hence a location factor is used to convert results in Equation 6 which are on USGC basis to an Alberta basis.

Using location indices for CCS processes, the material, equipment and labor (productivity and cost) factors for a 2018 plant in Canada are 1.18, 1.10 and 2.365 respectively [47], [48]. Since the cost item being scaled for (BEC) inherently embeds the equipment, material, labor, construction and supporting facilities costs [49] a single adjusted cost factor, weighing each of these factors according to their percentage in the BEC, can be used. Guthrie [50] gives info for chemical process equipment on the cost of material and labor as a percentage of equipment cost, where material cost is 71.4% and labor is 63% of the total equipment cost. Putting these as percentages of BEC, equipment, material and labor comprises 43%, 30% and 27% of the bare erected cost respectively. Taking this weighting scale and utilizing the above location factors after updating to 2020 an overall location factor to be multiplied in equation 1 is 1.48, another location factor for Fort McMurray was adopted from Towler and Sinnott [40] and updated to 2020 years giving 1.76 which will be utilized in Section 4.

2.2 Performance and Operating Expenses (OPEX) Estimation for Reforming Unit with CO₂ Capture

This section describes how mass and energy balances derived from fundamental thermodynamics and literature reported operating costs were incorporated in the techno-economic model to allow for screening between the different routes. Both the SMR pre- and SMR post-combustion capture options (locations 1 and 3 in Figure 3) were based on mass balance equations from the IEAGHG 2017 report [32], while the ATR with capture model adopted mass balance data from Skrenegne [51].
2.2.1 SMR with Pre-Combustion Carbon Capture Performance Estimation

As previously discussed, the baseline SMR process layout includes the pre-reformer, reformer, high-temperature water gas shift, and pressure swing adsorption (PSA) units. In the pre-combustion configuration, the carbon capture amine unit is situated upstream of the PSA as shown in Figure 9. The capture unit uses a liquid solvent (MDEA) which reacts with CO₂ forming new product components that reverse and release the acid gas during the solvent regeneration process; it has low energy requirements relative to other solvent systems however it is limited to high pressures streams (typically higher than 2 MPa)[52].

The main input to this model is hydrogen produced in (kg/day). The SMR conversion rate, PSA recovery, carbon capture efficiency, and water gas shift reactor conversion rate can be adjusted as well, but this would require calculating the composition of the Shifted Syngas and Treated Syngas streams into the model accordingly. In this case IEAGHG reported compositions were used. The parameters values used are provided in Appendix H.

By setting the input parameters mentioned above the model back calculates the molar and mass flowrates for the inlet stream to the PSA (Treated Syngas), the inlet stream to the amine unit (Syngas to CC) and the outlet stream from the WGS reactor (Shifted Syngas). The calculations are based on fixed molar compositions as stated before.

The SMR block encompasses both the reformer and pre-reformer reactors; the conversion rate set in the input section is the collective methane conversion for both units. For simplification the model was based on a 100% methane feed, though this assumption is not realistic, it gives sufficiently accurate mass flow and CO₂ emission rates required for cost assessment. An equation relating the hydrogen produced and the total natural gas consumption (with typical composition, 89% Mole CH₄) on an energy basis was regressed based on literature values, which is discussed in a following section.
Natural gas fuel consumption was set to 20% of the feedstock on a mass basis following the IEAGHG report’s set-up. The SMR unit inlet natural gas (NG) and outlet (Syngas) streams were calculated stoichiometrically based on the Shifted Syngas stream mole composition, WGS and SMR conversion rates.

A gate-to-gate CO₂ emission rate is calculated based on the CO₂, CO and CH₄ content in the inlet stream to the carbon capture unit and the fuel streams. The amount of CO₂ captured is based on the CO₂ removal efficiency, the Treated Syngas and Shifted Syngas streams mole compositions.

2.2.2 SMR with Post-Combustion Carbon Capture Performance Estimation

The implementation was similar to that in the previous section. The only difference is in the location of the carbon capture unit, where in this configuration the amine unit captures the CO₂ from the flue gas stream post it’s combustion in the furnace as shown in Figure 10. The solvent used in this model is an MEA solvent as opposed to the MDEA used for SMR-Pre which shows
high reactivity with CO₂ and is suitable for low pressure application hence it is used for the flue gas stream however it has high heat requirements for regeneration.

2.2.3 ATR with Pre-combustion Carbon Capture

As previously discussed the layout is similar to that of the SMR, but the ATR has a different reformer unit design which doesn’t require external heat input (from fuel gas), it also utilizes two reactors in the WGS section, high-temperature and a low-temperature shift reactor, which is necessary due to the chemistry involved [53]. Like the first SMR model, a pre-combustion carbon capture unit is situated upstream of the PSA unit. It should be noted that the only source of direct CO₂ emissions is from the tail gas combustion in the power island for energy generation. An air separation unit supplies oxygen to the ATR for partial reforming Figure 11.
Operating Cost Estimation

The operating expenses were divided into fixed and variable operating costs. The fixed costs were calculated as percentages of the labor cost [40], which were also adjusted to Canada costs using a labor location cost factor [47] (see Appendix [D]).

The variable costs consisted primarily of the cost of natural gas feedstock and fuel for furnaces, power consumption for the reforming plants, CO₂ capture and compression units, as well as hydrogen compression at the battery limit to raise from 2.5 MPa to the desired pressure when needed (in such cases a H₂ compressor is included in the capex model as well); it also included other consumables such as water, catalyst and chemicals. Catalyst and chemicals were taken as fixed a percentage of the TPC; costs of water consumed in the reforming unit were estimated from the mass balance.

A regressed relation was used to estimate the natural gas feedstock and fuel requirement as a function of produced Hydrogen using multiple literature sources [7], [32], [39], [54]–[56]. A graph of the regressed data points is shown in Figure 12. The linear model was a good fit with an R² value is 0.99, and a confidence interval of (0.0062–0.0068) at the 95% significance level.
A relation between the incremental amount of natural gas required to meet the thermal and electric demand of a carbon capture configuration was not evident when surveying literature.

Published studies used different modelling tools, operating parameters, capture systems, and did not always report key data that made developing a correlation difficult. Hence, an assumption of a 4% and 9% increase in natural gas consumption was made for the SMR pre- and post- combustion capture respectively based on the IEAGHG models [32] to account for the energy needs of the carbon capture and compression units while ATR was kept at the base consumption and the incremental demand was satisfied through power import as suggested by the Antonini et al. model [53]. Further discussion will follow in the coming section.

A second regression equation was used to estimate the power consumption of an SMR unit to balance the electricity available for the plant (this excludes any compressing requirement for hydrogen at the battery limit over 2.5 MPa hence some sources were excluded) as shown in Figure 13. The intercept was forced to zero, and the model was a good fit with an $R^2$ values of 0.98 and a confidence interval is (0.0055-0.0139) at the 95% significance level.

The increase in power consumption due to the addition of decarbonization facilities is represented by the capture and compression units power requirements. The power
requirement for the capture units using an MDEA technology was estimated to be 0.0125 MWh/tonne CO₂ by referring to the IEAGHG report and the IECM model [32], [57]. While the SMR post combustion unit uses an MEA technology with a power demand averaged at 0.025 MWh/tonne CO₂ captured [58].

![Figure 13: Regression equation for SMR power consumption per hydrogen production rate](image)

To estimate the power of compressor units for hydrogen and CO₂ isentropic work equations were used incorporating a properties tool package (Coolprops) to assess the thermodynamic properties of the fluid of interest. The equations and input parameters are listed in the appendix and further discussion on the Coolprops tool package will follow in the next chapter. For the ATR extra power costs for the ASU is evaluated at the rate of 256 kWh/tonne O₂[53].

*Energy and CO₂ Emission Balance*

Without carbon capture, both the SMR and ATR plants can produce surplus power/steam that can be exported over the fence of the plant. SMR plants in particular have significantly higher thermal output because of their furnace-reformer configuration which makes SMRs the “working horse” in a refinery converting this surplus energy into either into steam and/or
When carbon capture and compression facilities are added, this surplus in steam or power is redirected to meet the added unit thermal and electric requirements. In the case of SMR plants, both pre and post-combustion capture configurations require minimal fuel addition to satisfy these new energy demands but will have reduced energy export in comparison to an SMR without capture and hence the incremental energy input will have little environmental implications provided that hydrogen compression at the battery limit of the plant is not required [32], [53].

In contrast, the ATR breaks will require significantly higher increments in energy input when a CO₂ capture configuration is enabled as highlighted by Antonini et al. [53]. This is expected since the plant has higher power consumption in comparison to the SMR due to the ASU unit and at the same time limited steam is available to meet electric and thermal demand since a furnace-reformer configuration is not required in ATRs which is also why an ATR is generally used for syngas production rather than hydrogen. The possible options to meet the demand would need incremental natural gas or oxygen supplied to the cogeneration unit or importing power from the grid.

Due to the multiple parameters involved and possible configurations for the cogeneration unit, the amount of steam required (and, thus, natural gas and oxygen) is speculative without a detailed simulation model. The source of this power will not only have an economic impact but an environmental one as well, using an oxyfuel combustion configuration would increase the oxygen consumption (and amount of CO₂ captured per tonne H₂) [27] while using a fired heater configuration would increase the direct CO₂ emission rate and the natural gas consumption [53]. Using the grid power means crediting its carbon intensity to the produced Hydrogen reducing the avoided CO₂ potential in the case of a high carbon intensity grid such as Alberta.

The model used in this study accounts for this aspect by adopting the power generation abilities reported in literature and subtracting power consumption estimates for the respective units. In instances when power import is required, a grid calculator credits both the cost of electricity imported and the carbon intensity credited per kg of hydrogen taking Alberta’s grid intensity of
680 gCO$_{2eq}$/kWh (in 2018) [11]. By the same token in export cases the surplus power displace the 680 gCO$_{2eq}$/kWh and the avoided emissions are also credited per kg of H$_2$.

Upstream natural gas emissions were included in the analysis as they contribute significantly to the life cycle emissions of the plant. Estimates for these emissions vary widely in literature driven by varying leakage rates at the different segments of the natural gas supply chain [60]. Another influential factor when generally conducting life cycle emissions is the choice of the global warming potential GWP standard used to measure the impact of green house gases. Where choosing between a 100 year and 20 year basis would significantly change the estimated emissions, a 20 year GWP (GWP 20) attaches more weight on the impact of short-lived gases such as CH$_4$ and the difference between these standards can be major leading to different conclusions about technology performance [24], [61]. In this thesis a GWP100 standard was used to measure the impact of greenhouse gases and an upstream emissions rate of 4.2 gCO$_{2eq}$/MJ of natural gas was adopted which represents average data provided by a low-emission production company in West Canada, it is worth noting that this rate is significantly lower than other rates recorded in literature models, where B.C. rates have an average of 6.4-6.8 gCO$_{2eq}$/MJ while the US average is 13.8 gCO$_{2eq}$/MJ [62], [63]. Due to this variability the effect of upstream emissions on the levelized cost will be assessed.

### 2.3 Model Results: Cost of Blue Hydrogen in Alberta

Using the described tool, the cost of hydrogen production from the investigated routes is estimated for Alberta.

#### 2.3.1 Levelized Cost and Carbon Footprint Hydrogen Production

The equation for calculating the levelized cost of production LCOH is provided in Appendix F. The main input values used for the case study are shown in Table 5.
Table 5: Input parameters for Alberta case study

<table>
<thead>
<tr>
<th>Input Parameter</th>
<th>Unit</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrogen Production Rate</td>
<td>kgH₂/day</td>
<td>350,000</td>
</tr>
<tr>
<td>Battery Limit Pressure</td>
<td>MPa</td>
<td>10 (H₂)/11 (CO₂)</td>
</tr>
<tr>
<td>Grid Emission Intensity</td>
<td>gCO₂eq/kWh</td>
<td>680</td>
</tr>
<tr>
<td>Discount Rate</td>
<td>%</td>
<td>8</td>
</tr>
<tr>
<td>Natural gas price</td>
<td>CAD/GJ</td>
<td>2.83</td>
</tr>
<tr>
<td>Power Price</td>
<td>CAD/kWh</td>
<td>0.07</td>
</tr>
<tr>
<td>Plant Life</td>
<td>year</td>
<td>25</td>
</tr>
<tr>
<td>Upstream Emissions</td>
<td>gCO₂eq/MJ NG</td>
<td>4.2</td>
</tr>
</tbody>
</table>

The production rate represents a typical reforming plant capacity, the natural gas and power prices are reflective of the current utility prices in Alberta for industrial use. Pressures were set to satisfy pipeline requirements for long-distance transmission and the grid intensity is the provincial estimated for Alberta in 2018 as reported in the National Inventory report[11]. Figure 14 illustrates results from the model. The whiskers reflect a change in natural gas price from 4.25-1.42 CAD/GJNG. The SMR plant with pre-combustion capture represents the lowest cost alternative for production of hydrogen (absent any emission cost) but results in the highest direct GHG emissions. SMR with post-combustion capture and ATR with capture have markedly lower direct carbon intensities (1.0 to 1.3 kgCO₂eq/KgH₂), however these are more expensive investments compared to SMR with pre-combustion capture because of the larger capture units and ASU installation requirement. Interestingly, while the SMR with post-combustion capture has a larger capture system, its overall production cost is comparable to the ATR option due to the bigger capital investment of the main plant. With regards to costs the estimates are within range of other cost estimates for Canada, a study in British Columbia gives an estimate of 2 $/kgH₂ for a 100 t/day plant removing 80%CO₂, while an Asia Pacific study gives a cost of 1.14 $/kgH₂ for a 380 t/day plant however the details of capture were not stated and the plant assumed a 70% capacity factor and 3.4 US$/GJ price on natural gas [64], [65].
The direct carbon intensities are comparable for ATR with capture and SMR-Post with the ATR alternative having slightly higher emissions as the main plant requires more natural gas feedstock to produce the same amount of hydrogen owing to the reaction chemistry as previously discussed, in addition the emissions from the ATR’s power island are not captured. Both direct and life cycle carbon intensities are comparable to results from Antonini et al. [53] and Parkinson et al. [31], the ATR without capture is a worse alternative to the base SMR this is due to the increased CO₂ production owing to the reaction chemistry as previously discussed.

Figure 14: Levelized cost (US$/kgH₂) and carbon intensity (kgCO₂eq/kgH₂) of blue hydrogen from reforming units in Alberta (the bars are read against the left axis and the dots are read against the right axis) whiskers reflect sensitivity due to change in natural gas price between 4.25-1.42 CAD/GJ

In the results shown above, hydrogen is being compressed at the battery limit to a pressure sufficient for pipeline transport [66]. Thus, all pathways require power import with the selected power island configuration. When including the carbon intensity of the Albertan grid (680 gCO₂eq/kWh) this increases the overall emissions for the processes, making the environmental advantage of SMR with post-combustion over ATR more pronounced. This shows that ATR is a less favorable choice in a high carbon-intensity grid due to its reduced ability for power
generation in comparison to SMRs. However, should there exist a clean power source ATR may become more advantageous at larger scales, as will be indicated in following results, provided the plant has an optimized design for hydrogen generation.

When a similar analysis was made without H$_2$ compression at the battery limit the combined (Direct+Grid) intensity was slightly lower than the direct emissions intensity for the SMR with pre and post combustion capture as some emissions were off-set by export credits as discussed in a previous section, while for the ATR with capture the combined intensity remained significantly higher than the direct emissions.

Accounting for the upstream emissions and the grid credits/charges the life cycle carbon intensities for the case study were 4.71, 1.48 and 2.47 kgCO$_2$/kg H$_2$ for SMR pre-combustions, post combustion capture and ATR with capture respectively.

2.3.2 Cost of Avoided CO$_2$

The cost of avoided CO$_2$ represents the incremental capital and operational costs that are borne by a facility for every ton of CO$_2$ captured and stored. It is calculated by comparing the CO$_2$ emissions and the levelized cost of a plant with capture against the emissions and production cost of a reference plant without capture. In this analysis the reference plant was an SMR without capture as it represents the lowest-cost alternative for hydrogen production. The equation used to calculate this cost is provided in Appendix F. The costs calculated using only direct emissions do not reflect the best option as it neglects emissions that may have resulted outside the system which as indicated in the previous figure can be significant. For comparisons the cost of avoided CO$_2$ based on Direct, Direct+Grid and Direct+Grid+Upstream emission are shown in the table below. As can be noted when grid and upstream emissions are included the ATR becomes a less attractive alternative.
Table 6: Cost of avoided CO₂ CAC in US $/tCO₂ for reforming plants producing 350 tH₂/day

<table>
<thead>
<tr>
<th>Production Route</th>
<th>CAC Direct Emissions</th>
<th>CAC Direct + Grid</th>
<th>CAC Direct + Grid + Upstream (4.2 gCO₂e/MJ NG)</th>
</tr>
</thead>
<tbody>
<tr>
<td>SMR with pre-combustion capture</td>
<td>66</td>
<td>75</td>
<td>76</td>
</tr>
<tr>
<td>SMR with post-combustion capture</td>
<td>74</td>
<td>80</td>
<td>80</td>
</tr>
<tr>
<td>ATR with capture</td>
<td>75</td>
<td>89</td>
<td>90</td>
</tr>
</tbody>
</table>

2.3.3 Sensitivity Analysis

To assess the parameters to which the cost of production is most sensitive, an analysis was conducted by varying individual cost inputs one at a time and recording the percentage change in the levelized cost from the initial numbers shown in Figure 15 illustrates this analysis for the post combustion capture option. This figure shows that production costs are to be most sensitive to the location factor as it can significantly impact the capital investment cost. Changing the location factor to 1 (reflecting a U.S. Gulf Coast location) reduced the cost by over 20%. While assuming plant construction in Fort MacMurray increased production cost by 14%.

Other strong drivers of change were the cost of natural gas (as previously indicated) and the interest rate, as one significantly impacts operational cost and the other the economics of the capital investment.

As high-pressure compression of hydrogen for high flow rates is still under development, the efficiency of the compressor model was varied to reflect the impact of this uncertainty. Power prices are not expected to change drastically in Alberta and as such variation in the compressor efficiency didn’t have a significant impact on the cost of production.
The SMR with pre-combustion capture and the ATR with capture exhibited similar responses to the varied parameters and the results are attached in Appendix G.

![Figure 15: Sensitivity analysis on SMR with post combustion capture based on varying several input variables (0% reflects base case of 65% for H₂ compressor efficiency, 8% for interest rate, 2.83 CAD/GJ for NG price)](image)

2.3.4 Impact of Scale on Technology Choice
The scale of production will have an impact on the choice of technology where at higher scales ATR appears to be a lower cost alternative in comparison to SMR-Post as indicated in Figure 16. This trend is supported by findings in other literature sources [56], [57]. This result is caused by a combination of factors including the relatively less expensive ATR main reforming reactor in comparison to the SMR furnace configuration, the positive economies of scale of the ASU, and the increasingly larger cost and size of the post combustion capture unit for the SMR where it handles larger gas volumes. From the figure below it can be concluded that cost of blue hydrogen in Alberta will be in the range of 1.2 to 2 $/kg for large scale plants which falls within the expected range for Canada mentioned in other literature [2].
This impact of economies of scale is even greater felt when looking at the cost of avoided CO\textsubscript{2} in reference to a base plant SMR. Figure 17 shows that the marginal cost for capturing CO\textsubscript{2} reduces at higher scales indicating the economic benefit at capturing CO\textsubscript{2} at such scales. A crossover as the one observed in Figure 16 between the SMR-Post and the ATR-Pre is not seen for the studied range owing to the lower cradle-to-gate intensity of the SMR-Post which make the marginal cost spent on that alternative more effective.
Effect of Grid Intensity on Overall Emissions

The specific emissions from each technology will be impacted by the grid intensity as discussed in a previous section owing to the aspect of CO₂ credits and charges, however as the government of Alberta is aiming to reduce the overall grid intensity by phasing out coal-plants, deploying more natural gas power plants and incorporating renewables for power generation, this means that the current intensity is expected to decrease. To illustrate the effect of this aspect Figure 18 shows a carbon tax of 100 US$/tCO₂ applied on the cradle-to-gate emissions (which includes upstream, direct and grid credit or charge emissions) of each technology while varying the grid intensity from 0 gCO₂eq/kWh (net-neutral grid) to 800 gCO₂eq/kWh. The results show that generally, as the grid intensity reduces plants that would benefit from a carbon credit exhibit a reduction in that advantage and plants importing more power from the grid experience less impactful expenses. For the case study shown since H₂ is being compressed at the battery limit minimum power is available for power in the case of a baseline line SMR while all the plants incorporating capture units import power from the grid (recalling the results of Figure 14). Hence as the grid intensity reduces the baseline SMR becomes slightly more
expensive as it loses the ability to credit CO₂ on power. Since ATR with pre-combustion capture imports the most power it experiences the largest reduction when a cleaner grid is introduced, where at 200 gCO₂eq/kWh the ATR-Pre becomes lower cost than an SMR with pre-combustion capture. These crossover points would change should a different scale or carbon tax be used but the general impact of this aspect would be maintained.

Figure 18: Impact of varying grid intensity on the levelized cost of production for a plant capacity of 350 tH₂/day (a carbon tax of 100 US$/tCO₂ was applied on the cradle-to-gate emissions of each technology, y-axis starts at 1.75 US$/kgH₂)

Effect of Upstream Emission Intensity on Overall Emissions

Figure 19 shows that increasing the upstream emissions to 28 gCO₂eq/MJ from the adopted 4.2 gCO₂eq/MJ rate translates to a 20-22% increase in levelized cost of production when applying a 100 US$/tCO₂ carbon tax on the life cycle emissions.
These results highlight the significant impact of upstream emissions and how it can influence cost. This stresses the importance of having a low-intensity natural gas supply to production units to reap the environmental benefit of blue H₂ and avoid additional cost[68].
3  Comparing Costs of Moving H₂ and CO₂ in Alberta

This chapter will describe how the hydrogen pipeline model was developed from the McCoy and Rubin CO₂ pipeline model [69], the adjustments made to the original CO₂ and new hydrogen models to shift to an Alberta cost, and present a framework that compares both models’ pipeline costs.

The transmission cost assessment considered in this section is viewed as an extension of the analysis conducted by Saadi et al.[70] which compares the cost of transporting several energy forms (natural gas, oil, hydrogen and electricity) showing that alternative energy carriers like hydrogen and power have relatively higher transportation cost which could be reduced by incentivizing these systems using carbon pricing policies.

3.1  Review of Available Models and H₂ Pipeline Considerations

To the best of our knowledge there is no study that provides a transport framework comparing the cost of CO₂ and hydrogen transmission in Alberta. The closest work was by Olateju and Kumar [71], who estimated CO₂ and H₂ pipeline costs as part of a comparison between underground coal gasification and SMR with capture technologies for hydrogen production. This study did not set out to contrast the pipeline performance and cost for the two fluids and, as a result, they used fundamentally different models for CO₂ and H₂ pipelines. For CO₂, they used the techno-economic pipeline model from McCollum and Ogden [72] and, for H₂, they calculated pipeline diameter using the Panhandle-B equation (developed for natural gas) and pipeline cost using the work of Yang and Ogden [73]. These different approaches make it difficult to conduct a normalized comparison between the two transmission options, as it is not clear if the method of costing influences the results more than the properties of the fluids being transported.

Prior work has established that pipeline cost is a strong function of the pipeline diameter, however models available from literature use different diameter equations, incorporate varying input factors (e.g. terrain, system analysis over time) and different assumptions for the design inputs (e.g. temperature, density, velocity, friction factor equations and distance between
booster station) making the choice of the model highly dependent on the purpose it is used for [74].

3.2 H₂ Pipeline Performance Model

The hydrogen pipeline model developed in this study was inspired by McCoy and Rubin [69]. This model was based on US natural gas pipeline costs and was used to estimate the cost of CO₂ transport per ton for a range of flowrates across different regions in the US. In this thesis, the model was further developed to evaluate the hydraulic performance of hydrogen at different flow rates allowing for the assessment of pipe size, pressures and the addition of booster stations; from that the capital and operating costs for pipeline as well as booster compressor stations were evaluated and thereof the total cost of transport. Location factors were included within to the cost models for both pipeline tools (CO₂ and H₂) to adjust the costs from US based prices and reflect a Canadian price range accounting for the tougher terrains and higher cost of labor in comparison to the US context using CEPCI, UCCI and GCCSI indices.

Similar to the McCoy and Rubin model, the H₂ model developed in this study uses design parameter inputs such as pipeline length, design flowrates, input and desired output pressure, segment elevation and booster stations. The hydrogen pipeline model only accounts for H₂ flow, other materials were not considered as hydrogen is assumed to be transported at high purity 99.99% knowing that including any impurities can have negative impacts on the pipeline integrity [75]. Figure 17 illustrates the inputs required to estimate the pipe diameter and other outputs of the hydrogen model.
3.2.1 Materials choice for hydrogen pipelines

Hydrogen is known to have an embrittlement effect on pipelines, as it is absorbed into steel when compressed in a pipeline reducing the toughness and ductility. It interacts with steel at the tip of a flaw point inside the pipe causing the pipe to crack at a stress less than the yield strength of the material [76].

For this reason, hydrogen pipeline standards are more stringent and have higher safety margins in comparison to natural gas and CO₂. In practice, this means using thicker walled pipelines which increases the material cost per unit length, and consequently the cost of labor required due to increased welding hours and operating expenses [66]. As a result, extensive research is being conducted on investigating the effects of hydrogen on steel to modify standards and optimize cost without compromising on safety or performance.

Recently an update was proposed in the by the American Society for Mechanical Engineers (ASME) to update the code for hydrogen pipeline established in 2008 in section B31.12 [77] to improve the thickness equations for hydrogen pipeline design for higher steel grades. The section originally introduced a “material performance factor” imposing a higher thickness at a
given diameter and pipe pressure relative to natural gas pipes on pipelines with yield strength greater 358 MPa. This would impose a higher cost for high steel grade pipelines which is contrary to the general practice where high quality steel is used in order to reduce thickness and material cost due to its high durability in the case of natural gas and CO₂ [66], [78]. These requirements were eased in the updated code as more data became available on the interactions between hydrogen and steel.

Acknowledging the complexity of the matter and the limitation of data enabling an in-depth analysis of the appropriate steel grade and desired pressure, the study assumes the use of X52 steel grade as recommended in literature [79], [80] as it is considered a safe to use options with no extra consideration required for hydrogen for pressures below the limit of 10 MPa [66].

### 3.2.2 CoolProps for Physical Properties Estimation

Physical properties required for pipeline calculation such as compressibility, density and viscosity were estimated using Coolprops, which is available as an add-in for Microsoft Excel [81]. Coolprops is an open access thermophysical properties library that delivers state of the art formulations and employs a wide range of methods to estimate transport properties. The add-in functions employ user-input parameters such as temperature, pressure and allow the selection of a single fluid or a mixture to calculate thermodynamic properties required to assess the hydraulic performance of these fluids in pipeline.

The original McCoy and Rubin model was modified by substituting the cubic equation of state employed in the original model for the Coolprops formulations. As validation, this gave very close estimates of the transport properties for CO₂, which results in the same pipe size and model performance.

### 3.2.3 Pipe Segment Design model

The model adopts the engineering equation developed by McCoy and Robin as it represents accurate estimations of fluid behavior in turbulent flow. The final formulation was derived from
an energy balance on the fluid where the pipe diameter is calculated for a pipeline segment holding upstream and downstream pressures as constants. A pipeline segment here means the length of a pipeline where the inlet and the minimum outlet pressures are specified [69]. The energy balance assumes negligible change in kinetic energy hence constant velocity, and a compressibility that is averaged over the length of the pipeline. The final equation for optimum diameter calculation as derived by McCoy is shown below.

\[
D_i = \left\{ \frac{-64 \cdot Z_{ave}^2 \cdot R^2 \cdot T_{ave}^2 \cdot f_F \cdot m^2 \cdot L}{\pi^2 [M \cdot Z_{ave} \cdot R \cdot T_{ave} (p_2^2 - p_1^2) + 2 g p_{ave}^2 M^2 (h_2 - h_1)]} \right\}^{1/5}
\]

Where,

- \( Z_{ave} = \) Average fluid compressibility over the segment
- \( R = \) Universal gas constant (Pa.m\(^3\)/mol.K)
- \( T_{ave} = \) Average fluid temperature (K)
- \( m = \) Design mass flow rate (kg/s)
- \( M = \) Molecular weight of the fluid (kg/kmol)
- \( p_1 = \) Segment upstream pressure (Pa)
- \( p_2 = \) Segment downstream pressure (Pa)
- \( h = \) pipeline elevation (m)
- \( P_{ave} = \) Average pressure (Pa)
- \( f_F = \) fanning friction factor

It is worth noting that the operating pressures for the Hydrogen model employed in this work were in the range of 10 to 5 MPa as highlighted by Fekete to be the feasible range for long distance H\(_2\) transport [66], in the case of CO\(_2\) the model was operated between pressures of 11 and 9 MPa which maintains the fluid in a supercritical state and generally represents a good range for underground injection.

The average pressure is calculated for the upstream and downstream pressures using Equation 9.
\[ P_{ave} = \frac{2}{3} \left( p_2 + p_1 - \frac{p_2 p_1}{p_2 + p_1} \right) \] [9]

The fanning friction factor is a function of the pipe diameter and therefore cannot be solved for analytically, hence the explicit approximation given by Equation 10 is used [82].

\[
\frac{1}{2\sqrt{f_F}} = -2.0 \log \left( \frac{\varepsilon / D_i}{3.7} - \frac{5.02}{Re} \log \left( \frac{\varepsilon / D_i}{3.7} - \frac{5.02}{Re} \log \left( \frac{\varepsilon / D_i}{3.7} + \frac{13}{Re} \right) \right) \right) \] [10]

\( \varepsilon \) is the pipe roughness (m) which is estimated to be 0.0457 for commercial steel pipes [83]. \( Re \) is Reynolds number which is defined by Equation 11 [84].

\[
Re = \frac{4m}{\mu \pi D_i} \] [11]

Where \( \mu \) is the viscosity of the fluid (Pa.s) calculated by Coolprops. Equations 9, 10 and 11 are solved iteratively to determine the optimum pipe diameter for the input design flow rate and pressure conditions. Firstly, the Reynolds number is calculated using an initial guess of the diameter based on a velocity estimate of 15 m/s [70] this Re number is then used to estimate the fanning friction number which is in turn used to calculate the updated diameter. These calculations are repeated until the diameter converges to within 10\(^{-6}\) m.
Since pipelines are generally sized at standard diameters. The final converged internal diameter is adjusted to a larger available ID and consequently a new pipe thickness. The line pipe sizes are usually referred to as Nominal Pipe Size (NPS). The model developed in this study has 17 NPS values between 4 and 48.

The inside diameter (ID) is determined by estimating the wall thickness using Equation 12 as specified by the ASME code for hydrogen pipeline [77]. Although the “material performance factor” mentioned in an earlier section was incorporated in the thickness equation for the Hydrogen model, it was given the value of 1 as the case study assumed the use of X52 steel grade.
\[ t = \frac{p_{\text{mop}} D_o}{2SEFH} \] [12]

where,

\( t = \text{wall thickness (m)} \)
\( P_{\text{mop}} = \text{maximum operating pressure} \)
\( D_o = \text{outside pipe diameter (m)} \)
\( S = \text{specified minimum yield stress for the pipe material (Pa)} \)
\( E = \text{longitudinal joint factor} \)
\( F = \text{design Factor} \)
\( H = \text{material performance factor} \)

The maximum allowed pressure is assumed to be 11 MPa in this study for the case of hydrogen, this is to avoid issues that can be raised by high stresses that aid embrittlement as previously discussed. The minimum yield stress depends on the steel grade. The pipelines in this study are assumed to meet the American Petroleum Institute (API) 5L specification. As previously stated, based on the steel grade used for this study (API 5L X-52 pipe) the minimum yield stress is 359 MPa. The optimum internal diameter \( D_i \) calculated in Equation 8 is adjusted to the next largest available internal diameter for an available NPS. As a result of this adjustment the downstream pressure will also be adjusted and will be greater than the value specified in the initial design. As the adjusted diameter is greater than the optimum value calculated by the diameter equation.

### 3.2.4 Hydrogen Booster Compressor Modelling

Booster compression stations are necessary for long pipelines distance to counter pressure drop across pipeline, it also used when a pipeline experiences increase in elevation when going through mountainous terrains. They can also reduce transport cost by allowing the use of a smaller pipeline as will be illustrated in a following section. In this work the same compression
model was used for both booster station along the pipeline and at the battery limit of the reforming units when pressure increase is required as will be discussed in Section 4.

The design and size of hydrogen compressors can vary widely depending on its application but the main obstacle with hydrogen compression in comparison to other gases is its low density and molecular weight which for typical mechanical compressors causes leakage issues that reduce the attainable efficiency [85]. A lot of attention has been given to improve the technology for hydrogen compression in stationary, distribution and automotive applications as it expected to be an expensive component of the supply chain [86].

The Hydrogen Delivery Scenario Analysis Model (HDSAM) a hydrogen distribution model developed by Argonne Laboratory in the US assumes a centrifugal compressor for centralized compression in their model, suggesting its reliability as opposed to a reciprocating compressor. They utilize an isentropic efficiency of 85% reflecting the U.S. Department of energy targets 2020 targets [87]. However a panel report produced by the National Renewable Energy Laboratory (NREL) suggests the use of 65% due to the lack of published real world data on hydrogen energy consumption [88]. In this study the compressor work was modelled using the general isentropic work equation for compressors and turbines [89] and assuming an isentropic efficiency of 65% as suggested by the panel report. A two-stage compressor was utilized assuming a compression ratio of 2 which falls within the range suggested by HDSAM [90].

\[
W = Z \frac{RT_1}{M_w} \frac{n}{\eta n - 1} \left[ \left( \frac{P_2}{P_1} \right)^{\frac{n - 1}{n}} - 1 \right] \tag{13}
\]

Where,

\( M_w = \) Molecular weight (kg/kmol)

\( Z = \) compressibility

\( R = \) universal gas constant, 8.314 J/K.mol
\[ T_1 = \text{inlet temperature} \]

\[ W = \text{work done, J/kg} \]

\[ \eta = \text{the product of isentropic and mechanical efficiency} \]

3.3 Economic Model

The section describes the pipeline and booster stations capital and operating cost models for hydrogen. For the case of CO\(_2\), unless otherwise stated, the original cost assumptions in the McCoy CO\(_2\) cost model were maintained while cost figures were updated using the UCCI to 2020 US dollars.

Pipeline CAPEX Model

Given the extensive use of natural gas pipelines and the availability of cost data from regulated pipeline operators [66], [91], the cost of pipelines for any high pressure fluid – including hydrogen – are commonly derived from natural gas pipelines. While it is understood that changing from natural gas to hydrogen service will impact the material and labor cost as previously discussed, a detailed assessment of these impacts on cost do not exist. Models for hydrogen pipelines commonly assume a cost multiplier to account for these costs. For example, the H2A delivery model developed by NREL increases the cost by 10% across-the-board for hydrogen pipelines relative to natural gas pipelines [92]. Although it is not explicitly stated how this multiplier was calculated, it can be deduced that it is to account for the increased cost of material and labor. In contrast, Leighty et al. [75] suggest that hydrogen pipelines can be build at the same cost as that of natural gas provided that the embrittlement challenges are met.

Pipeline construction costs are categorized into four categories, material, labor, right of way and miscellaneous [91]. Describing each component briefly, material cost covers the cost of the pipeline, coating and cathodic protection. Labor cost covers transportation, welding and installation. Right of Way (ROA) includes cost for acquisition, compensation, or repair of land on which the pipeline passes. Miscellaneous cost includes that for engineering, contingencies, supervision, overhead and other allowances. Labor cost can be affected by material cost as
thicker pipes require more welding, high quality installations and longer working hours while ROA and miscellaneous costs are generally independent [66]. These costs are also impacted by pipe length and other factors such as terrains, population and location. Parker [91] states the breakdown for these costs based on historical data for natural gas pipelines is 45%, 26%, 22% and 7% for labor, material, ROA and miscellaneous costs respectively. Knoope et al. [74] highlight that prices of steel have exhibited large increases over recent years and continue to do so and, as such, historical data may not be the best representation of new pipeline costs. Knoope et al. recommends the use of cost equations based on weight of pipeline steel as an improvement to the available models [93]. Following this recommendation, the weight of steel equation in their study was used to estimate the material cost for both the hydrogen and CO₂ pipeline cost models.

\[
C_{\text{material}} = t \cdot \pi \cdot (OD_{NPS} - t) \cdot L \cdot \rho_{\text{steel}} \cdot C_{\text{steel}} \quad [14]
\]

Where,

\( C_{\text{material}} \) = material cost for the pipeline ($)
\( t \) = thickness (m)
\( OD_{NPS} \) = the outer diameter of the nominal pipe size (m)
\( L \) = length of the pipeline (m)
\( \rho_{\text{steel}} \) = density of steel (kg/m³)
\( C_{\text{steel}} \) = cost of steel ($/kg)

The density of steel was taken as 7900 kg/m³ while cost of steel was taken as 0.65 $/kg an updated price from Knoope et al. [93].

The cost models adopted from the work of McCoy and Rubin is based on logarithmic relations for each cost category, in the original study these relations were developed for several regions within the US using natural gas pipeline datasets from 263 projects containing information about cost, length, diameter and location. In this thesis the mid-west US region is used to represent
Alberta, and as such the binary variable used for region selection in the original study were dropped.

All costs categories (excluding material cost) were updated from 2004 to 2020 using UCCI index. The capital cost equation takes the general form as shown below.

\[
\log(C) = a_0 + a_1 \log(L) + a_2 \log(D_{nps}) \quad [15]
\]

Where,

- \( C = \) Pipeline capital cost in 2004 US $
- \( L = \) Pipeline length (km)
- \( D = \) Pipeline NPS

*Parameter estimates for labor, miscellaneous and ROA costs components are provided in Table 7[69].*

Table 7 Parameter estimates for the pipeline cost model

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>Cost Component</th>
<th>Labour</th>
<th>ROW</th>
<th>Miscellaneous</th>
</tr>
</thead>
<tbody>
<tr>
<td>( a_0 )</td>
<td>Labour</td>
<td>4.487</td>
<td>3.950</td>
<td>4.390</td>
</tr>
<tr>
<td>( a_1 )</td>
<td>Labour</td>
<td>0.820</td>
<td>1.049</td>
<td>0.783</td>
</tr>
<tr>
<td>( a_1 )</td>
<td>Labour</td>
<td>0.940</td>
<td>0.403</td>
<td>0.791</td>
</tr>
</tbody>
</table>

*Location factor for Alberta Cost*

As previously stated, all cost categories from the McCoy & Rubin model with the exclusion of material cost (since it was based on steel price) were updated from 2004 to 2020 Q3 using the UCCI index to account for historical inflation. An overall location factor was developed in a similar fashion to the one in Section 2 where a weighted overall location multiplier was employed combining the individual multipliers for material and labor cost adopted from the GCCSI report [48], since the material and labor percentages in the overall transportation costs would change as a function of distance the location multiplier would be weighted accordingly.
Compressor CAPEX Model

The capital cost of the hydrogen and CO₂ compressors were scaled on the power consumption following the cost to capacity methodology described in chapter 2. The scaling factors, base cost, installation factors and reference sources are provided in the appendix. It should be noted that battery limit compressors costs were estimated following the same approach.

Operating and Maintenance (O&M) Cost Estimation

McCoy and Rubin highlighted that the O&M cost of pipelines can be significant. These can include activities such as site maintenance, facilities and environmental protection, erosion control and stabilization, pipeline integrity assessment, repair and modification and landscaping. Knoope et al. suggests the use of 1.5% of capital an assumption which was incorporated in both the H₂ and CO₂ cost models [74].

Operating and maintenance cost for the hydrogen and CO₂ compressor (excluding energy costs) are taken as 5 and 4% of the capital cost for the hydrogen and CO₂ compressors respectively [74], [85].

3.4 Carbon Intensity for Pipelines

To have a wholesome understanding of the competing transmission alternatives, the carbon footprint of the two options should be assessed. There are multiple sources of emissions that can be considered throughout the lifetime of a pipeline such as emissions during the construction, material transportation and/or acquisition, these will not be considered here as they are beyond the scope of this analysis. The major source accounted for in this thesis is booster stations power footprint which is related to the carbon intensity of the grid calculated in a similar fashion to Section 2. Since H₂ pipelines would generally require more compression, compared to CO₂ to achieve lower transmission costs as shown in the previous figures it would be anticipated to have a higher carbon intensity as a result of this power demand.
CO₂ pipeline leakage is not expected to be a major source of emission, unlike the understood view on natural gas pipes which have varying ranges of upstream emissions due to methane leakage. The only expected occurrences for leakage in CO₂ pipeline would be during maintenance, in this thesis a 1% leakage assumption is made to account for this aspect. Pipelines carbon intensities will be further addressed in Section 4.

3.5 Combining Cost and Performance Models

As discussed, the cost model depends on the pipeline diameter. The performance model, thus, calculates the required diameter for each segment based on the input parameters as indicated in Equation 8. The number of segments is set by the numbers of booster stations added to the model where the number of segments is greater by one than the number of booster stations, for example if the total pipeline length is 100 km and one booster compressor is added it splits the pipeline into two segments and diameter will be calculated for two 50 km length segments.

The capital cost is the summation of all the capital cost all segments and the booster stations. Key results that are delivered by the model are total capital and annual operating costs, total levelized cost per km and levelized cost per tonne of hydrogen transported. A capital recovery factor specified by the user is used to annualize the capital plant cost over a project period also specified by the user. The levelized cost of transport ($/ton H₂) is obtained by dividing the total levelized cost by the annual rate of transport which is the mass flow rate multiplied by a capacity factor.

3.6 Model Results

The H₂ and the modified CO₂ model key outputs are compared in this segment to build a framework around the transport costs of H₂ and CO₂ in Alberta and identify an optimizing behavior of booster stations.
3.6.1 Comparison of H$_2$ and CO$_2$ pipe sizes and Cost in Alberta

Table 8 provides the design and economic inputs used to produce the results presented in Figure 19 and Figure 20.

Similar to the previous section the pressures were set to satisfy general transmission requirements for respective fluid, the capital recovery factor and project life values represent average used in literature for pipeline cost analysis and the roughness is taken for non-corroded carbon steel.

Table 8: Design and economic input parameters for the comparative cost study

<table>
<thead>
<tr>
<th>Model Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ground temperature ($^\circ$C)</td>
<td>13</td>
</tr>
<tr>
<td>Elevation change (m)</td>
<td>0</td>
</tr>
<tr>
<td>Inlet Pressure H$_2$/CO$_2$ (MPa)</td>
<td>10 / 11</td>
</tr>
<tr>
<td>Minimum Outlet Pressure H$_2$/CO$_2$ (MPa)</td>
<td>5 / 9</td>
</tr>
<tr>
<td>Pipe Roughness (mm)</td>
<td>0.0457</td>
</tr>
<tr>
<td>Number of booster stations</td>
<td>0</td>
</tr>
<tr>
<td>Capital Recovery factor (%)</td>
<td>15</td>
</tr>
<tr>
<td>Project Life (years)</td>
<td>25</td>
</tr>
</tbody>
</table>

Figure 19 and Figure 20 illustrate the use of the models described to present a comparison between H$_2$ and CO$_2$ transmission costs. For fixed inlet and minimum outlet pressures the NPS changes as function of pipeline length and flowrate. The NPS size increases with increasing flowrate and pipeline length exhibiting step increases which reflects the discrete available NPS in the model. Generally, increasing the inlet pressure would lower the trend lines of different flow rates across the y-axis, as compressing the fluid allows transporting more for the same diameter (this is assuming a pipeline without booster compression). On the other hand, setting a minimum output pressure to meet specific delivery requirements (e.g., for CO$_2$ underground injection for instance) would influence the actual output pressure where increasing the minimum, and hence the actual, output pressure would reduce the range allowed for pressure drop across the pipeline length causing faster discrete jumps in the curves to happen, which indicates higher transport cost as pipe length increases. These trends are similar for both models, however they are amplified in the case of hydrogen where the fluid is being
transported in a gaseous phase occupying more volume per unit mass and therefore requiring larger diameters to move similar mass flow rates in this case a flow rate of 10 Mt/yr could not be achieved without the use of booster compressors as shown in Figure 20, where the largest NPS was reached when moving 5 Mt/yr for 400 km, generally H₂ delivery pressure would depend on the downstream application, if H₂ is intended for liquification trucks or fueling stations it will be compressed to pressures in the range of 25-50 MPa, while pipeline transportation would require 10 to 5 MPa with a practical minimum of 3 MPa as stated previously, distribution networks operate on much lower pressures (0.2-0.7 MPa). For the given pressure ranges moving the same amount of Hydrogen incurs double the transport cost compared to moving CO₂ as illustrated by the area trends. The last observation is that moving more fluid would ultimately reduce the cost of transport as indicated by the area trends, taking the example of CO₂ being moved for 400 km doubling the flow rate from 1 to 2 Mt/yr reduced the transport cost from 25 to 16 $/tCO₂.

As these results provide a baseline for comparing H₂ and CO₂ transport costs, however when making these comparisons one should bear in mind two aspects. First is that hydrogen production from fossil fuels, whether using coal, biomass, natural gas or other hydrocarbon through gasification or reforming technologies, will results in a need to move more CO₂ for sequestration than H₂ to market on a mass basis. For example, the amount of CO₂ that needs to be handled as a result of producing blue hydrogen from reforming units will be in the range of 9:1 ratio.
Figure 22: CO₂ pipeline diameter and cost of transport as function of length for several flow rates in MtCO₂/yr (trend lines read against left axis and areas read against right axis)

Figure 23: H₂ pipeline diameter and cost of transport as function of length for several flow rates in Mt/yr (trend lines read against left axis and areas read against right axis)
The second aspect is how costs are reported and allocated depending on the value of the fluid being transported. In the results shown in Figure 19 and Figure 20 the cost of transport was normalized to the fluid being transported. However, this may not always be the case. For example, where CO₂ is treated as a waste product resulting from the hydrogen production, the costs of CO₂ transport must be borne by the produced hydrogen. In contrast, if there is an option to utilize the CO₂ – e.g., in enhanced oil recovery or other utilization projects – transport costs for CO₂ could be borne by the CO₂, as it could be seen as a feedstock delivered to a downstream user. Both of these mentioned aspects will be further addressed in Section 4.

The analysis conducted can be expanded further by investigating cost trade-offs when varying other parameters such as the input and out pressures and assessing the impact of using different steel grades as discussed in section 3.2.1. Such an analysis becomes scenario specific and requires more information regarding the allowable steel grades that can be utilized for each fluid and the regulations governing such decisions which is beyond the scope of this work.

3.6.2 Impact of Booster Compressors on Levelized Cost of Transport

As previously mentioned, adding booster station can reduce the cost of transport especially when moving more hydrogen for longer distances. This is illustrated by the case shown below for a 400 km pipeline moving 266 tH₂/day where adding more booster stations (taking a power cost of 0.07 CAD/kWh) reduces the pipe size required to move the same amount of Hydrogen at a marginal increase in operating cost due to power consumption (difference between zero and one booster station) further addition can then increase the outlet pressure for the same pipe size, however as more booster compressors are added the cost of transport becomes more sensitive to power prices where a 50% variation in power cost translated to 1% for a pipeline with one booster compressor and 6 % for a pipeline with 4 compressors.

Another possible strategy for utilising booster stations is to leverage capital and operating expenses (difference between 2 and 3 booster) where two comparable delivery costs are available one with larger pipe size and lower operating cost and another with a smaller capital but a marginal operating expense that counters the saving made by its addition. Though both
options may look like they cost the same, the return on investment for both options may be different depending on the nature of project funding, where a high project having high debt percentage can opt for the option with lower capital, bearing the risk of power price fluctuations.

![Figure 24: Impact of booster compressors on total levelized cost of H₂ transport moving 266 tH₂/day for 400 km (whiskers show effect of varying power price by ± 50%, y-axis starts from 0.2 US$/kgH₂)](image)

A similar effect is shown in the case of CO₂ transmission as shown in Figure 22 depicting the impact of booster compressors on a pipeline moving 2,384 tCO₂/day. However, as the fluid is being transported in a supercritical state it exhibits a mixture of liquid and gaseous behaviours hence having a smaller pressure drop and therefore the addition of booster stations do not have as much savings (pipeline volume is already fully utilized/occupied with high density fluid).
Figure 25: Impact of booster compressors on total levelized cost of CO₂ transport moving 2384 tCO₂/day for 400 km (y-axis starts from 0.04 US$/kgCO₂)

The smaller scale for the y-axis in comparison to the previous figure is due to the larger amounts moved for the same.

It is worth mentioning that electric motors are assumed to run both CO₂ and H₂ compressors and that their power supply comes from the grid. However, H₂ compressors can possibly be powered using turbines fueled by a fraction of the transported H₂. This can reduce the cost of transportation and the savings may be significant when moving a large scale provided the technology exists.

3.6.3 Sensitivity Analysis

A sensitivity analysis was conducted to assess the effect of varying some of the input parameters on the levelized cost of transport. Realizing that both pipeline models will behave similarly, and hydrogen was taken as an example to show which parameters is the transport cost most sensitive to.

Figure 23 shows that the cost of transport will be highly affected by the interest rate almost linearly, where a 50% increase will increase the cost by 38%. Transport would also be affected by
steel price where increasing the steel price by 50% increase the cost by 4.5% which confirms the comments made by Knoope et al. This impact would be even more should the pipeline be longer as the material cost would cover a larger percentage of the pipe.

Figure 26: Sensitivity analysis on levelized cost of transport for Hydrogen (0% correspond to base case of 15% interest rate, 0.76 $/kg for steel price and 1.5% of capital cost as fraction for pipe O&M )
4 Scenario Analysis: A case study for Canadian Oil Sands

Scenario analyses that can highlight the system and component level trade-offs across the supply chain between different production and delivery options for hydrogen, enable comparison with incumbent energy systems, and support decision making for development of node strategies. Scenario analyses and modelling tools have been developed by governmental organization such as NREL and Argonne labs [94] and in the academic domain [95], [96]. These focused primarily on the H₂ pipeline networks for different demand scenarios. NREL provides different models for varying feedstock and hydrogen production technologies, including natural gas, coal, nuclear and renewable resources [97]. However, their natural gas model employs only one technology variant (SMR with pre-combustion capture), limiting the climate and economic benefits for countries like Canada seeking to leverage their low-cost natural gas resource. This chapter carries out a scenario analysis similar in nature to the one conducted by Olateju and Kumar [71] where they investigated the techno-economics of hydrogen production from underground gasification of coal for upgrading bitumen in the oil sands. Their results showed that SMR with CCS is still a lower-cost alternative for hydrogen production. This complements the importance of investigating the trade-offs between the variants of the reforming and CO₂ configuration for hydrogen production and subsequent transmission alternatives.

This chapter seeks to develop a framework to investigate the economic and technical trade-offs between the different supply chain components by conducting scenario analyses contextualizing the various reforming routes and transmission options, using the models presented in the previous chapters to estimate and compare costs of hydrogen production, CCS and CO₂ and H₂ transport options for the purpose of delivering low-carbon intensity hydrogen.

4.1 Scenarios Tree Description

Two hydrogen demand scenarios and two delivery options were examined to identify and understand the cost and environmental trade-offs and drivers of variation in the system components. In the first scenario, referred to as Demand Scenario 1 (DS1), is based on demand
of 266 tons H₂/day to replace natural gas in once-through steam generators (OTSG) as a fuel in a steam assisted gravity drainage (SAGD) facility in Fort MacMurray. The demand for hydrogen in DS1 would be sufficient to produce 5700 m³ of crude bitumen per day, which is approximately the amount produced by the MacKay River project owned by Suncor Energy Inc. The second scenario, referred to as Demand Scenario 2 (DS2), is based on a demand of 110 tons H₂/day used to replace the diesel consumption for large mine trucks in an oilsands mining projects in Fort MacMurray, Alberta. The hydrogen demands estimates suggested in these scenarios are based on preliminary energy basis calculations using information from the Alberta Energy Regulator (AER) and Canadian Oil Sands Innovation Alliance (COSIA) reports [98]. The two scenarios for hydrogen utilisation in the oil sands investigated in this case study were reviewed by representatives of several oil sands producers and, in their view, represent realistic decarbonization options.

We then consider two options to supply this hydrogen to the oilsands. In Location Option 1 (LO1), new blue hydrogen production is constructed in the Industrial Heartland, Alberta, near existing CO₂ storage infrastructure (e.g., Shell Quest) and prospective geologic storage resources. Captured CO₂ is compressed to 11 MPa and geologically sequestered, while Hydrogen is compressed to 10 MPa and transported 400 km north by pipeline to meet to oil sands demand. In Location Option 2 (LO2), new blue hydrogen production is constructed in Fort MacMurray, close to demand where battery limit pressures satisfy the requirement for the end user and no H₂ compression is needed, and captured CO₂ is compressed to 11 MPa and transported 400 km south by pipeline to geological storage close to the Heartland region.

For ease of reference the production routes SMR with pre-combustion capture, SMR with post combustion capture and ATR with capture will be referred to as SMR-Pre, SMR-Post and ATR-Pre respectively in this section.
Figure 27: Description of scenario analysis hierarchy
4.2 Levelized Cost of Production

Tables 8 and 9 compares the performance of the three production technologies and their economic value for the scenarios under study, the same economic inputs from section 2.3.1 were used. As illustrated in Section 2, SMR-Post and ATR have lower direct carbon intensities (1-1.3
kgCO₂eq/kgH₂) however these are more expensive investments compared to SMR-Pre due to the larger capture units and ASU installation requirement. The results shown below are not considering GHG emissions pricing, as this aspect will be assessed in a following figure.

Table 8: Overview of performance metrics for blue hydrogen production technologies

<table>
<thead>
<tr>
<th>Metric/Parameter</th>
<th>Unit</th>
<th>SMR w/o CC</th>
<th>Pre</th>
<th>Post</th>
<th>ATR w CC</th>
</tr>
</thead>
<tbody>
<tr>
<td>Specific Direct Emission</td>
<td>kgCO₂eq/kgH₂</td>
<td>9.0</td>
<td>4.1</td>
<td>1.0</td>
<td>1.27</td>
</tr>
<tr>
<td>Mass Ratio CO₂captured/H₂</td>
<td>KgCO₂/KgH₂</td>
<td>-</td>
<td>5.1</td>
<td>9.0</td>
<td>8.2</td>
</tr>
<tr>
<td>Overall Plant Capture Efficiency</td>
<td>%</td>
<td>-</td>
<td>54</td>
<td>89</td>
<td>86.6</td>
</tr>
<tr>
<td>Specific NG consumption</td>
<td>kg NG/kgH₂</td>
<td>3.39</td>
<td>3.48</td>
<td>3.73</td>
<td>3.6</td>
</tr>
</tbody>
</table>

Table 9: Levelized cost of production for demand scenarios and location options (least expensive low emission alternative highlighted in green)

<table>
<thead>
<tr>
<th>Production unit</th>
<th>Levelized Cost of Production LCOH US$/kgH₂</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>LO1 (Heartland Production)</td>
</tr>
<tr>
<td>DS1 (266 tH₂/day)</td>
<td></td>
</tr>
<tr>
<td>SMR w/o CC</td>
<td>1.07</td>
</tr>
<tr>
<td>SMR-Pre</td>
<td><strong>1.41</strong></td>
</tr>
<tr>
<td>SMR-Post</td>
<td>1.70</td>
</tr>
<tr>
<td>ATR-Pre</td>
<td>1.70</td>
</tr>
<tr>
<td>DS2 (110 tH₂/day)</td>
<td></td>
</tr>
<tr>
<td>SMR w/o CC</td>
<td>1.25</td>
</tr>
<tr>
<td>SMR-Pre</td>
<td><strong>1.67</strong></td>
</tr>
<tr>
<td>SMR-Post</td>
<td>2.05</td>
</tr>
<tr>
<td>ATR-Pre</td>
<td>2.13</td>
</tr>
</tbody>
</table>
A breakdown of the levelized cost of production (which excludes the transportation costs) in the Heartland region (LO1) is shown in Figure 26 for the two demand scenarios. Compared to building the production units in the Heartland region (LO1) building the plant in Fort MacMurray (LO2) would increase capital investment by 13% due to the geographical constraints and the higher labor and equipment transport expenses this is due to the higher location factor used in the case of Fort MacMurray 1.76 as opposed to 1.48 for the Heartland region. Although the plants in Fort MacMurray (LO2) wouldn’t require large hydrogen compressors at the battery limit, the savings in compression (both capital and operational) do not offset the location factor impact and consequentially the LCOH is higher by 6 to 9% for all technologies and demand scenarios for LO2. The impact of scale is noticed where the cost of production is lower for DS1 with respect to individual routes. Since hydrogen will be compressed at the battery limit specifically in LO1 (Heartland production-with hydrogen pipeline) only the direct emissions are shown in this figure which shows comparable results for ATR and SMR-Post as the emissions from the ATR’s power island are not captured.

Figure 29: Levelized cost of H₂ and direct carbon emission intensity for scenarios
4.3 Levelized Cost of Transport (H₂ versus CO₂)

As discussed in Section 3, larger amounts of CO₂ would be transported when LO₂ is chosen (4-9 kg CO₂ transported/kg H₂ produced). As expected, the CO₂ pipeline had similar diameters to that of hydrogen, since CO₂ is being transported in a supercritical phase compared to hydrogen’s gaseous phase it resulted in the same pipe size for both fluids even at the 9 to 1 mass ratio as indicated in Figure 27.

The number of booster stations was optimized to achieve the minimum cost for each transport option, resulting in CO₂ transport costs that are slightly less expensive for the given pipeline distance of 400 km. In the analysis shown H₂ is considered the product and thus in the case of transporting CO₂ the costs is borne by hydrogen, in other words when considering CO₂ pipeline the cost of transport represents the cost of transporting the captured CO₂ per kg of H₂ produced which differs according to the technology. Another aspect that can be of influence is the interest rate of the project which can impact whether a pipeline with higher CAPEX and lower OPEX or vice versa would be more economic. When comparing other production capacities (or different technologies) and distances this may tip the scale between H₂ and CO₂ should compression become significant.
Figure 30: Levelized transport cost for H₂ and CO₂ pipelines under DS1 and DS2, for LO1 only H₂ is transported and for LO2, only CO₂ is transported regardless of the capture technology.

The carbon intensity of each transmission option was calculated as discussed in Section 3 by charging the emissions intensity of the power imported from the grid at 680 kgCO₂eq /MWh and accounting for CO₂ pipeline leakage. These intensities varied between options and scenarios where the higher intensities were borne by the DS1 due to the higher amounts of fluid being handled. Within the scenarios H₂ pipelines incurred a larger carbon footprint as it requires more booster compressors and hence consumed more power to deliver H₂ at a lower price. Table 10 summarizes the number of booster compressors, total power consumption and the resulting specific intensity for each transmission option in each scenario.
Table 10: Specific carbon intensity for pipeline due to booster compressors power consumption and CO₂ pipeline leakage

<table>
<thead>
<tr>
<th>Transmission option</th>
<th>CO₂ pipe SMR-Pre</th>
<th>CO₂ pipe SMR-Post</th>
<th>CO₂ pipe ATR-Pre</th>
<th>H₂ pipe</th>
<th>CO₂ pipe SMR-Pre</th>
<th>CO₂ pipe SMR-Pre</th>
<th>CO₂ pipe ATR-Pre</th>
</tr>
</thead>
<tbody>
<tr>
<td>DS 1 (266 tH₂/day)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>No. of boosters</td>
<td>4</td>
<td>1</td>
<td>5</td>
<td>4</td>
<td>3</td>
<td>1</td>
<td>3</td>
</tr>
<tr>
<td>Total power consumption MWₑ</td>
<td>9.4</td>
<td>.04</td>
<td>.38</td>
<td>.28</td>
<td>2.32</td>
<td>0.01</td>
<td>.09</td>
</tr>
<tr>
<td>GHG emissions (electricity) kgCO₂eq/kgH₂</td>
<td>.579</td>
<td>.006</td>
<td>.023</td>
<td>.017</td>
<td>0.334</td>
<td>.002</td>
<td>.014</td>
</tr>
<tr>
<td>GHG emissions (fugitives and venting) kgCO₂eq/kgH₂</td>
<td>-</td>
<td>0.051</td>
<td>.09</td>
<td>.082</td>
<td>-</td>
<td>0.051</td>
<td>0.09</td>
</tr>
<tr>
<td>Total Pipeline CI kgCO₂eq/kgH₂</td>
<td>.579</td>
<td>.057</td>
<td>0.113</td>
<td>0.099</td>
<td>.334</td>
<td>0.053</td>
<td>0.104</td>
</tr>
</tbody>
</table>

4.4 Scenario and Options Analysis

Production and transport costs and emission intensities were combined to identify the effect of technology alternatives and transmission options on the cost of H₂ delivery and the estimated carbon footprint for the suggested scenarios. Figure 28 shows that the lowest-cost pathway (without any emissions pricing) under both Demand Scenarios is building an SMR-Pre facility in the Heartland region and transporting hydrogen to the oilsands. The most expensive option for DS2 is to supply demand with an ATR built in the Fort MacMurray region while piping CO₂ south for storage (LO₂). For DS1 an SMR-Post built in Fort MacMurray is slightly more expensive than ATR.

The GHG impact from power import and export to the grid discussed in Section 2 becomes more significant in this context. Where the carbon intensity varies across technology and
options of transport. Although the direct emission intensity doesn’t change, when considering indirect emissions (from plant and pipeline power consumption and upstream emissions) the result is higher overall carbon intensities for plants built in the Heartland region (LO1). This is driven by the extra power required to pump the hydrogen at the battery limit to 10 MPa and throughout the pipeline in order to deliver it to Fort MacMurray and therefore increases the carbon intensity credits, an aspect which would not be necessary should the plants be build closer to demand in Fort MacMurray. Recalling the discussion in Section 2, this has particularly the worst impact on ATRs built in the Heartland region (LO1) which have significantly higher life cycle intensities (2.45-2.5 kgCO2eq/kgH2) compared to its original direct emission intensity (1.27 kgCO2eq/kgH2). A carbon tax will now play a critical factor on both the choice of technology and location of production.

![Figure 31: Total levelized cost and life cycle carbon intensity for blue hydrogen production and delivery for different routes and scenarios with zero carbon tax applied on emissions, primary y-axis starts from 1 $/kgH2](image)

Figures 31 and 32 show the effect of adding a carbon tax across the lifetime and on the life cycle emissions of the project, where every ton of CO2 emitted as a result of hydrogen production is taxed which will add to the expenses of the plant increasing the cost of production.
production. The range for carbon taxes is to demonstrate the effect of incrementing the tax. The current policy of the government of Canada will see the carbon tax reach 170 CAD/tCO\(_2\) in the year 2030.

The carbon tax penalizes the most carbon intensive pathways making them more costly, hence the steepest lines for both scenarios are SMR-Pre plants built in the Heartland region LO1, making it the worst alternative for DS1 and the third worst for DS2 which is a big shift from the results in Figure 28. These results also indicate that ATR plants are not an appealing alternative even under a 170 CAD carbon tax, this is owning to the high capital investments associated with these plants and the limited amount of CO\(_2\) avoided despite its high capture rates.

At the 170 CAD tax SMR-Pre in Fort MacMurray appears to be the lowest cost alternative, however, if the carbon tax increases, the appealing alternative (flattest slope) becomes SMR-Post built in Fort MacMurray (LO2) indicating that at the range of the studied scales (110 - 266 tonH\(_2\)/day) the carbon tax advantages the plant having the least direct intensities and minimal power requirement and for similar reasons the CO\(_2\) pipeline becomes a favorable alternative in these scenarios. At a 170 CAD tax the increased expenses for the Heartland location were large enough to offset the capital expenses that would be incurred for building the plant in Fort McMurray, making both options of equal value in DS2 and making Fort McMurray the appealing location for constructing large scale plants (DS1). Although not included in the main analysis the base SMR options were also considered where in LO1 the plant was constructed in Heartland with H\(_2\) pipeline to Fort MacMurray however in LO2 there was no CO\(_2\) pipeline included as there was non-captured. Interestingly for both DS1 and DS2 a baseline SMR (without capture) built in Fort MacMurray was still the least expensive option even with a 170 CAD $/tCO\(_2\) carbon tax indicating that the pipeline infrastructure savings were enough to offset the carbon tax expenses. A baseline SMR in LO1 became the most expensive option at a 115 CAD $/tCO\(_2\) tax for DS1 and 170 CAD $/tCO\(_2\) for DS2. Figures showing these results are provided in Appendix K.
Figure 32: Impact of carbon tax on the total levelized cost of hydrogen delivery (production +pipeline) for the three reforming technologies and the two locations options for demand scenario 1 (DS1), y-axis starts from 1.8 $/kgH₂

Figure 33: Impact of carbon tax on the total levelized cost of hydrogen delivery (production +pipeline) for the three reforming technologies and the two locations options for demand scenario 2 (DS2), y-axis starts from 2.2 $/kgH₂
The results of this chapter highlight that the extent of H₂ compression and the ability to be self-sufficient with power demand will be determining factors in the cost of produced hydrogen and thereof the most cost-effective route to follow. The impact of scale on increased carbon intensity driven by pipeline footprint would be significant enough to influence decision making should the full life cycle emissions taxes be borne by the production plants. Observing how independent generation influenced the cost of production an opportunity that unfolds now is the incorporation of a power generating SMR plants with carbon capture which allows for further leveraging of the carbon tax and getting credit by selling clean electricity displacing the grid intensity.

The analysis conducted in this section would need a comparison of the pathways discussed against costs and life cycle emissions of the incumbent systems to gain a holistic view of the best alternatives for energy delivery. This will be the target of future work and is beyond the scope of this thesis.
5. Conclusions and Recommendations

This thesis aimed to understand the trade-offs of different components within the blue hydrogen supply chain in an Alberta context by developing models to estimate costs at different capacities. The results show that the production of blue hydrogen can be achieved at relatively low costs when building large scale plants. Taking a baseline production of 350 tH₂/day the estimate costs were 1.6 US$/kgH₂ for SMR with post combustion capture and ATR with capture, and 1.2 US$/kgH₂ for SMR with pre-combustion capture, with direct intensities of 1.0, 1.3 and 4.1 kgCO₂eq/kgH₂, respectively. The low cost is primarily driven by the low natural gas price in the province. As the production scales increases ATR with pre-combustion capture becomes more favorable in comparison to SMR with post-combustion capture as the ASU which a capital-intensive expense benefits from economies of scale and the unit incremental expense on ATR falls, for the SMR, the large scale for capture increases the cost of the post-combustion capture unit faster than it does for the pre-combustion option in the ATR.

The effects of electricity generation on hydrogen production from merchant plants has not been fully considered in economic or environmental analyses, and particularly where CO₂ capture is integrated. Traditional SMR-based plants generate excess steam (or electricity) that can be exported. While ATR plants, too, may have electricity that can be exported, it is less than from an SMR due, in part, to the demands of the ASU. Thus, the impact of electric grid carbon-intensity impacts SMR and ATR plants differently as capture is added and, generally, reduces avoided CO₂ emissions over the life cycle. This is manifest in the cost of avoided CO₂, which were 66, 74 and 75 US$/tCO₂ avoided for SMR with pre-combustion capture, SMR with post-combustion capture, and ATR with capture, respectively, when accounting for direct emissions only, and 75, 80 and 89 US$/tCO₂ when electricity is included. Adding the upstream emissions didn’t have a significant effect on the cost of avoided CO₂. However ultimately low-emission suppliers of natural gas are necessary to reduce the life cycle emissions and minimize the impact of a carbon pricing.
Hydrogen compression and the ability to self generate power will determine the least carbon intensive production technology and the impact of these two factors would in turn determine the low-cost route if controlled by the carbon tax applied on the life cycle emissions. From the plant perspective, this means that ATR plants although have a capital cost advantage at higher scales, they may not be superior to SMR due to their limited power generation abilities when integrated with a carbon capture unit.

High levels of capture do, indeed, result in production of hydrogen with a low carbon intensity. When considering the cradle to gate emissions, the reduction is up to 84 % for SMR with post-combustion capture and 75% for ATR relative to an SMR without capture.

Moving \( \text{H}_2 \) is surprisingly up to four times more expensive than moving the resulting captured \( \text{CO}_2 \) when costs are ascribed per unit mass of fluid transported. This large difference is due to the transport of hydrogen in the gaseous phase, which requires larger pipe sizes and hence more expensive pipelines than \( \text{CO}_2 \), which is transported in a higher density supercritical state. However, when adding booster stations and associating cost of transport to the produced \( \text{H}_2 \) the cost of moving \( \text{H}_2 \) or the resulting \( \text{CO}_2 \) captured from \( \text{H}_2 \) production, is comparable. For example, the costs of moving 110 and 266 t\( \text{H}_2 \)/day or the resulting captured \( \text{CO}_2 \) for a distance of 400 km are approximately 0.4 US$/kg\( \text{H}_2 \) and 0.7 US$/kg\( \text{H}_2 \), respectively. This indicates that transmission cost alone is not an impacting factor in the supply chain analysis. However, from an environmental perspective the operational emissions associated with the compression of \( \text{H}_2 \) for transport can be significant in relation to the emissions for \( \text{CO}_2 \) transmission. Under a carbon tax, this tends to make a \( \text{CO}_2 \) pipelines a less expensive transmission option. Under the assumption that both pipeline systems are using grid electricity.

In the Alberta context, building hydrogen production in the Fort Saskatchewan area (Heartland region) and transporting \( \text{H}_2 \) north is a consistently lower cost option than building units in Fort MacMurray and moving \( \text{CO}_2 \) south across the scales investigated in the absence of a carbon tax. However, the prior option has significantly higher emissions due to the grid electricity needed to run compression that increases the pressure of hydrogen to a level sufficient for transmissions by pipeline. As a result, an increasing carbon tax applied on the full life cycle
emissions of the projects diminished the advantage of production in the Fort Saskatchewan (Heartland) area.

The findings from this thesis indicate that, all other things being equal, developing larger scale hydrogen plants with CCS reduces the cost of production and transportation and favors investments in central hub production and shared pipeline infrastructure. As the energy sector looks to a net-zero future with increasing carbon prices that will inflate the cost of current fossil commodities, blue H₂ represents a promising option to reduce emissions and blunt the impact of carbon pricing on production costs. This could also reduce the chance that natural gas resources become stranded in a net-neutral scenario and provide a transition pathway for the Province of Alberta and its oil and gas sector.

Recommendations

Some recommendations for future work to complement the analysis conducted in this study include:

- Modelling of ATR plants to better understand the emissions from pre-heating and carbon capture options therein.
- Analysis of pipeline transmission costs using different steel grades and incorporating spatial and temporal components in the pipeline models to gain more terrain-specific cost estimates and allow for uncertainty analyses in future transmission capacities.
- Conducting a full life cycle assessment for the proposed scenarios against the incumbent technologies and obtaining more accurate scales for the Demand Scenarios (DS) by conducting a technical assessment of actual requirements of the demanding units (Heavy hauler vehicles and SAGD facilities) for hydrogen.
- Conducting an analysis that assesses the techno-economic and environmental potential of synergistic integrations of SMRs with power generation cycles and carbon capture integration.
References


[104] “Alberta Historical Electricity Rates by Encor by EPCOR | EPCOR.”

Appendix A: Description of the NETL Capital Cost Aggregation Methodology

The NETL methodology of cost estimation is composed of four layers of capital cost. Cost components adding up to the Total Overnight Cost TOC are in constant currency as they don’t account for inflation effect which is needed for more realistic cost trend. Adding the aforementioned costs and accounting for escalation over the capital expenditure period gives the Total As Spent Cost (TASC). In this study however only the interest during construction was considered.

The cost terms mentioned are described briefly below[38], [49]:

**Bare Erect Cost (BEC)** consists of all the equipment costs of the plant, material and installation expenses, supporting infrastructure such as labs and offices, direct costs and indirect labor costs required for installation.

**The Engineering, Procurement and Construction Cost (EPCC)** which includes the BEC costs and the costs incurred by EPC contractor services. In this study these service costs were assumed to be 10% of the BEC.

**Contingency costs** are miscellaneous cost incurred as the project moves towards completion. These are divided into process and project contingencies. The former accounts for uncertainty related to the maturity of the technology, where technology at early stages of development have higher percentage of process contingency and lower percentages are given for commercial processes.

In this study the process contingency for the SMR and ATR plants were 10% of the process capital (BEC) while the CO2 capture systems were given a 35% as there has been limited plants which have integrated capture units in hydrogen reforming facilities.

Project contingency is an additional cost allowance that would be identified when the project implementation reaches a more detailed stage. As such this contingency is related to the class of the cost estimate as explained by Rubin et al. [49]. Since this study is categorized under the class 4 and 5 estimates a project contingency of 40% of the sum of BEC, EPCC and process contingency was used.
The **Total Plant Cost (TPC)** is the sum up value of all the previous cost components. The **Total overnight cost TOC** is the sum of the following items:

- TPC are reported above
- Pre-production costs taken as 2% of the TPC, plus 25% of the monthly fuel and feedstock consumption 3 month of operation and maintenance labour cost and 1 month of catalyst and chemicals cost.
- Working (inventory) capital which is 1 month of inventory and chemicals
- Owner’s cost which covers any additional which services needed such as feasibility studies, permitting costs, legal fees, taxes...etc. is assumed to be 15% of TPC.

The interest during construction was calculated assuming a 3-year plant construction schedule with a distribution of the TOC of 20-45-35% the interest rate equal to the discount factor. Adding up the interest during construction and the TOC the TASC (Total As-Spent Cost) is evaluated.

Other than capital costs there is also operational and maintenance costs. These are divided in variable and fixed cost. Variable costs cover feedstock and fuel consumption, power, water, chemical and catalyst. The consumption rates of these items were calculated by the performance model expect for chemicals and catalyst which were taken % of the TPC. Appendix B summarizes the economic inputs and assumption for the fixed and variable operating costs.
Appendix B: Description of the Cost Aggregation Method, Process Plant Layout and Operating Parameters of Different Studies

Nomenclature:
BEC: Bare Erected Cost, EPC: Engineering Procurement and Construction, HTS: High-Temperature Shift reactors, IDC: Interest During Construction, TIC: Total Installed Cost, TPI: Total Plant Investment, TPC: Total Capital Cost, TCR: Total Capital Requirement, TASC: Total As Spent Cost

Reference (A): NETL 2013 [99]
• Description: The report provided cost information on an SMR unit which represents a H₂ production unit in a Gas to Liquid (GTL) facility.
• Source of cost information: costs were scaled from previous NETL reports, these in turn obtained their information from engineering firms.
• Cost Aggregation:
  BEC covers equipment, supporting facilities, infrastructure and labor installation cost. Overnight cost covers inventory, working capital, production costs and owner’s cost
  TPC = BEC + EPC + Process contingency + Project contingency
  Total Overnight Capital = TPC + overnight cost
  TASC = TOC + IDC + escalation
• Reported Cost Year: 2011 (updated using CEPCI)
• Battery limit conditions: T: 38 ºC, P: 4.14 MPa
• Plant configuration:
The hydrogen production section included SMR, PSA, and other miscellaneous equipment. A pre-reformer was not included in that section, a shift reactor and purification (desulfurizer) unit costs were not mentioned explicitly but are assumed to be covered by the miscellaneous costs.
• Efficiency (conversion: SMR (CH4) = 78.5 % HHV)

Reference (B): NREL 2006 [41]
• Description: The report developed cost estimates for selected commercial technologies (FCC, SMR and Natural Gas Liquid (NGL) Expanders) at large and small scales.
• Source of cost information: Publications, licensors of SMR units and process modeling.
• **Cost Aggregation:**

An installation factor of 2.47 (based on a cited NREL Report “Biomass Syngas to Hydrogen Production Design Report”) was used to calculate the installed capital cost. EPC, construction, contingencies and fees (categorized as indirect cost) were accounted for using factors developed by NREL.

\[ TIC = \text{equipment cost} \times 2.47 \]

\[ TPI = TIC + \text{indirect cost factors} \]

• **Reported Cost Year:** 2005

• **Battery limit conditions:** T=54.4 °C, P=2.9 MPa

• **Plant configuration:**

Both plant scales include purification (desulfurizer) units, reformers, high-temperature shift reactors and PSAs. The major difference between the large- and small-scale facilities is the integration of a direct-fired heater for the large-scale plant. H₂ compression and storage costs were not included.

• **Efficiency:** Small: conversion: SMR= 71%, HTS = 95 %, PSA =75 % (recovery)

  Large: conversion: SMR= 77%, HTS = 92.3 %, PSA =86 % (recovery)

**Reference (C): Argonne 2003 [13]**

• **Description:** Report modeled H₂ production through SMR with CO₂ sequestration to produce economic and design benchmarks for comparison with other technologies (IGCC) for merchant H₂ production.

• **Source of cost information:** Published data and ASPEN simulation model

• **Cost Aggregation:**

\[ TPI = \text{Direct Capital cost} + \text{Installation costs} + \text{Inventories (catalyst, solvent)} \]

• **Reported Cost Year:** 2003, no mention of basis year hence the publication year was adopted as the basis

• **Battery limit conditions:** T=37.78 °C, P=1.38 MPa

• **Plant configuration:**

The simulation model included a purification unit, Pre-reformer, reformer, high and low-temperature shift reactors and PSA.
• **Efficiency:** (conversion: SMR= 75%, HTS = 95 %, LTS=90%, PSA =90 % recovery)

**Reference (D): Basye 1997** [55]

• **Description:** Report surveyed costs of different hydrogen production technologies

• **Source of cost information:** data from Steinberg & Cheng 1989[100]

• **Cost Aggregation:**
  
  Interest during construction (IDC) and start-up expenses were taken as 10% and 2% of the equipment and facility costs respectively, working capital represents 2 months of expenses.

  Total Capital Requirement (TCR)= equipment & facility costs + IDC + Start-up + working capital

• **Reported Cost Year:** 1997

• **Battery limit conditions:** not stated

• **Plant configuration:**
  
  Plant includes purification unit, reformer, shift reactor, PSA and other related equipment

• **Efficiency:** (conversion: SMR (CH4) = 78.5 % HHV)

**Reference (E): NREL 2002** [54]

• **Description:** Report evaluates hydrogen production technologies at three commercial scales:
  
  central stations (large scale) (2,500-500 tons H₂/day,) mid-size (100-10 tons H₂/day) and
distribution size (small scale) (10,000-100 kg H₂/day)

• **Source of cost information:** SFA Pacific database and H2 equipment vendors

• **Cost Aggregation:**
  
  Capital build-up is based on percentages of battery limit process unit cost. General facilities
  (20% of process units), Engineering Permitting and Start-up (10-15%) depending on scale,
  contingencies (10%) Working capital, land and miscellaneous (5-7%) depending on the scale.

  Total Capital Cost= total process unit costs + general facilities + Engineering permitting and
  start-up + Contingencies + Working capital

• **Reported Cost Year:** 2002

• **Battery limit conditions:** Mid/Central P =7.59 MPa, Distributed P = 40.53 MPa

• **Plant configuration:**
No detailed description of the process flow. The technology description section mentions feed purification units, reformers, shift reactors and H2 purification through PSA or chemical absorption. The cost estimates include the cost of H2 compressors.

**Efficiency:**

(Mid-size: SMR 72.0% LHV, compressor 0.7 kW/kg/hr)
(Central: SMR 76.2% LHV, compressor 0.5 kW/kg/hr)
(Distributed: SMR 60.0% LHV, compressor 2.0 kW/kg/hr)

Reference (F): IEAGHG 2017 [32]

**Description:** The report performs a techno-economic assessment of a hydrogen merchant SMR plant, compares several CO2 sequestration pathways and compares them to a base case

**Source of cost information:** Foster Wheeler In-house data, process modeling

**Cost Aggregation:**

Project contingency is 20% of the TPC for all process units.

TPC = Direct material + Construction + EPC services + Other Costs + Contingency

TCR = TPC + Spare parts cost + Start up costs + Owner’s cost + IDC + Working Capital

**Reported Cost Year:** Q4 of 2014 (initial costs in Euros)

**Battery limit conditions:** T = 40°C, P = 2.5 MPa

**Plant configuration:**

The report divided the plant into three units, Hydrogen plant (Desulfurizer, pre-reformer, reformer, High-Temperature shift reactor and PSA), Balance of Plant (instrument/plant air system, flare, Nitrogen and drain systems) and a power Island (steam turbine and generator)

**Plant performance:** 9.046K=kg CO2/kg H2, gross power output from Co-gen 11.5 MWe, export 9.918MWe

**Efficiency:** (conversion: Pre-reformer + SMR = 84.6%, HTS = 68%, PSA rec = 90%)

Reference (G): Foster Wheeler 1996 [39]

**Description:** The report examines H2 production processes from fossil fuel along with CO2 capture.
• **Source of cost information:** Foster Wheeler In-house data, information received from equipment suppliers and Icarus

• **Cost Aggregation:**
  
  construction costs include direct and indirect labor, management and supervision, temp facilities and overheads
  
  Total Capital Cost = Equipment & Materials + Installation + Engineering, management, commissioning and training + Catalyst and chemicals + Client’s cost + Contingency

• **Reported Cost Year:** Q4 of 1995

• **Battery limit conditions:** T= 45 °C, P= 6 MPa

• **Plant configuration:**
  
  The process includes desulfurizer, pre-reformer, reformer, high-temperature shift reactor, PSA and H₂ Compressor, turbo-generators for self-power generations (7175 kW, for a 94,000 Nm³/h of hydrogen)

• **Efficiency:** (Conversion (pre-reformer + reformer =87.7 %), PSA=88 % (recovery)
Appendix C: Cost and scaling factors used for estimating ATR, CO\(_2\) capture and \((\text{H}_2/\text{CO}_2)\) compression units’ capital costs.

<table>
<thead>
<tr>
<th>Auto Thermal Reformer Cost Estimating Parameters</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Cost Item</strong></td>
</tr>
<tr>
<td>SMR reforming Reactor</td>
</tr>
<tr>
<td>ATR reforming reactor</td>
</tr>
<tr>
<td>Air separation unit</td>
</tr>
<tr>
<td>Water gas shift (WGS) reactor cost</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>CO(_2) Compression unit cost</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Compressor Unit Cost</strong></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>CO(_2) Capture Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Capture unit Cost</strong></td>
</tr>
</tbody>
</table>
units respectively. Costs were updated to 2020 using CEPCI.
- Scaling factor was 0.7 and reference scale and cost were adopted from IEAGHG study.

<table>
<thead>
<tr>
<th>Hydrogen Compression unit cost</th>
<th>74.7 M$ (MEA)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Compressor unit Cost</strong></td>
<td></td>
</tr>
<tr>
<td>- Base cost for a plant scaled on per unit kW power consumption. Costs were updated to 2020 using CEPCI.</td>
<td></td>
</tr>
<tr>
<td>- Scaling factor was 0.8225 adopted from source.</td>
<td>$1962/kW</td>
</tr>
</tbody>
</table>
## Appendix D: Economic inputs assumptions for estimating the operating costs of the reforming units

<table>
<thead>
<tr>
<th>Cost Item</th>
<th>Unit/ description</th>
<th>value</th>
<th>Source</th>
</tr>
</thead>
<tbody>
<tr>
<td>Labor cost factor</td>
<td>Factor multiped by labor salary to adjust for location</td>
<td>1.3</td>
<td>[102]</td>
</tr>
<tr>
<td>Labor shift position</td>
<td></td>
<td>5</td>
<td></td>
</tr>
<tr>
<td>Operators per shift</td>
<td></td>
<td>3</td>
<td></td>
</tr>
<tr>
<td>Operator Salary</td>
<td>$/Year</td>
<td>US$50,000</td>
<td></td>
</tr>
<tr>
<td>Supervision Cost</td>
<td>25% of labor cost</td>
<td></td>
<td>[89]</td>
</tr>
<tr>
<td>Overhead</td>
<td>40% of Labor and supervision cost</td>
<td></td>
<td>[32]</td>
</tr>
<tr>
<td>Maintenance</td>
<td>3% of BEC</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Local tax and fees</td>
<td>0.5% of TPC</td>
<td></td>
<td>[40]</td>
</tr>
<tr>
<td>Insurance</td>
<td>0.5 of TPC</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Corporate overhead</td>
<td>65% of labor, supervision &amp; overhead</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Natural gas cost</td>
<td>CAD$/GJ</td>
<td>2.83</td>
<td>[103]</td>
</tr>
<tr>
<td>Process water cost</td>
<td>$/m$^3</td>
<td>$2</td>
<td>Assumed</td>
</tr>
<tr>
<td>Electricity cost</td>
<td>CAD$/kWh</td>
<td>$0.07</td>
<td>[104]</td>
</tr>
<tr>
<td>CO$_2$ storage cost</td>
<td>$/kgCO$_2$</td>
<td>$6</td>
<td>Assumed</td>
</tr>
</tbody>
</table>
Appendix E: Operating parameters for CO$_2$ compression model
The compressor model for CO$_2$ assumed 7 stages for compression from battery limit conditions (0.16-.29 MPa) to 11 MPa. Thermodynamic properties were estimated using Coolprops. Isentropic efficiency was taken as 85% and mechanical efficiency was 99%[105].
Appendix F: Equations for calculating the levelized costs of production and transport

The levelized Cost of Hydrogen (LCOH):

The levelized cost of hydrogen LCOH production is calculated in a similar manner to the levelized cost of electricity LCOE based on a discounted cash flow analysis, where the LCOH the is the cost that gives an NPV value of zero using an Internal Rate of Return (IRR) which is equal to the Discount Rate (DR) hence this metric gives a high-level economic evaluation. This analysis doesn’t consider cost variation and long-term inflation through the project life but considers interest during construction. Equation 16 was used to calculate the LCOH.

\[
LCOH = \frac{\sum_{n=1}^{N} \frac{CAPEX_n + OPEX_n}{(1+i)^n} \cdot H2\, produced}{\sum_{n=1}^{N} \frac{H2\, produced}{(1+i)^n}}
\]

Where:

LCOH is in $/kgH₂, N is project lifetime, CAPEX is the annualized capital expenditure in this case calculated using the TASC and a capital recovery factor, OPEX is the operational expenditure accounting for fuel, labor, maintenance and other operational cost.

When considering the total levelized cost of delivery the levelized cost of transport was added as a yearly operational expense (OPEX) in the equation above.

**CO₂ Avoidance Cost (CAC)**

As discussed the CO₂ avoidance cost (CAC) is evaluated by comparing the cost and specific emissions of a plant with capture facility against the respective values in a reference facility. The equation used is shown below:

\[
CAC = \frac{LCOH_{CCS} - LCOH_{Reference}}{CO2\,Emissions_{Reference} - CO2\, Emissions_{CCS}}
\]

Where:

CAC is in $/tCO₂, CO₂ emissions are in tons CO₂ per ton H₂ produced, the reference case is an SMR without capture facility.
Appendix G: Sensitivity analysis on the levelized production cost for 350 tH\textsubscript{2}/day reforming units

Figure 34: Sensitivity analysis on SMR with pre-combustion capture based on varying several input variables (base values (0%) correspond to 65% H\textsubscript{2} compressor, 8% interest, 2.83CAD/GJ NG price and 1.48 location factor)

Figure 35: Sensitivity analysis on cost of production from ATR with capture based on varying several input variables (base values (0%) correspond to 65% H\textsubscript{2} compressor, 8% interest, 2.83CAD/GJ NG price and 1.48 location factor)
Appendix H: Summary of input parameters used for analysis in Sections 2, 3 and 4

<table>
<thead>
<tr>
<th>item</th>
<th>value</th>
<th>unit</th>
<th>description</th>
<th>source</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>H₂ production</strong></td>
<td>350,000</td>
<td>kg/day</td>
<td>Scale of Hydrogen production for all units</td>
<td>assumed</td>
</tr>
<tr>
<td><strong>SMR conversion</strong></td>
<td>84.5</td>
<td>%</td>
<td>Includes pre-reformer conversion, same rate for Pre and Post capture configurations.</td>
<td></td>
</tr>
<tr>
<td><strong>SMR PSA recovery</strong></td>
<td>89</td>
<td>%</td>
<td>Recovery rate relative in to input stream, same rate for Pre and Post capture configurations.</td>
<td></td>
</tr>
<tr>
<td><strong>WGS conversion</strong></td>
<td>68.3</td>
<td>%</td>
<td>One reactor (HTS). Same rate for Pre and Post capture configurations.</td>
<td></td>
</tr>
<tr>
<td><strong>CO₂ capture rate (Post)</strong></td>
<td>90</td>
<td>%</td>
<td>Capture rate relative to input stream (Not overall CO₂ removal rate) mole basis</td>
<td>[32]</td>
</tr>
<tr>
<td><strong>Increment NG for CC (Post)</strong></td>
<td>9</td>
<td>%</td>
<td>Incremental increase of total NG consumption in post combustion capture relative to base case.</td>
<td></td>
</tr>
<tr>
<td><strong>CO₂ capture rate (Pre)</strong></td>
<td>98</td>
<td>%</td>
<td>Capture rate relative to input stream (Not overall CO₂ removal rate) mole basis</td>
<td></td>
</tr>
<tr>
<td><strong>Increment NG for CC (Pre)</strong></td>
<td>4</td>
<td>%</td>
<td>Incremental increase of total NG consumption in pre combustion capture relative to base case.</td>
<td></td>
</tr>
<tr>
<td><strong>ATR PSA recovery</strong></td>
<td>90</td>
<td>%</td>
<td>Recovery rate relative in to input stream</td>
<td></td>
</tr>
<tr>
<td><strong>WGS conversion</strong></td>
<td>94.45</td>
<td>%</td>
<td>Two reactors (HTS/LTS)</td>
<td>[51]</td>
</tr>
<tr>
<td><strong>CO₂ Capture rate ATR</strong></td>
<td>95</td>
<td>%</td>
<td>Capture rate relative to input stream (Not overall CO₂ removal rate) mole basis</td>
<td></td>
</tr>
<tr>
<td><strong>Increment NG for CC (ATR)</strong></td>
<td>0</td>
<td>%</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td><strong>Currency conversion</strong></td>
<td>1.26</td>
<td>%</td>
<td>To convert from CAD to USD 2020</td>
<td>-</td>
</tr>
<tr>
<td><strong>Catalyst &amp; Chemicals</strong></td>
<td>2.5</td>
<td>%</td>
<td>Taken as % from TPC of respective unit</td>
<td>[32]</td>
</tr>
<tr>
<td><strong>Interest rate</strong></td>
<td>8</td>
<td>%</td>
<td>Assumed</td>
<td>-</td>
</tr>
<tr>
<td><strong>Lifetime</strong></td>
<td>25</td>
<td>years</td>
<td></td>
<td>-</td>
</tr>
<tr>
<td><strong>Process contingency</strong></td>
<td>35</td>
<td>%</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Project contingency</strong></td>
<td>40</td>
<td>%</td>
<td>Taken as % from the Bare Erected Costs</td>
<td>[38], [49]</td>
</tr>
<tr>
<td><strong>EPC</strong></td>
<td>10</td>
<td>%</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>H₂ Compressor efficiency</strong></td>
<td>65</td>
<td>%</td>
<td>Isentropic efficiency</td>
<td>[88]</td>
</tr>
<tr>
<td><strong>CO₂ compressor efficiency</strong></td>
<td>85</td>
<td>%</td>
<td>Isentropic efficiency</td>
<td>[105]</td>
</tr>
<tr>
<td>Location factor</td>
<td>1.465</td>
<td>To transfer cost from Gulf coast basis to AB (Heartland)</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>----------------</td>
<td>-------</td>
<td>---------------------------------------------------------</td>
<td>---</td>
<td></td>
</tr>
<tr>
<td>Power consumption of MEA unit</td>
<td>0.025 kWh/kg CO₂</td>
<td>For SMR with Post-Combustion</td>
<td>[57]</td>
<td></td>
</tr>
<tr>
<td>Power Consumption of MDEA unit</td>
<td>0.0125 kWh/kg CO₂</td>
<td>For SMR with Pre-Combustion and ATR</td>
<td>[58]</td>
<td></td>
</tr>
<tr>
<td>Battery limit H₂ compression</td>
<td>10 MPa</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Battery Limit compression of CO₂</td>
<td>11 MPa</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
</tbody>
</table>
## Appendix I: Summary of input variables for analysis in Section 3 for H₂ and CO₂ pipeline model assumptions

<table>
<thead>
<tr>
<th></th>
<th>Value</th>
<th>Unit</th>
<th>Description</th>
<th>Source</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cost of Steel</td>
<td>0.764</td>
<td>USD/kg</td>
<td>Assumed to include fabrication (+15%) cost</td>
<td>[93]</td>
</tr>
<tr>
<td>Annual Pipe O&amp;M</td>
<td>1.5</td>
<td>%</td>
<td>Taken as % from total pipe CAPEX</td>
<td></td>
</tr>
<tr>
<td>Interest rate</td>
<td>15</td>
<td>%</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Lifetime (economic)</td>
<td>25</td>
<td>years</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Location factor</td>
<td>~2.3</td>
<td></td>
<td>Location factor changes according to pipe length and diameter</td>
<td></td>
</tr>
<tr>
<td>H₂ pipe outlet pressure</td>
<td>5.5</td>
<td>MPa</td>
<td></td>
<td></td>
</tr>
<tr>
<td>CO₂ pipe outlet pressure</td>
<td>9</td>
<td>MPa</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Appendix J: Comparison of CO₂ pipeline cost model against ACTL and Quest projects CO₂ costs

The exact design parameters of the reference projects could not be replicated in the model due to lack of data on specifications of steel grade and price quotations used. However, assuming an X52 steel grade with a design pressure of 15 MPa the model capital cost estimate for the ACTL pipeline was 0.94 M$/km versus 1.07 M$/km for the actual cost of the project giving a 14% error. While in the case of Quest project the model results were 0.83 M$/km versus 1.1 M$/km giving a 24% error.
Appendix K: Comparisons of all reforming options for DS1 and DS2 with a 680 gCO$_{2eq}$/kWh grid

Figure 36: Impact of carbon tax on the total levelized cost of hydrogen delivery (production +pipeline) including baseline SMR for demand scenario 1 (DS1) at 680 gCO$_{2eq}$/kWh grid intensity

Figure 37: Impact of carbon tax on the total levelized cost of hydrogen delivery (production +pipeline) including baseline SMR for demand scenario 2 (DS2) at 680 gCO$_{2eq}$/kWh
Figure 38: Impact of carbon tax on the total levelized cost of hydrogen delivery (production + pipeline) including baseline SMR for demand scenario 1 (DS1) at 420 gCO2eq/kWh grid intensity

Figure 39: Impact of carbon tax on the total levelized cost of hydrogen delivery (production + pipeline) including baseline SMR for demand scenario 2 (DS2) at 420 gCO2eq/kWh grid intensity
# Appendix L: Breakdown of life cycle specific emissions of produced hydrogen for the Demand Scenarios DS at 680 gCO\(_{2eq}\)/kWh

## Specific Carbon Intensity (kgCO\(_{2eq}\)/kgH\(_2\))

<table>
<thead>
<tr>
<th></th>
<th>Direct</th>
<th>Grid</th>
<th>Upstream (4.2 gCO(_{2eq})/MJ NG)</th>
<th>H(_2) &amp; CO(_2) Pipeline boosters</th>
<th>CO(_2) pipeline leakage</th>
<th>Total (rounded)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Baseline SMR</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>8.99</td>
<td>-0.053</td>
<td>0.66</td>
<td>0.579</td>
<td>0</td>
<td>10.18</td>
</tr>
<tr>
<td>LO2</td>
<td>8.99</td>
<td>-0.73</td>
<td>0.66</td>
<td>0</td>
<td>0</td>
<td>8.93</td>
</tr>
<tr>
<td><strong>SMR-Pre</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>4.14</td>
<td>0.57</td>
<td>0.68</td>
<td>0.579</td>
<td>-</td>
<td>5.97</td>
</tr>
<tr>
<td>LO2</td>
<td>4.14</td>
<td>-0.10</td>
<td>0.68</td>
<td>0.006</td>
<td>0.05</td>
<td>4.77</td>
</tr>
<tr>
<td><strong>SMR-Post</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>1</td>
<td>0.48</td>
<td>0.73</td>
<td>0.579</td>
<td>-</td>
<td>2.78</td>
</tr>
<tr>
<td>LO2</td>
<td>1</td>
<td>-0.13</td>
<td>0.73</td>
<td>0.023</td>
<td>0.09</td>
<td>1.71</td>
</tr>
<tr>
<td><strong>ATR-PRE</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>1.27</td>
<td>1.19</td>
<td>0.70</td>
<td>0.579</td>
<td>-</td>
<td>3.74</td>
</tr>
<tr>
<td>LO2</td>
<td>1.27</td>
<td>0.48</td>
<td>0.70</td>
<td>0.017</td>
<td>0.08</td>
<td>2.55</td>
</tr>
<tr>
<td><strong>DS 1 266 tH(_2)/day</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Baseline SMR</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>8.99</td>
<td>-0.07</td>
<td>0.66</td>
<td>0.344</td>
<td>0</td>
<td>9.92</td>
</tr>
<tr>
<td>LO2</td>
<td>8.99</td>
<td>-0.74</td>
<td>0.66</td>
<td>0</td>
<td>0</td>
<td>8.91</td>
</tr>
<tr>
<td><strong>SMR-Pre</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>4.14</td>
<td>0.55</td>
<td>0.68</td>
<td>0.344</td>
<td>-</td>
<td>5.71</td>
</tr>
<tr>
<td>LO2</td>
<td>4.14</td>
<td>-0.12</td>
<td>0.68</td>
<td>0.002</td>
<td>0.05</td>
<td>4.75</td>
</tr>
<tr>
<td><strong>SMR-Post</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>1</td>
<td>0.45</td>
<td>0.73</td>
<td>0.344</td>
<td>-</td>
<td>2.52</td>
</tr>
<tr>
<td>LO2</td>
<td>1</td>
<td>-0.15</td>
<td>0.73</td>
<td>0.014</td>
<td>0.09</td>
<td>1.68</td>
</tr>
<tr>
<td><strong>ATR-PRE</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>LO1</td>
<td>1.27</td>
<td>1.185</td>
<td>0.70</td>
<td>0.344</td>
<td>-</td>
<td>3.50</td>
</tr>
<tr>
<td>LO2</td>
<td>1.27</td>
<td>0.47</td>
<td>0.70</td>
<td>0.011</td>
<td>0.08</td>
<td>2.53</td>
</tr>
<tr>
<td><strong>DS 2 110 tH(_2)/day</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>