A Techno-Economic Assessment of Sustainable Large Scale Hydrogen Production from Renewable and Non-Renewable Sources

by

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Abstract

In recent times, the imperative to mitigate greenhouse gas (GHG) emissions that emanate from a multitude of sectors in the global energy economy has achieved unprecedented and widespread consensus. Depending on the energy resource and method used to produce hydrogen, it offers a compelling alternative to GHG intensive fossil-fuel based energy carriers. In this thesis, a techno-economic assessment of large scale, sustainable, hydrogen production pathways is addressed through the development of integrated techno-economic models. The hydrogen produced from the aforementioned pathways is used to displace hydrogen derived from natural gas - steam methane reforming (SMR), which dominates hydrogen supply, particularly in the bitumen upgrading industry in Western Canada, and oil refining complexes around the globe. As such, there is a considerable demand for low-GHG hydrogen production pathways that are cost competitive with SMR. Hydrogen production from wind energy, hydropower, natural gas and coal were assessed in the work carried out. In the case of wind energy, a wind-hydrogen plant with energy storage was evaluated. The hydrogen production cost from this pathway ranged from \$3.37 - \$15.06/kg H_2, depending on the electrolyser size and whether or not existing wind farm infrastructure is used. The optimal electrolyser-battery configuration for the plant consists of 81 units of a 3496 kW (760 Nm^3/hr) electrolyser and 360 MWh (60 units) of battery capacity. Additionally, it was observed that for a particular electrolyser-battery configuration, the minimum hydrogen production cost occurs when their respective capacity factors are approximately equivalent. For the hydropower-hydrogen plant the hydrogen production cost ranged from \$1.18 to \$5.35/kg H_2, depending on the electrolyser size and the use of existing hydropower assets. The optimal plant configuration consists of 90 units of a 3496 kW (760 Nm^3/h) electrolyser. In the case of coal and natural gas, integrated techno-economic models for underground coal gasification (UCG) and SMR with or without carbon capture and sequestration (CCS), were developed. The competitiveness of UCG and SMR is highly sensitive to the natural gas price. Hydrogen production from UCG without CCS (\$1.92/kg H_2) is slightly less competitive relative to SMR (\$1.87/kg H_2). Hydrogen production from UCG-CCS (\$2.28/kg H_2 to \$2.92/kg H_2) is slightly more competitive relative to SMR-CCS (\$2.31/kg H_2 to \$2.60/kg H_2). Overall, for the techno-economic conditions considered, hydrogen production from hydropower proved to be the pathway that is most competitive with SMR in Western Canada.

Preface

In Chapter 1, sections 1.2 to 1.4 are based upon a book chapter publication - *Olateju, B. and A. Kumar. Clean energy-based production of hydrogen: An energy carrier. Handbook of Clean Energy Systems; 2015. 1–30.* I am the principal author of this book chapter, with Kumar A. providing supervisory oversight, intellectual guidance and support with the manuscript composition.

Chapter 2 is one of three refereed journal publications, emanating from a body of work carried out in the area of grid connected wind-hydrogen plants. Chapter 2 has been published as a refereed journal publication - *Olateju, B., A. Kumar, M. Secanell. A techno-economic assessment of large scale wind hydrogen production with energy storage in Western Canada. International Journal of Hydrogen Energy, 2016. 41 (0): p. 8755-8776.* I was responsible for model development, analysis and manuscript composition. A. Kumar and M. Secanell both provided supervisory oversight, intellectual guidance and support with the model development and manuscript composition. The other two publications from the body of work in this area are:

Olateju, B., J. Monds, A. Kumar. Large scale hydrogen production from wind energy for the upgrading of bitumen from oil sands. Applied Energy, 2014. 118 (0): p. 48-56

Olateju, B., A. Kumar. Hydrogen production from wind energy in Western Canada for upgrading of bitumen from oil sands. Energy, 2011. 36 (11): p. 6326-6339

Chapter 3 has been submitted for publication and is under review - *Olateju, B. and A. Kumar. A techno-economic assessment of hydrogen production from hydropower in Western Canada for the upgrading of bitumen from oil sands. Energy, 2016 (in-review).* I was responsible for model development, analysis and manuscript composition. A. Kumar provided supervisory oversight, intellectual guidance and support with the model development and manuscript composition.

Chapter 4 has been published as a refereed journal publication - *Olateju, B. and A. Kumar. Techno-economic assessment of hydrogen production from underground coal gasification (UCG) in Western Canada with carbon capture and sequestration (CCS) for upgrading bitumen from oil sands. Applied Energy, 2013. 111: p.428 – 440.*

In Chapter 5, Table 5.1 is based on a book chapter publication - *Olateju, B. and A. Kumar. Clean Energy-Based Production of Hydrogen: An Energy Carrier. Handbook of Clean Energy Systems; 2015. 1–30*. I am the principal author of this book chapter, with Kumar A. providing supervisory oversight, intellectual guidance and support with the manuscript composition.

This thesis is dedicated to my Mom, Adeola Khadijat Olateju aka 'Iyabidemi' and my Dad, Monsuru Alade Olamide Olateju – you will always remain my foremost and dearest teachers in life.

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List of Abbreviations

AC	Alternating Current
AESO	Alberta Electric System Operator
ASU	Air Separation Unit
ATR	Auto-Thermal Reforming
BOP	Balance of Plant
bpd	Barrels Per Day
C	Carbon
CCS	Carbon Capture and Sequestration
CH_4	Methane
CO	Carbon monoxide
CO_2	Carbon dioxide
CO_{2e}	Carbon dioxide equivalent
CRIP	Continuous Retraction Injection Point
CSS	Cyclic Steam Simulation
DC	Direct Current
DCF	Discounted Cash Flow
DOE	Department of Energy
EOR	Enhanced Oil Recovery
FIT	Feed-In-Tariffs
FUNNEL	Fundamental Engineering Principles Based Model
g	Gram
GHG	Greenhouse Gas
GJ	Gigajoule
GOA	Government of Alberta

GTL	Gas-To-Liquids
GW	Gigawatt
H_2	Hydrogen
H_2O	Water
HHV	Higher Heating Value
HTE	High Temperature Electrolysis
IEA	International Energy Agency
IRENA	International Renewable Energy Agency
IRR	Internal Rate of Return
kg	kilogram
KOH	Potassium Hydroxide
kW	Kilowatt
kWh	Kilowatt-Hour
LHV	Lower Heating Value
LNG	Liquified Natural Gas
MDEA	Methyldiethanolamine
MJ	Megajoule
MPa	Megapascal
MW	Megawatt
MWh	Megawatt-hour
NaOH	Sodium Hydroxide
NREL	National Renewable Energy Laboratory
O_2	Oxygen
OEM	Original Equipment Manufacturing
O & M	Operating and Maintenance
PEM	Proton Exchange Membrane
POX	Partial Oxidation
PPA	Power Purchase Agreement

PSA	Pressure Swing Adsorption
ROV	Real Options Valuation
SAGD	Steam Assisted Gravity Drainage
SCO	Synthetic Crude Oil
SCOT	Shell Claus Off-gas Treating
SMR	Steam Methane Reforming
SOEC	Solid Oxide Electrolytic Cells
UCG	Underground Coal Gasification
US	United States
WACC	Weighted Average Cost of Capital
WCSB	Western Canadian Sedimentary Basin
WGS	Water gas shift
WTW	Well-To-Wheel

Chapter 1[1]

Introduction

1.1 Background & Motivation

Hydrogen is often regarded as an emerging energy carrier, which is capable of driving key sectors of the energy economy, sustainably, in a greenhouse gas (GHG) emissions constrained energy future. However, hydrogen can only be compatible with a GHG constrained energy future if it is produced using environmentally sustainable means and resources. The use of hydrogen as a transportation fuel, as well as its use in the power sector, has garnered significant research interest [1-11]. However, the consumption of hydrogen in the transportation sector is often based upon future oriented scenarios, where a significant demand for hydrogen exists (increased hydrogen vehicle penetration). While this long-term reality is probable, a significant demand for hydrogen currently exists in the bitumen upgrading and oil refining complexes in Western Canada and around the globe. Consequently, this creates ample opportunity to mitigate a formidable amount GHG emissions through the provision of 'green' hydrogen from sustainable pathways. This GHG mitigation opportunity exists in principle; it will only become a commercially viable endeavour if the economics of sustainable hydrogen pathways are competitive with the incumbent fossil fuel norm. As a result, the techno-economic assessment of sustainable and large scale hydrogen pathways is central to the work carried out in this thesis. It is important to mention that the body of work presented here is part of a broader research objective that seeks to assess hydrogen

[1] Part of this chapter is based upon a book publication. Please refer to: *Olateju, B. and A. Kumar. Clean energy-based production of hydrogen: An energy carrier. Handbook of Clean Energy Systems; 2015. 1–30*

pathways from a techno-economic and life cycle GHG emissions standpoint. Informed by the work presented in this thesis, subsequent research, which addresses the life cycle GHG footprint of hydrogen pathways, have been published [12-14]. It is also important to mention that the techno-economic models developed in this thesis are applicable to other sectors of the economy, where a significant hydrogen demand currently exists e.g. in the ammonia industry, or those that are anticipated in future e.g. the power and transportation sector.

1.1.1 The Bitumen Upgrading Industry in Western Canada

As of 2014, Canada harbored the third largest proven oil reserves in the world, with the oil sands[2] of Alberta accounting for about 97% of Canada's total oil reserves [17, 18]. Specifically, the total remaining established reserves of crude bitumen in Alberta, as of 2014, amounted to 166.3 billion barrels [19]; for the same year, bitumen production was 2.2 - 2.3 million barrels per day (bpd) [18, 19]. Alberta's oil sands play a vital role regarding Canadian and broader North American energy security. Moreover, they are of increasing significance in the global crude oil market as they constitute 60% of non-state-owned crude oil reserves in the globe [20] – providing a relatively accessible and competitive playing field, for the development of crude oil resources. Oil sands are currently produced via surface mining methods as well as in situ methods (the predominant in situ methods include: steam assisted gravity drainage (SAGD) and cyclic steam stimulation (CSS)). In this light, about 80% and 20% of oil sands reserves are recoverable via in situ and surface mining methods, respectively [19].

[2] Oil sands are mixture of sand, water, clay and bitumen. Bitumen is an unconventional hydrocarbon that is viscous (at room temperature) and dense, containing heavy hydrocarbon molecules in terms of their molecular weight, and higher quantities of sulphur, asphaltenes and metals in comparison to conventional crude oil grade [15, 16].

Oil sands industry operators are involved in the production of non-upgraded bitumen and/or upgraded bitumen in the form of synthetic crude oil (SCO). Bitumen is upgraded to reduce its viscosity and increase its hydrogen to carbon ratio [15, 21, 22]; the aforementioned properties facilitate pipeline transportation and increased market value respectively [21, 22]. As of 2014, about 47% (~ 1.1 million bpd) of the bitumen produced was upgraded to SCO in Alberta [19]. SCO production is expected to experience material growth in forthcoming years - 32.4% by 2024 and 45.4% by 2030 [18, 19]; however, these production forecasts will ultimately be governed by the degree to which oil prices rebound/collapse, and in particular, the future trend of the light vs. heavy oil price differential.

The production of SCO in the oil sands industry is of particular relevance to the body of work contained in this thesis, as the process of upgrading bitumen to SCO is highly hydrogen intensive; with the average hydrogen consumption, considering a multitude of bitumen extraction pathways and upgrading technologies, estimated to be 3.4 kg H_2/barrel of SCO [23]. Additionally, bitumen upgrading incurs a significant greenhouse gas emissions (GHG) footprint, accounting for 33% of total GHG emissions from the oil sands industry in 2012 [24]. It is worth mentioning that bitumen upgrading consists of primary and secondary upgrading operations [15, 22, 25]. Primary upgrading is used to increase the hydrogen to carbon ratio of bitumen, and can be achieved via carbon-rejecting or hydrogen additive processes i.e. coking or hydrocracking, respectively [15, 22, 25]. In secondary upgrading, hydrogen is used to displace unwanted impurities such as sulphur, nitrogen and metals [15, 22, 25].

The predominant means through which hydrogen is produced for the upgrading of bitumen, along with other heavy crude grades, world over, is via steam methane reforming (SMR) [25-29]. The dominance of SMR for upgrading purposes can be attributed to a multitude of factors; notwithstanding, the most notable of these is the abundance of relatively inexpensive natural gas in Alberta, and more broadly, North America. That said, although natural gas prices have been relatively low in recent years (for the most part, Henry Hub natural gas prices have remained below $5/GJ since 2011[30]), they are often difficult to predict into the future and have a marked history of price volatility [31]. Furthermore, SMR is a fossil fuel intensive process; resulting in a significant GHG emissions footprint ranging from 11,000-13,000 tonnes CO_{2e}/tonne H_2 [32-35]. Thus, the continued reliance on SMR as the principal hydrogen production pathway, raises questions about the economic risk and environmental sustainability of the bitumen upgrading industry, especially in the long-term. In this light, there is a need for alternative GHG-neutral hydrogen production pathways, which are economically competitive with SMR.

1.1.2 GHG emissions in Bitumen upgrading – A market access determinant

Against the backdrop of climate change mitigation efforts world over, the global energy market has become increasingly averse to energy commodities with a significant GHG emissions footprint, of which crude oil is no exception. With a number of jurisdictions imposing low carbon fuel standards[3] and an explicit levy on GHG emissions (according to the World Bank [36], 40 countries have put a price on CO_2 emissions or are planning to implement them), the life cycle GHG emissions of a given crude grade is becoming an increasingly important market access determining factor. Although it is increasing in importance, it is currently a relatively 'soft' market

[3] Prime examples include: California low carbon fuel standard and the European Union fuel quality directive.

access determinant. However, in the energy market of the future, which is likely to have more pervasive and stringent constraints on GHG emissions, life cycle GHG emissions could become a key market access determinant.

Previous research has shown that on a well-to-wheel (WTW) basis, SCO produced from steam assisted gravity drainage (SAGD) or surface mined bitumen has an incremental GHG footprint of 10-16g CO_{2e}/MJ gasoline, relative to other crude grades i.e. Bachaquero (Venezuelan) and Maya (Mexican) heavy crude [37]. Other authors have indicated that oil sands crude grades (including SCO) incur 17% more GHG emissions relative to conventional crudes imported by US refineries (which is the largest market for oil sands crude) [38]. For completeness, the same authors demonstrate that GHG emissions from crudes derived from oil sands, are within range or lower than crude grades such as Californian Kearn River (US) and Bonny Light (Nigeria) [37, 38]. Nonetheless, the need for enhancing the environmental (GHG) competitiveness of SCO in particular, vis-à-vis other crudes, remains evident.

In the context of upgrading emissions, some studies estimate that hydrogen production, as a component of bitumen upgrading emissions, accounts for 28.5% and 54.2% of the total emissions for delayed coking and hydrocracking upgrading technologies (including secondary upgrading), respectively [22] . Other authors have demonstrated that hydrogen production (in a delayed coking upgrader) accounts for 44% of the GHG emissions [39]. The aforementioned studies underscore the material impact that hydrogen production has on bitumen upgrading GHG emissions. With this in mind, the environmental competitiveness of SCO in the energy market can be enhanced by the injection of a GHG neutral, sustainably derived, hydrogen feedstock into the upgrading arm of the oil sands industry's value added chain.

The subsequent sections of this chapter address the sources, production processes, transportation modes and storage options for hydrogen. The appreciation of these elements will play an integral role in the development of a sustainable hydrogen economy, geared towards servicing the bitumen upgrading industry.

1.2 Hydrogen – Primary Sources

1.2.1 Hydrocarbons

Hydrocarbons such as natural gas, coal, oil, coke etc., constitute the predominant means through which hydrogen is produced [40-42]. Natural gas in particular is the single most prevalent feedstock for the synthesis of hydrogen; accounting for about 48% of production globally [40, 41]. Its high hydrogen to carbon ratio and reduced GHG footprint relative to coal and other hydrocarbons makes it particularly attractive. However, as mentioned in previous sections, natural gas has significant price volatility along with heterogeneous global prices with wide margins. As a result, coal and alternative hydrocarbons are the preferred sources of hydrogen in jurisdictions with relatively high natural gas prices. Moreover, due to its multi-faceted use in industry, the opportunity cost associated with natural gas as a hydrogen feedstock is significant [43].

About 95% of hydrogen production is captive [44, 45] – that is to say, the site of production and consumption is coincident. As a result, the need for a reliable, continuous, feedstock supply (which translates into high capacity factors for hydrogen plants), becomes evident. Apart from this, feedstocks with a high degree of mobility from their point source, which can facilitate large scale centralized[4] hydrogen production are sought, due to their superior economics. These are criteria that hydrocarbons readily meet. Furthermore, the widespread nature and abundance of hydrocarbon resources globally, along with the mature technology used to access and transport them in a cost effective manner, are some other key contributing factors to their dominance in the hydrogen supply market.

1.2.2 Biomass

Biomass has the advantage of being a renewable energy resource, which is considered nearly GHG neutral over its life cycle. However, due to the role of biomass as a life supporting resource, hydrogen is preferably produced from lignocellulosic biomass (e.g. agricultural, forestry or municipal waste etc.) [35, 46-48]. This is to negate potential competition with food supply in particular, and consequently, food prices, along with other existing commercialization activities of biomass resources. In similar fashion to fossil fuels, centralized large scale hydrogen production can be achieved with biomass, along with comparable capacity factors [49, 50]. However, the feedstock supply of biomass is dependent upon the scale of socio-economic activities by other industries (e.g. forestry, agriculture, construction, municipalities etc) which produce lignocellulosic biomass. Thus, a feedstock reliability risk is inherent to this source of hydrogen. In addition, although biomass is not a stranded resource, it lacks the mobility of hydrocarbons, and hence incurs an elevated transportation cost – this can be as high as 33-50% of the production cost [51, 52]. Development of innovative biomass transportation (e.g. pipeline transport of biomass in

[4] It is important to highlight the fact that decentralized hydrogen production is also possible with hydrocarbons; specifically, natural gas. However, this comes at an elevated production cost.

the form of slurries) and conversion technologies are needed to make this resource economically attractive.

1.2.3 Water

In similar fashion to biomass, water is renewable and is also a life supporting resource. Unlike hydrocarbons (natural gas), for the production of hydrogen, water cannot serve as both the fuel and the feedstock. In essence, it is dependent upon another energy carrier, electricity, to serve as the fuel for electrolytic hydrogen production. In the case where electricity is not used for splitting water into its constituent elements, process heat, as in the case of thermo-chemical water splitting cycles, is provided by an alternative source e.g. nuclear or solar energy [53-55]. However, because water is a hydrogen source with zero carbon content in its pure state, it holds significant promise as a clean energy pathway for hydrogen production.

1.3 Hydrogen – Production Processes

1.3.1 Thermo-chemical conversion processes

Thermo-chemical conversion processes are those in which heat energy is used to facilitate a chemical reaction that ultimately leads to the synthesis of hydrogen; these are applicable to hydrocarbon and biomass feedstock. The thermo-chemical processes considered in this study include: reforming, gasification and pyrolysis of the pertinent hydrogen feedstock highlighted in section 1.2. It is worth highlighting that all the aforementioned thermo-chemical processes share some salient unit operations as illustrated in the generalized process diagram (which includes the option of CCS) highlighted in Figure 1.1.

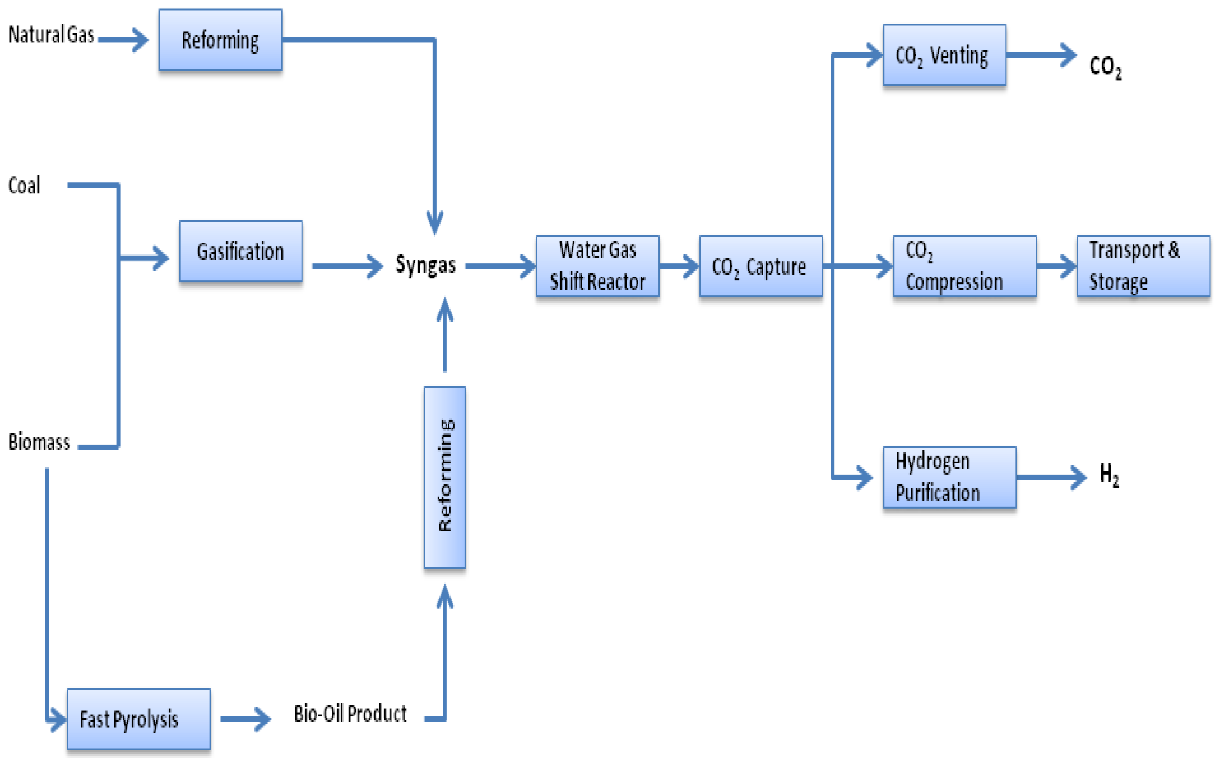

Figure 1.1: Thermo-chemical conversion processes – salient unit operations

1.3.1.1 Reforming

Steam methane reforming (SMR) involves the endothermic catalytic production of syngas, stemming from a reaction between natural gas (methane) and steam, under the influence of heat; syngas is ultimately converted to mainly hydrogen and carbon-dioxide, via the water gas shift reaction [56-58]. SMR is the principal means through which natural gas is converted to hydrogen [40, 41, 58-60] . However, it is worth mentioning that other SMR hybrids and variants such as auto-thermal reforming (ATR) and partial oxidation (POX) exist. The principal advantage with ATR is the fact that it is self-sustaining with regards to its thermal energy requirement, and thus requires no external heat source as in the case of SMR [58, 59, 61]. The attractive feature of POX systems relates to its increased tolerance to sulphur, and non-catalytic operation [59, 60] - which depresses operating costs and enhances process reliability as the risk of catalyst poisoning is eradicated.

The favorable features of ATR and POX systems are accompanied by some undesirable characteristics which contribute to their limited adoption in industry relative to SMR. First, ATR and POX systems have lower syngas quality in terms of hydrogen concentration [59, 62]. With SMR, hydrogen concentration in the syngas produced is in the order of 70-80%; while in the case of ATR and POX this varies between 40-50% [60, 62]. Due to their relatively low H_2 to CO ratio, POX systems in particular, are regarded to be more suitable for the Fischer-Tropsh process as opposed to hydrogen production [58, 59]. In addition, the operational problems caused by soot formation in both ATR and POX systems, along with the augmented costs associated with using oxygen in the case of POX, hinders their competiveness against SMR even further [58, 59].

One of the principal challenges associated with SMR has to do with the judicious use of natural gas as both a feedstock and a fuel for the process [57, 61], which often presents a dilemma, taking into account optimum economic and GHG emissions metrics. In SMR, purge gas from hydrogen purification (which contains CH_4, H_2, H_2O, CO_2, CO along with other species) is used to supplement fuel requirements [57, 61]; hence, the calorific value quality of this gas is important. However, increased hydrogen production in SMR comes at the expense of increased fuel consumption, due to the reduced calorific value quality of the purge gas [57]. Thus, depending on the pertinent operating and economic conditions, a delicate balance must be struck. On another note, desulphurization of the natural gas feed is also a key process component; as the catalysts involved in SMR are quite susceptible to sulphur poisoning [57, 59, 60] – which can have detrimental effects on process operation, cost and efficiency. That being said, feedstock preparation and syngas purification operations associated with SMR, are relatively less intricate and rigorous in comparison to other thermo-chemical processes.

As seen in Figure 1.1, in the context of thermo-chemical hydrogen production, the water gas shift (WGS) reaction plays a pivotal role in the conversion of syngas to hydrogen and CO_2. This is usually achieved in two stages via high temperature and low temperature shift reactors. The reason behind this is that equilibrium favors the reaction products at low temperature, while the reaction rate needs a high temperature to be practical [57, 59, 61, 63]. An inexpensive iron-based catalyst is used to promote the conversion of CO to CO_2; while a relatively costly copper-based catalyst is utilized for enhancing hydrogen production [46, 57, 64]. Upon completion of the WGS reaction, hydrogen and CO_2 are separated, using chemical or physical capture systems (e.g. MDEA, Selexol™ etc), with the CO_2 being vented or sequestered depending on the hydrogen system in question i.e. with or without CCS (see Figure 1.1). Hydrogen is then purified to the desired degree; this is carried out predominantly via pressure swing adsorption (PSA) systems which are capable of producing hydrogen with 99% purity [57, 63, 64]. Depending on the production pressure of hydrogen and its end-use, it can be compressed and stored, or compressed and transported.

In the subsequent thermo-chemical processes to follow, the unit operations downstream of syngas production will not be addressed; as they are of a similar nature to those discussed above (see Figure 1.1).

1.3.1.2 Gasification

Gasification, as the term suggests, involves the transformation of a hydrocarbon or biomass feedstock from a solid state to a gaseous state, using a limited supply of air or oxygen in tandem with steam to produce syngas [41, 52]; which can be used to produce hydrogen. As mentioned earlier, the gasification of coal and biomass draws a number of parallels with process operations of SMR, although some unique differences and subtleties are apparent.

The gasification of coal, depending on the gasifier type, occurs at elevated pressures (29 -70 bar) and temperatures (ranging from 1200^0C - 1800^0C) in comparison to SMR [63-65]; especially where oxygen serves as the gasification agent. In the case of entrained flow gasifiers (which are likely to be the most pervasive gasifier type), considering their relatively high hydrogen output pressure, coal can be introduced into the gasifier in the form of dry feedstock or slurry [44, 63-66]. The gasification of coal involves a detailed and rigorous syngas cleanup process mainly due to the levels of impurities (most notably, ash) and undesired by-product species being more pronounced in the case of coal relative to natural gas [64, 65]. Unlike SMR, the desulphurization of coal is carried out after syngas production via dedicated acid gas capture systems such as Selexol™, along with Claus and SCOT units [63, 65, 66]. Furthermore, coal serves as the feed, and indirectly, as the fuel of the process; the use of purge gas and heat recovery systems suffice for the mitigation of heat duties [63, 65, 66]. This is particularly true in systems that employ the separation of CO_2 (which can be vented or sequestered) along with sulphur from the syngas, leading to an increased calorific value of the purge gas [63, 65, 66].

In the case of biomass, gasification and pyrolysis represent the two thermo-chemical pathways with the highest potential for commercial-scale hydrogen production [41, 52]. Biomass gasification can be achieved using similar gasification technologies as in the case of coal; however, gasification temperature and pressure are reduced [52, 65]. Furthermore, special attention is paid to the reduction of biomass chip size and moisture content to facilitate optimal heat transfer during gasification [41, 52]. In similar fashion to coal, biomass gasification requires a more detailed syngas clean up system relative to natural gas. Additionally, in terms of the energy required, biomass gasification can be auto-thermal or allothermal [67, 68]. Auto-thermal gasification in a similar light to auto-thermal reforming, is self-sustaining in terms of the energy required; while allothermal gasification requires an external heat source [67, 68]. On another note, hydrogen yield from biomass gasification varies significantly (84.1 – 72.2 kg/dry tonne biomass), depending on the feedstock in question [52].

1.3.1.3 Pyrolysis

Pyrolysis of biomass involves heating at a modulated rate (i.e. fast or slow pyrolysis), in the absence of oxygen, to produce bio-oil, char, and vapor in varying concentrations [41, 46, 69, 70]. On average, fast pyrolysis leads to a yield of 75 wt%, 12 wt%, and 13 wt%, for bio-oil, char and vapor, respectively [46, 69, 70]. On the other hand, slow pyrolysis has a bias towards char production [46, 69, 70]. The premise behind bio-oil production in the context of hydrogen pathways is to increase the volumetric energy density and cost efficiency associated with the transport of biomass to the hydrogen production site [46]. Hydrogen is produced from the reforming of bio-oil, in a comparable manner to SMR; thus, it becomes evident that hydrogen production from biomass pyrolysis is a hybrid of two unique thermo-chemical processes that require their own unique infrastructure. This introduces increased process complexity and can potentially elevate costs.

On another note, with pyrolysis, a number of key technical conditions and parameters need to be satisfied for pragmatic bio-oil yield. First, in the same vein as biomass gasification, particular emphasis is placed on the minimization of biomass chip size to allow for effective heat transfer [46, 69, 70]. Additionally, a sufficiently high rate of heat transfer to ensure fast-pyrolysis is crucial [70]. More importantly, a limited window of opportunity exists to harness bio-oil from the initial vapor gas produced (operating temperature is usually in the range of 425 - 500 ^{0}C [46, 70]; this is because above 400^{0}C, the macro-polymer compounds contained therein are unstable and continue to experience thermal cracking as time progresses [46, 69]. As a result, in pyrolysis operations, limited vapor gas residence time and fast quenching (in the order of hundreds of milliseconds) are imperative to ensure optimal bio-oil production [46, 69, 70].

1.3.2 Electrochemical conversion processes

Electrochemical conversion processes are those in which electrical energy is utilized in the promotion of a chemical reaction that leads to the evolution of hydrogen as a product. In the context of hydrogen pathways, the electrolysis of water is the most widely adopted electrochemical process and the most mature from a technological standpoint; existing for over 200 years [42, 60, 71]. The basic principle behind electrolysis involves the passage of direct electric current through electrodes placed in an aqueous electrolytic solution [42]. With the aid of the electrolyte, the water molecule is split into its constituent elements of hydrogen (gas) and oxygen (gas) at the cathode and anode, respectively, via electrochemical reactions (see Figure 1.2) [42].

The performance of water-based electrolytic hydrogen production is underpinned by the cost, quality and reliability of the electricity supply. Furthermore, the hydrogen yield also depends on the electrolyser performance (which houses the electrolytic cell). The current electrolyser (electrolysis) technologies that exist in literature can be categorized into three main types namely: alkaline electrolysers, proton exchange membrane (PEM) electrolysers, and high temperature electrolysis (HTE) via solid oxide electrolytic cells (SOEC). Manage et al. (2011) [60] provides a summary of the electrochemical reactions and data involved in each technology (see Table 1.1).

The alkaline electrolyser usually has an aqueous solution of potassium hydroxide (KOH) and water as its electrolyte [60, 72, 73]; however, the use of sodium hydroxide (NaOH) with water, is also a possibility [60, 73]. The concentration of KOH in alkaline electrolysers is normally limited to the range of 20-30 wt% due to the competing factors of higher ionic conductivity (which increases efficiency) and corrosive effects, with increased concentration [60, 74]. Alkaline electrolysers require the purification of the hydrogen produced [60], which can be achieved by in-built dehumidifiers/driers in the electrolyser units [72]. In addition, they also need cooling water to maintain an operating temperature in the range of about 70ºC - 90ºC [60, 75]. Alkaline electrolysers have nominal efficiencies ranging from 64 - 85% (60.93 – 45.8 kWh/kg H_2 HHV of hydrogen) based on the values specified by several studies [60, 72, 75-78]; with an operational life of 15 or 20 years in the case of certain studies [76, 79]. However, the intermittency of power supply associated with renewable energy systems such as wind power, have adverse effects on the operational life of alkaline electrolysers [75, 76]. In some cases the operational life of the electrolyser is reduced by a factor of 2 [76]. With regards to costs, the capital costs of alkaline electrolysers are relatively cheaper in comparison to other technologies [76, 80], with no compromise on the purity of the hydrogen output. Furthermore, they are able to be produced in the megawatt scale [78, 81], which allows for large scale hydrogen production in comparison to other electrolytic pathways – potentially facilitating reductions in the cost per unit of hydrogen produced.

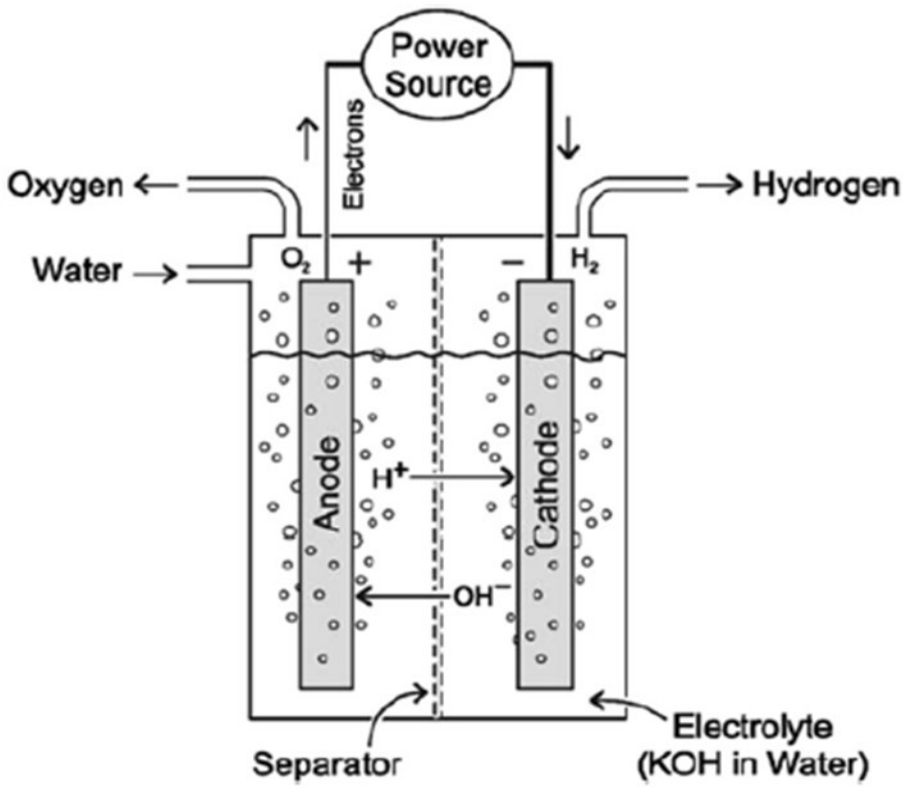

Figure 1.2: Typical alkaline electrolytic cell [82] – Reproduced with permission from Bhandari, Trudewind and Zapp (2014). © Elsevier B.V.

Table 1.1: Electrochemical reactions for Alkaline, PEM, and SOEC electrolysis technologies – Based on research by Manage et al. (2011) with minor modifications.

	Alkaline	**PEM**	**SOEC**
Operating temperature (^{0}C)	70-90	<100	500-1000
Electrolyte - Ion	OH^-	H^+	O^{2-}
Electrolyte - Material	$KOH_{(aq)}$, $NaOH_{(aq)}$	Sulfonated polymers e.g. Nafion™	Yttria stabilized zirconia (YSZ), Scandia-stabilized zirconia
Cathode - Reaction	$2H_2O + 2e^- \rightarrow H_2 + 2OH^-$	$2H^+ + 2e^- \rightarrow H_2$	$H_2O + 2e^- \rightarrow H_2 + O^{2-}$
Cathode - Material	Nickel with platinum catalytic coating	Platinum black	Nickel-YSZ cermet
Anode - Reaction	$2OH^- \rightarrow 0.5\ O_2 + H_2O + 2e^-$	$H_2O \rightarrow 0.5\ O_2 + 2H^+ + 2e^-$	$O^{2-} \rightarrow 0.5\ O_2 + 2e^-$
Anode - Material	Nickel or copper coated with metal oxides	Iridium oxide (IrO_2)	Perovskite oxides (e.g. lanthanum manganate)

The PEM electrolyser on the other hand has a solid polymer electrolyte, as opposed to an aqueous solution [60, 73, 83, 84]. The need for purification of hydrogen is avoided with this technology, and the efficiency of this electrolyser has been shown to be superior to an alkaline electrolyser of the same capacity [76]. Furthermore, they are more robust in comparison to alkaline electrolysers with regards to handling intermittent power supply [75, 76]. However, PEM electrolysers have higher capital costs [76, 80]; with relatively short operational lives [76, 79]. Nonetheless, the fact that PEM electrolysers are now transitioning from the kW to the MW scale [85], enhances their potential cost competitiveness against alkaline electrolysers significantly.

HTE is concerned with the electrolysis of steam as opposed to the use of water, with the use of SOEC [60, 73]. To the author's knowledge, this technology is still at the laboratory scale/demonstration stage with no evidence of industrial commercialization [60, 80]. The premise of HTE is to reduce the electrical energy requirement of the electrolysis process by utilizing the heat energy of steam in achieving the enthalpy of reaction needed to initiate the electrolysis reactions at the cathode and anode [73, 74]. Elevated temperatures have been proven to provide increased ionic conductivity of the electrolyte, as well as the enhancement of the electrode surface kinetics [60, 73, 74, 83]. Hence, this technology has the potential to attain increased electrolysis efficiency, and the production of hydrogen at a reduced cost. SOEC materials that are well suited for elevated temperatures are still the subject of research and development [60]. Notwithstanding, considerable progress has been made in recent decades, and SOEC technology is considered to be at the brink of commercialization by some authors [86].

1.4 Hydrogen – Transportation and storage alternatives

1.4.1 Hydrogen Transport

The low volumetric energy density and molecular weight of hydrogen (gas) at standard conditions present formidable challenges to its efficient and economical transport as well as storage [87]. With regards to transportation, the means by which hydrogen is delivered to the consumer is primarily a function of production scale and transport distance. The low volumetric energy density of hydrogen is usually compensated for by compressing hydrogen gas to elevated levels, or by transporting it in its denser cryogenic liquid phase. Three predominant hydrogen transmission modes are employed in practice: compressed (tube trailer) gas trucks, liquefied (cryogenic) hydrogen trucks, and hydrogen pipelines [87]. According to some authors [87] , for low production rates (< 600 kg H_2/day) compressed gas trucks are used, for moderate rates (600 kg H_2/day < flow rate < 2400 kg H_2/day) cryogenic trucks are utilised, and for flow rates greater than 2400 kg H_2/day, pipeline transmission is adopted. This thesis has a bias towards large scale transportation of hydrogen; hence, the use of hydrogen pipelines will be given particular focus.

The compression of hydrogen to elevated levels of about 60-70 bar is necessary for pipeline transport [63, 88]. For the same energy throughput, hydrogen compressors consume 2 - 3 times the amount of energy as natural gas compressors [75, 88]. Depending on the transportation distance and project specific pressure losses (due to section changes or changes in elevation etc), the requirement for booster stations will vary. With that being said, due to the lower pressure losses incurred with hydrogen pipelines, fewer booster stations are required relative to natural gas pipelines [75]. Apart from enhancing the volumetric energy density of hydrogen, the use of high pressure pipelines facilitates continuous supply of hydrogen and negates loading and unloading requirements; these features cannot be readily achieved with other modes of hydrogen delivery.

More importantly, pipelines have strong economies of scale inherent to their cost structure; hence, they are particularly attractive for large scale hydrogen transport. However, the need for specialized pipeline seals and the material specification for the pipeline itself will be vital in mitigating hydrogen leakage and the embrittlement of steel respectively [78, 89, 90]. These aforementioned parameters can impact costs significantly. Nonetheless, increasing experience is being gained considering the construction and operation of large scale hydrogen pipelines, with a number of projects currently in operation around the globe. The largest hydrogen pipeline network in the globe will transport an aggregate volume of 1.3 million Nm^3/day [91].

1.4.2 Hydrogen Storage

The safe and reliable storage of hydrogen is crucial for the emergence and sustenance of a hydrogen economy. For the large scale storage of hydrogen, the use of compressed hydrogen gas and liquefied hydrogen storage vessels are the two most developed and widespread technologies available today [89]. That being said, to a reduced extent, the use of underground storage media such as salt caverns, aquifers etc., is another alternative for large scale hydrogen storage [88, 89]. In the subsequent subsections, an overview of compressed hydrogen, liquefied hydrogen and underground hydrogen storage is presented.

1.4.2.1 Compressed Hydrogen Gas Pressure Vessels

Above ground pressure vessels for hydrogen storage are designed in accordance with pressure vessel theory, in cylindrical as well as spherical shapes [89], to maintain structural integrity and avoid stress concentrations associated with rigid corners and vertices. The materials used for their design include austenitic stainless steel (AISI 316 & 304), along with composite materials [89, 92, 93]. The use of austenitic stainless steel depresses the manufacturing and material costs incurred, while composite materials are likely to introduce prohibitive storage costs especially for large scale applications [89]. The primary advantage of using a composite material is mostly to do with the reduction of weight [89, 92, 93] - which is crucial for dynamic automotive applications, but a non-factor for the large scale stationary applications considered here [89]. The volume of these vessels depends on the application in question and can vary from the equivalent of hours of hydrogen production to months [88]. Pressures for compressed storage can extend from 70 – 185 bar depending on the storage volume [88, 89].

1.4.2.2 Liquified Hydrogen Cryogenic Storage

Hydrogen can be stored in liquefied form at a temperature of -253^{0}C and at pressures ranging from 1 - 5 bar [89, 94]. As a result, a significant amount of compression energy is saved in comparison with compressed gas storage. Furthermore, liquefied hydrogen storage has strong economies of scale, making it compatible with large scale centralized systems [88]. However, liquefaction comes with a substantial energy penalty; accounting for 33% - 40% of the energy content of the stored hydrogen [75, 88, 89, 92]. Moreover, hydrogen boil-off losses are inevitable with liquefied storage; due to internal heat transfer (stemming from the incomplete ortho-to-para hydrogen conversion during liquefaction) as well as external heat transfer from the surrounding environment [89, 94, 95]. Heat transfer to liquid hydrogen leads to the formation of vapor which can lead to storage tank over pressure and thus has to be vented periodically [89]. As a result, the minimization of heat transfer to liquid hydrogen is imperative for the practicality of this storage option.

Liquified hydrogen storage vessels are built in the form of a two-wall construction; the separate walls create shells which allow for the introduction of a vacuum gap between them [89, 95]. This gap is used to minimize convective and conductive heat transfer, with the placement of insulating and heat shielding materials such as perlite, silica aerogel, fused alumina, mylar etc, between the shells [89]. In the case of large vessels an additional outer wall is built, with the use of liquid nitrogen as a heat shield in the additional gap created [89].

1.4.2.3 Underground Hydrogen Storage

Although the underground storage of compressed hydrogen gas might appear to be a relatively novel industrial norm, the sub-surface storage of hydrogen rich gases has been in existence for decades [89, 95, 96]. The storage of town gas (a gas with a high concentration of hydrogen) in underground reservoirs, such as rock and salt caverns, porous rocks and aquifers, has been carried out in a number of jurisdictions globally [89, 95]. Thus, a significant degree of confidence and experience is associated with the storage of compressed hydrogen in these different media. However, underground storage (e.g. storage in aquifers) comes with a relatively reduced amount of control over the storage conditions of the gas; with issues of permeability, caprock integrity, and water contamination [97]. These aforementioned parameters need to be thoroughly understood, characterized and effectively managed. In addition, the selection process of suitable sites is a relatively rigorous and time consuming endeavor in comparison to compressed gas pressure vessels or cryogenic storage. Furthermore, logistical issues are more likely to come to the fore with this form of storage; as the site of production and storage does not necessarily overlap, thus requiring additional transportation infrastructure. From a more technical standpoint, the retrieval of hydrogen from underground reservoirs can be problematic. As much as 50% of the stored volume of hydrogen can remain as cushion gas; reducing the reliability of these systems [89]. Nonetheless, the fact that material and manufacturing costs are significantly reduced with underground storage, makes this a compelling alternative.

1.5 Problem Statement

The problem under investigation is concerned with the limitations of existing techno-economic models, prevalent in the literature, that are used in assessing sustainable hydrogen production pathways; specifically: wind-hydrogen systems and hydropower-hydrogen systems. Additionally, the lack of efficacious techno-economic frameworks that assess emerging sustainable hydrogen pathways – in this case, underground coal gasification (UCG) with carbon capture and sequestration (CCS), falls within the envelope of the problem under investigation. As such, the problem under investigation is multifaceted in nature and is situated in the literary context of each of the three hydrogen pathways being assessed (i.e. wind-hydrogen, hydropower-hydrogen and UCG-CCS hydrogen production), with SMR serving as a benchmark for comparisons. As a result, the detailed treatment of the problem under investigation is carried out elaborately in the subsequent chapters of this thesis. This section is intended to provide a succinct overview, such that readers can appreciate the rationale behind the thesis objectives. Figure 1.3 illustrates the hydrogen production pathways assessed in this thesis.

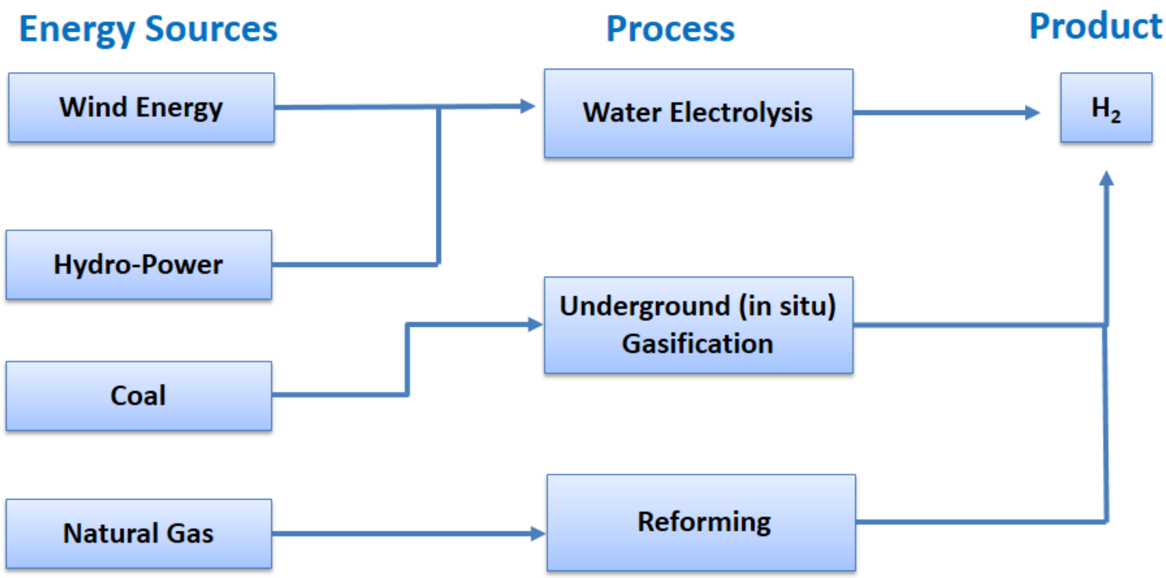

Figure 1.3: Hydrogen production pathways assessed

1.5.1 Techno-Economic Assessment of Grid-Connected Electrolytic Wind-Hydrogen Systems

The techno-economic modelling of wind-hydrogen systems encompasses the energy generation from the wind turbine(s) as well as hydrogen production from electrolyser(s); taking into account the context in which wind-hydrogen plants operate i.e. the electricity grid, along with the ancillary equipment and unit operations involved e.g. H_2 compressor, H_2 pipeline and H_2 storage/electrical energy storage etc.

A pervasive modeling approach in the existing literature is the characterization of the wind variability, via the use of statistical methods, most notably, a Weibull probability density function - to estimate energy, and consequently, hydrogen production [98-104]. While the efficacy of the Weibull function in resolving the variability of the wind speed is not disputed, a certain degree of error is inherent in the estimation of the probability of occurrence of a given wind speed magnitude. This in turn, hinders the accuracy of the energy production estimates. Moreover, the Weibull function is limited in its ability to accurately resolve bimodal or multimodal wind distributions which arise from unique climatic conditions [105, 106]. The model developed in this thesis utilizes real time hourly wind energy generation data, thus its accuracy is not hindered by the limitations of the Weibull function.

With regards to the electricity grid, the assessment of wind-hydrogen systems in existing studies often involves hydrogen production costs being ascertained using fixed/average electricity prices [77, 98, 107-109]. This modeling paradigm does not account for the dynamic pricing environment which is indicative of the increasingly liberalized electricity markets across the globe [110, 111]. As such, the hydrogen production cost estimates that result can be limited in accuracy. The dynamic pricing environment facilitates opportunities to take advantage of peak/premium electricity prices; owing to the electricity price differential that exists, depending on the time of day. Premium electricity prices provide an opportunity to enhance the competiveness of wind-hydrogen systems. To take advantage of these prices however, dynamic energy storage is required. In this thesis, batteries (electrical energy storage) are used to realize differential pricing opportunities on the electrical grid, as opposed to a hydrogen storage[5]–fuel cell pathway utilized

[5] For the model developed in this thesis, hydrogen demand variations are assumed to be negligible, hence, hydrogen storage when necessary, is facilitated through the use of a pipeline; this is in line with the options for hydrogen storage highlighted in previous studies [78, 88].

by other authors [10, 11, 99, 107, 112]. The use of hydrogen-fuel cell configurations incur a significant cost, and more importantly, a low round trip efficiency of about 25-30% [10, 11, 99, 107, 112]; this limits the added competitive advantage that can be harnessed from price differentials in a liberalized electricity market.

Another common norm in the modeling of wind-hydrogen systems is the direct coupling of the electrolyser unit to the wind turbine, without intermediate energy storage [98-101, 109, 112, 113]. By doing this, authors make the implicit assumption that the electrolyser units will perform at their nominal efficiencies, despite the perturbed and often transient nature of the power input – this is particularly pertinent to alkaline electrolysers, which is the focus of the models developed here. In practice, the variability of the wind energy input has an adverse effect on the nominal efficiency of the electrolyser as well as its operating life [109, 114]. Therefore, without energy storage to smoothen the erratic profile of the wind energy input and dispense this energy in a more uniform fashion to the electrolyser, hydrogen production has a likelihood of being over-estimated, with costs under-estimated. Thus, the significance of energy storage in wind-hydrogen systems becomes evident, with regards to economics and efficiency considerations.

Previous studies have also presented ‘element-level’ electrolyser models, where the authors have characterized the operating voltage of the electrolyser as a function of its operating temperature, current density and characteristic over-potentials (which are also temperature dependent), via a hybrid of electro-chemical and thermodynamic relations [103, 115, 116]. This modeling paradigm allows for a more robust and dynamic resolution of the electrolyser nominal efficiency, facilitating a more precise simulation of hydrogen yield from a given electrolyser. However, these aforementioned models are predicated upon empirical equations; the coefficients of which have to be ascertained for particular electrolyser via experiments [115, 116]. As such, the models cannot be readily generalized; limiting their utility in contexts where a broad number of electrolyser models/capacities and number of units need to be evaluated as part of an integrated energy system.

With this in mind, a systems-level approach is implemented in the modeling of the performance of the electrolyser, based on its salient characteristics and the energy input from the battery (which ultimately emanates from the wind turbine). This study assumes that the nominal efficiency of the electrolyser does not change materially during its operation, due to the role of the battery, which delivers a power supply with significantly reduced perturbation - in other words, the electrolyser operates at the constant, nominal current density. In essence, the framework adopted in this thesis facilitates modeling flexibility and generalization, without compromising on the accuracy of hydrogen yield (vis-à-vis element level models).

1.5.2 Techno-Economic Assessment of Grid-Connected Electrolytic Hydropower-Hydrogen Systems

The existing literature that pertains to the techno-economic modelling of hydropower-hydrogen systems is quite limited in the recent decade when compared to other hydrogen pathways. Notwithstanding, a multitude of systems have been proposed, which are assessed from a techno-economic standpoint with varying degrees of rigor. Each of these systems involves electrolytic hydrogen production using the electrolysis of water. The systems put forward in literature can be broadly categorized into three main themes. First, a number of studies have proposed small scale hydropower-hydrogen systems, where hydrogen is used to service the electricity/heat generation needs of remote off-grid communities [117-121]. Alternative models are premised upon the use of excess water from hydropower reservoirs, which are ‘spilt’ without harnessing their potential for hydrogen production [122, 123]. In these studies, the hydrogen produced is used in the electricity generation (peak-load applications/energy storage), transportation (fuel-cell vehicles) or in the value added industries i.e. food, pharmaceuticals and ammonia industries. Furthermore, the dedicated or off-peak use of hydropower plants for hydrogen production has been the basis of other models [124-129], where hydrogen has similar end uses as in the previous category of studies. Other related research to capitalize on hydropower-hydrogen potential, involve its use for methanol production [130].

From the perusal of previous studies, a number of noteworthy trends have been identified. With the exception of the model presented by Bellotti et al. 2015 [124], a number of models do not address the optimal sizing (to minimize cost) and configuration of the electrolyser plant. Having said that, Bellotti et al. (2015) [124] does not consider the impact of the hydropower-hydrogen plant functioning in a liberalized electricity market, and the effect of the dynamic electricity prices therein, on sizing considerations. Furthermore, hydrogen yield and electrolyser energy consumption are based upon idealized efficiencies, generic correlations, and assumptions of key metrics (e.g. electrolyser capacity factor) in some cases [117, 118, 122, 125, 126]. Moreover, fixed electricity prices which are not indicative of the dynamics of a liberalized electricity market, are often used to estimate hydrogen production costs [119, 123, 125, 126]. Additionally, some models proposed have limited transparency in terms of the key techno-economic data used, due to confidentiality and other factors [128]. Apart from this, a limited amount of studies present integrated hydropower-hydrogen models, which take a holistic account of all unit operations involved from hydrogen production, to its delivery to the end user.

1.5.3 Techno-Economic Assessment of Hydrogen Production from UCG with and without CCS

The synergy that can be realised from hydrogen production via UCG-CCS in the Western Canadian context, deserves research attention, as the conditions that exist in terms of the geology and resource wealth facilitate a fertile ground for its implementation. The published literature on the integration of CCS with large scale energy systems is in-depth and multi-faceted in nature; with technological, economic, environmental, and regulatory aspects being addressed exhaustively [131-138]. Contrastingly, in the case of UCG, much of the focus in the existing literature has been geared towards UCG process simulation and optimisation; as well as environmental impact monitoring [139-144]. While this is undoubtedly important, the appraisal of UCG-CCS from a techno-economic perspective in published literature is scarce. To the knowledge of the author, no previous research has provided an integrated techno-economic model for UCG hydrogen production with and without CCS. More often than not, the appraisal of UCG in a techno-economic context is qualitative with limited detail [145-147]. Admittedly, this is likely due to the infancy of the technology and the limited operational experience on a commercial large scale. The above ground (surface) gasification of coal and the subsequent processing of the syngas evolved for hydrogen production is a mature well understood technology. The process methodology for syngas-H_2 conversion is identical for both surface gasification and UCG plants. In both cases syngas is processed above ground, and the syngas composition, temperature and pressure are of similar magnitudes [145, 146]. As a result, this enables the techno-economic modelling of UCG to be carried out credibly, within reason. The techno-economic modelling of UCG-CCS with the explicit consideration of its apparent cost-competitiveness and environmental risks is needed to provide a quantitative and qualitative view of its utility as a hydrogen production pathway in comparison to conventional methods such as SMR-CCS. In addition, this will also facilitate the identification of areas of cost minimisation and key sensitivities.

1.6 Thesis Objective

The overall objective of this thesis is the development of integrated techno-economic models to ascertain the cost of hydrogen production from renewable and non-renewable energy resources. The Alberta bitumen upgrading industry serves as the platform of application for the models developed. The specific objectives are as follows:

1.6.1 Electrolytic Grid-Connected Wind-Hydrogen Systems

- The development of an integrated grid-connected wind-H_2 techno-economic model, with energy storage (battery), for the production of renewable hydrogen and estimation of costs, in a liberalized electricity market with dynamic prices.
- The development of an energy management algorithm for wind-H_2 plants with energy storage, which is a function of the wind turbine energy yield, dynamic electricity price, electrolyser and battery performance specifications.
- The development of a techno-economic framework for the determination of the optimal electrolyser size, number of electrolyser units and energy storage capacity, which yields a minimum hydrogen production cost, for wind-hydrogen systems with energy storage.

1.6.2 Hydrogen Production from UCG with and without CCS

- The development of an integrated techno-economic model for hydrogen production from underground coal gasification (UCG) with and without carbon capture and sequestration (CCS).
- A comparative techno-economic assessment of UCG and SMR hydrogen production pathways, for a multitude of scenarios.

1.6.3 Electrolytic Grid-Connected Hydropower-Hydrogen Systems

- The development of an integrated grid-connected hydropower-H_2 techno-economic model, for the production of renewable hydrogen and estimation of costs, in a liberalized electricity market with dynamic prices.
- The development of a techno-economic framework for the determination of the optimal electrolyser size and number of electrolyser units, which yields a minimum hydrogen production cost, for hydropower-hydrogen systems.

1.7 Scope and Limitations

- The energy resources assessed for hydrogen production in this thesis are those of particular pertinence to Western Canada; these include: wind energy, hydropower, coal and natural gas.

- The electrolysers utilized in the techno-economic models developed might not be indicative of the current state of the art – this is due to the proprietary nature of electrolyser performance specification data.
- Where applicable, the production cost of hydrogen is the sole metric used to define the optimal hydrogen plant configuration.
- The hydrogen production cost estimates provided in this thesis are reflective of Alberta conditions – Alberta and its bitumen upgrading industry, served as the platform of application for the models developed. Notwithstanding, the application and data inputs can be adjusted to suit the specificities of other jurisdictions.

1.8 Report Organization

This thesis consists of five chapters; it is a consolidation of refereed journal publications and a book chapter. Each chapter in this thesis is intended to be read independently. Due to the 'stand-alone' nature of the chapters, some overlap of data and concepts occur.

The current chapter provides the reader with an introduction to the thesis, its objectives and implications on the bitumen upgrading industry. It also provides the reader with an overview of the hydrogen economy; addressing the sources, conversion processes, transportation and storage alternatives for hydrogen. Furthermore, the problem statement being addressed in the thesis is outlined.

Chapter 2 presents a techno-economic assessment of large scale wind-hydrogen production with energy storage in Western Canada.

Chapter 3 discusses a techno-economic assessment of large scale hydropower-hydrogen production for bitumen upgrading in Western Canada.

Chapter 4 presents a techno-economic assessment of hydrogen production from UCG in Western Canada with CCS for upgrading bitumen from oil sands.

Chapter 5 articulates conclusions from the research and also highlights areas for future work along with future trends.

An appendix is provided at the end of the thesis with relevant equations and the publication list of the author is also included.

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Chapter 2[1]

A Techno-Economic Assessment of Large Scale Wind-Hydrogen Production with Energy Storage in Western Canada

2.1 Introduction

The production of hydrogen via steam methane reforming (SMR) in crude oil refining complexes is facing intense scrutiny [1-4]. Aside from regulations on greenhouse gas (GHG) emissions, the oil refining industry faces growing pressure to comply with increasingly stringent non-GHG environmental regulations – most notably, sulphur content in fuels [2-4]. Additionally, heavier crude oil grades with higher sulphur and nitrogen content are a growing portion of the global supply mix [2, 4]. To facilitate compliance with fuel regulatory standards via hydrogen intensive hydrotreating and hydrocracking processes, the oil refining sector has experienced a formidable rise in hydrogen demand [2, 4].

In Alberta, Western Canada, the bitumen upgrading industry has a considerable hydrogen demand for the production of synthetic crude oil (SCO); this need for hydrogen is expected to amount to 3.1 million tonnes / year by 2023 [5]. Steam-methane reforming is the single most prevalent pathway for hydrogen production; accounting for 48% of global supply [6]. While SMR is economically attractive, it produces a significant GHG emission footprint, i.e., in the range of

[1] This chapter is based upon a journal publication. Please refer to: *Olateju, B., A. Kumar, M. Secanell. A techno-economic assessment of large scale wind hydrogen production with energy storage in Western Canada. International Journal of Hydrogen Energy 2016. 41 (0). p:8755-8776*

11,000 -13,000 tonnes of CO_2 equivalent per tonne of hydrogen produced (based on HHV of H_2) [7-10]. Moreover, the industry-wide dominance of SMR as the hydrogen production pathway of choice creates significant economic exposure to natural gas prices; which although relatively low in recent times, have a history of significant price volatility (see Figure 2.1). Hence, in an increasingly GHG constrained energy market where economic competitiveness is increasingly coupled to environmental stewardship and the social license to operate, an alternative environmentally benign H_2 pathway, which remains economically palatable, is desired in the bitumen upgrading industry.

Wind powered electrolytic (via water electrolysis) hydrogen production is considered to incur the lowest life-cycle GHG emissions of all hydrogen pathways, amongst a number of authors [8, 9, 11, 12]. Furthermore, in the context of renewable energy pathways, with the exception of hydropower, wind energy has the lowest levelized cost of electricity ($/kWh) in most jurisdictions around the world [13-15]. Thus, a promising opportunity exists for cost-efficient, environmental benign, hydrogen production and GHG mitigation with this pathway. In Alberta, wind energy has an estimated generating potential of about 64 GW [16]. As of 2014, wind power accounted for about 9% of the electricity generation capacity of the province [17]; with coal power serving as the dominant base load electricity supply. In order to evaluate the techno-economic prospects of large scale wind-hydrogen production in Alberta, the installed wind energy capacity as of 2009 (563 MW) is utilized for electrolytic hydrogen production in this paper. This is to say that the hydrogen production costs determined in this paper are specific to a wind energy capacity of 563 MW, unless otherwise specified.

Previous studies that address hydrogen production from renewable and non-renewable studies in Alberta have been conducted [6, 7, 18-26]; in particular, Olateju, Monds & Kumar (2014) [6] as well as Olateju & Kumar (2011) [25], have addressed grid-connected wind-hydrogen systems at small and large plant scales. However, to the knowledge of the authors, no previous studies have addressed wind-hydrogen systems with the use of energy storage (battery energy storage) to ascertain the optimal plant configuration (minimum H_2 cost) for a given capacity of wind energy. More generally, the existing research regarding the techno-economic assessment of wind-hydrogen systems is quite extensive; with a multitude of modeling approaches, implicit and explicit assumptions and limitations therein. This paper aims to improve upon the limitations associated with the seemingly normative techno-economic modeling frameworks, widely adopted in the pertinent literature. Some of these modeling trends and their associated drawbacks are highlighted in the subsequent paragraphs vis-à-vis the methodology incorporated in this paper.

From an economic standpoint, the assessment of grid connected wind-hydrogen systems in existing studies often involves hydrogen production costs being ascertained using fixed/average electricity prices [25, 27-30]. This modeling paradigm does not account for the dynamic pricing environment which is indicative of the increasingly liberalized electricity markets across the globe [31, 32]. As such, the hydrogen production cost estimates that result can be limited in accuracy. The dynamic pricing environment facilitates opportunities to take advantage of peak/premium electricity prices; owing to the electricity price differential that exists, depending on the time of day (see Figure 2.2). Premium electricity prices provide an opportunity to enhance the competiveness of wind-hydrogen systems. To take advantage of these prices however, dynamic energy storage is required. In this study, batteries (electrical energy storage) are used to realize differential pricing opportunities on the electrical grid, as opposed to a hydrogen storage[2]–fuel cell

[2] In this paper, hydrogen demand variations are assumed to be negligible, hence, hydrogen storage when necessary, is facilitated through the use of a pipeline (see section 2.2); this is in line with the options for hydrogen storage highlighted in previous studies [15, 33].

pathway utilized in previous research [27, 34-37]. The use of hydrogen-fuel cell configurations incur a significant cost, and more importantly, a low round trip efficiency of about 25-30% [27, 34-37]; this limits the added competitive advantage that can be harnessed from price differentials in a liberalized electricity market.

A common norm in the modeling of wind-hydrogen systems is the direct coupling of the electrolyser unit to the wind turbine, without intermediate energy storage [6, 25, 30, 34, 35, 38, 39]. By doing this, authors make the implicit assumption that the electrolyser units will perform at their nominal efficiencies, despite the perturbed and often transient nature of the power input – this is particularly pertinent to alkaline electrolysers [6, 25, 30, 34, 35, 38, 39]; other authors assumed a lower efficiency based on the intermittent nature of wind energy [6, 20] – however, there is a degree of uncertainty with the efficiency value assumed. In practice, the variability of the wind energy input has an adverse effect on the nominal efficiency of the electrolyser as well as its operating life [30, 40]. Therefore, without energy storage to smoothen the erratic profile of the wind energy input and dispense this energy in a more uniform fashion to the electrolyser, hydrogen production has a likelihood of being over-estimated, with costs under-estimated. With this in mind, the wind-hydrogen model developed here addresses this issue, while providing a framework for the optimal sizing of the battery capacity (i.e. the capacity that yields a minimum H_2 cost) for a given electrolyser size.

Another pervasive modeling approach in the existing literature is the characterization of the wind variability, via the use of statistical methods, most notably, a Weibull probability density function - to estimate energy, and consequently, hydrogen production [25, 35, 38, 39, 41-43]. While the efficacy of the Weibull function in resolving the variability of the wind speed is not disputed, a certain degree of error is inherent in the estimation of the probability of occurrence of a given wind speed magnitude. This in turn, hinders the accuracy of the energy production estimates. Moreover, the Weibull function is limited in its ability to accurately resolve bimodal or multimodal wind distributions which arise from unique climatic conditions [44, 45]. The model developed in this paper utilizes real time hourly wind energy generation data, thus its accuracy is not hindered by the limitations of the Weibull function.

Considering the foregoing, the principal objectives/contributions of this paper are as follows:

- The development of an integrated grid-connected wind-H_2 techno-economic model with energy storage, for the production of renewable hydrogen and estimation of costs, in a liberalized electricity market with dynamic prices.
- The development of an energy management algorithm for wind-H_2 plants with energy storage, which is a function of the wind turbine energy yield, hourly wholesale electricity price (pool price), electrolyser and battery performance specifications.
- The development of a techno-economic framework for the determination of the optimal electrolyser size, number of electrolyser units and energy storage capacity, which yields a minimum hydrogen production cost, for wind-hydrogen systems with energy storage.

The model has been developed such that its inputs are not constrained to a particular jurisdiction. For instance, variables that can be readily adjusted to suit various jurisdictional contexts (e.g. peak electricity price hours and the wind energy generation profile) are used in the model. In this paper, Alberta serves as the case study of choice; with the hydrogen produced being 'customized' to service the bitumen upgrading industry. There is a scarcity of integrated wind-hydrogen models which consider the full supply chain of hydrogen from production to delivery, whilst incorporating the modeling features aforementioned. By circumventing the limitations associated with previous modeling approaches, the hydrogen production cost estimates provided by the model in this article are more indicative of 'real' costs. All costs indicated in this article are in 2014 Canadian dollars[3] unless otherwise specified.

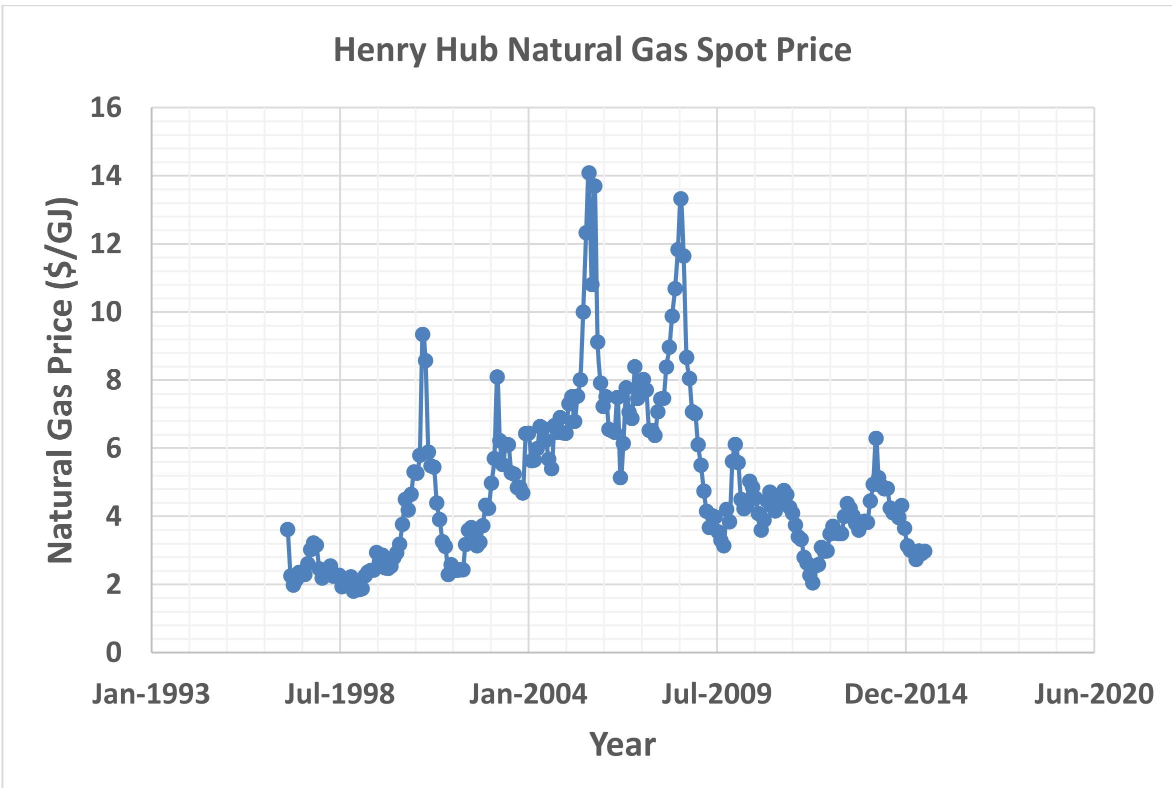

Figure 2.1: Historical natural gas price in Alberta 1997 – 2014 [46].

[3]Where necessary, an inflation rate of 2% has been used to convert all costs into 2014 $CAD. Furthermore, currency rates of $1CAD = $1US; $1.3CAD = €1; $1.6CAD = £1 are adopted in this paper.

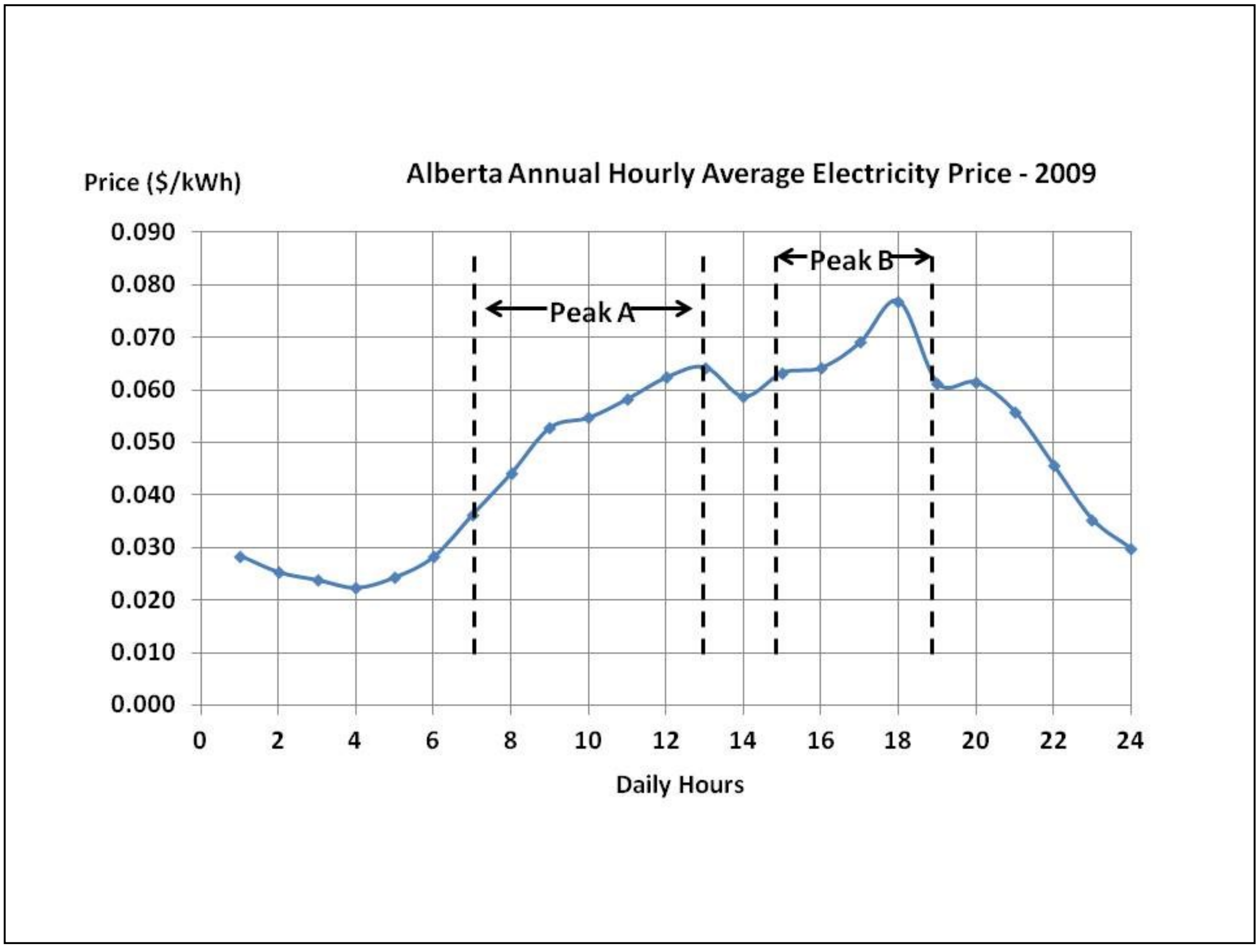

Figure 2.2: Alberta annual hourly average electricity (grid pool) price – 2009 [47]

2.2 Methodology & Scope

2.2.1 Site Selection and Energy Logistics

The wind energy generation potential is underpinned by the resource quality (mainly governed by the wind speed) and availability in a particular jurisdiction. For a given jurisdiction, the ideal site is such that the location of the wind resource and end-use hydrogen demand are coincident – this will increase the cost competitiveness of the plant. In the Alberta context, as shown in Table 2.1, as of 2009, the wind generation capacity of the province amounted to 563 MW [48, 49]. Alberta's wind energy endowment is located in the southern region of the province, as indicated by the wind farm development in this area (see Figure 2.3) [6]. For the plant proposed in this study, the energy from the network of wind farms is channelled via the existing transmission line system to the Summerview 1 wind farm in Pincher Creek; where the electrolyser farm is located for hydrogen production. Pincher Creek serves as the site for the electrolyser farm due to the high density of wind farms in this area relative to other regions in Southern Alberta (see Figure 2.3), as well as for comparative reasons with previous studies [6, 25]. Furthermore, the nature of the energy logistics pertaining to the plant, facilitates an enhanced capacity factor of the electrolyser farm - due to the geographically dispersed nature of the wind farm network on a localised level. This is considered to be a more efficient and pragmatic alternative to the option of having electrolyser farms situated at each wind farm location, where the capacity factor of the electrolysers are inhibited by the fact that they are constrained to the energy yield of a single wind farm as opposed to a broader localised network of wind farms.

Table 2.1: Grid-connected wind farm generation capacity in Alberta as of 2009 [48, 49]

<table>
<tr><th>Wind Farm Name</th><th>Period of Installation to 2009 Year End Capacity</th><th># of Wind Turbines</th><th>Wind Turbine Rated Power (kW)</th><th>Wind Farm Capacity (MW)</th></tr>
<tr><td>Blue Trail Wind</td><td>2009</td><td>22</td><td>3000</td><td>66</td></tr>
<tr><td>Castle River #1</td><td>1997-2001</td><td>59</td><td>660</td><td>40</td></tr>
<tr><td rowspan="2">Cowley Ridge</td><td rowspan="2">1993-2001</td><td>57</td><td>375</td><td rowspan="2">38</td></tr>
<tr><td>15</td><td>1300</td></tr>
<tr><td>Enmax Taber</td><td>2007</td><td>37</td><td>2200</td><td>81</td></tr>
<tr><td>Kettles Hill</td><td>2006-2007</td><td>35</td><td>1800</td><td>63</td></tr>
<tr><td>McBride Lake</td><td>2001-2003</td><td>115</td><td>660</td><td>75</td></tr>
<tr><td>Soderglen Wind</td><td>2006</td><td>47</td><td>1500</td><td>68</td></tr>
<tr><td>Summerview 1</td><td>2002-2004</td><td>38</td><td>1800</td><td>68</td></tr>
<tr><td>Suncor Chin Chute</td><td>2006</td><td>20</td><td>1500</td><td>30</td></tr>
<tr><td>Suncor Magrath</td><td>2004</td><td>20</td><td>1500</td><td>30</td></tr>
<tr><td>Taylor Wind Farm</td><td>2004</td><td>9</td><td>375</td><td>4</td></tr>
<tr><td></td><td></td><td></td><td>TOTAL</td><td>563</td></tr>
</table>

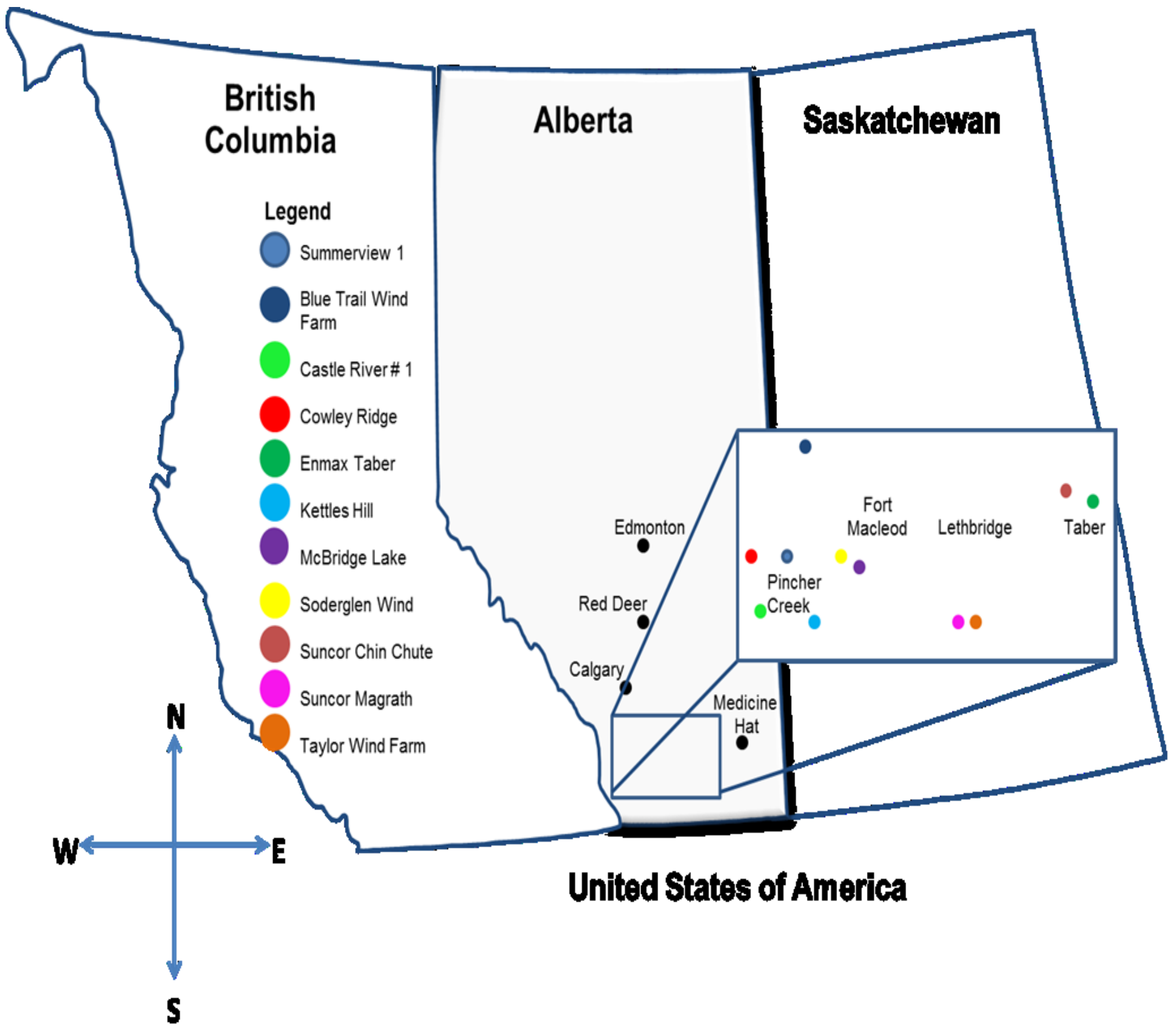

Figure 2.3: Geographical illustration of grid-connected wind farm locations in Alberta (2009) [6]

2.2.2 Wind-Hydrogen Plant Description

The FUNNEL – COST – H2 – WIND (FUNdamental eNgineering principlEs-based modeL for COST estimation of hydrogen (H_2) from WIND) plant model proposed in this paper, has a capacity of 563 MW which corresponds to the installed grid connected capacity in Alberta as of 2009 (see Figure 2.4). The plant has eight unique operating modes which are described in Table 2.2. A control unit is employed in the plant to govern its energy management, using a robust algorithm that determines its operational mode for any given hour of the year (see section 2.2.6 for further details). In addition, a rectifier and a buck converter are used for AC/DC conversion, depending on the operational mode of the plant. This is because, due to the energy from the wind farms being of a high voltage, after the rectifier has converted the energy from AC to DC, the buck converter steps down the high voltage DC to a lower voltage suitable for the battery/electrolyser units. A battery is used to smoothen the energy derived from wind, and feed the electrolyser unit with the energy requisite for hydrogen production. Alternatively, the battery is simply used for energy storage when required. Once hydrogen is produced, a compressor is then used to elevate its pressure so as to facilitate pipeline transportation (it is important to point out that the energy for compression is derived from the electrical grid and thus, constitutes a source of operational GHG emissions; the degree to which is governed by the emissions intensity of the grid). In turn, the pipeline transports the hydrogen produced to the bitumen upgrader for consumption.

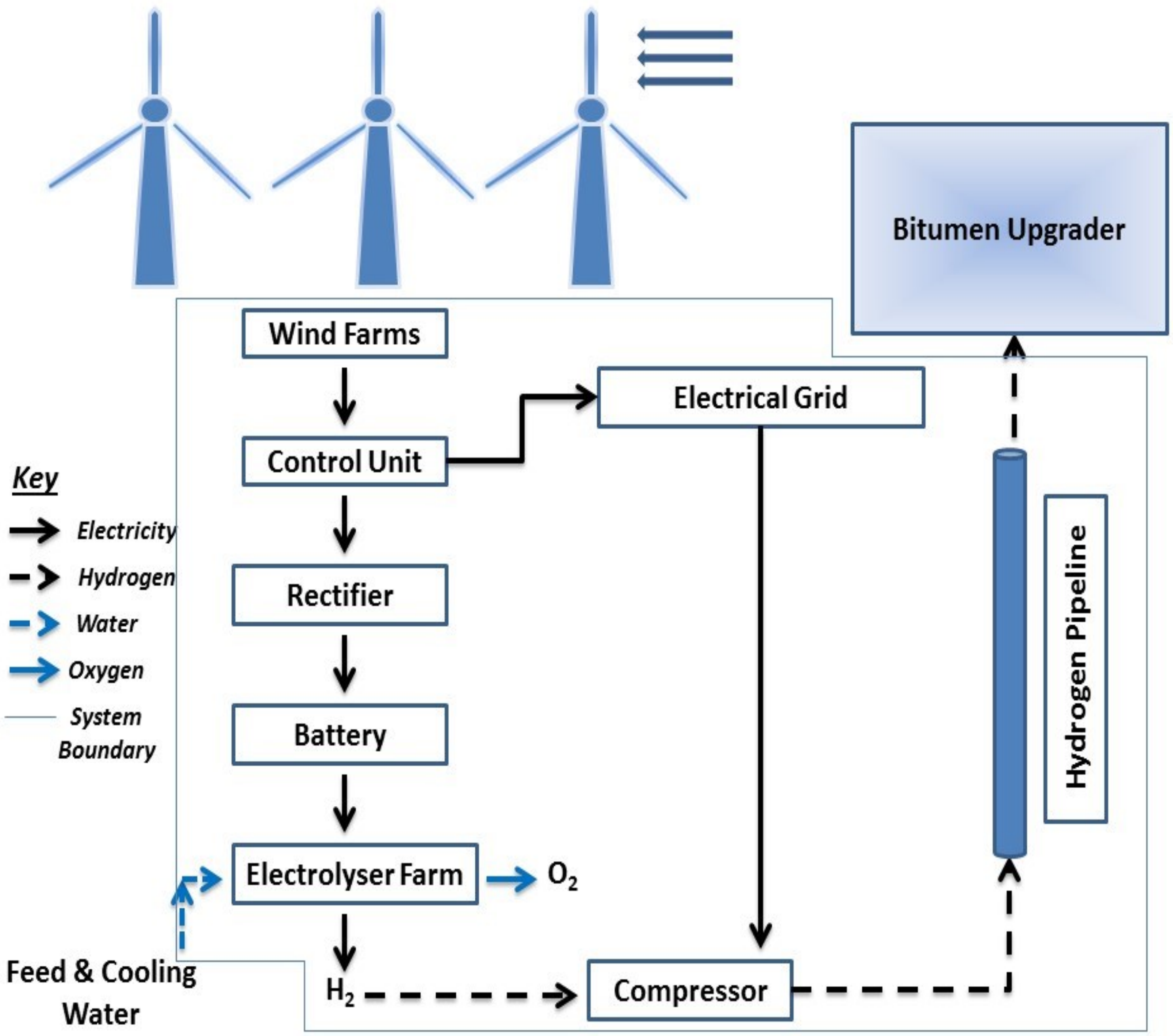

Figure 2.4: Conceptual schematic of the wind-hydrogen plant

Table 2.2: Wind-Hydrogen Plant – Modes of Operation

Operational Mode	Description	Comments
Mode A	Hydrogen production only.	The aggregate[a] amount of energy available to the plant is sufficient for hydrogen production only.
Mode B	Hydrogen production with energy storage.	The excess energy available is stored in the battery due to the unavailability of premium (peak) electricity prices.
Mode C	Hydrogen production with premium electricity sales.	The excess energy available is sold to the grid due to the availability of premium electricity prices.
Mode D	Hydrogen production with non-premium electricity sales.	This occurs when the battery capacity is undersized relative to the surplus energy that needs to be stored. Hydrogen is produced with the use of the grid as a dump-load for any surplus amount of energy. Here, grid sales occur irrespective of the pool price[b].
Mode E	Electricity is sold to the grid only.	This occurs for the following operating conditions: a) When the aggregate amount of energy in the plant does not meet the minimum charging threshold for the battery. b) When the aggregate amount of energy surpasses the minimum threshold for battery charging, however, the energy falls short of the minimum electrolyser energy requirement for H_2 production, and premium electricity prices are available on the grid. c) When the aggregate energy in the plant surpasses the battery capacity and the maximum energy that can be supplied by the

Operational Mode	Description	Comments
		battery does not meet the electrolyser minimum energy demand, along with premium prices being available on the grid.
Mode F	Energy storage only.	This occurs when the aggregate energy surpasses the threshold for battery charging, however, the energy falls short of the minimum electrolyser energy requirement for H_2 production, and premium electricity prices are not available on the grid.
Mode G	Energy storage occurs in addition to the use of the grid as a dump load.	This occurs when the aggregate energy in the plant surpasses the battery capacity and the maximum energy that can be supplied by the battery does not meet the electrolyser minimum energy demand. Here, grid sales occur irrespective of the pool price.
Mode H	Plant lull.	

Notes:

[a] This refers to the sum of the average energy produced by the wind turbine for a particular hour, and the energy contained in the battery for the same hour.

[b] This is the term used to refer to the hourly wholesale electricity price in Alberta's liberalized electricity market

2.2.3 Battery Selection

A summary of the salient characteristics of some battery types are presented below; however, a comprehensive review of batteries along with other energy storage technologies can be found in literature [50-52].

Lead-acid batteries are the most technologically mature and most widely used battery type [34, 51, 52]. The principal merits associated with these batteries are their inexpensive capital cost ($50-310/kWh), and relatively high roundtrip efficiency (75-80%) [51, 52]. However, lead-acid batteries, depending on the depth of charge, have short cycle lives in the range of 200 - 1800 full equivalent cycles [51, 52] – this stems from parasitic reactions such as positive plate corrosion along with the formation of lead sulphate instead of lead oxide (which occurs during normal operation) [53]. This is likely to result in significant replacement/maintenance costs over the wind-hydrogen plant's lifetime. Discharging constraints are associated with this battery type; their state of charge cannot be lower than 40% of their capacity [34]. Consequently, for the application of interest, this battery is prone to underutilization, which inhibits the battery capacity factor, resulting in cost inefficiencies. The battery capacity factor is defined in this study as the ratio of the energy supplied to the battery, and its maximum energy capacity, over an annual period (see section 2, equation 2, in Appendix).

Nickel-cadium batteries are of two types, sealed and unsealed [51, 52]. The sealed type is usually for everyday small-scale applications such as remote devices, lamps etc; hence, this type will not be discussed further. The unsealed type is used for large scale applications where weight and volume are important design constraints, a prime example being aviation [51]. Unlike their lead-acid counterparts, nickel-cadium batteries can be fully discharged, negating the need for a minimum state of charge [34]. That said, they incur considerably higher capital cost ($400-2400/kWh) in comparison to lead-acid batteries and also require periodic venting and water addition during charging, as a result of oxygen and hydrogen formation at the electrodes [51]. Furthermore they have a lower round trip efficiency of 60-72% in comparison to lead-acid batteries [51, 52]. More importantly, they are considered ineffective for peak shaving and energy management applications, and the toxicity associated with cadium is another concern [51].

Sodium-sulphur batteries are only second to lead-acid batteries with regards to cost effectiveness, with capital costs in the range of ($180-500/kWh) [50-52, 54, 55]. They have long cycle lives of up to 20,000 cycles depending on the depth of charge, as well as zero self-discharge [34, 52]. They have no minimum state of charge and can be fully discharged [56]. Furthermore, they have a relatively high efficiency of 75-92% and are considered particularly adept for large scale utility energy storage applications [51, 52, 57, 58]. Apart from this, they are especially suitable for economical energy management applications including: load leveling, power quality, peak shaving as well as renewable energy management and integration [34, 50-52]. The principal drawback of sodium-sulphur batteries is their requirement for high temperature operation (300-350^{0}C for optimal battery performance) and thermal management [34, 50-52]. Notwithstanding, some authors are of the contention that once Na-S batteries are running, the heat produced by charging and discharging cycles is sufficient to maintain operating temperatures and typically no external heat source is required [57, 58].

Taking the characteristics of the different battery types into consideration, sodium-sulphur batteries are adopted for the wind-hydrogen plant proposed. This is mainly due to their suitability for large scale energy storage and energy management applications, their inexpensive capital cost and high flexibility of charging/discharging depth. The superior performance of Na-S batteries is evidenced by their widespread application for large scale wind energy installations across the globe [51]. The specification of the sodium sulphur battery unit utilized in this study is provided in Table 2.3.

Table 2.3: Sodium-sulphur battery specification

Parameter	Value	Sources/Comments
Manufacturer	NGK Insulators ltd.	[59]
Energy rating (MWh)	6	[59]. A minimum energy threshold of 5% rated capacity was adopted in this study. This was done to limit the adverse effect of deep depths of discharge on the battery's operating life.
Power rating (MW)	1	[59]
DC Efficiency (%)	85	[59]
Operating Temperature (^{0}C)	300 -350	[59]. Operating temperature is assumed to be maintained by heat evolution during charging and discharging cycles; no external heat source required [57, 58].
Nominal operating life (yrs)	15	[59]. The nominal life corresponds to 300 charge-discharge cycles per year at full rated energy capacity (4,500 total cycles). As a conservative estimate, a service life of 10 yrs has been assumed in this study.

2.2.4 Electrolyser Selection

The current electrolyser (electrolysis) technologies that dominate the pertinent literature can be sub-divided into three main classes, namely: alkaline electrolysers, proton exchange membrane (PEM) electrolysers, and high temperature electrolysers (HTE) [25]. Relative to other electrolytic options, alkaline electrolysers are adopted in this study as a result of their superior technological maturity, large scale hydrogen flow rates, and relatively inexpensive capital cost [25]. For a more detailed examination of the aforementioned electrolyser pathways, the reader is referred to the work by Olateju & Kumar [25].

2.2.5 Electrolyser Modelling

Previous studies have presented 'element-level' electrolyser models, where the authors have characterized the operating voltage of the electrolyser as a function of its operating temperature, current density and characteristic over-potentials (which are also temperature dependent), via a hybrid of electro-chemical and thermodynamic relations [42, 60, 61]. This modeling paradigm allows for a more robust and dynamic resolution of the electrolyser nominal efficiency, facilitating a more precise simulation of hydrogen yield from a given electrolyser. However, these aforementioned models are predicated upon empirical equations; the coefficients of which have to be ascertained for a particular electrolyser via experiments [60, 61]. As such, the models cannot be readily generalized; limiting their utility in contexts where a broad number of electrolyser models/capacities and number of units need to be evaluated as part of an integrated energy system. Moreover, some authors have compared these element-level models to those where the nominal efficiency of the electrolyser is assumed to be independent of temperature – the difference in hydrogen yields were in the order of 3% [60].

With this in mind, in this paper, a systems-level approach is implemented in the modeling of the performance of the electrolyser, based on its salient characteristics and the energy input from the battery (which ultimately emanates from the wind turbine). This study assumes that the nominal efficiency of the electrolyser does not change materially during its operation, due to the role of the battery, which delivers a power supply with significantly reduced perturbation - in other words, the electrolyser operates at the constant, nominal current density. In essence, the framework adopted in this paper facilitates modeling flexibility and generalization, without compromising on the accuracy of hydrogen yield (vis-à-vis element level models).

A total of six different electrolyser sizes were considered in this study, the performance specifications of each electrolyser are outlined in Table 2.4. It is worth pointing out that the minimum electrolyser power requirement for all electrolysers has been determined based on a proportional relationship between the maximum flow rate and maximum power demand (rated power) of the electrolyser as shown in Eq. (1) [6]. The rationale behind this approach is the fact that the minimum operating threshold for electrolysers varies widely in the literature; ranging from 5-50% of their rated power [62, 63], depending on the scale and manufacturer of the unit. Thus, for reasons of consistency, this methodology has been adopted.

As opposed to the operation of the electrolyser at 73% of its nominal efficiency in previous studies by the authors [6, 25], the nominal efficiency of the electrolyser is assumed to be maintained in this study (see Table 2.4), due to the role of the battery (as explained previously); the sodium sulphur battery charge/discharge efficiency is assumed to be 85%; the rectifier and compressor efficiency have been taken as 95% and 70% respectively.

$$\boldsymbol{EP_{min}} = \frac{(\boldsymbol{EF_{min}} \times \boldsymbol{EP_{max}})}{(\boldsymbol{\eta} \times \boldsymbol{EF_{max}})} \qquad \text{(Eq. 1)}$$

Where: η represents the combined efficiency of the rectifier and battery; EF_{min} and EF_{max} represent the electrolyser maximum and minimum flow rates, respectively. EP_{max} represents the electrolyser rated power.

Table 2.4: Electrolyser size range [64, 65]

Electrolyser manufacturer/model	Min. H_2 flow rate (Nm^3/hr)	Max. H_2 flow rate (Nm^3/hr)	Energy requirement (kWh/Nm^3)	Nominal Efficiency (HHV) (%)[d]	Size (kW)	H_2 pressure (bar)	H_2 purity (%)
Norsk Hydro Atmospheric Type No. 5010 (5150 Amp DC) [64]	0[a]	50	4.8[b]	72.4	240	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5020 (5150 Amp DC) [64]	50	150	4.8[b]	72.4	720	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5030 (5150 Amp DC) [64]	150	300	4.8[b]	72.4	1440	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5040 (4000 Amp DC) [64]	300	377	4.8[b]	72.4	1810	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5040 (5150 Amp DC) [64]	300	485	4.8[b]	72.4	2328	1	99.9 ± 0.1
Industrie Haute Technologie (IHT) Type S-556 [65]	190	760	4.9[b,c]	70.8	3496	30	99.9 ± 0.1

[a]A minimum flow rate of $1Nm^3/hr$ was utilized in this study.

[b]Indicates the hydrogen production systems level energy requirement specified by the manufacturer [64].

[c]Average value of the energy requirement range (4.6-5.2 kWh/Nm^3) indicated.

[d]The nominal efficiency defined here is the ratio of the ideal energy consumption for water electrolysis (39 kWh/kg H_2) to the nominal energy consumption per unit of hydrogen produced for each electrolyser (at its rated power).

2.2.6 Integrated Techno-Economic Model Development

The FUNNEL – COST – H2 – WIND model utilized in this article was developed using MATLAB [66]. An integral part of the model is an energy management algorithm which considers the hourly average energy generated from wind, the hourly price (grid pool price) of electricity and the salient characteristics of the battery and electrolyser units (see Tables 2.3 & 2.4), to determine the operational mode of the plant for each hour in the year. The hourly average energy generated from wind and the hourly grid pool price of electricity, for the year 2009, were provided by the Alberta Electric System Operator AESO [47]. It is important to stress that other components of the plant including the pipeline and compressor, are sized in accordance with the performance of the electrolyser and battery units being evaluated by the model. The model aims to determine the plant configuration (i.e. the electrolyser size, number of electrolyser units and number of batteries) that will yield a minimum hydrogen production cost.

As shown in Figure 2.5, the energy management algorithm (see section 1 of Appendix for algorithm) uses the hourly wind generation data and the economic characteristics of the grid in terms of daily peak and off-peak prices (see Figure 2.2), to ascertain three principal variables of interest: the hourly amount of hydrogen production, electricity sales to the grid, and energy storage, for each plant configuration being evaluated. Each case is run for a duration of 8760 hrs so as to ascertain the corresponding annual values. Once these principal variables have been deduced along with other performance metrics, such as the electrolyser and battery capacity factors, auxilliary plant components are sized accordingly. Additionally, cost data including capital, labour, operating and maintenance costs associated with all plant components are utilized in a discounted cash flow (DCF) model. The DCF model allows for the determination of the hydrogen production cost. This process is repeated for all the plant configurations under consideration; i.e. for all the electrolyser sizes and number of units, as well as the number of battery units, for the determination of the optimal plant set up. The hydrogen production cost ($/kg) is the sole variable used to determine the optimal configuration, and an error margin of $0.01/kg H_2 is incorporated in the algorithm to halt/proceed with iterations after successive values of the production cost have been compared.

2.2.6.1 Determination of Optimal H_2 Cost

To appreciate the determination of the optimal H_2 cost in the plant, the relative sizing of the wind farm capacity and the electrolyser farm must be addressed. For a grid connected wind-hydrogen plant such as the one proposed in this paper, the electrolyser capacity (MW) must be undersized relative to the capacity of the wind farm (MW), for a competitive hydrogen production cost to be realized [28, 40]. This is because the undersized electrolyser operates at a higher capacity factor relative to that of the wind farm (the capacity factor of the wind farm used in the model is 30%). Thus, in this paper, the optimal electrolyser capacity (MW) for the fixed wind farm capacity of 563 MW, is assumed to exist between 1 and 563 MW. The model developed 'surveys' this solution space using 6 different electrolyser sizes (see Table 2.4) and number of units. For a particular electrolyser size, the number of units is varied incrementally in the model, using interval sizes, to traverse the lower and upper limits of the 'global' solution space (i.e. an electrolyser farm capacity of 1 to 563 MW). As illustrated in Figure 2.5, using an iterative process, the initial optimal electrolyser size is used to update the limits of the solution space, making it more localized to the region of the initial optimum. The interval size is also updated accordingly. This process continues until the difference between successive hydrogen production costs is less or equal to $0.01/kg H_2.

To the knowledge of the authors, there is no established paradigm for the optimal sizing of a battery (energy storage) relative to an electrolyser unit, in the context of wind-hydrogen systems. In this paper, the optimal sizing of the battery is carried out in tandem with the electrolyser. As in the case of the electrolyser, the number of battery units is varied incrementally using an interval size within a defined range (solution space). The power rating of the battery (each battery unit has a maximum power rating of 1 MW and energy capacity of 6 MWh – see Table 2.3) was used to define the solution space, and varied from 1 MW to 563 MW to conform to the electrolyser solution space. However, based on the hydrogen production costs yielded, the maximum of this range was adjusted to 100 MW/600 MWh (for the electrolyser farm sizes considered, H_2 costs were observed to escalate significantly beyond a battery capacity 100 MW/600MWh). As in the case of the electrolyser, the initial optimal battery size is used to 'localize' the solution space and update the interval size. This process continues until the difference between successive hydrogen production costs is less or equal to \$0.01/kg H_2. Section 2.4.2 elaborates upon the driving factors that govern the size of the battery relative to an electrolyser.

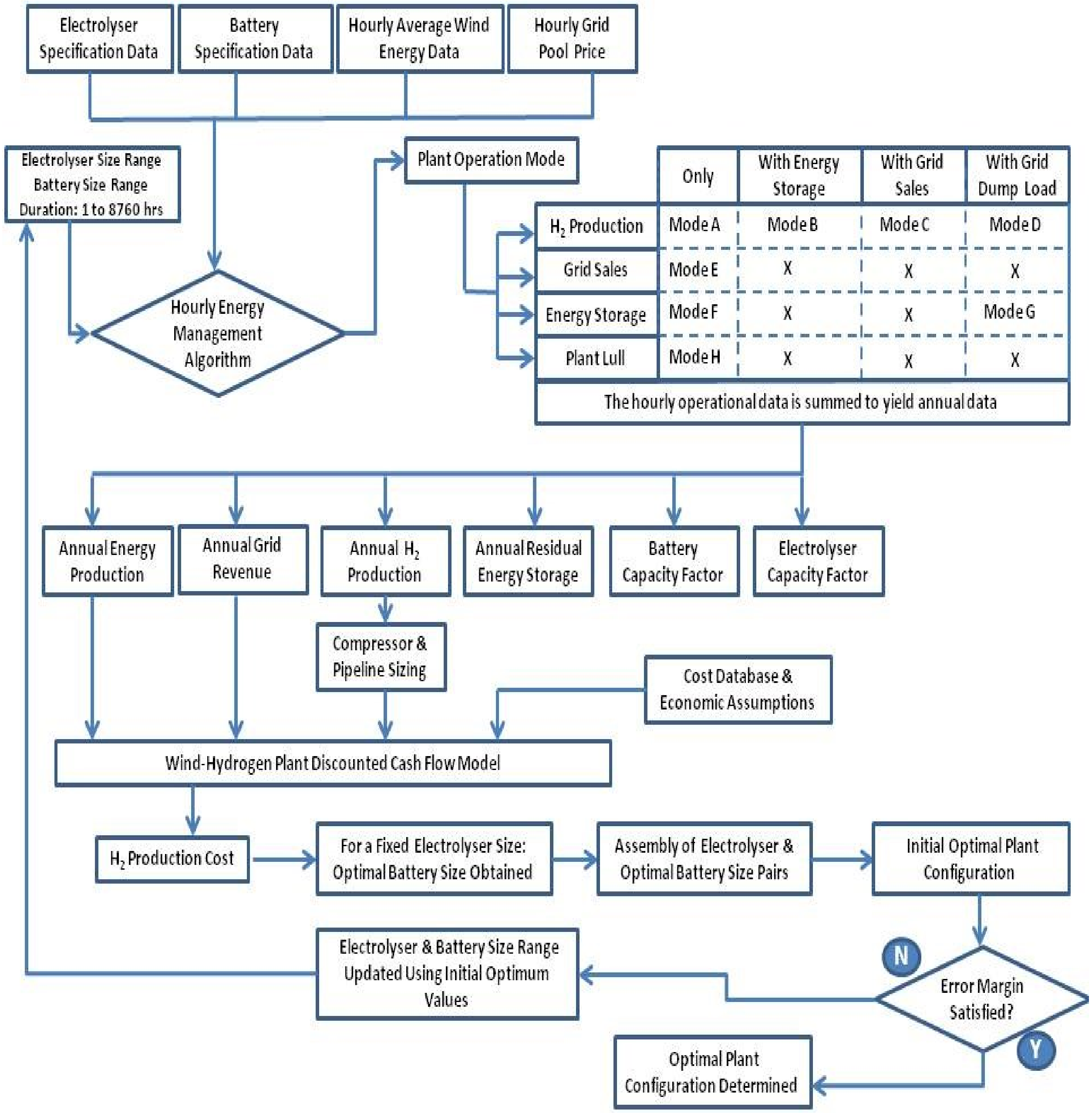

Figure 2.5: Techno-economic modeling framework

2.3 Cost Estimation

2.3.1 Electrolyser Capital Cost

An electrolyser capital cost model developed by Olateju & Kumar [25] (see Figure 2.6), which is based on data obtained from pertinent literature and industrial experts, was utilized in the techno-economic model. To put the electrolyser capital cost estimates into context, the US DOE had 2011 electrolyser capital costs estimates of $430/kW (2007 US Dollars) for its 'forecourt' production scale (62.5 kg/hr or 702 Nm^3/hr) [67]. For this same scale, the capital cost model developed here, yields an estimate of $445/kW (2010 Canadian Dollars) – making both model estimates, reasonably comparable. Furthermore, it is worth pointing out that although the current study involves the capital cost estimation of a significant number of electrolyser units with varying sizes, which will likely facilitate volume discounts along with cost/labour efficiencies, this is not factored into the model. A conservative approach is adopted in this paper, where none of the aforementioned efficiencies are realized. It is worth mentioning that the electrolyser capital cost model is specific to alkaline electrolysers and indicative of the state of the technology as of the early 2000s, not the state of the art. This is as a result of the limited availability of data. Specific capital cost data is considered proprietary by a number of electrolyser manufacturers. Nonetheless, the estimates provided by the model are within reason.

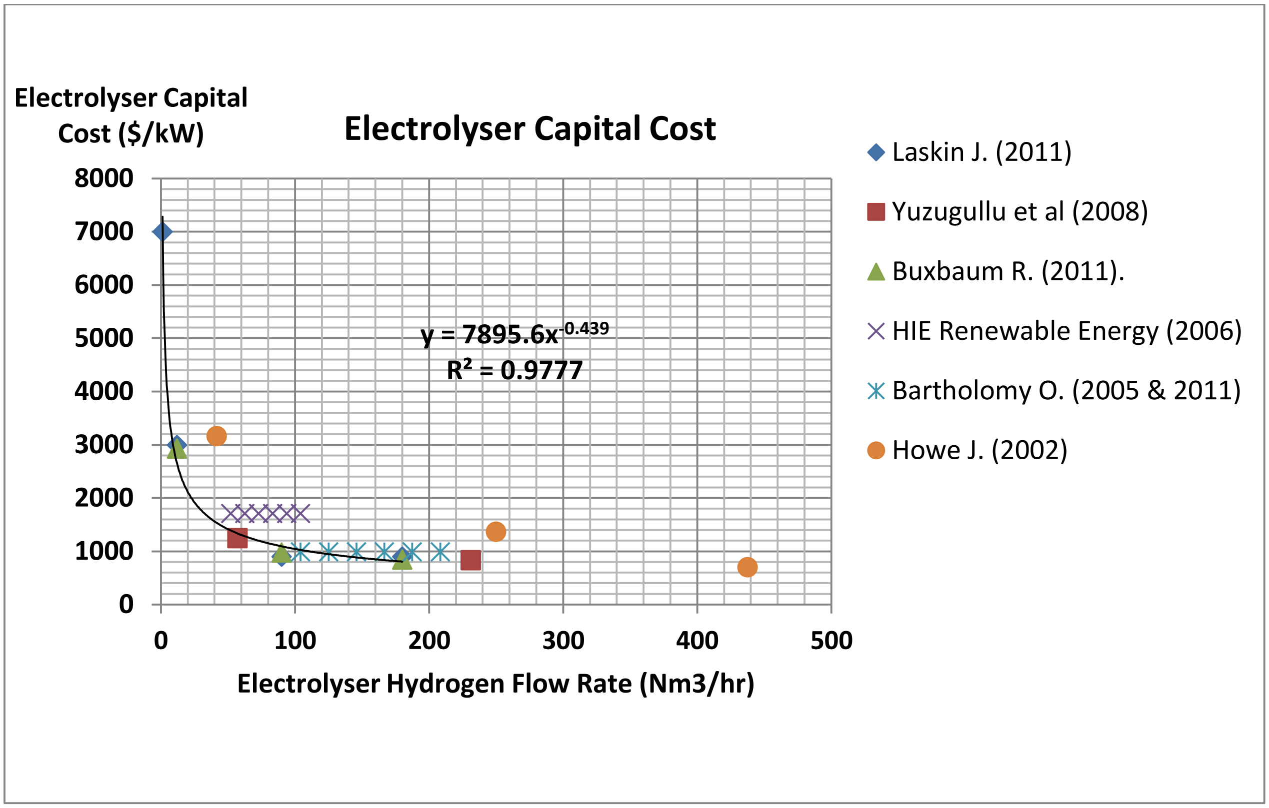

Figure 2.6: Electrolyser capital cost model [25]

Note: The model is based on data gathered from Laskin J. (2011) [68]; Yuzugullu et al. (2008) [69]; Buxbaum R. (2011) [70]; HIE Renewable Energy (2006) [71]; Bartholomy O. (2005 & 2011) [72, 73]; Howe J. (2002)[74]

2.3.2 Wind Turbine Capital Cost and Auxiliary Unit Cost

The wind turbine capital cost is afforded particular examination in this article, due to the capital intensive nature of wind power and its significant impact on the cost of electricity produced – which fuels electrolytic hydrogen production. The wind turbine itself accounts for about 64-84% of the installed capital costs incurred [75-79]. However, the wind turbine capital cost values provided in literature lack consensus and vary widely across jurisdictions and temporal standpoints; thus introducing a degree of uncertainty in the estimation of wind-hydrogen production costs. For instance, installed capital costs in the United States ranged between $1400 - $2900/kW as of 2011; the corresponding range for developed economies is between $1700 – $2150/kW; while for China this value is estimated to be $1300/kW [78, 79]. Furthermore, in the period spanning 2001-2004, the average installed capital cost in the United States was $1300/kW [79]. Thus, it becomes evident that capital cost estimates need to be specific with respect to the market year and jurisdiction they pertain to. With this in mind, an elucidation of the influential factors that govern the turbine's capital cost is duly warranted. In this light, the discussion given here is intended to provide context and insight into the underlying determining factors – thereby providing useful caveats for the capital cost estimate utilized in the model.

From the 1980s to the early 2000s, wind turbine capital costs experienced a dramatic decline; costs fell by more than 65% in the United States, with Denmark experiencing a 55% decrease [77]. However, in the United States, by the mid-2000s, installed capital costs had risen to about $2000/kW [79]. A number of drivers were behind the evolving nature of wind turbine capital costs over time and space.

Firstly, in the early 2000s, the demand and supply dynamics in the global wind energy market was in a state of excess supply, with the demand forecasts by the industry being over-exaggerated [75].

However, in the period between 2005-2008, demand for wind turbines grew considerably, by over 30% on an annual basis, creating a situation of excess demand in the market which translated into elevated prices [75]. In concert with the shift from excess supply to excess demand, raw material prices, most notably, steel and copper (see Figure 2.7), rose significantly over the same period – exerting additional upward pressure on prices [75, 76, 79]. Other raw material price increases included: lead (367%); aluminum (67%); and acrylonitrile (a raw material for the production of carbon fiber) (48%) [76]. Furthermore, the increased scale, complexity and sophistication of turbine design and their resulting components contributed to price increases [77]. That being said, there was one notable exception to this trend of commodity price volatility leading to considerable increases in the installed capital costs for wind energy projects. In the mid-2000s, China remained relatively insulated from these aforementioned market trends with wind turbine capital costs maintaining a level in the range of $1,100–$1,500/kW; owing to the development of a formidable original equipment manufacturing (OEM) base and the availability of labor at a relatively low cost [77].

In more recent times, it is important to highlight the fact that a reversing trend in wind turbine capital cost has been occurring since 2009-2010 [77, 80]. Increased competition between turbine manufacturers, increased manufacturing capacity and lower commodity prices have contributed to this downward trend [80]. In Denmark, capital costs decreased by 22% between 2009 and 2010, with an 8% reduction being observed for Europe as a single entity from 2007 to 2010 [77].

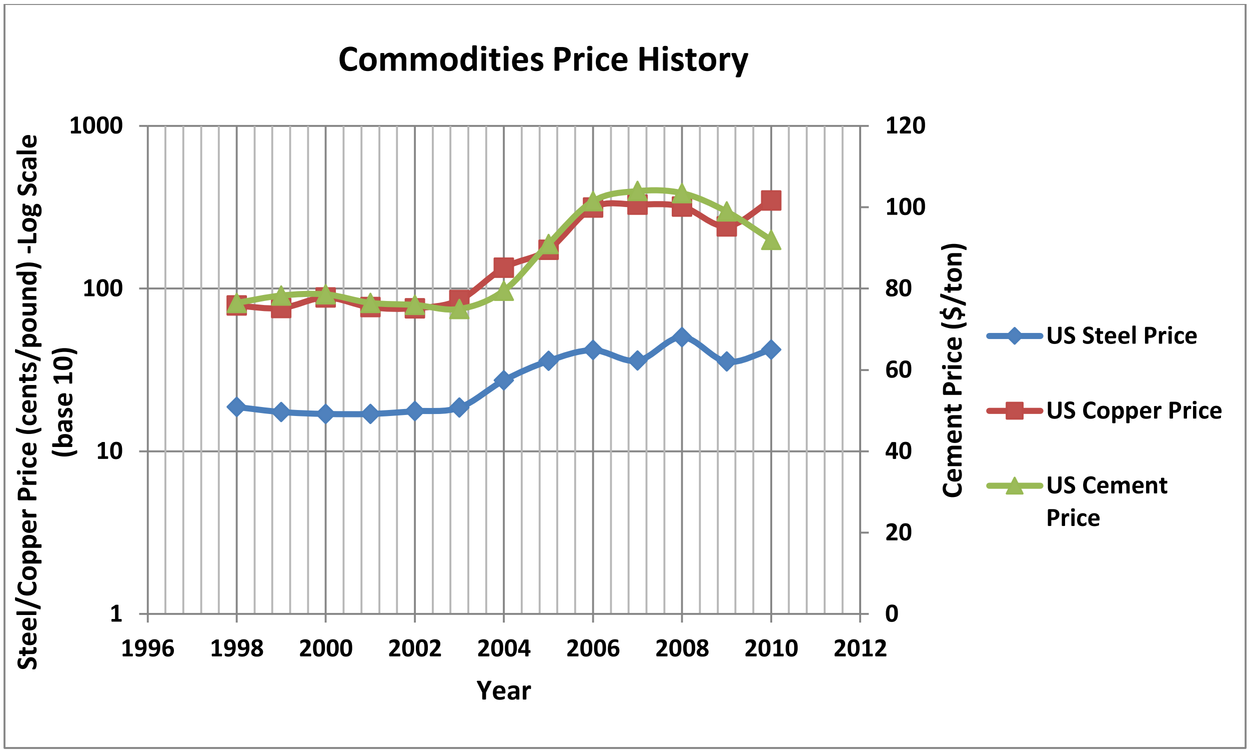

Figure 2.7: Wind turbine commodities price history (United States) – Steel, copper and cement [81]

2.3.3 Wind Turbine Capital Cost – Economies of Scale

In the existing literature, wind-hydrogen models seldom consider the economies of scale that pertain to wind turbines in an explicit fashion. The capital costs utilized are often generic and not specific to a particular wind turbine size [6, 15, 28, 39, 40]. The resolution of the economies of scale and its utilization in wind-hydrogen models will translate into improved (realistic) hydrogen cost estimates. A wind turbine (installed) capital cost model was developed for this research, and is illustrated in Figure 2.8. Figure 2.8 illustrates that smaller wind turbines have a higher specific capital cost ($/kW); the specific capital cost decreases considerably for larger turbine sizes. For a wind-turbine size greater than 1.5 MW, the economies of scale become relatively miniscule and the cost begins to increase gradually for units in the region of 3 MW or higher. It is worth mentioning that units greater than 3 MW are likely to be used offshore; this paper focuses on onshore wind turbines. In order to address some of the specificities of the Albertan economic context, such as higher transportation and labor costs, capital and labor costs are increased by a factor of 1.25. This value is lower in comparison to the magnitude utilized by the authors for a fossil fuel based hydrogen plant [24]. This is to reflect the reduced construction lead time, which can be in the order of months for wind farms as opposed to years for fossil fuel hydrogen plants. Furthermore, the construction of the wind-hydrogen plant is expected to be less labor intensive in comparison to the fossil fuel plant. For comparative purposes, the results yielded by the model were compared to the estimates provided by the United States National Renewable Energy Laboratory (NREL) [82]; the capital cost estimate for a 1.91MW wind turbine was 19% higher than the NREL estimate which, broadly speaking, is indicative of the elevated capital costs in Canada relative to the United States.

Due to the scale of the wind turbine units being considered in the model and the significant impact they have on capital cost expenditure, volume discounts are also taken into consideration. The volume discount model adopted in this article (see Eq. 2) is based on the work carried out by Mosetti et al. [83], with adjustments made to suit the magnitude of units being considered . The maximum volume discount achievable in the model amounts to one-third. For the plant size of 563 MW, six different wind turbine sizes were considered, with the developed capital costs for each turbine size given in Table 2.5. The 2.5 MW turbine had the lowest specific installed capital cost, hence it was utilized as the hypothetical turbine for the plant. It is important to stress that the selection of the 2.5 MW turbine is a real selection in economic terms but hypothetical in energy terms. This is because the real time energy generation data (along with the capacity factor) for the year 2009 is the energy input utilized in the model - this energy is generated from various wind turbine sizes as illustrated in Table 2.1. The rationale behind the consideration of different turbine sizes is to ascertain a minimum capital cost investment for the wind turbines used in the plant. With regards to the energy generated by the 2.5 MW turbine units, the assumption is that they will yield the same aggregate capacity factor of 30%, as is the case with the real time energy data.

$$\boldsymbol{Volume\ Discount} = \boldsymbol{N} \times \left(\frac{\mathbf{2}}{\mathbf{3}} + \frac{\mathbf{1}}{\mathbf{3}}\boldsymbol{e}^{-\mathbf{0.00001}\times \boldsymbol{N}^{\mathbf{2}}}\right) \quad \text{(Eq. 2)}$$

Where: N is the number of wind turbine units.

With regards to auxiliary plant costs, Table 2.6 also provides the O&M costs, as well as costs and service lives pertaining to auxiliary plant equipment and power electronics. Focusing on auxilliary plant costs, it is important to highlight the fact that the cost of purification of the feed water (via reverse osmosis) for the electrolysers, is miniature compared to the cooling water cost [25]. As a result, the latter has been assumed to account for the cost of purification.

Table 2.5: Wind turbine capital cost

Wind Farm Size (MW)	Wind Turbine (kW)	Capital Cost ($/kW)	Number of Units	Volume Discount	Total Installed Capital Cost ($M)	Total Installed Capital Cost ($/kW)
563	500	3864	1126	0.67	1,813	3220
	1000	2749	563	0.68	1,317	2339
	1500	2066	376	0.75	1,089	1934
	2000	1701	282	0.82	980	1741
	2500	1558	226	0.87	954	1694
	3000	1553	188	0.90	986	1752

Table 2.6: Wind turbine auxilliary plant costs

Cost components	Values	Sources/Comments
Wind farm network connection ($)	12.5 % of total wind turbine installed capital cost	[45]
Wind farm electrical infrastructure ($)	9.1 % of total wind turbine installed capital cost	[45]
Project development and management cost ($)	12.5 % of total wind turbine installed capital cost	[45]
Plant power electronics cost ($/kW) (including rectifier and control unit cost)	35	Estimated relative to the cost specified for a 1GW wind-hydrogen plant [15].
Electrolyser labour and installation costs ($)	Function of electrolyser size.	10 % of electrolyser capital cost.
Electrolyser O&M cost ($/kW/yr)	17	[73]
Electrolyser cell stack replacement cost	Function of electrolyser size.	30% of electrolyser capital cost [28].
Battery capital cost ($/kWh)	440	Capital cost is on the higher end of the capital cost range specified in literature ($180-500/kWh) [50-52, 54, 55].
Battery labour and installation costs ($)	Function of battery size	10 % of battery capital cost.
Battery O&M cost ($/kW)	14	[84]

Cost components	**Values**	**Sources/Comments**
Battery module replacement cost	Function of battery size	30% of battery capital cost
Wind turbine O&M cost (years 1-6) ($/yr)	3% of total installed capital cost	Estimated relative to values utilized in [6]
Wind turbine O&M cost (years 7-12) ($/yr)	5% of total installed capital cost	Estimated relative to values utilized in [6]
Wind turbine O&M cost (years 13-20) ($/yr)	8% of total installed capital cost	Estimated relative to values utilized in [6]
Pincher creek water cost ($/m3)	0.99	[25]
Wind turbine service life (yrs)	20	[28, 73]
Electrolyser service life (yrs)	10	[28, 85]
Inverter service life (yrs)	10	[86]
Control unit service life (yrs)	10	[25]

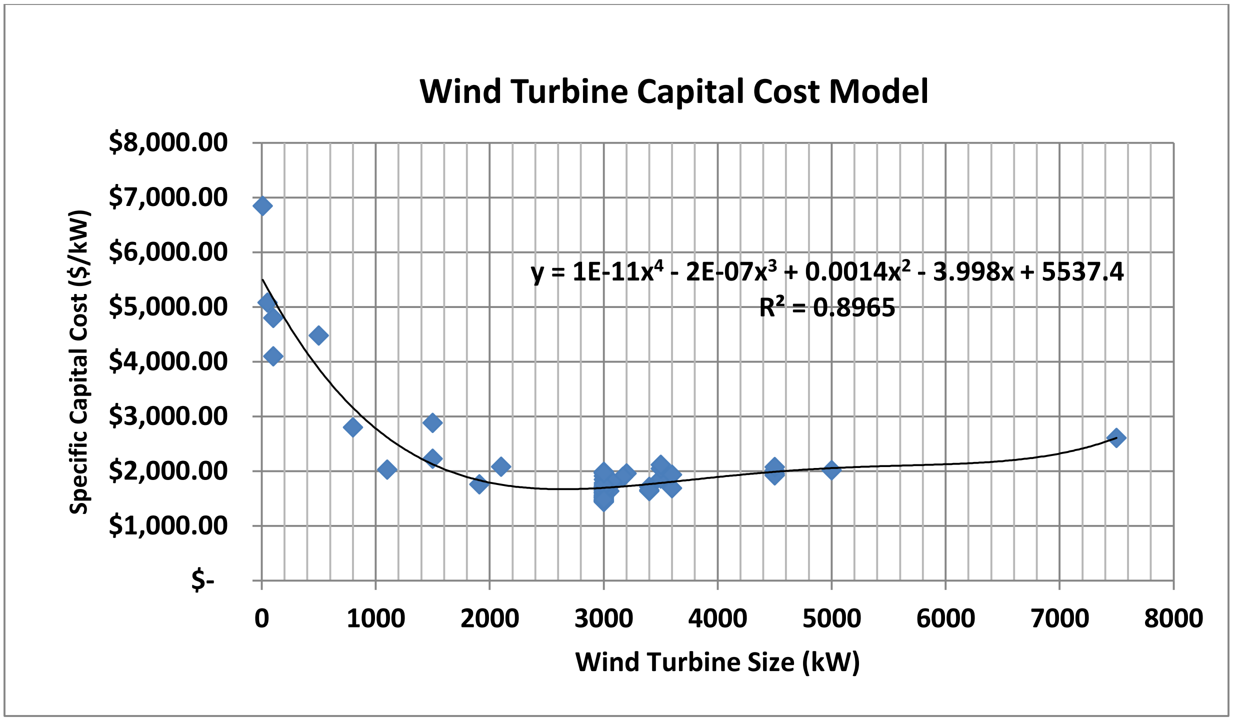

Figure 2.8: Wind turbine capital cost model

Notes: A total of 63 data points were utilized in the model, with data sourced from literature [78, 82, 87-91] .

2.3.4 Hydrogen Pipeline Costs

2.3.4.1 Hydrogen Pipeline Characterization

For a hydrogen production flow rate that surpasses 2,400 kg/day, pipeline transport of the hydrogen produced is regarded as the most cost efficient means of delivery to market [18, 92]. Taking into account the large scale flow rate of the plant, for each plant configuration evaluated in the model, the characterization of the appropriate pipeline dimensions are warranted. The characterization of the hydrogen pipeline required the determination of two principal pipeline parameters, i.e., the pipeline diameter and pipeline length. The diameter of the pipeline was ascertained with the use of the Panhandle – B equation [93]. In this regard, a reverse engineering approach was used to ascertain the diameter requisite for a given hydrogen flow rate. On the other hand, the pipeline distance from the electrolyser farm in Pincher Creek to the bitumen upgrader in the industrial complex in Edmonton, was estimated to be 450 km [25]. This estimate stems from the driving distance between these two locations; however, depending on the logistics of demand, the distance of hydrogen delivery can vary considerably.

2.3.4.2 Hydrogen Pipeline Capital Cost

A pipeline capital cost model developed by previous authors is adopted in this paper [92]. In addition, the capital cost estimate yielded by this model has been benchmarked against two other similar models provided by Johnson & Ogden [94] as well as Parker [95]. The difference in the resulting estimates ranged from 10-18%, which is considered to be a satisfactory range of consensus for the intended purpose. While the model provides a generic cost estimate, the technical, economic and social specificities of a particular hydrogen pipeline project, along with the quality of its construction execution and management, will go a long way in determining the costs realized. Hydrogen pipelines in general have an increased degree of operational risk in comparison to more conventional pipelines (e.g. natural gas), due to the tendency for leakage and embrittlement of steel. Characterizing the economic implication of these added risks will facilitate more robust capital cost estimates. Figure 2.9 provides capital cost estimates for the hydrogen pipeline (in 2010 dollars).

2.3.4.3 Hydrogen Compressor Cost

Typically, the desired pressure at which hydrogen should be delivered to the bitumen upgrader is 50 bar [18]. Consequently, in the model developed, hydrogen exits the compressor at 60bar (inlet pressure of pipeline) so as to be conducive for pipeline transport. For each plant configuration under consideration, a compressor is sized to suit the hydrogen flow rate, using a model provided in an earlier study [33]. In addition, it is important to mention that the hydrogen output pressure from the electrolyser has a significant effect on the cost of the compressor, as this determines the pressure ratio which the compressor will be subjected to. A two stage compressor with an efficiency of 70% and a specific capital cost of $970/kW is utilized in the techno-economic model [33].

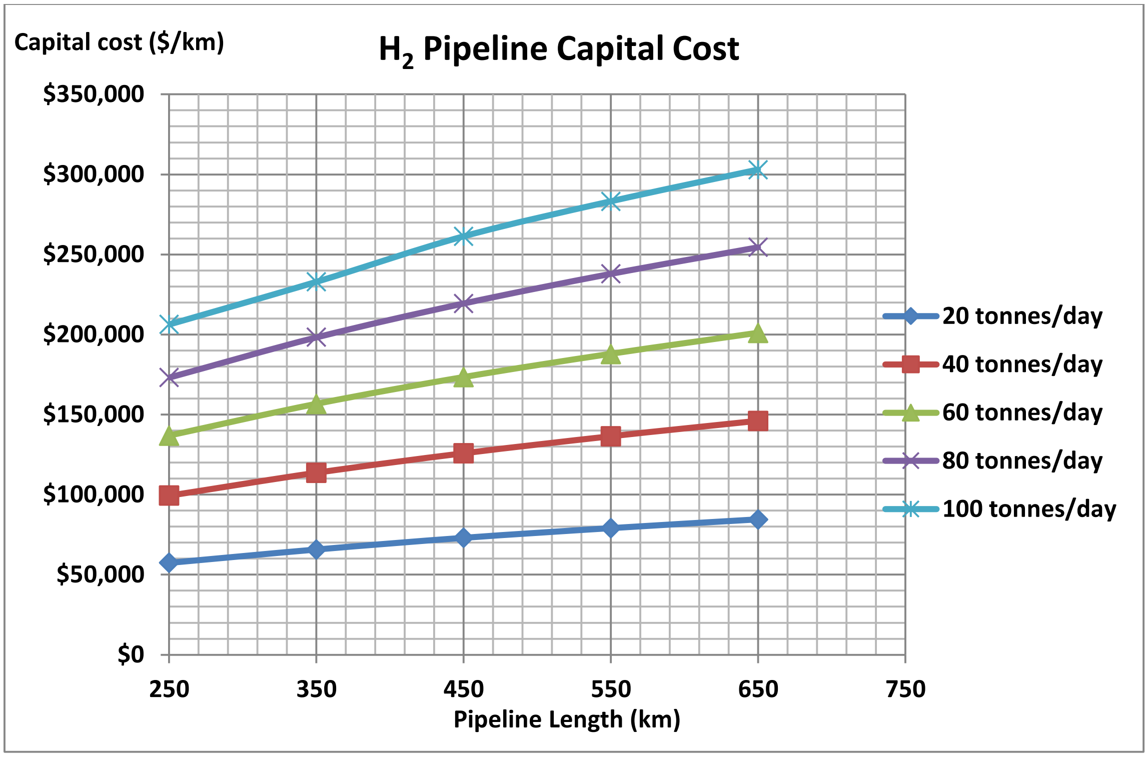

Figure 2.9: Hydrogen pipeline capital cost [6]

Notes: Cost shown are in 2010 Canadian dollars.

2.3.5 Principal Economic Data and Assumptions

In the model developed, a return on equity/internal rate of return (IRR) of 10% along with an inflation rate of 2% was adopted. The wind-hydrogen plant investment is assumed to be serviced by 100% equity; with an operating life of 20 years and a decommissioning cost with a negligible present value [25]. Furthermore, the duration of plant construction is considered to be one year. Another assumption is that the plant does not benefit from any renewable energy incentives such as feed-in-tariffs (FIT). In addition, oxygen, which is a co-product of the electrolysis process, is also considered as a revenue generation stream. It is important to stress that the price for oxygen varies substantially depending on the market in which it is sold, the scale of production and its level of quality (purity). Price quotes varied from $66.57/Nm^3 for medical grade (99.99% purity) oxygen from retail level vendors [96], to $0.078/Nm^3 for large industrial scale producers [97] . Furthermore, in the published literature a price of $2.77/Nm^3 (originally from Praxair Inc.) is cited by Becalli et al. (2013) [27], however the specific market in which oxygen is sold is not apparent.

The wind-hydrogen plant produces oxygen with a purity level that exceeds 99.99%; hence it is sufficient for medical grade applications in Canada, as evidenced by the specifications provided by Praxair Inc. [98]. In addition, medical grade oxygen trades at a significant premium to industrial application oxygen, which can aid the competitiveness of the plant. The demand for the high purity oxygen at the plant is assumed to come from hospitals, which purchase medical grade oxygen at the plant gate. With this in mind, based on the price quotes aforementioned, medical grade oxygen is assumed to trade at a price that has at least a 30% premium over the 'generic' oxygen price $2.77/Nm^3 provided in the existing published literature [27] – i.e. $3.60/Nm^3 . Having said that, other costs such as compression, storage, licensing, and handling, are likely to be significant for the sale of medical grade oxygen at the plant gate; hence, a profit margin of 20% is assumed i.e. $0.72/Nm^3 or $0.50/kg.

2.4 Results and Discussion

2.4.1 Hydrogen Production Cost

2.4.1.1 Electrolyser Farm Size

The minimum hydrogen production cost achieved for all the electrolyser farm[4] configurations evaluated, along with their corresponding optimal battery size, is illustrated in Figure 2.10. For the six different electrolyser sizes considered, the hydrogen production cost curve exhibits a similar non-linear variation. Initially, significant economies of scale are achieved as the hydrogen production flow rate is increased; however, the economies of scale progressively erode as the magnitude of the flow rate is amplified further. For a given electrolyser, in the vicinity of its maximum hydrogen flow rate, the minimum H_2 production cost is achieved; after this minimum cost, increments to the electrolyser farm size results in production cost increases. At a particular hydrogen flow rate, upon further increases in the number of electrolyser units, a corresponding increase in the hydrogen flow rate does not occur. The hydrogen flow rate remains constant – resulting in a significant rise in the H_2 production cost. These aforementioned trends can be explained as follows:

The economies of scale which are realized initially are attributed to the fact that the wind farm investment cost is fixed for all electrolyser farm configurations. Hence, as the hydrogen production flow rate is augmented with an increase in the electrolyser farm size, the unit cost of hydrogen production decreases accordingly. That being said, a point is reached where the incremental electrolyser doesn't yield additional hydrogen productivity. This is because at this point, the electrolyser farm is oversized relative to the wind farm energy yield. It is also worth highlighting

[4] An electrolyser farm consists of a fixed electrolyser size and a number of units.

that for a particular hydrogen flow rate to be produced in the plant, each electrolyser size requires significantly different numbers of units. Consequently, the cost to produce a certain hydrogen flow rate varies widely amongst the electrolyser sizes considered. The overall minimum hydrogen production cost of the plant is \$9.00/kg H_2, which corresponds to 81 units of the 3496 kW (760 Nm^3/hr) electrolyser and 360 MWh (60 units) of battery capacity. In comparison to SMR, this cost is uncompetitive. Olateju & Kumar [24] provide SMR hydrogen production costs for a number of scenarios in Alberta (with and without carbon capture and sequestration); costs range from \$1.87[5]/kg H_2 to \$2.60/kg H_2 (2014 dollars).

[5] The SMR production costs cited are based on an average natural gas price of \$5/GJ over the plant's 25 year lifetime [24].

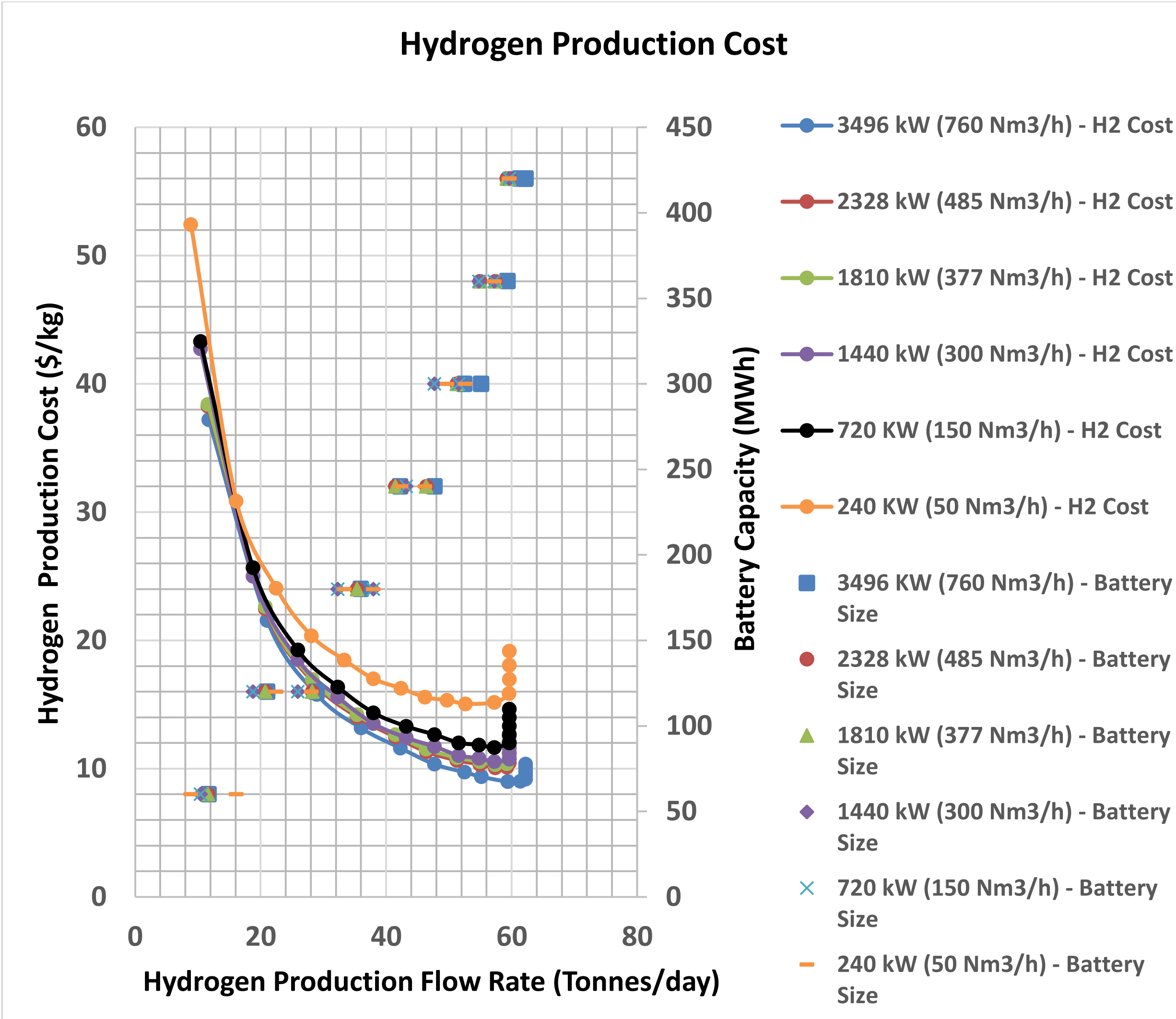

Figure 2.10: Hydrogen production cost

2.4.1.2 Battery Size

The optimal battery capacity for each electrolyser farm size assessed in the model is also shown in Figure 2.10. For a particular range of hydrogen flow rates, the optimal battery capacity for the different electrolyser farm sizes are coincident. With a further increase in the hydrogen flow rate beyond a given range, an increase in the optimal battery size occurs. In general, the optimal battery size increases as the hydrogen flow rate is increased. These observations are quite intuitive. A given battery size is sufficient to produce hydrogen at minimum cost for a specific range of flow rates. Irrespective of the size of the electrolyser or the number of units involved, as long as the hydrogen flow rate falls within range, it can serve as the optimal size. Accordingly, the optimal battery size for the various electrolyser sizes are identical within a particular flow rate range. The general trend of the optimal battery size increasing as the hydrogen flow rate rises is due to the fact that the battery supplies the electrolyser with energy and thus, as increased productivity (H_2 flow rate) is demanded from the electrolyser, the battery has to increase its capacity to deliver energy. Otherwise, the undersized battery will result in a low electrolyser capacity factor and the increased use of the grid as a dump load (i.e. electricity sales to the grid irrespective of the availability of premium electricity prices) – this hinders the cost competitiveness of H_2 production.

2.4.1.3 Hydrogen Production Cost Distribution

The contribution of the different plant components toward the minimum hydrogen production cost ascertained for each electrolyser size, is provided in Figure 2.11. The wind turbine capital cost accounts for the largest portion of the hydrogen production cost for all electrolyser sizes. For the minimum hydrogen production cost determined ($9.00//kg H_2), the wind farm accounts for 63% of this cost. Hence, if existing wind farm assets are used, such that the investment cost of building the wind-hydrogen plant does not include the wind farm costs, the hydrogen production cost is reduced to $3.37/kg H_2. For smaller electrolyser sizes, the electrolyser capital cost accounts for a relatively higher portion of the total cost in comparison to larger electrolysers. This is because smaller electrolysers require a significantly greater number of units and their specific capital costs are also higher. The cost contribution of the battery does not vary significantly for the different electrolyser sizes considered. This is also true for the pipeline and compression costs; however, the cost of compression for the largest electrolyser is relatively minute due to the elevated pressure at which hydrogen is produced (see Table 2.4). Lastly, the contribution of the power electronics cost is relatively insignificant.

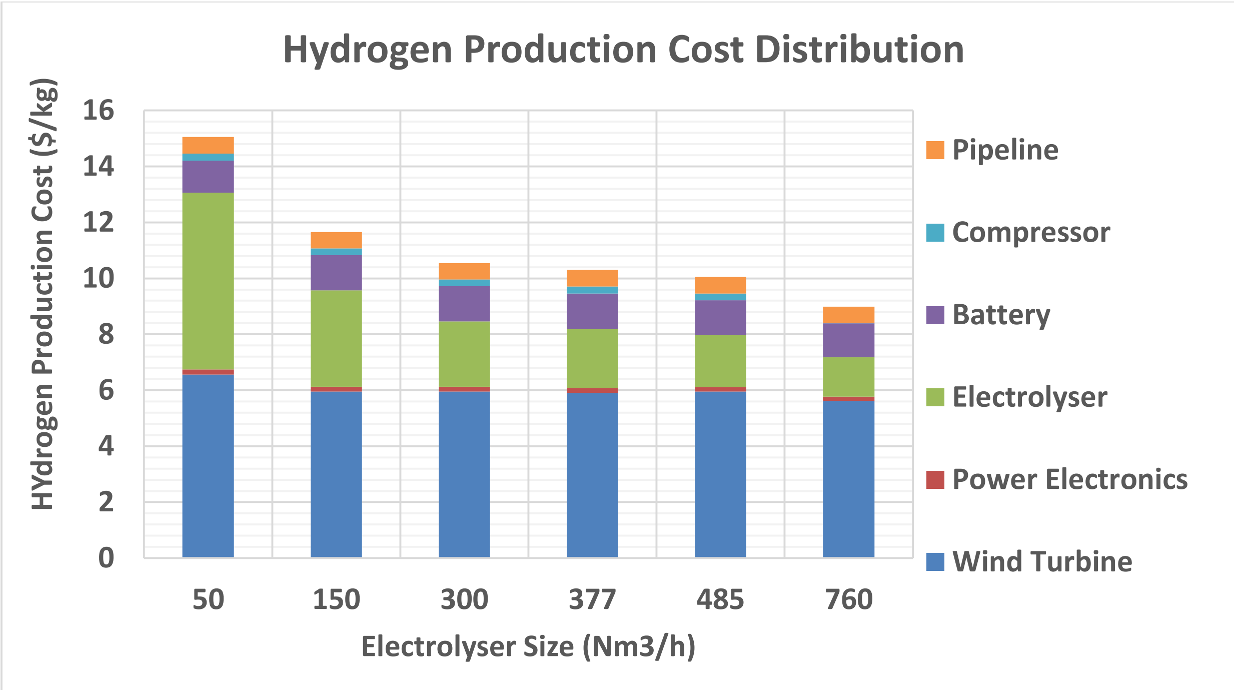

Figure 2.11: Hydrogen production cost distribution

2.4.1.4 Impact of Wind Farm Size

The installed capacity of wind power in Alberta has changed significantly over the past few years, from a capacity of 563 MW in 2009 to over 1 GW as of 2014. Thus, the effect of the wind farm capacity utilised in the model on the hydrogen cost is worthy of examination. To achieve this, a range of wind farm sizes were considered as shown in Figure 2.12. The hourly wind energy generated for the different sizes was assumed to have an identical profile (hourly capacity factor) as the data corresponding to the 563 MW base case. Additionally, the hourly grid pool price was kept constant for all sizes evaluated and a wind turbine size of 2.5 MW was utilized. As shown in Figure 2.12, while significant economies of scale are realised for smaller wind farm sizes, for wind farms greater than 900 MW, the economies are relatively small. This finding is analogous to the observation made by previous authors [79, 80], where the impact of wind farm scale on the cost of electricity is significant for smaller wind farm sizes, but is insignificant for larger scales.

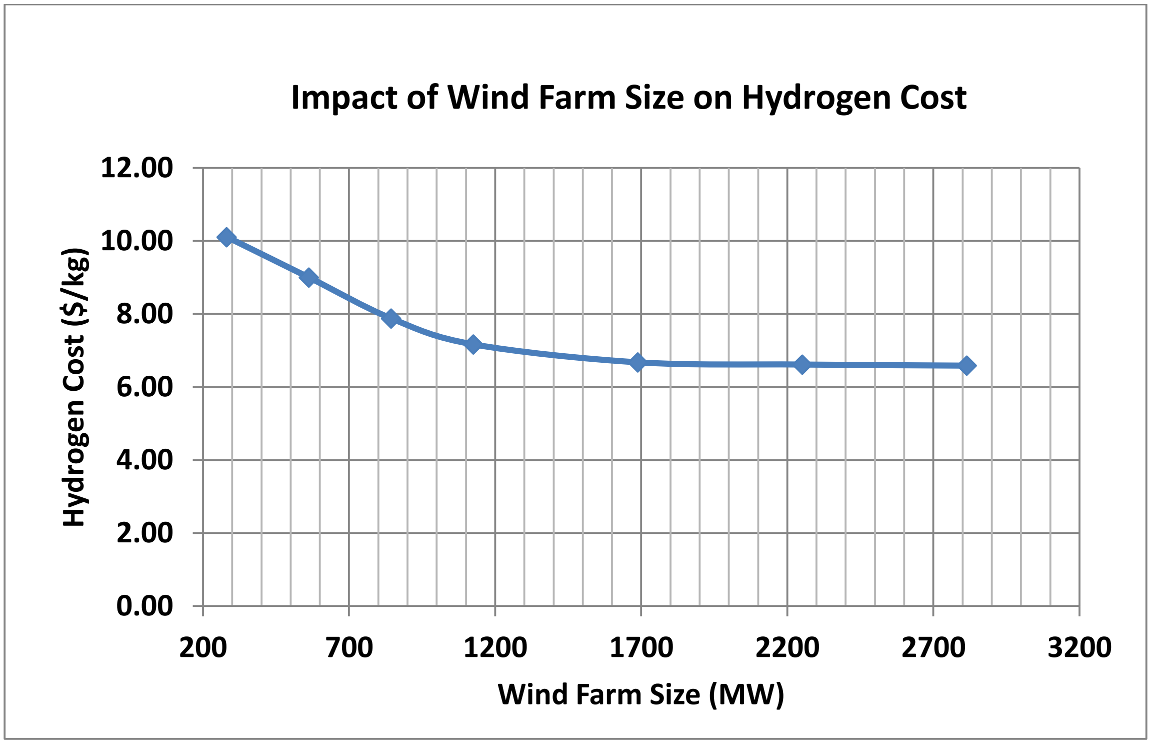

Figure 2.12: Impact of wind farm size on hydrogen production cost

2.4.1.5 Hydrogen Production Cost – Sensitivities

The sensitivity of the hydrogen cost estimates, to a number of model parameters, is illustrated in Figure 2.13. The sensitivity analysis is based on the optimal plant configuration - 81 units of the 3496 kW (760 Nm^3/hr) electrolyser and 360 MWh (60 units) of battery capacity. The effect of the battery and electrolyser efficiency are the most profound on the cost estimates; underscoring the importance of the plant's round-trip efficiency. The wind turbine capital cost has a considerable impact on the hydrogen cost estimates, reaffirming the need to consider the wind turbine capital cost ($/kW) value used in wind-hydrogen models, along with the importance of having a strong OEM base for a given jurisdiction (as this can lower capital costs, e.g., China). Negotiating competitive supply contracts from wind turbine manufacturers is also paramount. The internal rate of return (IRR) has a less significant impact relative to the wind turbine capital cost. The oxygen profit margin has a relatively modest effect on the hydrogen cost estimates.

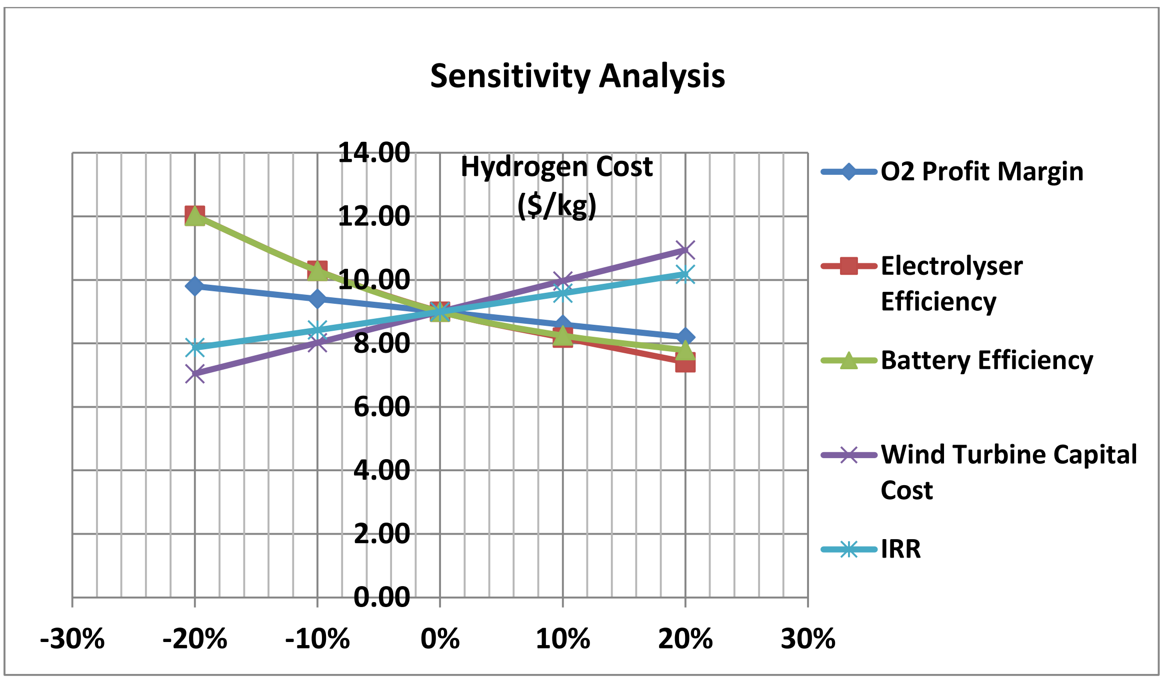

Figure 2.13: Hydrogen production cost sensitivities

2.4.2 Driving Factors for Electrolyser and Battery Sizing

The optimal battery capacity for a particular electrolyser size, is a strong function of their respective capacity factors as shown in Figure 2.14. Note: The graph shown in Figure 2.14 corresponds to a fixed electrolyser farm size of 3496 kW (760 Nm^3/hr) x 81 units. The minimum hydrogen production cost is realised when the capacity factor of the electrolyser and battery are approximately equivalent. The underpinnings of this notion stem from the relative sizing of the battery and electrolyser, and the impact it has on their performance. With this in mind, it is important to stress that the energy available to a given electrolyser, which in turn determines its productivity (and by extension, its capacity factor), is constrained by the energy capacity of the battery. Therefore, as shown in Figure 2.14, for a fixed electrolyser size, an increase in the battery size translates into an increase in the electrolyser capacity factor; however, this effect dissipates after a particular battery size is attained. This is because, at this juncture, increased storage capacity does not facilitate increased hydrogen production, as the electrolyser is no longer constrained by the battery size, but constrained solely by the energy production of the wind farms. On the other hand, as the battery size is augmented, intuitively, the capacity factor of the battery decreases. Thus, a balance between these two opposing forces facilitates a cost competitive point of operation which translates into a minimum cost. This assertion is further buttressed by the fact that the coincidence of the battery and electrolyser capacity factors leads to a minimum cost for the different electrolyser/battery sizes evaluated in the model.

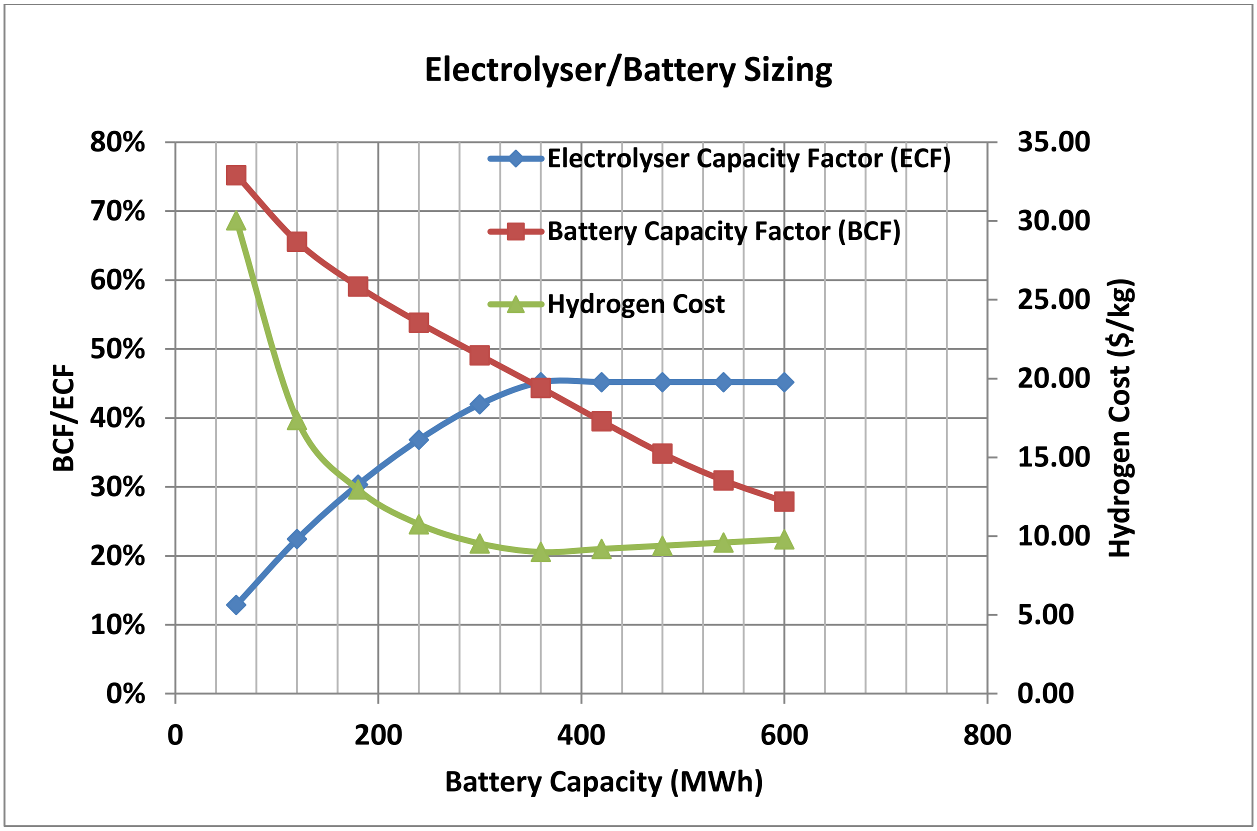

Figure 2.14: Electrolyser and battery sizing

2.4.3 Techno-Economic Impact of Energy Storage in Wind-H_2 Plants

In comparison to the production cost of an identical plant (without energy storage) outlined in a previous model [6], with updated model inputs consistent with the current model, the added element of energy storage has reduced the minimum hydrogen production cost from \$9.21/kg H_2 to \$9.00/kg H_2. This is a 2.3% decrease, which can be considered negligible. The impediment to increased cost efficiency is driven primarily by the 15% efficiency penalty associated with the battery (85% charge/discharge efficiency), and to a lesser extent by the added capital and operating costs incurred. It is important to stress that energy storage affords the plant two principal benefits: an enhanced electrolyser capacity factor and premium electricity sales. However, this benefit is realised, particularly, for smaller electrolyser farms relative to large ones – owing to the higher propensity for energy storage in the case of small electrolyser farms[6] (see Figure 2.15). Despite their enhanced capacity factors and premium electricity sales, smaller electrolysers suffer from reduced hydrogen production flow rates and increased specific capital costs, which do not justify the total capital expenditure of the plant.

[6] Energy storage is higher for smaller electrolyser farms as the electrolysers have a higher tendency to be undersized with respect to the battery capacity, facilitating an increased amount of excess energy which can be stored in the battery. Intuitively, the degree of energy storage diminishes as the electrolyser farm size is increased, owing to the significant drop in surplus energy available, as a result of the increase in the electrolyser energy demand.

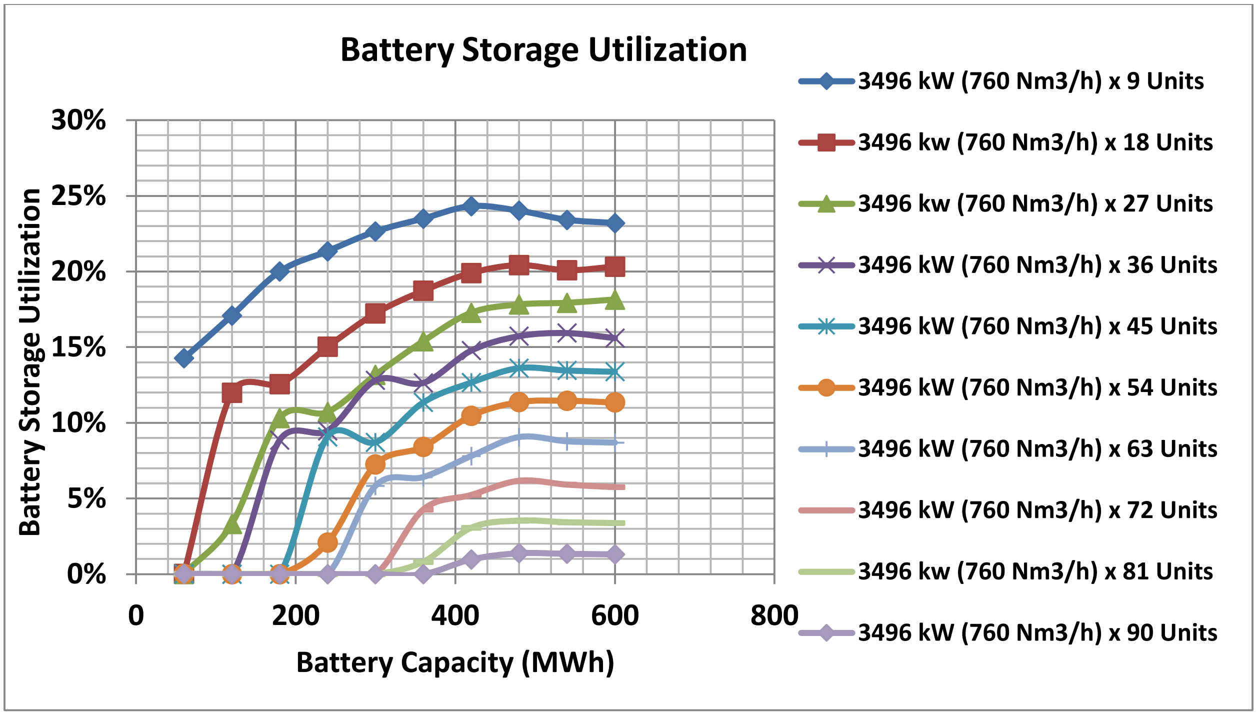

Figure 2.15: Battery storage utilization

Note: Storage utilization is defined here as the percentage of time in a year that energy is retained in the battery for storage purposes (see section 2, equation 1 in the Appendix).

2.4.4 Plant Operating Modes

The operating mode of the plant (see Table 2.2 for the description of plant operating modes) is governed by the electrolyser size, number of electrolyser units, the number of battery units, the energy generated from the wind farm, and the wholesale electricity (pool) price. As such, the operating modes vary considerably, depending on the electrolyser-battery configuration at hand. Broadly speaking, the plant operating mode will also differ from one jurisdictional context to another; as wind energy profiles and electricity pricing dynamics will vary. To give an illustration of the operating modes realized during the plant's operation, the optimal number and size of electrolysers i.e. 81 units of the 3496 kW (760 Nm^3/hr) electrolyser, is used as an example (see Figure 2.16). The impact of the battery capacity on the operating modes is also demonstrated in Figure 2.16. Two dominant and opposing trends in Figure 2.16 are worth highlighting. On one hand, Mode A (H_2 production only) becomes more prominent as the energy storage capacity of the plant is increased. On the other hand, Mode D (H_2 production only with non-premium electricity sales) becomes less dominant as the energy storage capacity is increased. This is because increasing the battery size reduces the need for non-premium electricity sales that arise due to the inability to store excess energy, as a result of the undersized nature of the battery. The increased battery size allows for the energy that would have been sold to the grid, to be utilized for hydrogen production. The trends exhibited by Mode C and Mode B also emanate from the battery sizing constraints. For smaller battery sizes, these modes are non-existent; however, once the battery size is large enough, the plant is able to make the autonomous decision to store or sell excess electricity at premium prices, alongside hydrogen production. Modes E and F have a negligible occurrence during the plant's operation, and Mode G doesn't occur at all. For about 4% of the year, the plant is at a lull; due to energy not being produced by the wind farms – Mode H. It is important to re-iterate that the trends exhibited by the plant, in terms of its operating modes, pertain specifically to the aforementioned electrolyser farm size.

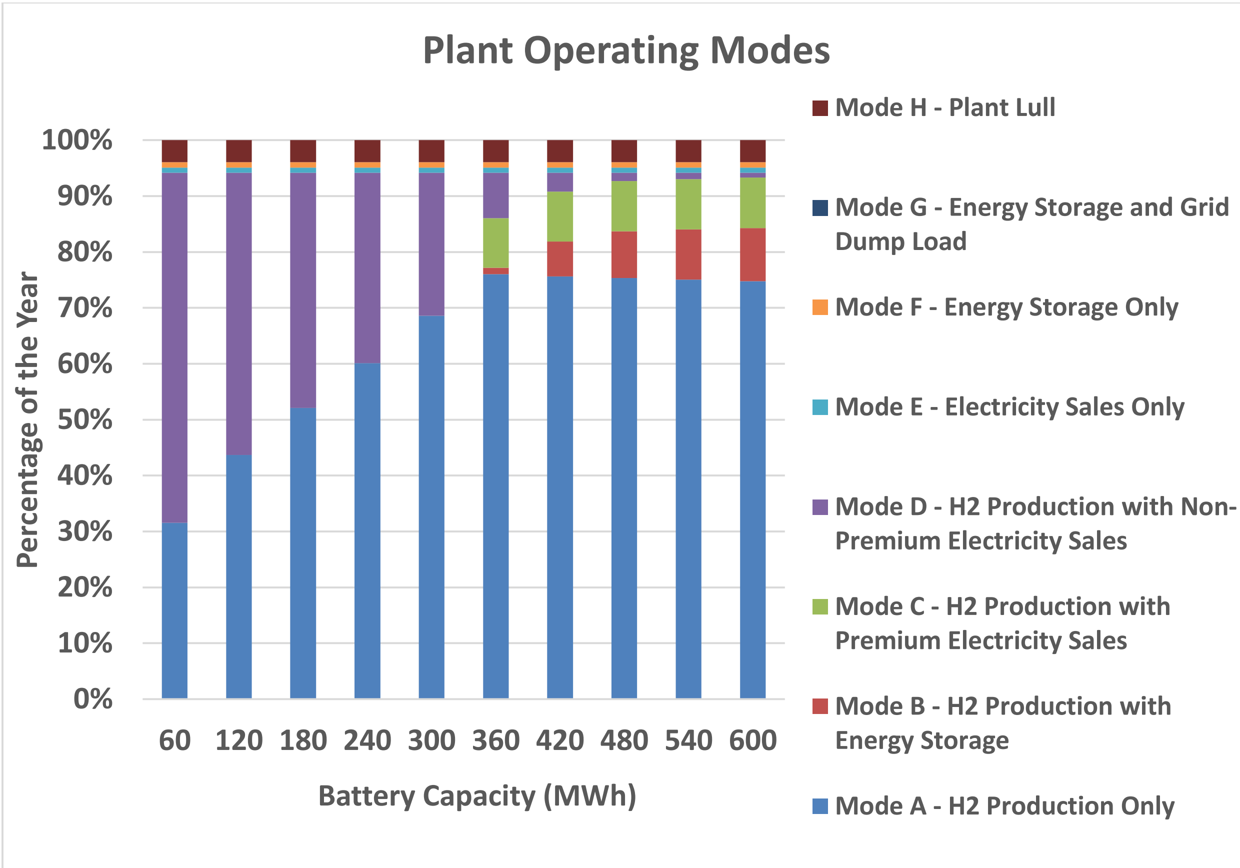

Figure 2.16: Plant operating modes

2.4.5 Modelling Methodology

The modelling methodology used to ascertain the optimal plant configuration and thus, the minimum hydrogen production cost, was effective, yielding intuitive results and new insights. In this regard, the coincidence of the electrolyser and battery capacity factors, for a minimum hydrogen production cost to be achieved, is an important finding. On another note, the use of variables that can be readily customized to suit various jurisdictional contexts (e.g. peak electricity price hours and the wind energy generation profile) in the model, provides significant flexibility for stakeholders.

A key challenge for the model was establishing the limits of the solution space for the optimal battery size. The approach taken to address this involved the initial assumption that the solution space of the battery is equivalent to that of the electrolyser (i.e. in MW). Based on observed results, the solution space for the battery is then adjusted (see section 2.2.6.1), to achieve more efficient computing. From the results yielded, this approach offers a pragmatic solution to the aforementioned challenge, be it less fluid. The main limitation of the methodology developed is that the minimum hydrogen production cost determined, is specific to a particular wind energy capacity and generation profile. That is to say, if the wind energy capacity or generation profile is changed, a different minimum hydrogen production cost would be found.

2.5 Conclusions

This paper involved the development of an integrated wind-hydrogen model termed FUNNEL – COST – H2 – WIND, which utilized real-time wind energy data to ascertain the optimal size of the electrolyser, the number of electrolyser units and battery (energy storage) capacities that would yield a minimum hydrogen production cost, whilst functioning in a liberalized electricity market with dynamic prices. The cost to produce a particular hydrogen flow rate varies widely amongst the electrolyser sizes considered. However, for a particular range of hydrogen flow rate, the optimal battery capacity for the different electrolyser sizes are coincident. The overall optimal configuration for the battery and electrolyser in the wind–hydrogen plant, comprised of 81 units of the 3496 kW (760 Nm^3/hr) electrolyser and 360 MWh (60 units) of battery capacity. This translated into a minimum production cost of \$9.00/kg H_2. The wind turbine accounts for a considerable portion of this cost i.e. 63% - hence if existing wind farms are used, the hydrogen production cost amounts to \$3.37/kg H_2.

For a particular electrolyser size, the optimal battery size occurs when the capacity factor of the electrolyser and battery are approximately equivalent. This observation was consistent for all the electrolyser and battery sizes evaluated in the model. Furthermore, the principal benefits of the battery (energy storage) on the wind-hydrogen plant, i.e., enhanced electrolyser capacity factor/premium electricity sales, are realized more readily for smaller electrolyser farms relative to larger ones. Despite the aforementioned benefits of the battery, the decrease in overall plant efficiency and to a lesser extent, increased capital costs, undermine the added benefits of energy storage. For the techno-economic conditions considered in the paper, hydrogen production from wind powered electrolysis is uncompetitive in comparison to SMR. However, depending on the volatility of natural gas prices and the cost of GHG emissions externalities (which is likely to rise in future), wind-hydrogen production can become more competitive.

2.6 References

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Chapter 3[1]

A Techno-Economic Assessment of Large Scale Hydropower-Hydrogen Production for Bitumen Upgrading in Western Canada

3.1 Introduction

Hydrogen is a vital feedstock for the heavy oil industry as it is used to upgrade[2] unconventional heavy crude to synthetic crude oil (SCO) (via hydro-treating and hydro-cracking processes). Relative to heavy crude grades, SCO has an increased market value owing to its reduced viscosity as well as sulphur, nitrogen and metal impurities [1]. In the broader conventional refining complexes world-over, hydrogen is used to enable compliance with fuel standards; most notably, sulphur content [2-4]. Thus, the demand for hydrogen in the refining sector of the oil industry (conventional or unconventional) is formidable and widespread. Alberta has a bitumen upgrading capacity of 1.35 million barrels per day (bpd), with an average of 3.4 kg of H_2 being consumed per barrel of SCO produced [5, 6]. With this in mind, steam methane reforming (SMR) has been the predominant pathway for the production of hydrogen in the bitumen upgrading industry in Alberta, Western Canada. The dominance of SMR in the hydrogen supply mix can be attributed to a multitude of factors; notwithstanding, the most notable of these is the abundance of relatively inexpensive natural gas in North America. While natural gas prices are currently low in Alberta

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[2] As opposed to hydrogen additive upgrading technologies mentioned in this paper, the upgrading of unconventional heavy crude can also be achieved via carbon rejecting technologies such as thermal coking.

and the broader North American market, they have a history of significant price volatility (see Figure 3.1). Moreover, the use of natural gas for bitumen upgrading has a significant opportunity cost. The increased penetration of natural gas fired plants in the electricity market, the development of liquefied natural gas (LNG) infrastructure to facilitate exports, along with efforts to facilitate the adoption of natural gas-to-liquids (GTL) automotive fuels in the transportation sector, have the potential to place upward pressure on natural gas prices in North America, particularly in the long term. Aside from this, the greenhouse gas (GHG) emissions footprint of SMR is significant, in the range of 11,000-13,000 tonnes of CO_2 equivalent (CO_{2e}) per tonne of hydrogen produced [7-10]. Considering the fact that recently announced environmental regulations will introduce an economy-wide carbon tax of $30/tonne CO_{2e} by 2018 [11], this creates an added incentive for the industry to utilize alternative, low-GHG hydrogen production pathways, which are economically competitive.

While costs are highly driven by project specific localized factors, hydropower has the lowest levelised cost of electricity amongst renewable generators in the majority of jurisdictions across the globe [12, 13]. Furthermore, the lifecycle GHG emissions associated with hydropower (1,128 - 2,000 CO_{2e}/tonne H_2), in similar fashion to wind energy, are considered to be one of the lowest amongst renewable pathways [14, 15]. As such, electrolytic hydrogen production powered by hydroelectric energy has the potential to produce hydrogen at a comparative cost to SMR, while mitigating a substantial amount of GHG emissions.

In Alberta, the estimated resource potential for hydropower ranges from 11.8 GW to 15 GW [16, 17] . Furthermore, 75% of this resource potential is situated in the Athabasca, Peace and Slave River basins in the province [17]. Figure 3.2 and Table 3.1 show the river basins in Alberta along with the installed capacity of existing hydro power plants as of 2011, respectively. As of 2014, the total installed hydropower capacity in Alberta amounted to 900 MW; accounting for 2.3 % of electricity production in the same year [18]. Research efforts to evaluate the potential of hydro-power based electrolytic hydrogen production in Canada is scarce; one of the possible explanations is the dominance of hydropower as a base load electricity generator in Canada's energy mix. As of 2012, hydropower accounted for 61% of the electricity generated in Canada [5]. Contrastingly, hydropower is used primarily for peak-load applications in the Alberta electricity market, evidenced by its low aggregate capacity factor[3] of 24% as of 2014 [18, 19]. Additionally, for the period of 2005 to 2013, electricity produced from hydropower in Alberta fell by 14.5%[4] [18]; this is in contrast with electricity demand growth which increased by 18.4% from 2005 to 2013 [18]. Hence, the underutilization of Alberta's hydropower capacity can be mitigated by the use of hydropower plants for renewable hydrogen production with a low GHG footprint. In this light, an integrated data-intensive techno-economic model termed FUNNEL – COST – H2 – HYDRO (FUNdamental eNgineering principlEs-based modeL for COST estimation of hydrogen (H_2) from HYDROpower) is developed in this paper, to provide a credible estimate of hydropower-hydrogen production costs in Western Canada.

Against the backdrop of the global climate change agreement achieved at the recent 2015 COP21 United Nations Conference on Climate Change held in Paris, the importance of techno-economic assessments that pertain to renewable sources of hydrogen with a low GHG footprint, is further emphasized. The authors have investigated a number of hydrogen production pathways from different perspectives [9, 20-29], with the exception of hydropower. The existing literature that

[3] It is important add that hydropower plants are also used for flood control and water management in Alberta. These two operations take precedence over energy production in hydropower plants; as a result, they can lead to low plant capacity factors.
[4] Considering 2014 data, the decrease in hydropower generation is more profound; amounting to a 21.5% drop from 2005 levels [18].

pertains to the techno-economic modelling of hydropower-hydrogen systems is quite limited in the recent decade when compared to other hydrogen pathways. Notwithstanding, a multitude of systems have been proposed, which are assessed from a techno-economic standpoint with varying degrees of rigor. Each of these systems involves electrolytic hydrogen production using the electrolysis of water. The systems put forward in literature can be broadly categorized into three main themes. First, a number of studies have proposed small scale hydropower-hydrogen systems, where hydrogen is used to service the electricity/heat generation needs of remote off-grid communities [30-34]. Alternative models are premised upon the use of excess water from hydropower reservoirs, which are 'spilt' without harnessing their potential for hydrogen production [35, 36]. In these studies, the hydrogen produced is used in the electricity generation (peak-load applications/energy storage), transportation (fuel-cell vehicles) or in the value added industries i.e. food, pharmaceuticals and ammonia industries. Furthermore, the dedicated or off-peak use of hydropower plants for hydrogen production has been the basis of other models [37-42], where hydrogen has similar end uses as in the previous category of studies. Other related research to capitalize on hydropower-hydrogen potential, involve its use for methanol production [43]. From the perusal of previous studies a number of noteworthy trends have been identified. With the exception of the model presented by Bellotti et al. 2015 [37], a number of models do not address the optimal sizing (to minimize cost) and configuration of the electrolyser plant. Having said that, Bellotti et al. (2015) [37] does not consider the impact of the hydropower-hydrogen plant functioning in a liberalized electricity market, and the effect of the dynamic electricity prices therein, on sizing considerations. Furthermore, hydrogen yield and electrolyser energy consumption are based upon idealized efficiencies, generic correlations, and assumptions of key metrics (e.g. electrolyser capacity factor) in some cases [30, 31, 35, 38, 39]. Moreover, fixed electricity prices which are not indicative of the dynamics of a liberalized electricity market, are often used to estimate hydrogen production costs [32, 36, 38, 39]. Additionally, some models proposed have limited transparency in terms of the key techno-economic data used, due to confidentiality and other factors [41]. Furthermore, a limited amount of studies present integrated hydropower-hydrogen models which take a holistic account of all unit operations involved from hydrogen production, to its delivery to the end user. The model developed circumvents the limitations highlighted above, translating into the following objectives:

- The development of an integrated grid-connected hydropower-H_2 techno-economic model for the production of renewable hydrogen and estimation of costs, in a liberalized electricity market with dynamic prices.
- The development of a techno-economic framework for the determination of the optimal electrolyser size and number of electrolyser units, which yields a minimum hydrogen production cost, for hydropower-hydrogen systems.

All costs indicated in this paper are in 2014[5] Canadian dollars unless otherwise specified.

[5]Where necessary, an inflation rate of 2% has been used to convert all costs into 2014 $CAD. Furthermore, currency rates of $1CAD = $1US; $1.3CAD = €1; $1.6CAD = £1 are adopted in this paper.

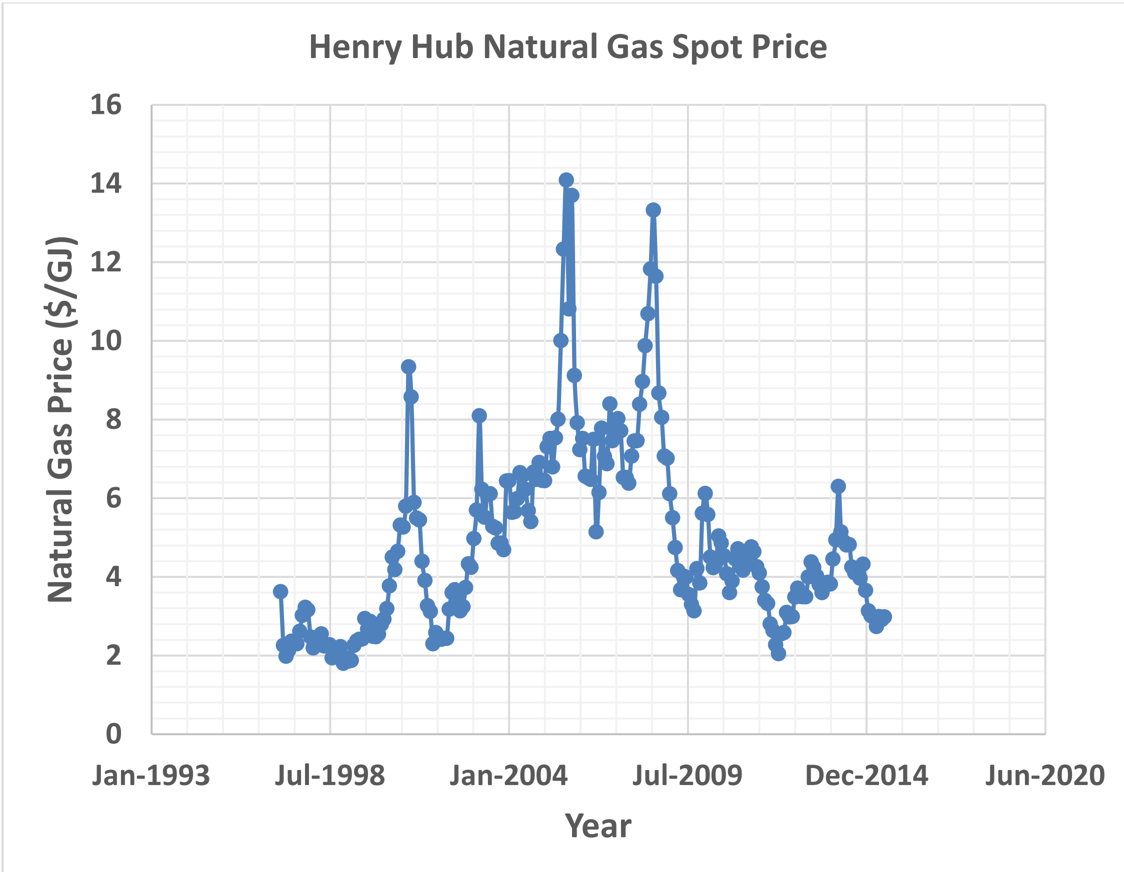


Figure 3.1: Historical natural gas price in Alberta [44]

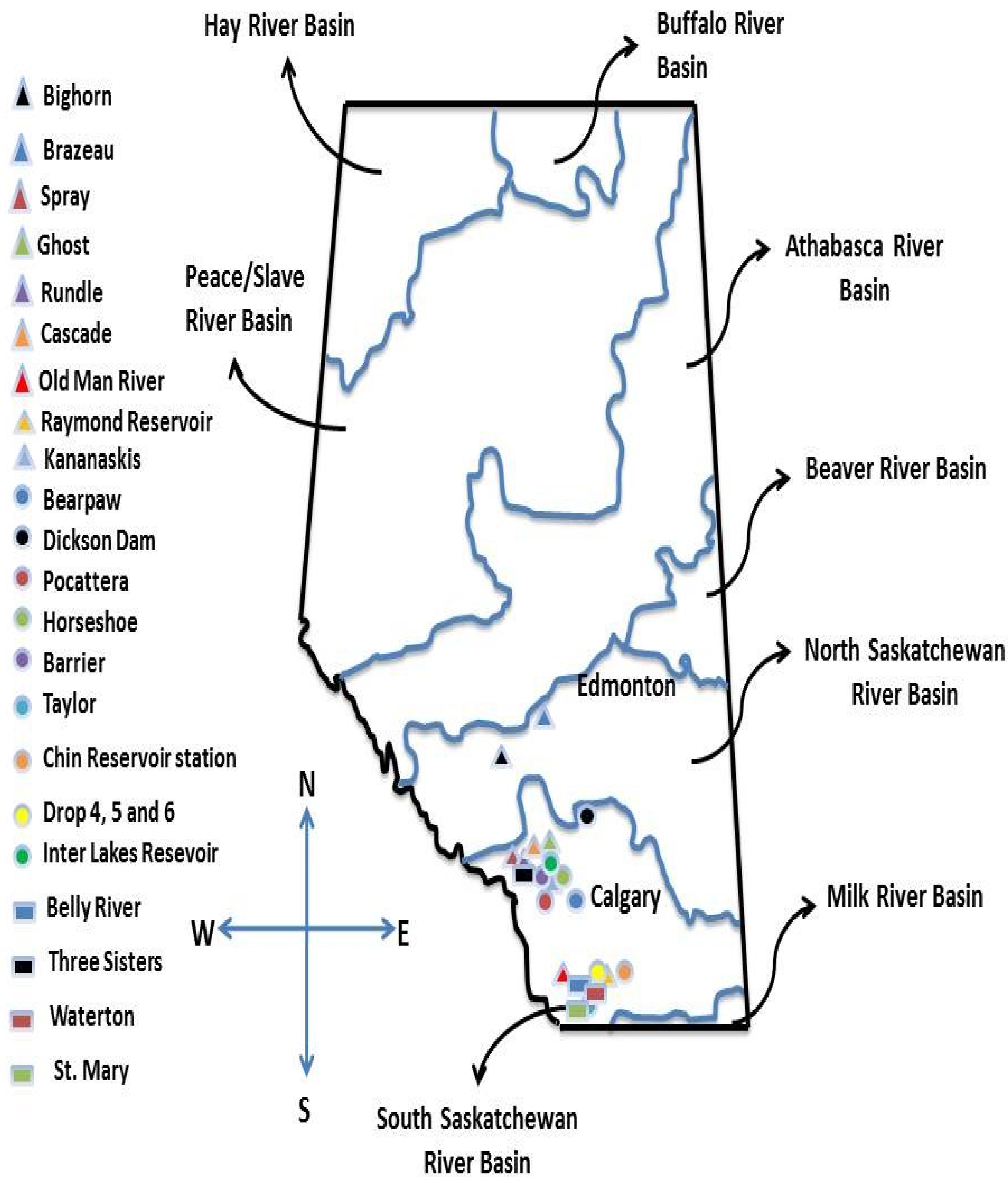


Figure 3.2: Alberta's river basins and existing hydropower plants (2011)

Table 3.1: Grid-connected hydropower capacity in Alberta as of 2011 [45]

Hydropower Plant Name	Plant Capacity (MW)
Brazeau	355
Bighorn	120
Spray	103
Ghost	51
Rundle	50
Cascade	36
Oldman River	32
Kananaskis	19
Raymond Reservoir	18
Barrier	13
Taylor	13
Chin Reservoir Station	11
Drop (4,5 & 6)	7
Inter Lakes Resevoir	5
Belly River	3
Waterton	3
St. Mary	2
TOTAL	**891**

3.2 Methodology and Scope

3.2.1 Hydropower-Hydrogen Plant Description

The technical details of a 436 MW hydropower plant proposed by Figueiredo and Flynn (2006) [46] are utilized in the model. The authors use the plant's pumped hydro storage capacity to take advantage of energy arbitrage opportunities on the electricity grid, and thus investigate the optimal sizing of the pump/generator relative to the reservoir (storage) capacity. However, the use of the plant in this current undertaking is for hydrogen production. The plant is located in Grand Cache, south-western Alberta (see Figure 3.3) – a conceptual schematic of the plant is shown in Figure 3.4. The hydrogen produced is transported to the Edmonton industrial heartland via a hydrogen pipeline, where a bitumen upgrader consumes the electrolytic hydrogen.

Unlike other intermittent renewables, the need for energy storage to smoothen the erratic profile of the energy generated, so as to allow the electrolyser achieve its rated efficiency and operational life [29], is not needed in the case of hydropower-hydrogen plants due to its non-intermittent energy generation (the hydropower plant in this model operates at constant baseload capacity). This highlights a significant competitive advantage. Apart from the cost savings realized from mitigating the need for energy storage, what is more significant is the higher roundtrip efficiency this affords the plant.

Lastly, as illustrated in Figure 3.4, the mechanism of hydrogen production is as follows: the hydropower plant (turbine) produces electricity which is converted from AC to DC by the rectifier. The DC energy produced, fuels the electrolyser which, while consuming feed water, produces hydrogen. Accordingly, the hydrogen produced is then compressed[6] to the required pressure amenable to pipeline transportation, which is eventually delivered to the bitumen upgrader.

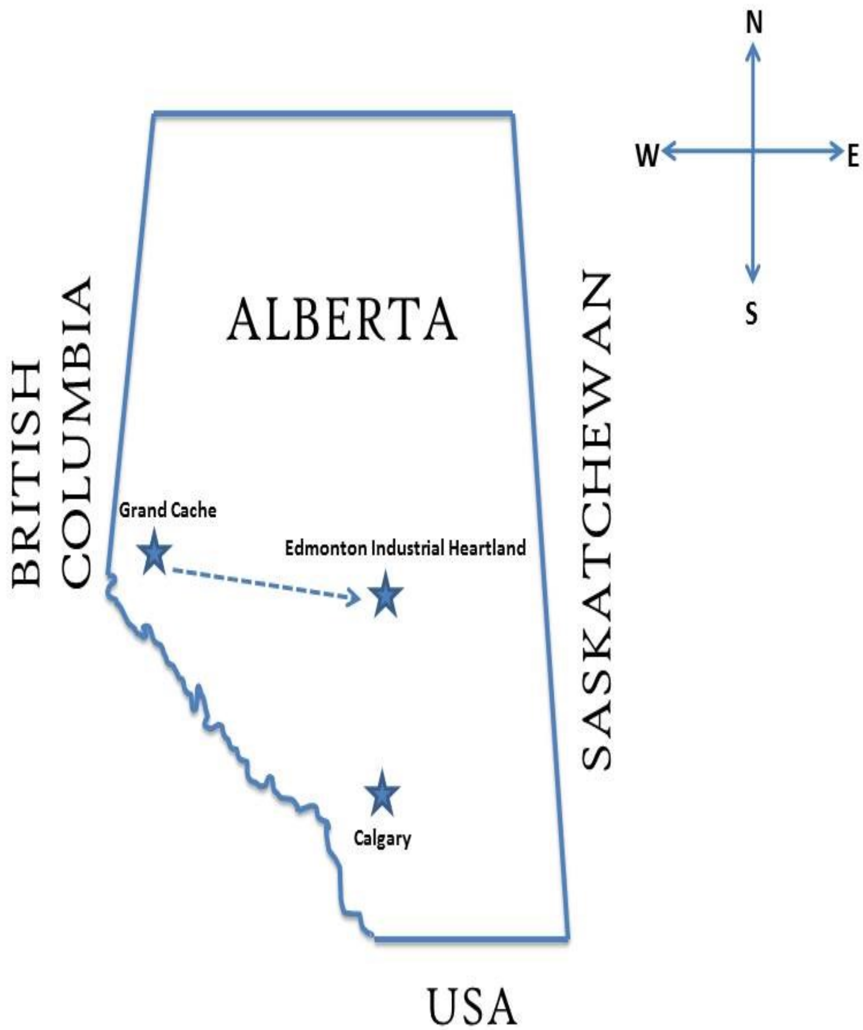


Figure 3.3: Proposed plant location (Grand Cache) relative to Edmonton industrial heartland (bitumen upgrader location).

[6] The energy for compression is sourced from the electricity grid at an assumed cost of \$70/MWh. This also represents an added GHG emissions footprint for the plant, which is a function of the emissions intensity of the grid.

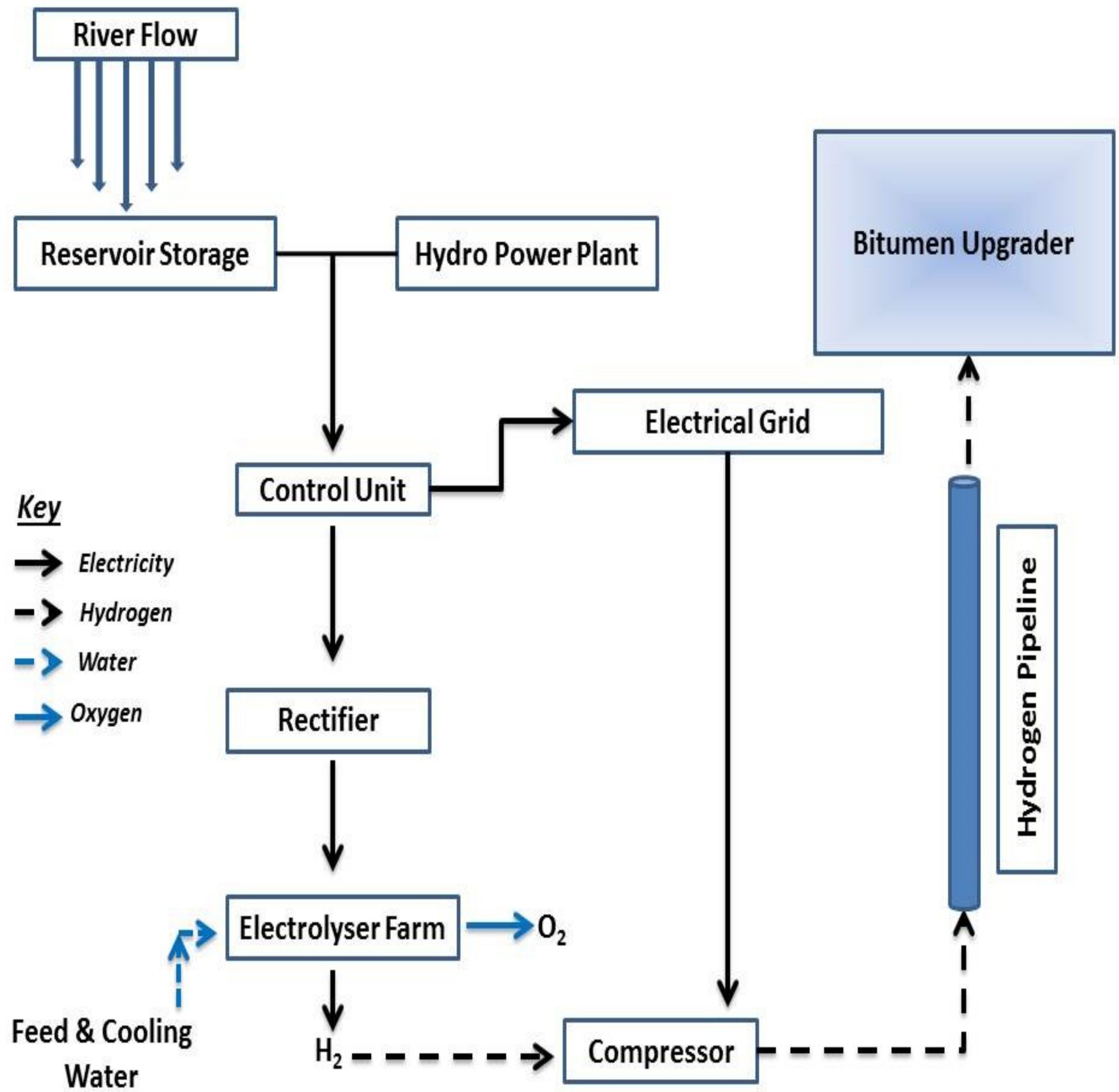


Figure 3.4: Conceptual schematic of the hydropower-hydrogen plant

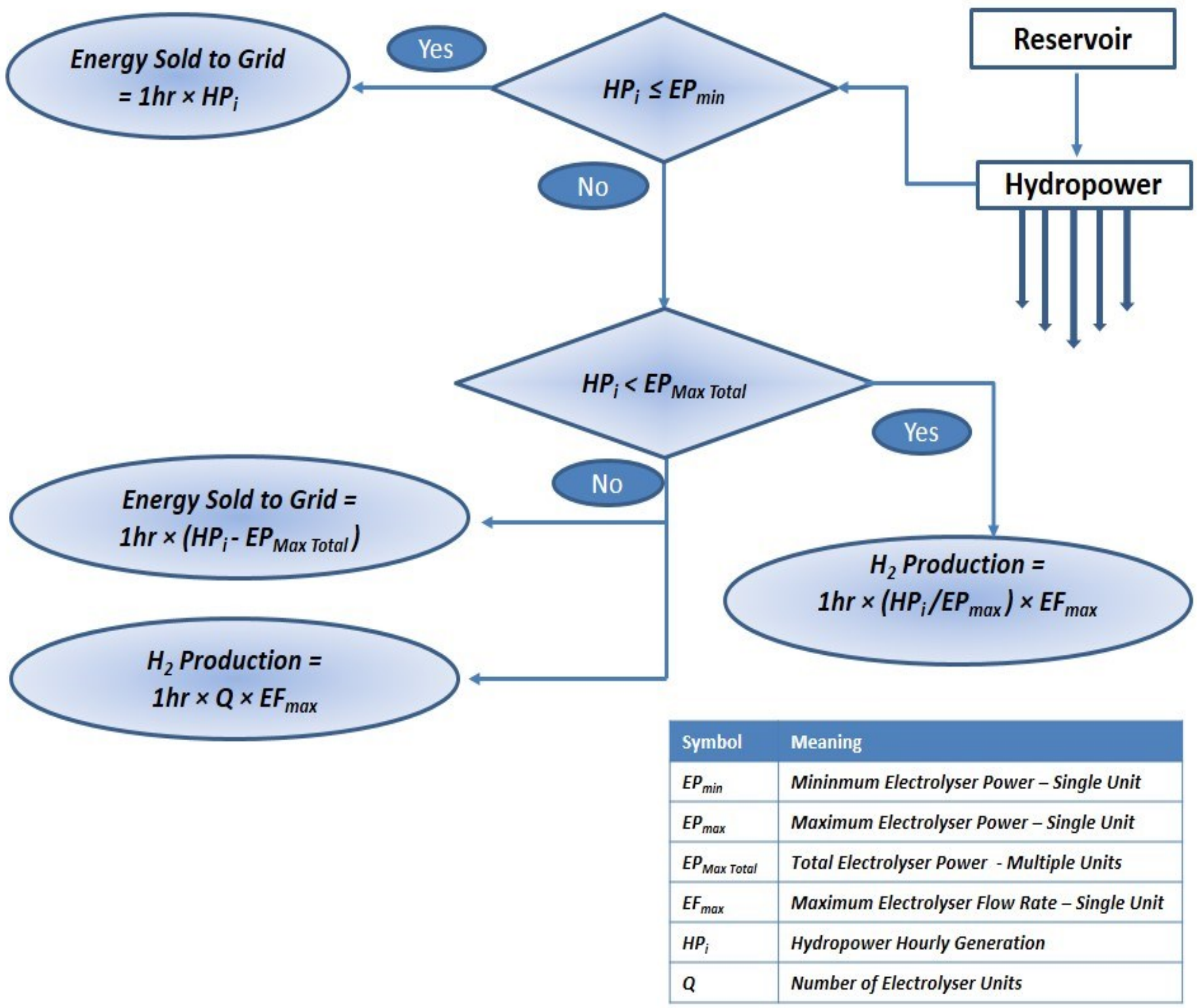
3.2.2 Energy Management and Quantification of Hydrogen Production

The plant's default operating mode is for hydrogen production only. However, due to its grid connection, instances where the amount of electricity generated by the plant creates an energy surplus relative to electrolyser demand, this is sold to the grid to enhance its economic competitiveness. Additionally, in the event that the electricity produced in the plant falls short of the threshold required for hydrogen production, this is also sold to the grid. The amount of hydrogen produced is a function of the energy output of the plant, the electrolyser energy demand (rated power), number of units, flow rate and efficiency[7]. In essence the plant has three possible operating modes, hydrogen production only; electricity production only; simultaneous production of hydrogen and electricity. The control unit determines the operating mode of the plant based on the energy management flow chart provided in Figure 3.5. It is also worth mentioning that the oxygen produced as a by-product of the plant (H_2:O_2 production is 2:1) is also sold at the plant gate to augment revenue – further details of the oxygen revenue stream are provided in section 3.4.

Depending on the operational mode, the hourly amount of energy generated in the hydropower plant is used to calculate the hourly amount of hydrogen/electricity produced for a period of 8760 hrs, i.e. one year. Furthermore, the hourly wholesale electricity (pool) price is used to calculate the energy revenue for each hour in the year where applicable. Data for the hourly pool[8] price corresponds to the year 2011, and was provided by the Alberta Electric System Operator (AESO) [47]. The summation of the hourly values of hydrogen production/electricity generation, yields the corresponding annual values. These annual values are then used within the FUNNEL – COST

[7] The difference in the values of efficiencies (energy consumed per unit of hydrogen produced) for the electrolysers evaluated in the model, can be considered negligible (see Table 3.2 for details). Hence, their relative performance is not dependent on their efficiency, but their size (rated power) in particular.
[8] Pool price refers to the hourly wholesale price of electricity in Alberta's liberalized electricity market.

– H2 – HYDRO model to calculate the hydrogen production costs (via an embedded discounted cash flow model of the plant) and other performance metrics.



Symbol	Meaning
EP_{min}	*Mininmum Electrolyser Power – Single Unit*
EP_{max}	*Maximum Electrolyser Power – Single Unit*
$EP_{Max\ Total}$	*Total Electrolyser Power - Multiple Units*
EF_{max}	*Maximum Electrolyser Flow Rate – Single Unit*
HP_i	*Hydropower Hourly Generation*
Q	*Number of Electrolyser Units*

Figure 3.5: Energy management flow chart

3.2.3 Electrolyser Selection and Modelling

The existing electrolyser (electrolysis) technologies that are prevalent in the pertinent literature can be broadly categorized into three, namely: alkaline electrolysers, proton exchange membrane (PEM) electrolysers, and high temperature electrolysis (HTE) [28]. Relative to other electrolyser technologies, alkaline electrolysers are utilized in the model presented here due to their superior technological maturity, large scale hydrogen flow rates and relatively inexpensive capital cost. For a more detailed examination of the aforementioned electrolyser pathways, the reader is referred to the work by Olateju & Kumar [28].

In this paper, a systems-level approach is implemented in the modeling of the performance of the electrolyser, based on its salient characteristics. The trade-offs involved with the use of systems-level models vis-à-vis 'element-level' models, has been addressed comprehensively by the authors [29]. This study assumes that the nominal efficiency of the electrolyser remains constant during its operation, due to the steady generation profile of electricity from the hydropower plant. It is worth pointing out that in previous hydrogen models from intermittent renewable sources developed by the authors, the electrolyser has been assumed to operate at 73% of its nominal efficiency [27, 28].

A total of six different electrolyser sizes were considered in this study, the performance specifications of each electrolyser is outlined in Table 3.2. Similar to the approach adopted in [27], the minimum electrolyser power requirement for all electrolysers, has been determined based on a proportional relationship between the maximum flow rate and maximum power demand (rated power) of the electrolyser as shown in Eq. (1). The rationale behind this approach is the fact that the minimum operating threshold for electrolysers varies widely in literature; ranging from 5-50% of its rated power [48, 49], depending on the scale and manufacturer of the unit. Thus, for reasons of consistency, this methodology has been adopted. On another note, the efficiency of the rectifier and compressor have been taken as 95% and 70% respectively [50, 51]. Furthermore, it is worth pointing out that the hydropower generator efficiency assumed in this paper is 90%.

$$\boldsymbol{EP_{min}} = \frac{(\boldsymbol{EF_{min}} \times \boldsymbol{EP_{max}})}{(\boldsymbol{\eta} \times \boldsymbol{EF_{max}})} \qquad \text{(Eq. 1)}$$

Where: η represents the efficiency of the rectifier; EF_{min} and EF_{max} represent the electrolyser maximum and minimum flow rates, respectively. EP_{max} represents the electrolyser rated power.

Table 3.2: Electrolyzer size range [52, 53]

Electrolyser manufacturer/model	Min. H_2 flow rate (Nm^3/hr)	Max. H_2 flow rate (Nm^3/hr)	Energy requirement (kWh/Nm^3)	Nominal Efficiency (HHV) (%)[d]	Size (kW)	H_2 pressure (bar)	H_2 purity (%)
Norsk Hydro Atmospheric Type No. 5010 (5150 Amp DC) [52]	0[a]	50	4.8[b]	72.4	240	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5020 (5150 Amp DC) [52]	50	150	4.8[b]	72.4	720	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5030 (5150 Amp DC) [52]	150	300	4.8[b]	72.4	1440	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5040 (4000 Amp DC) [52]	300	377	4.8[b]	72.4	1810	1	99.9 ± 0.1
Norsk Hydro Atmospheric Type No. 5040 (5150 Amp DC) [52]	300	485	4.8[b]	72.4	2328	1	99.9 ± 0.1
Industrie Haute Technologie (IHT) Type S-556 [53]	190	760	4.9[b,c]	70.8	3496	30	99.9 ± 0.1

[a]A minimum flow rate of $1 Nm^3/hr$ was utilized in this study.

[b]Indicates the hydrogen production systems level energy requirement specified by the manufacturer[52].

[c]Average value of the energy requirement range (4.6-5.2 kWh/Nm^3) indicated.

[d]The nominal efficiency defined here is the ratio of the ideal energy consumption for water electrolysis (39 kWh/kg H_2) to the nominal energy consumption per unit of hydrogen produced for each electrolyser (at its rated power).

3.3 Cost Estimation

3.3.1 Hydropower Capital Cost and Auxiliary Unit Cost

The hydropower capital cost and auxiliary unit costs for the plant are outlined in Table 3.3. As mentioned earlier, hydropower capital costs vary significantly and are highly site specific. The capital intensive nature of hydropower plants makes the capital cost value utilized in the model of vital importance. To put the specific installed capital cost adopted in the model in context, it is roughly twice the capital cost incurred for a wind farm of the same capacity [29]. The cost of $4000/kW is also comparable to the value of $3,788/kW specified in a detailed study for the United States National Renewable Energy Laboratory (NREL) [54]. Tables 3.3 and 3.4 put the specific capital cost utilized in this study into broader context.

Still focusing on the capital cost of the hydropower plant, it is worth highlighting the fact that some hydropower plants that are still in operation in Alberta (though underutilized) are likely to have had their capital cost fully recovered (e.g. the Brazeau 355 MW plant, which was built in 1965 [55]). As a result, the use of these types of plants to facilitate hydropower-hydrogen production is particularly promising from an economic perspective, not least because hydropower plants have relatively low operating costs and no fuel cost. However, a holistic economic evaluation that includes capital cost expenditure is undertaken in this paper, to facilitate comparisons with other hydrogen pathways investigated by the authors [26-29].

Table 3.3: Hydropower capital and auxiliary plant costs

Cost components	Values	Sources/Comments
Hydropower installed capital cost ($/kW)	4000	The range of installed capital cost specified for Canada by the International Renewable Energy Agency (IRENA) [56], ranges from $811 - $4870/kW. Alberta is likely to be closer to the upper end of this range. Moreover, the specific capital cost utilized in this paper falls within the range of recent capital cost estimates for greenfield hydropower projects in Canada (see Table 3.4).
Plant power electronics cost ($/kW) (including rectifier and control unit cost)	35	Estimated relative to the cost specified for a 1GW wind-hydrogen plant [57].
Electrolyser labour and installation costs ($)	Function of electrolyser size.	10 % of electrolyser capital cost.
Electrolyser O&M cost ($/kW/yr)	18.4	[58]
Electrolyser cell stack replacement cost	Function of electrolyser size.	30% of electrolyser capital cost [59].
Hydropower O&M cost ($/yr)	2.6 % of total installed capital cost	Based on values specified by [46]
Pincher creek water cost ($/m3)	0.99	[28]

Cost components	Values	Sources/Comments
Hydropower plant life (yrs)	40	A conservative estimate of the plant life time is adopted in this study.
Electrolyser service life (yrs)	10	[59, 60].
Inverter service life (yrs)	10	[61]
Control unit service life (yrs)	10	[28]

Table 3.4: Recent Canadian greenfield hydropower capital costs estimates [16, 62-64]

Name	Location	Cost ($ Millions)	Capacity (MW)	Installed Cost ($/kW)	Construction Activity
Site C [62]	British Columbia	8,775	1100	7977	In progress
Romaine Complex A [16, 63]	Quebec	6,500	1550	4193	In progress
Lower Churchill [16, 64]	Labrador	6,200	3000	2067	In progress

3.3.2 Electrolyser Capital Cost

The electrolyser capital cost incorporated into the model is based on the work carried out by Olateju & Kumar [28]. The authors aforementioned present a model that yields the specific capital cost of alkaline electrolysers as a function of electrolyser size. This model has been used to account for the capital expenditure for all the different electrolyser farm[9] configurations investigated. It is worth mentioning that volume discounts are likely to be achieved with electrolyser manufacturers in practice, as the purchase of a large number of units is likely to yield strong negotiating power in terms of supply contracts, which will facilitate a more competitive capital cost value. However, a conservative approach is adopted in this paper where none of the aforementioned economies are realized. The electrolyser capital cost model is specific to alkaline electrolysers and indicative of the state of the technology as of the early 2000s, not the state of the art. This is as a result of the limited availability of data. Specific capital cost data is considered proprietary by a number of electrolyser manufacturers. Nonetheless, the estimates provided by the model are within reason [29].

[9] An electrolyser farm consists of a specific electrolyser size and a number of electrolyser units.

3.3.3 Hydrogen Pipeline and Compressor Cost

Based on the hydrogen flow rate yielded by each electrolyser farm configuration assessed, a hydrogen pipeline is sized using the Panhandle B equation [65] – see sections 3 and 4 of Appendix. The pipeline distance estimate of 432 km from Grand Cache to the Edmonton industrial heartland where the bitumen upgrader is located, is based upon the driving distance [66]. The capital cost of the pipeline is accounted for using a model developed in a previous study [67]. The model utilized compares favorably with the estimates of alternative hydrogen pipeline models [68, 69]; with discrepancies falling within a range of 10-18%. Common to all pipelines, the capital cost will be determined by site-specific factors along with the properties of the transport fluid. In the case of hydrogen pipelines, measures to address the potential embrittlement of steel and hydrogen leakage will be particularly important. In general, there is an elevated risk of pipeline operation associated with hydrogen pipelines, relative to other industrial fluids (e.g. CO_2, natural gas etc.), which has to be factored into the cost estimates for improved accuracy [29]. On another note, a compressor is used to elevate the hydrogen pressure to 60 bar so as to facilitate pipeline transport [26, 27]. Similar to the pipeline, a compressor is sized for each electrolyser farm configuration assessed. Further details of the pipeline and compressor specification, as well as their corresponding cost estimates utilized in this paper, can be found in [26].

3.3.4 Principal Economic Data and Model Assumptions

In the FUNNEL – COST – H2 – HYDRO model, a return on equity of 10% along with an inflation rate of 2% was adopted. The hydropower-hydrogen plant investment is assumed to be serviced by 100% equity; with an operating life of 40 years and a decommissioning cost with a negligible present value. Another assumption is that the plant does not benefit from any renewable energy incentives such as feed-in-tariffs (FIT). Furthermore, the duration of plant construction is considered to be three years. As mentioned earlier, oxygen, which is a by-product of the electrolysis process, is also considered as a revenue generation stream. It is important to stress that the price for oxygen varies substantially depending on the market in which it is sold, the scale of production and its level of quality (purity). Price quotes varied from \$66.57/Nm^3 for medical grade (99.99% purity) oxygen from retail level vendors [70], to \$0.078/Nm^3 for large industrial scale producers [71] . Furthermore, in the published literature a price of \$2.77/Nm^3 (originally from Praxair Inc.) is cited by Becalli et al. (2013) [72], however the specific market in which oxygen is sold is not apparent.

The hydropower-hydrogen plant produces oxygen with a purity level that exceeds 99.99%; thus it is compatible with the standards for medical grade applications in Canada, as evidenced by the specifications provided by Praxair Inc. [73]. Furthermore, medical grade oxygen trades at a significant premium to industrial application oxygen, which can aid the competitiveness of the plant. The demand for the high purity oxygen at the plant is assumed to driven by oxygen consumption in Alberta hospitals and other institutions such as care homes, which purchase medical grade oxygen at the plant gate. In the model, an incremental oxygen revenue of \$0.50/kg O_2 is assumed, based on a selling price of \$3.60/Nm^3 [29].

3.4 Results and Discussion

3.4.1 Hydrogen Production Cost

The hydrogen production cost curves for the different electrolyser sizes evaluated, all exhibit a non-linear trend as shown in Figure 3.6. Significant economies of scale are realised as the hydrogen production flow rate of the plant is increased (it is important to mention that increases in the plant's hydrogen flow rate coincide with increases in the number of electrolyser units). As the flow rate is increased to larger magnitudes, the economies of scale realised decrease progressively, until a minimum hydrogen production cost is achieved. After the minimum cost is achieved, with further increases to the number of electrolyser units, the hydrogen flow rate remains constant; resulting in a rapid rise in the production cost (see Figure 3.6). This occurs because the electrolyser farm is oversized relative to the amount of energy produced by the hydropower plant.

Figure 3.6 also shows that for a particular hydrogen flow rate to be produced by the plant, the cost incurred varies significantly depending on the electrolyser size that is used. This is because in order to achieve a given flow rate magnitude, the number of electrolyser units required varies considerably, depending on the electrolyser size. Smaller electrolyser sizes require a significantly higher number of units in comparison to larger electrolysers. It is also worth mentioning that apart from providing more competitive production costs (as seen in Figure 3.6), the larger sized electrolysers in general allow for energy management, monitoring, operational and maintenance endeavours that are more pragmatic compared to smaller electrolysers, due to the significantly reduced number of units required.

The minimum hydrogen production cost for the hydropower-hydrogen plant amounts to \$2.43/kg H_2 – this corresponds to an electrolyser farm with 90 units of the 3496 kW (760 Nm^3/h) rated electrolyser. This cost is competitive with SMR/SMR-CCS production costs, which vary from \$1.87/kg H_2 to \$2.60/ kg H_2. This point is buttressed by the fact that if existing hydropower plants are used (hence negating hydropower capital costs), the minimum production cost amounts to \$1.18/ kg H_2 (see Figure 3.7).

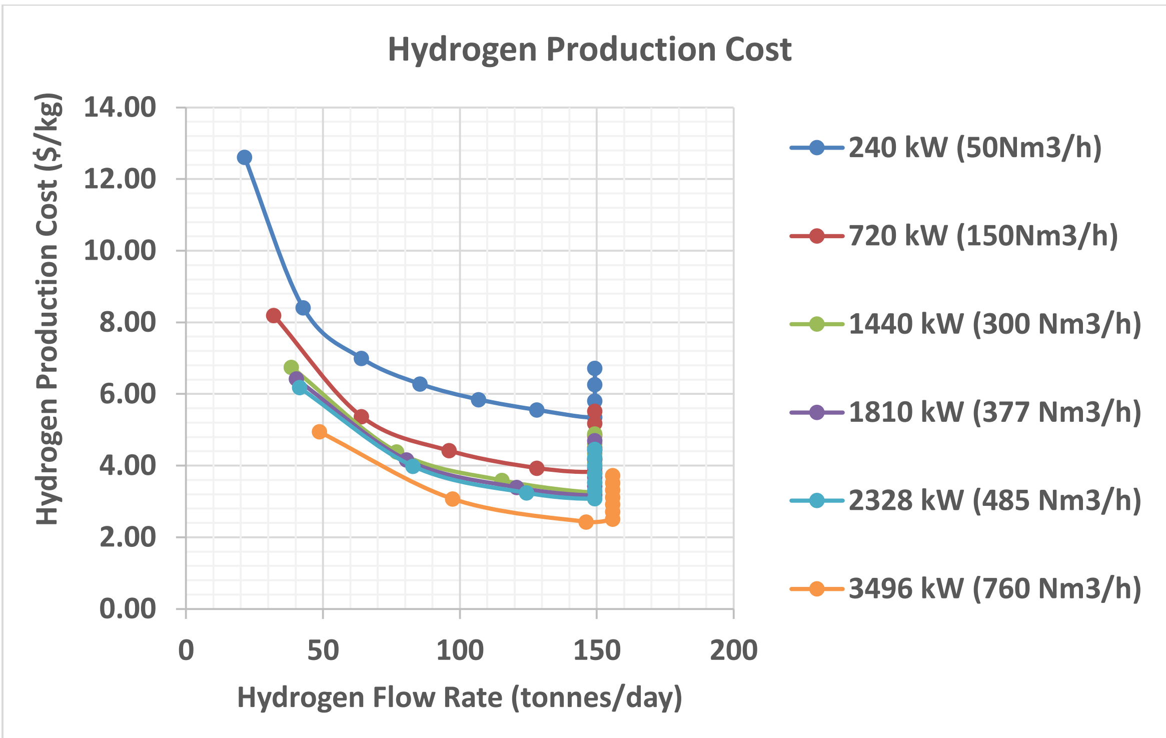


Figure 3.6: Hydrogen production costs

3.4.2 Hydrogen Production Cost Distribution

For each electrolyser size considered in the model, the minimum H_2 cost distribution is shown in Figure 3.7. For smaller electrolysers, the cost of the electrolysers units is the most significant cost component; accounting for 60% and 44% of the total hydrogen production cost in the case of the 50 Nm^3/h and 150 Nm^3/h electrolysers, respectively. This is due to the significant number of units required – 1400 and 500 for the 50 Nm^3/h and 150 Nm^3/h electrolysers, respectively; which is an order of magnitude greater than the 90 units required by the largest electrolyser. Apart from the sheer volume of units, the specific capital costs for smaller electrolysers are higher vis-à-vis their larger counterparts; hence, the total electrolyser investment cost is more significant – this also elevates the replacement cost of the electrolysers, which are replaced 3 times during the plant's 40 year lifetime. Moreover, as a result of the relatively high capacity factor of the hydropower plant (which results in a high electrolyser capacity factor despite the high number of units), the costs of running the electrolysers, including: water resource costs, operating and maintenance costs, are high in comparison to larger electrolysers.

The overarching trend indicated in Figure 3.7 is such that as the electrolyser size increases, the hydropower cost becomes increasing dominant, while the electrolyser cost decreases in significance. The pipeline and compressor cost are relatively consistent amongst the different electrolyser sizes, due to the fact that a similar magnitude of hydrogen flow rate (ranging from 146 – 149 tonnes H_2/day) is transported and compressed for the different minimum cost values of the electrolysers. The compressor cost for the largest electrolyser is relatively minute due to its high hydrogen production pressure (see Table 3.2). The power electronics cost is relatively insignificant for all electrolyser sizes.

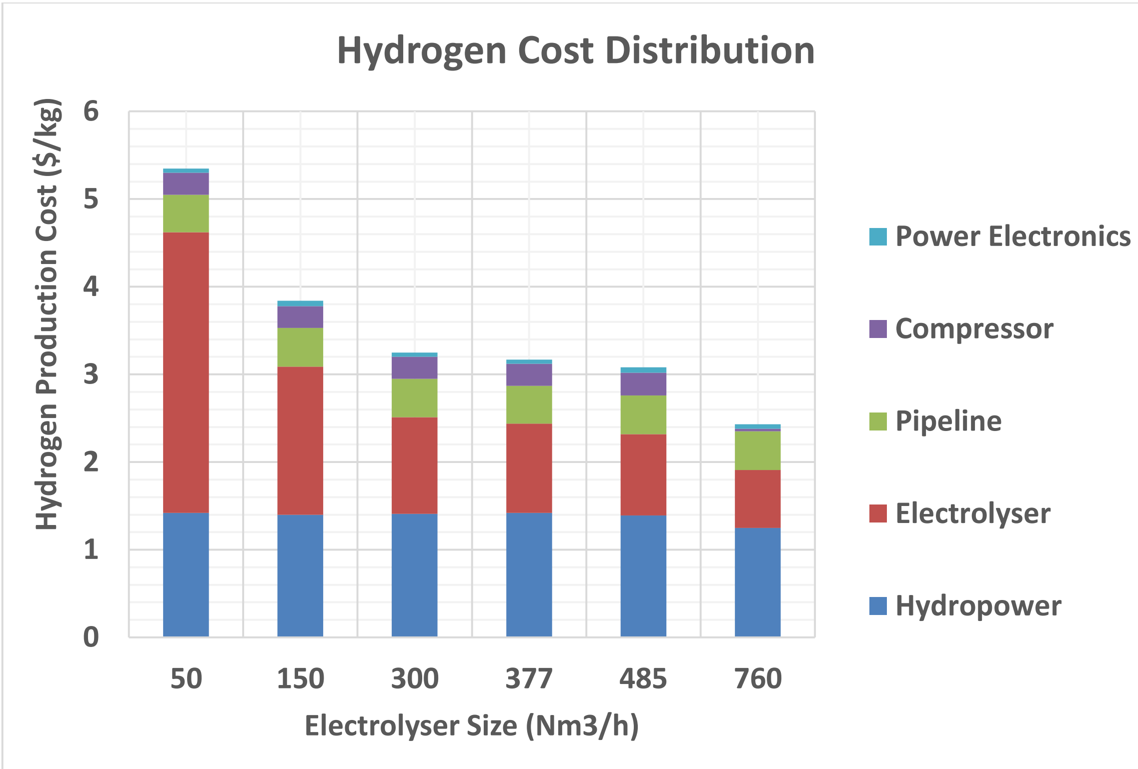


Figure 3.7: Hydrogen cost distribution

3.4.3 Hydrogen Production Cost – Sensitivities

The sensitivity[10] of the production cost estimate to key techno-economic parameters is shown in Figure 3.8. The electrolyser efficiency has the most significant effect on the cost estimates not least because of its dual impact; it influences the amount of hydrogen produced and the amount of surplus energy available to be sold to the grid. Intuitively, the installed capital cost estimate of the hydropower plant also has a formidable effect on the hydrogen production cost. As eluded to earlier, the sensitivity of the installed capital cost is indicative of the highly cost competitive

[10] The sensitivity analysis carried out is based upon the plant configuration which yielded the minimum hydrogen production cost – 90 units of the 3496 kW (760 Nm3/h) electrolyser.

production cost that can be achieved with existing hydropower plants in Alberta (with sunk capital costs) used for hydrogen production. Alternatively, upgrades to existing plants e.g. new generators etc, would also be reflective of a reduction in capital cost expenditure. Lastly, the IRR has a relatively moderate effect on the production cost, while the oxygen profit margin has the least impact of all the parameters considered.

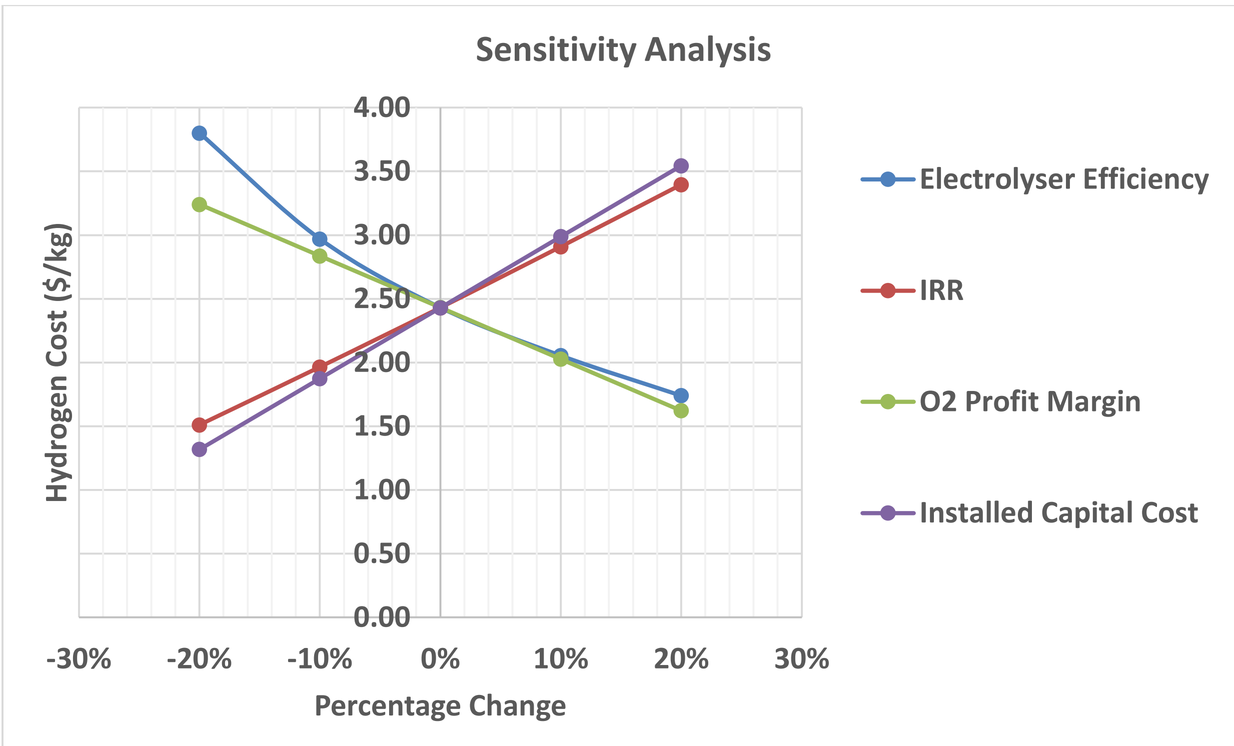


Figure 3.8: Hydrogen production costs - Sensitivities

3.4.4 Electricity Revenue

The amount of electricity revenue made from sales of energy to the grid, as a function of the electrolyser farm size, is illustrated in Figure 3.9. For the smaller electrolysers, electricity revenue is significant for a wide range of electrolyser farm sizes. This is because the surplus amount of energy available from the hydropower plant is sufficiently high enough to command meaningful revenue from the grid. In the case of larger electrolysers, electricity revenue is significant for a narrow range of electrolyser farm sizes. Intuitively, this is due to a reduced amount of surplus energy being available for sale in this context. Additionally, as a result of their increased power capacity (kW), a much smaller number of units will provide the same level of energy demand as in the case of a larger number of units for the smaller electrolysers.

The electricity revenue is also a function of the wholesale electricity (pool) price. Hence, if the value of energy in the electricity market appreciates, smaller electrolysers in particular will become more cost competitive, with the opposite being true. It is noteworthy to highlight the fact that the average annual pool price is currently experiencing a downward trend (see Figure 3.10), partly due to the growth of supply capacity exceeding demand growth in Alberta's electricity market. Thus, generally speaking, the sale of electricity as a by-product from the hydropower-hydrogen plant would not be as profitable as the current case evaluated here suggests[11], based on recent (2014) prices (see Figure 3.10). On another note, the model developed here assumes that electricity supply bids from the plant, offered into Alberta's deregulated market (merit order system), have a 100% success rate. In reality, the success of a supply bid made by the plant will be dependent on the bids of other generators in the electricity supply mix, along with demand and supply forces. That said,

[11]As stated earlier, wholesale electricity price data from 2011 was used in the model. Electricity in this period was valued relatively highly in Alberta's deregulated electricity market, with an annual average price of about $76/MWh.

the modelling of the merit order dynamics in a deregulated electricity market is beyond the current scope of work. The use of real time deregulated electricity market prices along with the 100% success rate of supply bids, is appropriate for the intended purpose of this paper. Moreover, in practice, measures such as power purchase agreements (PPA) and forward pricing mechanisms could be established with wholesale consumers of electricity, to limit exposure to the volatility and competitiveness of the deregulated electricity market.

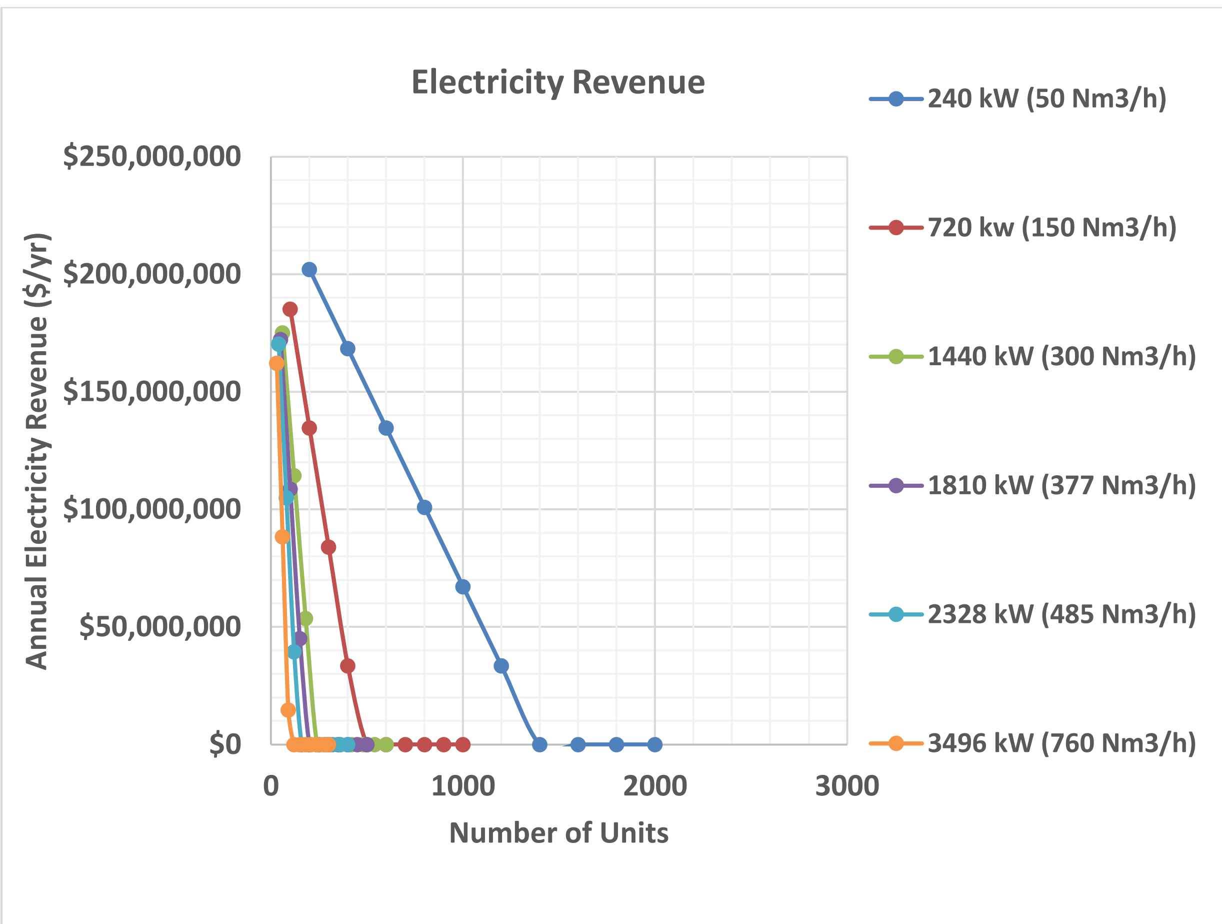


Figure 3.9: Energy sold to the electrical grid -Electricity revenue

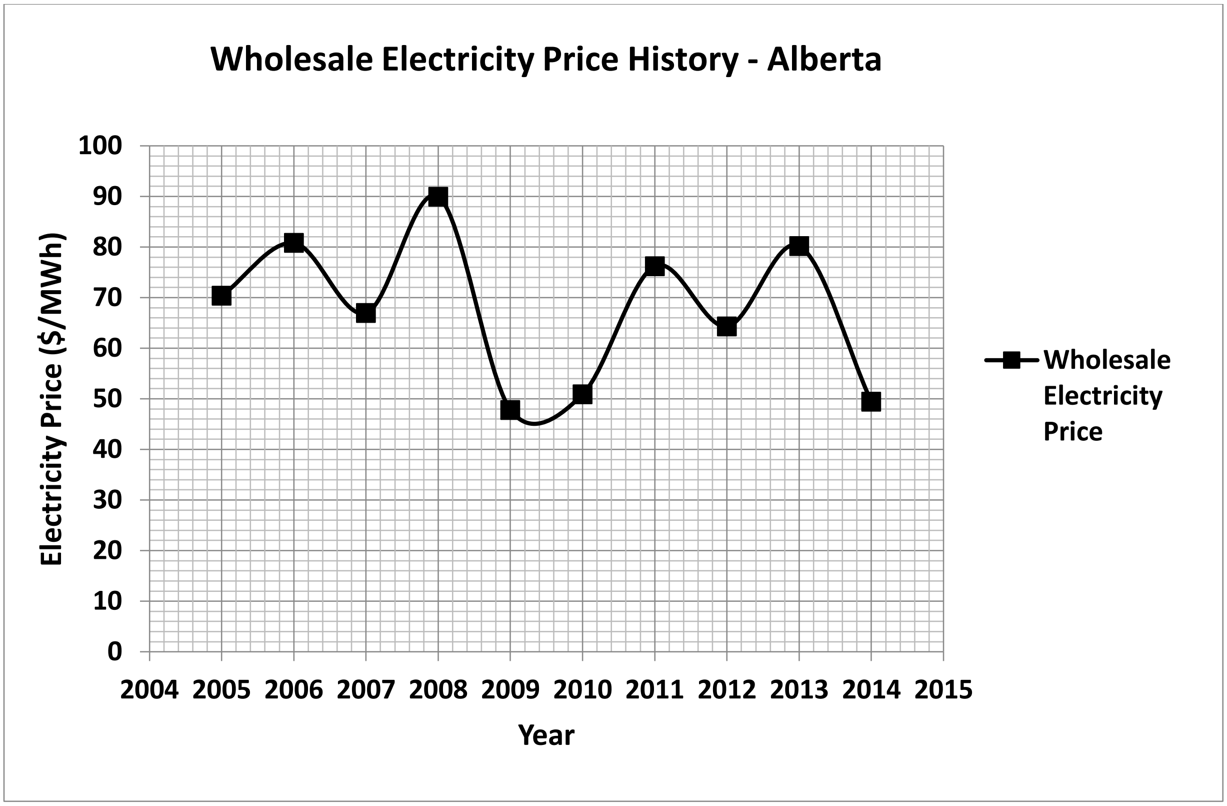


Figure 3.10: Alberta wholesale electricity (pool) price history: 2005 -2014 [47]

3.4.5 Electrolyser Capacity Factor

The electrolyser capacity factor variation with electrolyser farm size draws some parallels with the case of electricity sales (see Figure 3.11). Smaller electrolysers are able to maintain ideal (100%) capacity factors over an extended range of electrolyser farm sizes, while this range is much narrower for larger electrolysers. As the electrolyser farm size is increased, the capacity factor initially maintains an ideal value (because of the non-intermittent nature of hydropower generation). However, once the electrolyser farm size attains a magnitude such that the electrolyser demand for maximum hydrogen production supersedes the electricity supply, the capacity factor drops sharply with further increases in the electrolyser farm size. This trend is consistent for all the electrolyser sizes assessed in the model.

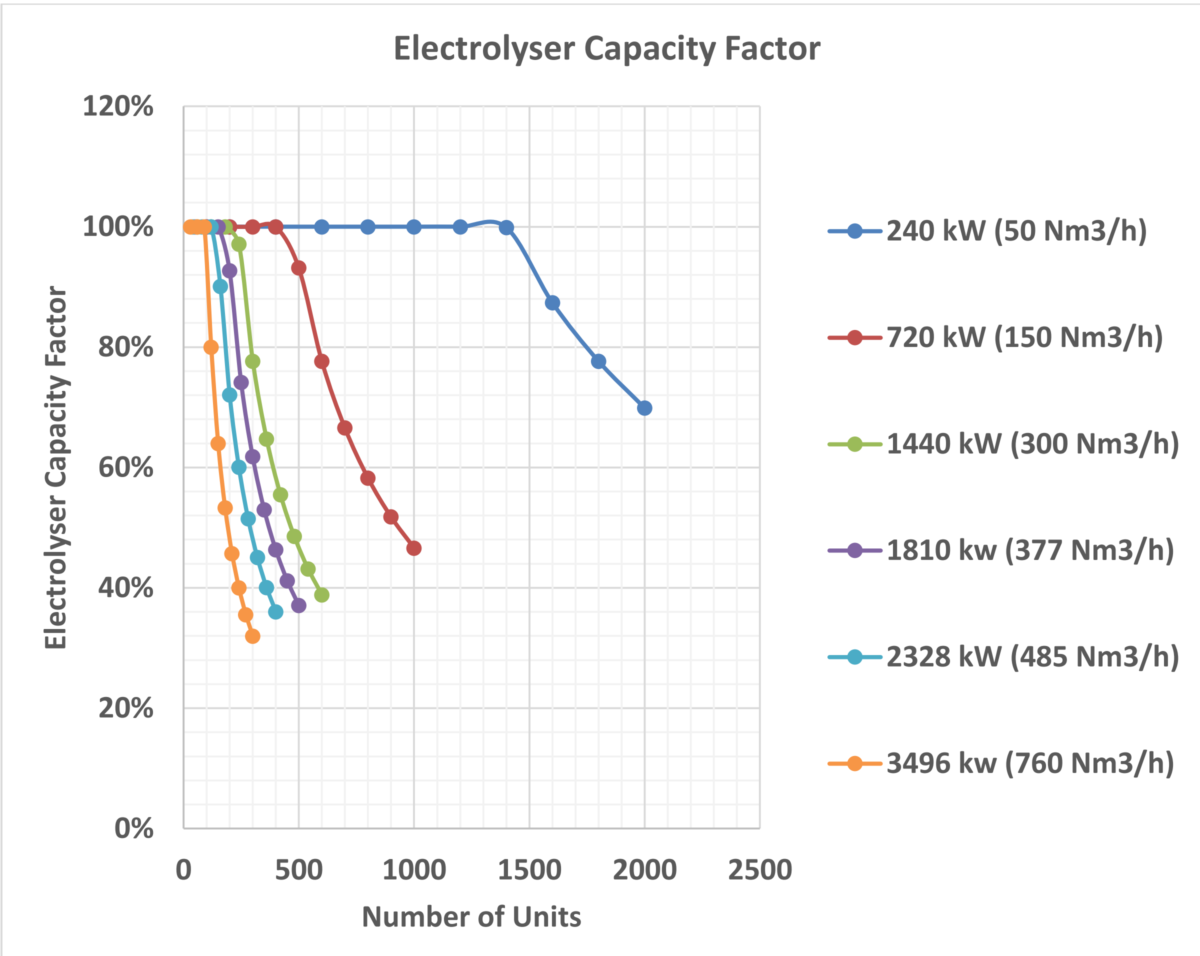


Figure 3.11: Electrolyser capacity factor

3.5 Conclusions

An integrated data-intensive techno-economic model termed FUNNEL – COST – H2 – HYDRO has been developed in this paper to estimate hydrogen production costs from a hydropower plant which operates in a liberalized electricity market. A number of electrolyser configurations (electrolyser farms) were assessed to determine the minimum cost of hydrogen production. The minimum hydrogen production cost for the hydropower-hydrogen plant amounts to \$2.43/kg H_2 – this corresponds to an electrolyser farm with 90 units of the 3496 kW (760 Nm^3/h) rated electrolyser. This cost is competitive with SMR/SMR-CCS production costs, which vary from \$1.87/kg H_2 to \$2.60/ kg H_2. This point is buttressed by the fact that if existing hydropower plants are used (hence negating hydropower capital costs), the minimum production cost amounts to \$1.18/ kg H_2.

In general, the smaller electrolysers exhibited significant electricity revenue and ideal capacity factors for a broad range of electrolyser farm sizes. However, the higher number of units and specific capital costs which these smaller sizes incur, impeded the achievement of cost efficient H_2 production costs. It is worth adding that the energy management, monitoring, operation and maintenance of relatively high numbers of electrolyser units, which pertain to smaller electrolysers, would be prohibitive in many cases. In contrast, even though their capacity factors and electricity revenue were not as extensive, larger electrolysers benefited from lower specific capital costs and a lower number of units – hence, translating into more competitive H_2 production costs.

The impact of the electrolyser efficiency and hydropower capital cost estimates on the hydrogen production cost is highly significant. Hydrogen from hydropower, under the techno-economic conditions considered here, is competitive in comparison to SMR. With the consideration of the cost of GHG emissions, the competiveness of hydropower-hydrogen against SMR will be further enhanced.

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Chapter 4[1]

Techno-Economic Assessment of Hydrogen Production from Underground Coal Gasification (UCG) in Western Canada with Carbon Capture and Sequestration for Upgrading Bitumen from Oil Sands.

4.1 Introduction

Hydrogen is a crucial feedstock to the oil sands industry as it is needed for the upgrading of bitumen to synthetic crude oil (SCO). Hydrogen demand is anticipated to reach 3.1 million tonnes/year in the industry by the year 2023 [1]; as oil sands production (a combination of SCO and non-upgraded bitumen) is expected to rise from 1.6 million bpd in 2010 to 3.95 million bpd in 2030 [2]. In Alberta, as is the case in much of the globe, hydrogen is predominantly produced via steam methane reforming (SMR) [3-5]; which has a significant life cycle greenhouse gas (GHG) emissions footprint of about 11,000 -13,000[2] tonnes of CO_2 equivalent per tonne of hydrogen produced [6-9]. Furthermore, the feedstock (i.e., natural gas) cost volatility [10, 11] and the intensity of natural gas usage in the oil sands industry as a whole (natural gas is a premium fossil fuel with a significant opportunity cost) - raise questions about the sustainability of hydrogen production via SMR especially in an increasingly competitive and carbon constrained energy market. There is a lot of interest in the production of hydrogen from conversion pathways which have low GHG footprints. As a result, research into the development of sustainable hydrogen

[1] This chapter is based upon a journal publication. Please refer to: *Olateju B., Kumar A. Techno-economic assessment of hydrogen production from underground coal gasification (UCG) in Western Canada with carbon capture and sequestration (CCS) for upgrading bitumen from oil sands. Applied Energy 2013; 111: 428-440.*
[2] Value based on the higher heating value (HHV) of hydrogen (141 MJ/kg).

production pathways, without compromising economic viability and maintaining a negligible GHG footprint is warranted to facilitate the sustainable growth of the bitumen upgrading oil sands industry in North America and elsewhere.

Underground coal gasification (UCG) is a technology that in recent times, has gained increased attention in the global energy mix; especially in the context of clean coal technologies and carbon neutral energy pathways [12, 13]. At the end of 2005, global coal reserves stood at approximately 850 billion tonnes [12]; UCG is expected to augment this figure by as much as 600 billion tonnes, an increase of about 71% [12]. In addition, some authors have estimated an increase in coal reserves in the range of 300 - 400% for particular jurisdictions such as the United States [13]. From a Canadian standpoint, the province of Alberta's proven coal reserves as of December 2014 amounted to 36.8 billion tonnes [10], with a considerable amount viable for UCG. The viability of UCG in Alberta is substantiated by the fact that it houses a UCG plant (still in its demonstration phase) with the deepest coal seam gasification depth in the world[3] (1400 metres below ground) [15].

UCG involves the in situ gasification of deep coal seams otherwise inaccessible by conventional coal mining methods, for the production of synthesis gas (syngas) which has a multitude of downstream industrial uses –hydrogen production, power generation, liquid fuels etc [13, 16, 17]. Syngas is mainly composed of hydrogen and carbon monoxide, with other species such as methane and carbon-dioxide to a lesser extent [16, 18]. UCG has a number of advantages over above ground coal gasification plants, which can potentially facilitate superior cost competitiveness. Unlike

[3] Due to the relatively low natural gas price environment that has persisted in North America, this project has been temporarily deferred. At the time of the project's deferral in 2013, the natural gas prices was below \$3/GJ. At this price, the cost of producing syngas which is a substitute feedstock for natural gas (and can be used to produce hydrogen), is uncompetitive [14]. Natural gas prices would need to rise to about \$5/GJ for the project to be economic [14].

surface gasification plants, the cost of a gasifier is mitigated, as the coal seam cavity created serves as the gasification unit. Furthermore, the cost associated with the purchase of coal, its transport and handling, along with its ash disposal are also abated with UCG [13, 16, 19]. The aforementioned economic gains associated with UCG come at the expense of reduced process control, along with productivity and environmental risks e.g. consistency of syngas quality and composition, land subsidence and ground water contamination [13, 16-18]. That being said, the aforementioned risks of ground water contamination and land subsidence are reduced with increased coal seam gasification depths as well as the management of the coal seam gasification pressure[4] [13, 16-18].

UCG lends itself as a formidable candidate for sustainable hydrogen production in Western Canada. In particular, the complementary nature of UCG and carbon capture and sequestration (CCS) technology is quite compelling. CCS technology includes the capture of CO_2 from the syngas generated through the UCG process and the injection of this CO_2 in a sink (such as underground geological storage or ocean storage) [20, 21] .UCG sites have been known to co-exist in close proximity to geological formations which are suitable for CO_2 sequestration [13, 18, 22]. A study conducted by Friedmann et al. (2009) [13] reports that greater than 75% of current, planned, and pilot stage UCG projects lay within 50 km of saline aquifers which possess the capacity for CO_2 sequestration; not to mention the added close proximity to depleted oil and gas fields which present an opportunity for CO_2 enhanced oil recovery (EOR) operations [13]. This characteristic of UCG is further verified in the Western Canadian context. The Western Canadian Sedimentary Basin (WCSB) which spans the majority of the province of Alberta is known to have

[4] Maintaining the UCG gasification cavity pressure below the surrounding hydrostatic pressure will reduce the risk of contamination. However, this has to be balanced against the effect on syngas quality e.g. calorific value, and other downstream processes which favour a higher syngas evolution pressure e.g. CO_2 capture [13, 18].

favourable CO_2 sequestration properties [23, 24]; furthermore, the vast coal resources of Alberta are also contained in the WCSB [25].

As of 2008, CCS implementation with large fossil fuel driven energy systems was the focal point of the Alberta government's strategy to reduce GHG emissions [25, 26]. The Government of Alberta (GOA) leadership at the time, aimed to reduce GHG emissions by 200 million tonnes in the year 2050 with respect to 2005 GHG emission levels – CCS was to account for about 69% (137 Mtonnes) of this reduction [25]. With the change of GOA leadership in 2015, the province's strategy towards achieving GHG emissions reduction appears to be taking a different route. The complete phase out of coal-fired power plants in Alberta's electricity sector by 2030, an economy-wide carbon price of \$30/tonne CO_2e by 2018, and an oil sands industry-wide annual GHG emissions cap of 100 Mtonnes CO_2e (with provisions for cogeneration and new upgrading capacity) are some notable policy changes that have been announced [27] . The phase out of coal power plants has the potential to incentivize alternative uses for coal in different sectors such as hydrogen production in the bitumen upgrading industry. Moreover, the increase in the price of CO_2 enhances the likelihood of CCS technology being competitive in the energy market, at least from a purely economic standpoint. While the new energy policy environment in Alberta is yet to be fully defined, it is likely that the potential of UCG-CCS to produce hydrogen in a low carbon future remains uncompromised.

The synergy that can be realised from hydrogen production via UCG-CCS in the Western Canadian context, deserves research attention, as the conditions that exist in terms of the geology and resource wealth facilitate a fertile ground for its implementation. The published literature on the integration of CCS with large scale energy systems is in-depth and multi-faceted in nature;

with technological, economic, environmental, and regulatory aspects being addressed exhaustively [20, 28-34]. Contrastingly, in the case of UCG, much of the focus in the existing literature has been geared towards UCG process simulation and optimisation; as well as environmental impact monitoring [35-40]. While this is undoubtedly important, the appraisal of UCG-CCS from a techno-economic perspective in published literature is scarce. More often than not, the appraisal of UCG in a techno-economic context is qualitative with limited detail [13, 16, 18] . Admittedly, this is likely due to the infancy of the technology and the limited operational experience on a commercial large scale. The above ground (surface) gasification of coal and the subsequent processing of the syngas evolved for hydrogen production is a mature well understood technology. The process methodology for syngas-H_2 conversion is identical for both surface gasification and UCG plants. In both cases syngas is processed above ground, and the syngas composition, temperature and pressure are of similar magnitudes [13, 16] . As a result, this enables the techno-economic modelling of UCG to be carried out credibly, within reason. The techno-economic modelling of UCG-CCS with the explicit consideration of its apparent cost-competitiveness and environmental risks is needed to provide a quantitative and qualitative view of its utility as a hydrogen production pathway in comparison to conventional methods such as SMR-CCS. In addition, this will also facilitate the identification of areas of cost minimisation and key sensitivities. Furthermore, the techno-economic insight gained has the potential to enhance the galvanisation of investor interest both from industry and governmental bodies.

As a result, the primary objective of this paper is the development of a data-intensive techno-economic model for UCG-CCS, which will yield a credible estimate of the cost of hydrogen production in a Western Canadian context. For comparative reasons, a SMR-CCS techno-economic model was also developed in similar fashion, to yield the cost of hydrogen production.

The techno-economic models developed in this research are partly based upon existing above ground coal gasification[5] and SMR plant models provided in earlier studies [41-43]. Adjustments and refining of the model architecture used [41-43], as well as the modification of data inputs were carried out to ensure specificity to Western Canadian conditions as reasonably possible. All costs specified in this paper are in 2010 Canadian dollars[6]. Seven different hydrogen production scenarios were assessed in terms of the cost of hydrogen production. These scenarios included the cost assessment of hydrogen production with CCS and without CCS. The details on these scenarios are given below.

4.2 Hydrogen Production with CCS – Scenarios in Western Canada

A multitude of practical and viable scenarios for hydrogen production with the added feature of CCS exists in Western Canada. To account for these various options in the implementation of a SMR-CCS plant and UCG-CCS plant, seven scenarios are considered in this study (see Table 4.1). For all the scenarios considered, the SMR-CCS plant location is in Fort-Saskatchewan Alberta, as this is an industrial heartland of the province suitable for plants of this nature (Figure 4.1 shows the location of Fort-Saskatchewan in Alberta). In similar fashion, the UCG-CCS plant is located in Swan Hills Alberta for all the scenarios considered (as shown in Figure 4.1). The distinction associated with each scenario relates to the location of the CO_2 sequestration site, and consequently, the length of the CO_2 pipeline**.** Also, the fate of the CO_2 evolved at the plant i.e.

[5]The above ground coal gasification model is comprised of a two-part study [41, 42].
[6]An inflation rate of 2.5% has been used to convert all currencies into 2010 $CAD. In addition, an exchange rate of $US 0.8 = $CDN 1 has been utilized in this study.

whether captured and sequestered, released into the atmosphere, or captured and sold for revenue, is another distinction.

It is important to stress that scenarios 1 and 4 represent the baseline scenarios for SMR-CCS and UCG-CCS, respectively. As a result, all the subsequent analysis conducted in this study is based on the plant configurations in these scenarios unless otherwise specified.

Table 4.1: Scenarios of hydrogen production with CCS

Scenarios	SMR/UCG based H_2 production	With/Without CCS	Description	Assumptions/Comments
Scenario 1	SMR	With CCS	H_2 production at Fort Saskatchewan, Alberta; with CO_2 sequestration in Thorhild, Alberta via 84 km CO_2 pipeline.	Based on CO_2 sequestration location and pipeline distance for the Shell Quest CCS project [44].
Scenario 2	SMR	With CCS	H_2 production at Fort Saskatchewan, Alberta; with CO_2 sequestration in Swan Hills, Alberta via 225 km CO_2 pipeline.	Based on the premise that the CO_2 sequestration reservoir is located within a 10 km radius of the UCG coal resource in Swan Hills, Alberta. A high potential for spatial co-location of UCG coal resources and CCS reservoirs is highlighted by [13, 18, 22]. Driving distance used to estimate pipeline length [45].
Scenario 3	SMR	Without CCS	H_2 production at Fort Saskatchewan, Alberta.	
Scenario 4	UCG	With CCS	H_2 production at Swan Hills, Alberta with H_2 delivery to Fort-Saskatchewan, Alberta via 225 km H_2 pipeline; along with CO_2 sequestration within a 10 km radius of the UCG plant.	Based on high spatial coincidence between UCG coal resources and CCS reservoirs (see scenario 2 assumptions). Driving distance used to estimate pipeline length [45].
Scenario 5	UCG	With CCS	H_2 production at Swan Hills, Alberta with H_2 delivery to Fort-Saskatchewan, Alberta via 225 km H_2 pipeline; along	Based on the premise that the high spatial coincidence does not hold true. Hence, CO_2 sequestration reservoir is located a relatively large distance away from the

Scenarios	SMR/UCG based H_2 production	With/Without CCS	Description	Assumptions/Comments
			with CO_2 sequestration in Thorhild, Alberta via 184 km CO_2 pipeline.	UCG plant. Driving distance used to estimate pipeline length [45].
Scenario 6	UCG	Without CCS	H_2 production at Swan Hills, Alberta with H_2 delivery (via 225 km H_2 pipeline) to Fort-Saskatchewan, Alberta	Driving distance used to estimate pipeline length [45].
Scenario 7	UCG	With CCS	H_2 production at Swan Hills, Alberta with H_2 delivery (via 225 km H_2 pipeline) to Fort-Saskatchewan, Alberta; along with the sale of CO_2 for enhanced oil recovery (EOR) operations.	EOR operators are assumed to be within a 10 km radius of the UCG plant. As of 2009, Alberta had an EOR storage capacity estimate of about 450 million tonnes [26].

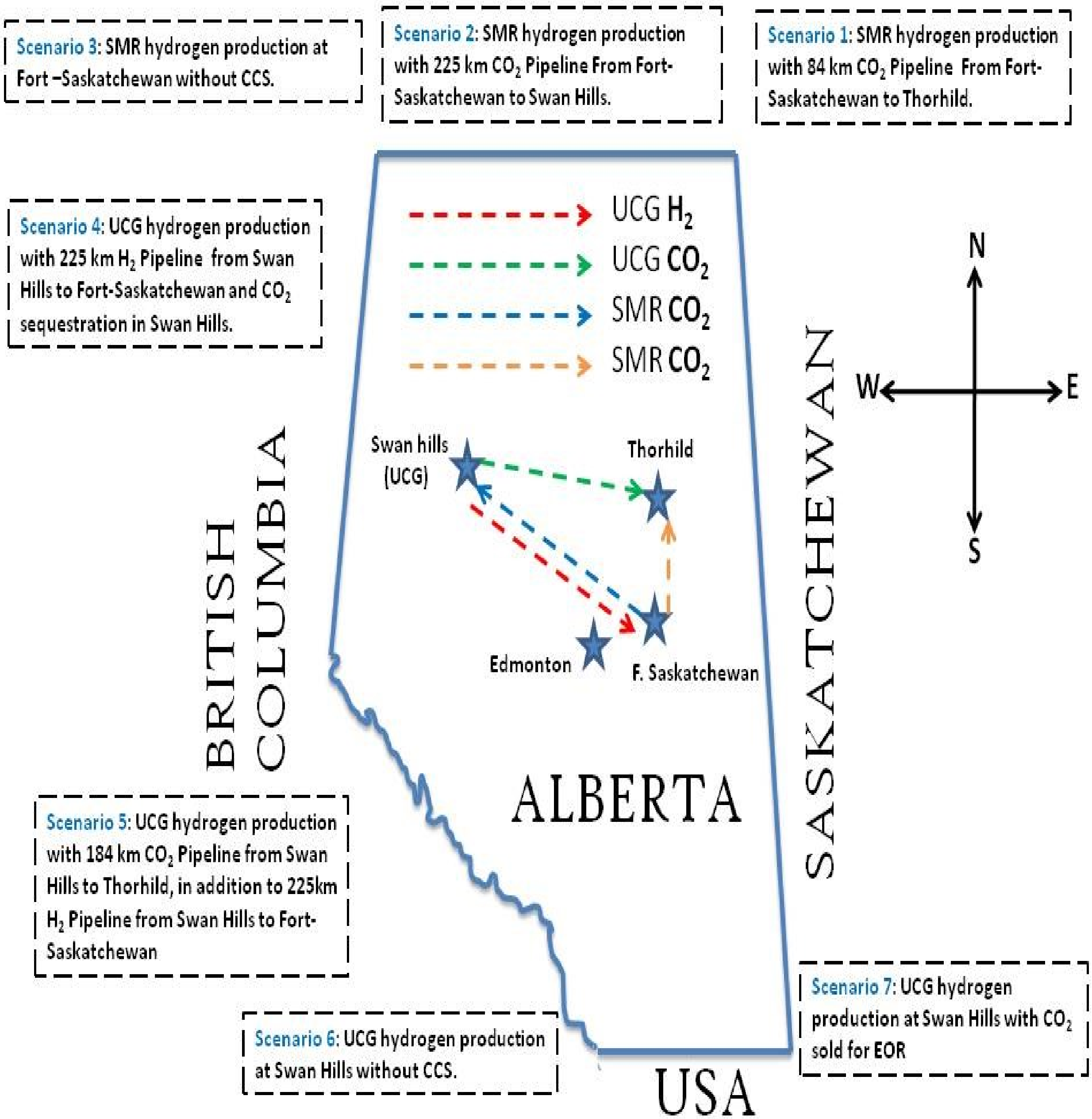

Figure 4.1: Geographical depiction of Hydrogen Production with CCS Scenarios

4.3 Hydrogen Production with UCG-CCS

4.3.1 UCG Technological Overview

Underground coal gasification involves the sub-surface gasification of a coal seam, with the use of oxygen/air and steam at elevated temperatures for the in-situ production of synthesis gas (syngas) [16-18]. The coal seam is accessed via the drilling of an injection well, and the syngas produced is channelled up a production well as shown in Figure 4.2. Advances in drilling technology have enabled the gasification of coal seams at depths previously considered as impractical and economically prohibitive. Prime examples of the advances in technology that enable access to deep coal seams and the (limited) control of the UCG process include the continuous retraction injection point (CRIP) process and the proprietary εUCG process[7]. The CRIP process was developed by Lawrence Livermore Laboratories over two decades ago [16]. The CRIP process involves the vertical drilling of a production well, and then the use of directional drilling to produce an injection well that connects to the production well [16]. Once a channel between the production and injection wells is established, a gasification cavity is created at the end of the injection well in the horizontal section of the coal seam, with the introduction of the gasification agents [16] (see Figure 4.2). As the gasification process proceeds, the coal seam in the cavity initially created will eventually be exhausted. The process is given continuity by the retraction of the injection point –achieved (preferably) with the burning of a section of its liner [16]. This creates a new gasification cavity which allows the continued production of syngas until the coal seam region accessed is used up.

[7] The εUCG process was developed by Ergo Exergy Inc.

A few competing factors and characteristics are associated with UCG which are worth highlighting. First, the coal seam thickness is a key parameter for the viability of a given UCG project. A coal seam thickness of two meters is considered to be the minimum threshold for economic viability [17]. The coal seam[8] (Mannville coal) to be utilized in the proposed UCG plant has been characterised with a seam thickness varying from 6-10 m [46]. This further demonstrates the quality of the resource that exists in Alberta.

Secondly, a balance between the amount of oxygen and steam introduced into the coal seam cavity must be maintained. Increased oxygen supply improves the concentration of CO_2 in the syngas stream, thus facilitating the ease of CO_2 capture downstream [17]. However, the increased CO_2 content decreases the calorific value of the syngas as CO_2 is inert [17]. Considering that the desired end use of the syngas in this study is hydrogen production with CCS, as opposed to power generation, there is likely to be a bias in the amount of oxygen added in practice. The addition of steam on the other hand increases the calorific value of the syngas; however, a surplus amount of steam can potentially extinguish the gasification process [17].

The permeability of the coal seam is also important as it facilitates the increased flow of the syngas produced [13, 17]. Coal seams with high permeability are desirable; however, this is seldom found in practice [17]. As a result, a coal resource of low-calorific value is often seen as ideal for UCG in the context of syngas flow. This is because these low grade coal types usually shrink upon

[8] In the techno-economic model developed, the characteristics of Illinois #6 coal in terms of its composition and calorific value have been adopted [41]. That being said, the caloric value of the coal utilized has a negligible effect on the model accuracy, as the coal feedstock cost utilized in this study is nil (see Table 4.3). In addition, the syngas $\rightarrow H_2$ & CO_2 conversion pathway for Illinois #6 coal provides a suitable benchmark for estimating the anticipated cost of UCG based hydrogen from Mannville coal in Western Canada. The syngas composition of both coal types is expected to be in the same order of magnitude.

gasification as opposed to expanding, which aids syngas flow [16, 17]. Mannville coal is a sub-bituminous coal with an average caloric value of 28.5[9]MJ/kg [47].

Another influential parameter is the pressure of the coal seam, which invariably dictates the pressure of the gasification process. Apart from influencing the reactions that take place in the coal seam, the partial pressure of CO_2 is also an important parameter for capture [30]. Higher partial pressures are desired for CO_2 capture [30]; consequently, a high gasification pressure is also ideal. The coal seam pressure reported by Swan Hills Synfuels for their Mannville coal seam is about 13 MPa [48].

[9] The average calorific value of Mannville coal considering a range of 26.8-30.2MJ/kg was utilized [47]. As a comparison, the calorific value of Illinois #6 coal amounts to 26.14 MJ/kg [41].

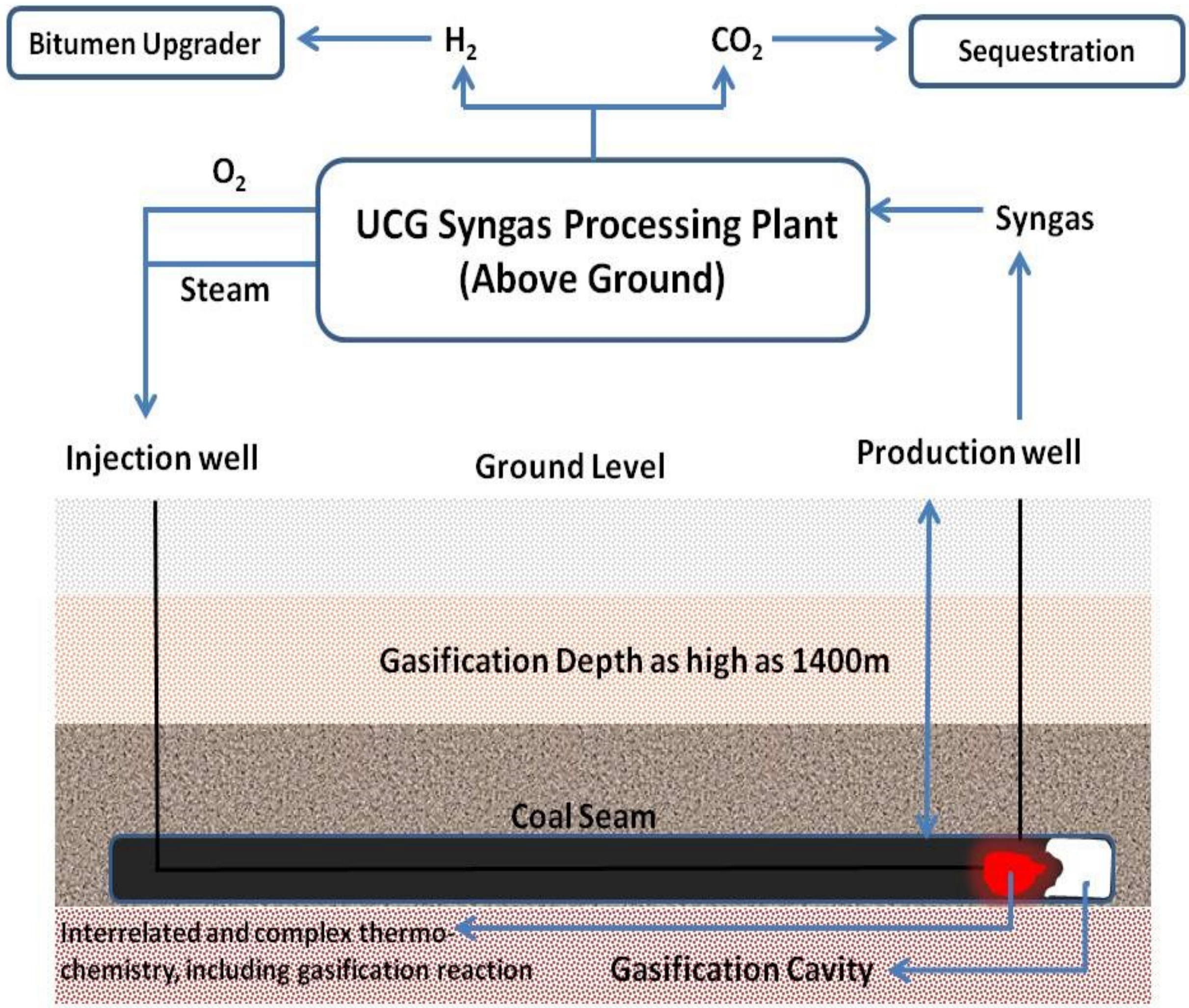

Figure 4.2: Schematic diagram of UCG-CCS

Notes: UCG involves a multitude of intricate reactions [16]; including:

(1) Gasification Reaction : $C + H_2O \rightarrow H_2 + CO$

(2) Shift Conversion: $CO + H_2O \rightarrow H_2 + CO_2$

(3) Methanation: $CO + 3H_2 \rightarrow CH_4 + H_2O$

(4) Hydrogenating Gasification: $C + 2H_2 \rightarrow CH_4$

(5) Partial Oxidation: $C + 0.5\ O_2 \rightarrow CO$

(6) Oxidation: $C + O_2 \rightarrow CO_2$

(7) Boudouard Reaction: $C + CO_2 \rightarrow 2CO$

4.3.2 UCG –CCS Plant Model

An above surface coal gasification plant model developed in earlier studies [41, 42] is utilized in this study as an approximation of the UCG-CCS process. The model was modified wherever required to replicate conditions specific to UCG.

The plant proposed in earlier studies [41, 42] has a number of configurations ranging from the production of mainly hydrogen or electricity, syngas cooling methods, and the venting or capture of the CO_2 produced. The configuration where mainly hydrogen is produced along with carbon capture is of particular relevance to this study. The plant configuration adopted from literature also involves the gasification of coal at a pressure of 12 MPa in an entrained flow gasifier with 0.5 kg of warm water consumed per kilogram of coal [41]. A gasification pressure of 12 MPa is a reasonable approximation, given that Mannville coal seam pressures have been reported to be in the range of 9-13 MPa [48, 49]. The warm water usage is however not applicable to UCG which requires steam. In this study, it is assumed that 0.5 kg of steam at an elevated temperature of 1227^0C is required per kilogram of coal. The injection of the oxygen and steam mixture for UCG is reported to be at a temperature of 1227^0C [17]. It is assumed in this study that the heat content of this mixture is solely due to steam. The energy used to raise steam at the aforementioned temperature is assumed to be provided by Mannville coal with an efficiency of 70% and a price of $0/tonne. In reality, this additional energy will likely be provided by the syngas evolved during UCG, and would mitigate the need to utilize coal. However, due to a scarcity of data on UCG syngas properties in Alberta, this assumption was made, as its effect on production cost is likely to be insignificant. Table 4.2 gives a summary of the UCG-CCS plant characteristics along with the bitumen upgrader plant specification. Some of the major base case assumptions of the study and input data are shown in Table 4.3 below.

Table 4.2: UCG-CCS plant specification

Plant Specifications	**Value**	**Sources/Comments**
UCG plant design capacity	660,000 kg H_2 /day	Determined based on a plant H_2 output of 1052.4 MW_{TH} [41]. A H_2 LHV of 119.96MJ/kg was adopted.
Plant capacity factor	85%	[43]
Coal consumption	5,574[a] tonnes/day	Determined using a Mannville coal calorific value of 28.5MJ/kg [47].
Water (steam) consumption	0.5kg/kg coal	[41]
UCG coal to H_2 efficiency	50%	Comparable to the range of values (51%-63%) estimated by a UCG simulation model [50]
Hydrogen production pressure	60 bar	[41]
CO_2 flow rate	11,302 tonnes/day	Based on a CO_2 flow rate of 554 tonnes/h [42], along with the consideration of the plant capacity factor.
Bitumen Upgrader Specification		
Shell Scotford Upgrader Capacity	290,000 bpd	Shell Canada has obtained approval from the Alberta Energy and Utilities Board for expansion of its upgrader to a capacity of 290,000 bpd [51].
Synthetic crude oil (SCO) production	246,500 bpd	A bitumen to SCO volumetric conversion ratio of 0.85 is utilised [52, 53]
H_2 consumption for SCO production	3.4 kg/barrel	An average of hydrogen consumption for a multitude of oil sands extraction and upgrading technologies is assumed [54]
H_2 Supply Required	838,100 kg H_2 /day	

[a] Includes the additional coal consumption used to raise steam.

Table 4.3: UCG-CCS model principal economic data

Parameter	Value	Sources/Comments
Base case coal price	\$0/tonne	Coal feedstock cost is considered to be negligible – due to in situ use of coal, zero feedstock transportation and handling costs.
Electricity cost	\$0.07/kWh	Average cost of electricity in Alberta utilized [55]
Inflation	2.5%	
Base case internal rate of return (IRR)	15%	An increased IRR for UCG relative to SMR is reflective of the technological infancy of UCG on a large commercial scale as well as its increased environmental and production reliability risks in comparison to SMR.
Plant equipment O&M factor	4% of Capital Cost	[56]
Albertan installation factor	1.65	The harsh Albertan climate and labour shortage warrants increased installation costs. Installation factor adopted is 15% higher than that for a North-American large scale fossil fuel plant proposed by Ogden (2004) [56]
Number of plant operators	8	Number of plant operators required is assumed to be equivalent to the SMR-CCS plant (see section 4.5).
Supervision and administration cost	\$1,300,000	Estimated to be 80% of annual operator labour cost [43].
Plant lifetime	40[10] years	
Financial year	2010	

[10]The only existing commercial scale UCG plant (Linc Energy's Yerostigaz plant located in Angren, Uzbekistan) has been in operation since 1961[57] . This plant is expected to maintain commercial operation for another 50 years [57]. However, a more conservative approach regarding plant lifetime estimation, which is reflective of typical coal-fired power plants, is adopted in this study.

4.4 Cost Estimation of UCG-CCS in Western Canada

4.4.1 Plant Equipment Capital Cost

The capital costs of the plant equipment for the UCG-CCS plant in this study are primarily based on the costs specified for the above ground coal gasification plant [40]. As mentioned earlier, the high degree of similarity between UCG and above ground gasification enables the reasonable estimation of the costs associated with UCG especially considering above ground plant infrastructure. The costs associated with the CCS element of the plant are estimated based on available data in literature, CCS models, and data derived from industry. The UCG-CCS plant equipment capital costs are listed in Table 4.4 (all costs are rounded up to the nearest million).

Table 4.4: UCG-CCS capital costs

Equipment	Cost ($CAD Millions)	Sources/Comments
Drilling of injection and production wells	68	[58][a]
Air separation unit (ASU)	142	[41][b]
O_2 compressor	29	[41][b]
Syngas quenching	231	[41][b]
WGS reactors & heat exchangers	92	[41][b]
Selexol H_2S removal and stripping	125	[41][b]
Sulphur recovery (Claus SCOT)	85	[41][b]
Selexol CO_2 absorption and stripping	88	[41][b]
PSA unit	34	[41][b]
PSA purge gas compressor	13	[41][b]
Siemens V64.3A gas turbine	47	[41][b]
Heat recovery steam generator (HRSG)	35	[41][b]
Steam turbine and condenser	84	[41][b]
CO_2 drying and compression	154	[59][c,d]
CO_2 pipeline	9	[59][c,d]
CO_2 injection and sequestration	9	[60][e]
Hydrogen pipeline	325	[61][d,f]
Total	1, 569	Estimate is comparable to the Swan Hills

Equipment	Cost ($CAD Millions)	Sources/Comments
		Synfuels project cost of $CAD 1.5 billion [62].

[a]Drilling cost obtained from TEXYN Inc, was provided by Williams & Heidrick (2009) [58]

[b]Cost quoted includes installation, apportioned balance of plant (BOP) and general facilities, engineering and process/project contingencies [42]; as a result, the installation factor was not applied to these costs to avoid exaggerated estimates.

[c]Cost was estimated using models provided by McCollum & Ogden (2006) [59].

[d]Installation factor applied.

[e]Cost is assumed to be equivalent to the CO_2 pipeline capital cost [60].

[f]Estimated using hydrogen pipeline cost model provided by [61].

4.4.2 Hydrogen Storage Cost

The requirement of hydrogen storage is not crucial as in the case of SMR. This is due to hydrogen production in the case of UCG not being captive[11]. The hydrogen produced will be pipelined to the bitumen upgrader as opposed to being consumed 'in-house' as is in the case of the SMR-CCS plant. Furthermore, in a UCG plant, the syngas produced is more likely to be stored as opposed to the storage of hydrogen. As a result, this study assumes hydrogen storage cost, if at all applicable, will be incurred by the bitumen upgrader in Fort-Saskatchewan.

4.4.3 Hydrogen Compression Cost

In the model developed for this study, hydrogen exits the pressure swing adsorption (PSA) unit at a pressure of 60 bar [41]. Thus, additional compression for pipeline transport is unwarranted. Hence, the cost of purchasing a hydrogen compressor is mitigated for the UCG-CCS plant.

4.4.4 CO_2 Compressor and Pump Cost

The compression of CO_2 to its supercritical phase is not just favorable for sequestration but also for its pipeline transport. CO_2 in its supercritical state has a viscosity about a hundred times lower than its liquid state [63]. This reduces frictional losses, which help to reduce fluid pressure losses in the pipeline. Furthermore, in its supercritical state, it exhibits the density of a fluid and thus allows for a higher throughput (mass flow rate) than would be the case in its gaseous form [63]. CO_2 is usually compressed to a pressure of about 150 bar for its supercritical pipeline transport

[11] Captive hydrogen production means that the hydrogen is consumed in the same place where it is produced.

[59, 63]; with a limitation on the maximum pressure level to ensure the structural integrity of the pipeline e.g. flanges [63].

A compressor is required to compress the gaseous phase CO_2 captured at the plant to its critical pressure of 73.8 bar[12] [59]. Once this critical pressure is achieved, the CO_2 in its 'dense' supercritical phase is then compressed using a pump to its final pressure of 150 bar [59]. Using models developed earlier [59] – see section 3 of Appendix, the cost of CO_2 compression and pumping applicable to both UCG and SMR is illustrated in Figures 4.3 and 4.4.

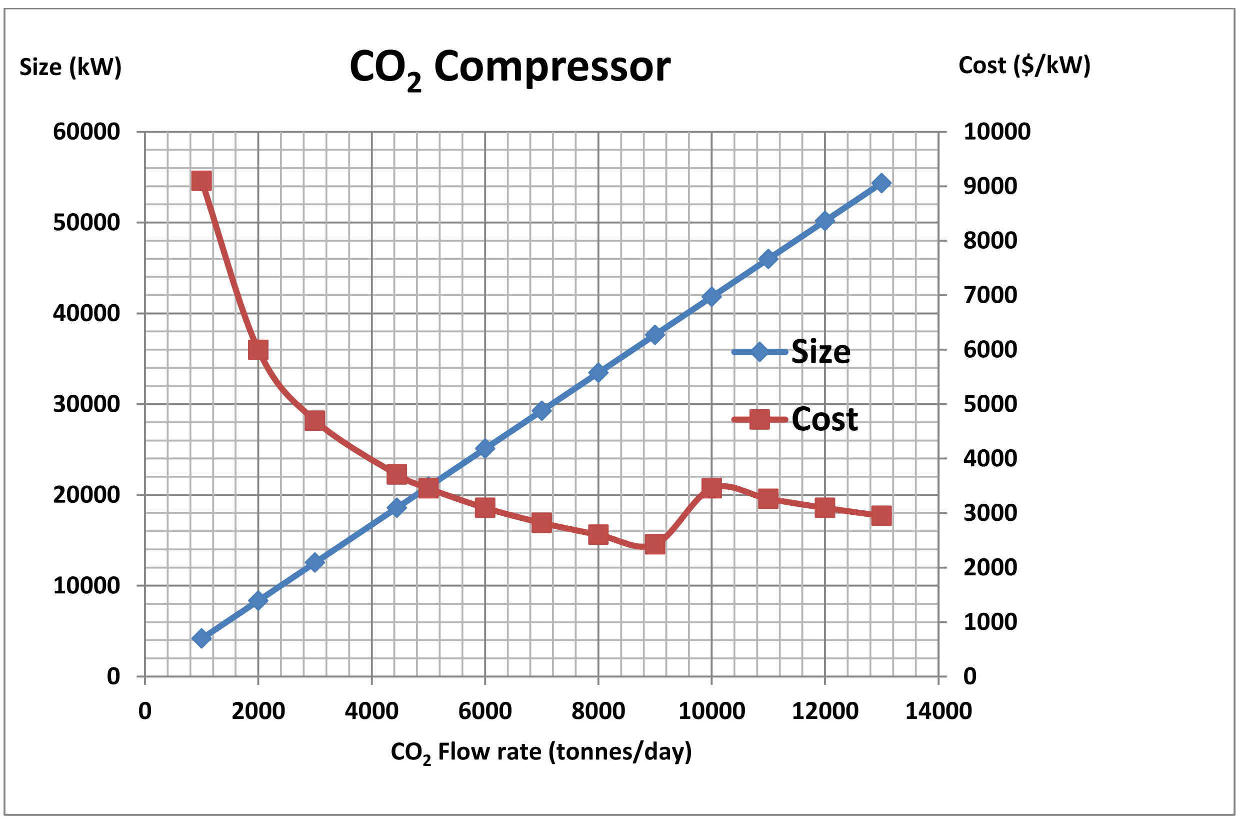

Figure 4.3: CO_2 compressor capital cost

[12] CO_2 has a critical pressure and temperature of 73.8 bar and 31.1 ^{0}C.

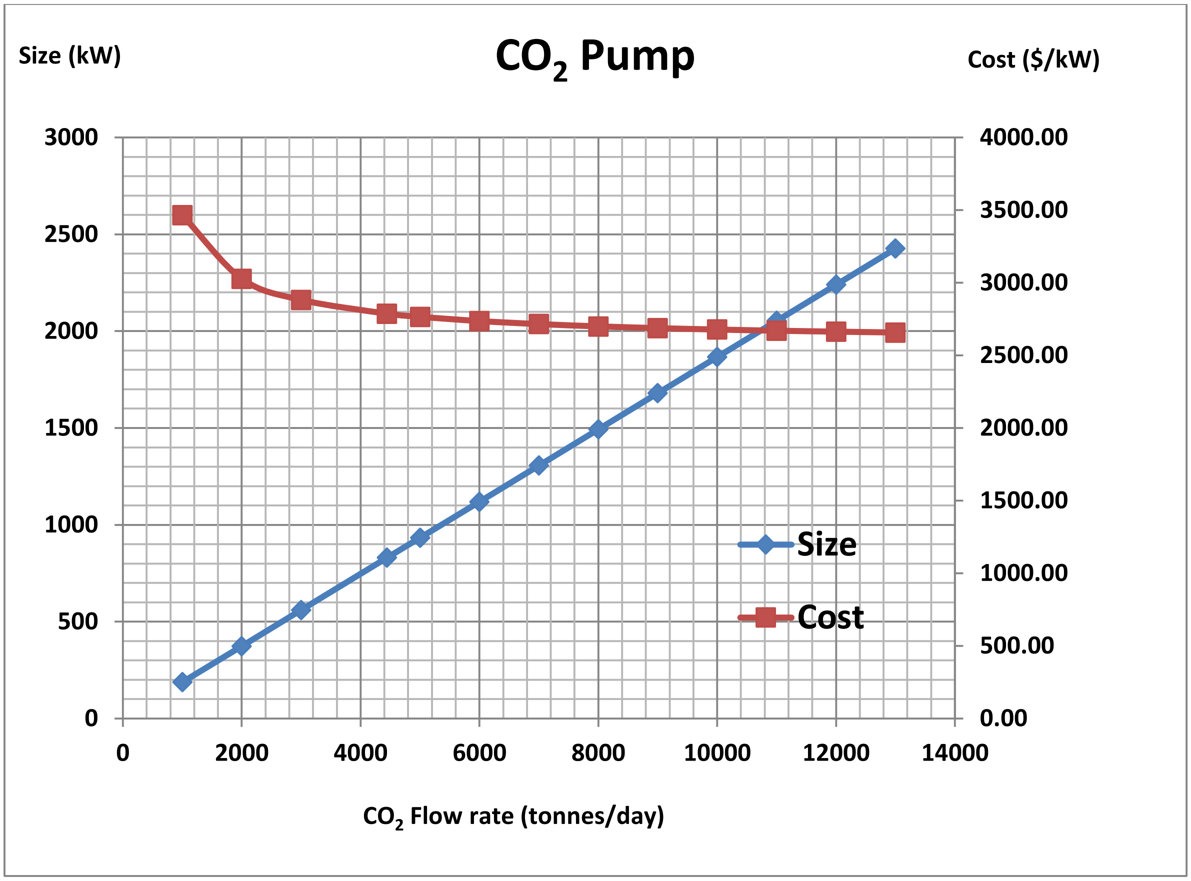

Figure 4.4: CO_2 pump capital cost

4.4.5 CO_2 Pipeline Characterization and Costs

For the UCG-CCS plant CO_2 flow rate of 11,302 tonnes/day [42], a pipeline diameter of 8.5 inches was determined using the iterative CO_2 pipeline model provided in an earlier study[13] [59] – see section 4 of Appendix. Recall that this diameter was determined for a 10 km CO_2 pipeline as stipulated in scenario 4 (see section 2). The CO_2 pipeline model adopted in this study stems from the average of seven other credible CO_2 pipeline models [56, 64-69] which were compared on a consistent model input basis [59]. Thus, the model is assumed to be realistic and fit for purpose for the characterisation of the pipeline along with the estimation of costs. The pipeline diameters presented in this study are given in their actual values without rounding up/down to nominal pipeline sizes so as to illustrate their effect on estimated costs. In actuality, the pipeline diameters will be increased when necessary during manufacture, to conform to nominal pipeline standards. Figure 4.5 illustrates the variation of the CO_2 pipeline cost with the pipeline flow rate and distance.

[13] The CO_2 pipeline characterization and cost for the SMR-CCS plant was determined in identical fashion.

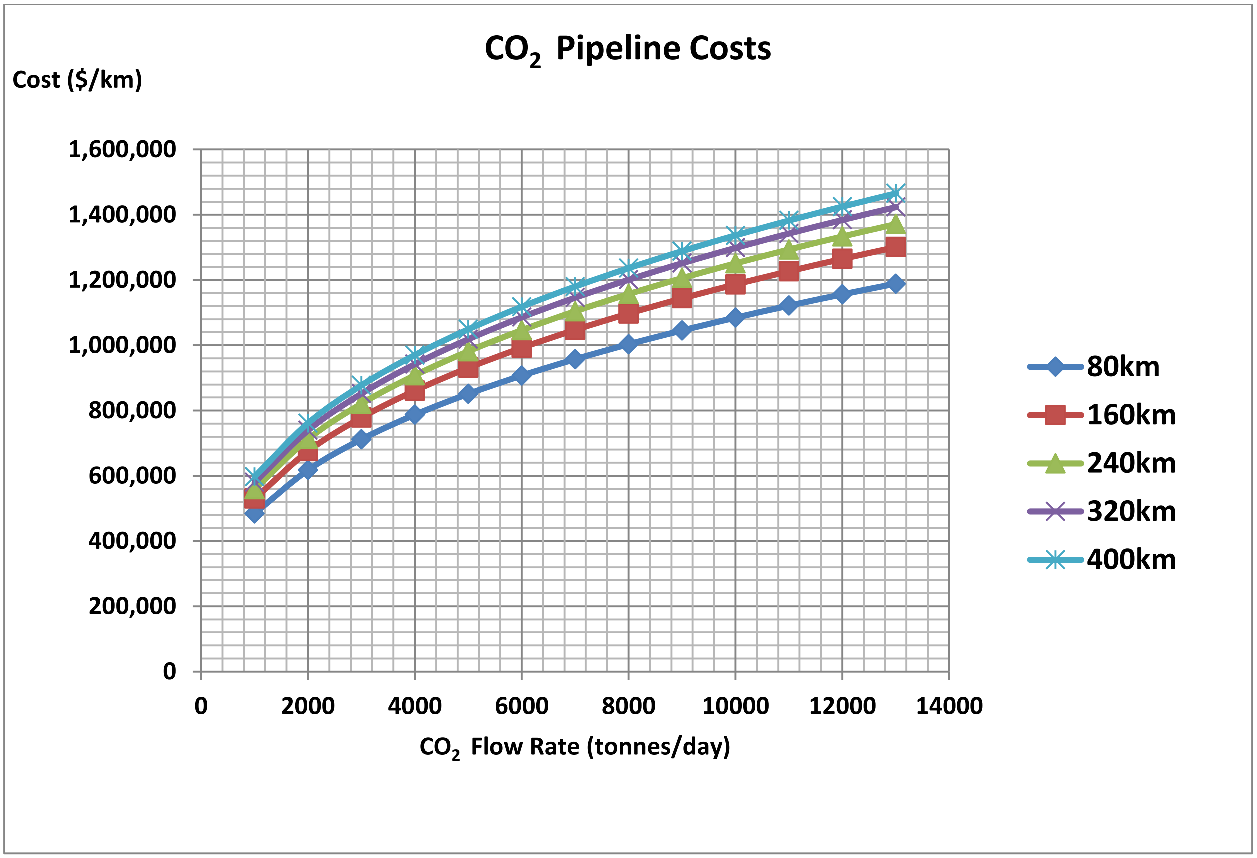

Figure 4.5: CO_2 pipeline capital cost

4.4.6 H_2 Pipeline Characterization and Costs

The characterization of the hydrogen pipeline required the determination of two principal pipeline parameters i.e. the pipeline diameter and pipeline length. The diameter of the hydrogen pipeline was calculated with the use of the Panhandle – B equation [70] (see section 4 of Appendix). This equation was solved with a reverse engineering approach to obtain the required diameter for the plant's hydrogen flow rate of 660 tonnes/day. As mentioned in preceding sections, the pipeline length was estimated with the use of the driving distance between Swan Hills and Fort-Saskatchewan [45]. The hydrogen pipeline cost model utilized in this study is based on a model provided in literature [61] (see Figure 4.6). The cost estimated using this model was benchmarked against alternative hydrogen pipeline capital cost estimation models [69, 71]. The difference in the pipeline capital cost between the model adopted in this paper and the models presented by Parker (2004) [69] and Johnson & Ogden (2012) [71] were 10% and 18% respectively. Similar to all pipelines, hydrogen pipeline capital costs will be highly site specific, and the consideration of the special pipeline seals required for hydrogen transport and the possible embrittlement of steel will play a significant role in determining costs [69]. In general, there is an increased risk of pipeline operation associated with hydrogen pipelines in comparison to other industrial fluids (e.g. CO_2, natural gas etc) which has to be factored into the cost estimates for improved accuracy. It is worth pointing that in 2010, the Government of Alberta approved the construction of a hydrogen pipeline by Air Products Inc. to transport hydrogen from the company's two production facilities to bitumen upgraders, refineries and chemical processors etc [72].

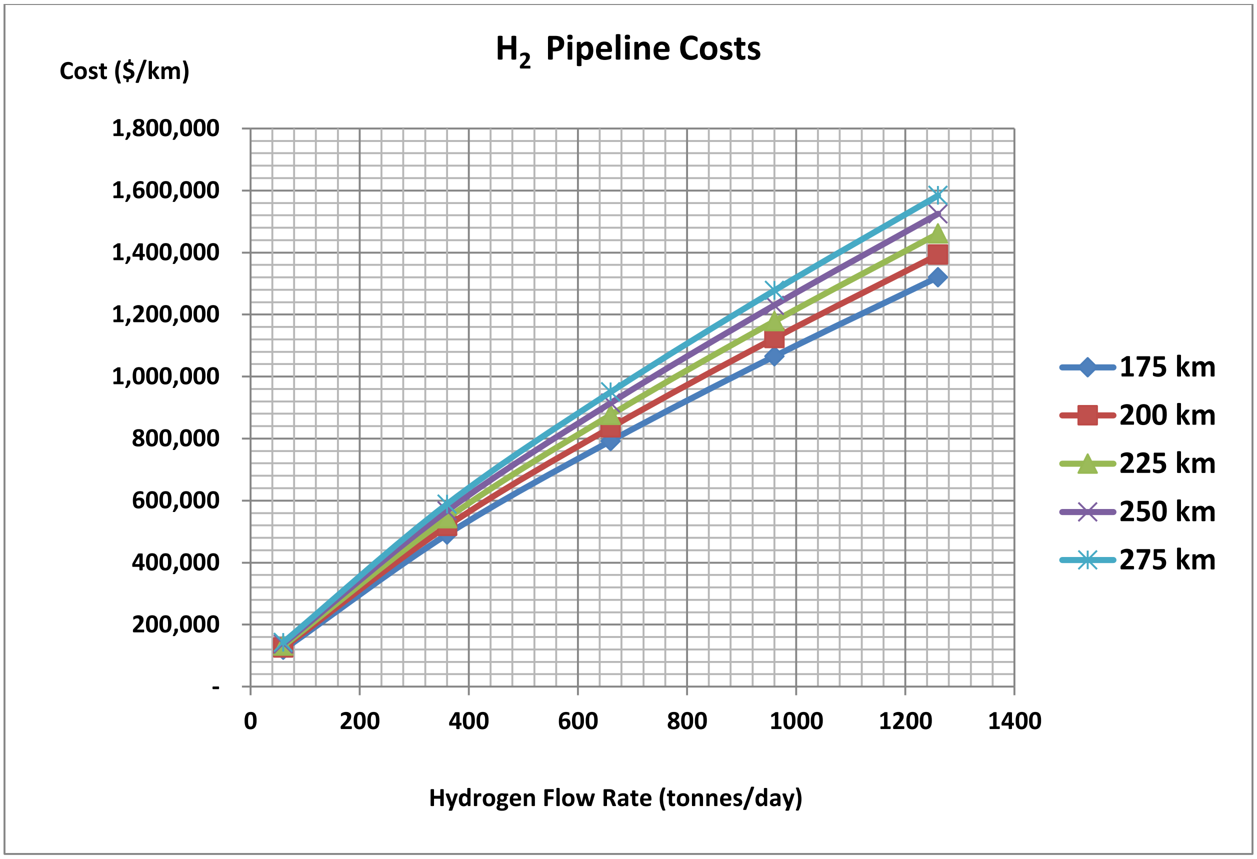

Figure 4.6: H_2 pipeline capital cost

4.4.7 H_2 Production Cost

A data intensive discounted cash flow (DCF) spreadsheet model of the UCG-CCS plant was developed to ascertain the hydrogen production cost for each scenario considered in this study. Having completed the sizing and cost estimation of the required equipment and processes, the use of the economic data and assumptions along with the plant specification data (see Tables 4.2 and 4.3 in the case of UCG-CCS) allowed for the deduction of the production cost via a DCF model. It is worth pointing out that the production cost for the SMR-CCS plant was determined with identical methodology.

4.5 Hydrogen Production with SMR-CCS in Western Canada

4.5.1 SMR Technological Overview

Steam methane reforming is a mature well understood technology in industry that accounts for the majority of hydrogen production in Alberta and the globe. For detailed technological insight into SMR and its use with CCS, the reader is referred to the work carried out by the International Energy Agency (IEA) [43] and the literature authored by Molburg & Doctor [73].

4.5.2 SMR-CCS Plant Model

A qualitative summary of the SMR-CCS plant model is provided in Figure 4.7. The plant model utilized in this study is based on a Foster Wheeler SMR-CCS plant developed for the International Energy Agency (IEA) [43]. Foster Wheeler is a leading player in the design and manufacture of industrial scale SMR plants; thus, the plant specified is assumed to be representative of the current technology. The Foster Wheeler plant considered here has a hydrogen production flow rate of 607 tonnes/day. Hence, it mitigates about 72% of the Shell Scotford Upgrader's hydrogen demand; details of the plant characteristics are provided in Table 4.5.

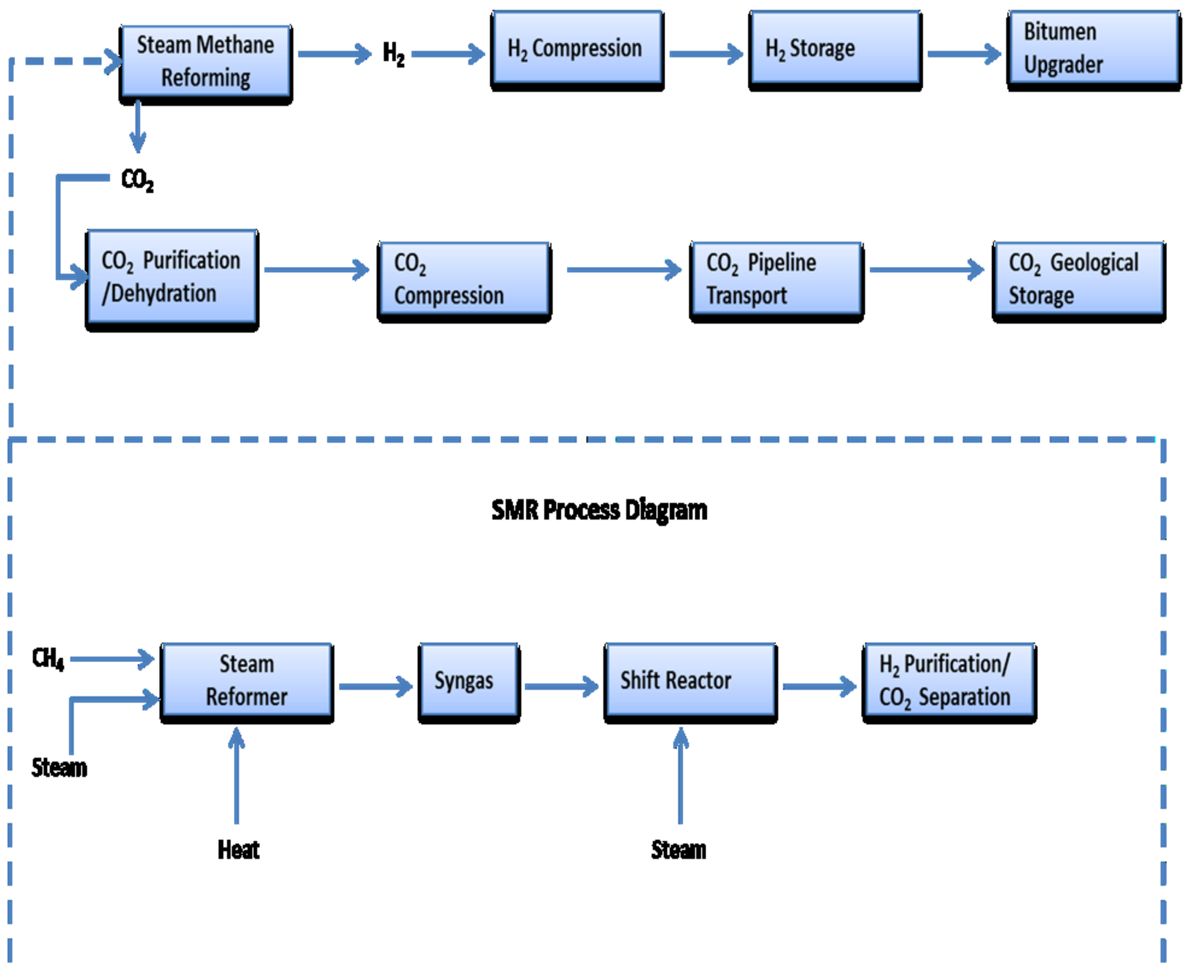

Figure 4.7: SMR-CCS plant model

Table 4.5: SMR-CCS plant specification

SMR Plant Specification	**Value**	**Sources/Comments**
SMR Plant Design Capacity	607,000 kg H_2 /day	[43]
Plant Capacity Factor	90%	[43]
Hydrogen Production Pressure	70 bar	Hydrogen is compressed from a PSA output pressure of 14 bar to 70 bar, with the aid of a hydrogen compressor.
Natural gas to H_2 Conversion Ratio (GJ/GJ H_2)	1.315	[43]
CO_2 Flow Rate	4406 tonnes/day	Based on a CO_2 flow rate of 204 tonnes/hr [43] along with the consideration of the plant capacity factor.
Bitumen Upgrader Specification		
Shell Scotford Upgrader Capacity	290,000 bpd	See Table 4.2
Synthetic Crude Oil (SCO) Production	246,500 bpd	See Table 4.2
H_2 Consumption for SCO Production	3.4 kg/barrel	See Table 4.2
H_2 Supply Required	838,100 kg H_2 /day	See Table 4.2

In the model developed, a number of plant equipment and downstream processes have been added to the earlier model [43] to ensure its specificity to the bitumen upgrading industry in Alberta. The costs of hydrogen storage and compression have been included in the model. In addition, the number of operators required for the SMR-CCS plant has been assumed to be eight; which is in good agreement with the 9.5[14] operators for a current CCS demonstration plant [74]. Also, remuneration of operators has been assumed to be more specific to Alberta with an operator's annual salary of $70,000. Furthermore, explicit determination of the cost of CO_2 compression and pipeline transport were not considered in earlier studies [41, 43]. Hence, the determination of the required sizes of certain equipment e.g. CO_2 pump and compressor, as well as the characterization of the CO_2 pipeline have been added to the current model developed in this study.

Some of the major base case assumptions and input data of the SMR-CCS model are given in Table 4.6 below.

[14]A shared electrician is the reason for the fraction of operators [74].

Table 4.6: SMR-CCS model principal economic data

Parameter	**Value**	**Sources/Comments**
Base Case Natural Gas Cost	$5/GJ	Estimated average value of the natural gas price over the plant's lifetime
Electricity Cost	$0.07/kWh	Estimated average value of the electricity price over the plant's lifetime [55]
Inflation	2.5%	
Base Case Internal Rate of Return (IRR)	10%	A reduced IRR for SMR relative to UCG is reflective of the technological maturity of SMR on a large commercial scale as well as its reduced environmental and production reliability risks in comparison to UCG.
Plant Equipment O&M Factor	4% of Capital Cost	[56]
Albertan Installation Factor	1.65	See Table 4.3
Plant Lifetime	25 years	[43]
Number of Plant Operators	8	
Plant Operator Salary per Annum	$70,000	Estimated average salary in Alberta for a plant operator. Operators are to have 3 daily 8hr shifts [43].
Supervision and Administration Cost	$1,300,000	Estimated to be 80% of annual operator labour cost [43].

4.6 Cost Estimation with SMR-CCS in Western Canada

4.6.1 Plant Equipment Capital Cost

The costs of plant equipment were derived from a number of studies and in consultation with experts. Wherever costs were not available these were developed based on suitable assumptions. A summary of the plant costs considered in this study is provided in Table 4.7 (all costs are rounded up to the nearest million).

Table 4.7: SMR-CCS capital costs

Equipment	Cost ($CAD Millions)	Sources/Comments
Reformer	120	[43]
H_2 Purification	103	[43]
WGS Reactor, Steam Generator etc [a].	309	[43]
Catalysts and Chemicals	16	[43]
Contingency Cost	69	[43]
H_2 Storage	407	Estimated using model provided by [56].
H_2 Compressor	41	Estimated using model provided by [56].
CO_2 Compressor & Pump	71	Estimated using model provided by [59].

Equipment	Cost ($CAD Millions)	Sources/Comments
CO_2 Pipeline	69	Estimated using model provided by [59].
CO_2 Sequestration	69	CO_2 sequestration capital cost is assumed to be equivalent to pipeline capital cost [60].
Total	**1,272**	

[a]This includes steam raising, power production, water gas reaction shift equipment, along with bulk materials such as piping, electrics and instrumentation.

4.6.2 Hydrogen Storage Cost

A storage capacity of 50% (i.e. 12 hrs of daily hydrogen production) is a reasonable estimate for the SMR-CCS plant, and is similar to that utilised in an earlier study [56]. It is important to point out that the storage capacity can vary from 12-24 hrs of production, which can have a significant impact on storage costs. As is predominantly the case in Alberta, 95% of hydrogen production in the world is captive [75]; hence, depending on specific conditions, e.g., frequency of hydrogen production lulls, the storage capacity required at a given plant will vary.

Above ground high pressure storage vessels are to serve as the storage medium for the plant. The hydrogen produced at the SMR-CCS plant leaves the PSA unit at a pressure of about 14 bar. It is then compressed to a pressure of about 70 bar to be suitable for storage in the pressure vessels. In this study, the storage specific cost amounted to $5,798/GJ H_2 [56].

4.6.3 H_2 Compression Cost

The compressor power required for hydrogen storage at the desired storage pressure of 70 bar was determined using a model developed in an earlier study [56] – see section 3 of Appendix. Details of the compressor characteristics and costs are provided in Table 4.8.

Table 4.8: Hydrogen compressor characteristics

H_2 Compressor	**Values**	**Sources/Comments**
Hydrogen Production Pressure from PSA (bar guage)	14	[43]
Required pressure for storage (bar guage)	70	[56]
Compressor efficiency	0.7	[56]
Number of compression stages	2	[56]
Compressor power requirement (MW)	19.32	[56]
Compressor capital cost ($CAD/kW)	937.5	[56]

4.7 Results and Discussion

4.7.1 CO_2 Compression Cost

The need for compression is common to all scenarios involving CO_2 capture and sequestration, regardless of the hydrogen production pathway i.e. UCG or SMR. The total power rating and pressure output of the CO_2 compression equipment determined for the UCG-CCS plant were 48.1 MW and 15 MPa. For the SMR-CCS plant, these values correspond to 19.2MW and 15MPa. The significant difference in the power rating of the aforementioned compressors is due to the fact that the UCG-CCS plant has a CO_2 flow rate which is considerably greater than its SMR counterpart. The power rating of the compressor is particularly relevant from an economic standpoint, as it determines the capital cost incurred. In comparison to industrial standards, the CO_2 compression model utilised in this study is quite accurate. For comparative purposes, it is worth mentioning that a current CCS demonstration plant uses a 17 MW compressor with a pressure output of 14 MPa [74]. This is comparable to the SMR-CCS plant parameters; however, the demonstration plant compressor was sized for a reduced flow rate of 3300 tonnes/day. Notwithstanding, the realistic nature of the compressor model is still apparent.

As shown in Figure 4.3, the cost of CO_2 compression decreases in a non-linear fashion as the CO_2 flow rate is increased. Hence, it can be inferred that CO_2 compression is more cost effective (in $/kW) at larger scale plants relative to smaller ones; as a result of benefits from economies of scale. Figure 4.3 also shows that at a CO_2 flow rate greater than 9,566 tonnes/day, the compressor capital cost rises. This is as a result of two compressor trains being required for a compressor size greater or equal to 40MW [59]. The pumping of CO_2 from its supercritical ‘dense’ phase to the final pressure of 150 bar entails significantly reduced capital costs compared to compression (as illustrated in Figure 4.4). This difference is mainly due to the high compression ratio associated with the compressor, which is greatly reduced with the pump. However, the economies of scale that can be achieved with the pump are less significant in comparison to the compressor. Figure 4.4 shows that for a CO_2 flow rate greater than 5,000 tonnes/day, the cost savings ($/kW) for the pump become relatively insignificant.

4.7.2 CO_2 Pipeline

The pipeline cost of transporting the CO_2 evolved at the plant to the sequestration location, for different pipeline lengths, is shown in Figure 4.5. The pipeline cost per unit distance increases with both an increase in distance and CO_2 mass flow rate. However, the gradient of the cost curves in Figure 4.5 gradually decreases as the magnitude of the flow rate is increased. Furthermore, it can be seen that at smaller CO_2 mass flow rates, the impact of length on the cost is less significant in comparison to higher flow rates. Figure 4.5 also illustrates that as the pipeline length increases, the differences in the capital cost per unit distance for each CO_2 mass flow rate considered, becomes increasingly reduced. Thus, it can be inferred that the pipeline cost becomes more economical as its scale is increased. Lastly, the cost per unit distance is shown to be more sensitive to the magnitude of the flow rate, as opposed to the pipeline distance. The diameter of the pipeline is invariably tied to the mass flow rate and thus determines how much material is required for the pipeline construction as well as manufacturing costs. This is reflected in the pipeline cost estimation model utilized in this study [59]; as the mass flow rate has a greater exponent in comparison to the pipeline length.

4.7.3 UCG Hydrogen Production Cost

At base case conditions (IRR of 15% and a coal price of \$0/tonne), the hydrogen production cost from UCG-CCS including hydrogen delivery is \$2.11/kg of H_2 and UCG (without CCS) including hydrogen delivery is \$1.78/kg of H_2. The hydrogen production cost without hydrogen delivery from UCG-CCS and UCG are \$1.75/kg of H_2 and \$1.42/kg of H_2 respectively. Thus, it becomes apparent that the additional cost of the hydrogen pipeline has a comparable effect on the UCG production cost as the CCS infrastructure. The sensitivities of key plant parameters on the production cost are shown in Figure 4.8. As seen in Figure 4.8, the installation factor, and to a greater degree, the UCG efficiency, have a relatively small effect on the price of hydrogen production. Note that the installation factor was only applied to selected plant equipment (see Table 4.4); hence, its effect is small. The UCG efficiency on the other hand has a relatively miniscule effect on production cost mainly due to the fact that the coal feedstock cost is negligible. Thus, the only financial penalty or benefit for an increased/decreased coal to hydrogen efficiency is reflected in the water (steam) consumption of the plant. The water resource cost utilised in this study amounts to \$0.99/m^3 [50]; hence, it has a relatively insignificant effect on the hydrogen production cost. In contrast, the capital cost estimate and the incremental rate of return (IRR) have a significant impact on the cost of hydrogen. The sensitivity of the IRR in particular shows that the commercial viability of UCG-CCS in Alberta, will be determined to a great extent by the perceived risks associated with the technology. In this regard, the deferral of the commercial scale deployment of the Swanhills UCG plant, due to low natural gas prices, highlights pertinent economic risks to UCG in Alberta – this is likely to have increased the perceived risk associated with the technology in Alberta.

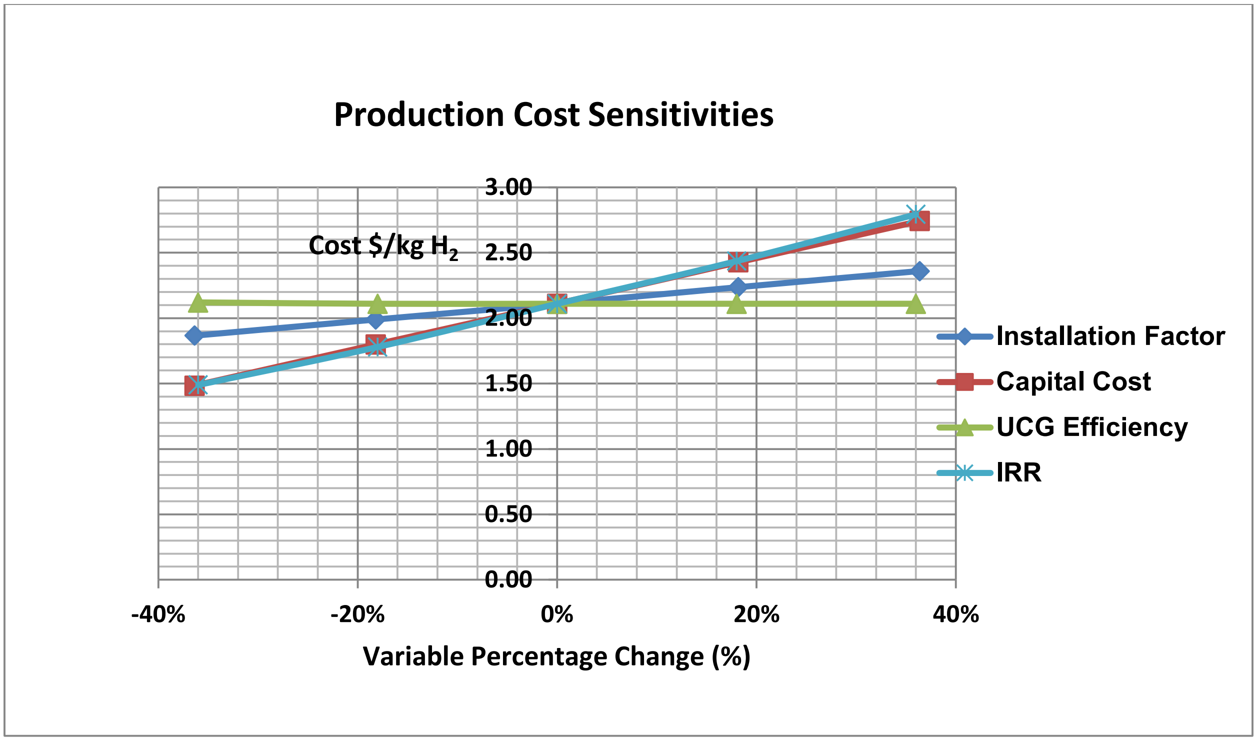

Figure 4.8: UCG-CCS hydrogen production cost sensitivities

4.7.4 SMR Hydrogen Production Cost

As seen in Figure 4.9, the hydrogen production cost at base case conditions (IRR 10% and natural gas price of $5/GJ) amounts to $1.73/kg of H_2 for SMR (without CCS) at Fort Saskatchewan, Alberta and $2.14/kg of H_2 for SMR-CCS with production at Fort Saskatchewan and CO_2 sequestration at Thorhild, Alberta. At an equivalent IRR to UCG i.e. 15%, these costs amount to $2.02 /kg$H_2$ and $2.57/kg$H_2$ for SMR and SMR-CCS, respectively. Thus, the impact of the IRR on production costs is significant. Intuitively, Figure 4.9 also shows that the hydrogen production cost increases linearly as the natural gas price is increased. The sensitivities of other parameters on the hydrogen production cost are illustrated in Figure 4.10. It can be seen that the plant capital cost and the IRR are the two most influential parameters on hydrogen production cost (see Figure

4.10). However, among all the input parameters, the plant capital cost is the single most influential factor, which is quite intuitive. The IRR is closely followed by the natural gas price with regards to the impact on production costs, and the installation factor has an intermediate effect on the production cost compared to other parameters. Finally, the significance of the electricity cost is quite small. This is attributed to the fact that it only applies to the compressor and pump equipment.

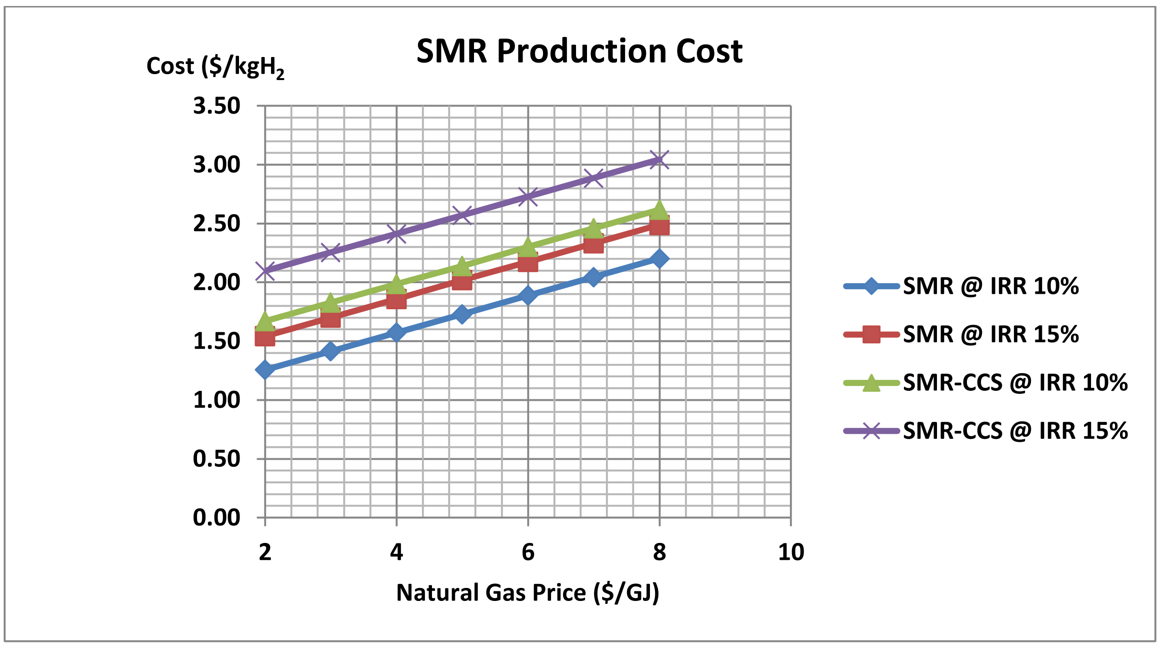

Figure 4.9: SMR-CCS hydrogen production costs

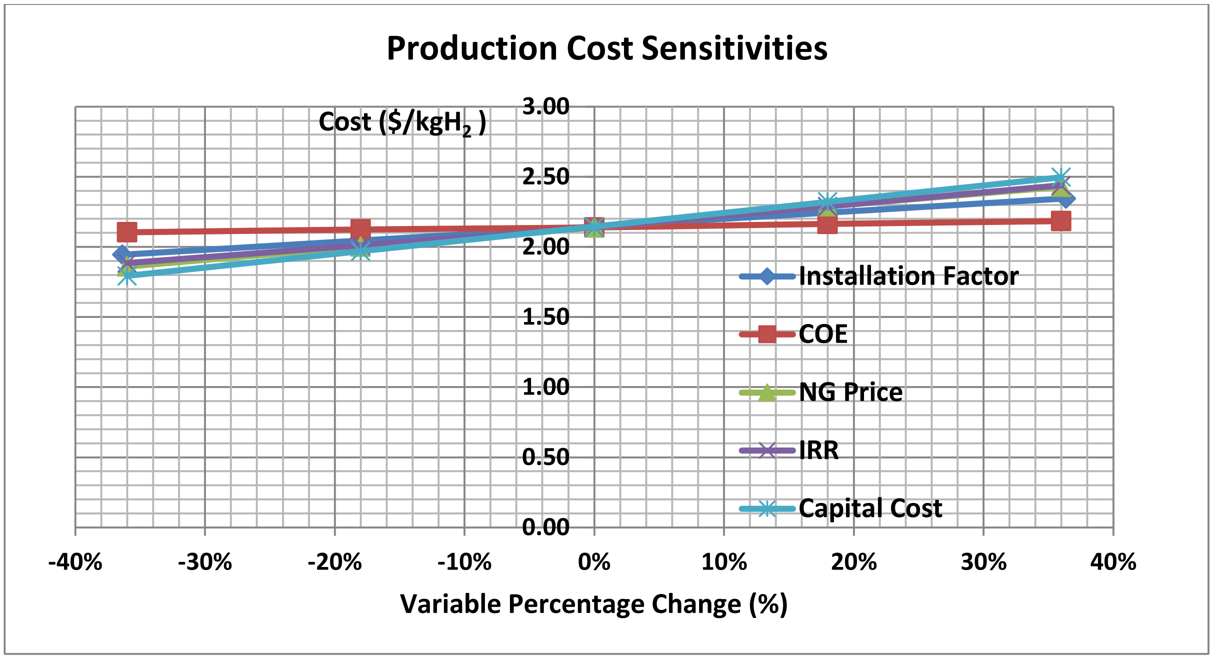

Figure 4.10: SMR-CCS hydrogen production cost sensitivities

4.8 UCG-CCS & SMR-CCS Comparative Costs

4.8.1 Production Cost Scenario Comparison

The comparison of production costs between UCG and SMR with and without CCS infrastructure is illustrated in Figure 4.11. Furthermore, the seven scenarios considered in this study are also included. It is worth re-iterating that in this study, the IRR for UCG is 15%; while the corresponding value for SMR is 10%. Thus, the results presented in Figure 4.11 should be viewed with full consciousness of this IRR differential. The effect of the IRR differential on the competiveness of both technologies is given greater scrutiny in section 4.8.2

For hydrogen production without CCS, at base case conditions, SMR is the more cost-effective option in comparison to UCG. This remains true for a natural gas price below \$5.30/GJ (see Figure 4.11). The reason for this advantage can be attributed to the fact that with SMR, the hydrogen production cost is highly sensitive to the natural gas price and IRR. That being said, the significant natural gas feedstock cost of SMR in comparison to the zero feedstock cost of UCG is offset by the increased investment risk for UCG, as reflected in the 5% IRR differential. Furthermore, a higher initial investment is required for UCG, mainly due to the expensive hydrogen pipeline; this helps to further offset the high feedstock cost of SMR.

However, as seen in Figure 4.11, for hydrogen production with CCS, UCG is the more economic option at the base case conditions despite the 5% IRR differential. For SMR-CCS to be competitive, the natural gas price would have to fall below \$4.80/GJ. Furthermore, considering scenario 7, the inclusion of the sale of CO_2 for EOR operations reduces the base case UCG-CCS price from \$2.11 to \$1.61/kg H_2. The selling price of CO_2 utilised in this study for EOR operations is assumed to be about \$47/tonne CO_2 [75, 76]. Furthermore, the incremental CO_2 flow rate of the UCG-CCS plant over the SMR-CCS alternative was utilised in the calculation of EOR revenues; so as to accommodate the possibility of EOR CO_2 sales from SMR-CCS. With the inclusion of EOR, the competitiveness of UCG-CCS over SMR-CCS is further enhanced. The natural gas price would have to fall below \$1.55/GJ for SMR-CCS to be the economically superior option. Table 4.9 gives the cost of hydrogen production for the seven scenarios considered.

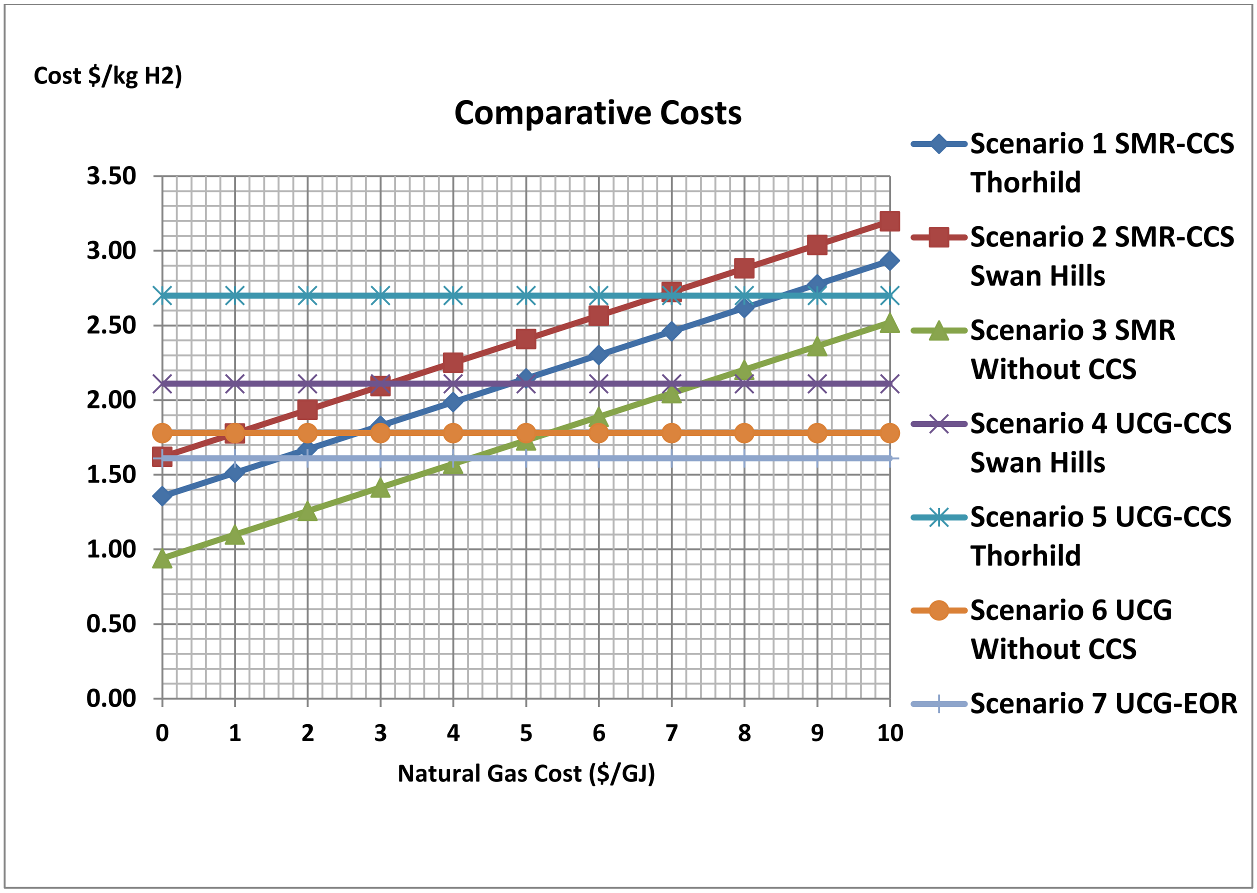

Figure 4.11: Hydrogen production cost comparison for scenarios 1 to 7

Table 4.9: Hydrogen production cost for scenarios 1 to 7

Scenarios	Description	Hydrogen Production Cost ($/kg of H_2)
Scenario 1	SMR-based H_2 production at Fort Saskatchewan, Alberta with CO_2 sequestration in Thorhild, Alberta	2.14
Scenario 2	SMR-based H_2 production at Fort Saskatchewan, Alberta with CO_2 sequestration in Swan Hills, Alberta	2.41
Scenario 3	SMR-based H_2 production without CCS	1.73
Scenario 4	UCG-based H_2 production at Swan Hills Alberta, with H_2 delivery to Fort-Saskatchewan, Alberta and sequestration in Swan Hills, Alberta	2.11
Scenario 5	UCG-based H_2 production at Swan Hills Alberta, with H_2 delivery to Fort-Saskatchewan, Alberta and sequestration of CO_2 in Thorhild, Alberta	2.70
Scenario 6	UCG-based H_2 production at Swan Hills, Alberta with H_2 delivery to Fort-Saskatchewan, Alberta without CCS	1.78
Scenario 7	UCG-based H_2 production at Swan Hills, Alberta with H_2 delivery to Fort-Saskatchewan, Alberta and sale of CO_2 for enhanced oil recovery (EOR)	1.61

4.8.2 Investment Risk Comparison

The degree of investment risk associated with UCG in comparison to SMR is pivotal in determining the competiveness of both technologies. In this study, the degree of risk of both technologies is assumed to be reflected in the minimum IRR required for investment. The impact of the IRR differential on the competitiveness of both technologies is illustrated in Figure 4.12. First, it can be seen in Figure 4.12 that at an equivalent IRR (0% differential), UCG with and without CCS is the more economical hydrogen production pathway by a relatively large margin.

In the case of hydrogen production without CCS, UCG remains the more cost-effective option up until the IRR differential between both technologies is above 4.6%. For hydrogen production with CCS, the competitive threshold is extended to an IRR differential above 5.4%. Thus, it becomes apparent that UCG is a financially sound technology particularly for hydrogen production with CCS.

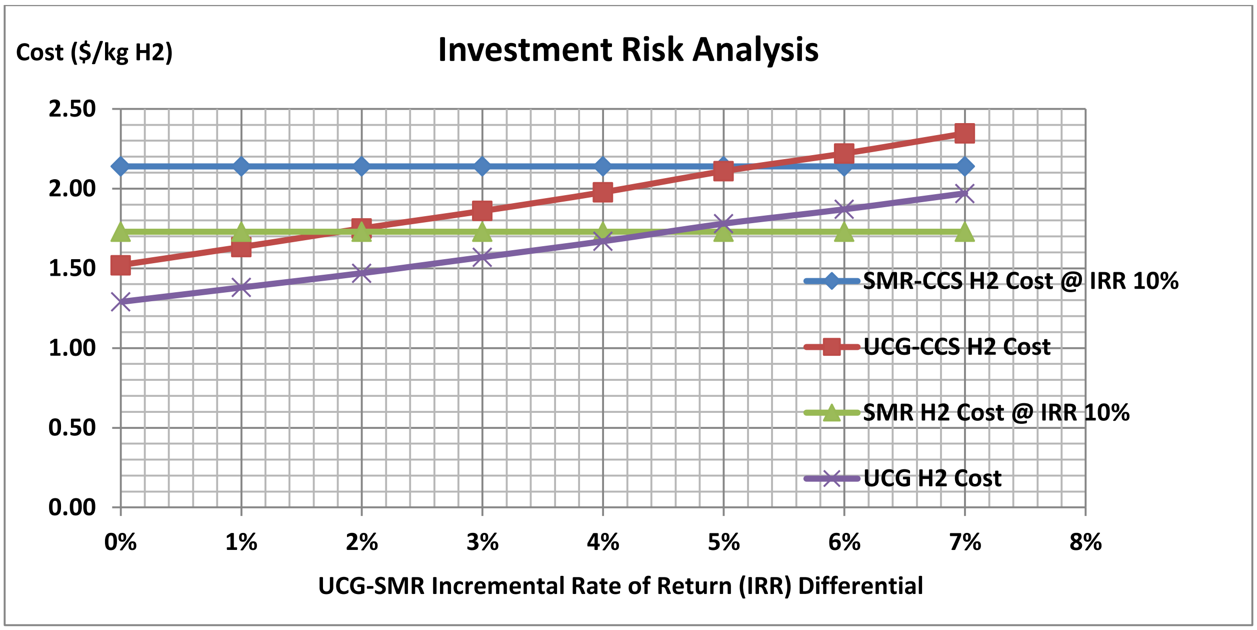

Figure 4.12: Investment risk analysis

4.9 Conclusion

From the techno-economic assessment conducted, a number of useful conclusions can be drawn. First, the sensitivity analysis conducted demonstrates that the natural gas price has a significant effect on the competitiveness of the SMR-CCS hydrogen production cost. In the case of UCG-CCS, no fuel costs are incurred. Thus, the profitability risk of the investment with regards to the feedstock cost is negligible for UCG, and quite significant for SMR. That being said, the feedstock cost volatility risk is counter-balanced by the fact that the technological infancy, environmental risks, and reduced reliability of UCG is greater than its SMR counterpart.

At base case conditions, for hydrogen production without CCS, SMR is the more cost competitive technology relative to UCG; however, for hydrogen production with CCS, UCG is more cost efficient. In particular, the effect of EOR on the competitiveness of UCG-CCS against SMR-CCS at base case conditions is quite significant. At base case conditions, the consideration of potential EOR revenue (scenario 7) from both technologies is the case where the competitive advantage of UCG-CCS against SMR-CCS is highest in magnitude.

The competiveness of UCG against SMR is highly sensitive to the perceived investment risk associated with UCG. If the IRR differential is less than 5.4%, UCG is the more cost effective technology for hydrogen production with CCS; if greater that 5.4%, SMR becomes the more economical pathway. For hydrogen production without CCS, the competitive margin of UCG is reduced to an IRR differential of 4.6%.

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Chapter 5[1]

Conclusions, Future Research and Trends

5.1 Conclusions

A multitude of low-GHG, renewable and non-renewable hydrogen production pathways were assessed from a techno-economic perspective, so as to evaluate their competitiveness in producing hydrogen that can be utilized in the energy economy of Alberta, as well as other jurisdictions. The assessment of these hydrogen production pathways involved the development of integrated techno-economic models. These models were built such that they satisfy a research deficit in the pertinent literature or address the limitations associated with existing research paradigms. Accordingly, the publications that stem from these models, which comprise the chapters of this thesis, have translated into novel contributions in their respective research contexts. Four hydrogen pathways were assessed in the body of work contained in this thesis – namely wind-hydrogen production, hydropower-hydrogen production, coal-based (UCG) hydrogen production and natural-gas (SMR) based hydrogen production. The key conclusions associated with each pathway are as follows:

[1] Part of this chapter is based upon a book publication. Please refer to: *Olateju, B. and A. Kumar. Clean energy-based production of hydrogen: An energy carrier. Handbook of Clean Energy Systems; 2015. 1–30*

5.1.1 A Techno-Economic Assessment of Large Scale Hydrogen Production from Wind Energy with Energy Storage for Bitumen Upgrading in Western Canada

- Hydrogen production from wind energy (with or without energy storage) is uncompetitive with SMR.
- The benefits of energy storage in grid connected wind hydrogen systems (enhanced electrolyser capacity factor and premium electricity sales) are limited by the efficiency penalty introduced by utilizing energy storage technology; in this case, a battery.
- For a particular electrolyser-battery configuration, the minimum hydrogen production cost occurs when their respective capacity factors are approximately equivalent.
- For the plant size evaluated (563 MW), hydrogen production costs from grid connected wind-hydrogen plants with energy storage range from \$3.37/kg H_2 to \$15.06/kg H_2, depending on the electrolyser size and the use of existing wind turbine assets.
- At base case conditions, the optimal electrolyser farm size and energy (battery) storage capacity consists of 81 units of a 3496 kW (760 Nm^3/h) electrolyser and 360 MWh of energy storage capacity (60 battery units).

5.1.2 A Techno-Economic Assessment of Large Scale Hydrogen Production from Hydropower for Bitumen Upgrading in Western Canada

- Hydrogen production from hydropower energy is competitive with SMR.
- For the plant size evaluated (436 MW), hydrogen production costs from grid connected hydropower-hydrogen plants range from \$1.18/kg H_2 to \$5.35/kg H_2, depending on the electrolyser size and the use of existing hydropower assets.
- At base case conditions, the optimal electrolyser farm size consists of 90 units of a 3496 kW (760 Nm^3/h) electrolyser.

5.1.3 A Techno-Economic Assessment of Large Scale Hydrogen Production from UCG with CCS for Bitumen Upgrading in Western Canada

- At base case conditions, hydrogen production from UCG without CCS (\$1.92/kg H_2) is slightly less competitive relative to SMR (\$1.87/kg H_2).
- At base case conditions, hydrogen production from UCG-CCS (\$2.28/kg H_2 to \$2.92/kg H_2) is slightly more competitive relative to SMR-CCS (\$2.31/kg H_2 to \$2.60/kg H_2).
- The competitiveness of both technologies relative to one another is highly sensitive to the natural gas price and perceived investment risk.

From the research conducted on the respective hydrogen pathways, a number of salient characteristics have been discerned; these are juxtaposed succinctly in Table 5.1, using a number of metrics. The metrics highlighted are not all-encompassing, as they will vary across stakeholders with different priorities and biases. Nonetheless, the juxtaposition of the challenges, competitive advantages, trade-offs and opportunities across a number of hydrogen pathways, which can service the bitumen upgrading industry, is helpful in providing a holistic view of each option.

Table 5.1: Hydrogen production pathways – Selected characteristics

	HYDROGEN PRODUCTION PATHWAYS			
METRICS	**Wind-hydrogen - Electrolysis**	**Hydro-hydrogen- Electrolysis**	**Natural Gas – SMR & SMR-CCS**	**Coal –UCG & UCG-CCS**
H_2 Cost (\$/kg of H_2)	3.37 – 15.06	1.18 – 5.35	1.87 & 2.31 – 2.60	1.92 & 2.28 – 2.92
Plant Scale (tonnes H_2/day)	59[1]	146[1]	607	660
Energy Logistics	• Stranded resource • Wind energy cannot be transported in its original form and requires conversion to electricity • Transmission line constraints to link supply and demand could be a formidable hurdle • Likely to be economically competitive only for captive H_2 production	• Stranded resource • Hydropower cannot be transported in its original form and requires conversion to electricity • Transmission line constraints to link supply and demand, could be a formidable hurdle • Likely to be economically competitive for captive and non-captive H_2 production	• Natural gas is highly mobile and is predominantly found in centralized locations with large volumes • Economically competitive for captive and non-captive H_2 production • Resource transportation infrastructure (pipelines) and operational experience is mature and readily available	• UCG coal seams are a stranded resource and are often found in large centralized volumes • Likely to be economically competitive for captive and non-captive H_2 production • Coal resource is utilized in situ; hence, transportation is not required.
Feedstock Supply/Availability	• Feedstock supply is renewable. • Intermittent feedstock (wind energy) supply.	• Feedstock supply is renewable. • Non-intermittent and predictable feedstock supply	• Feedstock supply is finite. Hence, supply diminishes over time. • Conventional natural gas production	• To a significant degree the availability of UCG coal seams will be dependent on the activities of surface

[1] Corresponds to the scale at which the minimum hydrogen production cost is achieved.

HYDROGEN PRODUCTION PATHWAYS				
METRICS	**Wind-hydrogen - Electrolysis**	**Hydro-hydrogen- Electrolysis**	**Natural Gas – SMR & SMR-CCS**	**Coal –UCG & UCG-CCS**
	• Availability of wind energy resource is limited to indigenous resource endowment.	• Availability of hydropower resource is limited to indigenous resource endowment.	growth rates are experiencing a general downward trend. Unconventional natural gas production growth rates are in the ascendancy. Future supply of natural gas will likely depend on the ability of technology to access unconventional reserves economically. • Widely available across jurisdictional and global markets	and underground mining operations which will leave 'unmineable' coal seams in their wake. • Availability of UCG coal seams is limited to indigenous resource endowment.
Techno-Economic and Environmental Merits & Challenges	• Large scale (> 50 tonnes H_2/day) and low cost H_2 production (< \$3.00/kg H_2) is relatively difficult to achieve. • Ability to yield consistent high purity hydrogen and oxygen • No feedstock cost or cost volatility	• Large scale and low cost H_2 production is readily achieved. • Ability to yield consistent high purity hydrogen and oxygen • No cost volatility associated with feedstock	• Large scale and low cost H_2 production is readily achieved. • High quality feedstock - high hydrogen to carbon ratio • Ability to yield consistent syngas and hydrogen quality and composition	• Large scale and low cost H_2 production is readily achieved. • Low opportunity cost – utilizes 'unmineable' coal resources. • Negligible feedstock cost and volatility.

HYDROGEN PRODUCTION PATHWAYS				
METRICS	**Wind-hydrogen - Electrolysis**	**Hydro-hydrogen-Electrolysis**	**Natural Gas – SMR & SMR-CCS**	**Coal –UCG & UCG-CCS**
	• Relatively low wind turbine capacity factors	• Relatively high capacity factors with base-load capability • Flooding of large expanses of land with hydropower incurs socio-environmental externalities e.g. displacement of indigenous peoples/remote communities etc • Damming of rivers can be politically tenuous, given the ubiquitous use of water and the number of stakeholders – small run-of-river hydro projects are likely to be less contentious	• Expected to be highly compatible with CCS deployment efforts • Significant opportunity cost of H_2 production – natural gas can be used to mitigate power and heat demands • Natural gas costs are quite volatile • Non-GHG environmental externality risks (e.g. ground water contamination) associated with unconventional gas production (e.g. hydraulic fracking) are becoming increasingly problematic for the social license of operators	• Expected to be highly compatible with CCS deployment efforts. • Fluctuations in syngas quality and composition consistency are likely. • Complex in situ gasification process and limited operational control. • Ground water contamination and land subsidence risks.

HYDROGEN PRODUCTION PATHWAYS				
METRICS	**Wind-hydrogen - Electrolysis**	**Hydro-hydrogen- Electrolysis**	**Natural Gas – SMR & SMR-CCS**	**Coal –UCG & UCG-CCS**
Technological Maturity	• Mature	• Mature	• Mature	• Demonstration phase

5.2 Future Work

5.2.1 Life Cycle Analysis (LCA)

The techno-economic analysis carried out for the various hydrogen production pathways addressed in this thesis should be complemented with a life cycle analysis (LCA) of pertinent environmental impacts. Of particular significance in this proposed LCA study will be the lifecycle GHG emissions from the different pathways. Research that addresses the LCA of integrated sustainable hydrogen pathways, particularly in Western Canada, is limited. There is an opportunity for increased understanding in this area, especially from a comparative standpoint.

5.2.2 Greenhouse Gas Emissions Abatement Cost

The combination of the techno-economic modelling and the LCA study proposed above, will allow for the GHG mitigation cost from each hydrogen pathway to be estimated. Against the backdrop of the GOA's announcement of an economy-wide carbon price of \$30/tonne CO_2e, the value of ascertaining the GHG mitigation cost for each pathway, as Alberta transitions into a low carbon economy, is significant.

5.2.3 Alternative and Emerging Hydrogen Production Pathways

Technologies with the potential to incur relatively low GHG emissions, which are capable of producing electricity at a competitive cost, and are non-intermittent, hold promise for the increased competitiveness of electrolytic hydrogen pathways. Geothermal energy falls under this characterization; moreover, there is an increased interest in Western Canada for producing sustainable energy from geothermal resources [1]. The cost of electricity from geothermal in some cases, is more competitive than renewable generators such as wind and biomass [2]. Having said that, costs and GHG emissions from geothermal are highly site specific and also depend on the geothermal technology employed [3]. Notwithstanding, the potential of geothermal hydrogen production is evident and worthy of thorough investigation.

5.2.4 Heterogeneous Electrolyser Configurations

In the hydropower and wind energy - based electrolytic hydrogen pathways, all the electrolyser farm configurations evaluated involved electrolysers of identical size (homogenous) and a particular number of units. Accordingly, the optimal number of units for a particular electrolyser size was ascertained. Future research should investigate the optimal combination of different sized (heterogeneous) electrolyser units i.e. the optimal combination of electrolyser size and number of units, in the wind-hydrogen or hydropower-hydrogen plants. Using different sized electrolysers in combination, has the potential to increase hydrogen productivity and enhance the capacity factor of the electrolyser farm. It is worth re-iterating that smaller electrolysers facilitate higher electricity sales to the grid and higher electrolyser capacity factors. On the other hand, larger electrolysers have lower specific capital cost and higher hydrogen productivity. Leveraging the characteristics of both large and small electrolysers in an optimal fashion could facilitate the determination of a more competitive hydrogen production cost.

5.2.5 Economic Model Enhancement

The integrated techno-economic models developed in this thesis, utilize discounted cash flow (DCF) methodology in ascribing economic value to the hydrogen production pathways, and consequently estimating production costs. Although DCF methodology is the most comprehensive and widespread economic valuation tool utilized in industry, government and academia, it has limitations that are worthy of note - readers are encouraged to consult the work done by [4-6] which provide a comprehensive overview of DCF limitations. As shown below, a key limitation pertinent to DCF methodology, has to do with accounting for cash flow risk in a differentiated fashion:

- **DCF Discount Rate is Applied Uniformly for Investment Costs and Future Cash Flows**: Apart from the fact that assigning the appropriate discount rate that is reflective of the risks associated with a particular project can be subjective, there are some inherent limitations in the manner in which the DCF technique assigns risks to investment costs and future cash flows. As explained by [4], the discount rate utilized in a DCF model is the sum of the risk free discount rate (e.g. the discount rate commensurate with a very low-risk government bond) and the market risk adjustment – which often amounts to the weighted average cost of capital (WACC) for the project proponent. Intuitively, the discount rate applied to a given cash flow should be reflective of the risks associated with it [4, 5]. Investment cost often involves risk that is 'internal' to the project proponent, not 'external' market risks [4]. Examples of internal risks include: construction lead times, efficient procurement of materials, equipment and labour etc. On the other hand, external market risks include: price fluctuations, demand and supply forces, new market entrants (competition), technology obsolescence etc [4]. Thus, it becomes apparent that internal risks associated with investment costs and external market risks associated with future cash flows, differ in the degree of risk that they incur. The DCF technique assigns a 'one-size fits all' discount rate to the investment cost and future cash flows, despite the fact that they have different risks [4, 5].

The limitations of the DCF technique do not render it obsolete or ineffective [4, 6], rather it provides ample opportunity for enhancing the robustness of the techno-economic models built in this body of work. In future research, complementary models to DCF methodology such as real options valuation (ROV) could be utilized. ROV considers the value associated with the resolution of uncertainty that pertains to a particular project, and uses options pricing theory for assessing that value [4]. Additionally, ROV incorporates a two-step approach in accounting for cash flow uncertainty; it adjusts cash flows for risk informed by inputs from the financial markets, and then discounts for the time value of money in a separate step [5]. The literature presented by [4-6] can be consulted by readers, to serve as a point of departure for future research in this area.

5.3 Future Trends

5.3.1 Technological Innovation - Partial Bitumen Upgrading

The principal economic incentive to upgrade bitumen to SCO is the light-heavy crude oil price differential in the North American energy market; however, the light-heavy differential is quite volatile (see Figure 5.1) and hence incurs a significant degree of economic risk. Moreover, a key drawback associated with the use of conventional 'full-upgraders' is that they are often large scale and centralized, involving considerable capital investment. Additionally, as mentioned in previous sections, the intensive use of hydrogen accounts for a significant portion of their GHG emissions. These characteristics are particularly undesired in an energy market that is increasingly GHG constrained coupled with the current low oil price environment. Consequently, the oil sands industry is making considerable efforts to develop partial upgrading technologies that can produce a 'pipeline-ready' SCO sour crude (this is a lower quality crude compared to SCO from full-upgraders), without the use of diluents to transport it to market [7-9]. An attendant implication of partial upgrading technologies is that they have the potential to significantly reduce the consumption of hydrogen in upgrading operations [7]. This competitive advantage of partial upgraders vis-à-vis full upgraders, has material cost and GHG emissions reduction impacts [10]. Partial upgraders can be deployed in a decentralized, modular fashion at the site of the bitumen resource, at much smaller scales than full-upgraders (scales as low as 10,000 bpd have been reported as possible [11]) - thereby also negating the need to transport the bitumen to a distant upgrader location. Examples of some oil sands companies making efforts to commercialize and deploy partial upgrading technology include: MEG Energy and Petrosonic Energy [11].

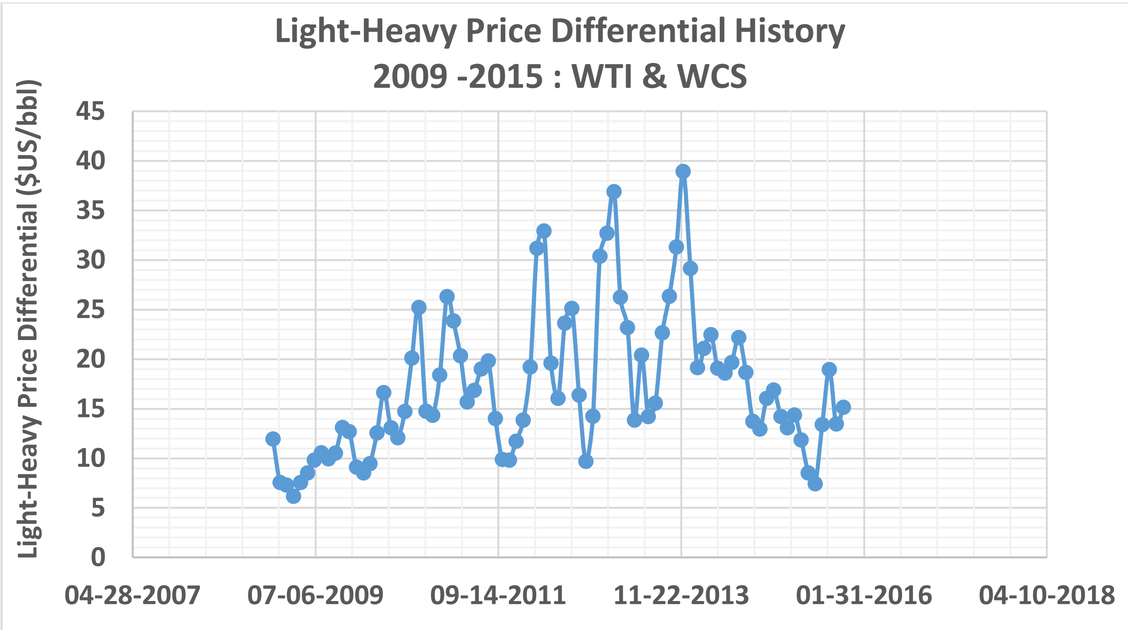

Figure 5.1: North American light-heavy crude oil price differential history [12]

At present, the commercialization of partial upgrading technology is yet to be realized in the bitumen upgrading industry. Notwithstanding, a future in which partial upgrading is adopted widely in the oil sands industry is possible. Assuming this future comes to fruition, the importance of alternative, economic and environmentally sound hydrogen pathways may appear to be diminished, due to the significant reduction of hydrogen use in upgrading technology. However, this is unlikely. Partial upgrading if successful and widely adopted, will likely ‘shift’ the hydrogen demand further downstream along the bitumen value-chain. Hydrogen will still be needed at the refineries where the partially upgraded bitumen is converted into value-added refined products such as gasoline, jet fuel and diesel. With GHG and non-GHG fuel standards (e.g. sulphur content in fuels) likely to be more stringent over time, sustainable hydrogen production pathways are likely to be of significant importance in the GHG constrained economies of the future.

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Appendix

Section 1: Wind- Hydrogen Model (With Energy Storage) - Energy Management Algorithm

The energy management algorithm determines the hourly operating mode of the plant for the entire year i.e. 8760 hrs. The equations for the different operating modes vary, based on the electrolyser farm size, the battery capacity, the hourly amount of energy generated from wind and the hourly pool price. Applicable equations for the different operating modes in the plant are provided in the form of nested 'if statements' (conditional statements with Boolean outcomes) with annotations below. The reader is encouraged to consult standard programming language literature, to enhance their understanding of 'if statements'.

if $\boldsymbol{WT_{Energy}(i) + B_{Storage}(i-1) < B_{Charge-Min}}$

$$E_{Grid}(i) = WT_{Energy}(i)$$

$$B_{Charge}(i) = 0$$

$$B_{Discharge}(i) = 0$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = 0$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode H – Plant Lull (this mode only occurred when the energy produced from the wind turbines was nil)

else

if $\boldsymbol{WT_{Energy}(i) + B_{Storage}(i-1) < B_{Capacity} \times B_{Number\ of\ Units}}$

if $\boldsymbol{WT_{Energy}(i) + B_{Storage}(i-1) < EP_{Min}}$

if $\boldsymbol{i = Peak\ A\ or\ Peak\ B}$

$$E_{Grid}(i) = WT_{Energy}(i) + B_{Storage}(i-1)$$

$$B_{Charge}(i) = 0$$

$$B_{Discharge}(i) = B_{Storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = 0$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode E – Electricity is sold to the Grid Only

else

$$E_{Grid}(i) = 0$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = 0$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = 0$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode F – Energy Storage Only

end

else

if $\boldsymbol{WT_{Energy}(i) + B_{Storage}(i-1) < EP_{Max} \times E_{Number\ of\ Units}}$

$$E_{Grid}(i) = 0$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i) + B_{Storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = \left(WT_{Energy}(i) + B_{Storage}(i-1)\right) \times EF_{Max} \times E_{Number\ of\ Units}/\left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode A – Hydrogen production only

else

if $\boldsymbol{WT_{Energy}(i) < EP_{Max} \times E_{Number\ of\ Units}}$

if $\boldsymbol{i = Peak\ A\ or\ Peak\ B}$

$$E_{Grid}(i) = B_{Storage}(i-1)$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i) + B_{Storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = WT_{Energy}(i) \times EF_{Max} \times E_{Number\ of\ Units} / \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode C – Hydrogen Production with Premium Electricity Sales

else

$$E_{Grid}(i) = 0$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = WT_{Energy}(i) \times EF_{Max} \times E_{Number\ of\ Units} / \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode B – Hydrogen Production with Energy Storage

end

else

if $\boldsymbol{i = Peak\ A\ or\ Peak\ B}$

$$E_{Grid}(i) = B_{storage}(i-1) + \left(WT_{Energy}(i) - \left(EP_{Max} \times E_{Number\ of\ Units}\right)\right)$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode C – Hydrogen Production with Premium Electricity Sales

else

$$E_{Grid}(i) = 0$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1) + \left(WT_{Energy}(i) - \left(EP_{Max} \times E_{Number\ of\ Units}\right)\right)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

<u>*Note: Plant Operating in Mode B – Hydrogen Production with Energy Storage*</u>

end

end

end

end

else

if $\boldsymbol{WT_{Energy}(i) + B_{Storage}(i-1) < EP_{Min}}$

if $\boldsymbol{i = Peak\ A\ or\ Peak\ B}$

$$E_{Grid}(i) = WT_{Energy}(i) + B_{Storage}(i-1)$$

$$B_{Charge}(i) = 0$$

$$B_{Discharge}(i) = B_{storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = 0$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode E – Electricity is sold to the Grid Only

else

if $\boldsymbol{WT_{Energy}(i) < B_{Capacity} \times B_{Number\ of\ Units}}$

$$E_{Grid}(i) = B_{Storage}(i-1)$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = 0$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = 0$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode G – Energy Storage occurs in addition to the use of the Grid as a Dump Load

else

$$E_{Grid}(i) = B_{Storage}(i-1) + \left(WT_{Energy}(i) - \left(B_{Capacity} \times B_{Number\ of\ Units}\right)\right)$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = B_{Storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = 0$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode G – Energy Storage occurs in addition to the use of the Grid as a Dump Load

end

end

else

if $\boldsymbol{WT_{Energy}(i) + B_{Storage}(i-1) < EP_{Max} \times E_{Number\ of\ Units}}$

if $\boldsymbol{WT_{Energy}(i) < B_{Capacity} \times B_{Number\ of\ Units}}$

$$E_{Grid}(i) = 0$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = \left(WT_{Energy}(i) + B_{Storage}(i-1)\right) \times EF_{Max}$$
$$\times E_{Number\ of\ Units} / \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode A – Hydrogen production only

else

$$E_{Grid}(i) = WT_{Energy}(i) - \left(B_{Capacity} \times B_{Number\ of\ Units}\right)$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = \left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = \left(\left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)\right) \times EF_{Max}$$
$$\times E_{Number\ of\ Units} / \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales

end

else

if $\boldsymbol{WT_{Energy}(i) < B_{Capacity} \times B_{Number\ of\ Units}}$

if $\boldsymbol{WT_{Energy}(i) < EP_{Max} \times E_{Number\ of\ Units}}$

$$E_{Grid}(i) = \left(WT_{Energy}(i) + B_{storage}(i-1)\right) - \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales

else

if $\boldsymbol{i = Peak\ A\ or\ Peak\ B}$

$$E_{Grid}(i) = \left(WT_{Energy}(i) + B_{storage}(i-1)\right) - \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

<u>Note: Plant Operating in Mode C – Hydrogen Production with Premium Electricity Sales</u>

else

$$\textbf{if}\ \left(\boldsymbol{WT_{Energy}(i) - (EP_{Max} \times E_{Number\ of\ Units})}\right) + \boldsymbol{B_{storage}(i-1) < B_{Capacity} \times B_{Number\ of\ Units}}$$

$$E_{Grid}(i) = 0$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = EP_{Max} \times E_{Number\ of\ Units}$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

<u>Note: Plant Operating in Mode B – Hydrogen Production with Energy Storage</u>

else

$$E_{Grid}(i) = WT_{Energy}(i) - (EP_{Max} \times E_{Number\ of\ Units})$$

$$B_{Charge}(i) = WT_{Energy}(i)$$

$$B_{Discharge}(i) = WT_{Energy}(i)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

<u>*Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales*</u>

end

end

end

else

if $\boldsymbol{WT_{Energy}(i) < EP_{Max} \times E_{Number\ of\ Units}}$

if $\boldsymbol{B_{storage}(i-1) + (B_{Capacity} \times}$

$\boldsymbol{B_{Number\ of\ Units}) < EP_{Max} \times E_{Number\ of\ Units}}$

$$E_{Grid}(i) = WT_{Energy}(i) - (B_{Capacity} \times B_{Number\ of\ Units})$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = B_{storage}(i-1) + (B_{Capacity} \times B_{Number\ of\ Units})$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = \left(\left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)\right) \times EF_{Max} \times E_{Number\ of\ Units}/\left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales

else

$$E_{Grid}(i) = \left(\left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)\right) - \left(EP_{Max} \times E_{Number\ of\ Units}\right) + WT_{Energy}(i) - \left(B_{Capacity} \times B_{Number\ of\ Units}\right)$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = B_{storage}(i-1) + (B_{Capacity} \times B_{Number\ of\ Units})$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales

end

else

if $\boldsymbol{B_{storage}(i-1) + (B_{Capacity} \times}$

$\boldsymbol{B_{Number\ of\ Units}) < EP_{Max} \times E_{Number\ of\ Units}}$

$$E_{Grid}(i) = WT_{Energy}(i) - (B_{Capacity} \times B_{Number\ of\ Units})$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = B_{storage}(i-1) + (B_{Capacity} \times B_{Number\ of\ Units})$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = \left(\left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)\right) \times EF_{Max} \times E_{Number\ of\ Units} / \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales

else

if $\boldsymbol{i = Peak\ A\ or\ Peak\ B}$

$$E_{Grid}(i) = \left(\left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)\right) - \left(EP_{Max} \times E_{Number\ of\ Units}\right) + WT_{Energy}(i) - \left(B_{Capacity} \times B_{Number\ of\ Units}\right)$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = B_{storage}(i-1) + (B_{Capacity} \times B_{Number\ of\ Units})$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

<u>*Note: Plant Operating in Mode C – Hydrogen Production with Premium Electricity Sales*</u>

else

if $\boldsymbol{B_{storage}(i-1) + \left(B_{Capacity} \times B_{Number\ of\ Units}\right) - \left(EP_{Max} \times E_{Number\ of\ Units}\right) < B_{Capacity} \times B_{Number\ of\ Units}}$

$$E_{Grid}(i) = WT_{Energy}(i) - \left(B_{Capacity} \times B_{Number\ of\ Units}\right)$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = EP_{Max} \times E_{Number\ of\ Units}$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

<u>*Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales*</u>

else

$$E_{Grid}(i) = WT_{Energy}(i) - \left(B_{Capacity} \times B_{Number\ of\ Units}\right) + \left(\left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)\right) - \left(EP_{Max} \times E_{Number\ of\ Units}\right)$$

$$B_{Charge}(i) = B_{Capacity} \times B_{Number\ of\ Units}$$

$$B_{Discharge}(i) = \left(B_{Capacity} \times B_{Number\ of\ Units}\right) + B_{storage}(i-1)$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$H(i) = EF_{Max} \times E_{Number\ of\ Units}$$

$$Energy\ Balance\ (i) = Energy\ In - Energy\ Out = 0$$

$$Energy\ In\ (i) = WT_{Energy}(i) + B_{storage}(i-1)$$

$$Energy\ Out\ (i) = ((H(i)/EF_{Max}) \times EP_{Max}) + B_{storage}(i) + E_{Grid}(i)$$

<u>*Note: Plant Operating in Mode D – Hydrogen Production with Non-Premium Electricity Sales*</u>

end

end

end

end

```
                end
            end
        end
    end
end
```

Section 2: Wind-Hydrogen (With Energy Storage) Model: Electrolyser/Battery Performance Metrics Equations

Equation 1:

$$Battery\ Storage\ Utilization = \frac{\left(\sum_{i=1}^{8760} B_{storage}(i)/B_{Capacity} \times B_{Number\ of\ Units}\right)}{8760}$$

Equation 2:

$$Battery\ Capacity\ Factor = \frac{\left(\sum_{i=1}^{8760} B_{charge}(i)/B_{Capacity} \times B_{Number\ of\ Units}\right)}{8760}$$

Equation 3:

$$Electrolyser\ Capacity\ Factor = \frac{AH}{EF_{Max} \times E_{Number\ of\ Units} \times 8760}$$

Where:

$$H(i) = Hourly\ hydrogen\ production\ in\ Nm^3/h$$

$$WT_{Energy}(i) = Average\ energy\ produced\ by\ the\ wind\ farm\ in\ hour\ (i)\ in\ kWh;$$

$$E_{Grid}(i) = Energy\ sold\ to\ the\ grid\ in\ hour\ (i)\ in\ kWh;$$

$$B_{storage}(i) = B_{Charge}(i) - B_{Discharge}(i) + B_{storage}(i-1)$$

$$B_{Charge}(i) = Energy\ the\ battery\ is\ charged\ with\ in\ hour\ (i)\ in\ kWh;$$

$$B_{Charge-Min} = Minimum\ amount\ of\ energy\ required\ for\ battery\ charging\ in\ kWh;$$

$$B_{Discharge}(i) = Energy\ discharged\ from\ the\ battery\ in\ hour\ (i)\ in\ kWh$$

$$B_{Storage}(i-1) = Energy\ stored\ in\ the\ battery\ from\ the\ previous\ hour\ (i-1)\ in\ kWh$$

$$(i)\ varies\ from\ 1\ to\ 8760;\ representing\ the\ number\ of\ hours\ in\ a\ year$$

$$B_{Capacity} = The\ energy\ capacity\ of\ the\ battery\ in\ kWh$$

$$B_{Number\ of\ Units} = The\ number\ of\ battery\ units$$

$$AH = Annual\ hydrogen\ production\ in\ Nm^3/year \rightarrow \sum_{i=1}^{8760} H(i)$$

$$EF_{Max} = Maximum\ electrolyser\ flowrate\ in\ Nm^3/h$$

$$EP_{Max} = Maximum\ electrolyser\ power\ in\ kW$$

$$EP_{Min} = Minimum\ electrolyser\ power\ in\ kW$$

$$E_{Number\ of\ Units} = The\ number\ of\ electrolyser\ units$$

$$Peak\ A = Hours\ of\ the\ year\ where\ electricity\ prices\ are\ at\ a\ premium$$

$$Peak\ B = Hours\ of\ the\ year\ where\ electricity\ prices\ are\ at\ a\ premium$$

Section 3: CO_2 & H_2 Compression Models

CO_2 Compressor Power Equation:

$$CP_{CO2(i)} = \left(\frac{1000}{24 \times 3600}\right) \times \left(\frac{mZ_S RT_{in}}{M\eta_{is}}\right) \times \left(\frac{k_s}{k_s - 1}\right) \times \left[(CR)^{\frac{k_s-1}{k_s}} - 1\right]$$

The above compression equation is based on the work carried out by McCollum & Ogden (2006) – see Chapter 4 reference list. The number of compression stages is assumed to be 5 and the optimal compression ratio is given as:

$$CR = \left(P_{cut-off}/P_{initial}\right)\hat{}\left(1/N_{stage}\right)$$

Where:

$$CP_{CO2(i)} = Power\ required\ per\ stage\ of\ compression\ in\ kW$$

$$CP_{CO2-Total} = \sum_{i=1}^{N_{stage}} CP_{CO2(i)}\ in\ kW$$

$$CR = Optimal\ compression\ ratio$$

$$k_s = Average\ ratio\ of\ specific\ heats\ for\ CO_2\ for\ each\ compression\ stage$$

$$i = Compression\ stage\ number$$

$$m = CO_2\ mass\ flow\ rate\ in\ tonnes/day$$

$$M = molecular\ weight\ of\ CO_2\ in\ kg/Kmol$$

$$N_{stage} = Number\ of\ compression\ stages$$

$$P_{initial} = Initial\ CO_2\ pressure\ from\ CO_2\ capture\ system\ in\ MPa$$

$$P_{cut-off} = Pressure\ at\ which\ compression\ switches\ to\ pumping\ in\ MPa$$

$$R = Universal\ Gas\ constant\ in\ \ kJ/Kmol - K$$

$$T_{in} = CO_2\ \ temperature\ at\ compressor\ inlet\ in\ K$$

$$Z_S = Average\ CO_2\ compressibility\ for\ each\ individual\ stage$$

$$\eta_{is} = Compressor\ isentropic\ efficiency$$

Values used in the CO_2 Compression Model

Parameter	**Value**	**Sources/Comments**
CO_2 initial pressure (MPa)	0.1	
CO_2 cut-off pressure (MPa)	7.38	Pressure at which compression switches to pumping
CO_2 flow rate (tonnes/day)	4406	SMR-CCS value
Number of compression stages	5	
CO_2 molar mass (kg/kmol)	44.01	
CO_2 inlet temperature (K)	313.15	
Universal gas constant (kJ/kmol K)	8.314	
Isentropic Efficiency	0.75	

Stage 1 Compression	Value	Comments/sources
Compressibility factor (Z)	0.995	Value corresponds to a compressor internal temperature of 356K and pressure range of 1 -2.4bar. McCollum & Ogden (2006)
Ratio of specific heats	1.277	Value corresponds to a compressor internal temperature of 356K and pressure range of 1 -2.4bar. McCollum & Ogden (2006)
Power Required (kW)	3785.21	
Stage 1 initial Pressure (MPa)	0.10	McCollum & Ogden (2006)
Stage 1 Cut-off Pressure (MPa)	0.24	McCollum & Ogden (2006)
Optimal compression Ratio per compression stage	2.36	McCollum & Ogden (2006)

Stage 2 Compression	Value	Comments/sources
Compressibility factor (Z)	0.985	Value corresponds to a compressor internal temperature of 356K and pressure range of 2.4-5.6 bar. McCollum & Ogden (2006)
Ratio of specific heats	1.286	Value corresponds to a compressor internal temperature of 356K and pressure range of 2.4-5.6 bar. McCollum & Ogden (2006)
Power Required (kW)	3756.29	
Stage 2 initial Pressure (MPa)	0.24	McCollum & Ogden (2006)
Stage 2 Cut-off Pressure (MPa)	0.56	McCollum & Ogden (2006)
Optimal compression Ratio per compression stage	2.36	McCollum & Ogden (2006)

Stage 3 Compression	Value	Comments/sources
Compressibility factor (Z)	0.970	Value corresponds to a compressor internal temperature of 356K and pressure range of 5.6 -13.2 bar. McCollum & Ogden (2006)
Ratio of specific heats	1.309	Value corresponds to a compressor internal temperature of 356K and pressure range of 5.6 -13.2 bar. McCollum & Ogden (2006)
Power Required (kW)	3721.61	
Stage 3 initial Pressure (MPa)	0.56	McCollum & Ogden (2006)
Stage 3 Cut-off Pressure (MPa)	1.32	McCollum & Ogden (2006)

Stage 3 Compression	Value	Comments/sources
Optimal compression Ratio per compression stage	2.36	McCollum & Ogden (2006)

Stage 4 Compression	Value	Comments/sources
Compressibility factor (Z)	0.935	Value corresponds to a compressor internal temperature of 356K and pressure range of 13.2 - 30.2bar. McCollum & Ogden (2006)
Ratio of specific heats	1.379	Value corresponds to a compressor internal temperature of 356K and pressure range of 13.2 - 30.2bar. McCollum & Ogden (2006)
Power Required (kW)	3649.89	
Stage 4 initial Pressure (MPa)	1.32	McCollum & Ogden (2006)
Stage 4 Cut-off Pressure (MPa)	3.02	McCollum & Ogden (2006)
Optimal compression Ratio per compression stage	2.36	McCollum & Ogden (2006)

Stage 5 Compression	Value	Comments/sources
Compressibility factor (Z)	0.845	Value corresponds to a compressor internal temperature of 356K and pressure range of 30.2 - 73.8 bar. McCollum & Ogden (2006)
Ratio of specific heats	1.704	Value corresponds to a compressor internal temperature of 356K and pressure range of 30.2 - 73.8 bar. McCollum & Ogden (2006)
Power Required (kW)	3511.03	
Stage 4 initial Pressure (MPa)	3.02	McCollum & Ogden (2006)
Stage 4 Cut-off Pressure (MPa)	7.38	McCollum & Ogden (2006)
Optimal compression Ratio per compression stage	2.36	McCollum & Ogden (2006)
Total Compressor Power Required (MW)	18.424	

CO_2 Pump Power Equation:

$$PP_{CO2} = \left(\frac{1000 \times 10}{24 \times 36}\right) \times \left[\frac{m(P_{final} - P_{cut-off})}{\eta_{pump} \times \rho_{CO2}}\right]$$

The above compression equation is based on the work carried out by McCollum & Ogden (2006). Where:

$$m = CO_2\ mass\ flow\ rate\ in\ tonnes/day$$

$$PP_{CO2} = Pump\ Power\ in\ kW$$

$$P_{cut-off} = Pressure\ at\ which\ compression\ switches\ to\ pumping\ in\ MPa$$

$$P_{final} = Final\ pressure\ of\ CO_2\ for\ intended\ application\ in\ MPa$$

$$\eta_{pump} = Pump\ efficiency$$

$$\rho_{CO2} = CO_2\ density\ in\ kg/m^3$$

Note that in the model developed, P_{final} corresponds to the inlet pressure for pipeline transportation - 15MPa.

Parameter	**Value**	**Sources/Comments**
CO_2 flow rate (tonnes/day)	4406	SMR-CCS Value
CO_2 liquid density (kg/m^3)	630	
CO_2 cut-off pressure (MPa)	7.38	Pressure at which compression switches to pumping
CO_2 final pressure (MPa)	15	
Pump efficiency	0.75	

H_2 Compressor Power Equation:

$$CP_{H2} = Q_{H2} \times \frac{0.164}{n_c} \times N \times \left[(P_{out}/P_{in})^{0.291}/_{N} - 1 \right]$$

The above compression equation is based on the work carried out by Ogden (2004) – see Chapter 4 reference list. Where:

$$CP_{H2} = H_2 \ compressor \ power \ in \ kW$$

$$Q_{H2} = H_2 \ flow \ rate \ in \ standard \ cubic \ feet \ per \ minute$$

$$n_c = Compressor \ efficiency$$

$$N = Number \ of \ compressor \ stages$$

$$P_{out} = Outlet \ pressure \ from \ compressor$$

$$P_{in} = Inlet \ pressure \ for \ compressor$$

Parameter	**Value**	**Sources/Comments**
H_2 flow rate (scf/min)	40,257	Value for Hydropower-H_2
Compressor efficiency	0.7	Ogden (2004)
Number of compressor stages	2	Ogden (2004)
Inlet pressure (bar)	30	Outlet pressure for 3496 kW electrolyser
Outlet pressure (bar)	60	Hydrogen pipeline inlet pressure
Compressor power (kW)	2002	

Section 4: Pipeline Transport of H_2 and CO_2 - Modelling Equations

Hydrogen Pipeline - Panhandle B Equation:

From the equations provided by Schroeder (2001) - see Chapter 4 reference list, the Panhandle B equation (assuming a negligible static head) is given below:

$$Q_{H2} = 737 \times \left(\frac{T_b}{P_b}\right)^{1.02} \times D^{2.53} \times e \times \left(\frac{P_1^2 - P_2^2}{L \times G^{0.961} \times T_a \times Z_a}\right)^{0.51}$$

Where:

$$D = Pipeline\ diameter\ in\ inches$$

$$e = Pipeline\ efficiency$$

$$G = Specfic\ gravity\ of\ hydrogen$$

$$L = Pipeline\ length\ in\ miles$$

$$P_b = Reference\ pressure\ in\ psi$$

$$P_1 = Pipeline\ inlet\ pressure\ in\ psi$$

$$P_2 = Pipeline\ outlet\ pressure\ in\ psi$$

$$Q_{H2} = H_2\ flowrate\ in\ standard\ cubic\ feet\ per\ day$$

$$T_b = Reference\ temperature\ in\ degrees\ rankine$$

$$T_a = Average\ gas\ temperature\ in\ degrees\ rankine$$

$$Z_a = Compressibility\ factor$$

Values used in the Panhandle B Model:

Parameter	Value	Sources/Comments
H_2 flow rate (scf/day)	261,879,101	Value for UCG-H_2
Pipeline efficiency	0.92	Sarkar & Kumar (2009) – Techno-Economic Model
H_2 specific gravity	0.0696	
Reference temperature (R)	530	
Reference pressure (psi)	14.7	
Compressibility factor	1.035273	Sarkar & Kumar (2009) Techno-Economic Model
Pipeline length (miles)	140	
H_2 pipeline inlet pressure (psi)	870.23	
H2 pipeline outlet pressure (psi)	725.18	
Gas average (H_2) temperature (R)	537	
Calculated pipe diameter (inches)	18.21	

CO_2 Pipeline Model:

The CO_2 pipeline model provided by McCollum & Ogden (2006) is solved iteratively with an initial educated guess for the pipeline diameter that corresponds to the flow rate of 11,302 tonnes/day – UCG-CCS value. The equation is given below as:

$$D = \frac{1}{0.0254} \times \left[\left(32 \times f \times \left(\frac{m \times 1000}{24 \times 3600} \right)^2 \right) / \left(\left(\frac{P_1 - P_2}{L} \right) \times \frac{10^6}{1000} \times \pi^2 \times \rho_{CO2} \right) \right]^{1/5}$$

Where:

$$f = \frac{1}{4\left[-1.8 \log_{10} \left(\frac{6.91}{Re} + \left(\frac{12(\varepsilon/D)}{3.7} \right)^{1.11} \right) \right]^2}$$

$$Re = (4 \times 1000/24 \,/\, 3600/0.0254) \times m/(\pi \times \mu \times D)$$

$$D = Pipeline\ diameter\ in\ meters$$

$$f = friction\ factor$$

$$L = Pipeline\ length\ in\ kilometers$$

$$m = \ CO_2\ mass\ flow\ rate\ in\ tonnes/day$$

$$P_1 = Pipeline\ inlet\ pressure\ in\ MPa$$

$$P_2 = Pipeline\ outlet\ pressure\ in\ MPa$$

$$Re = Reynolds\ number$$

$$\rho_{CO2} = CO_2\ density\ in\ kg/m^3$$

$$\varepsilon = pipeline\ roughness\ in\ feet$$

$$\mu = CO_2\ viscosity\ in\ Pa - s$$

Values used in the CO_2 Pipeline Model:

Parameter	**Value**	**Sources/Comments**
CO_2 length (km)	10	
CO_2 pipeline initial pressure (MPa)	15	
CO_2 pipeline outlet pressure (MPa)	10.3	
CO_2 critical pressure (MPa)	7.38	
CO_2 critical temperature (C)	31.12	
CO_2 pipeline internal pressure (MPa)	12.65	
CO_2 pipeline internal Temperature (C)	25	Heddle, G., H. Herzog, and M. Klett (2003) – See Chapter 4 reference list
CO_2 density (kg/m3)	884	Heddle, G., H. Herzog, and M. Klett (2003)
CO_2 viscosity(Ns/m2)	0.0000606	Heddle, G., H. Herzog, and M. Klett (2003)
Pipe diameter Guess (inches)	8.494	
Pipe diameter Guess (m)	0.2156	
Pipe surface roughness factor (ft)	0.00015	
Calculated Diameter (inches)	8.492	
Calculated Diameter (m)	0.2156	

Solution Iterations

Iteration No.	Diameter Guess (inches)	Diameter Calculated (inches)	Error (%)
1	10	8.439	-18.5
2	8.439	8.494	0.65
3	8.494	8.492	-0.02

Section 5: Discounted Cash Flow (DCF) Sheet Sample - Hydropower-H2

Year End	1	2	3	4	5	6	7	8
Investment Cost ($)	(215,753,107.31)	(1,186,642,090.20)	(755,135,875.58)					
Hydropower Plant O&M ($/yr)				(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)
Electrolyser O&M ($/yr)				(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)
Pipeline O&M ($/yr)				(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)
Compressor O&M ($/yr)				(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)
Water Resource Cost ($/yr)				(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)
Electricity Revenue ($/yr)				14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09
Hydrogen Revenue ($/yr)				129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27
Oxygen Revenue ($/yr)				215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00
Net Cash Flow ($/yr)	(215,753,107.31)	(1,186,642,090.20)	(755,135,875.58)	293,576,890.98	293,576,890.98	293,576,890.98	293,576,890.98	293,576,890.98
Present Value ($/yr)	(271,608,134.74)	(1,331,412,425.20)	(755,135,875.58)	261,654,983.05	233,204,084.72	207,846,777.82	185,246,682.55	165,103,995.14
Net Present Value ($)	**0.00**	**Nominal Discount Rate**	10%	**Inflation Adjusted Discount Rate**	12.2%			
Hydrogen Production Cost ($/kg H2)	**2.43**	**Inflation**	2%					

N.B: The above DCF sheet is continued on the next page.

Year End	9	10	11	12	13	14	15	16	17
Investment Cost ($)			(2,102,152.30)			(56,722,548.39)			
Hydropower Plant O&M ($/yr)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)
Electrolyser O&M ($/yr)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)
Pipeline O&M ($/yr)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)
Compressor O&M ($/yr)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)
Water Resource Cost ($/yr)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)
Electricity Revenue ($/yr)	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09
Hydrogen Revenue ($/yr)	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27
Oxygen Revenue ($/yr)	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00
Net Cash Flow ($/yr)	293,576,890.98	293,576,890.98	291,474,738.68	293,576,890.98	293,576,890.98	236,854,342.59	293,576,890.98	293,576,890.98	293,576,890.98
Present Value ($/yr)	147,151,510.82	131,151,079.16	116,053,452.86	104,180,432.16	92,852,435.08	66,766,702.18	73,757,737.08	65,737,733.58	58,589,780.38

N.B: The above DCF sheet is continued on the next page.

Year End	18	19	20	21	22	23	24	25	26
Investment Cost ($)	(2,102,152.30)						(56,722,548.39)	(2,102,152.30)	
Hydropower Plant O&M ($/yr)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)
Electrolyser O&M ($/yr)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)
Pipeline O&M ($/yr)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)
Compressor O&M ($/yr)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)
Water Resource Cost ($/yr)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)
Electricity Revenue ($/yr)	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09
Hydrogen Revenue ($/yr)	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27
Oxygen Revenue ($/yr)	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00
Net Cash Flow ($/yr)	291,474,738.68	293,576,890.98	293,576,890.98	293,576,890.98	293,576,890.98	293,576,890.98	236,854,342.59	291,474,738.68	293,576,890.98
Present Value ($/yr)	51,845,141.94	46,541,047.77	41,480,434.73	36,970,084.43	32,950,164.37	29,367,347.93	21,116,958.01	23,161,040.68	20,791,516.05

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Year End	27	28	29	30	31	32	33	34	35
Investment Cost ($)						(2,102,152.30)		(56,722,548.39)	
Hydropower Plant O&M ($/yr)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)
Electrolyser O&M ($/yr)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)
Pipeline O&M ($/yr)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)
Compressor O&M ($/yr)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)
Water Resource Cost ($/yr)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)
Electricity Revenue ($/yr)	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09
Hydrogen Revenue ($/yr)	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27
Oxygen Revenue ($/yr)	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00
Net Cash Flow ($/yr)	293,576,890.98	293,576,890.98	293,576,890.98	293,576,890.98	293,576,890.98	291,474,738.68	293,576,890.98	236,854,342.59	293,576,890.98
Present Value ($/yr)	18,530,762.97	16,515,831.52	14,719,992.44	13,119,422.85	11,692,890.24	10,346,848.04	9,288,298.40	6,678,866.88	7,378,200.37

N.B: The above DCF sheet is continued on the next page.

Year End	36	37	38	39	40	41	42	43
Investment Cost ($)				(2,102,152.30)				
Hydropower Plant O&M ($/yr)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)	(47,411,200.00)
Electrolyser O&M ($/yr)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)	(5,348,880.00)
Pipeline O&M ($/yr)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)	(6,320,654.85)
Compressor O&M ($/yr)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)	(1,311,531.37)
Water Resource Cost ($/yr)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)	(6,056,804.16)
Electricity Revenue ($/yr)	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09	14,661,940.09
Hydrogen Revenue ($/yr)	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27	129,657,781.27
Oxygen Revenue ($/yr)	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00	215,706,240.00
Net Cash Flow ($/yr)	293,576,890.98	293,576,890.98	293,576,890.98	291,474,738.68	293,576,890.98	293,576,890.98	293,576,890.98	293,576,890.98
Present Value ($/yr)	6,575,936.16	5,860,905.67	5,223,623.59	4,622,299.40	4,149,408.20	3,698,224.78	3,296,100.51	2,937,700.99

Author's Publication List

Book Chapter

1. **Olateju B.**, Kumar A. Clean Energy Based Production of Hydrogen – An Energy Carrier. In: Yan J. (Ed.). The Handbook of Clean Energy Systems 2015, John Wiley & Sons, Ltd., Chichester, United Kingdom, 1-30.

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1. **Olateju, B.**, Kumar A., Secanell M. A techno-economic assessment of large scale wind-hydrogen production with energy storage in Western Canada. *International Journal of Hydrogen Energy 2016; 41: 8755-8776*
2. **Olateju, B.** and Kumar A. A techno-economic assessment of hydrogen production from hydropower in Western Canada for the upgrading of bitumen from oil sands. *Energy, 2016 (in-review).*
3. **Olateju B.**, Monds J., Kumar A. Large scale hydrogen production from wind energy for upgrading bitumen from oil sands. *Applied Energy 2014; 118: 48-56*
4. **Olateju B.**, Kumar A. Techno-economic assessment of hydrogen production from underground coal gasification (UCG) in Western Canada with carbon capture and sequestration (CCS) for upgrading bitumen from oil sands. *Applied Energy 2013; 111: 428-440*
5. **Olateju B.**, Kumar A. Hydrogen production from wind energy in Western Canada for upgrading bitumen from oil sands. *Energy* 2011; 36: 6326-6339.
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1. **Olateju B.,** Kumar A. Development of baseline costs for hydrogen production, delivery, and carbon capture from Alberta based natural gas and coal. Final Report submitted to the Alberta Innovates – Energy and Environment Solutions (EES), 1800 Phipps Mckinnon Building, 10020 101A Avenue, T5J Edmonton, Alberta, February 2012.
2. Verma A., Nimana B. **Olateju B.**, Rahman M., Radpour S., Canter C., Subramanyam V., Paramashivan D., Vaezi M., Kumar A. Transportation of bitumen/synthetic crude oil (SCO) via rail or pipeline. Final Report submitted to Technical Advisory Committee (TAC) for NSERC/Cenovus/Alberta Innovates Associate Industrial Research Chair Program in Energy and Environmental Systems Engineering. May 2014.