Technoeconomic Modelling of Future Fuel Production Pathways: Summary Report

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IMPORTANT DISCLAIMER

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SUMMARY OF REPORT

This report is a summary of the work done in the Future Fuels Cooperative Research Centre's (FFCRC) Technoeconomic Modelling of Future Fuel Production Pathways project (RP1.2-02). The original work was done during 2019-21 in partnership with The University of Queensland, the University of Adelaide, and the University of Melbourne with advice from several industry partners, and then supplemented and revised in 2022 to refine the analysis of electrolysis pathways and hydrogen carriers. The project sought to develop technoeconomic models of low CO_2 emission hydrogen production processes. Models were developed for a variety of feedstocks to allow for quantitative economic comparison. This contributed to the FFCRC's mission statement of *"enable[ing] Australia's energy sector to adapt its infrastructure to net zero emissions fuels by providing new knowledge and facilitating its use by industry".*

A total of 22 production pathways and nine (9) hydrogen carrier scenarios were investigated. Production feedstocks included water (electrolysis), biomass (pyrolysis & gasification), natural gas (pyrolysis & reforming) and coal (pyrolysis & gasification). Cases involving a fossil feedstock required carbon capture and storage (CCS) to reduce emissions, for which energy penalties and associated capital and operating costs were accounted. For biomass pathways, the cost of adding CCS to yield negative emissions was investigated. Additional hydrogen carriers were considered including ammonia, methanol and liquid hydrogen.

Table 0.1 presents the most competitive technology and levelized cost of hydrogen (LCOH) for each feedstock based on consistent input assumptions at the time of the completion of the original study in 2021 (see Section 1.2 for an overview of key assumptions).

Table 0.1 Most competitive LCOH for each feedstock

Feedstock	Pathway	LCOH (\$/kg-H_2)[1]
Natural gas (NG) / Coal seam gas (CSG)	Autothermal reforming with CO_2 capture optimised for process efficiency	2.92
Coal	Gasification of Victorian Brown Coal in an entrained flow gasifier.	4.12
Biomass	Gasification of biomass in a dual fluidised bed gasifier.	4.66
Water	Alkaline electrolysis	6.38

Modelling of hydrogen carriers found ammonia to be the most cost-effective carrier on a mass basis, adding only \$1.55/kg-H_2. Liquified hydrogen and methanol[2] added \$2.27/kg-H_2 and \$2.06/kg-H_2 to the LCOH respectively. However, these costs did not include the cost of hydrogen production (only conversion to the carrier), transport as the carrier to the desired destination, or the cost of reversion back to molecular hydrogen at the destination.

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[1] As at study completion in 2021

[2] CO_2 used for methanol production was either biogenic in origin or from direct air capture. Fossil CO_2 was not used.

METHANE

An analysis of two key technology pathways for hydrogen production from methane were completed: reforming and methane pyrolysis. Reforming is the primary source of hydrogen today globally by volume; however, in this study only reforming pathways integrated with CCS where modelled. Methane pyrolysis, where methane is decomposed at high temperatures to form hydrogen and solid carbon, was considered as an emerging technology, rising in prominence due to recent laboratory investigations and development of pilot scale projects.

The reforming technologies considered were **steam methane reforming (SMR)** and **autothermal reforming (ATR)**. Hydrogen production was maximised in all cases using water gas shift (WGS) reactors. In addition, the **ATR-OP** scenario investigated a case where energy efficiency was maximised (through superior heat integration and recycling of off-gases) at the expense of CO_2 capture efficiency (which reduced from 90% to 84%).

For methane pyrolysis, the two pathways evaluated used either solid catalyst or molten catalyst systems. **Solid catalyst pyrolysis (SCP)** utilised a fluidised bed design while **molten media pyrolysis (MMP)** used bubble column reactors with a combined mixture of molten metals – the catalytic driver of the system, and molten salts – for carbon recovery and heat integration. Both pathways were assessed with both conventional natural gas (**NG**) and coal seam gas (**CSG**) feedstocks, as well as renewable electricity (**RE**) integration for heat provision. A key scenario also considered the production of a **low-purity**, raw hydrogen-methane blend for direct injection into the existing natural gas network (**MMP-BL**). The benefit of this scenario is that no hydrogen purification is required, resulting in a simplified process design with implications for operational and capital expenditure.

These processes were investigated using process modelling tools and all scenarios were considered on a 100 kta H_2 production basis. Electricity prices (grid-$85/MWh), natural gas ($6/GJ), and utility costs were uniform across all scenarios. The primary findings of the mass and energy balances are shown in Table 0.2 and Table 0.3. The process name is given by feedstock-technology, e.g. natural gas fed ATR is denoted NG-ATR.

Table 0.2 Reforming Mass and Energy Balance Summary

Result	NG-S-SMR	NG-C-SMR	NG-ATR	NG-ATR-OP
H_2 produced (kta)	100	100	100	100
Gas input (kta)	390	379	359	347
CO_2e (kg-CO_2/t-H_2)	930	920	750	1,400

Table 0.3 Methane Pyrolysis Mass and Energy Balance Summary

Result	NG-SCP	CSG-SCP	CSG-SCP+RE	NG-MMP	NG-MMP-BL	CSG-MMP	CSG-MMP+RE
H_2 produced (kta)	100	100	100	100	100	100	100

Gas input (kta)	576	533	438	595	267	550	458
CO_2e (kg-CO_2/t-H_2)	231	230	600	310	50	300	1130

For the reforming technologies, it was found that there is comparable performance regardless of configuration. The largest change came with optimization (for energy) of CCS integration with **ATR**, which reduced the natural gas input flowrate, but increased the overall emissions intensity. Combined and Standalone **SMR** had comparable performance, with the integration of **ATR** slightly reducing gas input and emissions intensity. For pyrolysis, it is apparent that SCP cases have low emissions intensities. **CSG-SCP+RE** has a higher emissions intensity as no CCS is used in this system due to the integration of renewable energy for heat provision, whereas >90% capture efficiency is used in other scenarios where tail gas provides the heat. **NG-MMP** produces a low emissions process for hydrogen production, with emissions only resulting from onsite combustion of fuel for heat and energy. Again, the integration of renewable energy with the omission of CCS results in a higher emissions process as no CCS is adopted, showing a low emissions pathway that negates reliance on CCS. Lastly, the **NG-MMP-BL** pathway, which did not consider hydrogen purification, was able to produce a maximum hydrogen concentration of approximately 79 mol% with 17.8 mol% of methane. This reduced purity ultimately means that there is a reduction of feed gas required. However, the reduced operating load of the purification units does mean that there is a reduced energy load for the process. This is represented by the reduced process emissions.

The process economics were analysed using a discounted cash flow model. Key learnings from the evaluation are as follows:

- **NG-MMP-BL** is an anomaly in performance, due to the $1.58/kg H_2 LCOH
- The low production cost of the **NG-MMP-BL** pathway is associated with the low purity, as there is no hydrogen purification which is 15%-28% of the CAPEX in the **MMP** alternatives
- This process design presents a low-cost hydrogen blending solution to partially decarbonise industrial and household gas use
- Natural gas price and CAPEX are the main economic drivers for natural gas to hydrogen pathways
- High natural gas prices significantly increase LCOH
- Lower CAPEX through deployment of technologies to improve learning rates has the potential to greatly enhance process economics for emerging technologies
- The sale price of the solid carbon by-product was a key economic driver for all pyrolysis scenarios
- Therefore, enhancing the carbon product value through further investigations is warranted noting that: the carbon sale price has high uncertainty around product quality; and, secondly, purification of the solid carbon is unlikely to be appropriately costed.

- Technologies that integrated renewable electricity (**CSG-SCP+RE** and **CSG-MMP+RE**) had lower CAPEX as CCS is not feasible on the small and dilute process emissions streams. Counterintuitively, these process schemes had a larger emissions penalty due to all process emissions
- The cheapest pure hydrogen pathway was **ATR-OP** with a LCOH of \$2.92/kg H_2
- The reduction in CCS CAPEX was the key driver for minimising the LCOH
- The lowest cost for emerging technologies (producing high purity hydrogen) was **CSG-SCP+RE** at \$3.68/kg H_2 and CSG-MMP+RE at \$3.71/kg H_2

COAL

Blue hydrogen can be produced through gasification of coal, with the implementation of carbon capture and storage. Coal gasification involves heating coal to high temperatures in the presence of an oxidising agent (water and/or oxygen) to produce syngas. The hydrogen content of the syngas is increased through a water gas shift reaction. CO_2 is then removed for compression and storage. The hydrogen undergoes purification through pressure swing adsorption before being compressed to 8 MPa.

Four (4) cases of hydrogen production through coal gasification were investigated; two (2) involving an **entrained flow gasifier (EFR)** and two (2) involving a **dual fluidised bed gasifier (DFB)**. EFR was investigated with both **Queensland black coal (QBC)** and **Victorian brown coal (VBC)** while DFB was investigated with only VBC. One DFB scenario was run with **in-situ carbon capture** (**VBC-DFB-CAP**) while the other implemented **Selexol carbon capture**. Both EFR scenarios utilised Selexol carbon capture.

The processes were investigated using process modelling tools with a production basis of 100kta of hydrogen. Electricity prices (grid-\$85/MWh), coal (\$1.5/GJ VBC and \$3.5/GJ QBC) and utility costs were uniform across all scenarios. The primary findings of the mass and energy balances are shown in Table 0.4. The process name is given by feedstock-technology, e.g. Queensland black coal fed entrained flow gasifier is denoted **QBC-EFR**.

Coal consumption of QBC was over **two times less** than VBC consumption in the respective processes due to the much higher moisture content of VBC. On a dry, ash-free basis coal consumption of VBC was 10-20% higher than QBC depending on the process route. Electricity consumption of the EFR cases was highest due to the air separation unit (ASU) demand; however, the DFB cases also had high auxiliary power demand. DFB cases utilise natural gas as fuel to supply the balance of energy to the gasifier unit and, for the in-situ carbon capture process, to regenerate the solid $CaO/CaCO_3$ sorbent.

CO_2 is produced during the gasification process, in the combustion of tail gas and, for DFB cases only, in the combustion of natural gas. As the majority of process CO_2 is produced in the gasification units for the EFR cases, the high pressure, high efficiency Selexol process is used. As such, these processes have very high CO_2 capture rates of >99%. The DFB cases have a lower capture rate of 94-96%. The QBC process has the lowest levelized CO_2 emissions **(0.18 t-CO_2/t-H_2)**, closely followed by the EFR gasification of VBC **(0.23 t-CO_2/t-H_2)**. The DFB process with Selexol CO_2 capture is much higher, at **0.68 t-CO_2/t-H_2**, while the DFB process with in-situ CO_2 capture lags at **1.08 t-CO_2/t-H_2**.

Table 0.4 Coal Gasification Mass and Energy Balance Summary

Result	**QBC-EFR**	**VBC-EFR**	**VBC-DFB**	**VBC-DFB-CAP**
H_2 produced (kta)	100	100	100	100
Coal input (kta, wet)	843	1959	2017	2119
CO_2e (kg-CO_2/t-H_2)	0.18	0.23	0.68	1.09

The capital expense of a DFB is much higher than an EFR, owing to the technology's novelty. Coupled with the requirement for CO_2 removal from syngas, the capital expense of the DFB scenario with Selexol CO_2 capture is the **highest**. Since QBC requires no drying, the capital expense savings associated with coal pre-treatment yields a very **similar capital cost between the EFR scenarios**. Both are **less expensive than the DFB** scenarios.

Utility costs for all scenarios are **dominated by electricity costs**. Cooling water make up and raw water demand both make up **very small portions** of utility cost in all scenarios. The natural gas demand of DFB scenarios increase their utility costs **beyond the EFR scenarios**. Despite having a slightly lower electricity cost than the DFB process employing Selexol capture, the DFB process with in-situ CO_2 capture has the highest utility cost due to its **greater natural gas demand**.[3]

Levelized cost of hydrogen (LCOH) is **driven largely by the capital cost**, accounting for 35-44% of the cost. Variable operational expenses also have a substantial impact on LCOH – the high utility cost of the DFB process with in-situ CO_2 capture stands out. **Coal feedstock costs are relatively insignificant** (<10%) to the LCOH, except for the QBC case where the cost of coal accounts for 22% of the LCOH. **Fixed operational expenses are roughly equal** between scenarios, except for the DFB with Selexol CO_2 capture which has higher fixed operational costs due to greater process complexity. Gasification of VBC in an EFR yields the lowest LCOH ($4.12/kg), followed by VBC gasification in a DFB with in-situ CO_2 capture ($4.48/kg). QBC gasification in an EFR gives an LCOH of $4.89/kg followed by VBC gasification in a DFB with Selexol CO_2 capture ($4.83/kg). QBC gasification in an EFR and VBC gasification in a DFB, both with Selexol CO_2 capture yield very similar LCOH at $4.89/kg and $4.83/kg respectively.

A sensitivity analysis found the LCOH to be most sensitive to variations in CAPEX, electricity price and coal price. QBC is only becomes competitive with VBC if the price of QBC drops by 70% (to $1.05/GJ). LCOH demonstrates minimal sensitivity to CO_2 storage price.

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[3] Additional natural gas is required to regenerate the CaO and remove CO_2 from the $CaCO_3$ after capture.

BIOMASS

Biomass has a great potential for conversion into green hydrogen. Two main thermochemical pathways exist for green hydrogen production from biomass: pyrolysis and gasification. Both methods involve the high temperature decomposition of biomass. Gasification adds steam into the reaction chamber while pyrolysis occurs without the addition of oxidising agents. Processing is required to remove tars produced through the process. Common methods are steam tar reforming and OLGA treatment. The former involves the addition of steam to tar to chemically convert it into syngas. OLGA treatment removes tar from the syngas through adsorption, after which the tar is combusted for additional steam generation in the plant which can be used for heat integration or additional electricity production.

Five (5) processes have been investigated for the production of green hydrogen from biomass; four (4) involve **pyrolysis** (P) while one (1) employs **gasification** (G). In the pyrolysis scenarios, the tar treatment (**steam tar reforming** (STM) or **OLGA treatment** (OLGA)) and carbon capture extent (LO/HI) was varied to create four (4) unique scenarios. As biomass is an inherently green feedstock, high carbon capture rates are not necessary, however a high level of capture could make the process carbon negative. Hence, two variations of carbon capture were investigated including capture of CO_2 from the **syngas only** (LO) or from the **syngas and flue gas** (HI). The gasification scenario utilised OLGA for tar treatment and removed CO_2 from the **syngas only**.

Syngas produced in each process also underwent a **water gas shift** to increase hydrogen yield and **pressure swing adsorption** for purification before compression to meet the required H_2 purity. For all scenarios, **sugarcane bagasse** was used as the feedstock, because it was one of the lowest cost, higher availability feedstocks that was sufficiently well characterised for both pyrolysis and gasification.

The processes were investigated using process modelling tools with a production basis of 25kta of hydrogen produced. Electricity prices (grid-$85/MWh), feedstock prices ($0.1/GJ) and utility costs were uniform across all scenarios. The primary findings of the mass and energy balances are shown in Table 0.7. The process name is given by feedstock-technology, e.g. pyrolysis employing steam tar reforming with carbon capture only on the syngas stream is denoted BIO-P-STM-LO.

Table 0.5 Biomass Mass and Energy Balance Summary

Result	**BIO-P-STM-LO**	**BIO-P-STM-HI**	**BIO-P-OLGA-LO**	**BIO-P-OLGA-HI**	**BIO-G**
H_2 produced (kta)	25	25	25	25	25
Biomass input (kta, wet)	597	597	1319	1319	870.5
CO_2e (kg-CO_2/t-H_2)	14.34	2.24	33.06	7.54	17.25
Net CO_2 Capture Efficiency (%)	46.8	91.7	36.2	85.4	48.5

Requirements for bagasse were lowest for pyrolysis cases with steam tar reforming. Compared to OGLA treatment where tar is combusted and **hydrogen in tail gases vented** to atmosphere, steam tar reforming **harnesses the hydrogen in tar** for production of additional hydrogen. Hence, the steam tar reforming process has a **greater biomass utilisation** than the other processes, almost twice that of the OLGA scenarios, and ~45% better than the gasification case (which has a **lower H_2 conversion rate** than pyrolysis, but still employs steam tar reforming).

Importantly, there is a substantial difference in CO_2 emissions between processes. The processes where **carbon capture is employed on both syngas and flue gas have the lowest levelized emissions**. Again, the steam tar reforming with pyrolysis scenario has less than half the emissions of the OLGA equivalent and is between 5 and 14 times better than the LO capture scenarios where CO_2 is only captured from the syngas.

Electricity dominates the utility demand of all scenarios. **Pyrolysis scenarios have higher electricity demand than the gasification scenario** owing to greater compressor requirements. OLGA scenarios generate significant amounts of extra electricity through combustion of the tars to produce steam. Hence, these scenarios have much lower total electricity demand than steam tar reforming scenarios. Scenarios with **higher rates of carbon capture have higher utility demands** than their respective counterparts. The steam tar reforming case where carbon capture is applied only to the syngas (BIO-P-STM-LO), has the lowest operating cost for the pyrolysis scenarios while the OLGA treatment with syngas and flue gas carbon capture (BIO-P-OLGA-HI) has the highest operating cost.

Despite utilising the same pyrolysis technology, the cost of the pyrolysis unit for OLGA cases is greater (more than double) than the steam tar reforming as a larger pyrolysis unit is required due to the lower biomass utilisation and therefore higher biomass feed rate. Despite the capital cost of the gasifier being greater than the smaller pyrolysis units, the gasification scenario has a similar capital cost to the cheapest pyrolysis scenario, owing to reduced costs around carbon capture equipment. **Carbon capture equipment** accounts for up to a **third of reduced capture (LO) scenarios** and over a **half of high carbon capture (HI) scenarios**.

The gasification has a lower LCOH than all pyrolysis scenarios, at **\$4.66/kg**. The most competitive pyrolysis case BIO-P-STM-LO produced hydrogen at **\$4.94/kg**. followed by BIO-P-OLGA-LO, at **\$6.38/kg**, and then BIO-P-STM-HI at **\$6.89/kg**. BIO-P-OLGA-HI was by far the least cost effective with an LCOH of **\$9.63/kg**. **CAPEX was the dominant component of LCOH** of all biomass scenarios. **Feedstock costs made a negligible contribution** to LCOH.

A sensitivity analysis indicated that only **electricity price** and **CAPEX** had notable impacts on LCOH. Biomass (feedstock) price, CO_2 sequestration incentives to generate negative emissions and CO_2 storage costs all had negligible impacts on LCOH over their respective ranges. **Electricity price** was the only parameter found to **change the competitiveness** order between technologies, with the LCOH of BIO-P-STM-LO lower than BIO-G when the grid electricity price was reduced to \$25/MWh.

ELECTROLYSIS

The electrolysis of water is an electrochemical process by which electricity breaks water (H_2O) into hydrogen (H_2) and oxygen (O_2). If this electricity is generated renewably, the hydrogen can be considered *green*. The two most prominent and commercially mature electrolysis processes are **proton exchange membrane (PEM) electrolysers** — using a solid polymer cation electrolyte, and **alkaline electrolysers (AE)** — using an alkali electrolyte (KOH or NaOH). These were the preferred technologies and formed the basis for the process and techno-economic modelling.

An electrolyser, at its simplest unit, is made up of electrolyser *cells*—including the anode, cathode, membrane, and electrolyte. These cells are then connected in series to form an electrolyser *stack*, which in turn are connected together until the target hydrogen production capacity is reached. To enable this system to continuously generate hydrogen, additional ancillary equipment (*balance of plant*) is required such as feed water treatment, oxygen separation, moisture removal, stack cooling, and power supply management. Each electrolyser had a discharge pressure of 10 bar and additional compression requirements to attain pipeline specifications (comparable with other scenarios in the study) was included in the energy balances and cost modelling.

Both technologies were modelled using process modelling tools for a production basis of 25 kta hydrogen production. Based on the assumed plant efficiencies (53.5% for PEM and 58.5% for AE) at the time of the study in 2021 and target production quantities, the required **electrolyser capacities were 187 and 171 MW$_e$ for PEM and AE respectively**.

Equipment costs were estimated for all equipment inside the battery limits by using a combination of industry surveys, literature sources, and process modelling tools. Current electrolyser capital costs are very dynamic with costs fluctuating with supply chain constraints and improved learnings. As of the time of completing this summary report, PEM electrolysers were more capital intensive than alkaline electrolysers by around \$200/W$_e$.

Operational costs included electricity, water, and the electrolyte in some cases. The electricity cost was the dominant operating cost, contributing **>85% of the operating cost per kilogram of hydrogen produced**. Due to the differences in stack efficiencies, electricity cost is also the differentiating factor between the cases, while the fixed maintenance, stack replacement and water/electrolyte costs are comparable between the cases.

The **LCOH was \$6.99 and \$6.38 for PEM and AE respectively**. The key drivers of this cost difference, and overall LCOH were:

1. **Operating cost**—contributing approximately 75% to the LCOH, this cost is dominated by the electricity cost.
2. **Capital cost**—contributing approximately 18% to the LCOH, this cost is dominated by the electrolyser stack and equipment costs.

This was validated using sensitivity analysis which varied key factors. It was found that electricity cost and capital expenditure led to the greatest change in LCOH, while changes to water cost had little to no impact. It is important to note that when aiming to achieve low LCOH (<\$2.0/kg-H_2), potential **future cost reductions due to technological advancements or scale up are unlikely to compensate for high electricity prices**.

Furthermore, the basis of the process and techno-economic modelling was a high utilisation factor (90%) such that the facility had to be grid connected and have a renewable supply agreement to ensure green hydrogen. However, in practice and at these scales, true renewable electricity is expected to come behind the meter and an additional[4] six power supply models were used to investigate the impact of this arrangement on (1) electrolyser sizing and (2) project economics. To meet the required production quantity using behind the meter renewable electricity (i.e. sacrificing utilisation factor to be equivalent to the corresponding renewable energy capacity factor), **electrolyser capacities ranged between 300MW_e and 600MW_e resulting in LCOHs between $6.19 and $11.88 per kg-$H_2$**.

CARRIERS

Hydrogen carriers are different physical or chemical forms of hydrogen (i.e. not compressed gas) which can be used to store and transport it in safer or more cost-effective ways. Potentially suitable hydrogen carriers include ammonia (NH_3), methanol (CH_3OH) and liquid hydrogen (LH2) due to their high volumetric energy density compared to compressed gaseous hydrogen.

Ammonia and methanol have been produced at commercial scale for over 100 years, however, the processes have not optimised for, nor typically even considered, low GHG emissions. While liquid hydrogen is also commercially available, it has not been scaled up to the size which would be required for a hydrogen economy, or even to match the design basis of the plants in this study (i.e. 100kta for blue hydrogen and 25 kta for green hydrogen). The levelized cost of producing these carriers with low to zero GHG emissions has been investigated and the outputs of the process modelling are summarised in Table 0.6, assuming 25 kta of green hydrogen input to each of the processes.

Table 0.6 Process modelling results for hydrogen carrier scenarios

Result	**Case**		
	Green Ammonia	**Green Methanol**	**Liquid hydrogen**
H_2 Consumed (kt/y)	25	25	25
CO_2 Consumed (kt/y)	0	185	0
N_2 Consumed (kt/y)	122	0	0
Electricity consumption (MW)	7.7[5]	31.2	43.8
CO_2 emitted (kt/y)	0	4	0
Carrier produced (kt/y)	136	108	25
Energy equivalent (PJ/y)[6]	3.06	2.48	3.54

Ammonia requires nitrogen as an input to the process, supplied from an air separation unit. Methanol requires CO_2 which was assumed to be supplied from a source with net zero emissions such as a biogenic source or direct air capture to ensure the resultant methanol product is net zero. Emissions from the methanol plant are associated with unconverted CO_2 leaving in the flue gas which is comprised of approximately 85 mol% hydrogen. High electricity consumption in the methanol case is due to the electrification of boilers which would

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[4] Beyond the scope of the original study and completed in 2022.
[5] Includes ASU and ammonia refrigeration package
[6] Based on HHV for each carrier

traditionally be heated using natural gas. This process change avoids the need for additional CO_2 capture from the boiler flue gases. Electricity consumption in the LH2 case is associated with the large number of compressors required for the refrigeration circuits. As with all electricity considered in this study, grid connected electricity is used to ensure high plant utilisation, with the expectation that green electricity would be purchased.

The levelized cost of the carriers has been summarised in Table 0.7. The cost of hydrogen feedstock into the process is not considered, so the levelized cost presented represents the marginal cost of production. The most comparable metric across the carriers is the levelized cost per gigajoule of product (rather than per tonne of product, which is a poor representation of the hydrogen carrying capacity due to its low mass density). This is the lowest for ammonia and highest for methanol. None of the LCOP's presented consider transport or conversion back to gaseous hydrogen at point of use.

Table 0.7 Summary of additional levelized cost of hydrogen carriers (excluding hydrogen feed cost which is assumed to be equal across each case)

	Green ammonia	**Green methanol**	**Liquified hydrogen**
Levelized cost ($/t product)	285	476	2,272
Levelized cost ($/GJ[7] product)	12.7	20.7	16.0
Levelized cost ($/t-$H_2$ feed)	1,550	2,056	2,272

This report highlights green ammonia as the most cost-effective carrier on both an energy and mass basis based on current technology and flowsheet configurations. However, liquid hydrogen is the next most cost-effective carrier on an energy basis, with methanol more cost effective on a per tonne product and per tonne H_2 feed basis.

Ammonia and methanol processes are already highly optimised for hydrogen from reforming and gasification, therefore radically new pathways for integrating green hydrogen that result in dramatically reduced cost are not expected. By contrast, liquid hydrogen production has only been operated for small scale, niche applications. Therefore, cost reductions (larger than those demonstrated here) associated with vastly increased plant size are still likely. Lastly, sourcing high-quality, low cost, non-fossil CO_2 feedstocks is also expected to be a significant challenge for green methanol, particularly at scale.

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[7] Based on HHV for each carrier

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KEY TERMS AND DEFINITIONS

Term	Definition
AE	Alkaline electrolysis. The oldest and most established form of electrolysis, involving an anode and cathode immersed in an alkaline bath to facilitate the transfer of hydrogen cations (H^+) from the anode to the cathode.
AEM	Anion exchange membrane electrolysis. An electrolysis technology which is facilitated by the transfer of hydrogen based anions (HX^-) across a membrane.
AGR	Acid gas removal. The process of removing CO_2 and H_2S from process streams. Removed CO_2 can be sequestered or vented to atmosphere (depending on emissions reductions objectives).
Anode	The electrode in an electrolyser where ionic hydrogen (chemical form is specific to the electrolyser technology) is converted into diatomic gaseous hydrogen (H_2).
Aspen Plus V11	A software used for process modelling, inclusive of thermodynamic and numerical convergence packages.
ATR	Autothermal reforming. A method of producing hydrogen from methane using heat and oxygen.
CAPEX	Capital expense.
Cathode	The electrode in an electrolyser where a current is applied to water to produce O_2 and ionic hydrogen (chemical form is specific to the electrolyser technology).
CCS	Carbon capture and storage. The process of capturing and sequestering CO_2 from a process stream (often to prevent release into the atmosphere). CCS is an all-encompassing term for a variety of technologies and process combinations.
CSG	Coal seam gas - a gas predominantly comprised of methane, found in seams within geological coal deposits.
DAC	Direct air capture. The process by which CO_2 is captured from the atmosphere. CO_2 captured can be sequestered or utilised. This technology is in early development.
DFB	Dual fluidised bed reactor – a reactor which facilitates biomass and /or coal gasification. The reactor has two chambers, one where the gasification occurs and the other where combustion occurs to pre-heat gasification media.
EFR	Entrained flow gasifier – a reactor in which fluidised pulverised coal flows alongside an oxidising agent (oxygen or steam) to facilitate chemical decomposition into syngas.

Electrolysis	An electro-chemical way of producing hydrogen. An electrical current is applied to decompose water into H_2 and O_2. A variety of electrolysis technologies exist, with slightly varying electro-chemical pathways, however the overall reaction is as follows: $H_2O \rightarrow H_2 + \frac{1}{2} O_2$
Gasification	A general term used for the decomposition of a solid form of carbon (i.e. coal or biomass) using an oxidising agent (often oxygen or water) into syngas.
ISBL	Inside battery limit. The equipment, operational responsibility and costs which are within the scope of the plant owner.
LCOH	Levelized cost of hydrogen. Typically expressed in \$/kg or \$/tonne or \$/GJ depending on context.
MMP	Molten metal pyrolysis – thermal decomposition of methane into syngas within a molten metal media (with the metal catalysing the decomposition).
NG	Natural gas – a gas predominantly comprised of methane.
OLGA	An oil based gas washing process (Dutch acronym), which removes tar entrained in gas streams.
OPEX	Operational expense. The cost of running a process, inclusive of utility costs (electricity, gas), feed costs (gas, coal), maintenance and administrative costs.
OSBL	Outside battery limit. The equipment, operational responsibility and costs which are not within the scope of the plant owner but lie with upstream or downstream entities.
PEM	Proton exchange membrane. Commercially developed form of water electrolysis for hydrogen production.
Pyrolysis	A general term for the thermal decomposition of carbon based material into syngas. The process occurs without an oxidising agent but is often accompanied by a catalysis to aide decomposition.
QBC	Queensland Black Coal. A high purity and more expensive coal.
Selexol process	An acid gas removal process (CO_2 + H_2S) which involves acid gas capture in an adsorption column using Selexol (a polymer solvent).
SMR	Steam methane reforming. A method of producing hydrogen from methane using steam and heat.
SOEC	Solid oxide electrolytic cell. Novel form of water electrolysis for hydrogen production.

Syngas	A gaseous mixture of H_2, CO and CO_2. Often an intermediate product or feedstock to produce H_2.
TIC	Total investment cost. The total cost request to complete the investment into a processing plant (capital costs, land costs, taxes, etc).
TRL	Technology readiness level – a nine-point scale used to evaluate the readiness of a technology. The scale goes from TRL1 (basic research) to TRL9 (system proven and ready for full commercial development).
VBC	Victorian Brown Coal. A low purity and less expensive coal.
WGS	Water gas shift reaction. A step commonly used to purify syngas (increasing H_2 yields) whereby the following reaction occurs: $CO + H_2O \rightarrow H_2 + CO_2$

1 INTRODUCTION

Hydrogen is a flammable, odorless, colourless gas which has been produced industrially for well over 100 years. Most commonly produced from fossil fuels, it is used in oil refining, fertiliser production, chemical production, and a variety of metallurgy applications. Decomposing into water upon combustion, it has gained interest in recent years as a potential clean energy carrier for the global clean energy transition, with potential applications in grid firming to long term energy storage to industrial heating to transport fuel. This report models zero to low greenhouse gas (GHG) emission production routes for hydrogen from various feedstocks to compare baseline metrics around GHG emissions, process efficiency and cost. It the rest of the report these zero to low GHG production pathways are referred to by the common colours of hydrogen – green (from renewable electricity or feedstocks) and blue (from fossil feedstocks but employing carbon capture and storage (CCS) to reduce emissions by >80%). The project did not model any grey (from natural gas) or brown (from brown coal) production routes (ie fossil feedstocks with no CCS). All blue hydrogen production routes attempted to maximise the CO_2 captured and thereby minimise CO_2 emissions. The report also conducts sensitivity analyses on the various pathways to highlight the most cost-effective areas of fundamental and applied research.

This report serves to provide a comprehensive summary of the modelling outcomes from the Future Fuels CRC project – Technoeconomic modelling of fuel production processes - which was completed in 2021. Methodological details are presented where they aide the readers understanding or assist with interpretations of results. However, all the detailed methodologies, design justifications and sensitivity analyses can be found in the detailed reports available through the Future Fuels CRC (project RP1.2-02).

1.1 PATHWAYS

There are many pathways to produce hydrogen with low carbon emissions, however, not all are at a level of technological maturity which warranted investigation. Technologies with a technology readiness level (TRL) of greater than 4 were considered. In selecting pathways, a prerequisite was that the technology must *genuinely* have low CO_2 emissions through technological implementations. Solutions which require carbon offsets were not included. Other factors which informed selection included resource availability and legislative constraints. For example, reforming of oil feedstocks with CCS (resource constraints in the Australian context), pink hydrogen a nuclear energy source (illegal in Australia), yellow hydrogen from solar photocatalysis (low TRL), white hydrogen extracted from naturally occurring reserves (resource constraints in Australia and low TRL) were all excluded.

Selected pathways were characterised by their primary production feedstock. Green pathways included **water (electrolysis)** and **biomass (pyrolysis & gasification)** as feedstocks while blue pathways utilised **natural gas (pyrolysis & reforming)** and **coal (pyrolysis & gasification)**. The report also considers hydrogen carriers – liquid fuels like **ammonia, methanol,** and **liquid hydrogen**.

1.2 DESIGN BASIS AND COSTING ASSUMPTIONS

To aid comparison between production pathways, a consistent design basis was set: green hydrogen production plants were sized to 25 kt-H_2/y (~100 MW_{th})[8] while blue hydrogen production plants were sized to 100 kt-H_2/y (~400 MW_{th})[9]. The quality of hydrogen produced by each process was defined using ISO 14687-2; 2012, with a H_2 purity of 99.97% and leaving the production facility at a pipeline pressure of 80 barg.

For utility services, several were assumed to be produced outside the plant battery limit (OSBL) and costs were included on a per unit basis. This included air (for cooling and nitrogen), cooling water, mains water and green electricity. In all core cases electricity was considered as grid connected electricity (to ensure high plant utilisation), but made 'green' using financial instruments (i.e. purchasing green electricity) to minimise scope 2 emissions. Utilities produced inside plant battery limits (ISBL) included steam and demineralised water, and capital costs for the necessary plant infrastructure and the associated operating costs were included. The quality of feedstocks for various pathways is detailed in Appendix A.

The capital cost of the individual process units was assessed using the cost of equipment, package, or unit, with literature data used to calculate the cost of reactors, separation units, pressure swing adsorption for hydrogen purification, air separation units and compressors. Literature data was adjusted for scale using a cost curve method and a scaling exponent appropriate for the unit type and adjusted to 2020 dollars using the Chemical Engineering Plant Cost Index. The size and cost of process heat exchangers was estimated by Aspen Exchanger Design and Rating (EDR), and the Aspen Plus database was used for costing generic equipment (e.g. pumps and tanks) in the process flowsheet. Spare equipment, including pumps and compressors, was considered as part of the CAPEX estimation for rotating equipment. Total inside battery limit costs (ISBL) were between 270-320% of the bare equipment cost depending on process complexity. Utilities and auxiliary facility costs (i.e. OSBL) were 40% of ISBL. Total direct costs (TDC) were classed as ISBL+OSBL. Total indirect costs (TIC) included:

- Engineering, procurement, construction and management costs were 12% of TDC and owner's costs were 8% of ISBL.
- Project and process contingency were adjusted based on the level of technological maturity with first of a kind plants (emerging technologies) incurring 50% TDC and project contingency 30% TDC to reflect the uncertainty associated with scaling lower TRL options. Mature technologies or so-called Nth of a kind plants attracted 10% of TDC for both project and process contingency respectively.

The overall fixed capital cost (FCC) were the sum of TDC and TIC. Working capital was assumed to be 8% of FCC and start-up capital 10% of FCC. Total capital cost (TCC) was

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[8] ~95MW based on the lower heating value of hydrogen (120MJ/kg) or ~112MW using the higher heating value of hydrogen (141.8 MJ/kg)

[9] ~380 MW_{th} using LHV and ~450MW_{th} using HHV

the sum of FCC + working capital + start-up capital. This equipment factored cost estimation approach was analogous to a class 5 AACE[10] estimate.

Operational costs included feedstock, electricity, water and carbon capture and storage inputs. Baseline values for unit costs for each are defined in Table 1.1. Maintenance, taxes and insurance were incorporated as a fixed percentage (3, 2 and 0.5 respectively) of the total capital cost (TCC).

Table 1.1 Utility and Feedstock Costs

Item	Cost	Units
Natural Gas (or CSG)	6	\$/GJ
Brown Coal	1.5	\$/GJ
Black Coal	3.5	\$/GJ
Biomass (bagasse)	0.1	\$/GJ
Electricity	85	\$/MWh
Water	0.1	\$/m^3
CO_2 transport	0.2	\$/t
CO_2 Storage	9	\$/t

Finally to report a levelized cost of hydrogen (LCOH) a discounted cashflow methodology was applied using the economic parameters in Table 1.2.

Table 1.2 Discounted cashflow parameters

Item	Cost
Year of Analysis	2020
Plant Lifetime	30 years
Construction Period (% breakdown of capital per year)	2 years (35% / 65%)
Start-up Period (capacity during start-up) **(Capacity during 1st year / 2nd year of operation)**	6 months (50%) (80% / 100%)
Land purchased	No value
Inflation	2.5%
Weighted average cost of capital	8.5%
Depreciation	15 years / straight line
Tax rate	30%
Net salvage value at end of project life	0

1.3 STRUCTURE OF REPORT

This work presents the technoeconomic results of each production pathway individually with each section containing: an introduction to the given pathway, followed by a description of the

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[10] Association for the Advancement of Cost Engineering (AACE)

process, modelling results and sensitivity analysis. Insights into the pathway as an outcome of the modelling conclude each section.

2 NATURAL GAS

2.1 INTRODUCTION

Natural gas is the primary feedstock for hydrogen production today. In this report, two broad groupings are considered for hydrogen production: conventional production and emerging technologies. Pipeline natural gas (NG) was considered the main feedstock, although coal seam gas (CSG) was explored in several scenarios due to the higher methane purity). A total of 5 primary scenarios[11] were considered:

- 3 for conventional reforming
    - Steam Methane Reforming (SMR) Combined in series with Autothermal Reforming (**NG-C-SMR**) of natural gas,
        - Standalone Steam Methane Reforming (**NG-S-SMR**),
        - Autothermal reforming (ATR) with CO_2 capture of >90% (**NG-ATR**), and
        - Autothermal reforming where CO_2 emissions generated from the syngas only were captured, with other flue gas emissions (predominately CO_2) vented to atmosphere (**NG-ATR-OP**).
- 2 for unconventional pyrolysis:
    - 3 variations of solid catalyst methane pyrolysis (**SCP**) with a solid iron oxide catalyst.
        - **NG-SCP** uses natural gas, five stage solid fluidised bed reactors, and a >90% CO_2 capture efficiency on flue gas streams where methane is combusted to produce heat.
        - **CSG-SCP** used the same process as NG-SCP but coal seam gas as a feedstock.
        - **CSG-SCP+RE** uses coal seam gas and renewable electricity for provision of process heat. This eliminates the need for heat provision via the combustion natural gas and the associated CCS units to minimise emissions. Nitrogen accumulation within the process must be managed via a purge stream (i.e. not all unreacted methane can be recycled back into the process).
    - 4 variations of methane pyrolysis using a molten catalyst technology, wherein a molten metal is used as the primary catalyst, with a molten salt used for carbon recovery and heat integration in the system.

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[11] The naming convention for each scenario is feedstock-process.

- **NG-MMP** uses natural gas >90% CO_2 CCS efficiency on flue gas streams where methane is combusted to produce heat for the pyrolysis reaction.
- **NG-MMP-BL** is functionally the same as **NG-MMP** except hydrogen purification to ISO 14687-2; 2012 specifications was not included. Rather, a blended hydrogen / methane mixture was the final product for direct injection into a blended natural gas pipeline. This reduces product value but also the overall plant footprint and cost of purification.
- **CSG-MMP** is functionally the same as **NG-MMP** but uses a feedstock of coal seam gas instead of natural gas.
- **CSG-MMP+RE** uses the same configuration as **CSG-MMP**, with the addition of renewable electricity for provision of process heat. This eliminates the need for heat provision via the combustion natural gas and the associated CCS units to minimise emissions. Nitrogen accumulation within the process must be managed via a purge stream (i.e. not all unreacted methane can be recycled back into the process).

To understand the basic process selection undertaken in this report, it is relevant to review the academic literature[12]. Primary considerations were preliminary technoeconomic potential, technology readiness level (TRL), and commercial deployment.

Reforming of natural gas is currently the dominant production technology for hydrogen. Coupled with Carbon Capture and Storage (CCS), reforming of natural gas, in principle creates a source of low GHG emissions hydrogen, often termed blue hydrogen. Steam Methane Reforming (SMR) is a mature technology (TRL 9) where steam is reacted with methane to form syngas (a mixture of H_2 and CO) which is then purified to produce H_2. SMR is deployed in the world's largest single train hydrogen plant. Autothermal Reforming (ATR) is another mature and deployed technology (TRL 9) where methane is partially combusted in the main reaction vessel to provide the heat required for the reforming reaction, as opposed to SMR where heat is provided external to the reaction vessel, creating two different streams containing CO_2. Simplified process flowsheets for all 4 cases are provided in Figure 2-1 through Figure 2-4.

The pyrolysis of natural gas via plasma has been deployed in the carbon black industry for roughly 100 years. In addition, the catalytic decomposition of methane to produce hydrogen and solid carbon has been investigated for a similar length of time in the academic literature. No oxygen is present when the methane is decomposed so CO_2 is not formed, eliminating the need for CCS on the pyrolysis products. In this study methane pyrolysis with solid catalysts was considered TRL 7/8 and the molten media pyrolysis technologies TRL 4/5.

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[12] A complete literature review was undertaken as part of the Milestone 2 Report for RP1.2-0.2. A copy of this literature review may be obtained by contacting the Future Fuels CRC.

2.2 PROCESS DESCRIPTION

2.2.1 Pre-treatment

All processes required pre-treatment for sulphur removal from the methane feedstock (natural gas or coal seam gas). NG and CSG enter at typical pipeline conditions (80 barg and 25 ^{o}C), before the pressure is reduced 36 bar and pre-heated to 380 ^{o}C for sulphur removal. All sulphur is hydrogenated to H_2S, which is then adsorbed using activated carbon for removal.

2.2.2 Reaction Pathways

Following sulphur removal, high-pressure steam is injected with the desulfurized natural gas into the pre-reformer at a steam to carbon ratio (**NG-C-SMR**: 2.8, **NG-S-SMR**: 2.2, **NG-ATR/OP**: 0.86). This cracks any C2+ hydrocarbons (e.g. ethane) in the pre-treated feedstock and helps to prevent coke formation in the main reformer. The output of the pre-reformer is heated to about 600°C and sent to the main reformer to generate syngas (a mixture of H_2/CO and small amounts of CO_2). The SMR typically uses an externally heated series of reformer tubes to provide the heat (i.e. through natural gas fired burners) for the reforming reaction with steam. The higher steam to carbon ratio produces a more H_2-rich syngas. ATR injects additional oxygen with a lesser amount of steam to combust some of the natural gas inside the reformer (thereby providing heat for the ATR reaction), which is conventionally a larger refractory lined vessel. The lower steam/carbon ratios produce a syngas with lower quantities of H_2. The syngas (regardless of reformer technology) is then sent to a Water Gas Shift (WGS) reactor to maximise the hydrogen content by reacting CO with water to produce H_2 and CO_2. Conditions in the WGS reactor vary depending on the process and detailed heat and mass balance information is provided in the appendices and Process Flow Diagrams. Steam is generated from a combination of heat integration and boilers fed by a slipstream of natural gas feed and H_2 rich waste gases (e.g. off-gas from CO_2 removal or tail-gas from the H_2 purification units). In 3 of the 4 cases excess steam is generated and used to make electricity which itself is either used in the process, reducing reliance on the grid or, if in excess, fed back into the grid. In the single reformer process NG-S-SMR, opportunities for heat integration are reduced and all the steam generated is used either for the reforming or as the heat source for the CO_2 removal units (to regenerate the solvent). Amine CO_2 removal was selected for all cases using a mixture of methyl diethanolamine (MDEA) and piperazine (PZ). Captured CO_2 is pressurised to 15.3 MPa through multi-stage compression. Intermediate dehydration using triethylene glycol is conducted at 2.1 MPa.

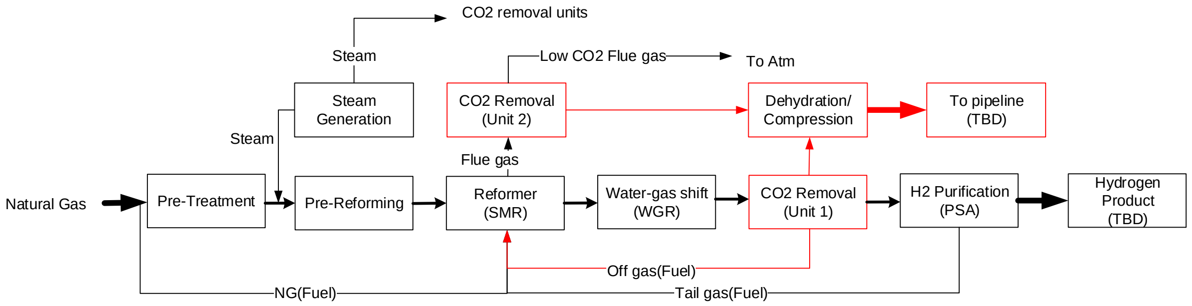

Figure 2-1 Simplified process diagram for NG-S-SMR

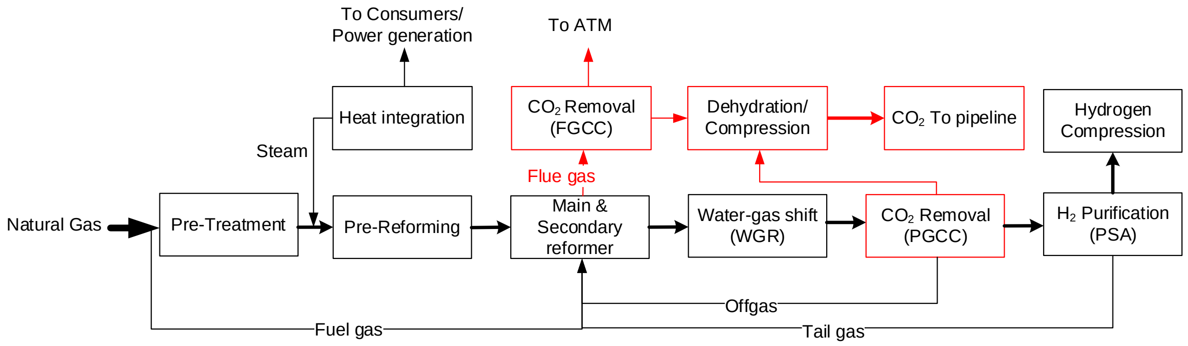

Figure 2-2 Simplified process diagram for NG-C-SMR

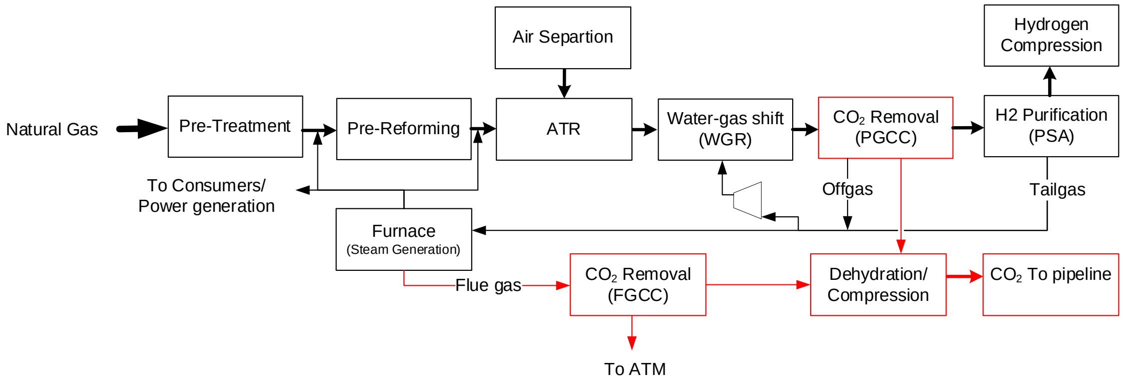

Figure 2-3 Simplified process diagram for NG-ATR

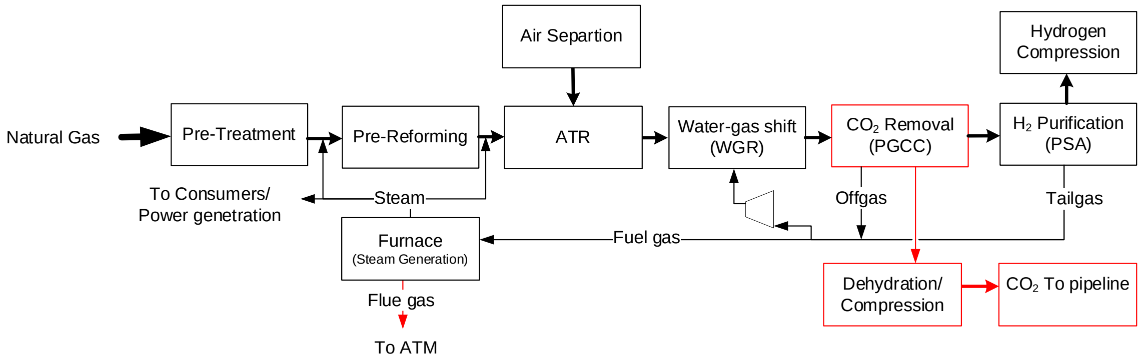

Figure 2-4 Simplified process diagram for NG-ATR-OP

The methane pyrolysis reaction pathways developed for the solid catalyst (SCP) as shown in Figure 2-5 and Figure 2-6, are based on the publicly available data of the Hazer process [1-4]. Catalyst dusting, wherein the solid catalyst particles are passed (in a counter current fashion) through fluidised bed reactors (FBR) of ascending pressures, allows for the catalyst particles to fracture and re-expose fresh catalyst surface. Heat is provided for the methane pyrolysis reaction by combusting the tail-gas from the H_2 purification stage. The flue gas from the FBR units contains some CO_2 and this is captured (>90% CO_2 capture efficiency). The key difference between **NG-SCP** and **CSG-SCP** is the use of coal seam gas feedstock (which has higher methane and lower impurities); whilst the **CSG-SCP+RE** integrates renewable electricity for the provision of process heat, which negates the need for CCS.

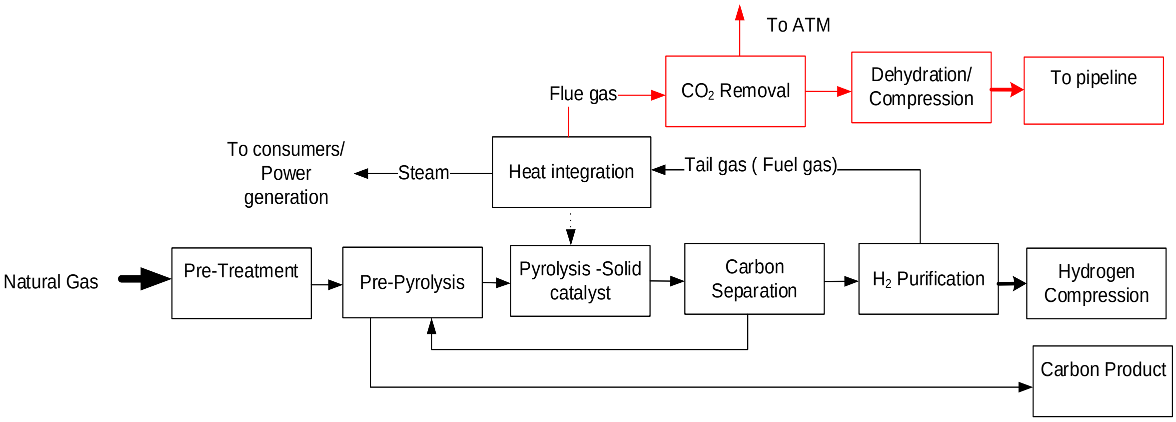

Figure 2-5 Simplified process diagram for NG-SCP / CSG-SCP

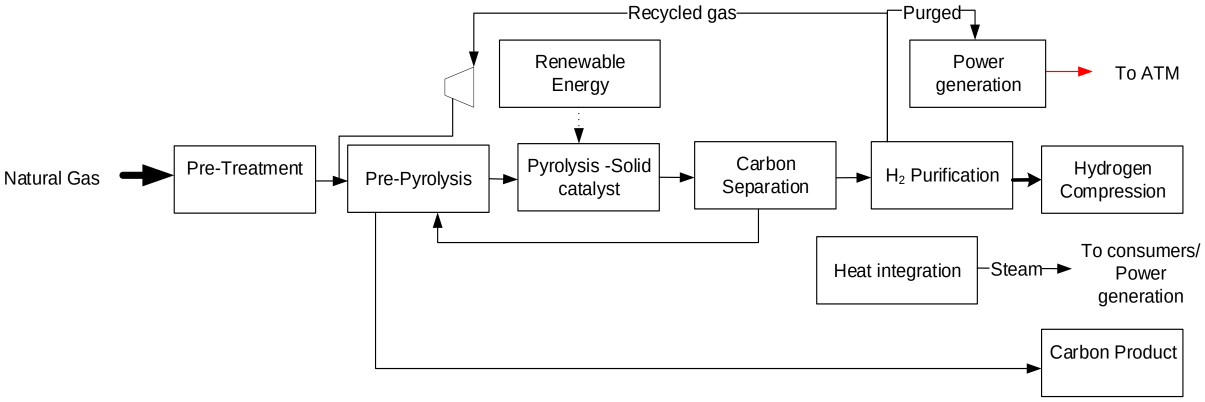

Figure 2-6 Simplified process diagram for CSG-SCP+RE

For all methane pyrolysis cases using molten media (MMP) cases, a molten catalyst system is used in bubble column reactor. The reactor system was modelled assuming a dense molten metal catalyst, and a low-density molten salt for heat integration and removal of carbon (assumed no catalytic affect). Due to differences in density, it is assumed that a carbon rich molten salt layer is looped in and out of the reactor, with the primary function of removing carbon for further processing and heat transfer (i.e. provision of heat for the methane pyrolysis reaction). Again, the feedstock is varied, either NG or CSG, and as with the SCP process, heat is provided for the methane pyrolysis reaction by combusting the tail-gas from the H_2 purification stage. The flue gas this combustion process contains some CO_2 and this is captured (>90% CO_2 capture efficiency). The exception is the CSG-MMP+RE which integrates renewable electricity for the provision of process heat, which negates the need for CCS.

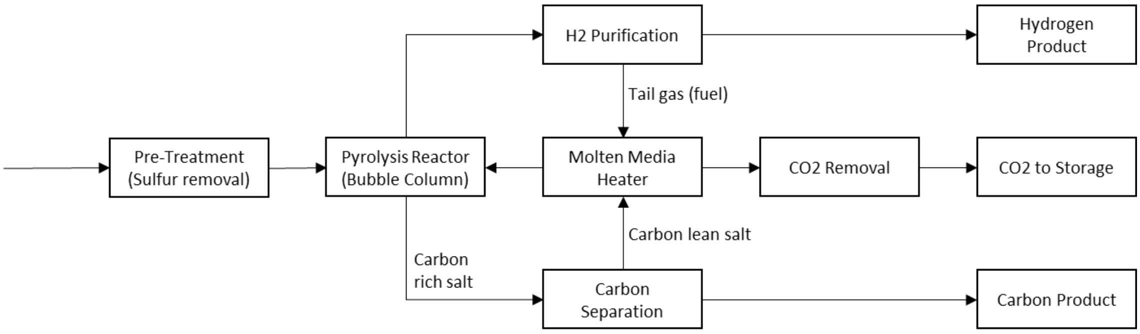

Figure 2-7 Simplified process diagram for NG-MMP / NG-MMP-BL / CSG-MMP

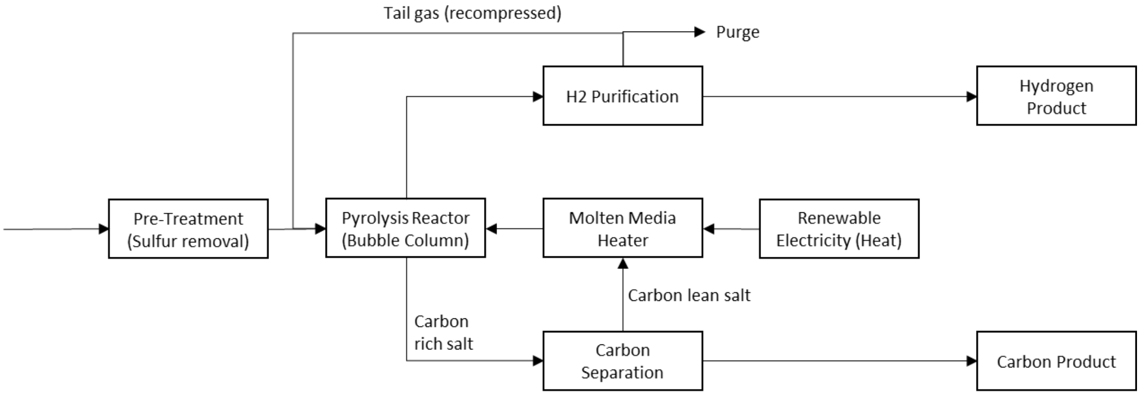

Figure 2-8 Simplified process diagram for CSG-MMP+RE

2.2.3 Hydrogen purification and compression

The hydrogen product stream in all cases undergoes purification to 99.97% H_2 through a pressure swing adsorption (PSA) unit. Following this, the hydrogen is compressed to 81 bar for pipeline injection. The exception is the NG-MMP-BL scenario where the reactor gas (a mixture of H_2/CH_4) is not purified, but rather directly compressed for injection (i.e. blending) into a natural gas pipeline.

2.3 PROCESS MODELLING

The mass and energy balances for all 5 scenarios were modelled using Aspen Plus V11® (hereafter referred to as Aspen) and the results are shown in Table 2.1 and Table 2.2. Units were modelling using built in blocks in Aspen. Where a unit is supplied as a package from a vendor, the unit is empirically modelled in Aspen.

Table 2.1 Process modelling results for conventional blue hydrogen derived gas scenarios

Result		**NG-S-SMR**	**NG-C-SMR**	**NG-ATR**	**NG-ATR-OP**
H_2 produced (kt/y)		100	100	100	100
Gas input (kt/y)		390	379	359	347
Electricity consumption (MW)	**ASU**	NA	4.45	17.87	17.26
	Auxiliary	26.7	24.61	21.44	20.01
Net energy efficiency (%)	**HHV**	71.60	73.70	77.80	80.40
	LHV	67.00	69.02	72.90	75.30
CO_2 (kt/y)	**Produced**	998	997	918	888

	Captured	905	905	843	748
	Emitted	93	92	75	140
CO_2e (t-CO_2/t-H_2)		0.93	0.92	0.75	1.4

Table 2.2 Process modelling results for unconventional blue hydrogen derived gas scenarios

Result		**NG-SCP**	**CSG-SCP**	**CSG-SCP+RE**	**NG-MMP**	**NG-MMP-BL**	**CSG-MMP**	**CSG-MMP+RE**
H_2 produced (kt/y)		100	100	100	100	100	100	100
Gas input (kt/y)		576	533	438	595	267	550	458
Electricity consumption (MW)		39.53	39.09	34.25	19.9	12.13	18.94	12.31
Net energy efficiency (%)	**HHV**	48.50	49.10	52.60	47.00	74.60	47.60	49.00
	LHV	45.50	46.00	48.80	44.00	71.80	44.70	53.50
CO_2 (kt/y)	**Produced**	240	238	60	323	58	317	113
	Captured	217	215	NA	292	53	287	NA
	Emitted	23	23	60	31	5	30	113
CO_2e (t-CO_2/t-H_2)		0.23	0.23	0.60	0.31	0.50	0.30	1.13

2.4 FINANCIAL ANALYSIS

2.4.1 Capital Cost

For the conventional blue hydrogen pathways from natural gas, the breakdown of the total capital costs is presented in Figure 2-9. The capital cost for the NG-S-SMR scenario is the lowest, closely followed by the NG-ATR-OP. The main factors that influence the TCC across all four scenarios are:

- **the selected reformer technology** with the ATR having a higher equipment cost than the SMR scenarios).
- **the location and number of CO_2 capture units** with the CO_2 removal from the flue gases the most significant capital equipment cost for the SMR pathways scenarios due to the indirect heating approach (approximately half the ISBL cost for the NG-S-SMR scenario is attributed to the CO_2 capture, dehydration and compression units.
- **the need for oxygen from an air separation unit** (ASU) with both ATR scenarios having 20-25% of the TCC in the ASU and O_2 compressors.

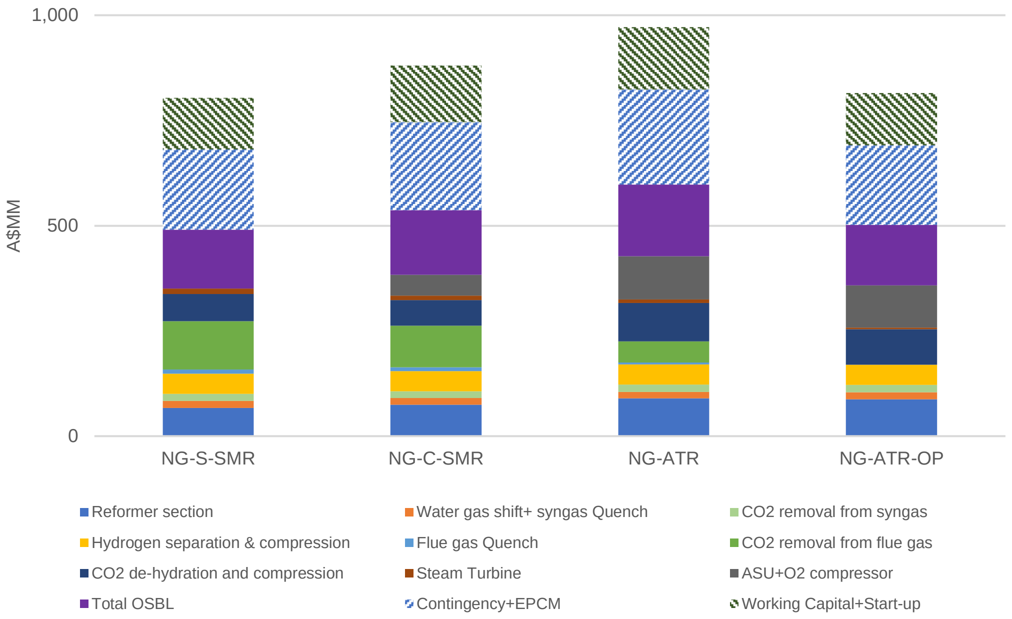

Figure 2-9 Capital cost breakdown for blue hydrogen from natural gas using conventional technologies

For the unconventional technologies, the only notable difference is that some equipment packages are not available in commercial software packages as the designs had not been constructed on a commercial scale. In these cases, designs and costs were adapted from the closest commercial analogue[13]. Renewable energy integration for both the CSG-SCP+RE and CSG-MMP+RE cases was through resistive heating coils in the reactor vessel. The cost of these coils was included in the capital cost of the reactor units themselves, although such configurations have never been demonstrated. The results are summarised in Figure 2-10 and Figure 2-11. The equipment costs for the pyrolysis scenarios is between 20% and 50% lower than the equipment costs of the conventional blue hydrogen from natural gas scenarios. This is largely a function of fewer units and a greatly reduced CO2 capture requirement. However, the low technology maturity and greater project / process contingencies actually means the TCC for the emerging technologies is up to ~40% higher than the SMR or ATR options.

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[13] Full details of this process are given in the Milestone 5a Report for RP1.2-02. A copy of this report may be obtained by contacting the Future Fuels CRC.

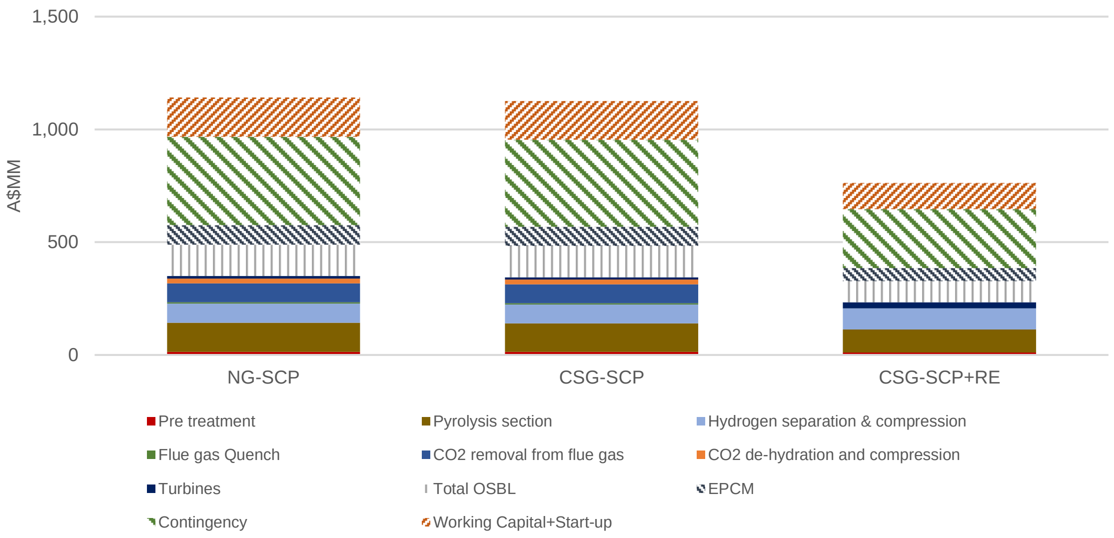

Figure 2-10 Capital cost breakdown for methane pyrolysis by solid catalyst

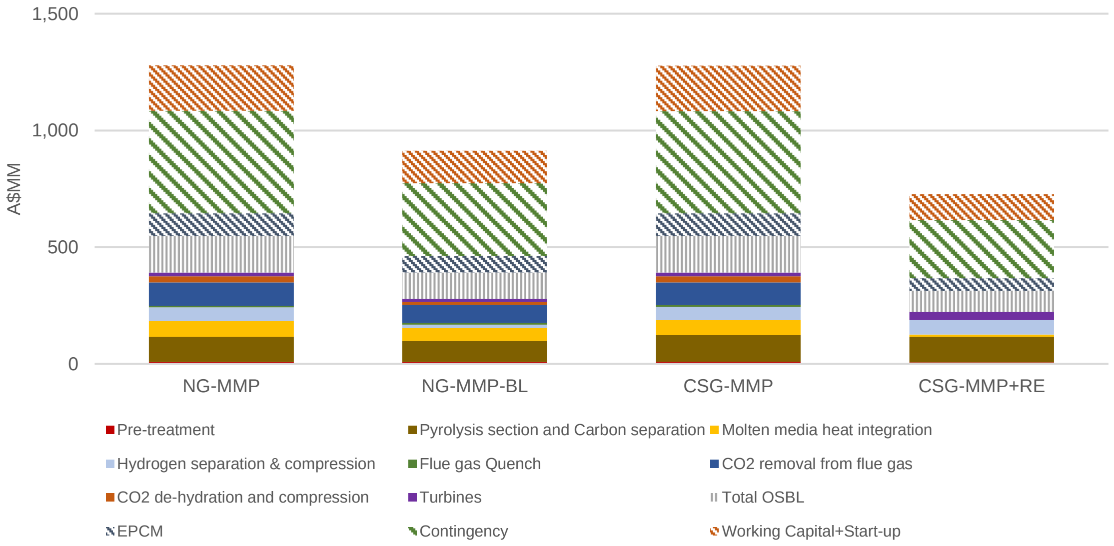

Figure 2-11 Capital cost breakdown for methane pyrolysis by molten catalyst

2.4.2 Operating Cost

Operating expenditure is summarised in Figure 2-12 and Figure 2-13. In all scenarios the operational expenditure is dominated by the feedstock cost which is approximately 65% of the total OpEx. The conventional technologies benefit from the water gas shift reaction which generates additional H_2 from reacting H_2O with CO in the syngas. As a result, they require less natural gas input compared to the various pyrolysis pathways (where H_2 is only produced from methane). This translates to feedstock costs that are on average ~45% lower for the conventional technologies compared to the methane pyrolysis pathways.

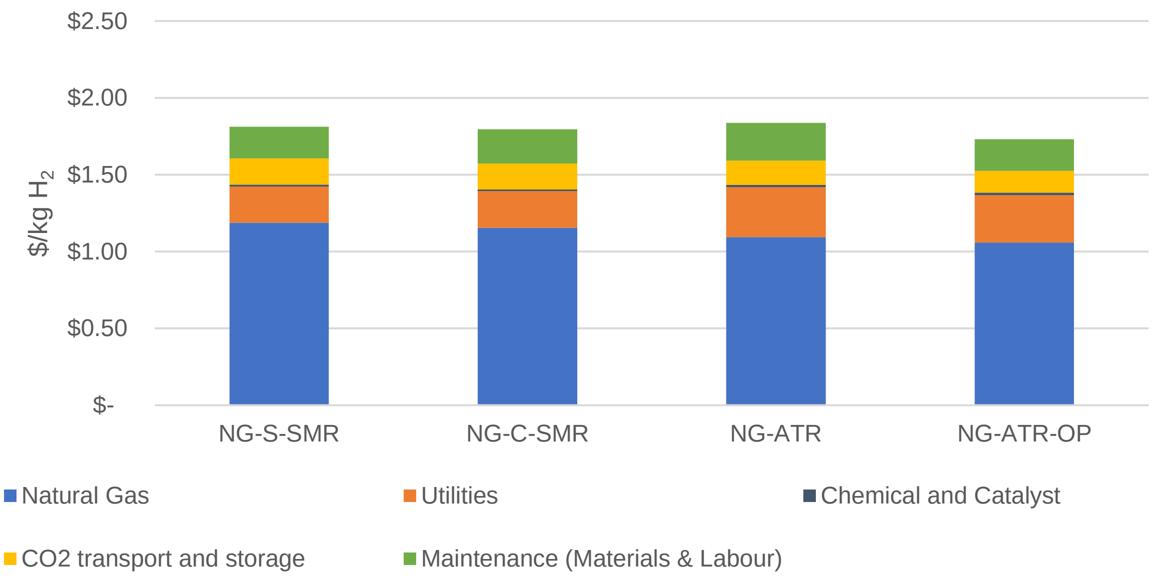

Figure 2-12 Operational expenditure for gas reforming pathways

Electricity is typically the second highest contributor to the operating cost for the methane pyrolysis scenarios. Integration of renewable electricity into either the SCP or MMP pathway for the provision of heat to the reactor, trades a reduction in feedstock cost for a significant increase in electricity costs. As a result, the CSG-MMP+RE and CSG-SCP+RE scenarios have the highest and second highest operational costs of any of the natural gas scenarios.

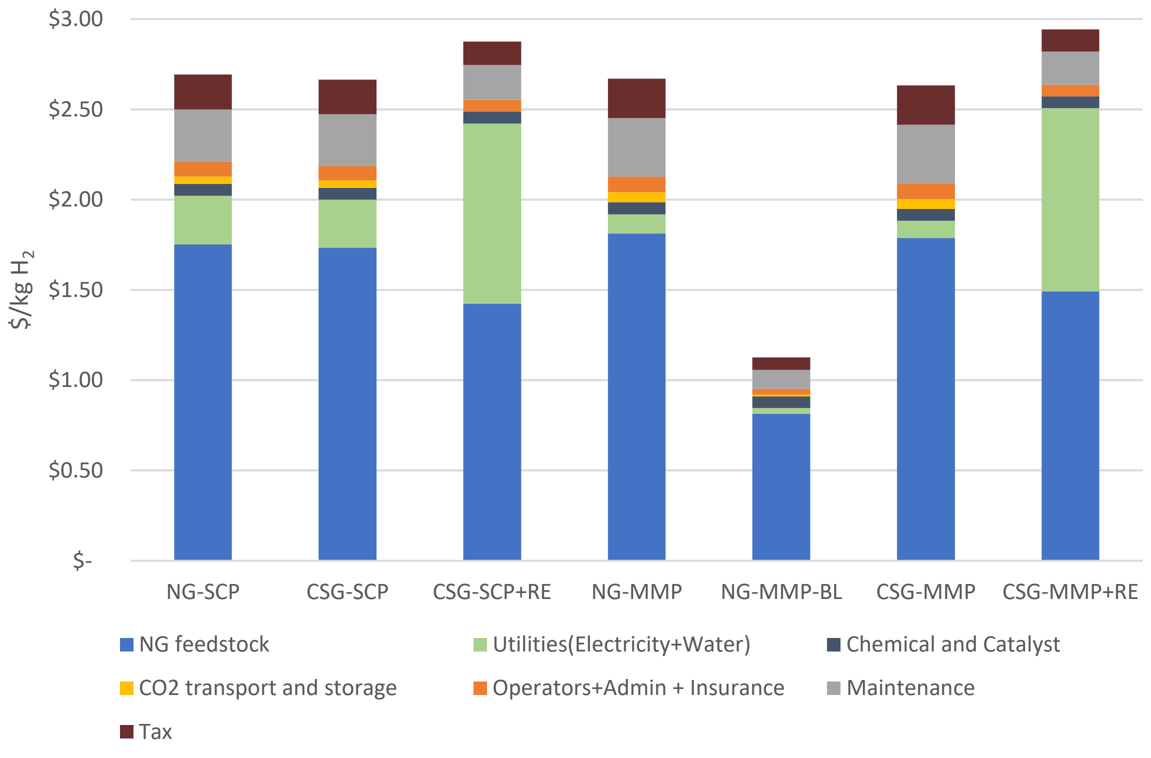

Figure 2-13 Operational expenditure for methane pyrolysis pathways

2.4.3 Levelized Cost

There was approximately an 11% difference in the levelized cost of hydrogen (LCOH) across the four conventional scenarios. The lowest cost being the NG-ATR-OP at \$2.92/kg H_2 whilst NG-ATR had the highest at \$3.25/kg H_2. The two SMR based technologies were \$2.98/kg H_2 and \$3.06 /kg H_2, for the NG-S-SMR and NG-C-SMR respectively.

As with grey hydrogen the natural gas cost remains the highest contributor to total production cost for the blue hydrogen variants, accounting for 34-40% of LCOH. CAPEX was the second major contributor, comprising 27-30% of the LCOH. Taken together these two factors comprise roughly 2/3 the total LCOH, highlighting how technology improvements, like electric furnaces which reduce both the plant footprint and complexity (CAPEX) and the natural gas demand, may still play a major role in reducing LCOH even for these established technology cases.

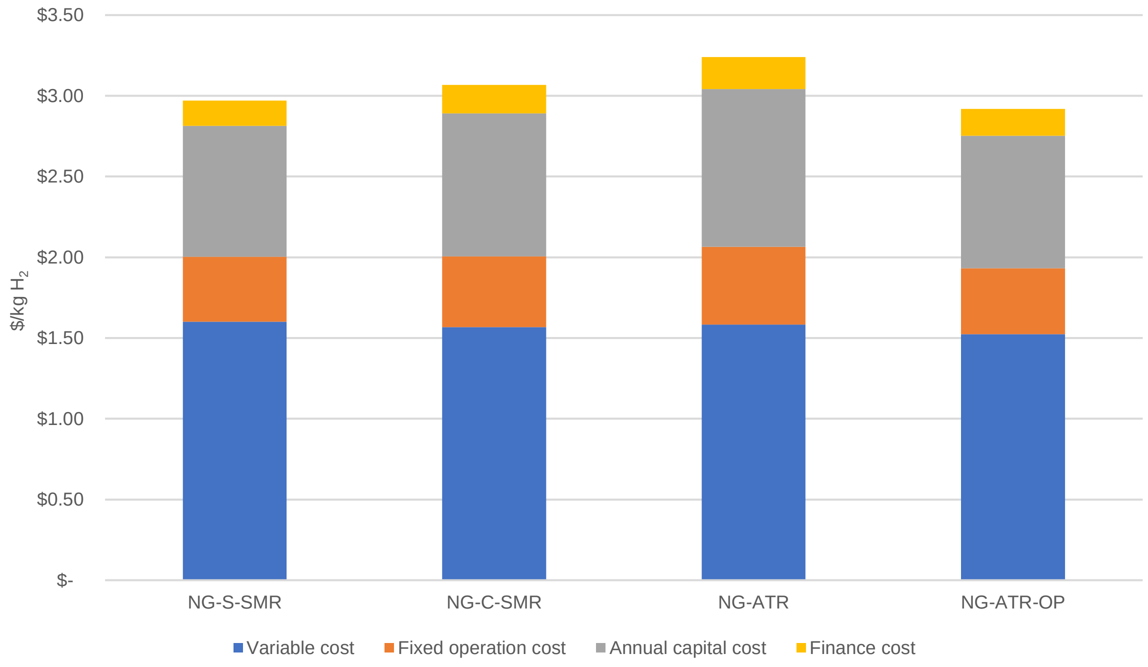

Figure 2-14 Levelized cost of hydrogen for blue hydrogen production from conventional technologies using natural gas

The levelized cost of hydrogen (LCOH) for emerging technologies was estimated to be between \$3.68 to \$4.11/kg H_2. Figure 2-15 shows the production cost breakdown for individual cases. Operational costs contribute the most to LCOH (37 - 44%), followed by the annual capital charge (20 - 31%). The change in process configuration to incorporate renewable electricity for process heat, rather than using natural gas with CCS, reduced the LCOH by 7% for the SCP and 9% for the MMP scenarios.

The low purity hydrogen case (NG-MMP-BL) can be considered as a standalone evaluation for hydrogen blending into natural gas pipelines. In this case the LCOH is \$1.5/kg product (79%mol H_2, which is equivalent to \$0.45/kg H_2).

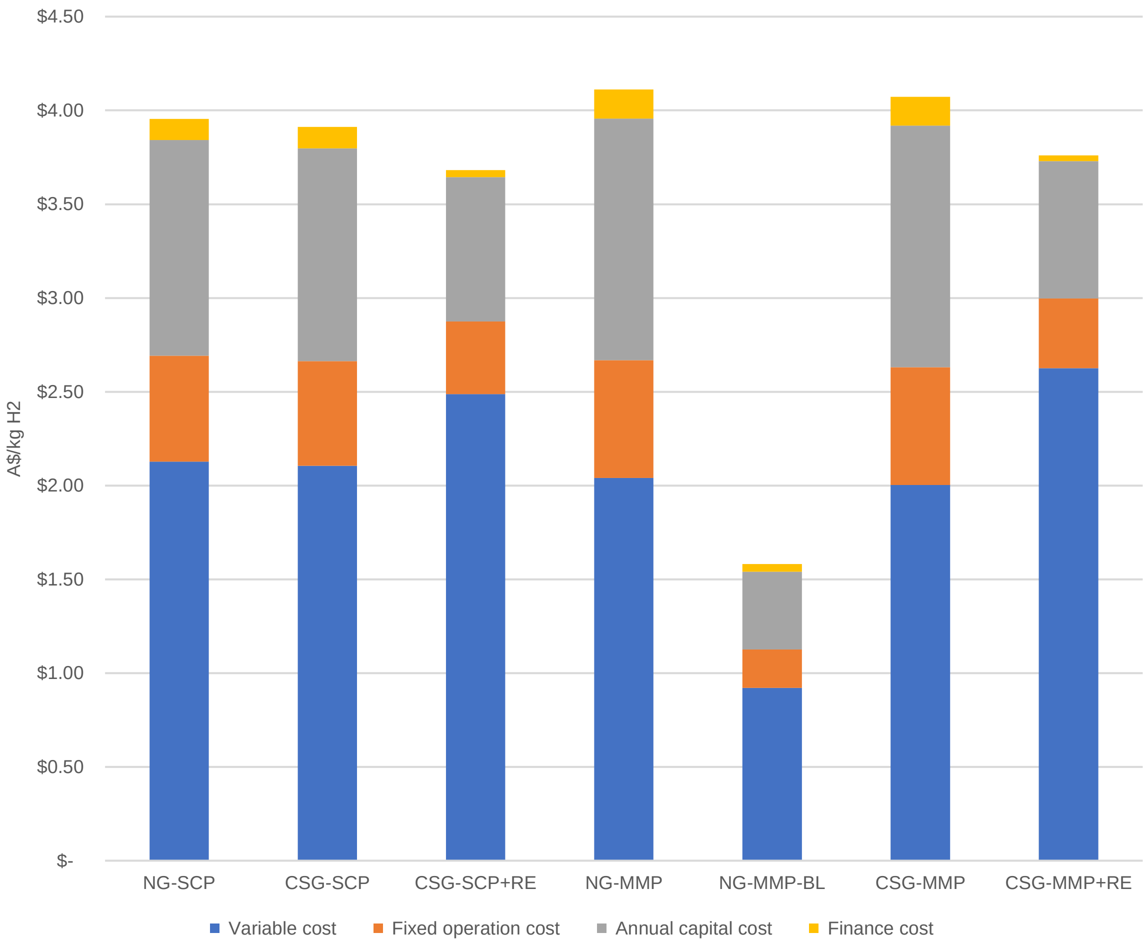

Figure 2-15 Levelized cost of hydrogen for blue hydrogen production from emerging technologies using natural gas

2.4.4 Sensitivity Analysis

A detailed sensitivity analysis[14] was conducted to better understand the impact of variations in the dominant cost factors (feedstock cost and capital cost) and to identify targets that may lead to hydrogen production costs below the $2/kg target originally expressed in the Low Emissions Technology Statement[15]. The results are summarised in Figure 2-16 and Figure 2-17. Crucially, LCOH values only approach the $2 /kg H_2 target when feedstock costs are $2.5/GJ and/or capital costs are reduced by 30%.

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[14] Full details of this process are given in the Final Milestone Report for RP1.2-02. A copy of this report may be obtained by contacting the Future Fuels CRC.

[15] LOW EMISSIONS TECHNOLOGY STATEMENT 2021 Australian Government Department of Industry, Science, Energy and Resources

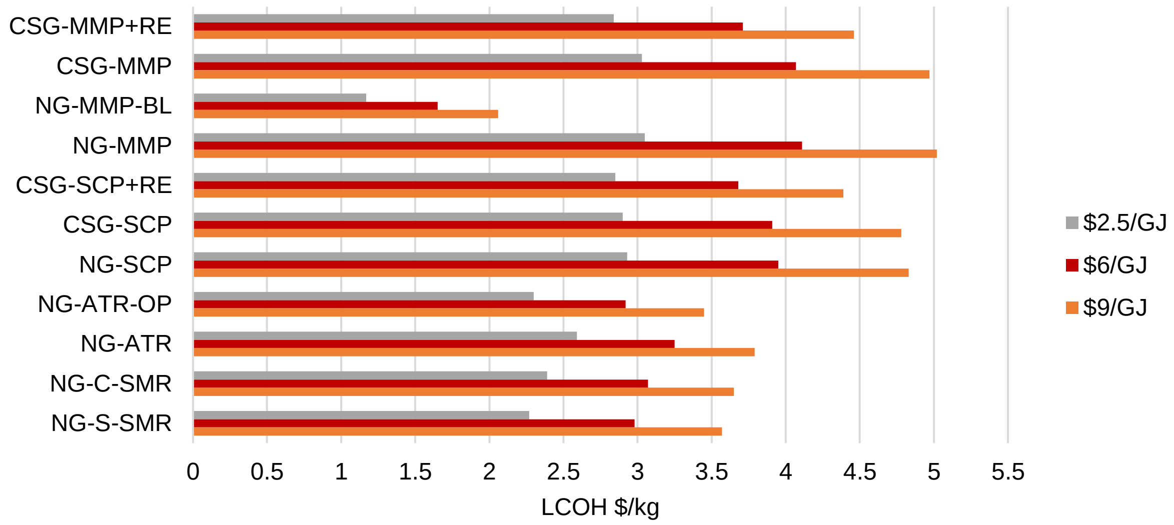

Figure 2-16 Feedstock price sensitivity analysis for blue hydrogen from methane

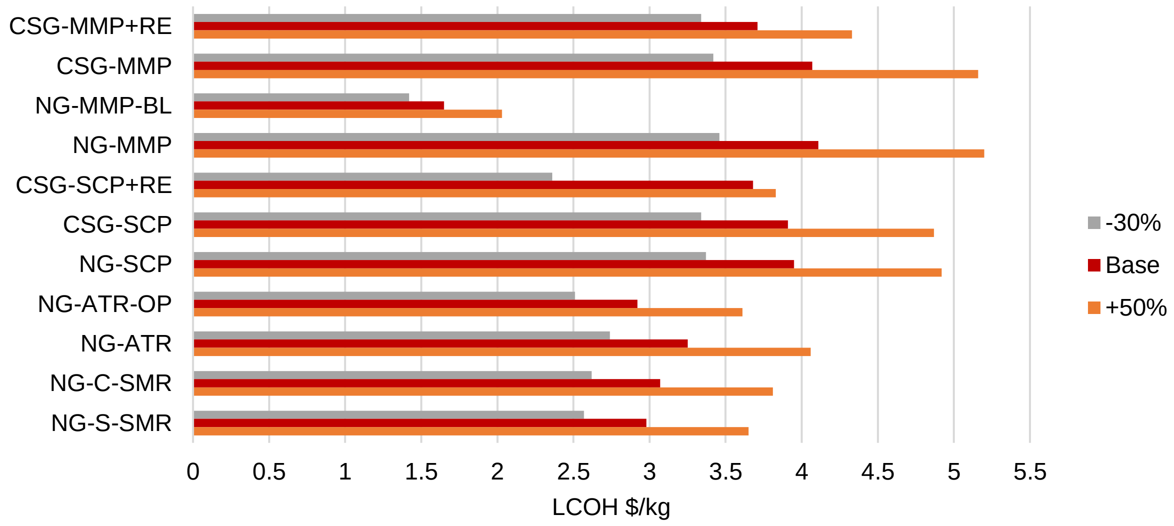

Figure 2-17 Capital cost sensitivity analysis for blue hydrogen from methane

The methane pyrolysis scenarios are also influenced by the potential sale price of the solid carbon by-product that is generated along with the H_2. Research on the quality of the carbon by-product is mixed at the lab-scale, presenting uncertainty in the resulting LCOH. A base carbon price of A$50/t was used initially, to assign some value to it, however; if products like high-quality graphite or carbon fibres can be produced, the value could be substantially higher. Therefore, we assessed the impact on LCOH between a price of $500/t and minimum of $0/t. The results showed that the carbon by-product sales price had a large impact on LCOH, even more than low natural gas prices; reducing the LCOH below $3 /kg H_2 for all cases and reaching as low as $2.36 /kg H_2 for CSG-SCP+RE.

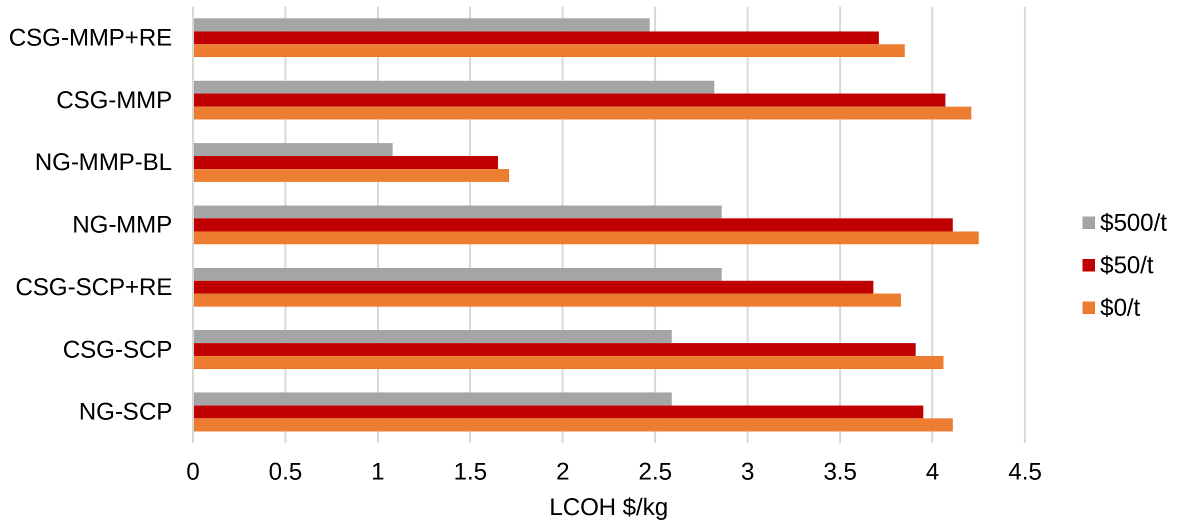

Figure 2-18 Carbon sale price for pyrolysis sensitivity analysis for blue hydrogen from methane

We note that in addition to the potential positive impact on process economics from producing a high value solid carbon product, there is also substantial uncertainty in the processes required for cleaning the carbon to obtain the desired quality. In each case further research is warranted to determine both how these qualities can be obtained and the associated costs. Currently, the process model only involves crude, gravity driven separation and minimal washing, which is likely inadequate for high quality, high purity carbons. Likewise, the costs of these downstream operations are not costed. Therefore, we recommend prioritising research that maximises the by-product carbon value and minimises the associated processing costs.

3 COAL

3.1 INTRODUCTION

Coal gasification is a process in which coal is oxidised at high temperatures to produce syngas. As with natural gas reforming, syngas produced through coal gasification is run through a water gas shift (WGS) reactor to increase H_2 yield. Traditional coal gasification is a mature technology which has predominantly been used for power generation in Integrated Gasification Combined Cycle (IGCC) plants or for chemical synthesis (e.g. methanol). However, there is substantial variation in plant performance based on coal type and gasifier configurations, and there are no commercial plants that currently incorporate carbon capture and storage (CCS) for the production of H_2. There is a significant potential to produce low emission blue hydrogen through coal gasification in Australia due to the co-existence of Australia's low-cost coal reserves and CO_2 storage reservoirs.

Four scenarios for H_2 production through coal gasification were investigated. Simplified process flowsheets for all 4 cases are provided in Figure 3-1 through Figure 3-4.

- The first (**QBC-EFR**) investigated the use of Queensland Black Coal (QBC) as a feedstock to an oxygen fed entrained flow gasifier.
- The second (**VBC-EFR**) used the same gasifier, however with Victorian Brown Coal (VBC) as a feedstock.

- The third **(VBC-DFB)** used VBC as a feedstock to a dual fluidized bed gasifier.
- The fourth **(VBC-DFB+CAP)** used VBC as a feedstock to a dual fluidized bed gasifier, and employed in-situ carbon capture as opposed to dedicated CO_2 capture units for the previous 3 processes.

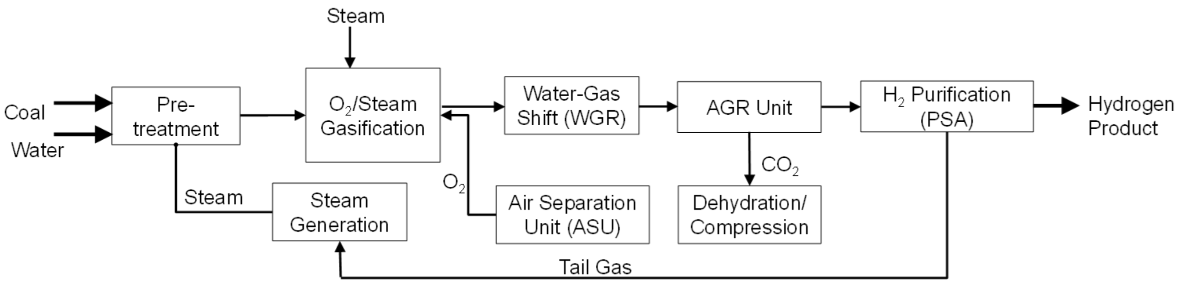

Figure 3-1 Simplified process diagram for oxygen/steam entrained flow gasification using Queensland black coal (QBC-EFR)

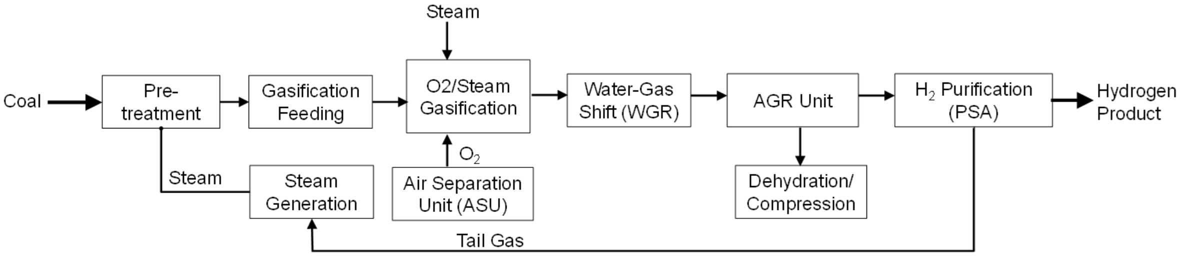

Figure 3-2 Simplified process diagram for oxygen/steam entrained flow gasification using Victorian brown coal (VBC-EFR)

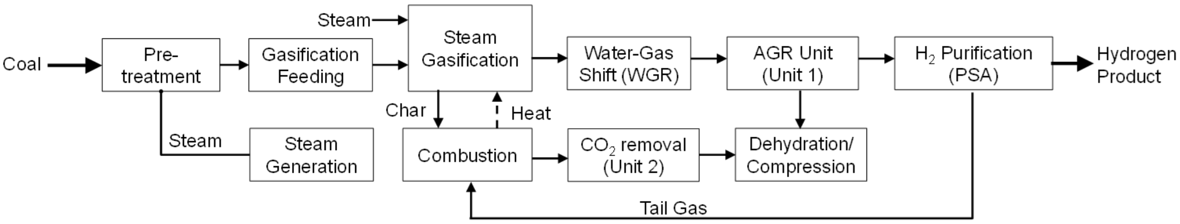

Figure 3-3 Simplified process diagram for fluidised bed gasification using steam and Victorian brown coal (VBC-DFB)

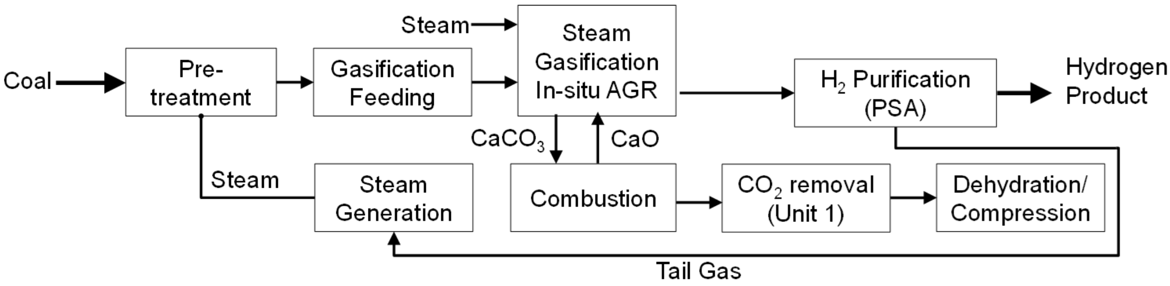

Figure 3-4 Simplified process diagram for fluidised bed gasification with in-situ acid gas removal using steam and Victorian brown coal (VBC-DFB+CAP)

3.2 PROCESS DECSCRIPTION

3.2.1 Pre-treatment

All processes involved pre-treatment of the coal feedstock. Coal was crushed prior to being fed into gasification units. QBC was converted to a 65% solid slurry after crushing while crushed VBC was kept dry and transported to the gasifier through a pneumatic conveyor (process CO_2 was used as a carrier gas).

3.2.2 Gasification

Coal, steam and oxygen are simultaneously fed to the entrained flow gasifier (EFR) which is operated at high temperature (1300 – 1350 ^{o}C) and pressure, along with extremely turbulent flow conditions to facilitate a high and rapid conversion. Oxygen was provided from an air separation unit (ASU).

Coal and steam were simultaneously fed to the dual fluidized bed gasification reactor (DFB) (VBC-DFB & VBC-DFB+CAP). The DFB is a recirculating unit operating at atmospheric pressure where half operates as a gasifier while the other half acts as a fluidized bed combustion reactor. Coal and steam are first fed to the fluidized bed comprised of hot inert material. Syngas then leaves the gasifier before the inert bed material and formed char are moved to the combustion chamber. This heats the bed material before it is fed back to the gasification chamber. Calcium oxide (CaO) is used in-situ to capture CO_2 (though the CaO–$CaCO_3$ equilibria) in VBC-DFB+CAP only.

3.2.3 Water gas shift reaction

The syngas stream is sent to a reactor where CO and H_2O are converted to H_2 and CO_2 through the water gas shift reaction (WGS). The reaction is catalysed by a Cobalt-Molybdenum (CoMo) catalyst, achieving efficiencies of 95-99%. A WGS reactor is not required for VBC-DFB+CAP as the inclusion of the CaO sorbent for CO_2 capture already forces the WGS equilibrium to its limit in the gasification unit (through removal of gaseous CO_2).

3.2.4 Acid gas removal

Acid gas (H_2S and CO_2) must be removed process and flue gas streams. For all cases bar VBC-DFB+CAP (where acid gas removal occurs within the gasifier), acid gas removal occurs in dedicated units.

In ERF scenarios, dry sorbent desulfurization using solid FeO was first used to remove H_2S from the syngas stream. Owing to the high sulfur content of QBC, the syngas stream was first passed through a catalytic hydrolysis reactor for the QBC case to convert carbonyl sulfide (COS) to H_2S.

The Selexol process was used for CO_2 removal from the syngas stream for all cases except VBC-DFB+CAP. This process utilises a mixture of polyethylene glycol dimethyl ethers to physically adsorb CO_2 at high pressures (30 bar) and cryogenic temperatures. A single stage Selexol process was used for EFR cases while VBC-DFB employed a two stage Selexol process for removal of both H_2S and CO_2.

Due to the low pressure of the flue gas stream, chemical (rather than physical) adsorption was used for flue gas treatment. Amine CO_2 removal was selected for all four cases using a mixture of methyl diethanolamine (MDEA) and piperazine (PZ).

Very high CO_2 capture rates (>99%) were modelled across all scenarios to deliver H_2 at similar emissions intensity to the methane scenarios. Captured CO_2 is pressurised to 15.3 MPa through multi-stage compression. Intermediate dehydration using triethylene glycol is conducted at 2.1 MPa. The CO_2 from the flue gas requires an extra stage of compression owing to its low capture pressure.

3.2.5 Hydrogen purification and compression

The product stream undergoes purification to 99.97% H_2 through a pressure swing adsorption (PSA) unit. Following this, the hydrogen is compressed to 81 bar for pipeline injection.

3.3 PROCESS MODELLING

The process modelling results are presented in Table 3.1 and the utility requirements are given in Table 3.2.

Table 3.1 Process modelling results for coal gasification scenarios

Result		**Case**			
		QBC-EFR	**VBC-EFR**	**VBC-DFB**	**VBC-DFB+CAP**
H_2 produced (kt/y)		100	100	100	100
Coal input (kt/y)	Wet (kt/y)	843	1959	2017	2119
	Dry, ash free (kt/y)	613	680	700	735
	HHV (MW)	767.0	782.2	805.3	846.2
	LHV (MW)	617.9	612.2	630.2	662.2
Electricity consumption (MW)	ASU	72.8	30	–	–
	Auxiliary	39.1	27.7	90.0	78.0
Net energy efficiency (%)	HHV	56.1	58.7	55.1	53.4
	LHV	57.1	62.2	57.9	56.3
CO_2 (kt/y)	Produced	1993	1634	1715	2091
	Captured	1975	1611	1647	1982
	Emitted	18	23	68	109
CO_2e (t-CO_2/t-H_2)		0.18	0.23	0.68	1.09

Table 3.2 Utility requirements for coal gasification scenarios

Item	**Unit**	**QBC-EFR**	**VBC-EFR**	**VBC-DFB**	**VBC-DFB+CAP**
Pumps and compressors	MW	111.90	57.70	90.00	78.00
Rotary dryer	MW	0.00	2.23	2.29	2.41
Coal handling	MW	1.37	0.43	0.44	0.46
Coal milling	MW	8.07	2.52	2.59	2.72
Air coolers	MW	0.22	0.19	0.24	0.49
Electricity generation	MW	-22.80	0.00	0.00	0.00

Total electricity required (MW_e)	MW	98.8	63.1	95.6	84.1
Total electricity per tonne of H_2 ($MWh_e/t\ H_2$)	MW/t-H_2	7.9	5.0	7.6	6.7
Natural gas	GJ/t-H_2	0	0	11.2	33.7

3.4 FINANCIAL ANALYSIS

3.4.1 Capital Cost

The breakdown of total capital cost is presented in Figure 3-5. The gasification island and the CO_2 capture and processing train are the two largest equipment components of the capital cost for all scenarios. The entrained fluidised bed gasifiers are more costly than the dual fluidised bed units, and are typically also associated with larger ASU requirements. However, the DFB reactors are less technologically mature and subject to a greater contingency. In addition, the requirement for a dual Selexol capture system for CO_2, rather than a single stage unit drives the VBC-DFB scenario to have the highest capital cost. The additional moisture content of the brown coal contributes to the substantially higher cost for coal handling, storage and pretreatment for all VBC scenarios compared to QBC.

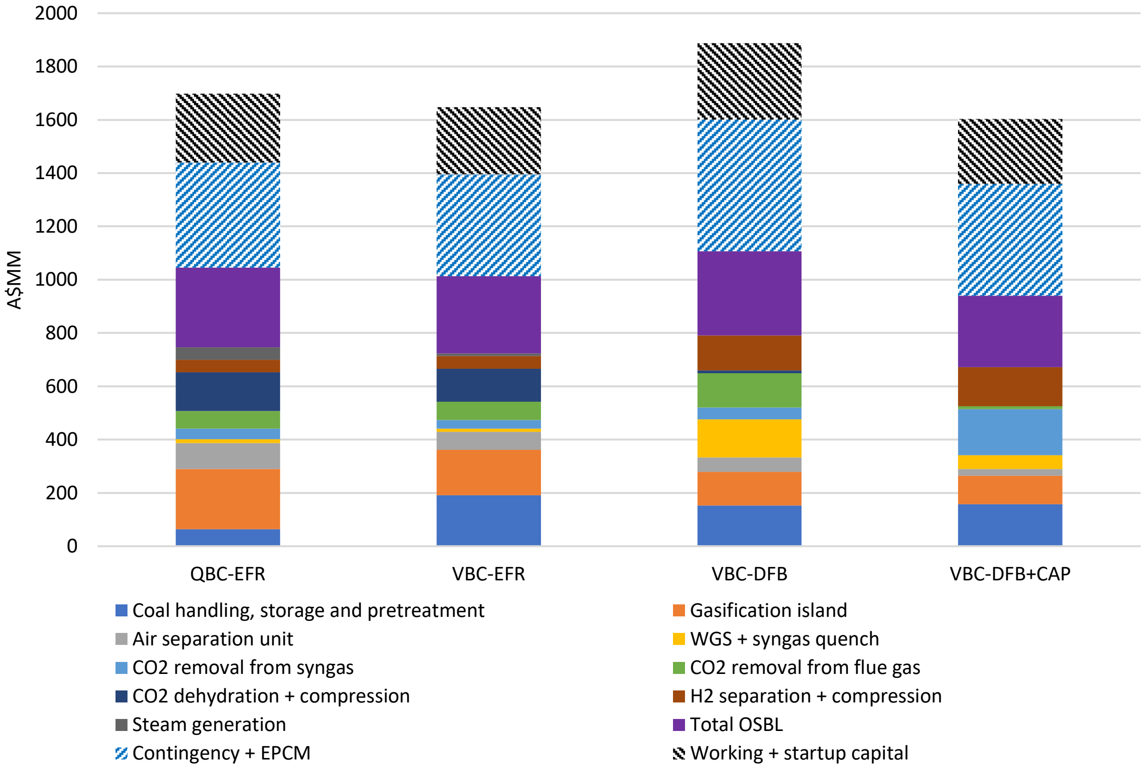

Figure 3-5 Capital cost breakdown for blue hydrogen from coal gasification

3.4.2 Operating Cost

The total utility cost for each case (in \$/kg-H_2) is detailed in Figure 3-6, with electricity demand dominating the utility cost and providing a differentiating factor between scenarios. Despite being the only scenario with ISBL power generation, the QBC scenario has the greatest electricity demand, owing to the larger power requirement of slurry pumping and black coal milling (compared to softer VBC). Likewise, the feedstock cost for the QBC is significantly more than the VBC scenarios reflecting the higher price of the higher quality Queensland black coal. The DFB scenarios have additional natural gas demands (to provide additional energy into the reactor and CO_2 capture systems) which raises their overall utility cost above the EFR scenarios.

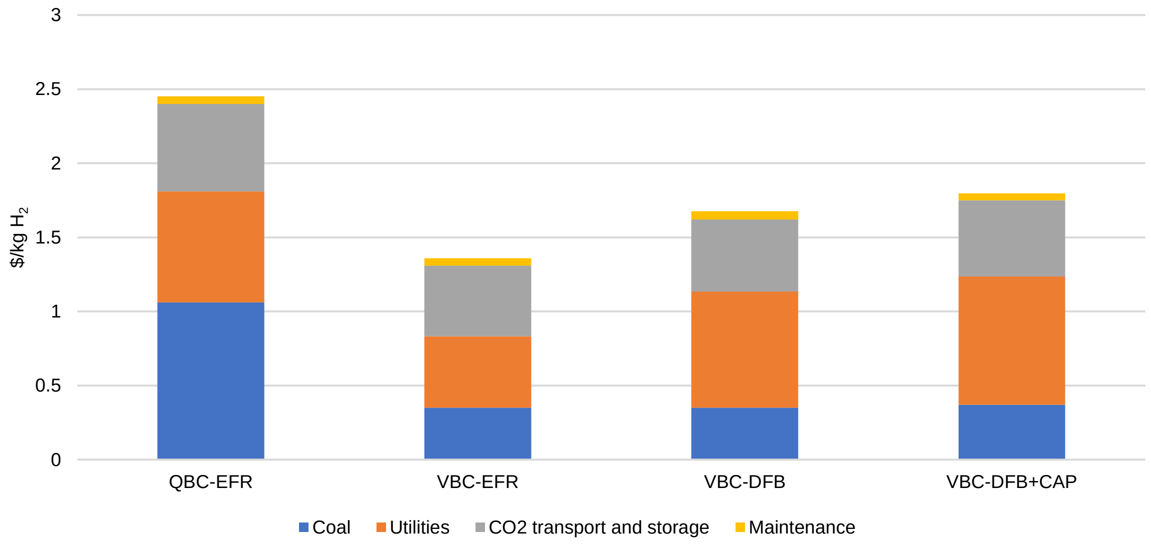

Figure 3-6 Operational expenditure for blue hydrogen from coal gasification

3.4.3 Levelized Cost

The final LCOH for each case is presented in Figure 3-7, with all scenarios ranging between $4 - 5/kg. Victorian Brown Coal (VBC) is the most cost-effective feedstock. Owing to its technological maturity, an entrained flow gasifier is the most cost-effective way to process VBC, resulting in an LCOH of $4.12/kg. Capital cost dominates all cases which use the VBC as feedstock; whereas the high feedstock and utility requirements of the QBC-FBR scenario mean it has the highest LCOH (QBC-EFR – $4.90/kg)[16]. The DFB scenario with Selexol carbon capture has the highest cost of the VBC scenarios, due to its high capital expense (VBC-DFB – $4.83/kg). While DFB with in-situ carbon capture offers notable savings in capital expense, the increased utility load make it less competitive (VBC-DFB+CAP – $4.48).

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[16] The finance cost is shown as negative here to reflect that the QBC-EFR scenario produces excess steam which was considered a by-product with a sale price.

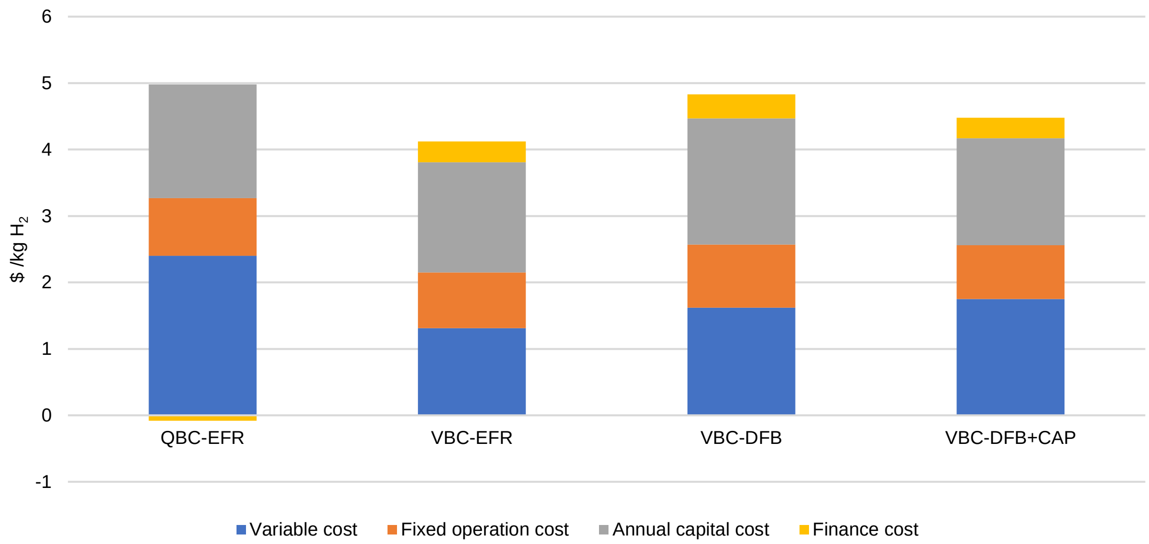

Figure 3-7 Levelized cost of hydrogen for blue hydrogen from coal gasification

3.4.4 Sensitivity Analysis

A detailed sensitivity analysis[17] was conducted to better understand the impact of variations in the dominant cost factors (feedstock cost and capital cost) and to identify targets that may lead to hydrogen production costs below the $2/kg target originally expressed in the Low Emissions Technology Statement[18]. The results are summarised in Figure 3-8 through Figure 3-10.

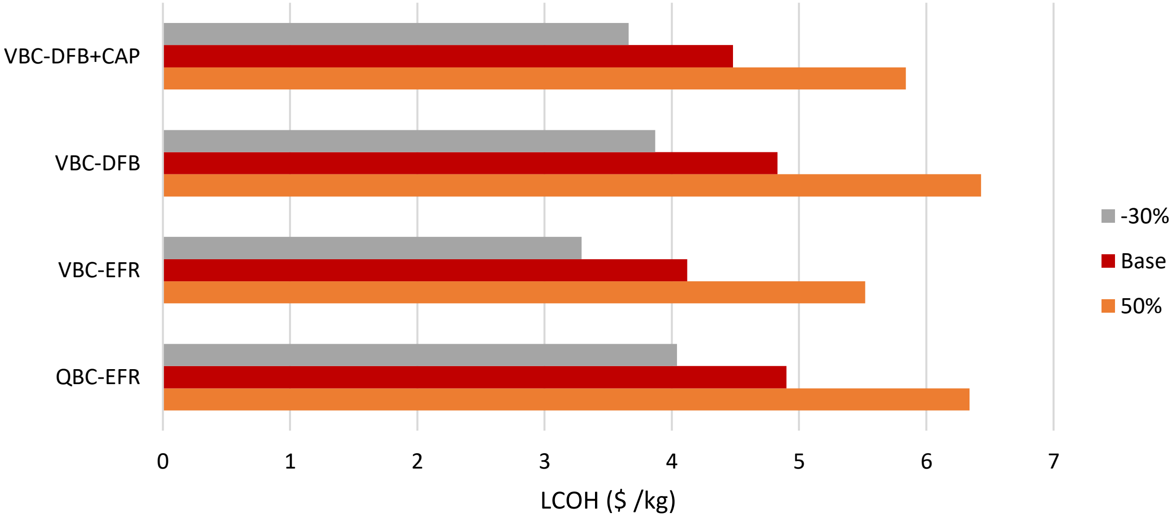

Figure 3-8 Capital cost sensitivity analysis for blue hydrogen from coal gasification

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[17] Full details of this process are given in the Final Milestone Report for RP1.2-02. A copy of this report may be obtained by contacting the Future Fuels CRC.

[18] LOW EMISSIONS TECHNOLOGY STATEMENT 2021 Australian Government Department of Industry, Science, Energy and Resources

Under a decreased capital cost scenario (-30%), the LCOH of the most competitive technology (VBS-EFR) drops to \$3.29/kg. The ranking of process competitiveness does not change when the capital cost is varied. However, it is worth noting that since the DFB is a novel and less mature technology, the capital cost for the DFB scenarios is subject to greater potential future decreases than for the EFR scenarios. Specifically the capital cost of the VBC-DFB+CAP scenario needs to decrease by ~22% for DFB as competitive as EFR (assuming no change in EFR capital costs).

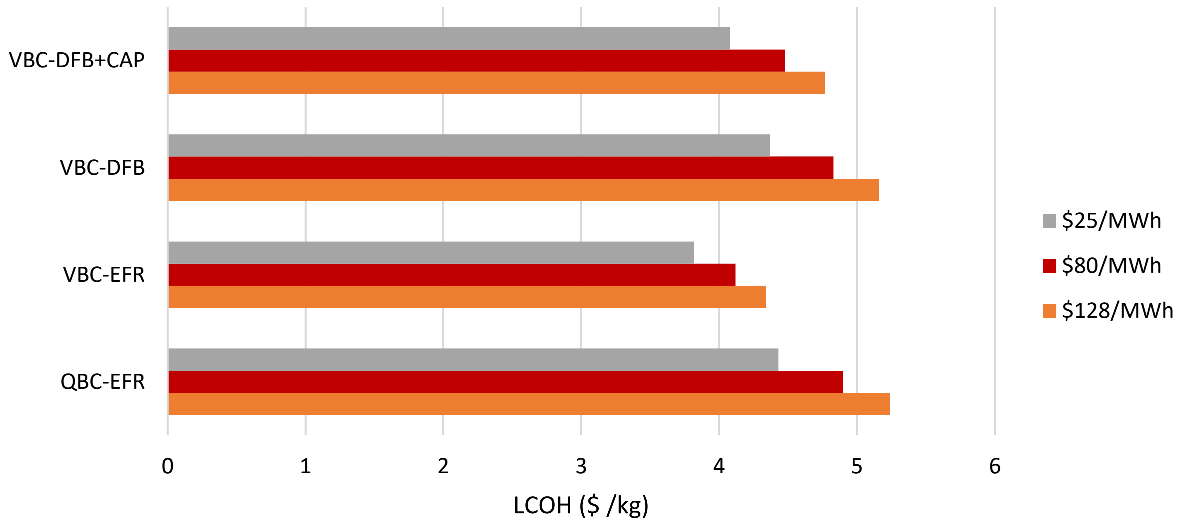

Figure 3-9 Electricity price sensitivity analysis for blue hydrogen from coal gasification

Electricity price has a tangible impact on the LCOH of all scenarios, although the variations explored here do not change the competitiveness ranking by process. Under high electricity prices (\$128/MWh), the LCOH of the most competitive scenario increases to \$4.34/kg and decreases to \$3.82/kg under low electricity prices (\$25/MWh).

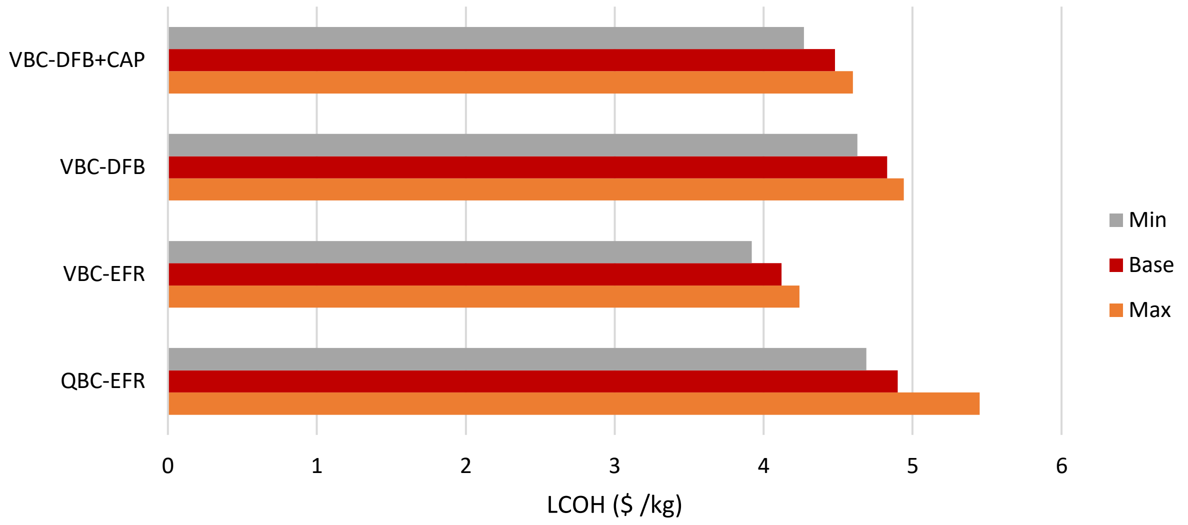

Figure 3-10 Coal feedstock price[19] sensitivity analysis for blue hydrogen from coal gasification

The results of the sensitivity of LCOH to coal price is given in Figure 3-10. Due to the inherently low price of VBC, the VBC scenarios demonstrate little sensitivity to the cost of coal. However, the QBC scenario is much more sensitive to coal price. There is no variation in process competitiveness ranking under a mutually changing coal price. For QBC to be competitive with VBC, the cost of QBC must decrease by 70% to \$1.05/GJ, without any change in the cost VBC (\$1.5/GJ).

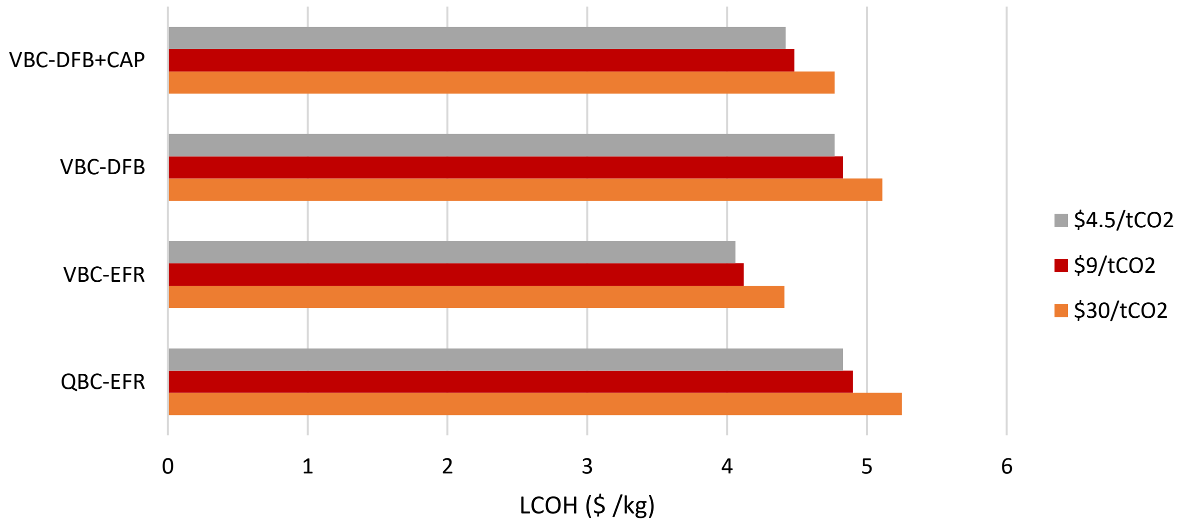

Figure 3-11 CO_2 storage price sensitivity analysis for blue hydrogen from coal gasification

The impact of varying CO_2 storage price (Figure 3-11) has a relatively minor impact on LCOH and does not change the rankings of process competitiveness. This is a similar result to the

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[19] QBC varied between \$2.8 – 5.3 /GJ. VBC varied between \$0.64 – 2/GJ.

blue hydrogen from methane scenarios (not shown in this report), despite the coal gasification scenarios needing to sequester more than double the CO_2 per tonne of hydrogen produced. Halving the CO_2 storage cost reduces the overall LCOH by less than 2% for all coal gasification scenarios, whilst a more than tripling of CO_2 storage cost (to $30/t$CO_2$) increases the LCOH by ~5-7%.

The sensitivity analysis revealed that no single factor is able to reduce the LCOH below $3/kg, much less the $2/kg target. Indeed, minimising all the major influences on the LCOH (i.e. -30% capital cost, $25/MWh electricity price, $0.64/GJ VBC and $4.5/tCO2 storage cost) reduces the LCOH for the VBC-EFR scenario to $2.7/kg H_2. It is therefore challenging to identify means by which blue hydrogen using coal gasification may meet H_2 production target costs. Whilst, not explored in the modelling reducing the CO_2 capture rate in line with the blue hydrogen from methane scenarios (i.e. to ~90% from ~99%) in combination with the most optimistic assumptions on capital and operational costs may be an additional means of cost reduction, with the obvious trade-off in sustainability.

4 BIOMASS

4.1 INTRODUCTION

Biomass is a popular feedstock for green hydrogen production as CO2 released from the combustion or gasification of the biomass is recaptured when the biomass if regrown. Hydrogen produced from biomass therefore requires very little carbon capture and storage to make the process truly carbon neutral. Further, CCS capture rates can be increased to make the hydrogen carbon negative. Doing so brings about an additional revenue stream, through carbon offsets. However, scrutiny is required around the source and lifecycle of the biomass feedstock to verify net lifecycle emissions.

Two predominant methods for converting biomass into hydrogen are pyrolysis and gasification. Both involve the high temperature decomposition of cellulosic biomass into syngas. Pyrolysis occurs without the addition of any oxidising agents (air and/or steam) while gasification involves adding steam/oxygen to partially oxidise the biomass. In addition to the syngas, a series of condensable gases, ash and tar are produced. Hence, the pyrolysis and gasification products require significant post-processing to render high purity hydrogen.

Four scenarios were investigated for the pyrolysis of biomass (BIO-P). The tar treatment method was varied between steam reforming (BIP-P-STM) and OLGA treatment (BIO-P-OLGA). Carbon capture was employed, such that all cases could be considered carbon negative. The carbon capture varied from syngas only to syngas plus flue gas. The four combinations of methods were labelled **BIO-P-STM-LO, BIO-P-STM-HI, BIO-P-OLGA-LO** and **BIO-P-OLGA-HI** respectively (where LO indicates capture of CO_2 from the syngas only while HI indicates capture from syngas and flue gas). A fifth case was investigated involving gasification of biomass. OLGA tar removal was utilised and CO_2 was removed from sygnas only. This case was labelled **BIO-G**. Bagasse was selected as a feedstock for all cases and a 25,000 t/y H_2 production basis was set.

4.2 PROCESS DESCRIPTION

The block flow diagrams are for each case are given in Figure 4-1 to Figure 4-3.

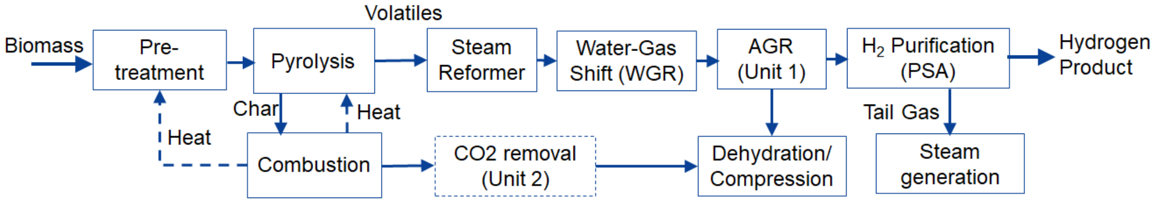

Figure 4-1 Simplified process diagram of BIO-P-STM-LO/HI: Biomass pyrolysis with steam tar reforming – CO_2 removal from flue gas (unit 2) exists in BIO-P-STM-HI only

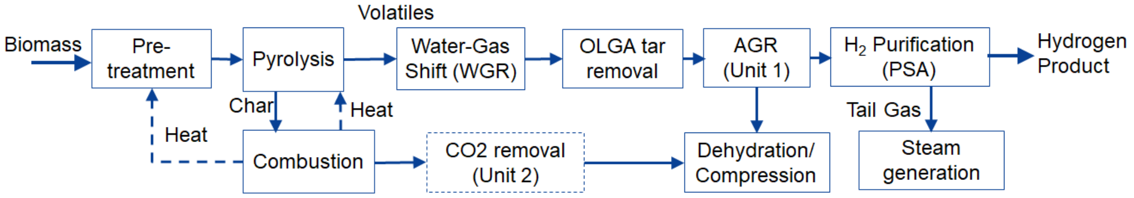

Figure 4-2 Simplified process diagram of BIO-P-OLGA-LO/HI: Biomass pyrolysis with OLGA tar removal – CO_2 removal from flue gas (unit 2) exists in BIO-P-OLGA-HI only

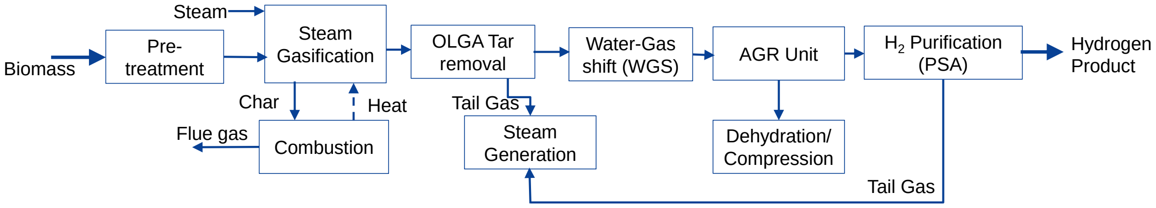

Figure 4-3 Simplified process diagram of BIO-G: Biomass gasification with OLGA tar removal

4.2.1 Pre-treatment

The bagasse is first dried to a moisture content of below 15 wt% – required heat is supplied from combustion of char in pyrolysis scenarios. After drying, bagasse is milled to a particle size of 2-3 mm.

4.2.2 Pyrolysis (excl. BIO-G)

Biomass is heated to above 650 °C in a fluidized bed reactor. Specifically, the Rapid Thermal Processing process from Ensyn Technologies was selected. The gases are rapidly cooled to prevent secondary cracking reactions, maximising yield.

4.2.3 Gasification (BIO-G only)

Gasification occurs in a dual fluidized bed gasifier (DFB), composed of separate gasification and combustion chambers. Biomass and steam enter the gasifier at temperatures below 900 °C where the biomass undergoes pyrolysis and heterogenous char gasification. Syngas and volatile gases leave the gasifier. The solid fluidised bed, along with the remaining char and other solid materials enter the combustion chamber where the char undergoes combustion. This reheats the sand before it re-enters the gasification chamber.

4.2.4 OLGA tar treatment (BIO-P-OLGA-LO/HI & BIO-G only)

OLGA tar removal uses Canola oil as a solvent to remove tar from the syngas at atmospheric pressure. The syngas must first be cooled before the solvent absorbs tar. The solvent is regenerated through steam stripping of the entrained tar. The tar is the combusted for process steam generation.

4.2.5 Water gas shift

The water gas shift reaction (WGS) is used to increase the H_2 yield of the syngas. A Cobalt-Molybdenum (CoMo) catalyst is used to convert water and CO to H_2 and CO_2. In cases using OLGA tar removal, WGS occurs following tar removal. In steam tar reforming scenarios, WGS occurs before tar removal.

4.2.6 Steam tar reforming (BIO-P-STM-LO/HI only)

Tar is reformed with steam at high temperature (>900 ºC) to produce syngas. A series of dolomite and nickel catalysts are used to facilitate the reforming. Heat is recovered from the tail gas for steam generation.

4.2.7 Acid gas removal

CO_2 is removed from the shifted syngas through a single stage Selexol process for all scenarios. As the process employs physical adsorption, it occurs at high pressures and cryogenic temperatures to ensure high capture rates. In flue gas CO_2 capture scenarios, CO_2 produced from the combustion of PSA tail gases and tar (BIO-P-OLGA-HI only) was explored. As the flue gas streams are at atmospheric pressure, amine stripping (methyl diethanolamine + piperazine) was used to remove CO_2. All captured CO_2 was dehydrated and compressed for storage.

4.2.8 Hydrogen purification

The remaining hydrogen stream is purified using pressure swing adsorption. The tail gas is combusted to generate process steam.

4.3 PROCESS MODELLING

The five (5) cases were modelled using Aspen Plus V11®. The process modelling results are presented in Table 4.1. The utility demand is presented in Table 4.2.

Table 4.1 Mass balance for biomass pyrolysis and gasification

Case		BIO-P-STM-LO	BIO-P-STM-HI	BIO-P-OLGA-LO	BIO-P-OLGA-HI	BIO-G
Bagasse feed (kg/kg-H_2)		23.88	23.88	52.75	52.75	34.82
CO_2 emissions (kg/kg-H_2)		14.34	2.24	33.06	7.54	17.25
Captured CO_2 (t/t-H_2)		12.61	24.70	18.77	44.28	16.27
Energy efficiency (%)	**LHV**	28.8	26.5	13.6	12.5	38.6
	HHV	56.2	46.9	28.1	23.3	20.6

The utility demand for each scenario is presented in Table 4.2.

Table 4.2 Electricity demand of biomass pyrolysis and gasification scenarios

Unit	Demand (MW)				
	BIO-P-STM-LO	BIO-P-STM-HI	BIO-P-OLGA-LO	BIO-P-OLGA-HI	BIO-G
Pumps and compressors	32.20	39.29	32.40	36.76	21
Rotary dryer	0.68	0.68	1.50	1.50	1
Biomass handling	0.13	0.13	0.29	0.29	0.19
Biomass milling	0.77	0.77	1.70	1.70	1.13
Air coolers	0.01	0.02	0.12	0.16	0.03
Electricity generation	-3.20	-3.20	-18.80	-18.80	-16.1
Total	30.59	37.68	17.20	21.60	7.25

There is a substantial difference in the biomass feed between the processes. Pyrolysis processes that use steam tar reforming require the least amount of bagasse feedstock per kilogram of hydrogen. The gasification case requires nearly 50% more bagasse to produce the same amount of hydrogen while the pyrolysis cases utilising OLGA tar removal require more than twice the amount of bagasse for the same amount of hydrogen production. The OLGA cases are far less efficient on a biomass feed perspective as the tar (which is derived from the biomass feedstock) is combusted rather than reformed into syngas, where it provides an additional source of H_2.

There are also substantial differences in CO_2 emissions between the processes. Naturally, the processes which capture CO_2 in the flue gas (HI) have substantially lower CO_2 emissions than the rest. However, the cases employing steam tar reforming have substantially lower CO_2 emissions than the OLGA cases as a result of reduced biomass input.

4.4 FINANCIAL ANALYSIS

4.4.1 Capital Cost

The breakdown of the total capital cost of all five biomass scenarios is given in Figure 4-4. The CO2 capture train (capture, dehydration and compression) is the largest cost equipment package across all scenarios, and in the BIO-P-STM-HI and BIO-P-OLGA-HI cases, more than 50% of the equipment cost. In both HI scenarios the additional amine unit required to capture CO_2 from the flue gas is more than the Selexol capture unit from the syngas, predominately due to the lower pressure operation and larger equipment size. This is a similar result to the coal gasification cases where the Selexol flue gas capture unit was a substantial capital cost component. The impact of these additional CO_2 capture units is explored in 4.4.3 where the resulting revenue stream is analysed (Figure 4-6).

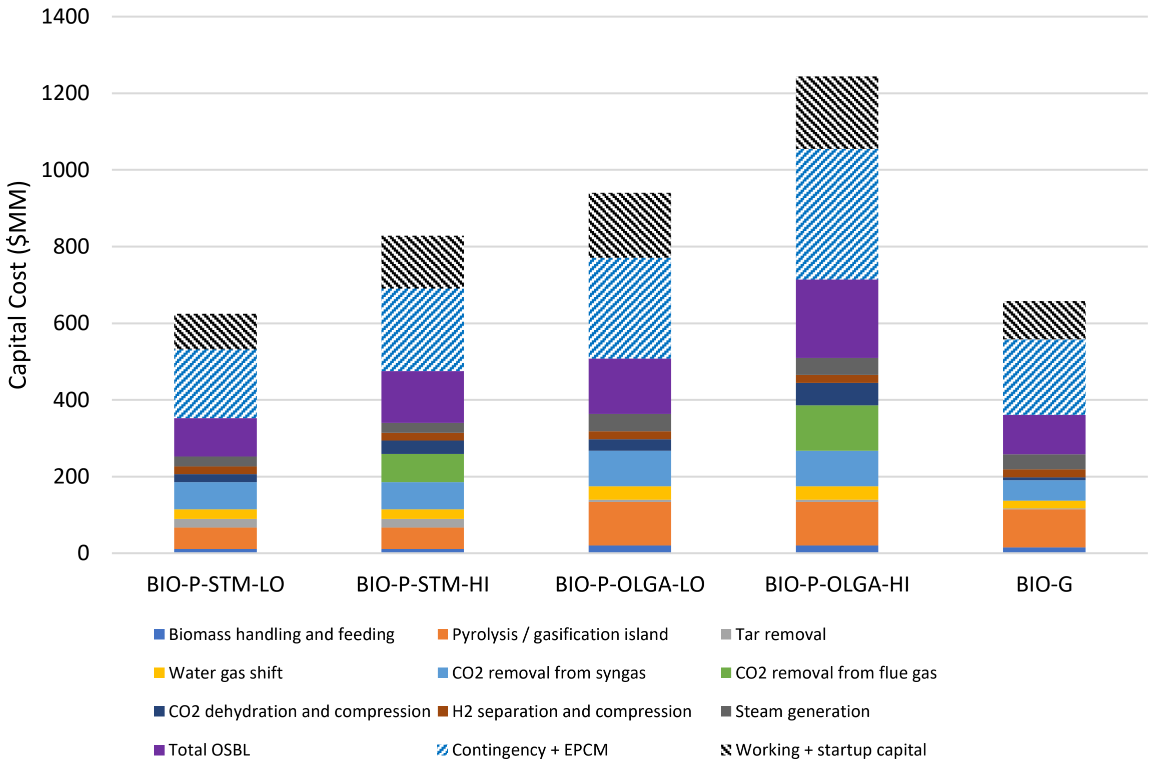

Figure 4-4 Capital cost breakdown for green hydrogen from biomass pyrolysis and gasification

The cases which employ OLGA tar removal have a substantially higher CAPEX than those that utilise steam tar reforming. This is predominantly due to the greater pyrolysis unit cost, which results from the much greater demand of biomass per unit hydrogen (as discussed in section 4.3). The OLGA tar removal itself is significantly less capital intensive (by a factor of 4 times) than the steam tar reforming. However, as the tar reforming units are only a small component of the equipment cost (<7%) this is not enough to offset the greater capital intensity of the pyrolysis unit. The gasification scenario had a slightly greater capital cost than the cheapest pyrolysis scenario. Being a dual fluidised bed gasifier, and therefore associated with higher degree of uncertainty, the gasifier cost drives the capital cost of this case beyond case BIO-P-STM-LO. The BIO-G scenario has the lowest tar removal cost of any scenario. As relatively immature technology with low TRL all scenarios had first-of-a-kind contingency factors applied.

4.4.2 Operating Cost

The variable operating costs (i.e. feedstock and utility costs) for the biomass pyrolysis and gasification scenarios are presented in Figure 4-5. Electricity is the dominant factor accounting for between 65% (BIO-G) and 85% (BIO-P-STM-HI/LO) variable operating costs. Electricity costs are reduced in the OLGA cases as the additional steam generation that occurs is used to produce electricity, offsetting approximately half the electricity requirements for those scenarios.

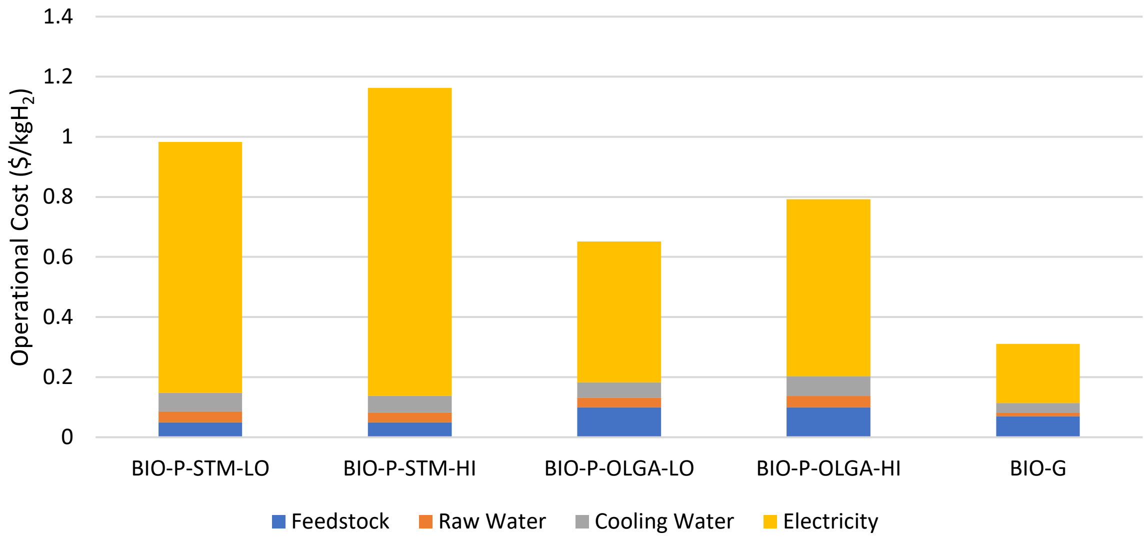

Figure 4-5 Operational costs for green hydrogen from biomass gasification and pyrolysis

Feedstock costs and cooling water demand are the next most significant component of the variable operating costs. Whilst the feedstock demand for the OLGA scenarios is more than double the STM scenarios, they are relatively small and do not have a significant impact on the overall cost competitiveness.

4.4.3 Levelized Cost

The levelized cost of hydrogen production for each scenario is presented in Figure 4-6, with values ranging from \$4.66 to \$9.62 /kg H_2. The most competitive scenario is the BIO-G at \$4.66 / kg H_2. The capital cost is the dominant factor influencing the LCOH for all technologies, which in turn makes the fixed operational costs (labour, maintenance, taxes and insurance) the second most influential factor as these are a fixed percentage of capital costs.

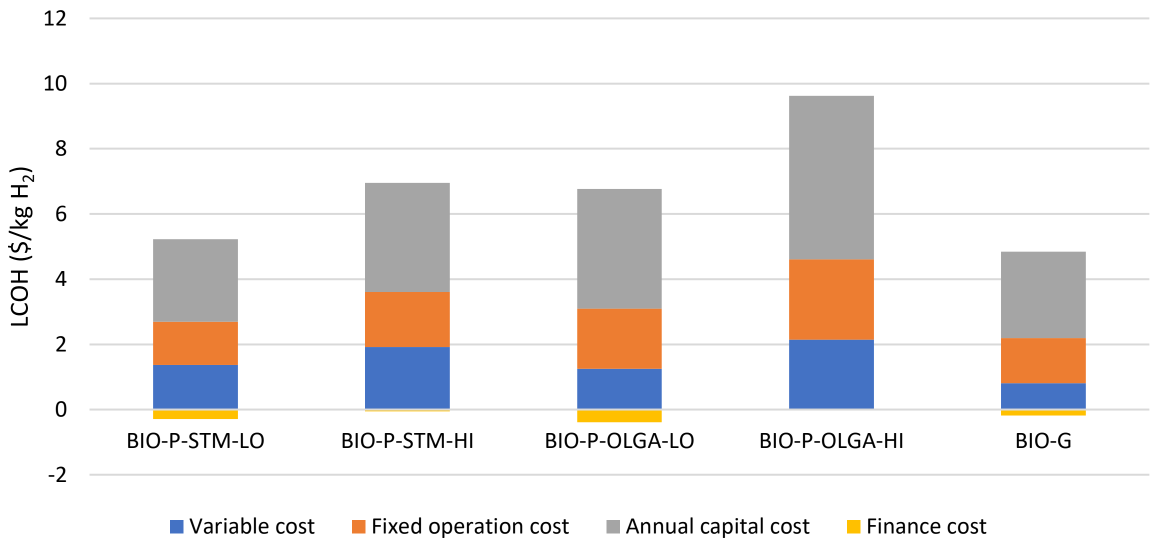

Figure 4-6 Levelized cost of hydrogen for green hydrogen from biomass gasification and pyrolysis

Finance costs are potentially negative (depending on scenario) as they also include the sale of the negative carbon emissions (from the CO_2 capture) as offsets (i.e. Australian Carbon

Credit Units, ACCUs). Only in the LO scenarios and the BIO-G scenario do the offsets (priced at \$25/tCO_{2e}) generate enough revenue to lower the LCOH. The additional capital and operating expense for the second capture unit from the flue gas in the HI cases is not offset by the ACCU revenue.

4.4.4 Sensitivity Analysis

[1]. The results are summarised in Figure 4-7 through Figure 4-9.

A detailed sensitivity analysis[20] was conducted to better understand the impact of variations in the dominant cost factors (feedstock cost and capital cost) and to identify targets that may lead to hydrogen production costs below the \$2/kg target originally expressed in the Low Emissions Technology Statement

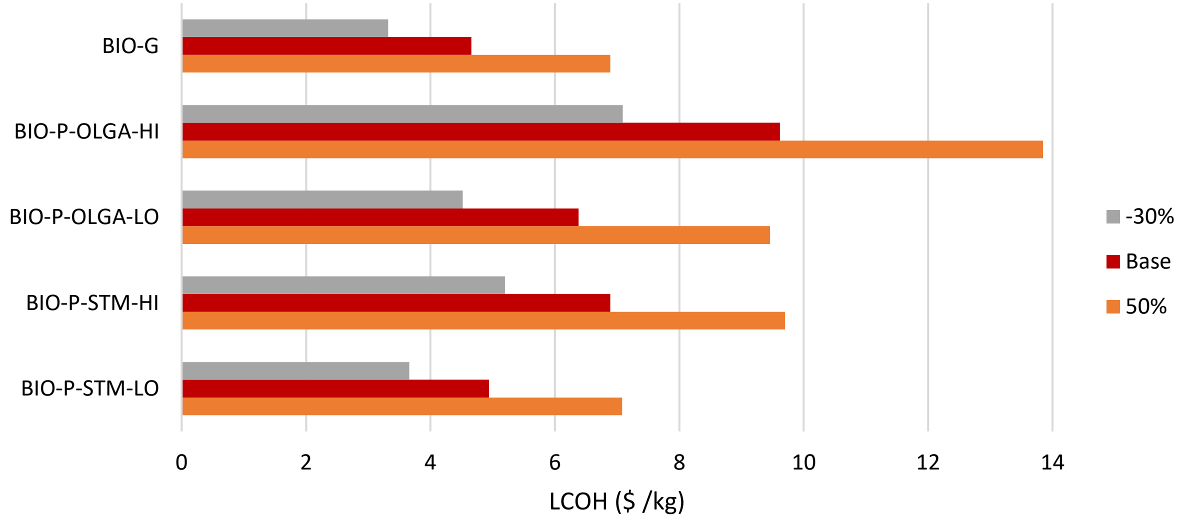

Figure 4-7 Capital cost sensitivity analysis of biomass gasification and pyrolysis

As the dominant influence on LCOH, the capital cost sensitivity analysis demonstrated the largest variability. Decreasing the capital cost by 30% saw the BIO-G and BIO-P-STM-LO cases yield levelized costs below \$4/kg H2, however, none approached the \$2/kg target range. The competitiveness between technologies didn't change with this sensitivity analysis, although the higher capital cost component of the OLGA scenarios had a more significant impact than for the STM scenarios. Indeed, a 50% capital cost increase reduces the difference between the BIO-P-OLGA-LO and BIO-P-STM-HI cases to ~2.5% (down from 8% in the base case), despite the substantial additional CO_2 capture costs associated with the HI scenario.

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[20] Full details of this process are given in the Final Milestone Report for RP1.2-02. A copy of this report may be obtained by contacting the Future Fuels CRC.

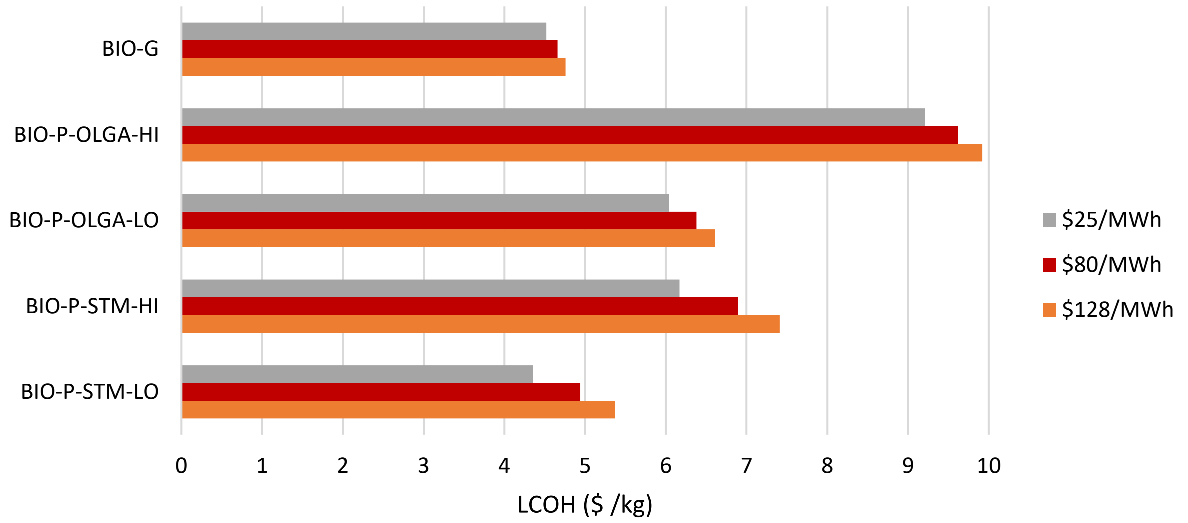

Figure 4-8 Electricity price sensitivity analysis of biomass gasification and pyrolysis

Electricity price did not have as large an impact on LCOH, however, the dramatic difference in electricity usage between the STM and OLGA scenarios, meant that the competitiveness of the technologies changed under this sensitivity analysis. Specifically, the lower electricity price reduced the LCOH for the BIO-P-STM-LO scenario below the BIO-G scenario, which was the lowest cost option under the base case assumptions. parameter which yielded a sensitivity in the competitiveness order of processes was electricity price. This is due to the relevant portion of LCOH that is driven by variable operating costs for the pyrolysis cases compared to gasification, where capital costs are more prominent. However, the sensitivity analysis did not identify that electricity price alone could reduce LCOH to the $2/kg target range, indeed, no scenario demonstrated an LCOH below $4/kg.

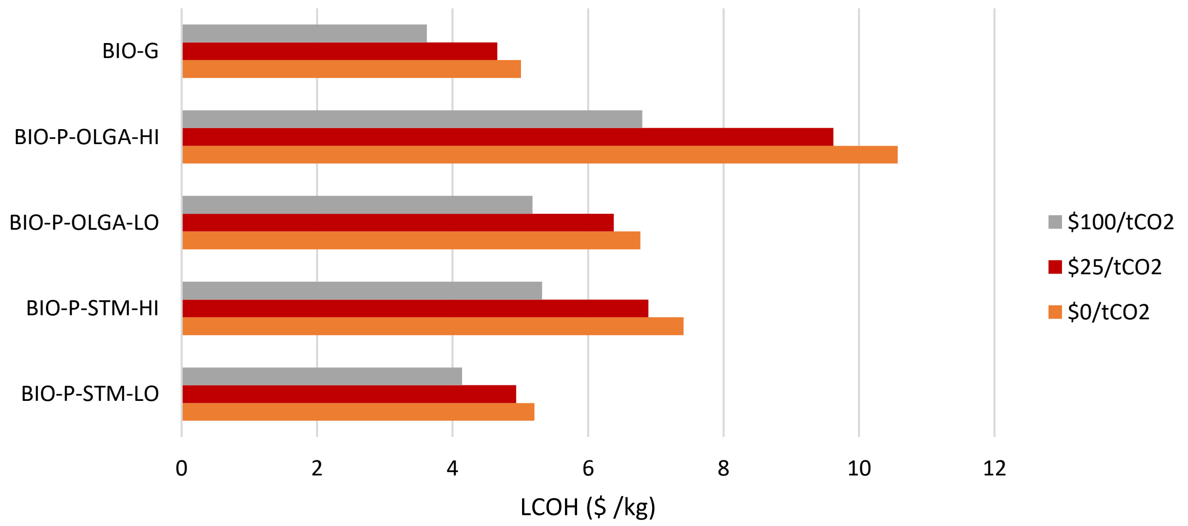

Figure 4-9 ACCU price sensitivity analysis for biomass gasification and pyrolysis

LCOH demonstrated some sensitivity to ACCU price, with $100/tCO2 offset price lowering the LCOH of the BIO-G sensitivity analysis to $3.62/kg. Importantly the sensitivity analysis and the underlying base case suggest the additional capital and operating costs imposed by attempting to generate negative emissions are not sufficient to lower the LCOH below $2/kg under realistic ACCU pricing levels. Indeed, the analysis strongly suggests negative abatement coupled with green hydrogen production from biomass, is less competitive than simply green hydrogen production. Minimal sensitivity in LCOH was demonstrated when varying CO_2 storage and biomass prices over the defined ranges.

5 ELECTROLYSIS

5.1 INTRODUCTION

The electrolysis of water is a process by which electricity splits water (H_2O) into hydrogen (H_2) and oxygen (O_2) gases. If this electricity is generated renewably, the hydrogen is considered *green*. This process is achieved by two half-reactions occurring simultaneously at two separate electrodes. These reactions are differentiated by whether electrons are lost through oxidation or gained through reduction. The electrode where oxidation occurs is called the anode, while the electrode where reduction occurs is called the cathode [1]. In an acidic environment, the two half-reactions for the electrolysis of water are [2]:

Anode (oxidation): $$2\mathrm{H_2O} + \mathrm{Energy} \rightarrow \mathrm{O_2} + 4\mathrm{H^+} + 2\mathrm{e^-} \quad \text{Eqn. 1}$$

Cathode (reduction): $$4\mathrm{H^+} + 4\mathrm{e^-} \rightarrow 2\mathrm{H_2} \quad \text{Eqn. 2}$$

Electrolysis cells also contain electrolytes which enable better transport of the electrons and mobile ions through solution. The two most prominent and commercially available electrolyser technologies are:

- **Proton exchange membrane (PEM) electrolysers—**using a solid polymer cation electrolyte, and
- **Alkaline electrolysers (AE)—**using an alkali electrolyte (typically KOH or NaOH) dissolved in water.

Alongside these stand less commercially-developed technologies, which include:

- **Anion exchange membrane (AEM) electrolysers—**using a solid polymer anion electrolyte, and
- **Solid oxide electrolysers (SOE)—**using solid ion-conducting ceramics at elevated temperatures.

A high-level overview of the four listed electrolyser technologies is given in the following sections, with Table 5.1 highlighting the significant differences between the four technologies.

Table 5.1 High level electrolyser technology comparison

Type	Mobile ion	Temperature (°C)	Discharge pressure (barg)	Stack life (h × 10^{-3})	System response	Efficiency (kWh/kg H_2)	CAPEX ($AUD/kW)
PEM	H^+	60–90	<40	20–60	Milliseconds	45–57	390–2870

AE	$\mathrm{OH^-}$	50–80	<0.3	60–90	Seconds	50–67	790–1760
AEM	$\mathrm{OH^-}$	60–80	<35	30	No data	57–80	No data
SOE	$\mathrm{O_2^-}$	800–1100	<25	<20	Seconds	45	940–3170

Of these four technologies, PEM and AE were considered for further analysis due to their commercial readiness and publicly available data.

5.2 PROCESS DESCRIPTION

5.2.1 Electrolyser Stack

An electrolyser, at its simplest unit of composition, is made up of a number electrolyser *cells*—a collective term which refers to the anode, cathode, membrane, and electrolyte. These cells are then connected in series to form an electrolyser *stack*, which in turn are connected together until the target hydrogen production capacity is reached [2].

Proton-exchange membrane (PEM) electrolysers (Figure 5-1) are the most widely known acid-electrolyte cell and use a solid polymer electrolyte to support the exchange of hydrogen ions from the anode to the cathode.

A unique feature of the PEM electrolysis cell (which also extends to other membrane-based cells, such as AEM and SOE cells) is that the solid electrolyte is bound on both sides to sheets of catalysed porous electrodes. This solid assembly enables the production of very compact electrolysis cells, which allows them to have higher energy efficiencies, higher hydrogen production rates, higher operating pressures, and a smaller footprint when compared to alkaline electrolysers [2]. PEM's biggest advantage, due to its ability to operate at near ambient conditions, is its wide operating range and short start-up time (~10 s), making it compatible with renewable energy sources (e.g. wind, solar) since it can quickly meet substantial changes in demand [3].

PEM electrolysis cells are, however, typically more expensive than alkaline electrolysers—due partly to their expensive polymer membrane and the noble metal catalysts used. There is also limited data on the reliable long-term performance of PEM cells, and on the performance degradation of the membrane and catalysts over time [2].

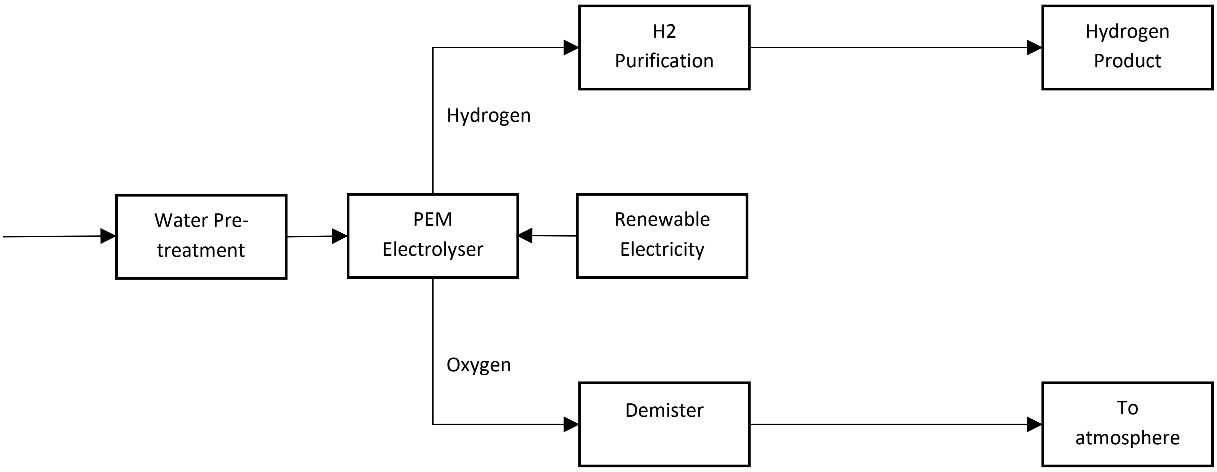

Figure 5-1 Simplified process diagram for PEM

Alkaline electrolysers (AE) (Figure 5-2are the most commercially mature and cost-effective electrolysis technology, having been used at an industrial scale since 1939. Unlike acid-electrolyte cells, an alkaline electrolysis cell supports the transfer of hydroxide ions to the anode using a 30–40 wt% solution of either sodium hydroxide (NaOH) or potassium hydroxide (KOH)—the cheapest and most abundant alkaline hydroxides.

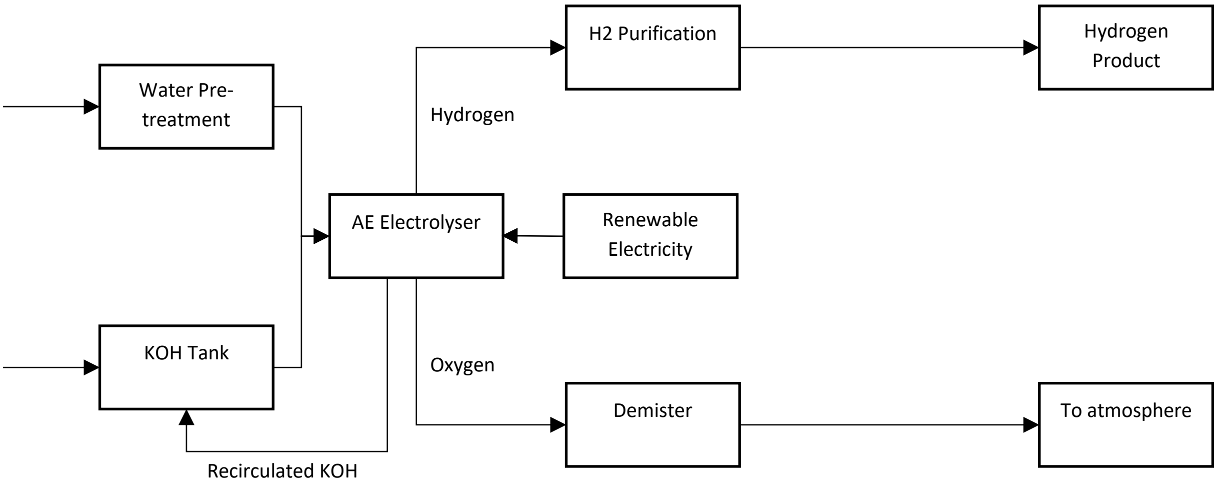

Figure 5-2 Simplified process diagram for AE

The key differences in an alkaline cell's hydrogen production process, when compared to a PEM cell, are [2]:

- Both the cathode and anode and are submerged in water, and
- A hydroxide-selective membrane is required to separate the produced gasses from each electrode since the liquid electrolyte is unable to do so.

The half-cell reactions at each electrode also differ, and these are shown in Equations 3 and 4 below.

Anode (oxidation): $$4OH^- \rightarrow O_2 + 4e^- + 2H_2O$$ Eqn. 3

Cathode (reduction): $$4H_2O + 4e^- \rightarrow 2H_2 + 4OH^-$$ Eqn. 4

Alkaline electrolysers are designed to operate with a fairly constant power supply and are typically unsuitable for use alongside renewable energy sources (unless specialised power control and conditioning equipment is installed). The water used to prepare the electrolyte must also be very pure, since impurities will otherwise accumulate over time or destroy the protective film that forms on the electrode surface [2].

Despite these limitations, alkaline electrolysers do offer technical advantages over their acidic counterparts. Compared to PEM electrolysers, the electrode reactions are both faster and more energetically favourable, enabling the electrodes to be plated with nickel-based catalysts rather than noble metals [2]. Recent technology developments in alkaline electrolysis have additionally demonstrated that state-of-the-art alkaline water electrolysers can deliver comparable performance to PEM water electrolysers, making them a competitive alternative technology option [4].

5.2.2 Ancillary Equipment

To enable this system to continuously generate hydrogen, additional ancillary equipment—such as feed water treatment, oxygen separation, moisture removal, stack cooling, and power supply management—is required. These additional pieces of equipment are often referred to as the *balance of plant*. It is for this reason that electrolyser vendor's often supply their equipment as a containerised *package*; this package often includes the electrolyser, the balance of plant equipment, and any control systems required to support operation [5].

A general process description explaining the typical operation of electrolyser packages is outlined as follows:

1. Potable water, generally sourced from a mains water supply, is first pumped to a water treatment unit to remove impurities. Typical water treatment options include an ion exchange column, a reverse osmosis (RO) unit, a cartridge filter, or a combination of the above.
2. The treated water is then pumped to the electrolyser where electrolysis takes place. Since electrolysers require a direct current (DC) power supply, electrolyser vendors will often include a transformer and rectifier to convert the input electrical current from AC to DC.
3. The formed oxygen gas will typically first pass through a gas separator or demister to remove entrained liquids before being vented to atmosphere.
4. The formed hydrogen gas will also typically first pass through a gas separator or demister to remove entrained liquids before being sent to a compressor.
5. When the electrolyser is operated at low pressures, a compressor may be required to compress and transport the gas to the two downstream purification stages: de-oxygenation and drying.
6. The de-oxygenation unit serves to remove any gaseous oxygen impurities from the produced hydrogen gas stream. Though alternative technologies exist (such as pressure or temperature swing adsorption and membrane-based purifiers), catalytic recombination purifiers are the most frequently used technology and remove oxygen by catalysing the reaction of oxygen and hydrogen gas to form water.
7. The drying unit is the final equipment item in the electrolyser package and serves to remove moisture from the product hydrogen stream. Several technology options exist, including coalescers, refrigeration dryers, membrane and molecular sieve dryers, and desiccant dryers.

In addition to the above, electrolyser packages also often include:

- A cooling system—to regulate the electrolyser's temperature,
- A control system—to manage start-up, shut-down, and normal operation of the package, and
- Utility connections—such as instrument air, nitrogen purge lines, and HVAC systems.

5.3 PROCESS MODELLING

As an emerging green hydrogen production route, the design basis for the process modelling was 25,000 tonnes per year (~3171 kg/h at 90% availability). Electrolysers have not yet become common unit operations in process modelling software so each unit was modelled empirically:

- **Alkaline electrolyser (AE)**—simulated using a custom model that provided add-in data to the HYSYS model, and
- **Proton exchange membrane (PEM) electrolyser**—simulated as a standard electrolyser, based on the enthalpy and Gibbs free energy of the reaction, HHV and LHV of the produced hydrogen gas, Faraday efficiency of the electrolyser (a function of the electrodes and PEM internal characteristics) as well as the polarization curve around the electrodes as a function of operating and maximum current density.

The mass balances are summarised in Table 5.2.

Table 5.2 Process modelling results for electrolysis

Result	Parameter (Units)	AE	PEM
Feed water	Flow rate (kg/h)	28,340	28,340
	Pressure (bara)	4.0 – 15.1	4.0 – 5.1
	Temperature (°C)	30 – 80	30 – 80
	Purity (%)	100%	100%
	Trace components	Salts	Salts
Hydrogen	Flow rate (kg/h)	3,171	3,171
	Pressure (bara)	11 (stack discharge pressure) 81 (battery limit)	11 (stack discharge pressure) 81 (battery limit)
	Temperature (°C)	25	25
	Purity (%)	>99.97%	>99.97%
	Trace components	—	—
Oxygen	Flow rate (kg/h)	25,170	25,170
	Pressure (bara)	11 (stack discharge pressure) 1 (Atmospheric discharge)	11 (stack discharge pressure) 1 (Atmospheric discharge)
	Temperature (°C)	25	25
	Purity (%)	>99	>99
	Trace components	Water	Water

The utility requirements are summarised in Table 5.3.

Table 5.3 Utility requirements for electrolysis

Result	Parameter (Units)	AE	PEM
Cooling water	Flow rate (kg/h)	121,200	241,700
	Pressure (bara)	2.9 – 4.1	2.9 – 4.1
	Temperature (°C)	30 – 65	30 – 65
	Purity (%)	100%	100%
	Trace components	Salts	Salts
Electrolyte (KOH)	Flow rate (kg/h)	0.2	—
	Pressure (bara)	3.9 – 15.1	—
	Temperature (°C)	35 – 80	—
	Purity (%)	11.4%	—
	Trace components	—	—
Efficiency	Stack (kWh/kg)	52.0	57.0
	Balance of plant (kWh/kg)	2.0	2.0
	Total (kWh/kg)	54.0	59.0

5.4 FINANCIAL ANALYSIS

5.4.1 Capital Cost

Equipment costs were estimated for all equipment inside the battery limits and presented in Figure 5-3 below. Costs were estimated using a combination of industry surveys, literature sources, the Aspen Plus database and Aspen Exchanger Design and Rating (EDR). The major equipment items were:

- Electrolyser stack—an industry survey illustrated a large range of pricing and scopes of supply, mostly driven by the package size, stack technology and deliver pressure. In most cases, some ancillary equipment, such as the water treatment, feed pump and gas purification, was included in this stack cost.
- Compressor train—the compressor trains (and inter-stage and post-train cooling) were costed based on industry estimates based on USA supply. The technology basis was reciprocating non-lubricated piston compressors.
- Cooling system—using the pricing database, technology basis was a water cooling tower.

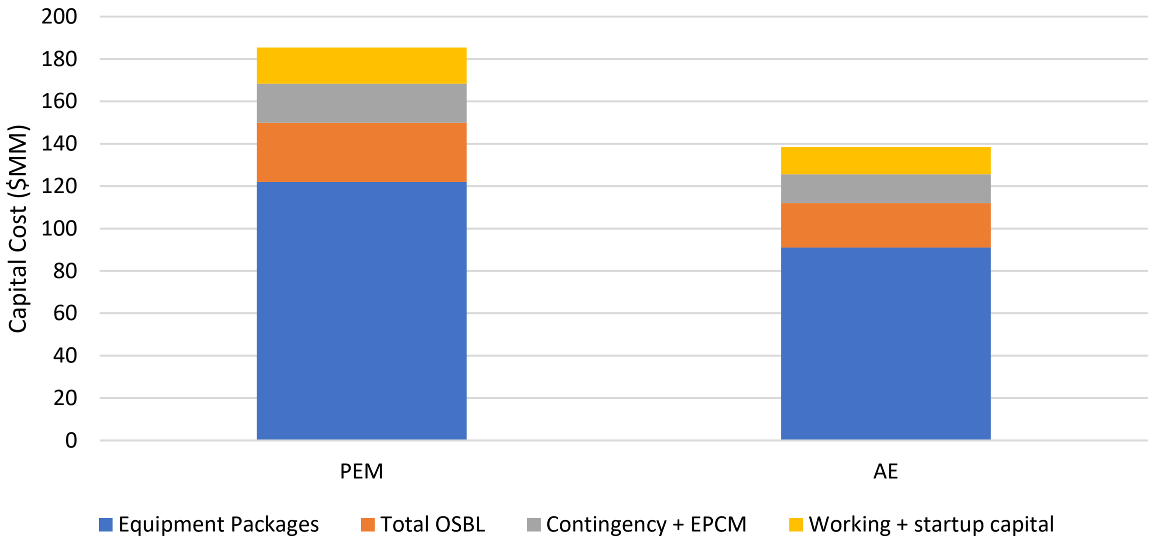

Figure 5-3 Total installed costs for hydrogen electrolysis scenarios

The capital costs for each technology were $185 and $138 million for PEM and AE respectively. Unsurprisingly, the dominating and differentiating cost for the facility were the major equipment packages (largely driven by the electrolyser stack with PEM the most expensive option overall)[21].

5.4.2 Operating Cost

Fixed operational costs include both stack replacement (~$0.48 million / $MW_{installed}$)[22] and administration, labour, maintenance and insurance. Variable operating costs included electricity, water and electrolyte in some cases and are shown in Figure 5-4.

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[21] A more detailed analysis of the impact of electrolyser discharge pressure can be found in the Final Milestone Report for Project RP1.2-02. Of note, LCOH more than doubles for the AE and almost triples for the PEM as the discharge pressure rises to 80 bar, although notably no commercial system has operated beyond 30 bar. A copy of the report can be obtained by contacting the Future Fuels CRC.

[22] These replacement costs were universally applied across both electrolyser scenarios

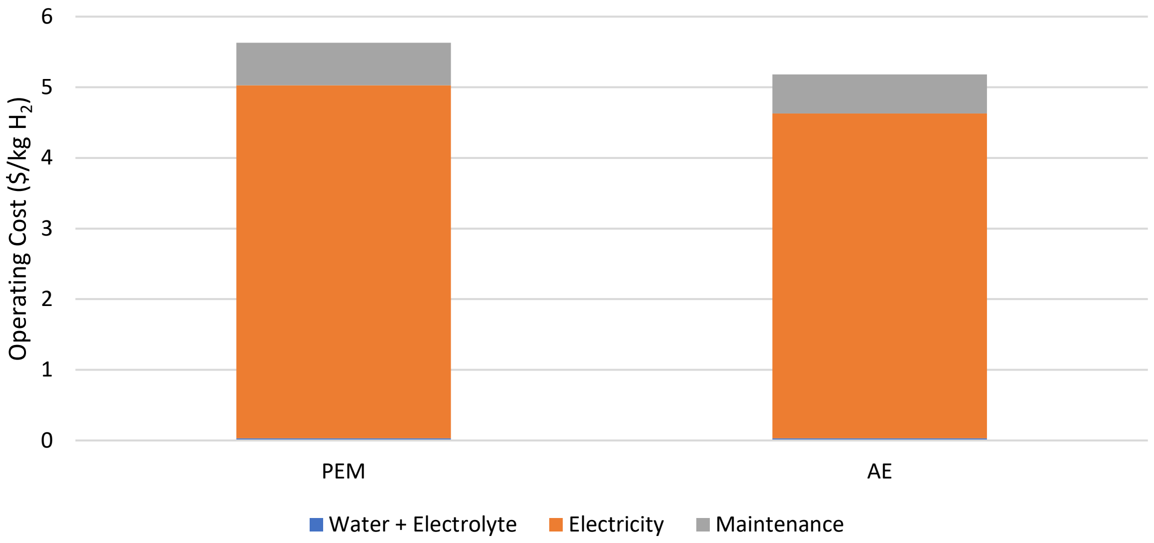

Figure 5-4 Operating costs for hydrogen electrolysis

As expected, the electricity cost is dominant, contributing >85% of the operating cost per kilogram. Due to the differences in stack efficiencies, electricity cost is also the differentiating factor between the two technologies.

5.4.3 Levelized Cost

The levelized cost of hydrogen are shown in Figure 5-5. There was approximately a 10% difference in levelized cost between electrolysers, with an LCOH of $6.99 /kg H_2 for PEM and $6.38 /kg H_2 for AE, respectively. The key drivers of this cost difference, and overall LCOH were:

1. Operating cost—contributing approximately 75% to the LCOH, this cost is dominated by the electricity cost (as illustrated in Figure 5-4).
2. Capital cost—contributing approximately 18% to the LCOH, this cost is dominated by the electrolyser stack and equipment costs (as illustrated in Figure 5-3).

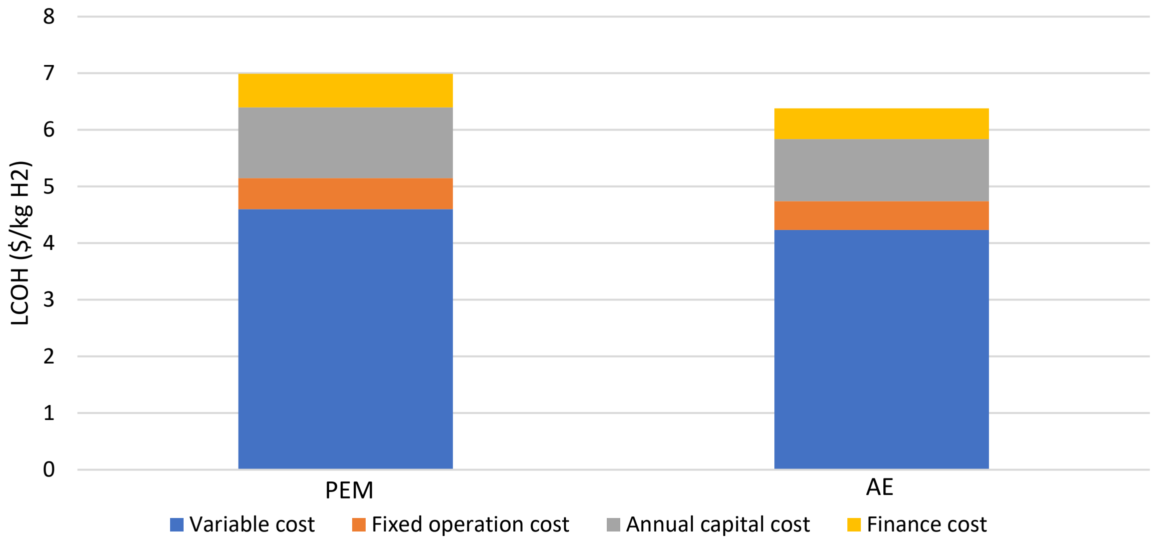

Figure 5-5 Levelized cost of hydrogen for electrolysis scenarios

Compared in this discrete case, the results indicate there is very little difference between the technologies at this stage. However, the design basis and case options fail to highlight some of the underlying drivers of cost, particularly capital cost reductions and electricity supply models. A sensitivity analysis was completed in Section 5.4.4 to understand the impact of these factors on project economics.

5.4.4 Sensitivity Analysis

A detailed sensitivity analysis[23] was conducted to better understand the impact of variations in the dominant cost factors (feedstock cost and capital cost) and to identify targets that may lead to hydrogen production costs below the $2/kg target originally expressed in the Low Emissions Technology Statement[24]. The results are summarised in Figure 5-6 and Figure 5-7. Across both technologies, varying electricity cost and capital expenditure led to the greatest change in LCOH, while changes to water cost had little to no impact and are not shown.

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[23] Full details of this process are given in the Final Milestone Report for RP1.2-02. A copy of this report may be obtained by contacting the Future Fuels CRC.

[24] LOW EMISSIONS TECHNOLOGY STATEMENT 2021 Australian Government Department of Industry, Science, Energy and Resources

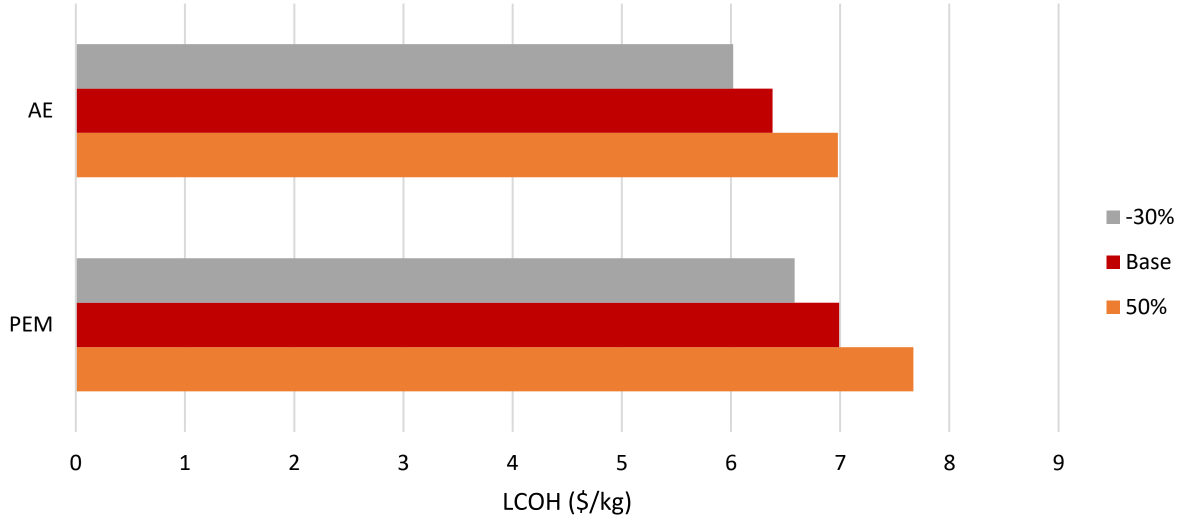

Figure 5-6 Capital cost sensitivity analysis for hydrogen from PEM electrolysis

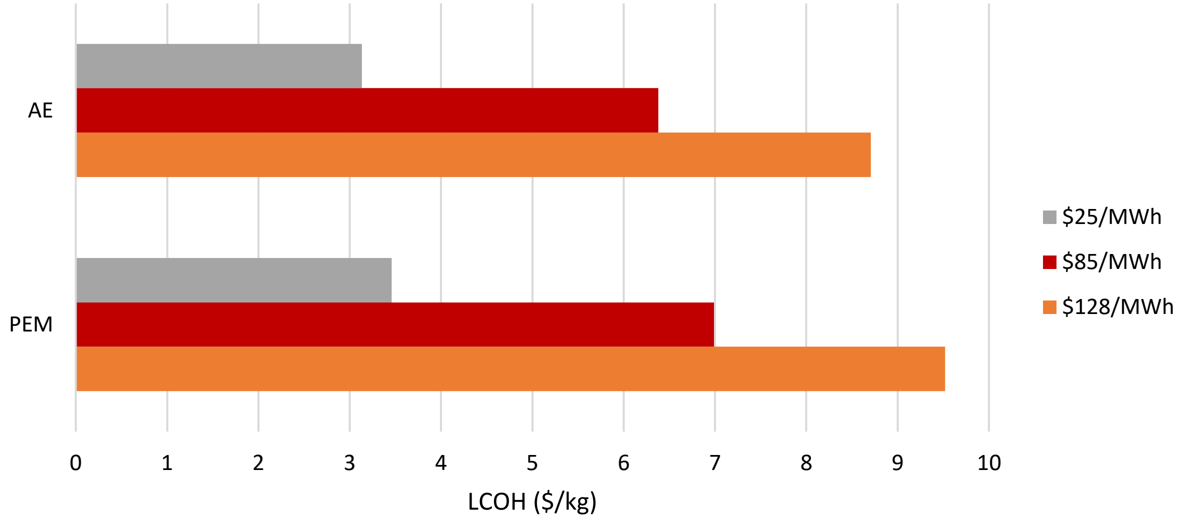

Figure 5-7 Electricity cost sensitivity analysis for hydrogen from AE electrolysis

Neither sensitivity case altered the competitiveness ranking between scenarios. Likewise, neither sensitivity case yielded a LCOH close to the $2/kg H_2 target. However, Figure 5-7 does demonstrate how sensitive the LCOH is to electricity cost and that potential future cost reductions in capital cost, due to technological advancements or learning curves, will struggle to compensate for high electricity prices. To investigate this further, an additional sensitivity analysis was completed to investigate the impact of power supply models.

Renewable energy is required to generate green hydrogen from electrolysis. The type and location of renewable power generation are the key drivers for cost and capacity factor. Four power supply models were used to investigate the impact of these characteristics on (1) electrolyser sizing and (2) project economics, see Table 5.4 for a summary of changed characteristics. It is important to note that unlike the base cases and sensitivity analyses above, these four new cases did NOT maintain the 90% utilisation factor for the electrolyser. In order to provide a level of comparison, the production basis of 25 kta was matched, necessitating an increase in electrolyser size. Additional hydrogen storage (that may be necessary to smooth out H_2 supply into the pipeline network) was not included in the updated costings.

Table 5.4 Summary of power supply models

Parameter	Capacity Factor (%)[25]	Power Price ($/MWh)[26]	PEM Size (MW)[27]	AE Size (MW)[28]
Solar PV	28%	43	601	550
Solar thermal	47%	147	358	328
Onshore wind	42%	45	401	367
Offshore wind	51%	90	330	302

The results of this sensitivity analysis are shown Figure 5-8.

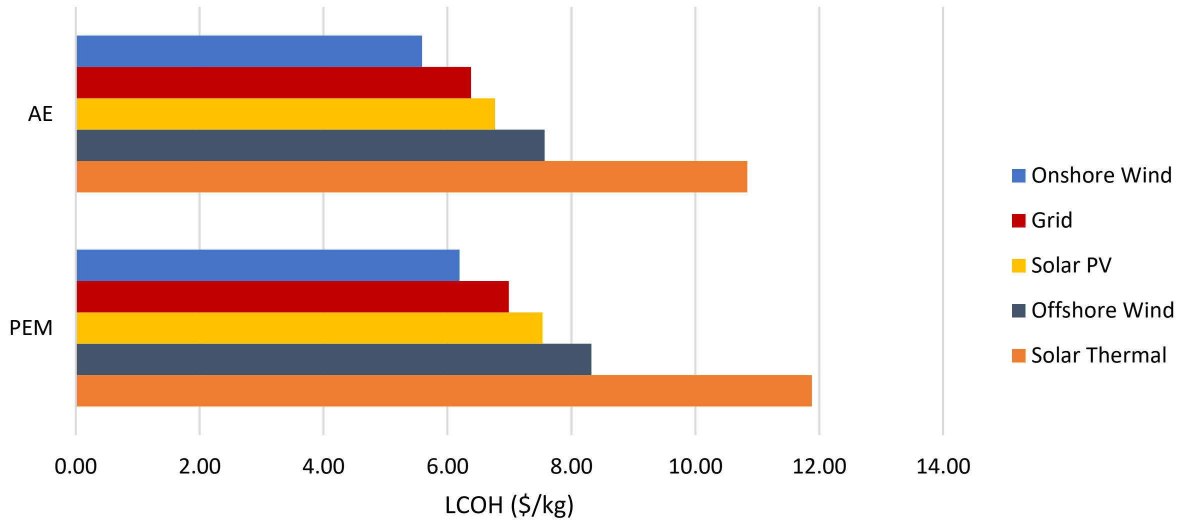

Figure 5-8 Extended sensitivity analysis of electrolysis scenarios

The original basis of the process and techno-economic modelling was a high utilisation (90%) such that the facility had to be grid connected, with a renewable electricity supply agreement in place to guarantee green electricity. However, in practice and at these scales, true renewable electricity may well come behind the meter and this sensitivity — based on nominal

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[25] The capacity factor does not incorporate storage.
[26] The power price given is the LCOE for the RE technology at the capacity factor shown. These power prices were incorporated.
[27] Compared to the original 187 MWe for the PEM base case. The CapEx component was adjusted to account for the change in electrolyser size.
[28] Compared to the original 171 MWe for the Ae base case. The CapEx component was adjusted to account for the change in electrolyser size.

Australian averages for various generators — illustrates how susceptible the LCOH and technical feasibility of this approach is.

6 CARRIERS

6.1 INTRODUCTION

Hydrogen carriers are different physical or chemical forms of hydrogen which can be used to store and transport it in safer or more cost-effective ways. Currently, the most prospective hydrogen carriers include ammonia (NH_3), methanol (CH_3OH) and liquid hydrogen (H_2). Methylcyclohexane ($CH_3C_6H_{11}$) (often simplified to MCH) is receiving more attention as a suitable hydrogen carrier but was not investigated in-depth as it is generally considered the least efficient and has the lowest TRL.

Ammonia has a mature production industry and market in Australia. Ammonia stores the most hydrogen (18wt%) compared to other chemical carriers such as methanol and MCH. Under refrigeration to -33.3 °C, ammonia remains liquid at atmospheric pressure. While existing ammonia plants typically use steam methane reforming to produce the hydrogen, Yara, Dyna Nobel and Queensland Nitrates have all announced plans to produce green hydrogen to make green ammonia. Global infrastructure for transporting ammonia is already in place with millions of tonnes traded globally via shipping.

Methanol has been produced at industrial scale since the 1800s, with mature and established process technology and is comprised of 12wt% hydrogen and is a liquid at atmospheric conditions. Methanol is traditionally produced via syngas pathways, although hydrogenation of a green source of CO_2 allows for a green methanol to be produced. Methanol plants are typically highly integrated with the upstream process of hydrogen production, currently dominated by natural gas and coal feedstocks. Global infrastructure for transporting methanol is already in place, with thousands of liquid chemical tanker ships in operation.

Liquefied hydrogen is a process which is has been achieved traditionally with a helium refrigerant at relatively small scale for scientific or military uses, however, scale-up is for global trade beginning to be considered. Hydrogen can be stored at atmospheric pressure aat a temperature of -253 °C. The largest plants are now producing hydrogen at about 30 tpd, with some proposals for >100 tpd plants in the feasibility phase.

Three scenarios for hydrogen carrier production were investigated. It was important to consider the additionality of the carrier production and as such the report only considers the transformation of a hydrogen (at pipeline specifications) into the relevant carrier. As such the source of the hydrogen (assumed to be green but could equally be blue) is not considered in the process modelling or subsequent economic analysis. All scenarios consider the same hydrogen input rate 25,000 tonnes per year, delivered via pipeline inside the battery limit. As such intermittency or storage requirements at the facility gate for green hydrogen production are not considered. **Green ammonia** considers ammonia produced using the Haber-Bosch process. The air separation unit required to deliver nitrogen for the Haber-Bosch process is considered inside the battery limit. **Green methanol** uses green hydrogen to hydrogenate CO2. Importantly, in the results presented here, the carbon dioxide is delivered at pipeline specifications inside the battery limit. It is assumed that the CO_2 source is either biomass or direct air capture to ensure the carbon neutrality of the methanol. Fossil CO_2 sources must be actively avoided to ensure the methanol does not simply because a means to move CO_2 emissions between producers and consumers. Finally, **Liquid hydrogen** investigates cryogenically cooling hydrogen to its liquid state to increase the volumetric energy density.

All scenarios model the same hydrogen conversion capacity of 25 ktpa (approximately 68 tpd) and the hydrogen cost was not considered the discounted cash flow.

6.2 PROCESS DECSCRIPTION

6.2.1 Green Ammonia

Industrial ammonia is produced using the Haber-Bosch process was developed in 1908. This reaction is dependent on the right mix of catalyst, temperature and pressure. The Haber-Bosch process demands temperatures above 400°C and pressures up to 30 MPa. The minimum theoretical energy required to produce 1kg of ammonia is 5.64 kWh (20.3 MJ) based on the lower heating value of ammonia [6]. This includes the total reaction process from water and air.

Over last century, the Haber-Bosch process has been continuously optimised, reducing the energy input from 60 MJ per kg of ammonia in the mid-1850s to current leading technology of 27.4-31.8 MJ per kg of ammonia [7]. The biggest efficiency gain has been through the replacement of coal feedstocks to methane derived syngas. Other technology developments, such as the introduction of large centrifugal compressors, has led to improved heat integration and allowed for dramatic scale-up of plant capacity. A 300 tonne per day ammonia production facility is considered as a minimum size for a conventional commercial plant.

The context for this report is to model ammonia production as a hydrogen energy vector. The model herein is based on data from Rouwenhorsta et. Al. [7], which considers hydrogen fed from large-scale electrolysis, and an ammonia model developed in Aspen’s ammonia guide particularly around the reactor design. Rouwenhorsta et. Al. (2019) looks at P2A2P where ammonia product and purge streams are also combusted to provide power generation for site use. Our model (Figure 6-1) includes air separation for nitrogen generation and the Haber-Bosch synthesis loop. The refrigeration package to produce liquid ammonia ready for shipping is included in our process modelling and economic analysis.

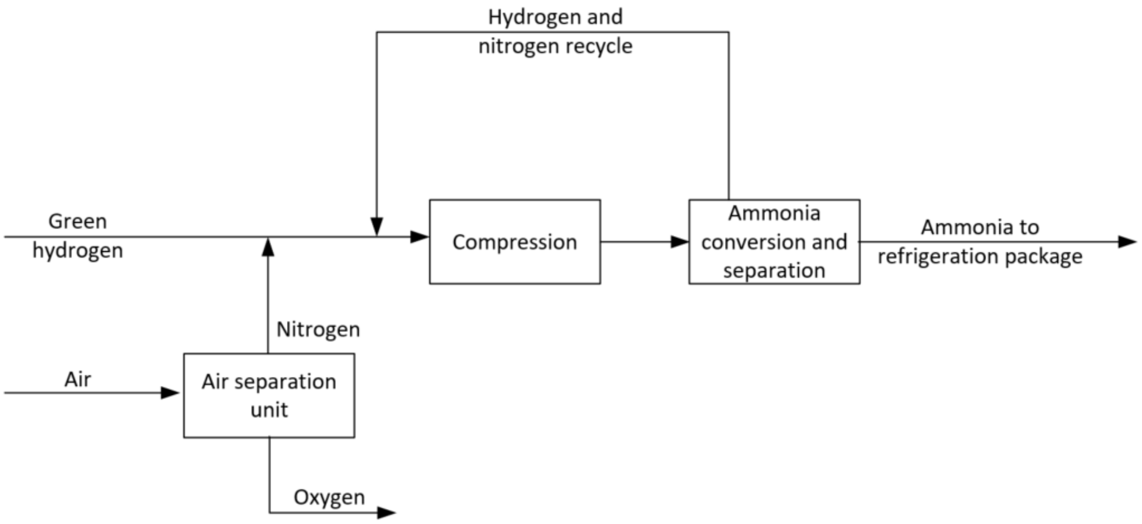

Figure 6-1 Simplified process diagram of green ammonia

6.2.2 Green Methanol

Methanol synthesis requires a carbon feedstock and a catalyst in an exothermic reaction and higher yields are given by lower temperatures and higher pressure conditions of around 80 bar. Currently most methanol plants are fed by natural gas, where hydrogen is produced through Steam Methane Reforming and the syngas by-product is used as the carbon feedstock. However, as methanol produces CO_2 when decomposed, it is important that the production of methanol is associated with green hydrogen (either as electrolysed hydrogen and green CO_2 or syngas from biomass). Green CO_2 can be sourced from biomass (either directly or from syngas from biomass gasification/pyrolysis or biogas) or direct air capture (DAC). The process model (Figure 6-2) does not include provision for CO_2 purification, rather it is assumed that the CO_2 is provided at pipeline specifications and the purchase cost (base case of $40/t$CO_2$) already includes the costs required to process the CO_2 to this purity.

Typically, there is a HP steam requirement for this process in the light end separation and the methanol product column to supply the heat for the two boilers. In this analysis, it has been assumed that these would be powered by renewable electricity, adding 20% in CAPEX to the boilers and with an electrical efficiency of 90%.

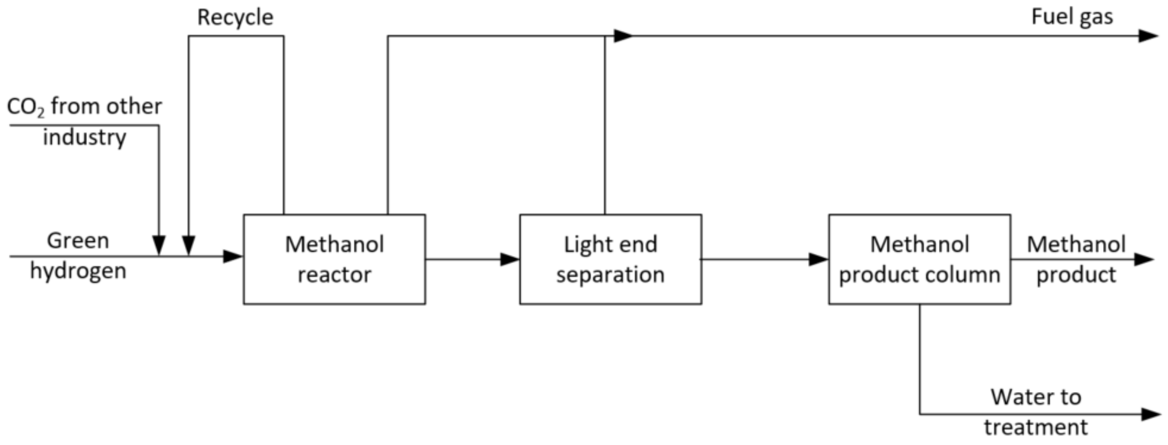

Figure 6-2 Simplified process diagram for green methanol

6.2.3 Liquid Hydrogen

The process (Figure 6-3) utilised a compression-cooling strategy with cooling duty supplied by a helium refrigerant, which in turn is cooled by a nitrogen refrigerant cycle coming from an air separation unit (ASU). This process design is standard for liquid hydrogen generation, though the scale is significantly larger than any comparable existing facilities globally. The process simulation presented here covers all aspects of the liquefaction process as well as helium and nitrogen refrigerant cycles; however, the ASU unit is excluded as the simulation of an ASU process is well known, with only the energy and economics of the ASU incorporated into process evaluation.

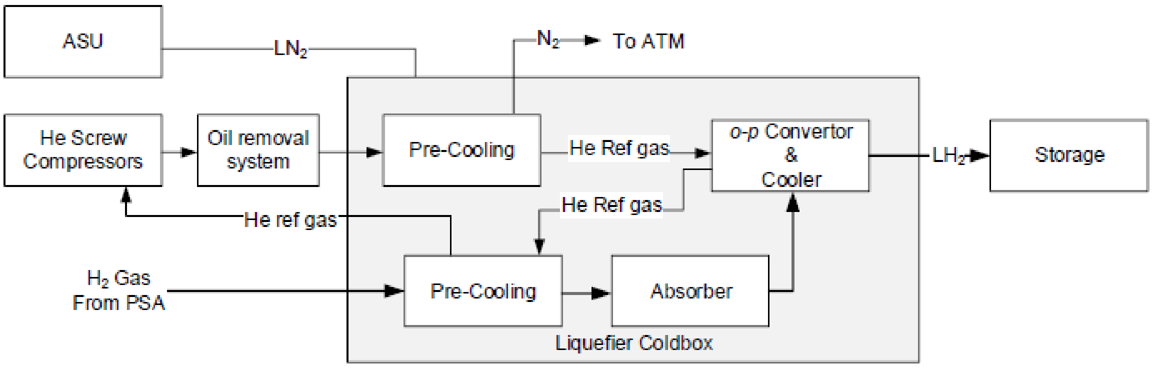

Figure 6-3 Simplified process diagram for hydrogen liquefaction

6.3 PROCESS MODELLING

The process modelling results are presented in Table 6.1. All plants have been sized for a hydrogen input of 25 ktpa and require additional different inputs depending on the technology. Due to the different product uses (i.e. ammonia and methanol can be used as other chemicals rather than carriers / fuels) and energy densities of the 3 carriers, it is useful to compare the energy equivalent of each carrier, showing that liquid hydrogen produces the carrier with the most energy and methanol the least.

Table 6.1 Process modelling results for hydrogen carrier scenarios

Result	**Case**		
	Green Ammonia	**Green Methanol**	**Liquid hydrogen**
H_2 Consumed (kt/y)	25	25	25
CO_2 Consumed (kt/y)	0	185	0
N_2 Consumed (kt/y)	122	0	0
Electricity consumption (MW)	7.7[29]	31.2	43.8
CO_2 emitted (kt/y)	0	4	0
Carrier produced (kt/y)	136	108	25
Energy equivalent (PJ/y)[30]	3.06	2.48	3.54

The utility requirements are broken down further in Table 6.2. The electrical utility requirement for green methanol is high due to the electrification of the boilers in the two product purification columns in the process. If the heat were to be supplied by natural gas, there would be significant emissions associated with the process, which would undermine the carbon neutrality of the carrier. With this electrical requirement, it makes it difficult for green methanol to be competitive from an efficiency standpoint as it requires the second largest amount of electricity, yet produces a product with the lowest energy equivalent.

Table 6.2 Utility requirements for hydrogen carrier production processes

Units	**Green Ammonia**	**Green Methanol**	**Liquid Hydrogen**
Pumps (kW)	210	120	NA
Compressors (kW)	7,500	180	39,600 (He) 2,900 (N_2) 1,270 (H_2)
Air coolers (kW)	0	430	NA
Reboilers (kW)	0	30,500	NA
Cooling water (m^3/h)	1,090	118	NA
Total electricity per tonne of H_2 processed (kWh_e/t H_2)	2,620	10,950	14,880

6.4 FINANCIAL ANALYSIS

6.4.1 Capital Cost

Equipment cost was estimated for all equipment inside battery limit and is presented, along with the full capital cost breakdown in Figure 6-4. The total capital cost of green ammonia is

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[29] Includes ASU and ammonia refrigeration package

[30] Based on HHV for each carrier

almost double that of green methanol. This is due to the substantially higher pressure requirements for the reactor (275 bara vs 80 bara for ammonia and methanol respectively) which increases material cost and compressor size. Ammonia also requires an air separation unit (ASU), accounting for about one quarter of the total capital cost to produce nitrogen from the air. A refrigeration unit is also required to store the ammonia at atmospheric pressure, and which is not required for methanol. The direct capital cost for LH2 construction is the second highest, due to the significant number of compressor trains required for both the helium and hydrogen cooling circuits.

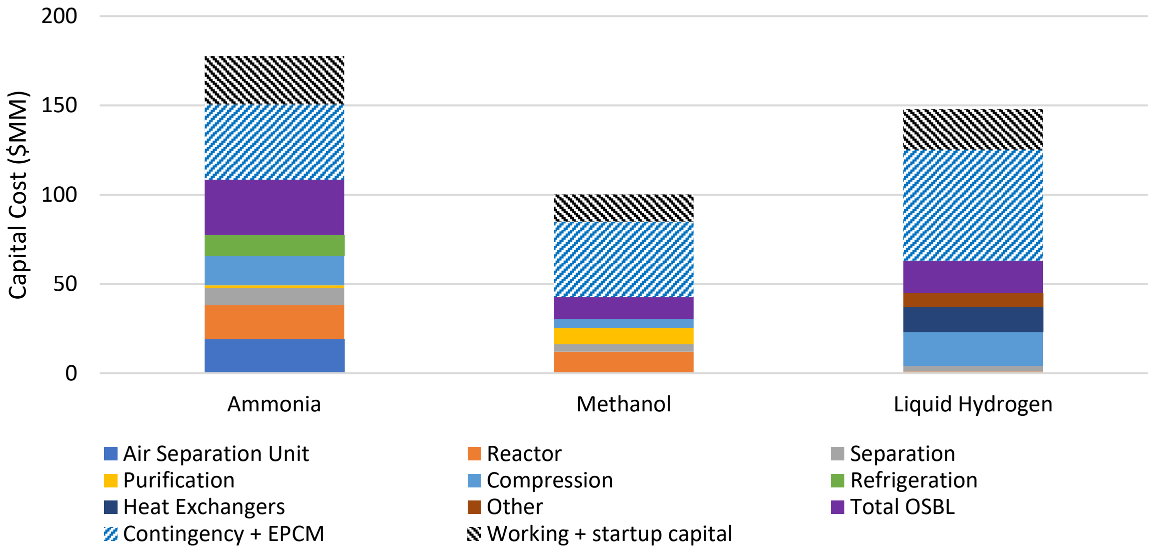

Figure 6-4 Capital cost breakdown for the hydrogen carriers

Across the three process options, compressors represent one of the most significant equipment cost categories. Improvement in compressor efficiency is expected to be a major influence on costs for commercial equipment packages. This is especially prudent given that hydrogen and helium-based compressors are less efficient than conventional air compressors. For all conversion-based processes, the reactor proportion of the capital cost is unlikely to reduce given the requirements of the conversion reactions and limitation in possible advances in this area, unless it is the development of novel catalysts, or process integration with the hydrogen generation process which haven’t yet been well developed when considered green hydrogen [8].

6.4.2 Operating Cost

The variable operating costs (i.e. utility costs) for the different carriers are presented in Figure 6-5. The relatively low electricity use in the ammonia conversion process helps this become more competitive over the life of the plant to help offset its high capital cost. Electricity makes up most of the green methanol and all the liquid hydrogen utility cost pathways.

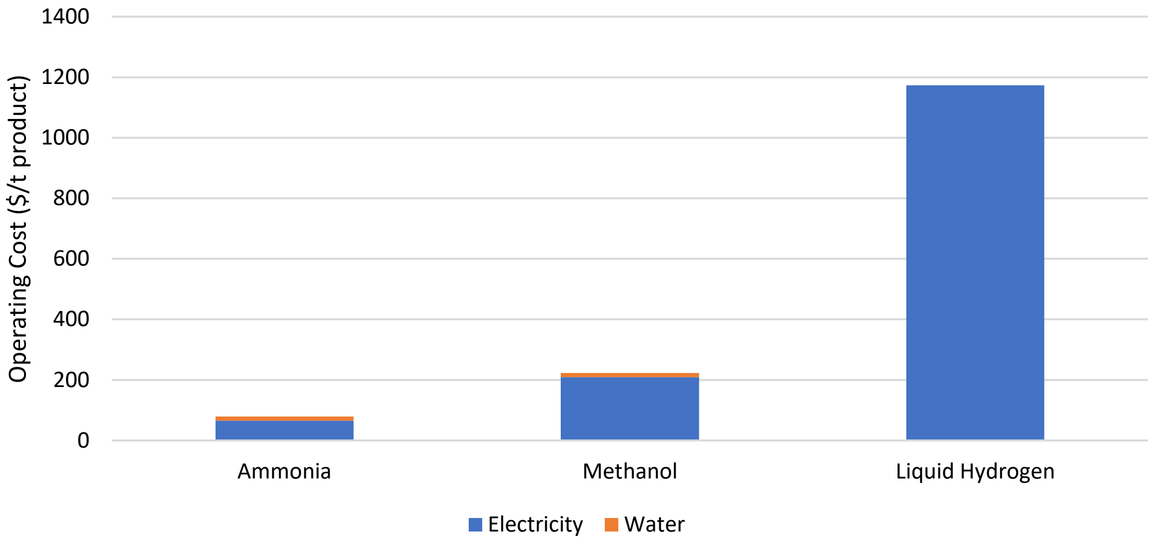

Figure 6-5 Utility cost breakdown for the hydrogen carriers

Additionally, fixed operational costs were calculated which included labour cost, maintenance tax and insurance. The latter three were at rates of 3%, 2% and 0.5% of the total production cost respectively while the labour cost was estimated as a function of the number of process units.

6.4.3 Levelized Cost

A summary of the levelized cost per tonne of product for ammonia, methanol and liquefied hydrogen is given in Table 6.3. It is also useful to consider the levelized cost on a per gigajoule of product basis, given the dramatic differences in mass density between the 3 carriers. And finally as the levelized cost only incorporated the cost of conversion and not the gate price of hydrogen, we report the levelized cost per tonne of H_2 feed. Green ammonia is the lowest cost carrier on an energy and per tonne of product or H_2 feed basis. Green methanol is a lower cost carrier than LH2 on a per tonne of product basis, but not on an energy basis. On a per tonne of H_2 feed basis it is marginally lower cost than LH2.

Table 6.3 Summary of levelized cost of hydrogen carriers

	Green ammonia	**Green methanol**	**Liquified hydrogen**
Levelized cost ($/t product)	285	476	2272
Levelized cost ($/GJ product[31])	12.7	20.7	16
Levelized cost ($/t$H_2$ feed)	1,550	2,056	2272
Global market price ($/t)[32]	500	400	N/A

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[31] HHV basis

[32] Indicative of 2020 pricing

6.4.4 Sensitivity Analysis

To identify major cost drivers, a sensitivity analysis was performed where the above levelized costs were used as base cases and varied in response to changes in capital cost and electricity price (the two most significant drivers of cost). Hydrogen cost is not considered in this analysis.

A sensitivity analysis for capital cost is shown in Figure 6-6 and is most significant for ammonia and liquid hydrogen, reflecting their higher capital cost. A sensitivity analysis for electricity price is shown in Figure 6-7. Green methanol is also much more sensitive to the impact of the electricity price due to the high electrical heating requirements (incorporated to avoid using natural gas fired heating for the methanol purification train). Liquid hydrogen is also sensitive to electricity price, which is expected given it is such a large component of the levelized cost (~30%). The variation in price for electricity variation is between $4.2 and $5.85 / GJ LH2, which indicates that LH2 process remains somewhat competitive as a storage option for hydrogen even at high electricity prices.

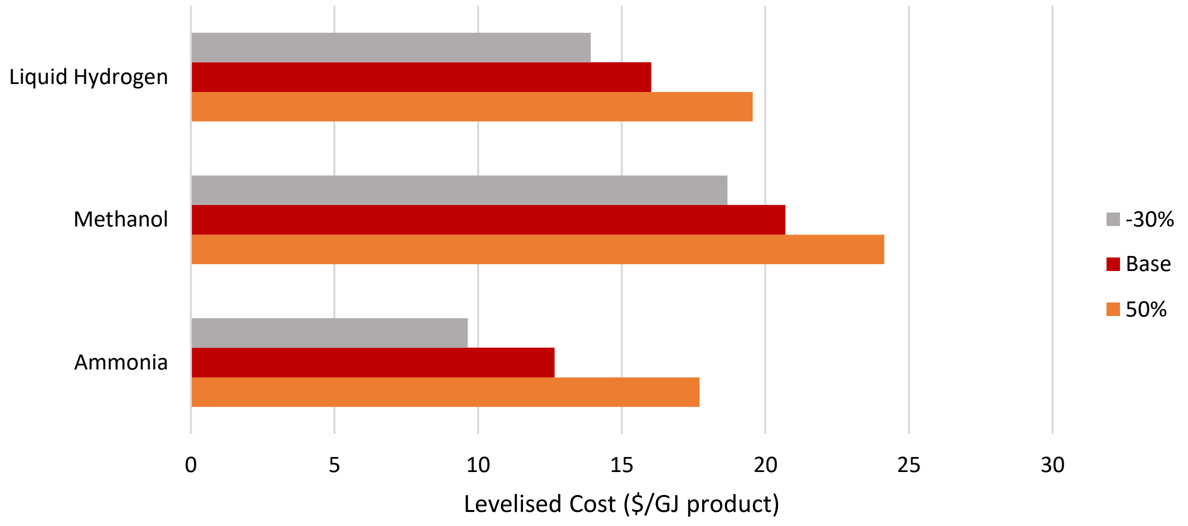

Figure 6-6 Capital cost sensitivity analysis for hydrogen carriers

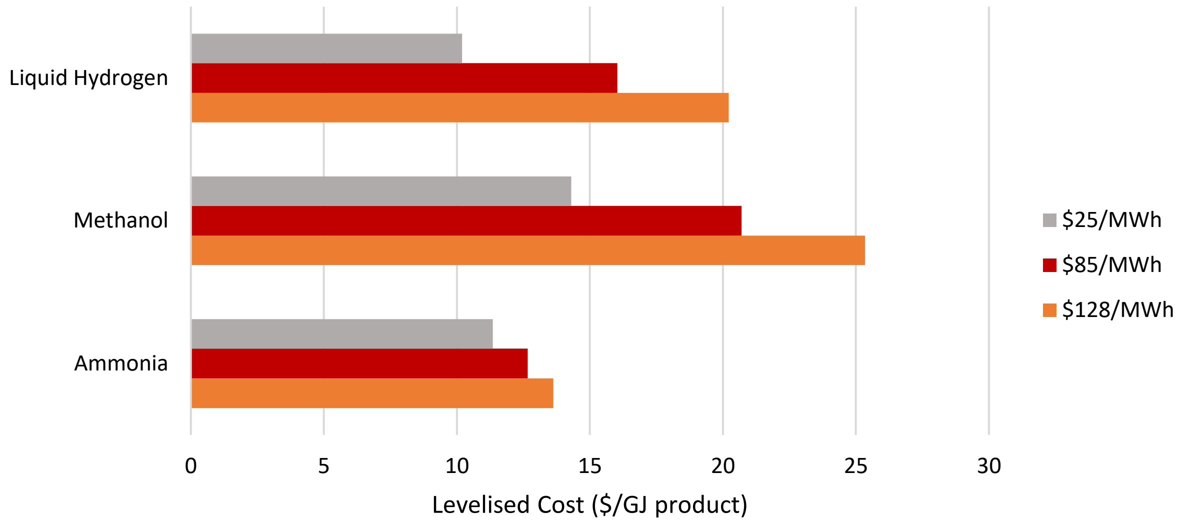

Figure 6-7 Electricity price sensitivity analysis for hydrogen carriers

Additional analyses were performed on the liquid hydrogen process which indicated that both the helium and liquid N_2 requirements were too minor to alter the utility costs, while the cooling water prices were negligible against the other utilities. In addition, a more than tripling of the CO2 purchase (from $40/tCO$_2$ to $150/tCO$_2$) increases the cost of methanol by ~30%.

This report highlights green ammonia as the most cost-effective carrier form, although it is possible that this would change with end-use application and transportation requirements. Green ammonia has significant advantages, but the conventional Haber-Bosch process based on a renewable hydrogen source presents challenges in terms of scaling and optimisation, given the long history of the process. This implies that cost savings will be difficult to establish. However, there are new electrochemical processes for generating green ammonia under development that represent a revolution in the ammonia production industry, and these have significant advantages to the conventional approach. For liquid hydrogen, the largest impact research and development will have is in the improvement to hydrogen compressors, as the low efficiencies represents significant cost wastage. There are significant research endeavours in this area and there is expectation that more efficient designs will be commercialised in the coming decade.

Sourcing quality carbon feedstocks is the main limitation to develop large-scale methanol production. Direct air capture (DAC) is also scaling for industrial use with several large plants in development. DAC stores CO_2 at ambient temperatures and pressures of 45 – 65 bar or in refrigerated, insulated tankers at temperatures of -35 to -15ºC and pressures of 12 – 25 bar. Carbon Engineering is developing a 1MtCO_2 per year capture plant in the US with plans for 2023 operation. This would equate to production of 600,000 tonnes of methanol per year (440 GWh of fuel energy).

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8 APPENDIX A: FEEDSTOCK SPECIFICATIONS

8.1 NATURAL GAS

The natural gas composition has been specified through communication with industry partners. Three scenarios have been defined using available data: normal, maximum CO_2 with higher heavier hydrocarbon and maximum methane content. Table 8.1 gives the composition of available natural gas at the plant battery limit. The coal seam gas scenario (CSG) is assumed to be provided from an existing plant with a higher methane concentration.

Table 8.1 Natural gas specifications and condition

	Pipeline Australia	**Pipeline Maximum CO_2**	**Pipeline CSG (Kogan North)**	**Australian Gas Pipeline AS4564-2011**
Methane content	Normal*		High	
Composition mol%				
Methane	91.04	91.32	98.68	
Ethane	3.53	4.49	0.06	
Propane	0.65	0.68	0.0	
Butane-i	0.06	0.07	0.0	
Butane-n	0.08	0.09	0.0	
Pentane-n	0.0	0.02	0.0	
Pentane-i	0.01	0.03	0.0	
C6+	0.0	0.03	0.0	
Nitrogen	3.88	0.75	0.94	Inert Max. 7 mol %
CO_2	0.75	2.52	0.32	CO_2 Max. 3 mol %
Total	100	100	100	
Oxygen content				0.2 mol%
Water Content (mg/m^3)				Max.73 (at 15000kPa)
Hydrogen Sulfide (mg/m^3)				Max. 5.7
Mercaptan sulphur (mg/m^3)				Max.5.0
Total Sulphur (mg/m^3) (Include Odorant)				Max.50
High Heating Value (MJ/m^3)	37.5	38.45	37.3	Max:42.3
Wobbe index MJ/m^3	48.31	49.03	49.78	Min. 46- Max. 52
Temperature ºC	25	25	25	Min. 10- Max. 50
Pressure	8.0MPag	8.0MPag	8.0MPag	
Hydrocarbon Dew Point ºC				2 ºC at 3500 kPag

The reported value is at 15 ºC and 1 atm
* No LPG blending

8.2 BROWN COAL

Within Gippsland, the Latrobe Valley has an estimated measured resource of close to 65 billion tonnes, equivalent to ~25% of the world's known brown coal reserves. Victoria's brown coal (VBC) is typically low in ash, sulphur, heavy metals and nitrogen. However, its high moisture content - which ranges from 48-70% reduces its effective energy content (average

8.6 MJ/kg on a net wet basis or 26.6 MJ/kg on a gross dry basis). Typical characteristics and composition of VBC are summarised in Table 8.2 [1].

Table 8.2 Typical characteristics of Victorian brown coal

Coal characteristic	
Energy Value (Net wet)	5.8-11.5 MJ/kg
Energy Value (Gross dry)	25-29 MJ/kg
Water	48-70%
Carbon	65-70%
Oxygen	23-30%
Hydrogen	4-5.5%
Ash	<4%
Nitrogen	<1%
Sulfur	<1%

The Loy Yang and Yallourn mines produce the majority of brown coal in the Latrobe Valley. For simulation purposes, the project will use Yallourn brown coal was used as an input to the model [2] and the plant location for the gasification processes examined is the Latrobe Valley in Victoria. The composition and Particle size distribution (PSD) shown in Table 8.3 and Table 8.4 respectively.

Table 8.3 Ultimate, Proximate and Sulfur Analyses of Victorian brown coal

Ultimate Analysis %		Proximate Analysis %		Sulfur Analysis %	
Ash	0.91	Moisture	65	Pyritic	0.3
Carbon	65.4	FC*	16.4	Sulfate	0
Hydrogen	4.4	VM*	17.7	Organic	0
Nitrogen	0.6	ASH	2.61		
Chlorine	0				
Sulfur	0.3				
Oxygen	29.3				

*FC: Fixed Carbon, VM: Volatile Matter

Table 8.4 Wet Victorian brown coal particle size distribution

Interval	Lower limit (μm)	Upper limit (μm)	Weight fraction	Cumulative weight fraction
1	0	106	0.085	0.085
2	106	300	0.228	0.313
3	300	600	0.202	0.515
4	600	1000	0.02	0.535
5	1000	4000	0.167	0.702
6	4000	8000	0.137	0.839
7	8000	20000	0.161	1

8.3 BLACK COAL

The majority of the black coal mined in Australia belongs to the bituminous category, and the majority of Australia's black coal Economic Demonstrated Resources (EDR) is located in Queensland (61%) and New South Wales (36%). Of this, 34% and 29% of recoverable EDR are located in the Bowen (Queensland) and Sydney (New South Wales) basins, respectively [3].

For this project Queensland black coal from the Bowen basin was chosen as the feedstock due to the proximity to potential CO_2 storage reservoirs [4]. The ultimate, proximate and sulfur analyses of a typical Queensland black coal is presented in Table 8.5 [5]. The particle size distribution of a typical black coal is also presented in Table 8.6 [6].

Table 8.5 Ultimate, Proximate and Sulfur Analyses of Queensland Black Coal

Ultimate Analysis %		Proximate Analysis %		Sulfur Analysis %	
Ash	22.82	Moisture	5.75	Pyritic	3.94
Carbon	68.07	FC*	55.29	Sulfate	0
Hydrogen	3.66	VM*	16.14	Organic	0
Nitrogen	0.24	ASH	22.82		
Chlorine	0				
Sulfur	3.94				
Oxygen	1.27				

*FC: Fixed Carbon, VM: Volatile Matter

Table 8.6 Particle Size Distribution of a typical black coal

Interval	Lower limit (μm)	Upper limit (μm)	Weight fraction	Cumulative weight fraction
1	0	20	0.113	0.113
2	20	40	0.042	0.155
3	40	60	0.06	0.215
4	60	80	0.097	0.312
5	80	100	0.146	0.458
6	100	120	0.108	0.566
7	120	140	0.052	0.618
8	140	160	0.046	0.664
9	160	180	0.059	0.723
10	180	200	0.278	1

8.4 BIOMASS SPECIFICATION

Biomass feedstock can be collected from diverse sources and exhibits variable chemical composition, which has an influence on their specific properties and applicability [7]. The choice of plant species depends upon the end-use and conversion option of interest. Some plant species such as rapeseed can be processed through almost all of the potential conversion technologies while others such as wood and cereal crops are suitable for combustion, gasification, pyrolysis and fermentation [8]. Aquatic plants and manures are intrinsically high-moisture materials and more suited to wet processing techniques such as anaerobic digestion. High-moisture content biomass are normally more suited to wet/aqueous conversion process such as fermentation while a dry biomass such as woodchips is more economically suited to gasification, pyrolysis or combustion. Usually aqueous processing is used when the energy required for drying would be large compared to the energy content of the product formed [7]. However, there are other factors that must be taken into account in determining the selection of the conversion process. Some of these factors are including the ash, alkali and inorganic contents and trace elements, differences and irregularity in biomass, the low bulk density and high oxygen content. These factors may impact thermal conversion adversely [9].

Numerous crops have been proposed or being tested for commercial energy farming which includes woody crops and grasses/herbaceous plants, starch and sugar crops and oilseeds. The main important characteristics of the ideal energy crops are:

- High yield (maximum production of dry matter per hectare)
- Low energy input to produce
- Low cost
- Least contaminants in the composition
- Low nutrient requirements

Different types of biomass feedstock have been tested to evaluate their hydrogen production potential. The main issues to consider in the selection of biomass as a feedstock for hydrogen production are the spectrum of cultivars available in each country and the growth time, the demands for each cultivar, harvesting, transport and pre-treatment cost. Agricultural residues, peanut shell, post-consumer wastes such as plastics, trap grease, mixed biomass and synthetic polymers and rapeseed have been widely tested for hydrogen production through pyrolysis and gasification. In order to solve the problem of decreasing reforming performance caused by char and coke deposition on the surface of catalyst fluidised catalysts beds are usually used [10]. However, the main current feedstock for gasification is lignocellulosic biomass, namely wood and residues from forestry and agricultural activities [11]. Around 80% of the commercial and operating biomass gasification plants run on woody biomass mostly wood chips [12].

8.4.1 Potential Biomass Feedstock for bioenergy production in Australia

The potential to increase the energy produced from biomass relies on different factors, including the feedstock and the resources available. While the resources are abundant in Australia, but are currently underutilised. The biomass resources come in the form of specifically grown crops, or by-products generated in agriculture such as slurries and manures or from industrial applications such as paper, wood, furniture manufacturing and municipal solid waste. However, the majority of bioenergy comes from the combustion of sugarcane bagasse [13]. Bioenergy resource sector in Australia is summarised in Table 8.7 [14].

The biomass conversion routes can determine whether or not a project is commercially viable and the costs for these conversion processes are often very site and project specific. They vary with the source of raw biomass, its moisture content, the transport distance, the complexity of the process involved, the plant scale, the value of any co-products, the savings of disposal cost if a waste, the reduction in greenhouse gas emissions, the market value for the bioenergy, and whether there are subsidies and incentives available. Careful analysis and risk assessment are therefore required to get a good overview of what is involved and the chance of commercial success for each project [15, 16]. Over time it is expected that bioenergy project costs will reduce as industry knowledge increases with regard to feed materials, technical alternatives for processing, and operating characteristics.

Table 8.7 Bioenergy resources sector in Australia

Resource sector	Definition
Agricultural-related wastes	Agricultural-related wastes are a very diverse resource including resources such as crop and food residues and livestock wastes
Energy crops	Energy crops refer to short rotation crops that are coppiced or other crops grown for the purpose of bioenergy production as well as woody weeds.
Landfill gas	Landfill gas captures the methane emitted from landfills which is produced from mainly municipal solid wastes and industrial wastes, to generate bioenergy.
Sewage gas	Sewage gas captures the methane emitted from the solid organic components of sewage collected by water utilities to produce bioenergy.
Sugarcane	The fibre of processed sugarcane, known as "bagasse", contributes to sugar mill electricity exports. The trash, tops and leaves resulting from the harvesting can potentially be used to add to electricity generation.
Urban Biomass (including urban timber wastes)	Urban biomass consists of food-related wastes, garden organics, paper and cardboard material, and also includes urban timber (i.e. from construction and demolition timber).
Wood related wastes	This resource includes wastes produced in the harvesting and processing of wood such as sawmill and pulp-mill residues. (Additional potential is in urban waste such as urban timber.)

Table 8.8 shows a summary of Australian biomass resources, projected to both 2020 and 2050 from the Clean Energy Council's Bioenergy Roadmap [17]. Although Table 8.8, took a conservative approach in prediction of future biomass resources, it shows there are considerable potential for generating heat and power from stubble-grain and cotton crops, forest-based industry wastes and bagasse [18-22].

Table 8.8 Biomass resources and bioenergy production potential in Australia

Biomass Source	Quantity (Mt/y) – Year 1999-2004	Quantity (Mt/y) – Year 2010	2010 (GWh/y)	2020 (GWh/y)	2050 (GWh/y)
Poultry		94	-	297	1055
Cattle - Feedlot		0.87	-	112	442
Pigs		1.8	1	22	205
Dairy Cows		1.4	-	22	89
Abattoirs		1.3		337	1773
Stubble- grain and cotton crops		24			47000
Bagasse	10.6	5	1200	3000	4600
Sugar cane trash, tops and leaves	9.25	4	-	165	3200
Oil mallee Eucalypts		-	-	112	484
Camphor laurel				83	20

Forest residues (native forests, plantations, processing residues)	23	9	79	2442	4554
Black liquor	0.25		285	365	365
Wood waste	6.9				
Crop wastes	44				
Other pulp and paper wastes			74	141	141
Urban food wastes		2.9	29	267	754
Garden Organics		2.3	29	121	461
Urban Paper and cardboards		2.3		38	1749
Urban timber/wood waste		1.6	45	295	1366
MSW **C&I** **C&D**	8.7 10.3 13.5				
Landfill gas			772	1880	3420
Sewage Gas			57	901	929
Energy crops **Cereals** **Other**	 30 8				

Based on the literature review, the potential biomass feedstock for studied technologies and the potential location of the plant is summarised in Table 8.9. The ultimate and proximate analyses of the potential biomass feedstocks as an input to the model is presented in Table 8.10 [23, 24].

Table 8.9 Potential biomass feedstock, available amount, delivered cost and potential location and conversion technologies

Biomass group	Biomass type	Potential Location	Available amount (Mt/y)	Biomass delivered Price	Potential technology
Agricultural wastes	Sugar cane bagasse	North Queensland/ North NSW	10.6	$0 - 0.20/GJ	Gasification pyrolysis
Agricultural wastes	Crop stubbles	South Australia, western Victoria and central NSW	27.7	$4.80 - 8.00/GJ	Gasification pyrolysis
Forest residue **Wood waste**	Pulpwood, woodchips and sawlog residues	Green Triangle and Murray Valley	23	$2.00 – 3.20/GJ	Gasification pyrolysis
Wastes	Co-digestion of food wastes with sewage sludge	Major cities	2.9	-	Co-digestion Anaerobic digestion

Table 8.10 Proximate and ultimate analyses of typical biomass feedstocks used as an input to the model

Biomass Group	Typical Samples	Proximate Analysis (wt%)				Ultimate Analysis (wt%) db				
		M	VM	ASH	FC	C	H	N	S	O
Agricultural wastes	Wheat straw	4.38	68.52	12.91	14.2	40.36	5.95	0.55	0.27	52.87
Agricultural wastes	Sugar Cane Bagasse	46.0	41.77	6.69	5.54	45.45	5.26	0.23	0.05	36.65
Pulpwood, woodchips and sawlog residues	Sawmill wood residue	22.2	59.75	1.01	17.04	49.7	5.9	0.15	0.03	42.9
Wastes	Food waste	81.5	17.21	1.02	0.28	49.7	5.9	1.86	-	39.3
Wastes	Sewage Sludge	8.07	48.90	39.61	3.42	28.4	5.29	4.65	2.66	25.58

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