Low Carbon Hydrogen Supply 2 Stream 1 Phase 1

BEIS Purchase Order No. 415000049193

Optimised Hydrogen Liquefaction Feasibility Study

1.0 Abbreviations / Definitions

AACE	American Association of Cost Engineers
ALPEMA	Aluminium Plate-Fin Heat Exchanger Manufacturers' Association
BAHX	Brazed Aluminium Heat Exchanger
BOG	Boil Off Gas
C1	Methane
CAPEX	Capital Expenditure
DOE	Department of Energy (USA)
EPC	Engineering, Procurement and Construction
H&MB	Heat and Material Balance
HP	High Pressure
IP	Intellectual Property
ISO	International Organization for Standardization
J-T	Joule-Thomson
KOH	Potassium Hydroxide
LH2	Liquid Hydrogen
LK	Lee Kesler
LNG	Liquefied Natural Gas
LP	Low Pressure
MBWR	Modified Benedict Webb Rubin
MEL	Master Equipment List
MP	Medium Pressure
NIST	National Institute of Standards and Technology
OHL	Optimised Hydrogen Liquefaction
OPEX	Operating Expenditure
PDP	Process Design Package
PFD	Process Flow Diagram
PFHE	Plate Fin Heat Exchanger
PR	Peng Robinson
PSA	Pressure Swing Adsorption
RefProp	Reference [Fluid] Properties
te/d	Tonnes per day
tpa	Tonnes per annum
TIC	Total Installed Cost
UA	Product of overall heat transfer coefficient (U) and Area (A)
VIP	Vacuum Insulated Pipe

2.0 Executive Summary

Optimised Hydrogen Liquefaction (OHL) is a high efficiency, large scale bulk hydrogen liquefaction process and was developed by Gasconsult Limited in response to 2050 net zero carbon emissions targets. Hydrogen liquefaction can be an enabler for many hydrogen-based net zero schemes as the reduction of gas volumes by a factor of 800 by liquefaction will:

- lower regional hydrogen distribution costs
- facilitate large-scale intercontinental energy transfer, comparable to current LNG practice
- enable long term energy storage for use with intermittently available renewable power

Hydrogen liquefiers currently in operation typically have capacities of 5 to 30 te/d with a design that minimizes capital cost at the expense of energy efficiency. They are unsuited to plants in the capacity range of 100 to 500 te/d which will be required to make a meaningful reduction in the current 4 billion tpa global consumption of liquid fossil fuels. Gasconsult's patent pending OHL design seeks to to realise economic liquefaction by a) reducing current power demand (typically 10 to 15 kWh/kg LH2) to around 7 kWh/kg LH2 and b) to develop a design an order of magnitude higher than current practice to meet market size requirements. Without these developments the use of hydrogen will not meet energy transition requirements.

Gasconsult engaged McDermott to conduct an independent technical verification of the OHL process covering both the process design and the availability of equipment for plants an order of magnitude higher in capacity than current practice; and to develop a capital cost estimate for the scheme.

McDermott's early work identified modifications to the initial concept to mitigate potential issues. These were incorporated into the design as part of a Base Case 300 te/d LH2 design. The Base Case was used to solicit technical proposals and cost estimates of the main equipment items from suppliers. Based on this a detailed Process Design Package was completed together with a cost estimate for the liquefier.

The verification process demonstrated that OHL can achieve a low specific energy of 6.8 – 7.1 kWh/kg LH2, whilst meeting all design criteria including LH2 product specifications. This calculated specific power is 25 to 40% lower than currently operating technologies. No significant step-outs in equipment duties were identified for the Base Case scheme.

The cost estimate for the liquefier is $500 million, providing potentially 20% savings relative to current technologies[1]

The work additionally covered development of a smaller scale bridging capacity of 50 te/d. This indicated no significant reduction in performance relative to the Base Case.

The OHL process can be modified to use nitrogen precooling in place of methane. An initial assessment indicates that a specific power of ~7 kWh/kg LH2 can be achieved with the nitrogen configuration; comparable to the Base Case.

3.0 Basis of Design

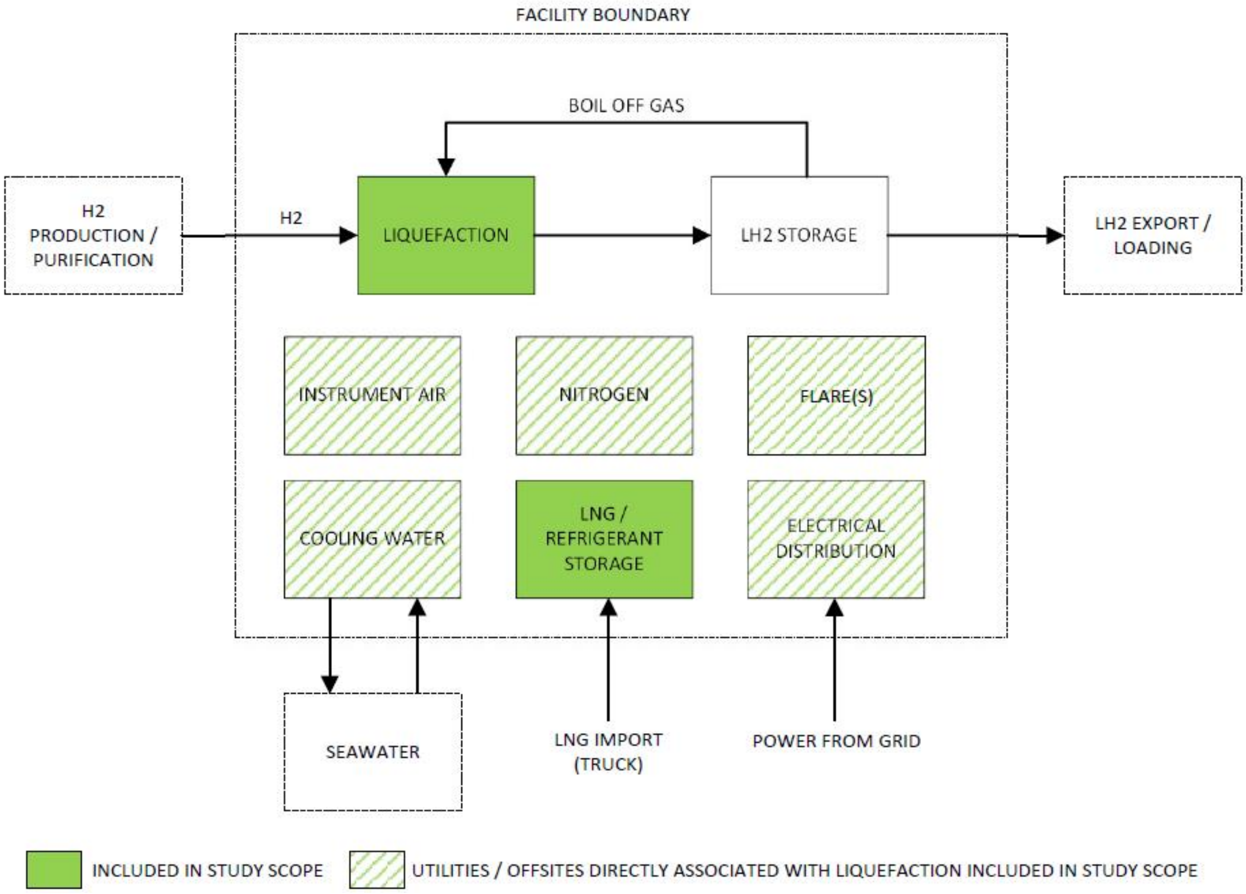
3.1 Background and Scope

The plant would be located in NW Europe with a capacity of 300 te/d LH2 and based on Gasconsult's OHL process operating in methane cycle pre-cooling mode. Construction would be based on Asian built pre-fabricated modules. The plant is indirectly sea water cooled via a circulating closed cooling water circuit.

The hydrogen feed to the facility would be either blue or green hydrogen produced upstream of the liquefaction unit. The hydrogen production and pre-treatment facilities required by the cryogenic unit and to meet LH2 product specifications are outside the scope of the study.

An outline of the overall facility boundaries in the hydrogen supply chain is shown in Figure 1. The study scope for verification of the OHL technology and preparation of the cost estimate covers the liquefaction block, any utilities directly associated with liquefaction, and the required refrigerant loading, storage and transfer.

Figure 1 – Hydrogen Liquefaction Facility and Study Scope Boundaries



3.2 Basis of Design Summary

Parameter	Unit	Value
H2 Feed Flowrate	te/d	300
H2 Feed Composition	Mol%	>99.99
H2 Feed Pressure	kPa absolute	2500
H2 Feed Temperature	ºC	30
Sea Water Temperature	ºC	11
LNG refrigerant storage	m^3	2 x 100

3.3 Product Specification

Liquefaction itself does not impact the hydrogen product specification. A product specification in line with ISO 14687 was selected (for Type II liquid fuels, Grades C and D) with a minimum 95% para hydrogen content to minimise boil off from ortho to para hydrogen in LH2 product storage.

3.4 Refrigerant Import

The Gasconsult OHL process uses a methane precooling refrigerant circuit which will require an external methane supply.

For the purpose of this study, LNG was selected as the methane refrigerant source as it is available within the UK, can easily be transported and is already of a purity suitable for use within a cryogenic system.

The OHL process can be designed with a nitrogen precooling refrigerant circuit as an alternative to methane. For the nitrogen alternative LNG storage would not be required.

4.0 Process Package Verification

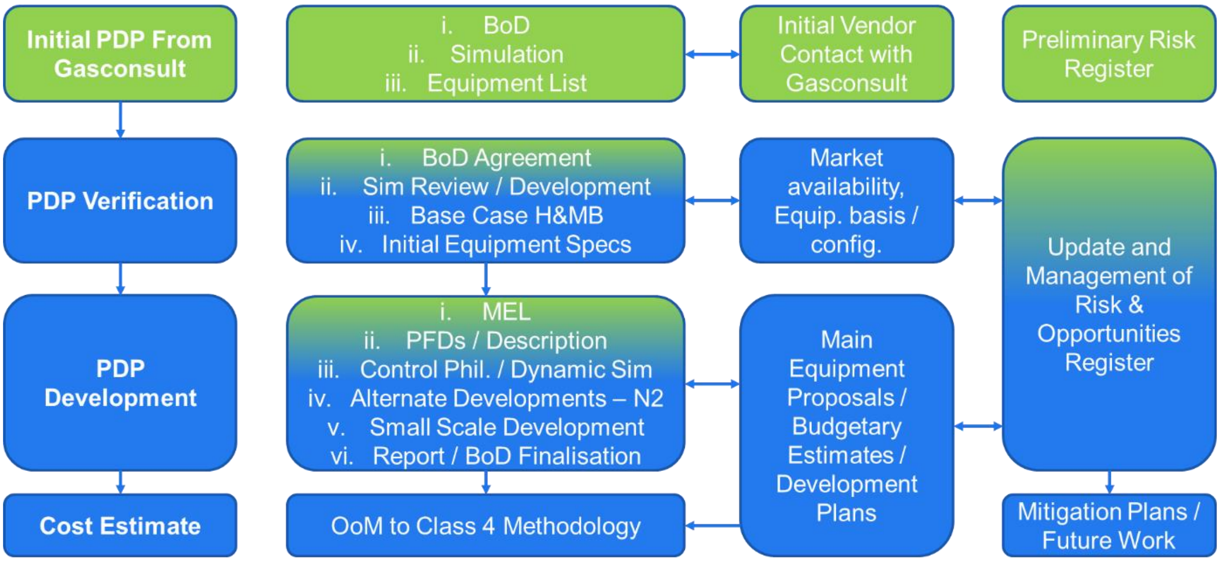
4.1 Background

Gasconsult provided McDermott a basic PDP as the starting point for developing the Base Case hydrogen liquefaction process (300 te/d). The PDP consisted of:

- Preliminary Basis of Design
- Process simulation performed using Aspentech HYSYS software.
- Preliminary Equipment Specifications
- Preliminary listing of equipment
- Initial risk register.

The work process for verification of the design package and development of the process design is shown in Figure 2.

Figure 2 - Design Package Verification and Development Work Process



4.2 Verification

McDermott first completed the verification of the initial process design concept provided by Gasconsult. The verification process began with an initial flow scheme review to support setting of the design basis and to identify items for further investigation.

The verification went on to review the simulation platform and property package selection, comparing these to available data from literature. McDermott evaluated and updated the simulation model considering the practicality of the pressure and temperature profile, equipment availability and performance, the ortho-to-para hydrogen conversion with inclusion of the heat of reaction, and achievable refrigerant compositions. McDermott engaged with multiple suppliers to confirm the availability, performance, configuration options, level of maturity of the required equipment and identification of key risks. As part of this initial review and supplier engagement McDermott, in collaboration with Gasconsult, identified modifications to the initial concept to mitigate potential issues. Some key highlights of the verification are captured below.

4.2.1 Simulation Platform and Property Package Selection

McDermott evaluated Gasconsult's heat and material balance using v12 of Aspentech HYSYS software package and updated the equations of state as follows:

- For the precooling section of the liquefaction, the Peng Robinson equation of state has been used with the Lee Kesler enthalpies rather than the standard equation of state package. This has been selected based on McDermott's experience with simulating the cryogenic sections of LNG facilities and is considered to provide a better, though slightly more conservative prediction of the methane cycle.
- The hydrogen liquefaction section has been assessed using the RefProp model developed by NIST. This is widely reported in literature to be the most accurate model to use for pure components studied by NIST; and the operating conditions of the hydrogen liquefier are within the temperature and

pressure ranges covered by this property package. This has the added advantage that the implementation of the RefProp equation of state within HYSYS can be cross checked against open literature data available from NIST.

4.2.2 Pressure Profile

The initial PDP included pressure drop allowances for equipment items and piping losses. In addition to verifying these allowances, McDermott has also identified areas where additional pressure drop may be required to enable a more practical facility design. The evaluation considered McDermott's EPC contracting experience from LNG and was refined following completion of preliminary hydraulic calculations.

4.2.2.1 Methane Recycle Compressor Suction Pressure

The initial PDP had the Methane Recycle Compressor low pressure stage suction at sub-atmospheric pressure. Allowing this pressure to be sub-atmospheric potentially enables the ingress of air into a closed loop refrigeration cycle where, over time, it could accumulate to levels that cause a significant safety concern. Consequently, the suction pressure of this compressor was limited to a minimum of 1.05 bara to maintain positive pressure in the circuit and allow some margin for process control.

4.2.2.2 Hydrogen Feed Gas Pressure

A sensitivity analysis was performed to determine the impact of reducing the feed gas pressure on the process specific power and the cold box exchanger UAs. The sensitivity analysis was performed over a hydrogen feed gas pressure range from 90 bara to 60 bara.

In terms of specific power, the assessment found that a reduction in feed gas pressure resulted in a specific power increase of approximately 3% (i.e. 1% per 10 bar decrease) with the increase following a linear trend across the pressure range evaluated.

The exchanger UA was predicted to increase by 0.7% for every 10 bar decrease in feed gas pressure (again following a linear trend) over the feed gas pressure range.

Following these sensitivity assessments, and in order to mitigate the risk around the ortho to para hydrogen performance at high pressure (above currently proven operation conditions), the feed gas pressure was reduced to 25 bara. This made the scheme consistent with available cryogenic exchanger references in hydrogen service.

4.2.2.3 Other

Other minor changes to the original Gasconsult design basis included:

- Methane refrigerant circuit composition to align with agreed basis
- Revised temperature profile arising from agreed atmospheric conditions
- Precooling and liquefaction heat exchanger approach temperatures
- Rotating equipment efficiencies to align with supplier information
- Split of ortho-para heat of reaction across exchangers based on catalyst conversion information

- Switch to letdown of hydrogen rundown pressure from a hydraulic turbine to simple J-T valve to mitigate development risk

5.0 Mechanical Equipment

This section summarises key discussion points around the main equipment items in respect of their performance and level of maturity under OHL process conditions.

5.1 Compressors

5.1.1 Overview

The Gasconsult OHL process has four main sets of compressors:

a) The Hydrogen Refrigerant Compressors which circulate the hydrogen refrigerant and provide the pressure required to generate the cooling.

b) The Hydrogen Feed Gas Recycle Compressors which re-compress low pressure hydrogen vapours from the feed circuit (resulting from the final LP flash) and return these to the main hydrogen feed supply.

c) The Methane Refrigerant Compressors which circulate the methane refrigerant and provide the pressure required for the expanders to generate cooling.

d) The Methane Recycle Compressors which return the low pressure methane refrigerant at the cold end of the cold box to the main methane refrigerant compressor suction.

Current hydrogen liquefaction plant refrigerant cycles typically use either oil injected rotary screw compressors or reciprocating compressors for duties (a) and (b) depending on the selected cycle, service and plant capacity. These types of machines are usually specified due to the compression duties fitting into standard commercially available low capacity size ranges with suitable turndown flexibility.

Oil injected rotary screw compressors have a relatively low efficiency and require oil removal / recovery facilities downstream. It is therefore unlikely that they will be a suitable selection for large scale commercial hydrogen liquefaction facilities.

Initial market analysis indicated reciprocating compressors were likely to reach volumetric flow limits and hence require parallel machine configurations to meet the anticipated duty requirements for a large scale facility, especially in hydrogen refrigerant compression service. Reciprocating compressors can maintain higher efficiencies at higher capacities and are less impacted by low molecular weight gases. However they have shorter operational periods between maintenance due to the required overhaul of the piston rider rings, piston rings and packing.

Centrifugal or axial compressors may offer a better fit in terms of potential compressor size range (volumetric) and efficiency while providing a higher reliability / overall availability. For a scale-up in capacity, it is likely that a single compressor train can meet the duty; however, a string arrangement may be preferable to improve overall availability and provide additional operational flexibility. A move to large centrifugal compressors for the hydrogen refrigerant service may eliminate the necessity to use parallel machine configurations (although this may be preferable for reasons of plant availability) but will present challenges on a technical level due to the low molecular weight of the refrigerant and resulting requirement to have a large number of compressor casings in series to achieve the required head. Hydrogen compression intrinsically features limited compression ratios and potentially onerous sealing requirements.

The main methane refrigerant compressor (c) duty is very similar to the mixed refrigerant or methane refrigerant duty of operational LNG facilities. The duty, performance, configuration, control and operation is well established with many similar references. Similarly the methane recycle compressor (d) with the lower cryogenic suction temperatures is also very well proven in numerous LNG facilities across end flash gas and BOG compressor applications. The methane cycle machines are thus assessed as not presenting a technical risk.

5.1.2 Hydrogen Refrigerant Compressors

Hydrogen refrigerant compression was a particular concern at the start of the verification. However, early engagement with the suppliers indicated that there was considerably more experience with hydrogen rich fluids (>95mol%) than first assumed.

Recent modernisation of Kuwait National Petroleum Company's refineries has seen the application of multiple parallel strings of the largest reciprocating compressors (order of 16 MW) in the world, demonstrating the feasibility of reciprocating machines at approximately the required capacity for 300 te/d LH2.

Additionally the main suppliers of centrifugal machines have already set out development programs to address the key issue in hydrogen service, namely the low compression ratio. Although not yet available, 1st and 2nd generation developments are in place to increase the current proven compression ratios from around 1.1-1.3 up to 2, and then possibly even 3. This is achieved by increasing the impeller tip speeds, with resultant developments required to the impeller design and casings.

Offers were ultimately received for both reciprocating and centrifugal machines for the 300 te/d LH2 duty. The reciprocating compressors required a high number in parallel however they are more efficient than centrifugal compressors, consuming less power for the given duty (~60MW versus ~70MW). They have less stages requiring intercooling, resulting in a lower pressure drop, and are cost competitive on an equipment supply basis. The large number of compressors in parallel however would probably require additional pressure drop to balance the flows, but could offer improved turndown while maintaining overall compression efficiency. The main areas of concern would be in the swept volumes required at the low pressure end and the resultant footprint. This may not be conducive to a modularised design.

The centrifugal machines are less proven and have a lower efficiency than the reciprocating machines, however they do offer a number of solutions that may be attractive for given a given project. There is a wide range of configurations with definite opportunity for improvement in both cost and footprint through supplier's development programs. These improvements may be realised by the time a large scale hydrogen liquefaction project is commercialised.

Initial observations from the review and supplier feedback indicate that 300 te/d may be close to a capacity breakpoint between reciprocating and centrifugal machines for hydrogen refrigerant compression.

5.1.3 Hydrogen Feed Gas Recycle Compressors

The hydrogen feed gas recycle compressors have similar areas of concern as the main refrigerant compressors, however with a smaller duty. The overall power requirements were about 20 MW; about a third of that required for the main hydrogen refrigerant duty (but with 2/3rds of the suction volume flow due to the lower suction pressure). Suppliers offered options for reciprocating and centrifugal machines, or hybrids thereof. These machines were similar to those proposed for the main Hydrogen Refrigerant Compressors, but with smaller casings / fewer parallel strings.

5.2 Expander-Compressors

The base case Gasconsult OHL process has four main expander duties across the two refrigerant cycles serving the precooler and liquefier:

- Precooling
  - Warm and Cold C1 Expander-Compressor
- Liquefaction
  - Warm and Cold H2 Expander

The methane service warm and cold duties are well proven in numerous applications (NGL Recovery for example), with a number of suppliers able to provide these machines. They do not represent a technical risk.

5.2.1 Hydrogen Refrigerant Service

The optimum speed for the warm and cold hydrogen refrigerant expanders is high due to the low fluid molecular weight and flow / head ratio; and results in a large overall enthalpy drop. As such all suppliers stated that to achieve the target performance multiple machines in series were required to manage the enthalpy drop and speed. All suppliers advised that the warm expander duty required 3 to 4 stages, and the cold expander duty required 2 stages, to be operated in series, with an isentropic efficiency of >80% and a speed of ~30,000 rpm. This was irrespective of whether specifying an expander-generator or expander compressor arrangement.

The machines were well within the available frame sizes of the suppliers, although the overall experience in hydrogen service was not extensive.

Suppliers do not see significant technical risk in moving to hydrogen as materials are understood, and current seal / bearing designs can be applied. Additionally control of a number of expanders in series is proven and has been applied in a number of services. Other than the suppliers approached there are other potential suppliers working in hydrogen who can also offer proven solutions, although at lower capacities.

Performance risk is one of the key areas that needs consideration. The performance is achievable at the design point, however may drop off significantly with change in the flow / turndown. This needs to be reviewed by dynamic simulation to understand achievable production across the design range.

In terms of scalability increasing the capacity should help improve performance as the flow increases for a given head; whereas a drop off in flow for a smaller capacity would potentially result in poorer performance while still requiring the same number of machines in series.

5.3 Hydraulic Turbine

Gasconsult's original PDP utilised hydraulic turbines for final refrigeration and letdown of the hydrogen product rundown. This allowed improved refrigeration relative to a J-T valve arrangement due to energy recovery by the turbine.

To evaluate the availability of the equipment in this service a number of hydraulic turbine suppliers were approached. Feedback indicated there are no market references for an LH2 hydraulic turbine, and in most cases suppliers did not have plans for development of such a product.

5.4 Cold Box Design

The precooling and the liquefaction exchanger designs, the insulation concept and ortho-para conversion, are important factors impacting cost effective, robust and efficient scale-up.

Current applications use 2 or 3 separate PFHEs / BAHXs located in cold box(es) with a number of Licensors / Equipment Suppliers having operational references in hydrogen liquefaction service for a range of capacities from ~5 to ~30 te/d. The applications involve:

1. Precooling of the hydrogen to 80 to 110K
2. Final cooling and liquefaction to 20 to 25K

The precooling exchanger can be accommodated in a standard cold box with perlite or equivalent insulation. However liquefaction cold boxes currently utilise vacuum insulation to minimise heat in-leakage and are subject to dimensional constraints. The precooler duty is not an onerous cryogenic application and has temperature ranges in which significant operational experience exists. The precooling exchanger is therefore not expected to give rise to risk or scale up issues.

5.4.1 Cold End (Liquefier) Exchanger

The PFHE or BAHX design and manifolding is well understood and materials are designed for the operating and design conditions associated with liquefaction, with design covered by clear ALPEMA guidelines. As such, scale-up could be managed simply by increasing the number of cores in parallel. However, physical limitations on the cold box size due to vacuum insulation, or transportation, becomes a factor.

The challenge for scale-up is how best to approach the design and insulation concept of the liquefaction cold box given the limitations on vacuum cold box dimensions and the increase in size of the exchanger arising from increased plant capacities. Further, allowing also for associated equipment / piping that needs to be included within the cold environment.

Licensors / suppliers are developing scale-up solutions and some consider this as IP, making access to detail information difficult. However budgetary estimates of the exchangers / cold boxes have been secured to support the order of magnitude estimate. The proposals however do not specify how the ortho para catalyst is to be incorporated in the exchanger.

Further work is required on the inclusion of catalyst, the modification of the enthalpy curve, ensuring the volume of catalyst achieves the required conversion, and how this will need to be specified on a datasheet to allow the exchanger supplier to provide meaningful guarantees. This in itself is not a barrier to scale up as the risk can be mitigated by design conservatism / external catalysts vessels. However a test program to secure more definitive data is recommended.

5.5 Cooling Water Exchangers

Due to the potential number of compression stages and resultant requirements for intercooling and aftercooling the design of these exchangers can have a noticeable impact on the specific power of the process. The two key design parameters are temperature approach and allowable pressure drop. Water cooled exchangers are proven in hydrogen service and the selected basis for the pressure drop and approach temperature are consistent with standard shell and tube designs available from a wide range of suppliers. These exchangers therefore present no technical risk.

5.6 Other Items

A high level market availability check was carried on vacuum insulated piping and valving in liquid hydrogen service. The main findings are:

- There is proven experience and supply for piping and valving up to 8in, driven by the current demand.
- VIP piping fabrication and experience for LNG service is available for upwards of 20 - 24in, in some cases 30in from limited suppliers (process piping).
- The current limitations for maximum size of VIP in liquid hydrogen service is in the order of 16in.
- Vacuum insulated valving available up to approximately 12in with current designs.

Initial sizing indicates that the liquid / supercritical hydrogen lines outside the cold box are expected to be within current proven references with the maximum hydrogen rundown line estimated at 8in.

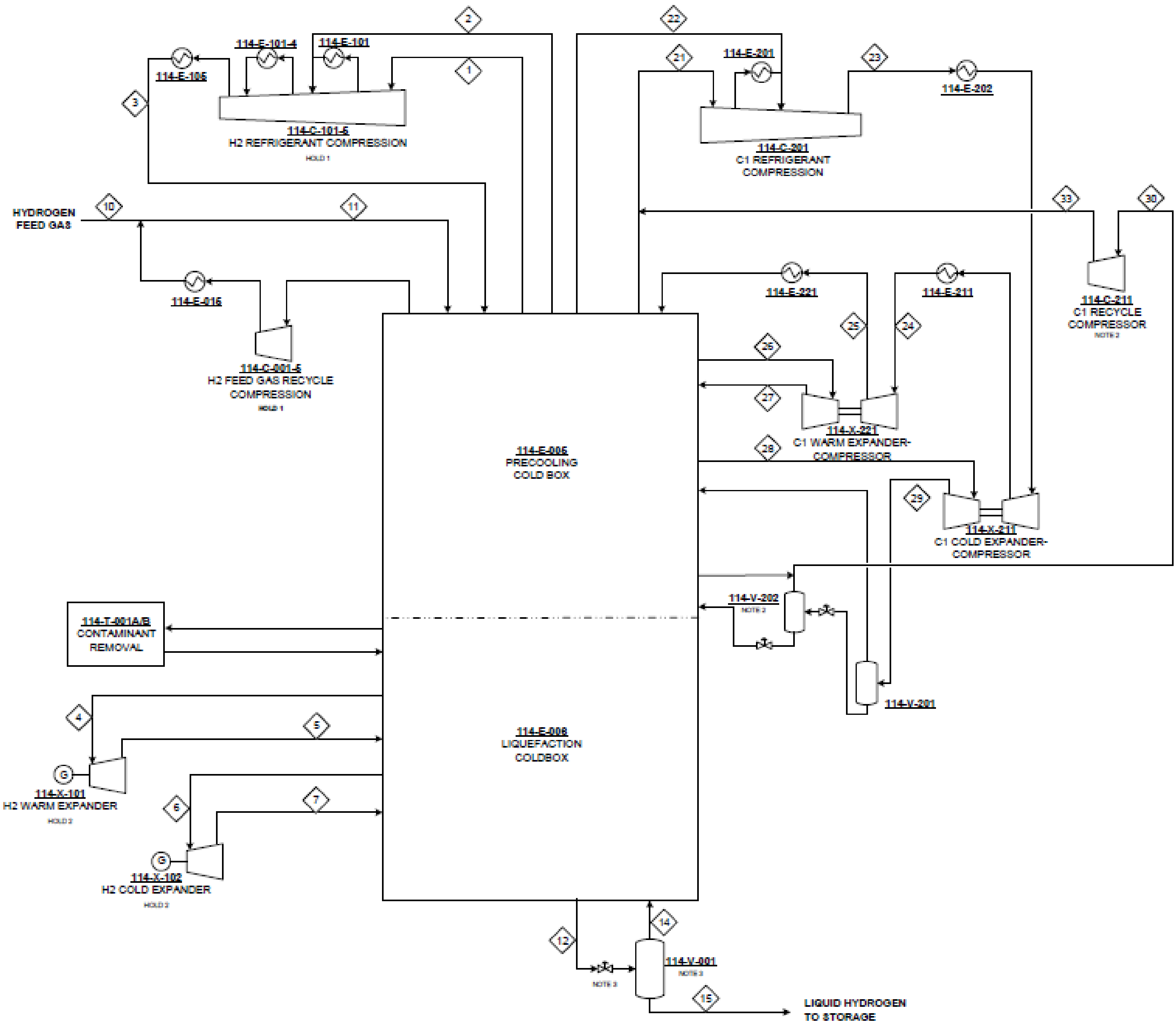
6.0 Process Design Development

The initial flowscheme developed by Gasconsult was critically reviewed by McDermott. The following issues were identified impacting the process configuration:

- Source of methane pre-cooling refrigerant was modified to incorporate use and storage of replacement inventory in the form of LNG
- The level of risk for inclusion of a hydraulic turbine was considered too great and it was replaced with a J-T valve, with recycle of the flashed hydrogen vapour to the feed gas (refer 5.3)
- The hydrogen feed gas pressure was reduced to 25 bara to bring the scheme into existing experience levels for catalyst containing cryogenic hydrogen exchangers

The flowscheme arising from these adjustments plus input from equipment suppliers was used as the basis of process design development and is shown in Figure 3. This, under the process conditions detailed in the Basis of Design (3.0) formed the basis of final process simulations to establish the overall heat and material balance and equipment sizing. Final equipment sizing allowed suppliers to provide updated technical proposals and pricing for development of a capital cost estimate.

Figure 3 – Base Case Overall Configuration



6.1 Process Description

As indicated in Figure 3 the process comprises two refrigeration circuits for pre-cooling the hydrogen feed and subsequent liquefaction of the precooled hydrogen. A further circuit combines incoming feedgas with recycled hydrogen refrigerant prior to processing the combined feed in the refrigeration circuits

6.1.1 Hydrogen Feed Gas Circuit

Hydrogen feed gas is supplied to the liquefaction unit at 25 bara and 19°C from the upstream pre-treatment units and is considered suitable for liquefaction. The specific pre-treatment units will vary depending on the type of hydrogen production (electrolysis or methane reforming) and are outside the scope of this study. The hydrogen feed gas is essentially pure hydrogen.

The hydrogen feed gas is combined with the recycled hydrogen refrigerant from the outlet of the H2 Feed Gas Recycle Compression (114-C-001-005) before it is routed to the Precooling Cold Box (114-E-005) where it is cooled to -150ºC using a methane refrigerant circuit. Some of the cooling is also provided by the returning hydrogen refrigerant vapours.

From the precooling cold box, the hydrogen feed gas stream is routed to the contaminant removal section where any trace contaminants (mainly hydrocarbons and nitrogen) are removed to (i) protect the ortho-para catalyst (ii) prevent freeze out in the liquefaction section and (iii) achieve the required hydrogen liquid specification. Removal is by cryogenic adsorption in the H2 Contaminant Adsorption Beds (114-T-001A/B). This is a regenerative process with multiple beds in parallel to allow for online regeneration with a slipstream of warmed feed gas.

The precooled, purified hydrogen feed gas is routed into the Liquefaction Cold box (114-E-006) where it is further cooled and contacted against the ortho to para hydrogen conversion catalyst. The enthalpy rejected from cooling of the gas and the heat of reaction is transferred into the hydrogen refrigerant stream. The final cooling step is achieved by flashing the liquid hydrogen across a J-T valve prior to entering, at reduced pressure, the LP H2 Rundown Flash Drum (114-V-001). The flash gas is routed back to be combined with the feed gas stream via the H2 Feed Gas Recycle Compressor (114-C-001 to 005) and the liquid hydrogen is run down to storage.

6.1.2 Methane Precooling Circuit

The OHL process utilises a dual-expander methane precooling cycle to provide the precooling duty for hydrogen liquefaction. The loop contains small amounts of nitrogen, ethane, propane, and butanes as is sourced, and made up, from LNG.

The methane cycle operates as a closed circuit with the methane refrigerant compressed and circulated by electrically driven centrifugal compressors. The Main C1 Refrigerant Compressor (114-C-201/202) compresses the LP methane and MP Methane, from the Warm C1 Expander (114-X-221) and the Cold C1 Expander (114-X-211) outlets respectively. The heat of compression is removed by the Main C1 Refrigerant Compressor Inter and After Coolers (114-E-201 & 202 respectively).

The methane stream is then routed through, first the C1 Cold Expander-Compressor (114-X-211) and then the C1 Warm Expander-Compressor (114-X-221) compressor stages which are direct coupled and driven by their associated expander. Each compressor includes an aftercooler (114-E-211 & 221 respectively) which removes the heat of compression. The compressed methane refrigerant then enters the warm end of the pre-cooling cold box.

A slipstream of cooled methane refrigerant is withdrawn from the cold box and sent to the C1 Warm Expander (114-X-221) where it is let down in pressure and then routed back to the precooling cold box to provide the warm end cooling. The Stream is then routed to the MP Suction of the Main C1 Refrigerant Compressor (114-C-201).

The flow split to each expander is controlled based on a flow ratio control with the hydrogen feed gas.

The remainder of the methane refrigerant is further cooled in the cold box and routed to the C1 Cold Expander (114-X-211) where pressure is let down, resulting in partial liquefaction of the stream with the vapour / liquids separated in the MP C1 Flash Drum (114-V-201). The cold vapours are routed back to the cold box and the liquid stream is then sequentially flashed to lower pressure levels with separation in LP C1 Flash Drum (114-V-202). The remaining liquid phases are routed back to the cold box to provide the refrigeration for the cold end of the precooling cold box. The low pressure liquids are vaporised in the cold box and vapours routed to the C1 Recycle Compressor (114-C-211). The vapour phases resulting from the sequential flashes are routed directly to the C1 Recycle Compressor. The C1 Recycle Compressor provides sufficient head for the low-low pressure methane vapours to rejoin the low pressure methane refrigerant stream at the suction of the 1st stage of the Main C1 Refrigerant Compressor (114-C-201). It is then compressed and mixed with the returning MP methane refrigerant stream from the cold box before being compressed in the final compression stages and recycled.

6.1.3 Hydrogen Refrigerant Circuit

The hydrogen refrigerant cycle operates as a closed circuit with the hydrogen refrigerant compressed and circulated by electrically driven centrifugal compressors. The H2 Refrigerant Compressor (114-C101 to 105) compresses the recycled hydrogen returned from the cold box. The compressor has two feeds that float on the Warm and Cold H2 Expander (114-X-101 & 102) outlet pressures. The heat of compression is removed by the H2 Refrigerant Compressor Inter and After Coolers (114-E-101 to 105 respectively). The hydrogen stream is then routed to the warm end of the Precooling Cold Box where it is cooled against the methane refrigerant circuit.

The precooled hydrogen refrigerant stream is then split with approximately half of the stream routed to the Liquefaction Cold Box (114-E-006) and with the other half routed to the Warm H2 Expanders (114-X-101). The warm expander duty is let down across three (3) stages of expansion to provide the 1st section cooling for the Liquefaction Cold Box.

The remaining hydrogen refrigerant is further cooled before being let down in pressure across the Cold H2 Expanders (114-X-102) to the LP hydrogen refrigerant pressure, providing the final level of cooling for the Liquefaction Cold Box.

Both the LP and MP hydrogen refrigerant return streams are routed back through the liquefaction and precooling cold boxes to the Hydrogen Refrigerant Compressor suction, with the LP Hydrogen Refrigerant entering the compressor at the LP suction and MP Hydrogen Refrigerant at the inlet to the second stage.

6.2 Design Development

Enhanced PFDs and a MEL were developed for the scheme. A process control philosophy was developed and reflected on the PFDs together with major line sizes. The equipment list included duty, design conditions, materials of construction, power and dimensions (all as appropriate to the equipment type).

A dynamic analysis was performed to confirm OHL's transient response to a number of predefined scenarios including steady operation, turndown / turn-up, equipment trips, and start-up from ambient.

The objectives for the analysis were to:

- Confirm the proposed configuration of equipment and controls
- Confirm the operability of the proposed design
- Confirm a preliminary sequence for starting up the OHL process.
- Confirm the interactions between the rotating equipment and the main cold boxes during unsteady operating conditions

The dynamic model was initially set-up to reflect the 300 te/d steady state simulation model to validate that the model provided an acceptable representation of the process design. Following this, the predefined transient cases were analysed to assess the system response. The main findings indicated:

- The control scheme was shown to work well for system throughput changes down to 50% and back up to 100%.
- Main expander trips were survivable, with the control system able automatically reduce the LNG production to a new steady state. The transient temperature excursions could be managed within acceptable limits for the main exchangers, based on ALPEMA guidelines and subject to discussions with exchanger suppliers.
- Start-up time from ambient conditions could be achieved in less than 20 hours.

A utilities summary was prepared covering consumptions of cooling water, electric power, instrument air, nitrogen and demineralised water.

7.0 Project Estimate

7.1 Estimate Scope

Site Location: UK / Northern Europe site – Fabrication Yard in South East Asia

Estimate Class: Class 4

McDermott prepared a Cost Estimate to an AACE Class 4 methodology as part of the Study to determine the order of magnitude cost for the Liquefaction Module and supporting LNG refrigerant storage / treatment and transfer.

The estimate is based on costs prevailing 4Q2021 and excludes the following:

• Owners costs	• Contingency
• Forward Escalation	• Turnkey margin
• Import Duties	• Spare Parts
• Licence Fees & Permits	• Module connection and tie-ins
• Insurance costs	• Offsite including LH2 storage

7.2 Estimate Summary

Description	US Dollars
Equipment	145,800,000
Bulk Materials	69,700,000
Freight	17,600,000
Site Construction	7,000,000
Module Fabrication	117,700,000
Module Transport	4,300,000
Vendor Representative	7,300,000
Construction/Fabrication Management	9,100,000
Engineering	79,000,000
Pre-commissioning & Commissioning	11,000,000
Other	32,300,000
TOTAL COST	**500,800,000**

8.0 Summary Outcomes

McDermott completed a verification of the initial process design concept provided by Gasconsult and engaged with multiple suppliers to confirm the availability, performance, configuration options, level of maturity for the required equipment and for identification of key risks.

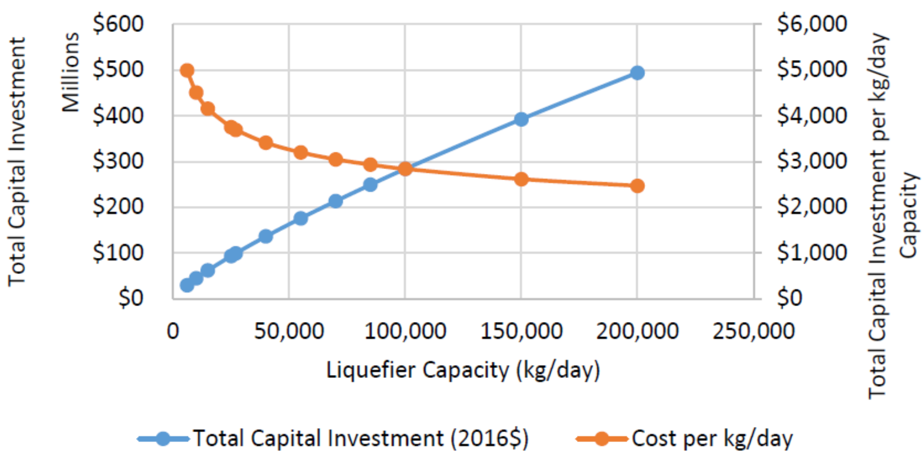
The verification process and initial supplier engagement identified modifications to the initial concept which were incorporated into the Base Case design and form part of the study outcomes.

From the work carried out by McDermott during this study, Gasconsult's OHL process has the capability to achieve an overall performance of ~6.8 to 7.1 kWh/kg LH2 for a 300 te/d capacity (depending on selected compressor type / configuration). This represents a power demand reduction of 25-40% relative to current operating practice (resulting in a reduction in CO2 emissions of ~0.4 Mtpa and ~0.165 Mtpa relative to coal and natural gas derived power respectively[2]). It is important to recognise that this reduction in power demand allows production of some 40-60% more LH2 from equivalent sized compression equipment, and thus has a marked impact on the capital efficiency of the process, realising a capital cost of ~$1670 per daily kg of installed LH2 capacity.

The overall CAPEX for the 300 te/d facility is estimated to be US$ 500 Million (Class 4). According to the US Department of Energy (DOE) study into Hydrogen Liquefaction Costs[1] the total capital investment for hydrogen liquefaction in 2016 ranged from US$30 to US$490 million for capacities between 6 te/d and 200 te/d of LH2. This includes for land costs at $12.35/m^2 for the liquefier only (negligible), and 12% for owner's cost. The cost curve is shown in Figure 4.

Extrapolating the curve data, and assuming a 2% average annual inflation rate[3] for the five years from 2016 to 2021, it is estimated that a 300 te/d liquefier would cost in the order of US$600 million (excluding owners and land costs). This is US$100 million more than the US$ 500 million estimated CAPEX for the OHL process at this capacity, and as such implies that the OHL process could offer a ~20% reduction compared to scaling up current, low efficiency, open nitrogen precooling technologies.

Figure 4 – Total Capital Investment of Liquefier by Capacity[1]



SOURCE: **DOE Hydrogen and Fuel Cells Program Record – Current Status of Hydrogen Liquefaction Costs, 6th August 2019**[1]

Research into alternative developing LH2 technologies indicates a number of other OHL advantages. Competing schemes have been promoted which will be comparable in energy efficiency to OHL, but with the following disadvantages:

- A dependence on high cost and scarce refrigerants such as helium and neon
- More complex pre-cooling through use of mixed refrigerants which carry the burden of high refrigerant infrastructure costs to store and blend a range of liquid hydrocarbon refrigerants
- More complex operations – mixed refrigerant processes require regular composition adjustment to run at their design point
- A requirement to buy liquid nitrogen pre-cooling refrigerant

In summary OHL appears an efficient, commercially competitive option for future large-scale hydrogen liquefaction facilities. Assuming capital amortisation over a 5-year period and power costing $0.03 per kWh, liquefaction costs compute to $1.40/kg LH2. This compares to $2.23 computed by the US DOE HADSAM (Hydrogen Delivery Scenario Analysis Model[4]) for current commercial practice.

The major risk is the uncertainty in the performance of the ortho to para conversion catalyst and the impact on liquefier exchanger performance. This risk can be mitigated by conservative design but further investigation (physical lab scale testing – please refer to 10.1) is recommended to eliminate this risk.

9.0 Market Potential

There is universal agreement that hydrogen, as a zero emissions fuel, has outstanding technical credentials as a contributor to 2050 net zero objectives. There is limited consensus however as to the extent to which hydrogen will displace fossil fuels. This lack of consensus as to future hydrogen demand arises from uncertainty regarding the future cost of hydrogen and the future level of carbon pricing. There is clearly huge potential however, given the world consumes over 4 billion tpa of liquid fossil fuels.

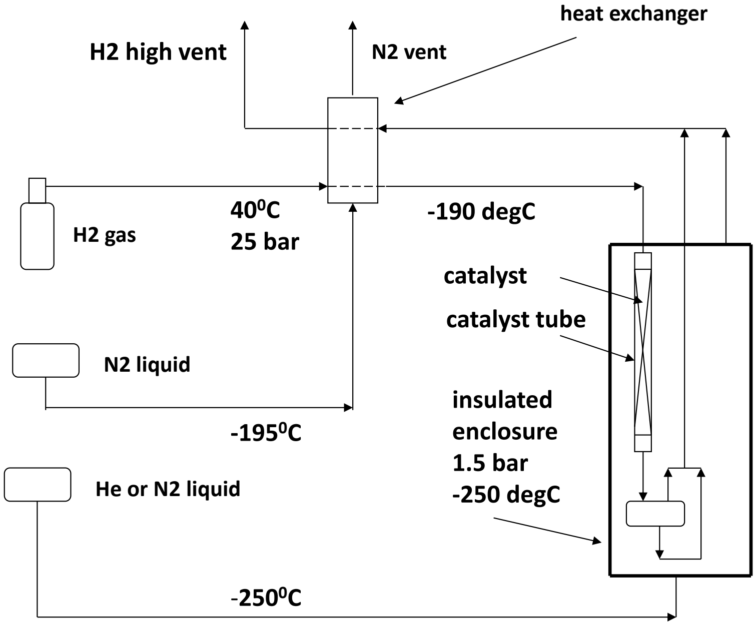
The Hydrogen Council and DNV have both produced estimates of 2050 demand.

The Hydrogen Council projects 650 million tpa. DNV projects 200 million tpa. Assuming 5% of this demand is in liquid form LH2 demand in 2050 would be between 10 million tpa and 32 million tpa. This is in contrast to current global LH2 production of only 150,000 tpa; mainly used as rocket fuel.

To make a material impact, substituting the liquid fossil fuels market will require new LH2 production plants with a capacity an order of magnitude higher than the existing plants serving the rocket fuel market. Gasconsult foresees a progressive increase in plant capacities up to 300 te/d, or beyond, with potentially 500 new liquefaction plants built by 2050. These, like oil refineries, will be built world-wide with a total CAPEX in the order of £180 billion.

10.0 Way Forward

10.1 Phase 2 Testing



When storing LH2 it is necessary to convert the ortho-hydrogen isomer into the para isomer. Reduction of ortho-hydrogen is essential as its conversion to para in storage is exothermic and would lead to excessive hydrogen gas boil off. Data from the most experienced conversion catalyst supplier is not sufficient to accurately size the conversion reactor system and establish its operating conditions. Although this would not prevent design of a demonstration plant it would lead to a conservative risk mitigated design, compromising competitiveness. Gasconsult's Phase 2 project will involve the building of a test rig and operating it under controlled conditions to establish the required design data.

To provide the most economical full-scale design under the most advantageous process conditions Gasconsult proposes to install the conversion catalyst in the liquefier's cold end heat exchanger. This exchanger, in practice, will comprise multiple heat transfer channels of approximately 600 square mm area. A single tube test rig of the same dimensions will therefore mitigate scale-up risk. A schematic of the proposed test rig is shown in the inset.

In addition to providing the needed data this work will also generate proprietary information and would be valuable Intellectual Property within the overall OHL sales package.

10.2 Further Engineering Development

10.2.1 Process Configuration

- Feed gas and refrigerant pressure levels should be further investigated to understand the impact of the change on the overall process performance and complexity.
- The temperature approaches and pass pressure drop allowances for the main exchangers should be reviewed in combination with the work on required catalyst volume to optimise the exchanger design
- Possible combination of the H2 recycle gas duty with the main hydrogen refrigeration loop should be reviewed. This offers the opportunity to reduce the overall number of rotating / reciprocating bodies and hence CAPEX, however the impact of the change to the ortho-para hydrogen equilibrium in the hydrogen feed to the precooler needs to be reviewed with respect to overall conversion and heat of reaction to be removed.

- Sensitivities on the number of levels and location of the rundown flash should be carried out to understand impact to overall performance versus equipment count and exchanger passes.

10.2.2 Facility Layout

It is recommended to perform further work to develop the liquefaction unit plot plan and layout. This will reduce the level of uncertainty in the hydraulic design and pressure profile of the unit and allow safety studies and risk assessments to be performed. As part of this layout work, the potential for integration between the storage/loading and liquefaction units should be assessed. This, taking account of the required facility spacing and risk profile, will determine the potential for BOG integration with the liquefaction unit for larger scale facilities.

The layout work will additionally allow module constructability studies and by facilitating bulk material take-offs allow enhancement of the capital cost estimate to a higher level of accuracy.

10.2.3 Insulation Concept

For larger scale facilities, the vacuum insulation concept needs to be reviewed in further detail as this potentially limits the liquefier cold box capacity and is an area of high cost. It is recommended to review the cold box designs with suppliers to determine if vacuum insulation is the optimum concept for the larger scale facilities.

10.2.4 Supplier Engagement

It is evident that large-scale processing of high purity hydrogen to support a changing energy market is an area of significant interest and development within the industry. Many suppliers have current and planned development work to improve equipment performance, increase the scale and capacity of equipment and to identify new solutions where a market need has been identified. Continued engagement with the suppliers is required to ensure that the latest technologies and developments are included in the OHL process.

10.3 Route to Commercialisation

The industrial gases companies, who supply the current LH2 market are potential customers for OHL. However, their current LH2 operations (a combined 150,000 tpa) are insignificant relative to global transportation and industrial fuel use of 4 billion tpa. These LH2 incumbents have limited distribution networks and customer bases in the fuels market. The oil majors have the most to lose in the energy transition and will fiercely defend their existing markets. They are also:

- the largest producers and consumers of hydrogen
- the largest producers/distributors of transportation/industrial fuels

A major motivator for an oil major to involve itself in the technology development would be to preserve their position in the fuels market by ensuring ongoing direct access to a competitive LH2 process in what is at present, a very tightly controlled technology landscape. The oil majors also have the technical knowledge to assess risk on a first of a kind project.

Without excluding other opportunities that may arise, the oil majors will thus be targeted by Gasconsult for sales of OHL, and approached with a view to their building a commercially viable demonstration plant based on the completed BEIS feasibility study.

It is anticipated that a selected partner organisation may participate in the Phase 2 testing outlined in 10.1, further engineering development per 10.2 and initiate an up-front due diligence on the OHL process. Depending on the outcome of the due diligence, two routes to commercialisation are possible. Selection of the route will depend on the partner's assessment of risk:

1. construction of a pilot plant, or
2. construction of a commercial scale demonstration plant

Whichever route is selected the project would then progress through the following phases:

3. preparation of a FEED package and associated cost estimate (likely cost £5 million and for the partner's account with completion 2 years after completion of the BEIS feasibility study)
4. approval by the partner's Board to proceed with the design and construction of either the pilot or commercial demonstration plant
5. detail design, procurement and construction of the pilot or commercial demonstration plant (at a cost of ~£25 million or ~£100 million respectively and with completion 5 years after completion of the BEIS feasibility study)

10.4 Monetising OHL

In terms of monetising the OHL technology Gasconsult will follow a low-risk licensing model. Typically, capital projects are developed by preparing a front-end package comprising key process data and preliminary engineering adequate to support a ~10% accurate capital cost estimate. This package is then used to solicit engineering/procurement or design/construct offers from contractors.

Gasconsult, together with selected process contractors, will target selling the initial front-end package incorporating the OHL technology Licence. This will realise revenues of ~10% of the plant CAPEX of which Gasconsult's share will be ~5% (4% Licence fee + 1% engineering). Depending on the extent to which the LH2 market develops and based on a 10% LH2 market share, Gasconsult predicts 1 to 2 licence sales each year. Gasconsult would ultimately employ ~15 people and would remain profitable even on the lower DNV LH2 demand forecasts.

10.5 Marketing OHL

OHL is more energy efficient than current LH2 plants and less complex than newly emerging technologies. These OPEX and CAPEX advantages will be Gasconsult's USPs when marketing, supported by the data available from the feasibility study subject of this report.

A schedule will be drawn up to maintain regular contacts and interact with key players in the hydrogen sector. New contacts will be identified and new areas of business activity explored where the OHL process might find application (renewable power operators, storage terminal operators). Gasconsult will maintain industry access underpinned by presentations at industry forums and publications in technical journals.

During the course of 2022 Gasconsult has made over 15 high level promotions of the OHL technology to industry leaders. See below.

POWERPOINT OR SIMILAR ZOOM PRESENTATIONS MADE ON OHL		
Company		Ongoing Actions
ExxonMobil	13/07/2022	Update being scheduled
Messer Group	05/01/2022	Due to revert on completion of feasibility study
Confidential Oil Major	01/02/2022	Draft collaboration agreement in process
Cryopeak	19/01/2022	
Vopak	13/01/2022	Due to revert on completion of feasibility study
Plug Power	01/02/2022	Due to revert on completion of feasibility study
Babcock USA	02/02/2022	Due to revert on completion of feasibility study
Chart	11/02/2022	Due to revert on completion of feasibility study
Snam	18/02/2022	Due to revert on completion of feasibility study
Snam	18/03/2022	Due to revert on completion of feasibility study
Aecom	01/03/2022	
Magnum Development	04/04/2022	Due to revert on completion of feasibility study
Schlumberger	26/05/2022	Due to revert on completion of feasibility study
Next Era	27/05/2022	Due to revert on completion of feasibility study
Fortescue Future Industries	07/10/2021	Due to revert on completion of feasibility study
Hycap	10/12/2021	
BP Energy Partners	19/01/2022	
Strital	19/04/2022	
Tractebel	04/01/2022	

Progress will also be measured in an internal strategy meeting once per year when new concepts and strategies will also be discussed and developed for future roll-out.

10.6 Intellectual Property

Gasconsult has applied for a number of patents in respect of OHL. Inter-alia these cover:

- The liquefying expander methane (or nitrogen) pre-cooling concept
- Optimisation scheme to mitigate close temperature approaches on the hydrogen liquefier exchanger
- An enhanced insulation concept to handle the extremely low temperatures inherent in hydrogen liquefaction.

11.0 References

1. DOE Hydrogen and Fuel Cells Program Record #19001 6 August 2019
2. US EIA https://www.eia.gov/tools/faqs/faq.php?id=74&t=11 (4th November 2021)
3. https://tradingeconomics.com/united-states/inflation-cpi (10th June 2022).
4. https://hdsam.es.anl.gov/index.php?content=hdsam