Programme Area: Carbon Capture and Storage

Project: Hydrogen Turbines

Title: Hydrogen Storage and Flexible Turbine Systems WP1 Report – Hydrogen Power Production

Abstract:

The purpose and focus of the Hydrogen Turbines project is to improve the ETI's understanding of the economics of flexible power generation systems comprising hydrogen production (with CCS), intermediate hydrogen storage (e.g. in salt caverns) and flexible turbines, and to provide data on the potential economics and technical requirements of such technology to refine overall energy system modelling inputs. The final deliverable (D2) comprises eight separate components. This document is D2 WP1 Report – providing results of a techno-economic assessment of options for hydrogen power production from coal, coal/biomass and gas.

Context:

This £300k project, led by global engineering and construction company Amec Foster Wheeler, in collaboration with the BGS, assessed the economics of a range of flexible power generation systems which involve the production of hydrogen (with CCS) from coal, biomass or natural gas, its intermediate storage (e.g. in salt caverns deep underground) and production of power in flexible turbines. The work included mapping of potentially suitable hydrogen storage salt cavern sites in and around the UK and provided the ETI with a flexible economic modelling tool to assess the range of possible options. The ETI's energy system modelling work suggests that systems such as these could provide a valuable contribution to the future energy mix, filling the gap between base load nuclear plant and low carbon power generation.

Contract No.:	1.17.13058
Client's Name:	The Energy Technologies Institute
Project Title:	Hydrogen Storage and Flexible Turbine Systems
Project Location:	Generic UK

REVISION	**Rev O1 (Preliminary)**	**Rev O2 (Preliminary)**	**Rev A1 (Final)**		
DATE	22 March 2013	4 April 2013	12 July 2013		
ORIG. BY	S. Ferguson	R. Ray	S. Ferguson	Ferguson	
CHKD BY	G. Skinner	G. Skinner	G. Skinner	G Skinner	
APP. BY	T. Abbott	T. Abbott	T. Abbott	Tim Abbott .	

Hydrogen Storage and Flexible Turbine Systems
WP1 Report – Hydrogen Power Production

CONTENTS

ATTACHMENTS

DISCLAIMER

The information contained herein is provided by Foster Wheeler Energy Limited (FWEL) to Energy Technologies Institute LLP (ETI), solely to assist ETI in improving its understanding of flexible power generation systems comprising of hydrogen production, storage and turbines, and to enable ETI to refine its Energy System Modelling Environment (ESME) model.

FWEL has not made any independent verification of data and information contained herein that has been supplied by ETI or other third parties. This report is intended for the sole use of ETI and FWEL makes no representation or warranty, express or implied, and assumes no obligation or liability, whatsoever, to any third party with respect to the veracity, adequacy, completeness, accuracy or use of any information contained herein.

The information provided is not, and should not be construed as, a recommendation by FWEL that any recipient provide finance to the project. Each recipient of this document should make its own independent evaluation of the project and of the relevance and accuracy of the information contained herein, and should make such other investigations as it deems necessary to determine whether to extend credit to the project.

1. EXECUTIVE SUMMARY

1.1 Introduction

Fossil fuel based power generation currently plays a key part in providing for the UK's energy demands. The development and implementation of Carbon Capture and Storage (CCS) technologies is an important option in reducing the associated CO2 emissions, but adding CCS to conventional power systems impacts their ability to respond to power demand fluctuations, since the column systems for CO2 removal work best at steady state conditions and are inefficient in turndown operation. Adding intermediate storage of hydrogen-rich fuel gas to a pre-combustion carbon capture scheme could be an attractive way of achieving flexible low-carbon power generation for the UK: the upstream carbon capture system would normally operate at a steady, base load capacity for maximum efficiency, while the hydrogen store would provide buffer capacity to allow the downstream hydrogen based power generation scheme to respond to demand fluctuations.

The purpose and focus of this project is:

- To improve the ETI's understanding of the economics of flexible power generation systems comprising hydrogen production (with CCS), intermediate hydrogen storage (e.g. in salt caverns) and flexible turbines; and
- To focus on the potential, economics and technical requirements for salt cavern storage and flexible turbines, to enable refinement of the ETI Energy System Modelling Environment (ESME) model in order to confirm or adjust ESME findings.

1.2 Scope

This report covers the work undertaken in the execution of WP1 – Hydrogen Power Production.

The aim of WP1 is to review the technical options and economics for hydrogen production with CCS and hydrogen fired turbine power generation, against the likely energy system requirements of 2030.

The scope of WP1 consists of:

- A review of options for hydrogen production;
- Techno-economic definition of the following four options for hydrogen production:
  - gasification of coal;
  - gasification of a coal/biomass mix;
  - autothermal reforming of natural gas; and
  - steam reforming of natural gas;
- Characterisation of basic design requirements for cost effective hydrogen stores;
- A review of requirements / options for hydrogen turbines, being:
  - Power generation gas turbines;
  - Gas expansion turbines;
- A brief study of the economics of hydrogen pipelines;
- A brief discussion of the effects of hydrogen purity.

1.3 Key Findings of Work Package 1 – Hydrogen Power Production

The following items describe the key findings of this section of the Hydrogen Storage and Flexible Turbine Systems Project.

1.3.1 Options for hydrogen production

There are a number of well established hydrogen production schemes suitable for producing a low carbon, or carbon free, fuel gas at a scale capable of supplying a commercial gas turbine combined cycle power plant. These include:

- Gasification of coal, oil, petcoke & biomass,
- Reforming of natural gas via partial oxidation, autothermal reforming (either using oxygen, air or a mixture) or steam methane reforming.

Gas turbines (GTs) currently available for firing a low/no carbon hydrogen-rich fuel gas require dilution of the fuel gas with nitrogen and/or steam, primarily to limit exhaust NO_x content.

As most of the hydrogen production processes listed above require an Air Separation Unit (ASU) for oxygen production, resulting in simultaneous “free” nitrogen production, nitrogen is the generally assumed dilution option. The dilution nitrogen produced in the ASU will require storage, similar in volume to the hydrogen store, during the time that the power plant is not operating.

Liquid nitrogen (LIN) storage with an evaporator was eliminated from consideration since the operation of an evaporator to meet the expected continuously varying power plant load demand may be problematic, and there will be a high energy penalty associated with the liquefaction and subsequent vaporisation.

An air-blown ATR would not need an ASU, which has both significant cost and operational inflexibility associated with it. Since the air-blown ATR naturally results in a premixed fuel gas product containing almost exactly the required nitrogen to hydrogen mixture required for the GT, it was decided to assume co-storage for this case. It was considered that an oxygen-blown ATR would not benefit from these advantages and would therefore be less favourable.

In order to preserve purity of the hydrogen-rich gas and nitrogen where possible, it was decided to assume separate nitrogen and hydrogen storage for all other cases.

An Australian steam coal was considered as the base case gasifier fuel. A case including some (wood pellet) biomass firing was also desired, so a coal & biomass based gasification case was also selected. Petcoke and oil firing were eliminated as they would not be economically sensible options in the UK.

A steam methane reforming (SMR) case was selected due to the prevalence of this method of hydrogen production in existing plants. The SMR plant was expected to be lower in capital cost but more challenging in general due to complexity of flue gas decarbonisation and lack of nitrogen production as a process by-product.

A standalone ASU would be required for the SMR case in order to provide the nitrogen required for the GT, with the by-product oxygen being vented. As 100% steam dilution gas for the GT is possible, a sub-case (4b) was added, in which there is no ASU and the GT uses only steam for dilution, using steam generated within the power island,

The scale of the plant is another variable. It was agreed (for WP1) that all cases would be based on operation of one GE Frame 9 (Syngas) gas turbine.

1.3.2 Techno-economic definition of the selected hydrogen production options

The techno-economic definition provides a high-level comparison of the four methods of hydrogen-rich fuel gas production. For a fair comparison between technologies, the operating pattern of the plant must be the same for all cases. For the purposes of calculating overall plant efficiency and other technical data a steady state basis is also required. For the purpose of the technology comparison within WP1, it has been assumed that the hydrogen production plant operates continuously (i.e. 24 h/day, 365 days/y), and is sized at the capacity required to supply gas to a single GE gas turbine operating continuously at full load.

For the overall scheme (including hydrogen production plant, salt cavern storage facility and power island) to be assessed, the above hydrogen plant sizing basis will need to be scaled based on the total number of gas turbines and operating regime being considered. This will be considered in the WP4 modelling.

It has been assumed that the three main plant elements (hydrogen production, gas storage caverns and the hydrogen-fired “power island”) are all located at a common site in a generic UK location, with minimal lengths of interconnecting pipe and no heat integration.

In all cases, except Case 3 (the air-blown ATR case), >85 mol% purity hydrogen is produced from the hydrogen production unit for storage or direct use in the gas turbine(s).

To ensure a fair comparison between the cases, it was assumed that the hydrogen production plant delivery pressure is constant and at the same pressure as that required for the gas turbine 33.4 bar (abs).

Additional design basis data, such as feedstock compositions, climatic data and utility assumptions can be found in the Basis of Design, Attachment 1.

Case 1 – Hydrogen Production via Gasification of Coal

Refer to Figure 1 – BFD for Case 1 (page 33)

The process comprises coal milling and drying, gasification to form a synthesis gas (syngas), made up largely of hydrogen and carbon monoxide, shifting the syngas with steam to produce additional hydrogen and convert the carbon monoxide to carbon dioxide (CO_2), heat recovery and separation of the syngas into a hydrogen rich stream, a CO_2 rich stream and an H_2S rich stream. Sulphur is then recovered from the H_2S rich stream and the remaining components are recycled to the syngas. The hydrogen rich stream is compressed and sent to the power island or to underground storage. The CO_2 stream is dried, compressed and exported for sequestration.

Case 2 – Hydrogen Production via Gasification of a Coal/Biomass Mix

Refer to Figure 2 – BFD for Case 2 (page 35)

The biomass specified for this case is wood pellets, with characteristics described in the Basis of Design (see Attachment 1). The process comprises coal and biomass co-milling and drying, gasification to form a raw synthesis gas (syngas), shifting the syngas with additional steam to produce additional hydrogen and convert the carbon monoxide to carbon dioxide (CO_2), heat recovery and separation of the syngas into a hydrogen rich stream, a CO_2 rich stream and an H_2S rich stream. Sulphur is then recovered from the H_2S rich stream and the remaining components are recycled to the syngas. The hydrogen rich stream is then compressed and sent to the power island or to underground storage. The CO_2 stream is dried, compressed and exported for sequestration.

Case 3 – Hydrogen Production via Autothermal Reforming of Natural Gas

Refer to Figure 3 – BFD for Case 3 (page 38)

The process involves desulphurising the natural gas before mixing it with steam and feeding it to the autothermal reformer (ATR), in which the natural gas / steam mixture is reacted with preheated air and is converted to syngas. The ATR is operated with an outlet pressure of 28.5 bara. The syngas from the ATR is then shifted with the residual water vapour contained in the ATR product stream to produce additional hydrogen and convert the carbon monoxide to CO_2. After heat recovery the shifted syngas is separated using an amine-based acid gas removal unit (AGR) into a hydrogen/nitrogen stream and a CO_2 rich stream. The hydrogen and nitrogen rich stream is then compressed and sent to the power island or to underground storage. The CO_2 stream is dried, compressed and exported for sequestration.

Case 4 – Hydrogen Production via Steam Methane Reforming of Natural Gas

Refer to Figure 5 – BFD for Case 4 (page 43)

The process involves pre-treating the gas before mixing it with steam and feeding it to the steam reformer, which comprises a large number of externally heated tubes containing a nickel-based catalyst. In the catalyst tubes the mixed natural gas/steam feed is converted to a syngas made up largely of hydrogen and carbon monoxide. The SMR is operated with a catalyst tube outlet pressure of 27.7 bara. The syngas from the SMR is then shifted with water to produce additional hydrogen and convert the carbon monoxide to CO_2. Heat is then recovered and the syngas is separated using a PSA into a hydrogen rich stream and a CO_2 rich stream. The hydrogen rich stream is then compressed and sent to the power island or to underground storage. The CO_2 rich stream, also containing hydrogen, CO and residual methane is fed as fuel to the reformer, along with supplementary natural gas fuel. The reformer flue gas is cooled and scrubbed with an amine solvent, so as to capture 90% of the carbon content of the natural gas feed to the SMR unit.

Applying a pre-combustion carbon capture scheme to an SMR unit only achieves capture of the CO_2 content of the shifted gas, and it does not address the CO_2 emissions from firing of natural gas and PSA tail gas in the reformer. Typically this supplementary natural gas fuel for reformer firing accounts for nearly 20% of the total feed so it is essential to capture carbon from this fuel in order to achieve 90% carbon capture overall. A pre-combustion capture scheme only captures approximately 60% of the total carbon fed to the SMR. Therefore, whilst a pre-combustion capture scheme may result in a lower capital cost, a post-combustion capture scheme has been considered for this study, as shown in Figure 5.

Case 4b – Hydrogen Production via Steam Methane Reforming with Steam to GT

Refer to Figure 6 – BFD for Case 4b (page 44)

This scheme is essentially identical to the SMR-based scheme in Case 4, except that nitrogen required for the power island gas turbine is fully replaced with steam generated in the gas turbine heat recovery steam generator. This change allows the deletion of the ASU and its nitrogen compressor from the scheme as well as the need for nitrogen storage. These items contribute significantly to savings in both capital cost and continuous power requirements, and hence a scheme avoiding these items may prove beneficial.

The overall performance of the Case 4b scheme has been estimated based upon the assumption that nitrogen is replaced by flow rate of steam advised by GE.

Technical Performance

Table 1 summarises the key numerical findings from the high level simulations of power generation schemes based upon the hydrogen production options considered.

For the purposes of calculating overall plant efficiency a steady state basis is required. The figures below represent the energy balance if the plant were operating at the full capacity of one GT at all times.

Table 1 - Summary of technical performance results

Steady state basis with the plant running at 100% GT capacity			Coal Gasification	Coal & Biomass Gasification	ATR	SMR	SMR with Steam to GT
			Case 1	Case 2	Case 3	Case 4	Case 4b
Hydrogen Production							
	Feedstock flow rate	te/h	136.52	152.74	65.08	76.63	73.86
	Total Feedstock LHV	MWth	986.00	1002.50	859.29	1011.79	975.19
	Carbon in feeds	te/h	88.46	90.00	47.57	56.35	54.31
	Carbon captured	te/h	79.62	80.99	43.10	50.95	49.11
	Carbon captured		90.0%	90.0%	90.6%	90.4%	90.4%
	Oxygen consumption (Note 1)	te/h	102.75	97.20	0.00	103.84	0.00
	Syngas Product LHV	MWth	666.0	665.8	665.7	666.0	666.0
Power Balance							
	Hydrogen production total	kWe	-49713	-48069	-33932	-55585	-2410
	Pre-treatment gasification/reforming & shift	kWe	-8940	-9799	-45475	-1522	-1445
	Heat recovery & steam turbine	kWe	59047	60334	42598	27599	28303
	Acid gas removal	kWe	-7808	-7808	-4665	-3078	-2966
	CO2 dehydration and compression	kWe	-25120	-25543	-17543	-20399	-19661
	Sulphur recovery and tail gas treatment	kWe	-1729	-1882	0	0	0
	Air separation unit	kWe	-34295	-32462	0	-19148	0
	N2/H2 compression	kWe	-27499	-27409	-6208	-34638	-2905
	Fresh cooling water	kWe	-954	-995	-775	-1546	-1313
	Sea cooling water	kWe	-2416	-2506	-1863	-2854	-2423
	Cavern storage	kWe	0	0	0	0	0
	Power Generation	kWe	391193	391152	389650	392788	353572
	Offsites & Utilities	kWe	-2264	-2286	-2338	-2338	-2338
	Net Power Export	kWe	339216	340797	353380	334865	348825
	Plant Efficiency (LHV)		34.40%	33.99%	41.12%	33.10%	35.77%

Note 1: Oxygen produced in the ASU in Case 4 is not used by the process, and can either be vented or potentially exported.

With the exception of the SMR cases (Cases 4 and 4b) the plant efficiencies in Table 1 above are broadly in line with historical studies of pre-combustion capture power generation processes with 90% carbon capture, which show approx 40% with natural gas feedstock and approx 35% LHV efficiency with bituminous coal feedstock. The lower efficiency of the SMR cases is mainly attributable to the relatively low steam generation pressure in the reforming unit (35 bar) and the high reboiler heat duty of the flue gas CO2 capture unit (amine process).

Previous studies indicate that plant efficiency is not significantly affected by the selection of oxidant. Foster Wheeler has simulated an oxygen-fed ATR (outside the scope of the this study) giving an overall efficiency approximately 41% - same as for Case 3 above with an air-fed ATR. For coal feed it may be noted that the Mitsubishi-designed IGCC operating in Japan with air-fed gasifier is considered to have superior efficiency to oxygen-fed IGCC schemes.

From Table 1, the following observations can be made:

- The carbon content in the natural gas fed cases is much lower than the coal fed cases since a significant portion of the energy in the natural gas feed is provided by the hydrogen content of natural gas.
- The oxygen production/consumption is very similar in the two gasification cases. Slightly less oxygen is required for the biomass case since more oxygen is contained in the biomass feedstock.
- In all cases the power recovered by using excess process heating for steam generation enables the hydrogen production facility to meet approximately 50% of its own parasitic demand. In the SMR case with steam dilution of the fuel gas and hence no ASU, however, the hydrogen production plant is close to self-sufficient in power.
- The very high power demand in the reforming section of the ATR plant is due to the air compressor. Cases 3 and 4b have much lower power demand for the H2/N2 compression unit as these two schemes do not require a nitrogen compressor.
- Net power produced by the power island is similar in all cases since the impact on the power island between cases is very small, due only to minor changes in fuel gas compositions. In case 4b the power produced is lower since more energy internal to the power island is required to provide the steam used for fuel gas dilution and NOx control.
- The total quantity of net power available for export is very similar in all cases, with the ATR and SMR with steam dilution cases exporting slightly more power than the other cases.
- The overall efficiency figures show that the above differences in performance combine to put the ATR case ahead of the other 4 cases by a margin of nearly 5.4 LHV efficiency % points. Compared to previous similar studies this is a very significant margin. The efficiency differences between the two gasification cases and between the two SMR cases are small in comparison.
- The amount of natural gas consumed in ATR case (Case 3) is expected to be around 30% higher per kWh than the mainstream CCGT + post-combustion CCS technology for which the plant efficiency is now typically 53%. This 30% increase in fuel cost may cause concern at a time when natural gas storage capacity and national security of gas supply are in focus.

The superior performance of the ATR case does not necessarily mean that a natural gas fed ATR with carbon capture is the best choice for the hydrogen storage and flexible turbines system, as the thermal efficiency of the plant is only one of a number of key factors. Feedstock price, availability and overall plant economics will have a big impact on whether or not the ATR case would be the best option for this project moving forward.

GE has advised that substitution of nitrogen dilution with steam is feasible, but the GT would be de-rated to mitigate the expected higher maintenance costs associated with steam dilution. To make an allowance for such derating, Foster Wheeler has made preliminary simulations of steam injection, using guidelines from GE, in place of nitrogen. For the SMR Case 4b, steam injection results in a gain in overall efficiency of about 2.7% points, due to a significant contribution from elimination of the ASU and N_2 compressor. It is estimated that for Cases 1 & 2, steam injection will result in a smaller increase in overall efficiency of <0.5% points.

Capital Costs

The capital cost estimates generated for this study are factored estimates, based on the scaling of previous similar estimates prepared using Foster Wheeler in-house data, and are considered to have an overall project cost accuracy of +/-40% at best. For all of the cases reported the source estimate data has been adjusted to provide figures on a consistent and comparable) 1Q 2010 UK Basis.

It can be seen that there is a small incremental cost - circa 1% - associated with the Case 2 (coal + biomass fed gasification) process when compared with the Case 1 (coal fed gasification) process.

Table 2 – Capital Costs Estimate Summary

DESCRIPTION	CASE 1 COAL GASIFICATION	CASE 2 COAL & BIO GASIFICATION	CASE 3 ATR	CASE 4 SMR	CASE 4B SMR	POWER ISLAND
	Million £	Million £	Million £	Million £	Million £	Million £
MAJOR EQUIPMENT	193.7	196.0	92.7	188.4	112.0	108.2
DIRECT BULK MATERIALS	89.3	91.9	110.7	61.5	57.1	32.6
DIRECT MATERIAL & LABOUR CONTRACTS	40.5	40.7	96.6	36.9	34.6	12.3
LABOUR ONLY CONTRACTS	95.7	96.5	80.6	95.3	59.4	18.2
INDIRECTS	29.6	30.0	34.7	26.1	18.0	12.1
EPC CONTRACTS	50.8	51.1	62.3	36.4	31.0	8.3
INSTALLED COST	499.7	506.2	477.7	444.6	312.1	191.7
LAND COSTS, 5%	25.0	25.3	23.9	22.2	15.6	9.6
OWNERS COSTS, 10%	50.0	50.6	47.8	44.5	31.2	19.2
CONTINGENCY, 25%	124.9	126.6	119.4	111.2	78.0	47.9
TOTAL PROJECT COST	**699.6**	**708.7**	**669.0**	**622.5**	**436.9**	**268.4**

The Case 3 ATR estimate gives a total project cost of £669m, which is £31m cheaper than Case 1. There are notable differences in the proportional costs within each of the line items in Case 3 (when compared to other cases). This is due to the way the ATR scheme estimate was derived with fewer large packages (Gasifiers, ASU) contributing to the major equipment cost, and an increase to the direct materials and installation costs.

Conversely, the Case 4 SMR estimate gives a total cost of £622m, which is £78m cheaper than Case 1. The SMR option consists of relatively simple and highly packaged process units, which shows in proportionally lower direct materials and installation costs.

The Case 4b SMR without ASU demonstrates the impact on the SMR process of having to produce nitrogen solely for use as a dilution gas for the GT. Removing the ASU reduces the total project cost by £185m to £437m.

Operating Costs

O&M costs are allocated as variable and fixed costs. Variable operating costs are directly proportional to the number of kilowatt-hours produced and are sometimes referred as incremental costs. They may be expressed in £/kWh. Fixed operating costs are essentially independent of the quantity of kilowatt-hours produced. They may be expressed in £/h or £/year.

Table 3 shows that the operating costs are dominated by the cost of purchasing feedstock. Differences caused by considerations such as maintenance and different numbers of personnel required to operate the plant are relatively minor in comparison. Since natural gas is the highest cost feedstock the natural gas fed cases exhibit higher total operating cost, with the coal & biomass case showing similar operating cost and the coal fed gasification case showing the lowest.

The figures shown do not include a cost for imported electricity consumed by the hydrogen production units. This is considered as part of the economics within Table 5.

Table 3 - Operating Costs Estimate Summary

Million £/y	Case 1 Coal Gasification	Case 2 Coal-Biomass Gasification	Case 3 ATR	Case 4 SMR	Case 4b SMR with Steam to GT	Power Island - All Cases
Fixed Costs						
Direct Labour	4.40	4.40	3.00	3.75	3.00	2.00
Administration / General Overheads	1.32	1.32	0.90	1.13	0.90	0.60
Maintenance	15.58	15.89	12.11	12.62	12.11	8.80
Insurance & Local Taxes Allowance	9.99	10.13	8.89	9.55	8.89	6.24
Total Fixed Costs	**31.3**	**31.7**	**24.9**	**27.0**	**24.9**	**17.6**
Variable Costs						
Feedstock	71.2	93.5	102.6	120.8	113.9	0.0
Solvent, Catalysts and Chemicals	1.90	1.90	1.25	1.00	1.00	0.38
Waste Disposal	0.95	0.95	0.63	0.50	0.50	0.19
Total Variable Costs	**74.00**	**96.32**	**104.43**	**122.25**	**115.35**	**0.56**
TOTAL OPERATING COSTS	**105.3**	**128.1**	**129.3**	**149.3**	**140.3**	**18.2**

Project Execution Schedules – Hydrogen Production

The developed schedules show the main differences in the project execution time frame, from the end of “Appraise” phase up to “H_2 Ready for Storage”, as a basis for comparison of the four hydrogen production options under consideration. Schedules above are subject to the qualifications and assumptions presented in section 4.10.

Table 4 – Project Execution Schedule Summary

Phase Description	Case 1 Coal Gasification Duration (Months)	Case 2 Coal/Bio Gasification Duration (Months)	Case 3 ATR Duration (Months)	Case 4 SMR Duration (Months)
Select	6	6	6	6
Define / FEED	9	9	9	9
EPC Bid Award	10	10	10	10
EPC / Start-up	39	39	39	35*
Overall	**64**	**64**	**64**	**60**

The Select, Define / FEED and EPC Bid Award phases will have the same duration, for all of the cases under consideration, as the amount of engineering deliverables to be produced is similar. The critical path for the project execution of the hydrogen production plant is therefore driven by the delivery time span of the main Long Lead Items (LLI)s.

Case 4 uses a single stage steam turbine, while the previous cases are based on multi-stage turbine. The single stage turbine has a manufacturing and delivery time span approximately four months less than a multi-stage and therefore, the Overall Project Execution Schedule is four months shorter for Case 4.

Conclusions

Table 5 summarises the numerical findings on the simplified basis of a plant running continuously as a base-load plant at the capacity of a single GT, in order to demonstrate the differences between the five cases studied.

The differences between the technical performance, capital and operating costs have been discussed in previous sections with the following headline conclusions:

- The overall efficiency figures show that the differences in performance combine to put the ATR case ahead of the other 4 cases by a margin of nearly 5.4 LHV efficiency % points. The efficiency differences between the two gasification cases and between the two SMR cases are small in comparison.
- The capital cost of the two gasification cases is almost identical. The Case 3 ATR estimate gives a total project cost of £669m, which is £31m cheaper than Case 1; the Case 4 SMR estimate gives a total cost of £622m, which is £78m cheaper than Case 1; while the Case 4b SMR without ASU demonstrates the impact on the SMR process of having to produce nitrogen

solely for use as a dilution gas for the GT. Removing the ASU reduces the total project cost by £185m to £437m.

- The operating cost figures are dominated by the price of the various feedstocks, which results in the natural gas fed cases exhibiting the highest total operating cost. The coal & biomass cases show similar costs with the coal fed gasification case giving the lowest operating cost.

The high efficiency of the ATR case means that it is a stronger option than the SMR case with nitrogen dilution to the GT. This is because less feedstock is required to produce the same quantity of product for only a marginal increase in capital cost. However, the SMR with steam dilution to the GT gives the best overall LCOE, as the elimination of the ASU both saves significant capital cost and operating cost through reduced parasitic electricity demand.

Once the capital cost of the cavern is included, Case 4b should look better still, as less storage will be required.

The coal fed gasification case also looks worth further consideration, since its operating cost is substantially lower than the ATR case, even though its overall plant efficiency is just 34.4% compared to the ATR's 41.1%.

In order to quantify the relative merits of operating costs versus capital cost and efficiency it is necessary to perform a high level calculation of the project economic performance over a number of years of operation. In order to do this two simplified levelised cost calculations have been performed for each case. Based on the hydrogen plant capital cost, hydrogen plant operating cost, fixed imported electricity cost and an assumed figure for the value of CO_2 produced by the plant over a 30 year period with a discount rate of 10%, we can calculate:

- Levelised cost of hydrogen production ($LCOH_2$) – this values the technology as a producer of hydrogen; and
- Levelised cost of electricity export (LCOE) – this values the technology as a producer of electrical power.

  n.b. the LCOE calculation estimates the break-even price of electricity produced by the power island, assuming a single gas turbine operating continuously at full load, and with no buffer storage. As such it is only useful as a comparison between these cases, and is not directly comparable with LCOE in other studies.

The $LCOH_2$ calculations show that the cheapest way to produce hydrogen would be the SMR case 4b scheme at £35.1/MWth, where an ASU is not needed for the overall power production scheme. Cases 1 and 3 produce the next cheapest hydrogen at £40.4/MWth.

The LCOE calculation shows that, once the whole power production scheme is included, the SMR Case 4b is the strongest with a lowest LCOE value of £83.3/MWh. This value is comparable to its nearest rival, the ATR Case 3 with LCOE value of £84.2/MWh. A contributing factor to this result is the increased availability of the natural gas fed processes compared to the coal fed cases.

Table 5 - Summary of numerical results for all cases

Steady state basis with the plant running at 100% GT capacity			Coal Gasification	Coal & Biomass Gasification	ATR	SMR	SMR with Steam to GT
			Case 1	Case 2	Case 3	Case 4	Case 4b
Hydrogen Production							
	Feedstock flow rate	te/h	136.52	152.74	65.08	76.63	73.86
	Total Feedstock LHV	MWth	986.00	1002.50	859.29	1011.79	975.19
	Carbon in feeds	te/h	88.46	90.00	47.57	56.35	54.31
	Carbon captured	te/h	79.62	80.99	43.10	50.95	49.11
	Carbon captured	%	90.0%	90.0%	90.6%	90.4%	90.4%
	Oxygen consumption	te/h	102.75	97.20	0.00	103.84	0.00
	Syngas Product LHV	MWth	666.0	665.8	665.7	666.0	666.0
Power Balance							
	Hydrogen production	MWe	-49.71	-48.07	-33.93	-55.59	-2.41
	Cavern storage	MWe	0.00	0.00	0.00	0.00	0.00
	Power Generation	MWe	391.19	391.15	389.65	392.79	353.57
	Offsites & Utilities	MWe	-2.26	-2.29	-2.34	-2.33	-2.34
	Net Power Export	MWe	339.22	340.80	353.38	334.87	348.82
	Plant Efficiency (LHV)	%	**34.40%**	**33.99%**	**41.12%**	**33.10%**	**35.77%**
Costs							
	Hydrogen production capital (TPC)	Million £	**699.6**	**708.7**	**669.0**	**622.5**	**436.9**
	Storage cavern capital (TPC)	Million £	0	0	0	0	0
	Power island capital (TPC)	Million £	268.4	268.4	268.4	268.4	268.4
	Total scheme capital (TPC)	Million £	968	977.1	937.4	890.9	705.3
	Capital Intensity	Million £/MWe	2.85	2.87	2.65	2.66	2.02
	Hydrogen production total opex	Million £/yr	**105.3**	**128.1**	**129.3**	**149.3**	**140.3**
	Storage cavern total opex	Million £/yr	0	0	0	0	0
	Power island total opex	Million £/yr	18.2	18.2	18.2	18.2	18.2
	Total scheme opex	Million £/yr	123.5	146.3	147.5	167.5	158.5
	Opex intensity	Million £/yr/MWe	0.36	0.43	0.42	0.50	0.45
Other							
	Execution schedule duration	months	64	64	64	60	60
	Approximate lifetime	years	>30	>30	>30	>30	>30
	Availability	%	85%	85%	92%	92%	90%
Simplified LCO Estimates							
	Import Electricity Price (Note 1)	Million £/MWh	72.00	72.00	72.00	72.00	72.00
	Cost of Import Electricity to H2 Units	Million £/yr	26.65	25.77	19.69	32.25	1.37
	Cost of Import Electricity Total	Million £/yr	27.87	27.00	21.05	33.61	2.70
	CO_2 Price	£/te	10	10	10	10	10
	Income from CO_2 Export	Million £/yr	5.9	6.0	3.5	4.1	3.9
	Project Life	years	30.0	30.0	30.0	30.0	30.0
	Discount Rate	%	10%	10%	10%	10%	10%
	Levelised Cost of Hydrogen	£/MWth	40.38	44.98	40.36	45.36	35.06
	Levelised Cost of Electricity Export	£/MWh	85.18	93.01	84.24	92.09	83.26

Note 1: Data from DECC, 2012.

1.3.3 Characterisation of basic design requirements for a cost effective hydrogen store

Hydrogen Production Requirements

By using a number of trains, the syngas production plant can produce any quantity of hydrogen the project requires, with a sensible minimum size of syngas plant equivalent to an electrical generating capacity of around 175 MWe, corresponding to continuous generation of hydrogen in the syngas plant and operation of one GE Frame 9 gas turbine for twelve hours per day.

Hydrogen Consumption Requirements

Hydrogen usage rates, assuming the only hydrogen user is the power island, will be determined by the operating pattern assumed for the power plant. This, in turn, will be determined by the pattern of power demanded by the National Grid.

The vast majority of the burden of intermittent operation currently falls to CCGT plants. With significantly more wind capacity expected to come on-line in the near future, these plants will have to cope with the intermittency of the grid demand compounded by increasingly intermittent supply. The total reserve capacity required by the grid is expected to increase by approximately 30% by 2030. This means that the demand for flexible power generation should be significantly higher in 2030.

Total power demand in 2012 varied from a maximum of 55.5 GW to a minimum of 17.9 GW, while the average daily demand ranged from 26.5 GW in summer to 45.9 GW in winter (173% increase).

If we consider a plant with 2 gas turbines, in order for flexible operation to be applied to respond to seasonal variations in power demand the following annual mode of operation could be considered:

- Both GTs off for 4 months in summer
- 1 GT running in autumn/spring
- 2 GTs running 3 months in winter

In order to respond to the weekend drop in power demand, the plant could be designed to operate with 2 GTs during the week and one GT at the weekend, sizing the syngas production to produce sufficient fuel to support this operating pattern.

A similar variation between maximum and minimum applies to the daily variation in 2012. The shape of the demand curve shows that it would be sensible for the plant to be operating with 2 GTs at full load for 18 hours, from 06:00 until 24:00 or perhaps 15 hours from 06:30 until 21:30 then reduce to either 1 GT or both GTs on minimum stable generation (MSG) during the night.

The daily variation scenario would result in the lowest volume syngas store, however the following practicalities of diurnal operation would need to be considered:

- Would the low plant efficiency during the night (e.g. MSG) have a severe effect on the overall economics?
- Would the plant be off during the night, and what is the efficiency impact of start-up and shutdown on a daily basis?
- Is the GT life or maintenance impacted by ramping up and down frequently?

It is known that a natural gas fired CCGT can provide very flexible operation since they currently perform the vast majority of the load balancing for the UK grid.

Matching Power Supply and Demand

In order to maintain the frequency of the UK electricity grid at 50Hz, electricity supply must match electricity demand closely at all times. In case of a sudden drop in electricity production, National Grid has a number of contingencies which are able to respond to stabilise the grid until a replacement large unit can come online. These contingencies are designed to cope with the failure of one of the two largest single electricity sources on the UK grid.

Excluding trips, National Grid can forecast with some accuracy the likely power demand at any time. Operators can plan to have their most cost-effective plants running first, then bring more expensive to operate plant on-line as demand increases. Although demand varies considerably on an hourly basis, it is not unpredictable. It should therefore be reasonable to expect a fairly regular pattern of operation for a low-carbon plant with hydrogen storage and flexible turbine system.

Storage Volume Requirements

Approximate syngas storage volumes can be calculated based upon flow rate requirements for a base load gasification plant. Gas turbines operate most efficiently at full load, so in order to calculate a syngas consumption rate it is sensible to assume full load of a single machine.

For cases 1, 2 & 4, dilution nitrogen will be produced by a continuous process and therefore requires storage, similar in volume to the hydrogen storage, in order to supply an intermittently operating power island. For case 3 the hydrogen–nitrogen generated by the ATR unit is stored as a mixture. Considering the approximate flow rates it is possible to calculate approximate storage volumes (per GT) required for a number of different operating scenarios.

The stability requirement of the salt cavern puts limitations on the shape, storage volume and the maximum and minimum operating pressures. Maximum pressure should be typically 0.7-0.8 of lithostatic pressure so that the pressures do not exceed a safe limit below the fracture pressure of the halite. Storage pressure and percentage of cushion gas in the storage site are two major factors which will dictate the size of the salt cavern. Minimum pressure should be maintained above 0.3 of lithostatic pressure with a volume of stored gas used as a cushion to prevent instability and limit salt creep.

In this study, we are looking into several diurnal variations which give the smallest cavern sizes, as well as other operating regimes up to seasonal variations. For so-called “fast churn” facilities, the working volume available for withdrawal per day is a maximum of about 10% of the total stored gas. For other operating modes, the cushion gas requirement will vary in the range of 40 to 80 % depending on the type of gas stored. For this study, it has been assumed that only 10% of the total stored gas is available to withdraw per day in diurnal operating modes and an average of 60% cushion gas is required for seasonal operational modes. For weekly operational mode assuming daily withdrawal over 5 days, the cushion gas requirement has been calculated at 67% of the total stored gas; so that on 5th day, it will be 10% withdrawal of total stored gas in cavern.

Gas working volume (using proven storage pressures from Table 14) and total storage volume for salt caverns at various operating pressures have been calculated and summarised in Table 15, Table 16, Table 17 & Table 18.

Salt cavern operating temperature is also an important operating factor in order to preserve the structural integrity of the storage cavities. For this study, a cavern operating temperature of 45°C has been assumed for all cases.

To accommodate the seasonal cases, several caverns will be required if gas is being stored at low storage pressure. As the storage pressure increases, the storage volume decreases substantially. So at higher storage pressures, the project will be more flexible to accommodate weekly variation using a single cavern and seasonal variations with lesser number of multiple caverns. Overall cavern construction costs are not expected to vary proportionally with cavern size, so there should be significant economies of scale associated with projects using larger (or multiple) caverns.

It would seem sensible to build a bigger storage facility at the highest storage pressure within the lithostatic pressure limit required to maintain structural integrity. This will increase the flexibility of the power plant facility to operate in a wider variety of patterns to maximise revenue.

Other Considerations

Size is only one requirement for a salt cavern to be a well functioning, safe and economic gas store. Other important considerations include:

- Safe containment, Structural Stability and Salt Creep

  When storing hydrogen, safety is a big concern, as the gas is highly flammable, explosive and difficult to contain because of the very small dimensions of hydrogen molecules. Suitable storage pressures need to be selected to make the cavern safe and economical.

  Once a salt cavern has been created, its gas tightness has to be maintained and to ensure containment of the stored gas, a periodic mechanical integrity test should be performed as well as underground infrastructure should be monitored.

- Contaminants

  It is understood that in the first months of operation of the salt caverns it is likely that the gas leaving the caverns will be wetter than gas entering the caverns. This is due to residual water being present in the cavern as a result of the solution mining process. It is also anticipated that this additional water content will be saturated with salt.

  Other contaminants may be possible due to impurities present in the salt, and these will be specific to any given halite formation, and will need to be considered on a location specific basis.

- Project / Storage Location

  The geographical location of a salt cavern is also an important consideration, as a project will generally be more viable if it does not have to involve significant transportation of feedstocks, products, wastes or intermediates. Items to consider include proximity to:

  - a location for brine sale or disposal at sea (within a reasonable distance);
  - fuel source or fuel transportation infrastructure
    - national gas grid connection;
    - coal / biomass import terminal;
  - electricity grid connection;
  - CO2 storage location, or connection to a CO2 transportation hub;
  - source of cooling water - sea water is preferred for maximum power plant efficiency.

1.3.4 Requirements / options for hydrogen turbines

High Hydrogen Fired Gas Turbines

The availability of large gas turbines capable of firing undiluted hydrogen is some way into the future. Existing types of gas turbines with power outputs >300MWe from GE and Mitsubishi require significant dilution of the hydrogen-rich fuel gas with nitrogen or perhaps steam. This requirement for dilution of hydrogen fuel is mainly due to the need to limit NOx emissions.

The modelling of hydrogen production and firing for the study is based on information provided by GE for the Frame 9F Syngas variant, giving a thermal efficiency of 34.4% using nitrogen as the diluent.

Mitsubishi have provided an estimate of performance for their 701F gas turbine. Overall plant power output and thermal efficiency (34.9%) are marginally higher than the information from GE, but with higher NO_X at < 100 vpmd, which may require provision of a SCR unit downstream in the HRSG – a feature which may become more generally needed with the move towards higher firing temperatures with hydrogen fuel.

Gas turbine efficiency will be comparable or even slightly improved with steam dilution of the fuel, with the.lower firing temperatures needed to offset perceived higher gas turbine maintenance costs being largely offset by the elimination of nitrogen compression. Initial indications are that plant thermal efficiency with gasification may be 0.5% higher than with nitrogen dilution, without taking into account any de-rating of the gas turbine. Less attractive features of steam dilution include high water consumption (the stack gas contains approx 18 -20 % water vapour), increased tendency to a visible exhaust plume, perceived higher maintenance costs for the gas turbine and perhaps the HRSG also.

Hydrogen Power Recovery Turbines (Expanders)

Higher storage pressures may make attractive the recovery of power by expansion of stored gas in an expansion turbine.

Foster Wheeler has received two budget proposals from GE Nuovo Pignone.

- For an expansion turbine to let down 93 kg/s of H_2/N_2 mixture from 50 bar to 33 bar, with 89% isentropic efficiency. This offer included a 4500 kW coupled electric generator. The scope of GE's supply comprises the radial-flow expansion turbine itself, integral reducing gear and a 4500 kW asynchronous generator, together with lubricating oil console and local controls.
- For a much larger expander, required to expand the hydrogen fuel plus nitrogen diluent from 253 bar to 30 bar, with a power developed of over 30MW .GE's proposed solution provides for six stages of expansion in two separate integrally geared expander-generator machine groups. While this type of expander technology is well proven, a judgement would have to be made at some point on the economic benefit and complexity of future gas - expansion schemes.

Moreover if steam is substituted for nitrogen as the gas turbine fuel diluent, the resulting reduction in molecular weight of the stored fuel gas will make mechanical energy recovery significantly more difficult to achieve with existing types of expander.

1.3.5 Economics of hydrogen pipelines

The investment cost of a pipeline includes materials, labour, right-of-way fees and miscellaneous costs. The cost of the pipeline material is then determined by the pipeline length, diameter, wall thickness and material of construction.

For the case of transporting gaseous hydrogen it is necessary to consider hydrogen embrittlement in the material selection and wall thickness. The wall thickness and diameter are then affected by the operating pressure (or pressure range) and the flow rate of gas.

In this instance, two design cases were considered to calculate order of magnitude costs for a range of pipe sizes (6” to 60”) to cover all anticipated conditions in the system:

- hydrogen at ambient temperature (design temperature) and 10 bar (design pressure);
- hydrogen at 120°C and 200 bar;

It is assumed that water is present only in the vapour phase which, along with the addition of a corrosion inhibitor, allows carbon steel with a 3 mm corrosion allowance to be used.

Summaries of costs/km for various diameters of onshore hydrogen pipeline are provided in Attachment 22.

The estimates represent an indicative cost in a generic NE England location for the supply and below ground installation of the lines. No allowance is included for any pipe crossings or bridges, which can be a significant contributor to TIC.

Brine Pipeline Costs

Another significant cost which is not captured elsewhere is the brine pipeline cost. Depending on the proximity of the proposed plant and cavern location to the sea, brine pipelines may be required to transport seawater for cooling and/or brine for solution mining.

Order of magnitude costs per km have been derived for carbon steel pipelines transporting de-aerated brine at 40°C and 20bar, with various diameters (from 6” to 60”):

Summaries of costs/km for various diameters of onshore brine pipeline are provided in Attachment 23.

Offshore Pipeline Costs

If an offshore option of the hydrogen pipeline is selected/favoured, the submarine pipeline has to be strong enough to withstand all the loads that will be applied to it both during its construction and when in operation. When it goes into operation, it must withstand the internal pressure from the hydrogen it is carrying, external pressure from the sea (water depth expected to be less than 50 meters), and stresses induced by temperature changes. In addition, the submarine pipeline needs be able to withstand external impact loads from anchors and fishing gear.

In general, offshore pipeline costs are considerably greater than those for onshore pipelines, and are dominated by installation costs, as the rental of the barges and crew are expensive, and the laying rate and availability of pipelaying barges is critical. This means the correlation between diameter and cost is less strong.

1.3.6 Effects of hydrogen purity

Hydrogen Production

The level of hydrogen purity produced in the proposed hydrogen generation processes varies. The purest hydrogen stream is produced from Case 4 (SMR) – 100 mol% hydrogen followed by Cases 1 and 2 (Gasification) – 89 mol% hydrogen. The least pure stream is produced in Case 3 (ATR) – 53 mol% hydrogen.

There are some limitations on the commercial availability of machines to compress pure, or nearly pure, hydrogen to very high pressures (circa 285 bara). For Case 4 SMR, producing pure hydrogen from the PSA unit, it appears feasible in the medium term to compress the hydrogen product to around 100 bar in a centrifugal compressor, but a reciprocating compressor will probably be needed for the final stage of compression from 100 to 285 bara.

For the gasifier-based cases (Cases 1 & 2), in which the produced hydrogen is impure and has a molecular weight around 5.5, it can be assumed that compression to 285 bara can be achieved in a centrifugal compressor using existing technology.

The purity of hydrogen produced by each technology case discussed should not be significant for the correct function of downstream equipment, as the storage and combustion processes are flexible. The volume of dilution gas can generally be varied during operation depending on hydrogen purity to achieve the same input composition to the gas turbine.

If the hydrogen syngas were to be used for applications other than for combustion for power generation, complications arise. Consequently, the high hydrogen syngas produced from every design case excepting Case 4 would require further purification, e.g. by pressure swing adsorption (PSA), before it could be used for any of the highlighted alternative technologies.

Hydrogen Storage

The salt cavern will contain residual brine from the solution mining process, which for an initial period of operation will saturate the stored gas and exit along with the gas when the cavern is evacuated along with traces of salt. Chloride is the usual problem with the brine/ salt caverns as it will corrode the pipe work, so suitable provision needs to be made for this.

When the gas contains CO2, it will dissolve in the water to form carbonic acid. Carbonic acid will not affect the salt cavern, but this saturated, acidic, corrosive gas output stream can cause downstream complications unless protective measures are put in place. The presence of carbonic acid will affect the material of construction of the well head, downstream pipework and associated processing equipment.

Water should only be present in the outlet gas for an initial period whilst the newly excavated cavern is drying out after the solution mining process. Eventually, the cavern will dry out and the level of water in the outlet gas will be small.

Hydrogen Combustion

As the mole fraction of syngas impurities increases, the calorific value of the gas mixture decreases. A decrease in syngas calorific value will upset the ratio of hydrogen to dilution gas, which is set to achieve 25 ppmvd NO_x at the maximum allowable gas turbine firing rate.

The composition of the hydrogen-rich fuel itself pre-dilution can vary to a limited extent, as the gas turbine inlet conditions can be controlled by varying the flow of dilution gas. The currently available high-hydrogen GTs are designed to burn a high hydrogen syngas/dilution gas mixture covering a wide composition range, and are also capable of starting up and running on natural gas whenever syngas is unavailable.

The presence of certain impurities also causes other complications. If the high hydrogen syngas extracted from the salt cavern is highly water/brine-saturated, traces of carbonic acid and salt can degrade the working surfaces of the gas burners and turbine blades over time, resulting in a decrease in system efficiency and the requirement for more regular maintenance/replacement.

During the design of such a system, the working limits in terms of hydrogen purity and impurity levels should have a clearly defined range and operation conditions should closely resemble these conditions to achieve the optimum system efficiency and power output.

HSE Impact

Hydrogen purity variation generally does not introduce any additional HSE issues in terms of hydrogen production as this part of the process is a closed system. The main HSE impact arises from the processing of toxic impurities, leading to release of toxic gas to atmosphere. This is unlikely under normal operation conditions and many typical impurities will be oxidised to a harmless form in the turbines, however, certain toxic contaminants when accidently entrained into the fuel gas stream (for example H_2S) can be oxidised to more harmful forms (SO2).

There are also complications associated with the alternative end-uses of hydrogen, in particular with open system use in a domestic environment. End-use applications in the home, safely modified for pure hydrogen fired operation, are perfectly acceptable from a HSE stand point. However if certain toxic impurities such as CO are present in the gas stream, an HSE risk is introduced. These aspects are covered in more detail in the WP3 report.

2. INTRODUCTION

The Energy Technologies Institute (ETI) is a public private partnership between global industry members - BP, Caterpillar, EDF, E.ON, Rolls-Royce and Shell with the UK government. The ETI brings together projects that accelerate the development of affordable, clean, secure technologies needed to help the UK meet its' legally binding 2050 targets. The ETI's mission is to accelerate the development, demonstration and eventual commercial deployment of a focused portfolio of energy technologies, which will increase energy efficiency, reduce greenhouse gas emissions and help achieve energy and climate change goals.

The ETI's modelling, using its Energy System Modelling Environment ("ESME") shows that flexible power generation systems comprising hydrogen generation with Carbon Capture and Storage ("CCS"), intermediate storage (particularly using salt caverns) and flexible turbines are attractive components in a future UK Energy system. In such a system, hydrogen is supplied from coal and biomass fired gasifiers and steam methane reformers, with carbon dioxide ("CO_2") captured for storage. This permits the use at high load of capital intensive and relatively inflexible conversion and CCS equipment, filling hydrogen storage when power is not needed, and releasing hydrogen at short notice through turbines when power is at a premium. Superficially there are no barriers to using salt caverns as stores; as such stores are in use in the USA. However, these are for high value added applications and not for use in power where loss of efficiency is a more serious drawback. The ETI currently lacks sufficient data and knowledge to build a good representation of costs or efficiency (particularly relating to hydrogen storage) in ESME.

The purpose and focus of this project is:

- To improve the ETI's understanding of the economics of flexible power generation systems comprising hydrogen production (with CCS), intermediate hydrogen storage (e.g. in salt caverns) and flexible turbines; and
- To focus on the potential, economics and technical requirements for salt cavern storage and flexible turbines, to enable refinement of the ETI Energy System Modelling Environment (ESME) model in order to confirm or adjust ESME findings.

2.1 Scope of Study

The Hydrogen Storage and Flexible Turbine Systems Project is split into five work packages. The first three work packages (WP1, WP2 & WP3) are focused on data collection and research in order to derive a basis for techno-economic analysis in WP4. Using the output from the WP4 modelling, a representative system will be selected. In WP5, this representative system will be compared against a post combustion CCGT case:

- WP1 – Hydrogen Power Production;
- WP2 – Hydrogen Storage;
- WP3 – Supporting Studies;
- WP4 – Development of a Flexible Modelling Tool;
- WP5 – Identification of a Representative System and Comparison of CCGT with CO_2 Buffer Storage.

This report covers the work undertaken in the execution of WP1 – Hydrogen Power Production.

2.2 Scope of WP1 – Hydrogen Power Production

The activity in Work Package 1 is intended to characterise the requirements for storage for “power scale” use, in terms of offering tactical (diurnal) or strategic levels of storage (and what that means in terms of a delivery pattern of hydrogen) and to estimate what storage pressures are of interest.

The aim of WP1 is to review the technical options and economics for hydrogen production with CCS and hydrogen fired turbine power generation, against the likely energy system requirements of 2030.

The scope of WP1 consists of:

- A review of options for hydrogen production;
- Techno-economic definition of the following four options for hydrogen production:
    - gasification of coal;
    - gasification of a coal/biomass mix;
    - autothermal reforming of natural gas; and
    - steam reforming of natural gas;
- Characterisation of basic design requirements for cost effective hydrogen stores;
- A review of requirements / options for hydrogen turbines, being:
    - Power generation gas turbines;
    - Let-down (expansion) turbines;
- A brief study of the economics of hydrogen pipelines;
- A brief discussion of the effects of hydrogen purity.

The WP1 report forms a part of the Final Report deliverables for the Hydrogen Storage and Flexible Turbine Systems Project.

3. OPTIONS FOR HYDROGEN PRODUCTION

3.1 Objective

The objective of this section of WP1 is provide a high level screening review of the potential options for hydrogen production, in order to derive a basis for the four cases selected for techno-economic definition.

3.2 Discussion of Technologies and Option Screening

There are a number of well established hydrogen production schemes suitable for producing a low carbon, or carbon free, fuel gas at a scale capable of supplying a commercial gas turbine combined cycle power plant. These include:

- Gasification of coal, oil, petcoke & biomass,
- Reforming of natural gas via partial oxidation, autothermal reforming (either using oxygen, air or a mixture) or steam methane reforming.

Very early screening and recent Foster Wheeler experience concluded that gas turbines (GTs) currently available for firing a low/no carbon hydrogen-rich fuel gas require dilution of the fuel gas with nitrogen and/or steam, primarily to limit exhaust NO_x content. As most of the hydrogen production processes listed above require an Air Separation Unit (ASU) for oxygen production, resulting in simultaneous “free” nitrogen production, nitrogen is the generally assumed dilution option.

Provision of nitrogen to the GT will be required on the same non-continuous basis as the low carbon fuel gas. This presents a challenge since ASU’s are currently all distillation based and require continuous operation, and where oxygen is needed for the syngas production process, this must also be supplied continuously. Thus the dilution nitrogen produced in the ASU will require storage, similar in volume to the hydrogen store, during the time that the power plant is not operating. This can be achieved by:

- Liquid nitrogen (LIN) storage with an evaporator
- Gaseous nitrogen storage in an adjacent salt cavern
- Gaseous nitrogen co-storage with the low carbon fuel

LIN storage was eliminated from consideration since the operation of an evaporator to meet the expected continuously varying power plant load demand may be problematic, and there will be a high energy penalty associated with the liquefaction and subsequent vaporisation.

A significant advantage of using an ATR process stems from an air-blown ATR not needing an ASU, which has both significant cost and operational inflexibility associated with it. Since the air-blown ATR case naturally results in a premixed fuel gas product containing almost exactly the required nitrogen to hydrogen mixture required for the GT is was decided to assume co-storage for this case. This avoids the energy and capital costs that would be required to separate the two gases, would be easier to operate, and provides a co-storage reference case.

Most existing ATR plants use oxygen, however, but development work carried out by FW in a Polish fertiliser plant in the1990's showed that air-based ATR is feasible, and two subsequent natural gas based CCS projects developed by FW were based on air as the selected oxidant. A preliminary comparison of air versus oxygen for a natural gas based ATR operation showed almost identical thermal efficiency of the hydrogen production unit. Therefore since it was considered that an oxygen-blown

ATR would not benefit from the above advantages, it would be less favourable than the air-blown ATR case. It is possible, however, that once the economics of storing nitrogen have been included an oxygen blown case with steam dilution in the GT may look more attractive.

In order to provide a storage method comparison, and preserve purity of the hydrogen-rich gas and nitrogen where possible, it was decided to assume separate nitrogen and hydrogen storage for all other cases.

Most coal gasifiers use oxygen, although Mitsubishi has developed an air based gasification process and are running a >200 MWe IGCC unit in Japan. According to their calculations, air gasification is more efficient and less expensive than oxygen in a normal IGCC project. However, as with the ATR discussed above, the elimination of nitrogen from the downstream storage may favour oxygen gasification in the future.

The Gasification Cases will need an air separation unit (ASU) to provide the oxygen required by the gasifier which will also be used as a source of nitrogen. An Australian steam coal, from the Basis of Design for the original gasification study for the ETI by Foster Wheeler, was considered to be the most appropriate base case gasifier fuel.

A case including some biomass firing was also desired, so a coal & biomass based gasification case was also selected. To enable best comparison with the coal only case, these two cases were kept as similar as possible.

Petcoke and oil firing would not be economically sensible options in the UK. Heavy oils are generally upgraded to high value products or sold as bunker fuel, both of which are more lucrative uses than sale for combustion for power generation. Very little oil is used in the UK for power generation, and then only on sites which have oil available without access to cheaper fuels such as coal or natural gas. The UK only has one coker, which makes anode grade coke which is also far more valuable than either coal or natural gas. In northern Europe there are very few cokers due to the generally lighter nature of crudes processed, so petcoke would have to be imported from some distance.

The final case selected is to be based on steam methane reforming (SMR), rather than partial oxidation, due to the prevalence of this method of hydrogen production in existing plants, and to prove or disprove the expectation that this case would not be very suitable for hydrogen-based power production. The SMR plant was expected to be lower in capital cost but more challenging in general due to complexity of flue gas decarbonisation and lack of nitrogen production as a process by-product.

A standalone ASU would be required for the SMR case in order to provide the nitrogen required for the GT, with the by-product oxygen being vented. If the oxygen were to be considered for onward sale, the ASU would have to be designed to produce high purity oxygen, which would require additional CAPEX and operating power requirement.

A case using 100% steam instead of nitrogen dilution for the GT is possible. A sub-case (4b) was added in which there is no ASU and the GT uses only steam for dilution, using steam generated within the power island.

The scale of the plant is another variable to be considered as a number of trains of the hydrogen production plant and of the power island can be used. It was agreed (for WP1) that all cases would be based on the net power production and heat rate of one GE Frame 9 (Syngas) gas turbine, based on a performance prediction provided by GE for a broadly similar application. This will then be used to scale the capacity and estimate the performance of the overall project configurations within the WP4 modelling.

4. TECHNO-ECONOMIC DEFINITION OF THE SELECTED HYDROGEN PRODUCTION OPTIONS

4.1 Objective

The objective of this section is to perform an outline techno-economic definition of the following options for hydrogen production:

- Case 1 - gasification of coal,
- Case 2 - gasification of a coal/biomass mix,
- Case 3 - air-fed autothermal reforming of natural gas, and
- Case 4 - steam reforming of natural gas,
    - Case 4b - steam reforming of natural gas with steam to GT.

Data will be reported for each of the above cases so that their suitability and performance for power generation with CCS can be compared. This data will include block flow diagrams, process descriptions, outline heat and mass balances, outline utility summaries, comparative project execution schedules, capital and operating cost estimates, unit lifetimes and availability. A qualitative conclusion of the suitability of each technology for the proposed application will also be provided.

4.2 Basis

The purpose of this task in WP1 is to provide a high-level technical and economic comparison of four methods of hydrogen rich fuel gas production. For a fair comparison between technologies, the operating pattern of the plant must therefore be the same for all cases.

For the purposes of calculating overall plant efficiency, and other technical data, a steady state basis is also required. For technology comparison within WP1 only, it has been assumed that the hydrogen production plant operates continuously (i.e. 24 h/day, 365 days/y), and is sized at the capacity required to supply gas to a single gas turbine (GE Frame 9F) operating continuously at full load.

For the overall scheme (including hydrogen production plant, salt cavern storage facility and power island) to be assessed, the above hydrogen plant sizing basis will need to be scaled based on the total number of gas turbines and the operating regime being considered. This will be considered in WP4 modelling.

In all cases, except Case 3 (the air-blown ATR case), >85 mol% purity hydrogen can be drawn from the hydrogen production unit for storage or direct firing in the gas turbine.

If industrial users of hydrogen were being considered (for example on an oil refinery), a PSA unit would typically be required for final purification of the exported hydrogen. In Case 3 a similar arrangement could be used, although the capacity of the PSA unit would be much larger due to the need to reject nitrogen.

It has been assumed that the three main plant elements (hydrogen production, gas storage caverns and the hydrogen-fired “power island”) are all located at a common site in a generic UK location with minimal lengths of interconnecting pipe, and no heat integration. In reality there is no particular objection to the three elements being remote from each other, for example a coal gasifier at a mine-head, gas storage where caverns can be produced, and the power island close to an industrial area. In such circumstances, allowances would be necessary for the cost of interconnecting pipelines and resulting pressure losses.

To ensure a fair comparison between the cases, it was assumed that the hydrogen production plant delivery pressure is constant and at the same pressure as that required for the gas turbine 33.4 bar (abs).

Additional design basis data, such as feedstock compositions, climatic data and utility assumptions can be found in the Basis of Design, Attachment 1.

4.3 Case 1 - Hydrogen Production via Gasification of Coal

4.3.1 Introduction

Refer to Figure 1 – BFD for Case 1 (page 33)

This route to hydrogen production, incorporated into Integrated Gasification Combined Cycle schemes, is one of the most widely studied routes for power production from coal with carbon capture.

The process typically comprises coal milling and drying, gasification to form a synthesis gas (syngas), made up largely of hydrogen and carbon monoxide, shifting the syngas with steam to produce additional hydrogen and to convert the carbon monoxide to carbon dioxide (CO_2), heat recovery and separation of the syngas into a hydrogen rich stream, a CO_2 rich stream and an H_2S rich stream. Sulphur is then recovered from the H_2S rich stream and the remaining components recycled to the syngas. The hydrogen rich stream is compressed and sent to the power island or to underground storage. The CO_2 stream is dried, compressed and exported for sequestration.

4.3.2 Process Description

The following process description relates to a coal gasification process based on entrained-flow dry-feed gasification processes such as the Shell process.

Coal Milling and Drying

Raw coal as-received is milled and dried to a particle size typically <100 µm and water content <2%. Fluxant is added to adjust the ash melting temperature of the coal.

The pulverized coal and flux from the milling section is pressurized with high pressure nitrogen and fed into an entrained flow gasifier, in which it is gasified with oxygen.

Air Separation Unit (ASU)

The gasification oxygen is produced in an ASU which also provides nitrogen for use as fuel diluent in the power island downstream. The oxygen is produced by the ASU at a pressure of around 50 bar, which is higher than the gasifier pressure, while the nitrogen is produced at low pressure. A multistage compressor is used to compress the nitrogen to the pressure required for the nitrogen storage cavern, nominally at 33 bar (abs).

Gasification and Syngas Cooling

The dry pulverised coal is gasified using oxygen with moderating steam to produce a raw synthesis gas (syngas) containing mainly CO and hydrogen.

The entrained flow gasifier is a membrane wall reactor installed inside a pressure vessel. In the membrane wall, absorbed heat is used to produce saturated MP steam. The operating temperature of the gasifier zone is about 1500-1600°C. At this temperature, ash from the coal is converted into molten slag, which runs down the gasifier walls to the slag removal zone, where it is contacted with water and solidifies. The operating pressure of the gasifier is 41 bar (abs).

Hot syngas from the gasifier is initially quenched with recycle syngas to approximately 800°C. The combined gas stream then enters a heat recovery steam generator comprising an HP steam superheater, an HP steam generator and an MP steam generator. The gas leaves the heat recovery steam generator duct at 270°C. The cooled gas then flows to a candle- type filter which removes most of the entrained solids. Finally the filtered gas is scrubbed with water in a wash column which removes the remaining fly ash as a slurry in the recirculating water, along with organic acids and ammonia.

Shift Conversion Unit

The cooled scrubbed syngas from the gasification unit at 145°C flows to the Shift Conversion Unit where CO and steam present in the syngas are converted to CO_2 and H_2. In addition, COS and HCN in the syngas are also hydrolysed.

There are three CO shift reactors in series. The syngas is first mixed with additional superheated MP steam and then preheated in the shift interchanger, recovering heat from the syngas leaving the third shift reactor. The syngas/MP mixture enters the first shift reactor at 270°C. The outlet stream at 500°C is used to generate HP and MP steam before entering the second shift reactor at 275°C. The gas leaving the second reactor at 318°C is used to generate additional MP steam, before entering the third reactor at 265°C. The gas leaving the third reactor at 272°C is used to preheat the shift feed as described above and to provide reboil heat for the downstream acid gas removal.

Acid Gas Removal Unit

The AGR Unit removes the H_2S and CO_2 from the shifted syngas by washing with a solvent (DEPG) in order to produce a hydrogen-rich fuel gas.

Rich solvent from the absorber is flashed in two stages in order to recover the dissolved CO_2 at two different pressures. The rich solvent from high pressure flash is routed to the H_2S Stripper where it is heated by LP steam to produce an H_2S-rich stream overhead product. This H_2S-rich stream is sent to a Claus-type Sulphur Recovery Unit (SRU). The low-pressure flash releases most of the dissolved CO_2, which is routed to the CO_2 Dehydration/Compression Unit.

After take-off of a small fraction (approx 8%) of the decarbonised hydrogen + nitrogen fuel gas product for firing of the HP and MP steam superheater, the remainder of the product is delivered to the hydrogen storage cavern without compression at a nominal pressure of 33 bar (abs).

Sulphur Recovery Unit

The H_2S-rich acid gas from the AGR unit is treated in the Sulphur Recovery Unit (SRU) where H_2S in the acid gas is converted into elemental liquid sulphur.

The SRU comprises a thermal oxidation stage followed by two catalytic stages with elemental sulphur being removed between the stages by condensation. The tail gas from the SRU is hydrogenated to convert sulphur components into H_2S. After hydrogenation, the tail gas is quenched with process water, compressed and recycled back to the inlet of AGR absorber.

CO_2 Purification and Compression Unit

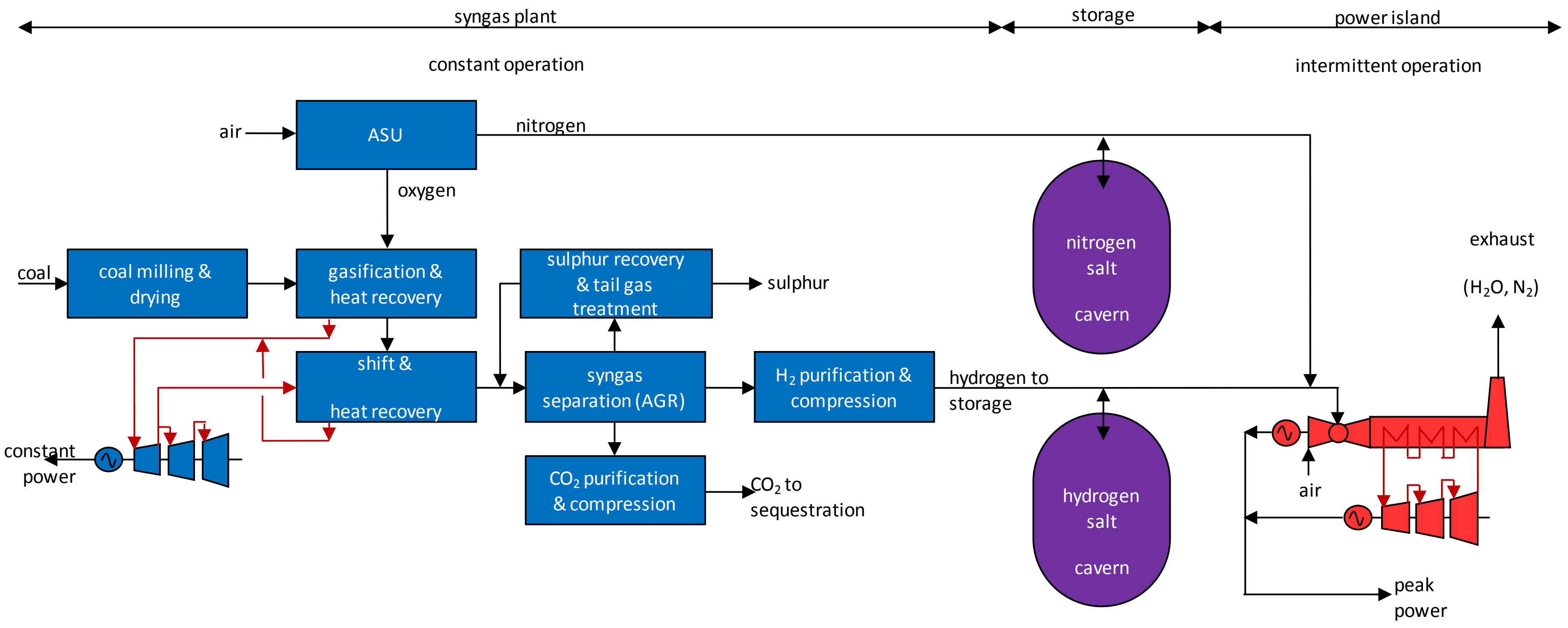
The CO_2 streams from AGR unit flow to the CO_2 Dehydration/Compression Unit where the CO_2 product is initially compressed to 34 bar (abs) and dried to < 50 vppm water using a molecular sieve adsorption process, before being further compressed to 151 bar (abs) for export to sequestration.

Steam System

125 bar, 530°C steam (HP) and 45 bar, 530°C steam (MP) is generated in the gasification and shift units and superheated as described above in the fired steam superheater. These steam flows supply a condensing steam turbine generating electric power for internal plant use. About 40% of the inlet steam is extracted at 9 bar for miscellaneous heating duties.

The high level technical performance of power generation based upon this method of hydrogen production can be understood by studying the utility summary results in Attachment 3. The highlights from the utility summary and material balance are compared against the other methods of hydrogen production in Section 4.7.

Figure 1 – BFD for Case 1 – Coal-fed Gasification with CCS and separate H_2 / N_2 storage



The Heat & Material Balance for this scheme is provided in Attachment 2.

The Utility Summary for this scheme is provided in Attachment 3.

4.4 Case 2 - Hydrogen Production via Gasification of a Coal / Biomass Mix

4.4.1 Introduction

Refer to Figure 2 – BFD for Case 2 (page 35)

This route to hydrogen production is similar to the coal-only route described in Section 4.3, above with additional requirements for feedstock receiving and storage upstream of the gasifier.

The biomass specified for this case is wood pellets, with characteristics described in the Basis of Design (see Attachment 1). The process comprises coal and biomass co-milling and drying of, gasification to form a raw synthesis gas (syngas), shifting the syngas with additional steam to produce additional hydrogen and convert the carbon monoxide to carbon dioxide (CO_2), heat recovery and separation of the syngas into a hydrogen rich stream, a CO_2 rich stream and an H_2S rich stream. Sulphur is then recovered from the H_2S rich stream and the remaining components are recycled to the syngas. The hydrogen rich stream is then compressed and sent to the power island or to underground storage. The CO_2 stream is dried, compressed and exported for sequestration.

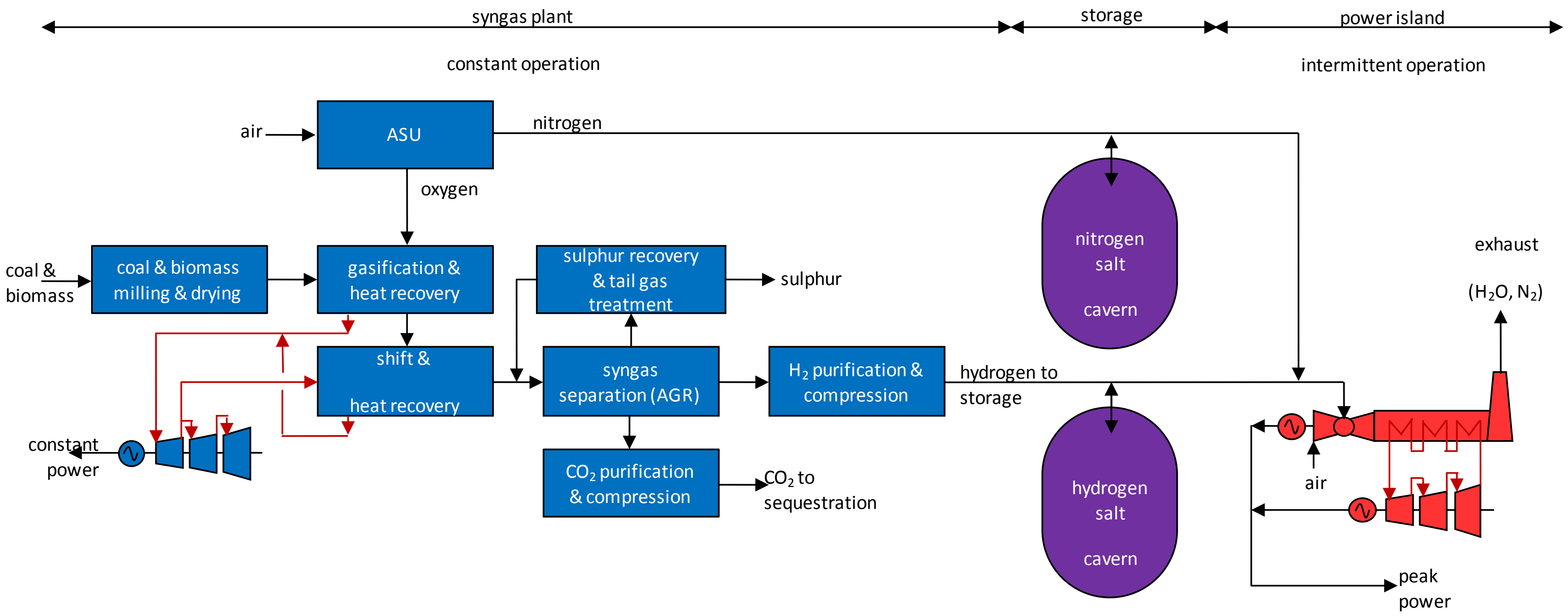
4.4.2 Process Description

As specified by the ETI, 22% of the lower heating value (LHV) of the total coal/biomass feed is made up of wood pellet biomass in this case. Experience from operators shows that for this proportion of biomass of this nature there is no requirement for additional pre-treatment facilities. It is therefore assumed that the biomass is fed into the coal milling and drying system along with the coal.

From that point onwards, the process is identical to the coal-only flow scheme described in section 4.3.2, although there are some differences in capacities of some of the units, such as the sulphur recovery unit, and in the overall flow rates. These differences can be seen in the overall heat and material balance for this case.

The high level technical performance of power generation based upon this method of hydrogen production can be understood by studying the utility summary results in Attachment 5. The highlights from the utility summary and material balance are compared against the other methods of hydrogen production in Section 4.7.

Figure 2 – BFD for Case 2 – Coal and Biomass-fed Gasification with CCS and separate H_2 / N_2 storage



The Heat & Material Balance for this scheme is provided in Attachment 4.

The Utility Summary for this scheme is provided in Attachment 5.

4.5 Case 3 - Hydrogen Production via Autothermal Reforming of Natural Gas

4.5.1 Introduction

Refer to Figure 3 – BFD for Case 3 (page 38)

This route to hydrogen production involves the autothermal reforming process route usually applied for the generation of syngas and its products from natural gas, particularly for ammonia and methanol production. While most autothermal reformers use oxygen or oxygen-enriched air, in this case air is used as the ATR oxidant.

The process involves desulphurising the natural gas before mixing it with steam and feeding it to the autothermal reformer (ATR), in which the natural gas / steam mixture is reacted with preheated air and is converted to syngas. The ATR is operated with an outlet pressure of 28.5 bara. The syngas from the ATR is then shifted with the residual water vapour contained in the ATR product stream to produce additional hydrogen and convert the carbon monoxide to CO_2. After heat recovery the shifted syngas is separated using an amine-based acid gas removal unit (AGR) into a hydrogen/nitrogen stream and a CO_2 rich stream. The hydrogen and nitrogen rich stream is then compressed and sent to the power island or to underground storage. The CO_2 stream is dried, compressed and exported for sequestration.

4.5.2 Process description

Natural Gas Pre-Treating

Natural Gas is pre-treated in a hydro-desulphurisation reactor in which the feed gas is mixed with a small stream of recycled hydrogen and is preheated to 350°C. The gas is then passed through a Cobalt-Molybdenum catalyst which converts organic sulphur into H_2S and then through a zinc oxide H_2S adsorbent.

Autothermal Reforming and Syngas Production

Desulphurised feed from the pre-treating section is mixed with superheated steam at steam/carbon ratio 1.5, the combined feed is then heated to 550°C in the mixed feed preheater. The mixed feed then enters the ATR reactor, in which it reacts over a nickel catalyst with air, which has been preheated to 600°C in a fired preheater. The resulting syngas exiting the ATR at 28.5 bar (abs) and 950°C is first cooled to 650°C in the reformed gas boiler, generating steam at 128 bar (abs). The gas stream leaving the boiler is then divided with part flowing to the mixed feed preheater and part to a steam superheater, where the 128 bar (abs) steam is superheated to 510°C.

Shift Conversion Unit

The process gas then flows at 350°C to the Shift Conversion Unit, in which the CO content of the syngas reacts with residual steam to produce CO_2 plus an equal molar quantity of additional hydrogen. The shift reactions are carried out in two reactors in series. The first shift reactor contains an iron-based catalyst. The outlet gas from this reactor, at around 425°C, is first cooled in a steam superheater and then in a boiler feed water (BFW) heater, before entering the second shift reactor, containing copper/zinc catalyst, at 200°C. The shifted gas leaving the second reactor, at 235°C, is used to heat BFW and then to provide reboil heat for the downstream amine CO_2 removal.

Acid Gas Removal Unit

The AGR Unit removes the CO_2 from the shifted syngas by washing with an amine solvent (BASF aMDEA®) in order to produce a substantially carbon-free hydrogen-rich fuel gas.

The heat required for regeneration of the rich solvent is partly provided by cooling the shifted syngas and partly by low pressure steam.

After deduction of a small fraction (7%) of the hydrogen rich fuel gas product for firing of the process air preheater, the remainder of the product is compressed and delivered to the hydrogen storage cavern at a nominal pressure of 33 bar (abs).

CO_2 Purification and Compression

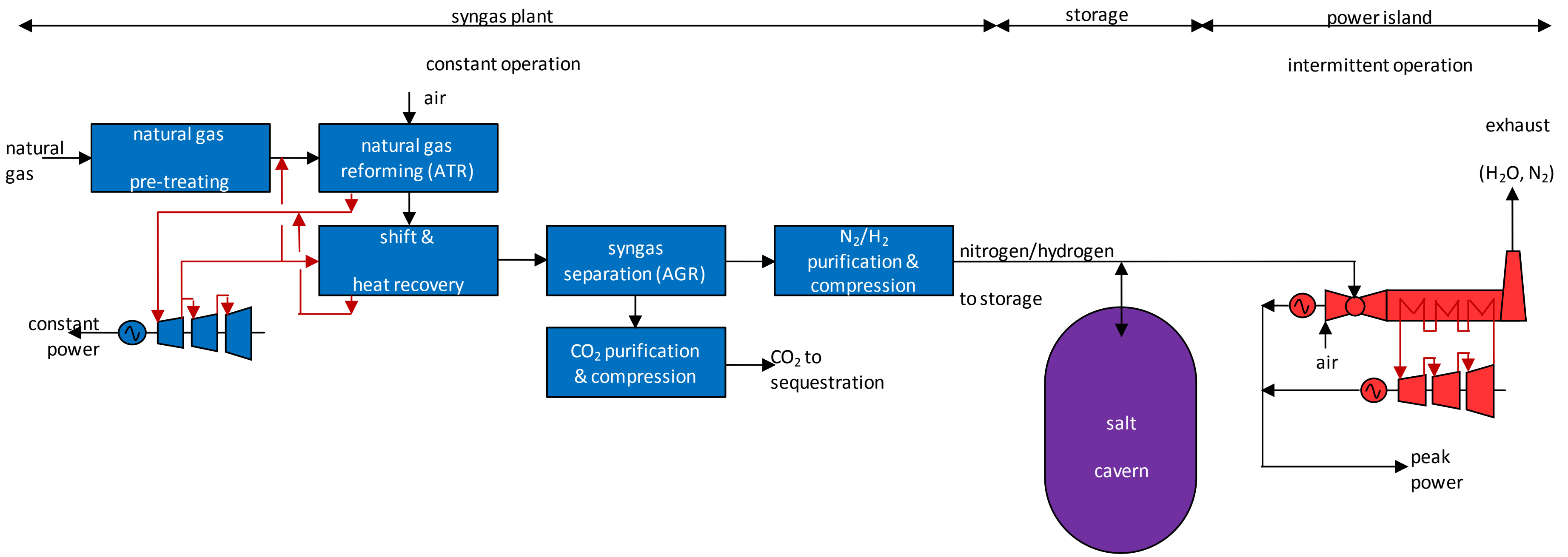
The CO_2 streams from AGR unit flow to the CO_2 Dehydration & Compression Unit where the CO_2 product is initially compressed to 29 bar (abs), then dried with molecular sieve to < 50 vppm water content, before being further compressed to 151 bar (abs) for export to sequestration.

Steam System

128 bar (abs), 510°C steam is generated in the reformed gas boiler and superheated as described above. This steam supplies a condensing steam turbine generating electric power for internal plant use. About 40% of the inlet steam is extracted at 30 bar (abs), initially to strip shift process condensate and then as process steam for the ATR.

The high level technical performance of power generation based upon this method of hydrogen production can be understood by studying the utility summary results in Attachment 7. The highlights from the utility summary and material balance are compared against the other methods of hydrogen production in Section 4.7.

Figure 3 – BFD for Case 3 – Autothermal Reforming with CCS and combined H_2 / N_2 storage



The Heat & Material Balance for this scheme is provided in Attachment 6.

The Utility Summary for this scheme is provided in Attachment 7.

4.6 Case 4 - Hydrogen Production via Steam Methane Reforming of Natural Gas

4.6.1 Introduction

Refer to Figure 5 – BFD for Case 4 (page 43)

The steam methane reforming (SMR) route to hydrogen production is the route usually employed for chemical processing uses, particularly on oil refineries, where the hydrogen is required for hydrotreating. The most common hydrogen purification step employed in modern plants in conjunction with SMR hydrogen production is to use pressure swing adsorption (PSA).

The process involves pre-treating the gas before mixing it with steam and feeding it to the steam reformer, which comprises a large number of externally heated tubes containing a nickel-based catalyst. In the catalyst tubes, the mixed natural gas/steam feed is converted to a syngas made up largely of hydrogen and carbon monoxide. The SMR is operated with a catalyst tube outlet pressure of 27.7 bara. The syngas from the SMR is then shifted with water to produce additional hydrogen and convert the carbon monoxide to CO_2. Heat is then recovered and the syngas is separated using a PSA into a hydrogen rich stream and a CO_2 rich stream. The hydrogen rich stream is then compressed and sent to the power island or to underground storage. The CO_2 rich stream, also containing hydrogen, CO and residual methane is fed as fuel to the reformer, along with supplementary natural gas fuel. The reformer flue gas is cooled and scrubbed with an amine solvent, so as to capture 90% of the carbon content of the natural gas feed to the SMR unit.

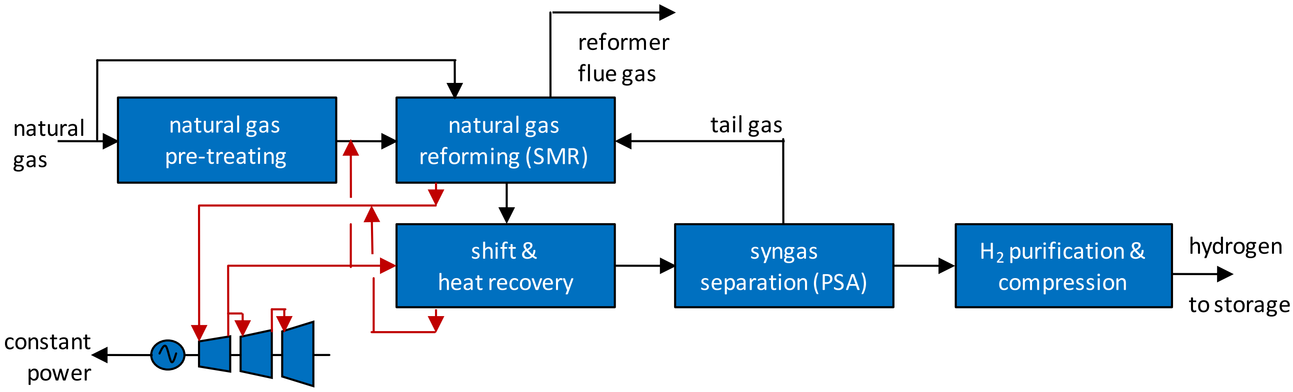


Figure 4 - BFD of hydrogen production via SMR, without carbon capture.

There are two main routes for capture of CO_2 emissions from an SMR hydrogen plant with PSA:

- CO_2 removal between the shift and the PSA (pre-combustion scheme)
- CO_2 removal from the reformer flue gas (post-combustion scheme)

CO_2 emitted from an SMR hydrogen plant has two sources; the combustion of the CO, CO_2 and residual methane in the PSA tail gas fired in the reformer, and the combustion of additional natural gas also required for reformer firing. Applying a pre-combustion carbon capture scheme to an SMR unit only achieves capture of the CO_2 content of the shifted gas, and it does not address the CO_2 emissions from firing of natural gas and PSA tail gas in the reformer. Typically this supplementary natural gas fuel for reformer firing accounts for nearly 20% of the total feed so it is essential to capture carbon from this fuel in order to achieve 90% carbon capture overall. A pre-combustion capture scheme only captures approximately 60% of the total carbon fed to the SMR. Therefore, whilst a pre-combustion capture scheme may result in a lower capital cost, a post-combustion capture scheme has been considered for this study, as shown in Figure 5.

4.6.2 Process description

Natural Gas Pre-Treating

Natural Gas is pre-treated in a hydro-desulphurisation reactor in which the feed gas is mixed with a small stream of recycled hydrogen and is preheated to 350°C. The gas is then passed through a Cobalt-Molybdenum catalyst which converts organic sulphur into H_2S and then through a zinc oxide H_2S adsorbent.

Steam Reforming and Syngas Production

Desulphurised feed from the pre-treating section is mixed with superheated steam at a steam to carbon ratio of 2.7, then heated to 620°C in the first flue gas heater in the reformer convection section. The heated mixed feed then flows to the catalyst-filled reformer tubes. The gas mixture flows from top to bottom through tubes arranged in vertical rows. The tubes are heated externally causing the hydrocarbon/steam mixture to react, forming a syngas containing hydrogen and carbon monoxide.

The catalyst tube outlet temperature is 860°C. The process gas at the outlet of the reformer is cooled down in waste heat boiler to generate 35 bar steam and is then routed to the downstream Shift Conversion Unit.

The overall effects of reforming reactions are endothermic and the necessary heat is provided by external firing of the reformer radiant section. The reformer mainly uses tail gas from the PSA unit as fuel with supplementary firing provided by feed natural gas. Flue gas from the radiant section is cooled down in a series of coils in the convection section. The convection section heating duties comprise reformer mixed feed heater, MP steam superheater, MP steam generator and combustion air preheater.

Shift Conversion Unit

The cooled syngas from the waste heat boiler flows to the Shift Conversion Unit where CO and steam present in the syngas are converted to CO_2 and H_2.

Shift reactions are carried out in two reactors in series. The syngas enters the first shift reactor, which is a high temperature shift reactor operating with inlet temperature of 350°C. The syngas from the outlet of the first reactor is cooled through a series of heat exchangers before entering the second shift reactor, which is a low temperature shift reactor at 200°C. The heat of reaction from the first reactor is used to heat boiler feed water (BFW) and the feed to the desulphurisation reactor. The heat of reaction from the second reactor is used to preheat BFW.

Syngas Separation

The cooled, shifted syngas is routed to a PSA unit where pure H_2 is produced by separation from methane, CO and CO_2. The recovery (yield) of hydrogen in the PSA is 88 mol%. The total heating value of the combined reformer fuel comprises approximately 65% from the PSA tail gas with 35% from supplementary firing of natural gas.

The pure hydrogen product is then compressed to 33 bar (abs) and is delivered to the hydrogen storage cavern.

Reformer Flue Gas Acid Gas Removal (AGR) Unit

The AGR Unit removes CO_2 from the reformer flue gas at near atmospheric pressure. A generic MEA based system is used to remove CO_2 from the flue gas in an absorber in order to produce an almost CO_2 free flue gas. The CO_2-rich solvent from the bottom of the absorber is routed via heat recovery to a regenerator column where heat is applied to release the CO_2 from the solvent. The AGR process requires a large heating duty for the regenerator, which is primarily provided by LP steam supplemented by heating with the shifted syngas in this scheme.

CO_2 Dehydration and Compression

The CO_2 stream from the AGR unit flows to the CO_2 Dehydration and Compression Unit where the CO_2 product is initially compressed to 29 bar (abs), then dried with molecular sieve to < 50 vppm water content, before being further compressed to 151 bar (abs) for export to sequestration.

Steam System

Steam at 35 bar (abs) and 400°C is generated in the reformer convection section and superheated as described above. This steam supplies a back pressure steam turbine exhausting at 4 bar (abs) and generating electric power for internal plant use. The 4 bar (abs) steam from the exhaust of this turbine provides the LP steam for the AGR reboiler.

Air Separation Unit

An air separation unit (ASU) is required in this scheme in order to provide the nitrogen needed by the GT for dilution of the hydrogen rich fuel gas and NOx control. Since an ASU cannot operate on the flexible scheme assumed for the power island the ASU is considered as part of the hydrogen production plant, although it will have no interaction with the rest of the process units other than to use steam, power and cooling water from this section of the overall scheme.

The high level technical performance of power generation based upon this method of hydrogen production can be understood by studying the utility summary results in Attachment 9. The highlights from the utility summary and material balance are compared against the other methods of hydrogen production in Section 4.7.

4.6.3 Case 4b – Hydrogen Production via Steam Methane Reforming with Steam to GT

Refer to Figure 6 – BFD for Case 4b (page 44)

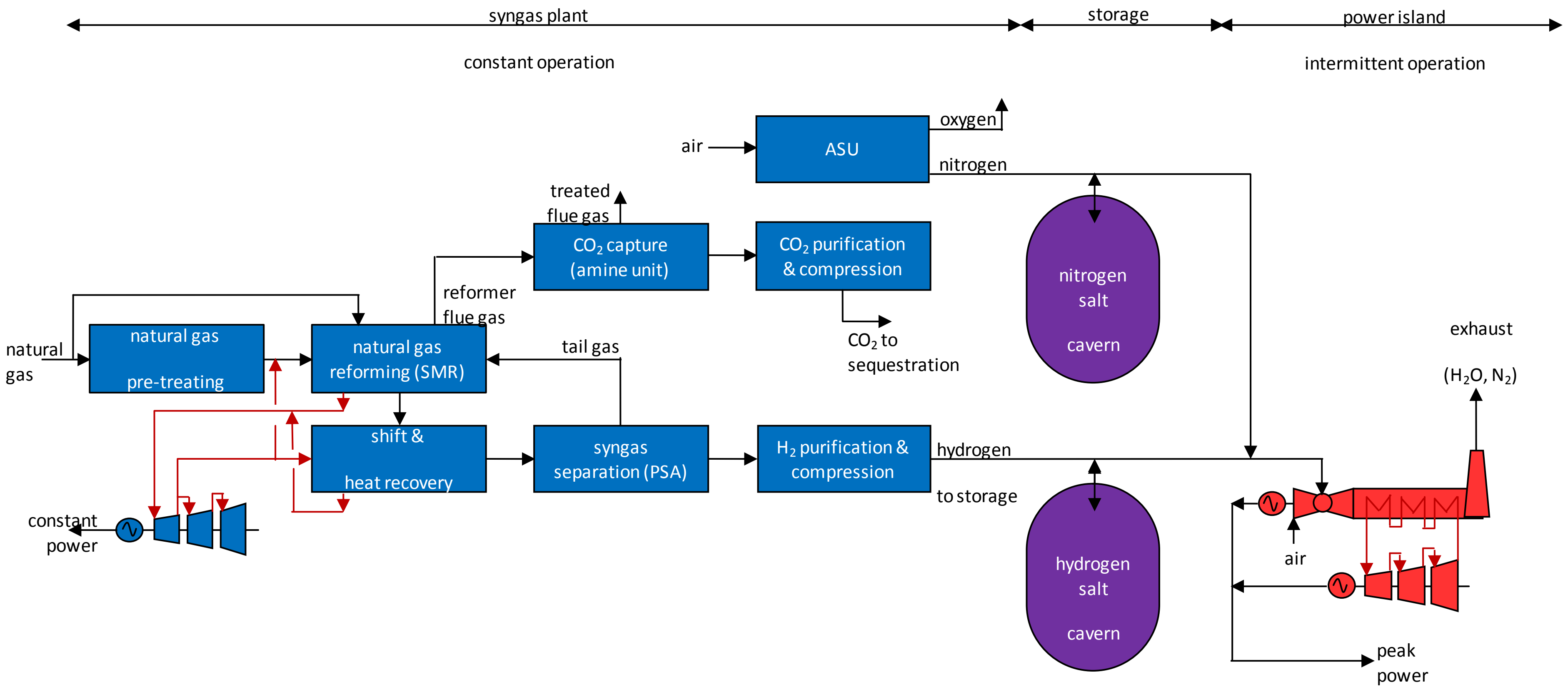
This scheme is essentially identical to the SMR-based hydrogen production scheme in Case 4 described above, except that nitrogen required for the power island gas turbine is fully replaced with steam generated in the gas turbine heat recovery steam generator. This change allows the deletion of the ASU and its nitrogen compressor from the scheme as well as the need for nitrogen storage. These items contribute significantly to savings in both capital cost and continuous power requirements, and hence a scheme avoiding these costs may prove beneficial.

The overall performance of the Case 4b scheme has been estimated based upon the assumption that nitrogen is replaced with steam at a flowrate advised by GE. The production of the necessary quantity of steam has been modelled as injection of boiler feed water (BFW) into the hydrogen-rich fuel gas received from storage at near-ambient temperature. This mixture is then heated in a coil in the heat recovery steam generator to a temperature of approximately 200°C, at which temperature the water has been fully evaporated. The resulting hydrogen/steam mixture is supplied to the burners of the gas turbine.

It should be noted that while this method gives an estimation of the overall power generation scheme impact of switching nitrogen dilution completely for steam it does not represent a fully developed or optimal power island design. Other issues such as significant increase in plant water demand, increased exhaust plume visibility and increased GT maintenance requirements may reduce the attractiveness of this option.

The high level technical performance of power generation based upon this method of hydrogen production can be understood from the utility summary results in Attachment 11. The highlights from the utility summary and material balance are compared against the other methods of hydrogen production in Section 4.7.

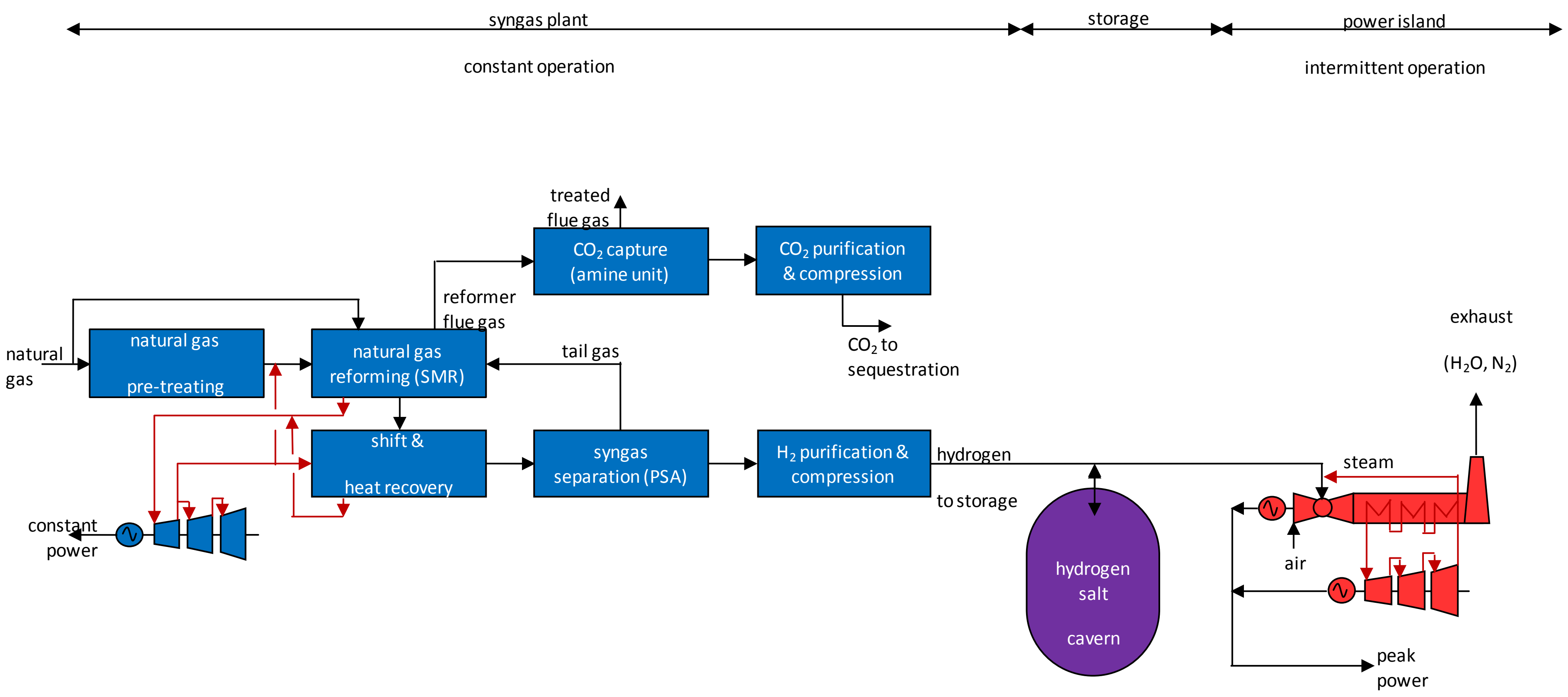
Figure 5 – BFD for Case 4 – Steam Methane Reforming with post-combustion CCS and separate H_2 / N_2 storage



The Heat & Material Balance for this scheme is provided in Attachment 8.

The Utility Summary for this scheme is provided in Attachment 9.

Figure 6 – BFD for Case 4b – Steam Methane Reforming with post-combustion CCS, H_2 storage and steam to GT



The Heat & Material Balance for this scheme is provided in Attachment 10.

The Utility Summary for this scheme is provided in Attachment 11.

4.7 Technical Performance Results

Table 6 summarises the key numerical findings from the high level simulation of power generation schemes based upon the hydrogen production options considered.

For the purposes of calculating overall plant efficiency a steady state basis is required. The figures below represent the energy balance if the plant were operating at the full capacity of one GT at all times.

Table 6 - Summary of technical performance results

Steady state basis with the plant running at 100% GT capacity			Coal Gasification	Coal & Biomass Gasification	ATR	SMR	SMR with Steam to GT
			Case 1	Case 2	Case 3	Case 4	Case 4b
Hydrogen Production							
	Feedstock flow rate	te/h	136.52	152.74	65.08	76.63	73.86
	Total Feedstock LHV	MWth	986.00	1002.50	859.29	1011.79	975.19
	Carbon in feeds	te/h	88.46	90.00	47.57	56.35	54.31
	Carbon captured	te/h	79.62	80.99	43.10	50.95	49.11
	Carbon captured		90.0%	90.0%	90.6%	90.4%	90.4%
	Oxygen consumption (Note 1)	te/h	102.75	97.20	0.00	103.84	0.00
	Syngas Product LHV	MWth	666.0	665.8	665.7	666.0	666.0
Power Balance							
	Hydrogen production total	kWe	-49713	-48069	-33932	-55585	-2410
	Pre-treatment gasification/reforming & shift	kWe	-8940	-9799	-45475	-1522	-1445
	Heat recovery & steam turbine	kWe	59047	60334	42598	27599	28303
	Acid gas removal	kWe	-7808	-7808	-4665	-3078	-2966
	CO2 dehydration and compression	kWe	-25120	-25543	-17543	-20399	-19661
	Sulphur recovery and tail gas treatment	kWe	-1729	-1882	0	0	0
	Air separation unit	kWe	-34295	-32462	0	-19148	0
	N2/H2 compression	kWe	-27499	-27409	-6208	-34638	-2905
	Fresh cooling water	kWe	-954	-995	-775	-1546	-1313
	Sea cooling water	kWe	-2416	-2506	-1863	-2854	-2423
	Cavern storage	kWe	0	0	0	0	0
	Power Generation	kWe	391193	391152	389650	392788	353572
	Offsites & Utilities	kWe	-2264	-2286	-2338	-2338	-2338
	Net Power Export	kWe	339216	340797	353380	334865	348825
	Plant Efficiency (LHV)		34.40%	33.99%	41.12%	33.10%	35.77%

Note 1: Oxygen produced in the ASU in Case 4 is not used by the process, and can either be vented or potentially exported.

With the exception of the SMR cases (Cases 4 and 4b) the plant efficiencies in Table 1 above are broadly in line with historical studies of pre-combustion capture power generation processes with 90% carbon capture, which show approx 40% with natural gas feedstock and approx 35% LHV efficiency with bituminous coal feedstock. The lower efficiency of the SMR cases is mainly attributable to the relatively low steam generation pressure in the reforming unit (35 bar) and the high reboiler heat duty of the flue gas CO2 capture unit (amine process).

Previous studies indicate that plant efficiency is not significantly affected by the selection of oxidant. Foster Wheeler has simulated an oxygen-fed ATR (outside the scope of the this study) giving an overall efficiency approximately 41% - same as for Case 3 above with an air-fed ATR. For coal feed it may be noted that the Mitsubishi-designed IGCC operating in Japan with air-fed gasifier is considered to have superior efficiency to oxygen-fed IGCC schemes.

From Table 1, the following observations can be made:

- The carbon content in the natural gas fed cases is much lower than the coal fed cases since a significant portion of the energy in the natural gas feed is provided by the hydrogen content of natural gas.
- The oxygen production/consumption is very similar in the two gasification cases. Slightly less oxygen is required for the biomass case since more oxygen is contained in the biomass feedstock.
- In all cases the power recovered by using excess process heating for steam generation enables the hydrogen production facility to meet approximately 50% of its own parasitic demand. In the SMR case with steam dilution of the fuel gas and hence no ASU, however, the hydrogen production plant is close to self-sufficient in power.
- The very high power demand in the reforming section of the ATR plant is due to the air compressor. Cases 3 and 4b have much lower power demand for the H2/N2 compression unit as these two schemes do not require a nitrogen compressor.
- Net power produced by the power island is similar in all cases since the impact on the power island between cases is very small, due only to minor changes in fuel gas compositions. In case 4b the power produced is lower since more energy internal to the power island is required to provide the steam used for fuel gas dilution and NOx control.
- The total quantity of net power available for export is very similar in all cases, with the ATR and SMR with steam dilution cases exporting slightly more power than the other cases.
- The overall efficiency figures show that the above differences in performance combine to put the ATR case ahead of the other 4 cases by a margin of nearly 5.4 LHV efficiency % points. Compared to previous similar studies this is a very significant margin. The efficiency differences between the two gasification cases and between the two SMR cases are small in comparison.

The superior performance of the ATR case does not, however, necessarily mean that a natural gas fed ATR with carbon capture is the best choice for the hydrogen storage and flexible turbines system, as the thermal efficiency of the plant is only one of a number of key factors. Feedstock price, availability and overall plant economics will have a big impact on whether or not the ATR case would be the best option for this project moving forward.

GE has advised that substitution of nitrogen dilution with steam is feasible, but the GT would be de-rated to mitigate the expected higher maintenance costs associated with steam dilution. To make an allowance for such derating, Foster Wheeler has made preliminary simulations of steam injection, using guidelines from GE, in place of nitrogen. For the SMR Case 4b, steam injection results in a gain in overall efficiency of about 2.67% points, due to a significant contribution from elimination of the ASU and N_2 compressor. IT is estimated that for Cases 1 & 2, steam injection will result in a smaller increase in overall efficiency of <0.5% points.

4.7.1 Consumption of Natural Gas

The gas-to-power efficiencies for the natural gas based cases are 41% for Case 3 Autothermal Reforming and 33% for Case 4 Steam Reforming. These efficiencies are low compared with the efficiency expected from a natural gas fired combined cycle with typical post combustion carbon capture such as amine wash.

The suppliers of large gas turbines (Alstom, GE, Mitsubishi, Siemens) can now provide natural gas fired combined cycle plant with over 60% LHV efficiency under UK climatic conditions. The efficiency of combined cycle + CCS can be expected to have reached 52-54 %, so the efficiency penalty for CCS appears to be in the range of 6 to 8 % points.

Consequently, the amount of natural gas consumed in Case 3 is expected to be around 30% higher per kWh, and in Case 4 60% higher, than the main-stream CCGT + post-combustion CC technology, for which the plant efficiency is now typically 53%. This 30% increase in fuel cost may cause concern at a time when natural gas storage capacity and national security of gas supply are in focus.

4.8 Availability and Life of Plant

For the natural gas based cases (Cases 3, 4 and 4b) an almost indefinite life can be expected for most of the large equipment and for the installation as a whole. The catalyst tubes in an SMR plant are usually designed for 100,000 hours life, so require periodic replacement. The technology is based on the numerous natural gas based ammonia, hydrogen and methanol plants designed and installed worldwide over the past 40 years. As examples, the two large autothermal synthesis gas installations in Poland have been in operation, essentially with only normal maintenance, since the mid 1960s and 1970s; and several large steam reforming ammonia plants installed in North & South America and Russia are still in operation and are being upgraded after 35-40 years of service. Plant availability and down-time vary with the standard of maintenance, but very long run times can be achieved, a steam reforming plant in the Netherlands was recently running non-stop for three years.

Gasification plants (Cases 1 and 2) are based on many of the equipment types and sub processes used in the natural gas based syngas plants, but the overall plant availability has tended to be rather lower (<90%) due to the need for scheduled and unscheduled attention to high temperature parts of the gasifiers, particularly the burners and refractory linings. The licensors are continuously improving their technologies in these respects, but in the absence of more licensor recommendations and records, an initial plant availability of 85% may be assumed. As with the natural gas based plants, the life of most of the equipment and the gasification plant as a whole can be considered to be almost indefinite, at least 50 years.

Unlike natural gas fired plants, coal-fired power plants traditionally have large reserves of raw coal on site, stored on the ground. Historically these have proved invaluable, particularly during periods of industrial unrest. While conditions in the industry have changed, there could be resistance to any drastic cutting back of site coal reserves.

4.9 Capital and Operating Cost Estimates

4.9.1 Capital Cost Estimating Basis

The capital cost estimates contained within this study report have been based on the technical definition generated for each of the cases under consideration. These are factored estimates and are based on the scaling of previous similar estimates prepared using Foster Wheeler in-house data, including previous ETI Studies.

For all of the cases, reported the source estimate data has been adjusted to provide figures on a consistent and comparable) 1Q 2010 UK Basis.

Capital cost estimates prepared using this methodology and associated qualifications/exclusions are normally considered to have an accuracy of +/-40% at best. This accuracy is considered on the overall project cost (not individual lines items on the summary).

Estimate Format

The Work Breakdown Structure (WBS) used for the estimates is as follows:-

- Coal handling, storage, milling and gasification or natural gas pre-treatment and reforming;
- Air separation / oxidant supply;
- Acid gas removal;
- Sulphur plant;
- Syngas treatment unit;
- CO_2 compression and dehydration;
- Power block (hydrogen production units area only);
- Common facilities.

The common facilities area includes the following major items, as appropriate:

- Interconnecting piping;
- Electrical switchgear/transformers;
- 275 kV cables to new switchyard;
- DCS system;
- Demineralised water system;
- Natural gas system;
- BFW chemical injection;
- Condensate polishing package;
- Chemicals;
- Water treatment;
- Flare package;
- Instrument/utility air package;
- Fire fighting system.

Currency

The estimates are reported in GB Pounds (GB£).

When in-house data is available in a different currency, the following currency conversion rates have been used for conversion:

GB£ 1.00 = US$ 1.52

Escalation

The estimates have been escalated to the date of the reference project (1Q 2010), based on Foster Wheeler experience. No allowance has been made for future escalation.

Major Equipment

The equipment costs have been factored from major equipment of previous (carbon capture) projects of a similar nature. The costs are inclusive up to FOB.

Direct Materials

The estimated material costs reflect worldwide procurement, therefore no allowance for possible savings by locally purchasing of direct materials and associated reductions in shipping costs have been made.

Bulk Materials

The bulk material costs have been factored from the major equipment costs using factors derived from a more detailed study for a very similar plant. The costs include piping, instrumentation, electrical, catalyst and chemicals, spares (commissioning and two years operational) and shipping and freight.

Direct Material and Labour Contracts

These costs include for civil, steelwork, buildings and protective cover and have been factored from projects of a similar nature.

Where necessary, adjustments have been made if source data is from a different location to the UK.

No allowance has been made for seawater intake / outfall and associated piping.

Labour Only Contracts

These costs include for mechanical, electrical and instrumentation, pre-commissioning trade labour support and scaffolding labour costs and have been factored from projects of a similar nature.

Indirect Costs

These costs include for temporary facilities, heavy lifts, commissioning services and vendors engineers and have been factored from projects of a similar nature.

EPC Contracts

These costs include for home office engineering and procurement and construction management and have been factored from projects of a similar nature.

Land / Site costs

No Site specific costs have been included. The site has been assumed to be a generic site clear and level and free from underground obstructions.

Land costs have been included (as specified by ETI) at a rate of 5% of the total installed costs for all cases.

Owner’s Costs

Owner’s costs have been included (as specified by ETI) at a rate of 10% of the total installed costs for all cases.

Contingency

Contingency has been included (as specified by ETI) at a rate of 25% of the total installed costs for all cases.

Exclusions

The following costs have been specifically been excluded from this estimate:

- Seawater intake / outfall and associated piping;
- Import duties;
- Capital / insurance spares;
- Financing;
- Royalties & process guarantees;
- Piling;
- Removal of unseen/unidentified underground obstructions;
- Operating costs; (which are covered separately)
- Statutory authority & utility company costs & permits;
- Currency fluctuations;
- PMC costs;
- Contractors fees;
- Contractors all risk insurance;
- Taxes;
- Metal pricing movements.

4.9.2 Operating and Maintenance Cost Estimating Basis

Operating and Maintenance (O&M) costs include the following;

- Feedstocks;
- Chemicals;
- Catalyst;
- Solvents;
- Direct labour;
- Maintenance;
- General Overheads.

O&M costs are generally allocated as variable and fixed costs. Variable operating costs are directly proportional to the amount of kilowatt-hours produced and are referred as incremental costs. They may be expressed in £/kWh. Fixed operating costs are essentially independent of the quantity of kilowatt-hours produced. They may be expressed in £/h or £/year.

Variable costs

The variable costs include the consumption of catalysts, chemicals and solvents. These costs are annual, based on the expected equivalent availability of the plant. The variable costs mainly include the following:

- Feedstocks (natural gas, coal or biomass),
- Solvent consumption for the chemical or physical removal of the acid gases,
- Catalyst consumption for the CO shift reaction and the Claus/Scot unit,
- Chemicals for water/steam treatment and waste water treatment,
- CO_2 emissions (cost assumed to be zero),
- Waste disposal.

The following feedstock costs have been specified by the ETI for use on this project:

- Coal - £70/te (0.94p/kWh)
- Gas - $6.6/MMBTU (1.5p/kWh)
- Biomass £6/GJ HHV (2.16p/kWh)

Fixed costs

The fixed costs mainly include the following:

- Direct labour
- Administrative and general overheads
- Maintenance

Direct Labour

The yearly cost of the direct labour has been calculated assuming, for each individual, an average cost equal to £50,000 / year. The number of personnel engaged for the different alternatives has been evaluated on the basis of the following tables.

Based on Table 7 & Table 8 below it has been assumed that the Power Island section of the plant will require approximately 40 personnel. Subtracting 40 personnel from the figure required for a gasification plant with carbon capture plant gives an estimate of 88 personnel required for the hydrogen production plant for the two gasification cases (Cases 1 and 2). The reforming cases (Case 3, 4 and 4b) are somewhat simpler plants so a figure of 60 personnel has been selected, with an additional 15 personnel for the SMR case including and ASU (Case 4).

Table 7 - Personnel basis for Gasification plants with CO_2 capture

Operation	ASU	Gasification	CCU & Utilities	Total	Notes
Area Responsible	1	1	1	3	daily position
Assistant Area Responsible	1	1	1	3	daily position
Shift Superintendent	5			5	1 shift position
Electrical Assistant	5			5	1 shift position
Shift Supervisor	5	5	5	15	3 shift position
Control Room Operator	5	10	10	25	5 shift position
Field Operator	5	25	20	50	10 shift position
Subtotal				**106**	
Maintenance					
Mechanical group	4			4	daily position
Instrument group	7			7	daily position
Electrical group	5			5	daily position
Subtotal				**16**	
Laboratory					
Superintendent + Analysts	6			**6**	daily position
Total				**128**	

Table 8 - Personnel basis for Combined Cycle Gas Turbine plants without CO_2 capture

Operation	Total	Notes
Area Responsible	1	daily position
Assistant Area Responsible	1	daily position
Electrical Assistant	5	1 shift position
Shift Supervisor	5	1 shift position
Control Room Operator	10	2 shift position
Field Operator	10	2 shift position
Subtotal	**32**	
Maintenance		
Mechanical group	3	daily position
Instrument group	3	daily position
Electrical group	2	daily position
Subtotal	**8**	
Laboratory		
Superintendent + Analysts	**4**	daily position
Total	**40**	

Administrative and General Overheads

These costs include all other Company services not directly involved in the operation of the Complex, such as:

- Management
- Personnel services
- Technical services
- Clerical staff

These services vary widely from company to company and are also dependent on the type and complexity of the operation.

Based on an EPRI study, Technical Assessment Guide for the Power Industry, an amount equal to 30% of the direct labour cost has been considered.

Maintenance

A precise evaluation of the cost of maintenance would require a breakdown of the costs amongst the numerous components and packages of the complex.

Since these costs are all strongly dependent on the type of equipment selected and statistical maintenance data provided by the selected supplier, this type of evaluation of the maintenance cost is premature at this stage of the study.

For this reason, the annual maintenance cost of the complex has been estimated as a percentage of the installed capital cost of the facilities.

Different percentage factors have been applied to the different units, based on the following criteria:

- 4.0% for solid handling units;
- 2.5% for gaseous and liquid handling units;
- 1.7% for utilities and offsites;
- 5.0% for the Power Island (to take into account the gas turbine maintenance cost based on the assumption of a Long Term Service Agreement (LTSA) with the selected gas turbine manufacturer).

4.9.3 Summary of Capital Cost Estimates

Table 9 – Capital Costs Estimate Summary

DESCRIPTION	CASE 1 COAL GASIFICATION	CASE 2 COAL & BIO GASIFICATION	CASE 3 ATR	CASE 4 SMR	CASE 4B SMR	POWER ISLAND
	Million £	Million £	Million £	Million £	Million £	Million £
MAJOR EQUIPMENT	193.7	196.0	92.7	188.4	112.0	108.2
DIRECT BULK MATERIALS	89.3	91.9	110.7	61.5	57.1	32.6
DIRECT MATERIAL & LABOUR CONTRACTS	40.5	40.7	96.6	36.9	34.6	12.3
LABOUR ONLY CONTRACTS	95.7	96.5	80.6	95.3	59.4	18.2
INDIRECTS	29.6	30.0	34.7	26.1	18.0	12.1
EPC CONTRACTS	50.8	51.1	62.3	36.4	31.0	8.3
INSTALLED COST	499.7	506.2	477.7	444.6	312.1	191.7
LAND COSTS, 5%	25.0	25.3	23.9	22.2	15.6	9.6
OWNERS COSTS, 10%	50.0	50.6	47.8	44.5	31.2	19.2
CONTINGENCY, 25%	124.9	126.6	119.4	111.2	78.0	47.9
TOTAL PROJECT COST	**699.6**	**708.7**	**669.0**	**622.5**	**436.9**	**268.4**

It can be seen that there is a small incremental cost - circa 1% - associated with the Case 2 (coal + biomass fed gasification) process when compared with the Case 1 (coal fed gasification) process.

The Case 3 ATR estimate gives a total project cost of £669m, which is £31m cheaper than Case 1. There are notable differences in the proportional costs within each of the line items in case 3 (when compared to other cases). This is due to the way the ATR scheme estimate was derived with fewer large packages (Gasifiers, ASU) contributing to the major equipment cost, and an increase to the direct materials and installation costs.

Conversely, the Case 4 SMR estimate gives a total cost of £622m, which is £78m cheaper than Case 1. The SMR option consists of relatively simple and highly packaged process units, which shows in proportionally lower direct materials and installation costs.

The Case 4b SMR without ASU demonstrates the impact on the SMR process of having to produce nitrogen solely for use as a dilution gas for the GT. Removing the ASU reduces the total project cost by £185m to £437m.

Capital Cost Summaries for each case are provided in the following Attachments:

- Att 12, Capital cost summary – Case 1
- Att 13, Capital cost summary – Case 2
- Att 14, Capital cost summary – Case 3
- Att 15, Capital cost summary – Case 4
- Att 16, Capital cost summary – Case 4b
- Att 17, Capital cost summary – Power Island

4.9.4 Summary of Operating Cost Estimates

Table 10 - Operating Costs Estimate Summary

Million £/y	Case 1 Coal Gasification	Case 2 Coal-Biomass Gasification	Case 3 ATR	Case 4 SMR	Case 4b SMR with Steam to GT	Power Island - All Cases
Fixed Costs						
Direct Labour	4.40	4.40	3.00	3.75	3.00	2.00
Administration / General Overheads	1.32	1.32	0.90	1.13	0.90	0.60
Maintenance	15.58	15.89	12.11	12.62	12.11	8.80
Insurance & Local Taxes Allowance	9.99	10.13	8.89	9.55	8.89	6.24
Total Fixed Costs	**31.3**	**31.7**	**24.9**	**27.0**	**24.9**	**17.6**
Variable Costs						
Feedstock	71.2	93.5	102.6	120.8	113.9	0.0
Solvent, Catalysts and Chemicals	1.90	1.90	1.25	1.00	1.00	0.38
Waste Disposal	0.95	0.95	0.63	0.50	0.50	0.19
Total Variable Costs	**74.00**	**96.32**	**104.43**	**122.25**	**115.35**	**0.56**
TOTAL OPERATING COSTS	**105.3**	**128.1**	**129.3**	**149.3**	**140.3**	**18.2**

Table 10 shows that the operating costs are dominated by the cost of purchasing feedstock, as would be expected. Differences caused by considerations such as maintenance and different numbers of personnel required to operate the plant are relatively minor in comparison. Since natural gas is the highest cost feedstock the natural gas fed cases exhibit higher total operating cost, with the coal & biomass case showing similar operating cost and the coal fed gasification case showing the lowest.

The figures above do not include a cost for electricity consumed by the hydrogen production units. This is considered as part of the economics within Table 12.

4.10 Project Execution Schedules

The attached schedules show the main differences in the project execution time frame, from the end of "Appraise" phase up to "H2 Ready for Storage", as a basis for comparison of the four hydrogen production options under consideration.

The project execution schedule is broken down into four (4) phases:

1. **The "Select" phase** is primarily used to select the licensors for the main Packages and produce an assessment of the time and cost of the project approximated to +/-30%. Upon completion of this phase, the shareholders will have sufficient information related to economics/ finance to decide whether or not to proceed to the next phase.
2. **The "Define / FEED" phase** is primarily used to develop the design in more detail and add evaluation criteria and design data that enable the project team to achieve an approximation of the project time and costs of +/- 15%. As part of this phase the ITB package to be sent to potential main contractors for the EPC bid phase is also defined.
3. **The "EPC Bid Award" phase** is for technical / commercial evaluation of potential EPC contractors. During this phase an upgrade of the engineering documents is also planned in order to be aligned with technical clarifications raised during the bid phase.
4. **The "EPC / Start-up" phase** refers to the project execution from start of Detailed Design up to H2 Ready for Storage, through materials procurement, subcontracting and construction.

4.10.1 Qualifications and Assumptions

The main qualifications and assumptions used to estimate the Overall Project durations are described below.

The durations of tasks shown in the programme, are based on in-house information held by Foster Wheeler, based on historical project data, previous similar project experiences and material supply and installation benchmark data time frames. The programme is based on the assumption that at the date zero, the main contractor who will be responsible for the execution of the Select, Define / FEED and PMC support during the EPC Bid Award phases, has already been selected.

The project execution strategy has been assumed to be based on a lump sum approach for the execution of the EPC Phases. The time span shown under procurement phase is inclusive of delivery at site.

Only on-shore plant installation execution has been considered.

It has been assumed that the plant will be handed over to operations in a system-wise manner as each system is completed. The mechanical completion date shown in the schedules, refers to the last system to be mechanically complete and ready for commissioning and start-up.

The duration of the activities shown may vary considerably depending on the period in which the project will be actually executed, mainly due to market conditions with regards to the materials supply and manpower availability.

The plant final location, in respect to the H2 storage area and brine discharge, is another fundamental that could change the project durations, mainly due to the pipeline part of work that can vary considerably.

Taking this into consideration, these Comparative Project Execution Schedules should be used for comparative purposes only, for considering the differences between potential hydrogen production options, and not used as the basis for setting firm schedule expectation.

4.10.2 Comparison of Schedules

The Select, Define / FEED and EPC Bid Award phases will have the same duration, for all of the cases under consideration, as the amount of engineering deliverables to be produced is similar.

As such, the critical path for the project execution of the hydrogen production plant is driven by the delivery time span of the main Long Lead Items (LLI)s.

Long Lead Items – Cases 1 & 2 – Gasification

- Gasifier Unit – 21 Months
- SRU Unit – 18 Months
- Compressors – 20 Months
- Titanium Plate Heat Exchanger – 18 Months
- Steam Turbine – 24 Months

Long Lead Items – Case 3 – ATR

- AGR Unit – 18 Months
- Auto thermal Unit – 20 Months
- Compressors – 20 Months
- Titanium Plate Heat Exchanger – 18 Months
- Steam Turbine – 24 Months

Long Lead Items – Case 4 – SMR

- PSA Unit – 18 Months
- Reformer Unit – 20 Months
- Compressors – 20 Months
- Titanium Plate Heat Exchanger – 18 Months
- Steam Turbine – 20 Months
  (Four months less compared to the others cases due the short delivery lead time for the single stage Steam Turbine.)

4.10.3 Summary of Overall Project Execution Schedules

Table 11 – Project Execution Schedule Summary

Phase Description	Case 1 Coal Gasification Duration (Months)	Case 2 Coal/Bio Gasification Duration (Months)	Case 3 ATR Duration (Months)	Case 4 SMR Duration (Months)
Select	6	6	6	6
Define / FEED	9	9	9	9
EPC Bid Award	10	10	10	10
EPC / Start-up	39	39	39	35*
Overall	**64**	**64**	**64**	**60**

Case 4 uses a single stage steam turbine, while the previous cases are based on multi-stage turbine. The single stage turbine has a manufacturing and delivery time span approximately four months less than a multi-stage and therefore, the Overall Project Execution Schedule is four months shorter for Case 4.

Project Execution Schedules for each case are provided in the following Attachments:

- Att 18, Comparative Project Execution Schedule – Case 1
- Att 19, Comparative Project Execution Schedule – Case 2
- Att 20, Comparative Project Execution Schedule – Case 3
- Att 21, Comparative Project Execution Schedule – Case 4

4.11 Conclusions

It has been shown in the preceding sections that there are many factors to consider when selecting the most appropriate method of hydrogen production to apply to the hydrogen storage and flexible turbine scheme.

Table 12 summarises the numerical findings on the simplified basis of a plant running continuously as a base-load plant at the capacity of a single GT, in order to demonstrate the differences between the five cases studied. Clearly the economics of the flexible operating system will be more favourable than the base-load operation, however, it can be reasonably assumed that the improvement to the economics would be the same for all the cases.

The differences between the technical performance, capital and operating costs have been discussed in previous sections with the following headline conclusions:

- The overall efficiency figures show that the differences in performance combine to put the ATR case ahead of the other 4 cases by a margin of nearly 5.4 LHV efficiency % points. The efficiency differences between the two gasification cases and between the two SMR cases are small in comparison.
- The capital costs of the two gasification cases are almost identical. The Case 3 ATR estimate gives a total project cost of £669m, which is £31m cheaper than Case 1. The Case 4 SMR estimate gives a total cost of £622m, which is £78m cheaper than Case 1, while the Case 4b SMR without ASU demonstrates the impact on the SMR process of having to produce nitrogen solely for use as a dilution gas for the GT. Removing the ASU reduces the total project cost by £185m to £437m.
- The operating cost figures are dominated by the price of the various feedstock which results in the natural gas fed cases exhibiting the highest total operating cost. The coal & biomass cases show similar costs, with the coal fed gasification case giving the lowest operating cost.

The high efficiency of the ATR case means that it is a stronger option than the SMR case with nitrogen dilution to the GT. This is because less feedstock is required to produce the same quantity of product for only a marginal increase in capital cost. However, the SMR with steam dilution to the GT gives the best overall LCOE, as the elimination of the ASU both saves significant capital cost and operating cost through reduced parasitic electricity demand.

Once the capital cost of the cavern is included, Case 4b should look better still, as less storage will be required.

The coal fed gasification case also looks worth further consideration, however, since its operating cost is substantially lower than the ATR case, even though its overall plant efficiency is just 34.4% compared to the ATR's 41.1%.

Table 12 - Summary of numerical results for all cases

Steady state basis with the plant running at 100% GT capacity			Coal Gasification	Coal & Biomass Gasification	ATR	SMR	SMR with Steam to GT
			Case 1	Case 2	Case 3	Case 4	Case 4b
Hydrogen Production							
	Feedstock flow rate	te/h	136.52	152.74	65.08	76.63	73.86
	Total Feedstock LHV	MWth	986.00	1002.50	859.29	1011.79	975.19
	Carbon in feeds	te/h	88.46	90.00	47.57	56.35	54.31
	Carbon captured	te/h	79.62	80.99	43.10	50.95	49.11
	Carbon captured	%	90.0%	90.0%	90.6%	90.4%	90.4%
	Oxygen consumption	te/h	102.75	97.20	0.00	103.84	0.00
	Syngas Product LHV	MWth	666.0	665.8	665.7	666.0	666.0
Power Balance							
	Hydrogen production	MWe	-49.71	-48.07	-33.93	-55.59	-2.41
	Cavern storage	MWe	0.00	0.00	0.00	0.00	0.00
	Power Generation	MWe	391.19	391.15	389.65	392.79	353.57
	Offsites & Utilities	MWe	-2.26	-2.29	-2.34	-2.33	-2.34
	Net Power Export	MWe	339.22	340.80	353.38	334.87	348.82
	Plant Efficiency (LHV)	%	**34.40%**	**33.99%**	**41.12%**	**33.10%**	**35.77%**
Costs							
	Hydrogen production capital (TPC)	Million £	**699.6**	**708.7**	**669.0**	**622.5**	**436.9**
	Storage cavern capital (TPC)	Million £	0	0	0	0	0
	Power island capital (TPC)	Million £	268.4	268.4	268.4	268.4	268.4
	Total scheme capital (TPC)	Million £	968	977.1	937.4	890.9	705.3
	Capital Intensity	Million £/MWe	2.85	2.87	2.65	2.66	2.02
	Hydrogen production total opex	Million £/yr	**105.3**	**128.1**	**129.3**	**149.3**	**140.3**
	Storage cavern total opex	Million £/yr	0	0	0	0	0
	Power island total opex	Million £/yr	18.2	18.2	18.2	18.2	18.2
	Total scheme opex	Million £/yr	123.5	146.3	147.5	167.5	158.5
	Opex intensity	Million £/yr/MWe	0.36	0.43	0.42	0.50	0.45
Other							
	Execution schedule duration	months	64	64	64	60	60
	Approximate lifetime	years	>30	>30	>30	>30	>30
	Availability	%	85%	85%	92%	92%	90%
Simplified LCO Estimates							
	Import Electricity Price (Note 1)	Million £/MWh	72.00	72.00	72.00	72.00	72.00
	Cost of Import Electricity to H2 Units	Million £/yr	26.65	25.77	19.69	32.25	1.37
	Cost of Import Electricity Total	Million £/yr	27.87	27.00	21.05	33.61	2.70
	CO_2 Price	£/te	10	10	10	10	10
	Income from CO_2 Export	Million £/yr	5.9	6.0	3.5	4.1	3.9
	Project Life	years	30.0	30.0	30.0	30.0	30.0
	Discount Rate	%	10%	10%	10%	10%	10%
	Levelised Cost of Hydrogen	£/MWth	40.38	44.98	40.36	45.36	35.06
	Levelised Cost of Electricity Export	£/MWh	85.18	93.01	84.24	92.09	83.26

Note 1: Data from DECC, 2012.

In order to quantify the relative merits of operating costs versus capital cost and efficiency it is necessary to perform a high level calculation of the project economic performance over a number of years of operation. In order to do this two simplified levelised cost calculations have been performed for each case. Based on the hydrogen plant capital cost, hydrogen plant operating cost, fixed imported electricity cost and an assumed figure for the value of CO_2 produced by the plant over a 30 year period with a discount rate of 10%, we can calculate:

- Levelised cost of hydrogen production ($LCOH_2$) – this values the technology as a producer of hydrogen; and
- Levelised cost of electricity export (LCOE) – this values the technology as a producer of electrical power.

  n.b. the LCOE calculation estimates the break-even price of electricity produced by the power island, assuming a single gas turbine operating continuously at full load, and with no buffer storage. As such it is only useful as a comparison between these cases, and is not directly comparable with LCOE in other studies.

The $LCOH_2$ calculations show that the cheapest way to produce hydrogen would be the SMR case 4b scheme at £35.1/MWth, where an ASU is not needed for the overall power production scheme. Cases 1 and 3 produce the next cheapest hydrogen at £40.4/MWth.

The LCOE calculation shows that, once the whole power production scheme is included, the SMR Case 4b is the strongest with a lowest LCOE value of £83.3/MWh. This value is comparable to its nearest rival, the ATR Case 3 with LCOE value of £84.2/MWh. A contributing factor to this result is the increased availability of the natural gas fed processes compared to the coal fed cases.

It should be noted that the amount of natural gas consumed in Case 3 is expected to be around 30% higher per kWh than the main-stream CCGT with post-combustion carbon capture technology. This 30% increase in fuel cost may cause concern at a time when natural gas storage capacity and national security of gas supply are in focus.

5. CHARACTERISATION OF BASIC DESIGN REQUIREMENTS FOR A COST EFFECTIVE HYDROGEN STORE

5.1 Objective

The objective of this section is to identify the salt cavern design requirements imposed by the hydrogen production and consumption processes, likely usage patterns and resultant storage capacities required to achieve these usage patterns. From this review, the characteristics of a well functioning, safe and economic salt cavern storage site for a hydrogen power project with CCS will be identified. The conclusions of this section of the scope of work will form the starting point for WP2 – Hydrogen Storage.

5.2 Hydrogen Production Requirements

By using a number of trains, the syngas production plant can produce any quantity of hydrogen the project (and its store) requires, with a sensible minimum size of syngas plant equivalent to an electrical generating capacity of around 175 MWe. This value corresponds to continuous generation of hydrogen in the syngas plant and operation of one GE Frame 9 gas turbine for twelve hours per day.

The syngas production plant will require the store to be available for 100% of the time that the syngas plant is running and/or the CCGT is running, depending on the relative capacities of the syngas and power islands.

No other requirements of the store imposed by the hydrogen production process are foreseen.

5.3 Hydrogen Consumption Requirements

Hydrogen usage rates, assuming the only hydrogen user is the power island (and therefore excluding consideration of any hydrogen product export), will be determined by the operating pattern assumed for the power plant. This, in turn, will be determined by the pattern of power demanded by the National Grid. Figure 7 shows the electricity price, grid demand pattern, and proportion of power supplied by intermittent generation via renewables for a typical month in 2009.

It is clear from Figure 7 above that the vast majority of the burden of intermittent operation falls to CCGT plants. It can also be assumed that with significantly more wind capacity expected to come on-line in the near future (wind will contribute to around 9% of UK supply at 2020) these plants will have to cope with the intermittency of the grid demand compounded by increasingly intermittent supply.

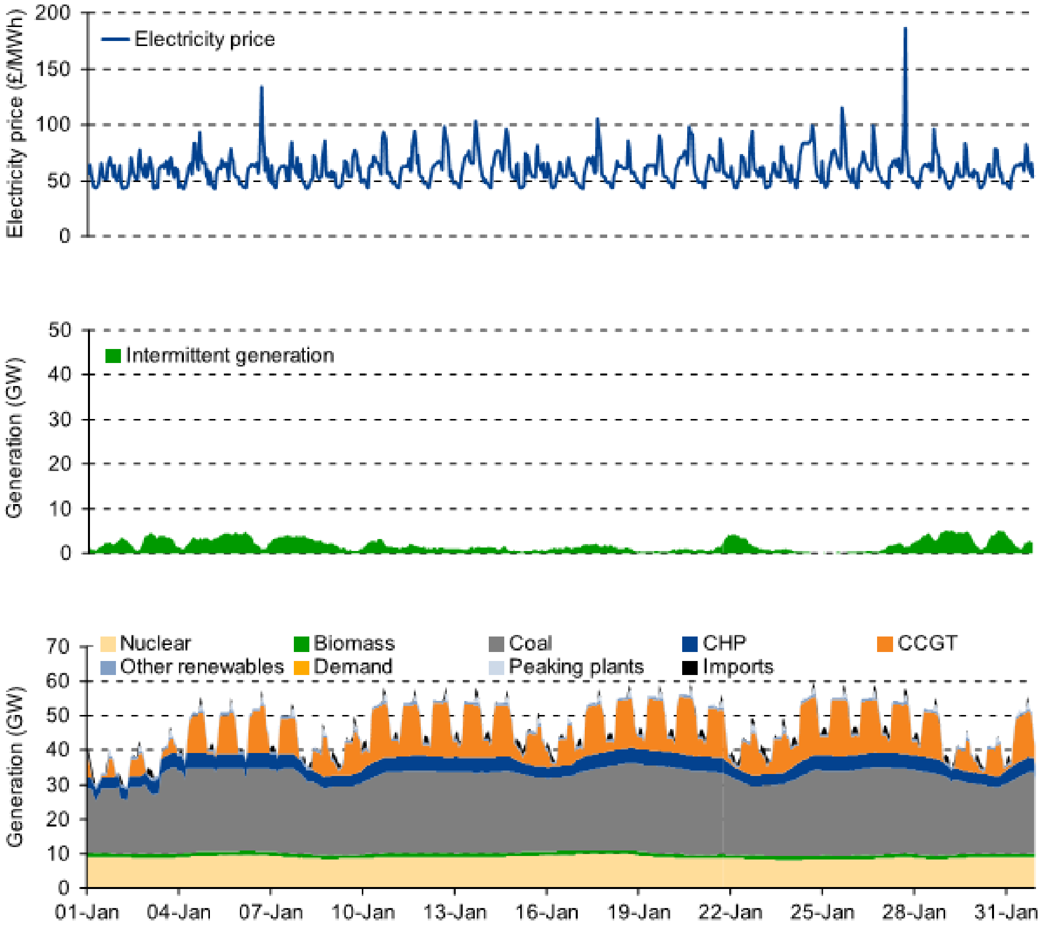


Figure 7 - Typical patterns of electricity price and plant dispatch, January 2009, Poyry(2)

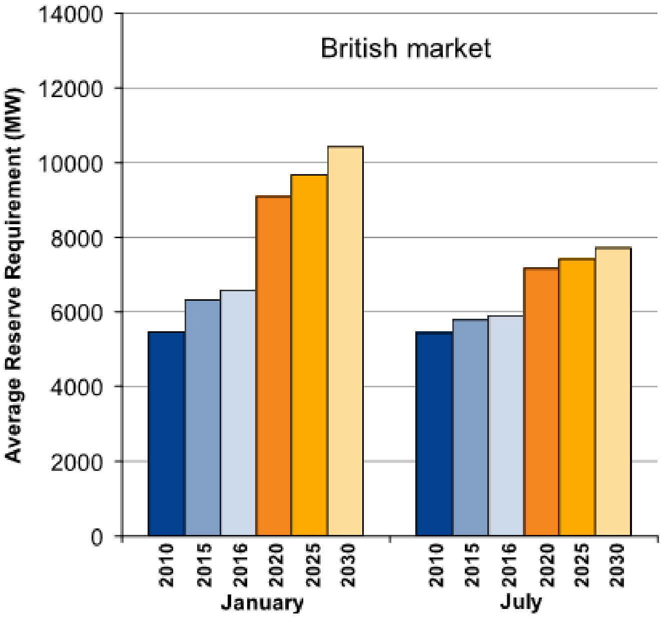


Figure 8 - Grid Capacity Reserve Requirements, Poyry(2)

Figure 8 shows that with increasing wind generation capacity coming on-line, the total reserve capacity required by the grid to cope with times when wind power is unavailable is expected to increase by approximately 30% by 2030. This means that the demand for flexible power generation should be significantly higher in 2030 than it is today.

It should be noted that as the Poyry report dates from 2009, its results may be impacted by a number of key events which can reasonably be expected to reduce the impact of intermittent wind generation:

- Expansion of the smartgrid system to medium and small users should enable much better alignment of demand and supply by allowing the grid to load-shed during periods of high demand and/or low wind power output,
- UK Energy Market Reform is designed to incentivise capacity provision to reduce peaking plant dependence on extreme price events for financial viability,
- Developments and deployment of energy storage at a significant level will also help to minimise future price spikes, such systems include pumped hydroelectric power, compressed air energy storage,
- It is assumed that carbon capture will only be applied to coal plants and that these will be base-load operation. The possibility of CCGTs with post-combustion capture and amine storage is not considered.

5.4 UK Electricity Demand

5.4.1 Data Source

From National Grid public data(1) it is possible to view the dramatic changes in electricity demand that the grid experiences. The data used here is referred to as the INDO figures, which exclude Station Load, Pumped Storage Pumping and Interconnector Exports. Thus it is representative of the demand placed on the grid by its customers, not taking into account grid balancing technologies such as pumped storage or import/export to/from the UK. A 12 month time period has been considered dating back from the most recently available data, plotting demand every 30 minutes (November 2012).

Peaks and troughs range from 55.5 MW at 18:00 on Thursday the 8th February 2012 to just 17.9 MW at 06:00 on Saturday the 28th July 2012 and are most noticeable over 3 time periods:

- Annual
- Weekly
- Daily

5.4.2 Annual Variations

The following chart (Figure 9) was generated by summarising the half-hourly National Grid data into a minimum, maximum and mean value for each day, thus plotting 3 rather than 48 figures for each day of the year shown.

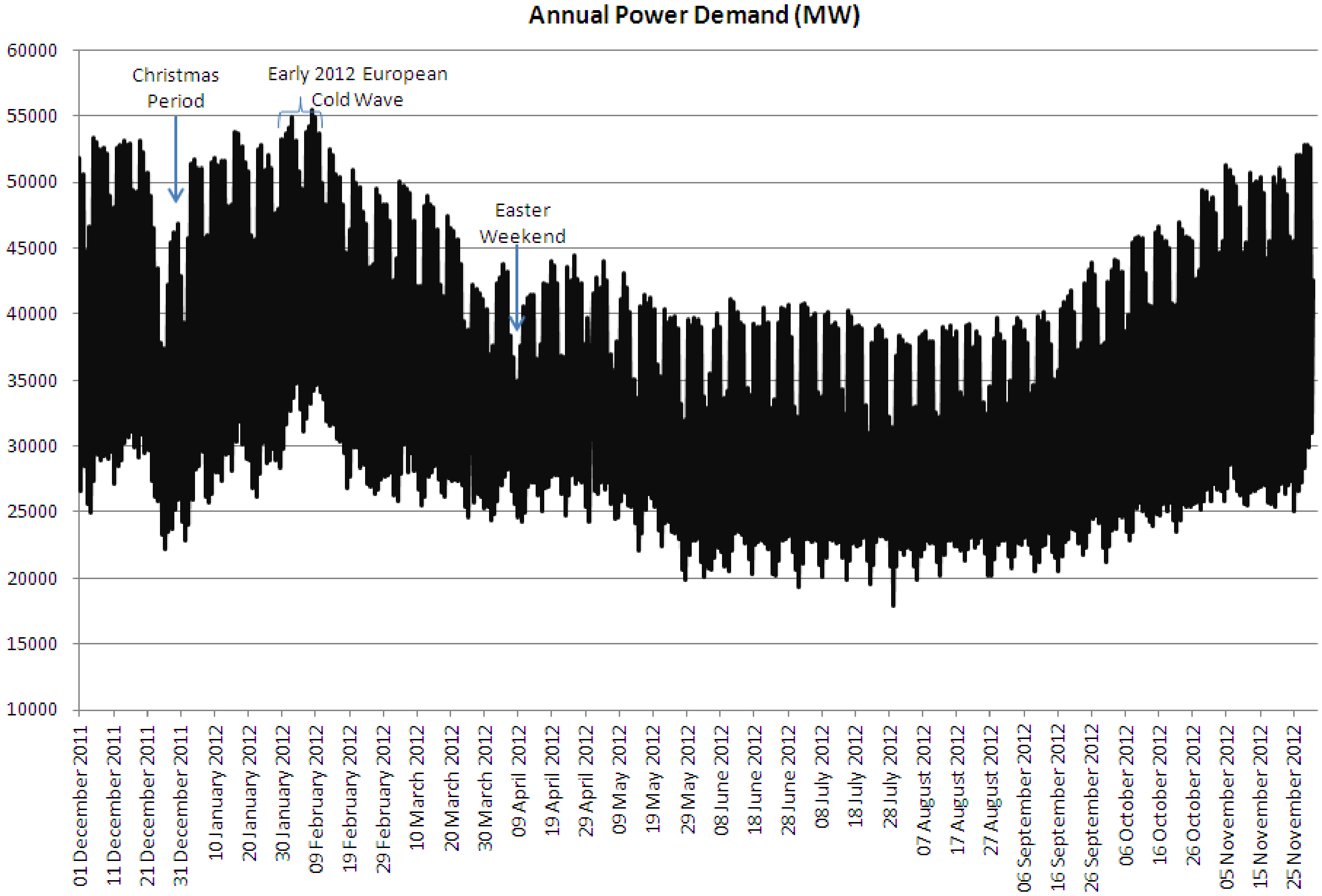


Figure 9 - National Grid data for UK annual power demand

It can be seen that UK power demand is lower in the summer than in the winter and that the difference between minimum and maximum daily demand seems to be somewhat reduced in the summer months. The average daily demand ranges from 26.5 GW in summer to 45.9 GW (173% increase) in winter while the extremes of demand vary from a 17.9 GW trough in the summer to a 55.5 GW peak (310% increase) in the winter.

Figure 9 also illustrates that there is a significant reduction in power demand on a weekly basis, over the weekend, but also that the largest variation in demand is on a daily basis.

There are two periods which appear to vary from the prevailing annual pattern of demand:

- The Christmas and New Year period shows a marked reduction in power demand, and
- A 2 week period in February appears to have a rather elevated demand for power, this timing coincides with an extreme weather event across Europe which has been referred to as the Early 2012 European Cold Wave, from 27th January to approximately 10th February, the last few days of which in the UK were marked by high snowfall, record low temperatures (-16°C) and significant disruption.

Long weekends such as the Easter weekend are also visible, but are much less obvious than the Christmas period.

The summer trend period of lower demand appeared to last from approximately mid-March to mid-September with average demand in summer of around 26.5 GW and 45.9 GW in the winter, nearly double in the winter compared to the summer. The period of highest demand was from late November to mid-February, depending on timing of particularly cold weather. The period of lowest demand was from mid-May to early September.

5.4.3 Weekly Variations

The following two charts (Figure 10 & Figure 11) allow the weekly variation in power demand to be visualised, for a month in summer and a month in winter:

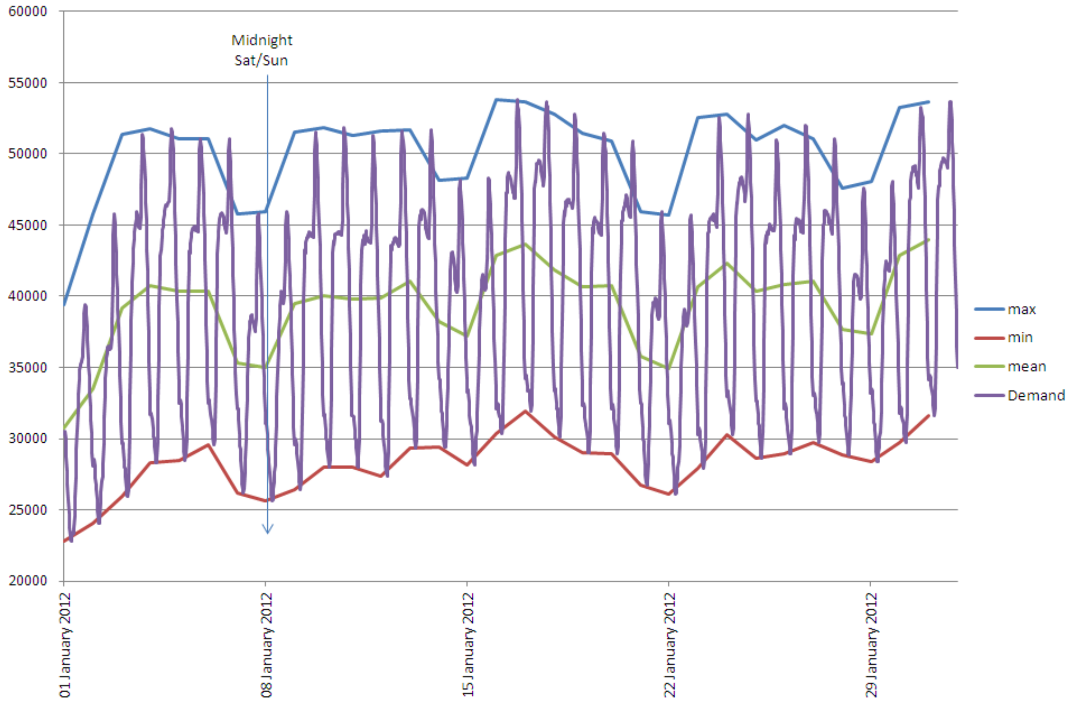


Figure 10 - Power Demand for January 2012 (MW)

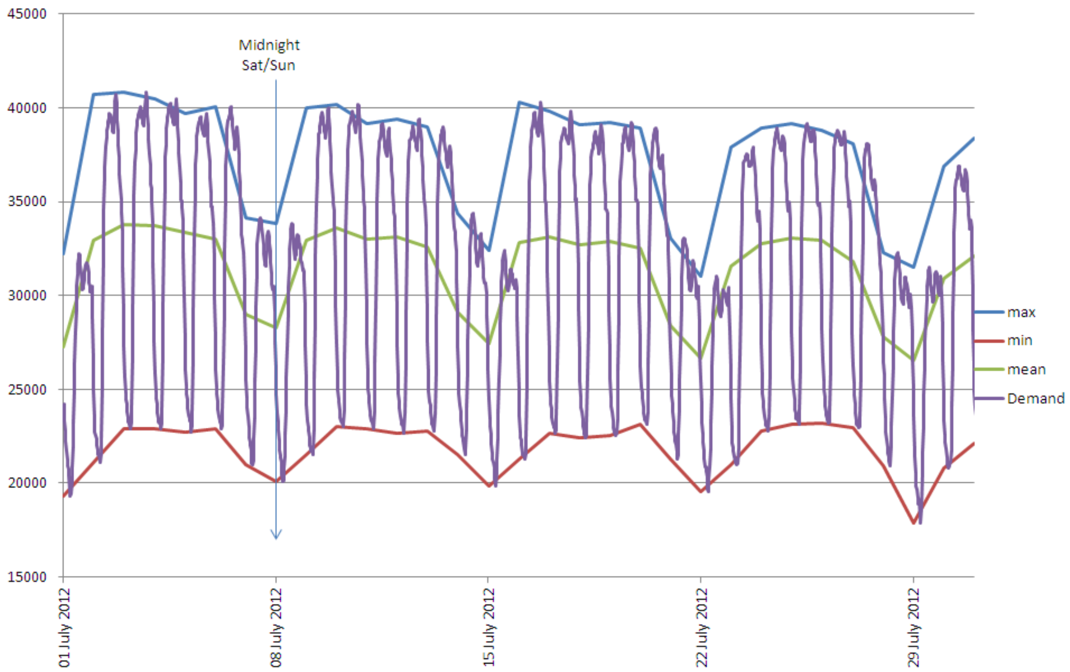


Figure 11 - Power Demand for July 2012 (MW)

It is clear from the above data that the UK power demand reduces at the weekend, by about 6 to 16% in the winter and 14 to 16% in the summer as a daily average. The daily weekend peak demand in winter is much closer to the week-day peak demand, however.

5.4.4 Daily Variations

The following two charts (Figure 12 & Figure 13) allow the daily variation in power demand to be visualised for the summer and winter cases.

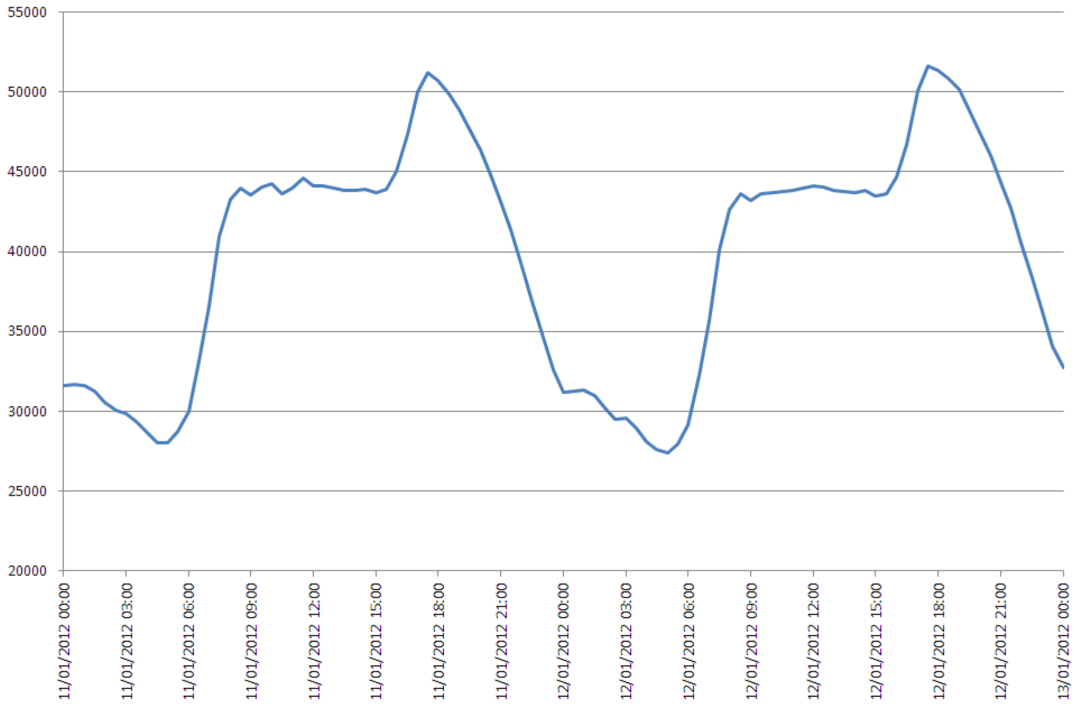


Figure 12 - Power Demand for 2 days in January 2012 (MW)

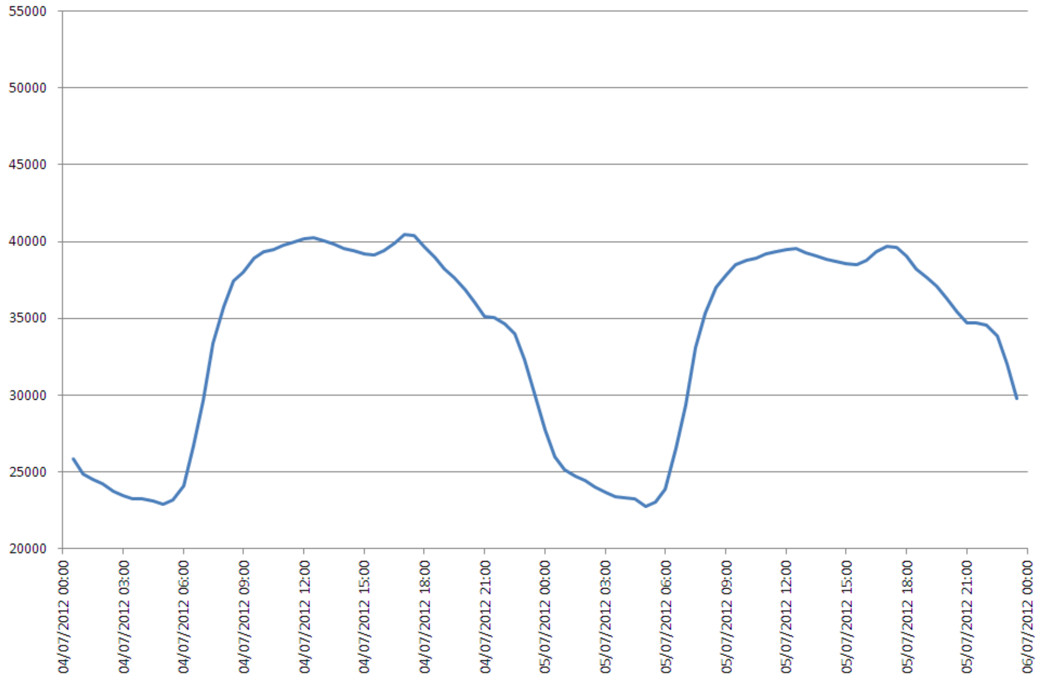


Figure 13 - Power Demand for 2 days in July 2012 (MW)

It can be seen from Figure 12 & Figure 13 above that both the shape and the magnitude of daily power demand variations changes between summer and winter. In the winter the curve is characterised by a very significant peak at 18:00 every day, including at the weekends. This peak is still visible in the summer, as a small blip which may or may not exceed the daily peak at around midday. For both winter and summer, the lowest power demand is experienced at about 05:00.

In the winter the daily peak demand is about 51 GW, with a minimum daily demand of about 28 GW, and average of 40 GW, so the peak is 180% of the minimum. In the summer the daily peak demand is about 39 GW, with a minimum daily demand of about 23 GW, and average of 33 GW, so the peak is 170% of the minimum.

5.4.5 Power Demand Conclusions

Total power demand in 2012 varied from a maximum of 55.5 GW to a minimum of 17.9 GW, while the average daily demand ranged from 26.5 GW in summer to 45.9 GW in winter (173% increase).

If we consider a plant with two gas turbines, in order for flexible operation to be applied to respond to seasonal variations in power demand the following annual mode of operation could be considered:

- Both GTs off for 4 months in summer
- 1 GT running in autumn/spring
- 2 GTs running 3 months in winter

In order to respond to the weekend drop in power demand the plant could be designed to operate with 2 GTs during the week and one GT at the weekend, sizing the syngas production to produce sufficient fuel to support this operating pattern.

A similar variation between maximum and minimum applies to the daily variation in 2012. The shape of the demand curve shows that it would be sensible for the plant to be operating at full load for 18 hours, from 06:00 until 24:00 or perhaps 15 hours from 06:30 until 21:30 then reduce to 1 CCGT or one or both CCGTs on minimum stable generation (MSG) during the night.

The daily variation scenario would result in the lowest volume syngas store, however, the following practicalities of diurnal operation would need to be considered:

- Would the low plant efficiency during the night (e.g. MSG) have a severe effect on the overall economics?
- Would the plant be off during the night, and what is the efficiency impact of start-up and shutdown on a daily basis?
- Is the GT life or maintenance impacted by ramping up and down frequently?

It is known that a natural gas fired CCGT can provide very flexible operation since they currently perform the vast majority of the load balancing for the UK grid. A certain amount of pumped storage and import/export already helps the UK grid to cope with matching supply with demand, but its total magnitude is small compared to a daily swing of 15 to 25 GW.

5.5 Matching Power Supply and Demand

In order to maintain the frequency of the UK electricity grid at 50Hz, electricity supply must match electricity demand closely at all times. National Grid is responsible for achieving this and is continually forecasting the likely demand. In case of a sudden drop in electricity production, National Grid has a number of contingencies which are able to respond to stabilise the grid until a replacement large unit can come online. These contingencies include:

- Fast Reserve: rapid delivery within two minutes of increased generation or reduced demand, sustained for a minimum of 15 minutes.
- Fast Start: units that start from standstill to deliver power within five minutes automatically, or seven minutes of a manual instruction, maintained for a minimum of four hours.
- Demand Management: reduction of demand of at least 25MW from large users of power, for at least an hour.
- Short Term Operating Reserve (STOR): generation of at least 3MW, from a single or aggregation of sites, within four hours of instruction and maintained for at least two hours.
- BM Start-Up: mainstream major generation units maintained in either an energy readiness or hot standby state.

These contingencies are designed to cope with the failure of one of the two largest single electricity sources on the UK grid, Sizewell B nuclear power station (1,260 MW) or one cable of the HVDC Cross-Channel interconnector (1,000 MW). A Blackout occurred in May 2008 when Longannet and Sizewell B tripped within 5 minutes of each other causing a widespread loss of power for 20 minutes while the grid re-stabilised.

Excluding trips, the National Grid can forecast with some accuracy the likely power demand at any time. Operators can plan to have their most cost-effective plants running first, then bring more expensive to operate plant on-line as demand increases. Nuclear plants are used to provide base-load power generation since they are the least flexible. Plants can even be mothballed and de-mothballed to cope with longer term variations such as an expected high winter peak 6 months in the future.

It can therefore be concluded that although demand varies considerably on an hourly basis, it is not unpredictable. It should therefore be reasonable to expect a fairly regular pattern of operation for a low-carbon plant with hydrogen storage and flexible turbine system.

5.6 Storage Volume Requirements

Approximate syngas storage volumes can be calculated based upon flow rate requirements for a base load gasification plant. Gas turbines operate most efficiently at full load, so in order to calculate a syngas consumption rate it is sensible to assume full load of a single machine.

Foster Wheeler experience has shown that the syngas required for a single GE Frame 9F syngas machine at full load, for example, is as follows:

Table 13 - Typical Feed Rates of Hydrogen Rich Fuel and Dilution Nitrogen

	Hydrogen Rich Fuel	Dilution Nitrogen
Flow Rate	12,287 kmol/h	11,067 kmol/h
	72,000 kg/h	310,000 kg/h
Molecular Weight (kg/kmol)	5.86	28.01
Composition (mol %)		
Hydrogen	86.7	0.0
Nitrogen	6.7	100.0
Carbon Monoxide	4.0	0.0
Carbon Dioxide	1.8	0.0

For cases 1, 2 & 4, dilution nitrogen will be produced by a continuous process and therefore requires storage, similar in volume to the hydrogen storage, in order to supply an intermittently operating power island. For case 3 the hydrogen–nitrogen generated by the ATR unit is stored as a mixture.

Considering these approximate flow rates it is possible to calculate approximate storage volumes (per GT) required for a number of different operating scenarios.

The stability requirement of the salt cavern puts limitations on the shape, storage volume and the maximum and minimum operating pressures. Maximum pressure should be typically 0.7-0.8 of lithostatic pressure so that the pressures do not exceed a safe limit below the fracture pressure of the halite. The minimum pressure is governed by the need to keep cavern wall convergence (salt creep) to acceptable limit. The minimum pressure is maintained using cushion gas which is the volume of gas intended as permanent inventory in a storage reservoir to maintain adequate pressure and deliverability rates throughout the withdrawal season. In caverns, the cushion gas is also necessary to ensure stability.

Storage pressure and percentage of cushion gas in the storage site are two major factors which will dictate the size of the salt cavern. Wide variation of cushion gas percentages have been reported in literature and also identified by British Geological Survey (BGS) based on operational projects. Plaat et al. (2009) stated typical cushion gas requirement of one third of total gas stored in a full cavern with the working gas comprising around two thirds. Hart (1997) and Taylor et al. (1986) reported that the cushion gas that occupies the underground storage volume at the end of the discharge cycle be as much as 50% of the working volume.

Three companies in the USA, ConocoPhillips, Praxair and Air Liquide, currently store hydrogen commercially underground in salt caverns constructed within salt domes in Texas (Parsons Brinckerhoff, 2013):

- The ConocoPhillips storage facility (storing Hydrogen since 1980) has a total H2 storage capacity of 64,400,000 m³ with a working volume (usable hydrogen capacity) of 30,200,000 m³. This facility uses 53% cushion gas to maintain the minimum pressure of storage site. The maximum cavern discharge rate is 1,420,000 m³/day which is 4.7% of working volume.
- The Air Liquide storage facility is designed to have a working volume capacity of 85,000,000 m³ of hydrogen, with another 57,000,000 m³ of cushion gas (40% of total capacity) required to ensure that the cavern remains structurally stable during operations (Parsons Brinckerhoff, 2013). The cavern is designed with the capability of an instantaneous peak withdrawal of 3,680,000 m³/day which is 4.3% of working volume or a sustained average of about 1,700,000 m³/day which is 2% of working volume (Air Liquide, 2008).

There is also an operational salt cavern in Holford, Cheshire, UK (storing natural gas) which is a “fast churn storage facility”. The gas withdrawal rate per day is 10-15% of the storage capacity to provide high flexibility, fast cycling products capable of supplying ‘peak gas’. This fast cycle withdrawal facility obviously makes a significant difference to the volume of gas need to be stored and makes the diurnal caverns significantly bigger.

In this study, we are looking into several diurnal variations which give the smallest cavern sizes, as well as other operating regimes up to seasonal variations. It can be surmised from the information obtained from the literature and also from existing operational facilities that for fast churn facility (daily withdrawal), working volume available for withdrawal is maximum of 10% of the total stored gas. For other operating modes, the cushion gas requirement will vary in the range of 40 to 80 % depending on the type of gas stored.

Table 14 summarises several existing and planned salt cavern including operational pressure regime. For this study, it has been assumed that only 10% of the total stored gas is available to withdraw per day in diurnal operating modes and an average of 60% cushion gas is required for seasonal operational modes. For weekly operational mode assuming daily withdrawal over 5 days, the cushion gas requirement has been calculated at 67% of the total stored gas; so that on 5th day, it will be 10% withdrawal of total stored gas in cavern.

Gas working volume (using proven storage pressures from Table 14) and total storage volume for salt caverns at various operating pressures have been calculated and summarised in Table 15, Table 16, Table 17 & Table 18. For diurnal operation, 90% cushion gas is used (as only maximum of 10% withdrawal/day is possible) whereas for seasonal variation, cushion gas requirement is used as 60% (based on literature). For weekly operational mode assuming daily withdrawal over 5 days, cushion gas being calculated to be 67% of the total stored gas; so that on 5th day, it will be 10% withdrawal of total stored gas in cavern.

Salt cavern operating temperature is also an important operating factor in order to preserve the structural integrity of the storage cavities. Caverns must be operated within a strict temperature envelope. The operating temperature of a Teesside salt cavern storing propane varies between -5 - 27°C, whereas a salt cavern storing butane varies between 0 - 29°C. Operation outside of these temperature ranges (and sudden changes in temperature) can increase the risk of a cavity becoming unstable, resulting in roof falls etc. Such damage has the potential to put affected cavities permanently beyond use. An existing salt cavern under operation in Portugal storing natural gas operates at 45°C. For this study, a cavern operating temperature of 45°C has been assumed for all cases.

Table 14 - Operational and Planned Salt Cavern (Information from BGS)

Area	Storage Gas	Storage Capacity (Mcm)	Approximate Cavern Volumes (m^3)	Number of Cavern	Operating Pressure ranges (bara)
Clemens Terminal, Texas, USA (ConocoPhillips) (1)	Hydrogen	64.4	580,000	1	135-70
Moss Bluff Field, Texas, USA (Praxair) (1)	Hydrogen	0.56 (3)	566,000	1	148-31
Teesside, UK	Hydrogen	210 (70 mcm each)	70,000	3	50-45 (Operates in brine compensated mode)
Halford, Cheshire Basin, UK (1)	NG	160-170	300,000	8	105-40
Aldbrough, East Yorkshire, UK (1)	NG	325	270,000	9	270-120
King Street, Cheshire Basin, UK (2)	NG	600	500,000-850,000	11	60 (Fast cycle)
North of Aldbrough, East Yorkshire (2)	NG	400-420	270,000	10	320-100
Preesall, NW England, UK (2)	NG	900	310,000	19	Shallowest Cavern: 61 - 22 Fast cycle) Deepest Cavern: 92-33 (Fast cycle) (Based on 0.83 and 0.3 lithostatic pressure)

Notes: (1) Operational salt Cavern

(2) Planned Salt Cavern

(3) FW suspect this is a cavern volume rather than a stored gas volume

Table 15 - Approximate Storage Volumes at 60 bara and 45°C

Operating Mode	GT offline	Hydrogen Store Volume		Nitrogen Store Volume		Combined H2/N2 Store Volume		Syngas Plant Capacity (4)
(single GT)	(hrs)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	kg/h H_2 rich fuel gas
Reference case (100% 24 hours)	0	-	-	-	-	-	-	72,000
Diurnal (100% 18 hours)	6	26,087	260,870	21,824	218,242	47,896	478,963	54,000
Diurnal (100% 15 hours)	9	32,609	326,087	27,280	272,802	59,870	598,704	45,000
Diurnal (100% 12 hours)	12	34,783	347,826	29,099	290,989	63,862	638,618	36,000
Weekly (off weekends)	48	198,758	596,273	166,279	498,838	364,925	1,094,774	51,429
Seasonal (off 4 months)	2920	11,285,024	28,212,560	9,440,968	23,602,420	20,719,606	51,799,016	48,000

Seasonal (off 6 months)	4380	12,695,652	31,739,130	10,621,089	26,552,722	23,309,557	58,273,892	36,000

Notes: (4) Assuming syngas plant operating constantly throughout the year.

Table 16 - Approximate Storage Volumes at 105 bara and 45°C

Operating Mode	GT offline	Hydrogen Store Volume		Nitrogen Store Volume		Combined H2/N2 Store Volume		Syngas Plant Capacity (4)
(single GT)	(hrs)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	kg/h H_2 rich fuel gas
Reference case (100% 24 hours)	0	-	-	-	-	-	-	72,000
Diurnal (100% 18 hours)	6	15,154	151,543	12,568	125,676	27,695	276,945	54,000
Diurnal (100% 15 hours)	9	18,943	189,429	15,709	157,095	34,618	346,182	45,000
Diurnal (100% 12 hours)	12	20,206	202,058	16,757	167,568	36,926	369,261	36,000
Weekly (off weekends)	48	115,462	346,385	95,753	287,259	211,006	633,018	51,429
Seasonal (off 4 months)	2920	6,555,659	16,389,149	5,436,637	13,591,592	11,980,452	29,951,130	48,000
Seasonal (off 6 months)	4380	7,375,117	18,437,792	6,116,216	15,290,541	13,478,009	33,695,022	36,000

Notes: (4) Assuming syngas plant operating constantly throughout the year.

Table 17 - Approximate Storage Volumes at 135 bara and 45°C

Operating Mode	GT offline	Hydrogen Store Volume		Nitrogen Store Volume		Combined H2/N2 Store Volume		Syngas Plant Capacity (4)
(single GT)	(hrs)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	kg/h H_2 rich fuel gas
Reference case (100% 24 hours)	0	-	-	-	-	-	-	72,000
Diurnal (100% 18 hours)	6	11,925	119,249	9,859	98,587	21,932	219,316	54,000
Diurnal (100% 15 hours)	9	14,906	149,061	12,323	123,233	27,415	274,145	45,000
Diurnal (100% 12 hours)	12	15,900	158,999	13,145	131,449	29,242	292,422	36,000
Weekly (off weekends)	48	90,857	272,570	75,114	225,341	167,098	501,294	51,429
Seasonal (off 4 months)	2920	5,158,631	12,896,577	4,264,782	10,661,955	9,487,454	23,718,636	48,000
Seasonal (off 6 months)	4380	5,803,460	14,508,649	4,797,880	11,994,700	10,673,386	26,683,465	36,000

Notes: (4) Assuming syngas plant operating constantly throughout the year.

Table 18 - Approximate Storage Volumes at 270 bara and 45°C

Operating Mode	GT offline	Hydrogen Store Volume		Nitrogen Store Volume		Combined H2/N2 Store Volume		Syngas Plant Capacity (4)
(single GT)	(hrs)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	Working (m^3)	Total (m^3)	kg/h H_2 rich fuel gas
Reference case (100% 24 hours)	0	-	-	-	-	-	-	72,000
Diurnal (100% 18 hours)	6	6,312	63,121	5,290	52,901	11,638	116,385	54,000
Diurnal (100% 15 hours)	9	7,890	78,901	6,613	66,126	14,548	145,481	45,000
Diurnal (100% 12 hours)	12	8,416	84,161	7,053	70,535	15,518	155,179	36,000
Weekly (off weekends)	48	48,092	144,277	40,306	120,917	88,674	266,022	51,429
Seasonal (off 4 months)	2920	2,730,567	6,826,417	2,288,459	5,721,148	5,034,710	12,586,775	48,000
Seasonal (off 6 months)	4380	3,071,888	7,679,719	2,574,516	6,436,291	5,664,049	14,160,122	36,000

Notes: (4) Assuming syngas plant operating constantly throughout the year.

Storage capacity listed in Table 14 is the volume of the gas at STP that can be stored in a cavern. Actual cavern size is much smaller depending on their corresponding storage pressure.

According to Plaat et al.(3), however, caverns created in salt domes usually range in size from 300,000m^3 to 700,000m^3, while those created in salt beds usually have volumes in the range of 100,000m^3 to 300,000m^3.If only salt beds are available for many suitable locations for a power plant in the UK then this may limit the project to diurnal and weekly operation rather than seasonal hydrogen rich fuel storage.

To accommodate the seasonal cases, several caverns will be required if gas is being stored at low storage pressure. As the storage pressure increases, the storage volume decreases substantially; cavern size will be around 4 times smaller if storage pressure increases from 60 bara to 270 bara. So at higher storage pressure, the project will be more flexible to accommodate seasonal variations with lesser number of multiple caverns. Overall cavern construction costs are not expected to vary proportionally with cavern size, so there should be significant economies of scale associated with projects using larger (or multiple) caverns.

It would seem sensible to build a bigger storage facility with higher storage pressure within the lithostatic pressure limit required to maintain structural integrity.

5.7 Other Requirements

Size is only one requirement for a salt cavern to be a well functioning, safe and economic gas store. Other important considerations include:

- Safe containment
- Short and long term structural stability.
- Limited volume decrease due to salt creep.
- Contaminants

- Location

5.7.1 Safe Containment, Structural Stability and Salt Creep(3)

As this project deals with hydrogen, of course safety is big concern: the gas is, after all, highly flammable, explosive and difficult to contain because of the very small dimensions of hydrogen molecules. Suitable storage pressure needs to be examined carefully to make the cavern safe and economical.

In order to ensure long term structural stability the cavern needs to be constructed in as smooth a shape as possible and then be operated within a minimum and maximum allowable pressure range (approximately 0.3 - 0.8 of the lithostatic pressure, determined by the cavern depth). It is important to prevent loss of containment through the salt bed or minor fractures. Once a salt cavern has been created, its gas tightness has to be maintained and to ensure containment of the stored gas, a periodic mechanical integrity test should be performed as well as underground infrastructure should be monitored. This section will be discussed in more detail in WP2.

5.7.2 Contaminants

It is necessary to understand if any contaminants are likely to be found in the gases extracted from salt cavern storage so that the technical and economic implications of these contaminants can be taken into account in the plant design.

It is understood that in the first months of operation of the salt caverns it is likely that the gas leaving the caverns will be wetter than gas entering the caverns. This is due to residual water being present in the cavern as a result of the solution mining process. It is also anticipated that this additional water content will be saturated with salt, the materials of construction will therefore need to be selected considering this corrosive service, see also section 8.2.2.

For caverns which then operate in a dry regime (i.e. not using brine displacement injection/withdrawal to maintain cavern pressure as hydrogen is withdrawn/injected) it can be assumed that the composition of gas leaving the cavern will be the same as that entering the cavern.

If brine displacement is used to maintain the cavern at constant pressure then the withdrawn gas can be assumed to be saturated with water, which is in turn saturated with salt. For the downstream power island proposed for the flexible turbine operation, this would necessitate a water washing step between the cavern and the GT to remove the salt content.

Other contaminants may be possible due to impurities present in the salt, and these will be specific to any given halite formation, and will need to be considered on a location specific basis.

5.7.3 Other Location Factors

The geographical location of a salt cavern is also an important consideration, as a project will generally be more viable if it does not have to involve significant transportation of feedstocks, products, wastes or intermediates. Items to consider include proximity to:

- a location for brine sale or disposal at sea (within a reasonable distance);
- fuel source or fuel transportation infrastructure
  - national gas grid connection;

    - coal / biomass import terminal;
- electricity grid connection;
- CO_2 storage location, or connection to a CO_2 transportation hub;
- source of cooling water, sea water is preferred for maximum power plant efficiency.

6. REQUIREMENTS / OPTIONS FOR HYDROGEN TURBINES

6.1 Objective

The objective of this section is to review the current and future requirements for turbines being considered as part of the power island. Depending on the pressure of the salt cavern storage, two types of turbines may be required for the envisaged scheme.

For all salt cavern storage pressures, power generation gas turbines suitable for firing on a high-hydrogen syngas at the specified 350MWe scale will be required. The challenges and potential design solutions for hydrogen fired gas turbines (including the flexibility to burn both natural gas and mixtures with high hydrogen concentrations), and the availability of suitable machines will be characterised.

For a high pressure hydrogen storage case, in excess of 35 barg, installation of expansion turbine(s) may be justifiable, to recover much of the energy which was used to compress the hydrogen (and, in most cases, nitrogen also) into storage. The machines currently available on the market which could be used for this duty will be reviewed, and feedback will be obtained from potential vendors on development needed to bring such machines to the market place before 2030.

6.2 High Hydrogen Fired Gas Turbines

The availability of large gas turbines capable of firing undiluted hydrogen is some way into the future.

Existing types of gas turbines with power outputs >300MWe from GE and Mitsubishi require significant dilution of the hydrogen-rich fuel gas with nitrogen or perhaps steam. This requirement for dilution of hydrogen fuel is mainly due to the need to limit NOx emissions.

Smaller E-class gas turbines with large external combustion chambers such as Alstom GT11N2 and Siemens 2000E/Ansaldo V94.2 may require less dilution of the fuel, but they have not been investigated in this study due to their relatively low power output and thermal efficiency.

The modelling of hydrogen production and firing for this study is based on information provided by GE for the Frame 9F Syngas variant, giving a thermal efficiency of 34.4% using nitrogen as the diluent. The exhaust gas contains 25 vpmd NO_X at 15% O_2 content.

Mitsubishi have provided an estimate of performance for their 701F gas turbine. When inserting this MHI data into our model the resultant overall plant power output and thermal efficiency (34.9%) are marginally higher than the original information from GE (34.4%) but with higher NO_X at < 100 vpmd. This may require provision of a Selective Catalytic Reduction unit downstream in the HRSG.

So far, SCR in the UK has not been used to any extent if at all to treat gas turbine exhaust. However, SCR may be used in the future when gas turbines are brought into use to fire hydrogen-rich fuels, if after evaluation, the gain in GT output and thermal efficiency resulting from increased firing temperature outweighs the CAPEX and OPEX of the SCR catalyst reactor "basket" and its auxiliaries.

All the above-mentioned representations from the gas turbine suppliers are state -of - the -art in early 2013. Between now and the time hydrogen fired gas turbines are ordered around 2020, there will for sure be normal improvements in the capacity and thermal efficiency and power output of natural gas fired gas turbines from all the suppliers. Meanwhile, there could be parallel improvements with hydrogen-fired

GT's, but it seems likely that their efficiency shortfall relative to natural gas fired GTs will, if anything, grow over this period. Consequently, with natural gas feed, the already large gap in efficiency between conventional post combustion carbon capture efficiency (around 53%) and our ATR-based hydrogen fired Case3 (41%) is also likely to increase over the period. By 2020, other manufacturers, particularly Alstom, may be able to provide large hydrogen-fuelled gas turbines.

Cases 1 and 2, above will already require an air separation unit (ASU) to provide the oxygen required by the gasifier, and the nitrogen required for the gas turbine can be obtained from the ASU as a by-product. The Case 3 air-based ATR process delivers a syngas that already contains approx 45 mol% nitrogen, and consequently can be fired in the GT without need for further dilution. Only Case 4 will require an ASU specifically to provide nitrogen for dilution purposes.

Substitution of steam for nitrogen as the fuel diluent is of considerable interest to ETI, due the potential for elimination of nitrogen compression and nitrogen storage caverns. Foster Wheeler has made a preliminary simulation of replacing nitrogen with steam taken from the HRSG based on data received from GE. Gas turbine efficiency will be comparable or even slightly improved with steam dilution of the fuel, with the.lower firing temperatures needed to offset perceived higher gas turbine maintenance costs being largely offset by the elimination of nitrogen compression. Initial indications are that plant thermal efficiency with gasification may be 0.5% higher than with nitrogen dilution, without taking into account any de-rating of the gas turbine. Less attractive features of steam dilution include high water consumption (the stack gas contains approx 18 -20 % water vapour), increased tendency to a visible exhaust plume, perceived higher maintenance costs for the gas turbine and perhaps the HRSG also.

6.3 Hydrogen Power Recovery Turbines (Expanders)

The optimum pressure at which the hydrogen fuel and nitrogen diluent (if used) are to be stored is dependent on final location and depth of halite, amongst other things.

Higher storage pressures may make attractive the recovery of power by expansion of stored gas in an expansion turbine.

An expansion turbine is a centrifugal or axial flow turbine through which a high pressure gas is expanded to produce work. Because work is extracted from the expanding high pressure gas, the expansion is approximated by an isentropic process (i.e., a constant entropy process) and the low pressure exhaust gas from the turbine is at a very low temperature, depending upon the operating pressure and gas properties.

Interestingly, hydrogen exhibits a reverse Joule-Thomson effect, which means that instead of the temperature dropping along with the pressure, the temperature actually increases. This holds true when hydrogen is expanded through a valve over at typical pressure range of 50 to 30 bar results in a small temperature rise of about 1°C, but with expansion through a turbine over the same pressure, the gas is cooled from 50°C to around 15°C as the extraction of heat as mechanical work outweighs the much smaller J-T effect.

Foster Wheeler approached two potential suppliers of hydrogen expansion turbines: GE Nuovo Pignone & Atlas Copco. No response was received from Atlas Copco. Two budget proposals were received from GE Nuovo Pignone.

- For an expansion turbine to let down 93 kg/s of H_2/N_2 mixture from 50 bar to 33 bar, with 89% isentropic efficiency. This offer included a 4500 kW coupled electric generator. The scope of GE's supply comprises the radial-flow expansion turbine itself, integral reducing gear and a 4500 kW asynchronous generator, together with lubricating oil console and local controls. The budget cost of this equipment was € 2.2 million. This commercially available expansion turbine technology was originally developed for natural gas treatment applications.

- For a much larger expander, required to expand the hydrogen fuel plus nitrogen diluent from 253 bar to 30 bar, with a power developed of over 30MW .GE's proposed solution provides for six stages of expansion in two separate integrally geared expander-generator machine groups. While this type of expander technology is well proven, a judgement would have to be made at some point on the economic benefit and complexity of future gas - expansion schemes.

From this information, it appears feasible to recover power by expansion of a range of a hydrogen-containing and diluted hydrogen streams, as may be needed for the ETI study and subsequent developments.

Moreover if steam is substituted for nitrogen as the gas turbine fuel diluent, the resulting reduction in molecular weight of the stored fuel gas will make mechanical energy recovery significantly more difficult to achieve with existing types of expander.

Foster Wheeler believes that the 105 bara storage pressure case should not have any issue, but the pressure ratio and or inlet pressure may be an issue for the Yorkshire (270 bara) storage case. For the high pressure scenario, if necessary, a valve may be positioned upstream of the expander to take the high end of the total pressure drop.

7. ECONOMICS OF HYDROGEN PIPELINES

7.1 Objective

The objective of this section is to investigate the order of magnitude cost of hydrogen pipelines and risk factors impacting this cost, so that the impact of there being a significant distance between hydrogen source, store and power island on the project economics can be judged.

7.2 Hydrogen Pipeline Design

The investment cost of a pipeline includes materials, labour, right-of-way fees and miscellaneous costs. The cost of the pipeline material is then determined by the pipeline length, diameter, wall thickness and material of construction.

Typical design parameters required for calculations include the fluid composition, operating temperature, design temperature, operating pressure, design pressure and fluid flow rate. Other useful information includes the presence of any contaminants (particularly those which are toxic/corrosive) and other chemicals such as corrosion inhibitors. The pipework line size (outer diameter), wall thickness and MOC can then be established.

For the case of transporting gaseous hydrogen it is necessary to consider hydrogen embrittlement in the material selection and wall thickness. The wall thickness and diameter are then affected by the operating pressure (or pressure range) and the flow rate of gas.

The operating and design pressure / temperature of the hydrogen pipeline is dependent on the operating pressures of the salt cavern and the gas turbine.

In this instance, two design cases were considered for a range of pipe sizes (6” to 60”) to cover all anticipated conditions in the system:

- hydrogen at ambient temperature (design temperature) and 10 bar (design pressure);
- hydrogen at 120°C and 200 bar;

Based on previous hydrogen pipeline experience, it is assumed that water is present only in the vapour phase, which along with the addition of a corrosion inhibitor eliminates the requirement for corrosion resistant materials or corrosion resistant alloy (CRA) lined pipework, allowing a carbon steel with a 3 mm corrosion allowance to be used.

The pipeline wall thickness calculations follow pipeline code ASME B31.8 and are to some extent dependent on design factors. Pipeline design depends not only upon fluid conditions, but also upon the physical location of the pipework. The factors are applied depending on location, population density and dwellings in a defined area (generally the higher the population density, the lower the design factor value and the greater the wall thickness). Three design factors (0.72, 0.6 and 0.5) were considered in the calculations to highlight the effect on wall thickness.

Temperature is not defined within the calculations until above 121 °C, at which point a de-rating factor is considered, however it is not considered in this study, as initial research has suggested that cavern storage temperatures should be kept below 120°C.

Other key assumptions include:

- API 5L X65 grade specified at 18" and above,
- longitudinal weld design factor 1.0 applied,
- maximum gas design temperature below 120 °C.

7.3 Hydrogen Pipeline Costs

Order of magnitude costs per km have been derived for two pipeline scenarios, with design factors of 0.5, 0.6 and 0.72 and with various diameters (from 6” to 60”):

- Gas @ Ambient temperature and 10bar
- Gas @ 120°C and 200bar

Summaries of costs/km for various diameters of onshore hydrogen pipeline are provided in Attachment 22.

The estimates represent an indicative cost in a generic NE England location for the supply and below ground installation of the lines. No allowance is included for any pipe crossings or bridges, which can be a significant contributor to TIC.

7.4 Brine Pipeline Costs

Another significant cost which is not captured elsewhere is the brine pipeline cost. Depending on the proximity of the proposed plant and cavern location to the sea, brine pipelines may be required to transport seawater for cooling and/or brine for solution mining.

For the brine pipeline, there are multiple options available depending on fluid conditions:

- Considering aerated brine < 10 bar, high density polyethylene (HDPE) or glass reinforced plastic (GRP) may be used;
- Considering aerated brine > 10 bar, fusion bonded epoxy (FBE) internally lined pipework may be used;
- Considering de-aerated brine, carbon steel with a 3 mm corrosion allowance is deemed appropriate as in the case of the hydrogen pipelines. It is also assumed that the stream is free of CO_2 and other corrosive substances (apart from salt).
- The pipeline wall thickness calculations for liquids follow pipeline codes ASME B31.4. Three design factors (0.72, 0.6 and 0.5) were considered in the calculations without any net effect on wall thickness.

Order of magnitude costs per km have been derived for carbon steel pipelines and HDPE pipelines transporting de-aerated brine at 40°C and 20bar, with various diameters (from 6” to 60”):

Summaries of costs/km for various diameters of onshore brine pipeline are provided in Attachment 23.

The estimates represent an indicative cost in a generic NE England location for the supply and below ground installation of the lines. No allowance is included for any pipe crossings or bridges, which can be a significant contributor to TIC.

7.5 Offshore Pipeline Costs

If an offshore option of the hydrogen pipeline is selected/favoured, the submarine pipeline has to be strong enough to withstand all the loads that will be applied to it both during its construction and when in operation. The design must be able to withstand functional, environmental, and construction loads and anticipated combinations of these loads.

When it goes into operation, it must withstand the internal pressure from the hydrogen it is carrying, external pressure from the sea (water depth expected to be less than 50 meters), and stresses induced by temperature changes. In addition, the submarine pipeline needs be able to withstand external impact loads from anchors and fishing gear.

The design of the offshore pipeline cannot be undertaken without due consideration being given to the requirements of, and impact on, the necessary pump/compressor facilities, terminals and associated equipment and facilities. During the initial pipeline design phase the offshore pipeline system configuration needs to be addressed.

An important factor is that during the material selection phase due regard shall be given to the limitations of some materials that may impose on the choice of installation vessel and method, e.g., the use of CRA or clad pipe or thick wall thickness (above 30 mm) reduces the lay rate and increases installation costs.

Selection of the pipeline codes or standards to be used for the hydrogen pipeline design, construction, and operation is principally influenced by:

- The country in which the pipeline is to be built/installed – in this case UK
- Regulatory constraints
- Partner/Client preferences

If cost is the determining factor in the choice of the pipeline code for a project, it is recommended that a preliminary pipeline cost estimate is prepared for each recommended pipeline code during the conceptual phase.

The estimated length of pipeline, design pressure and temperature, pipeline location, the number of major crossings along the route; need to be estimated before determining the design parameters and related quantities and costs.

In general, offshore pipeline costs are considerably greater than those for onshore pipelines, and are dominated by installation costs, as the rental of the barges and crew are expensive, and the laying rate and availability of pipelaying barges is critical. This means the correlation between diameter and cost is less strong.

On a recent offshore pipeline project within Foster Wheeler’s experience, involving a carbon steel pipeline of 48” diameter, 22mm wall thickness and material grade of API X65, the cost of materials, fabrication, construction and installation amounted to 7.5 million USD per km (£5m/km). Of that, approx £1.5m/km would be the materials cost, so the installation cost will be approx £3.5m/km, more than 3 times the equivalent onshore installation costs.

8. EFFECTS OF HYDROGEN PURITY

8.1 Objective

The objective of this section is to provide a brief discussion of the techno-economic and HSE impacts of hydrogen purity on the production, storage, and consumption processes involved in the proposed scheme.

Topics of discussion initiated here will have an effect on the technologies discussed in WP3, some of which are very sensitive to variations in hydrogen purity.

8.2 Techno-economic Impact

8.2.1 Hydrogen Production

The level of hydrogen purity produced in the proposed hydrogen generation processes varies. The purest hydrogen stream is produced from Case 4 (SMR) – 100 mol% hydrogen followed by Cases 1 and 2 (Gasification) – 89 mol% hydrogen. The least pure stream is produced in Case 3 (ATR) – 53 mol% hydrogen.

There are some limitations on the commercial availability of machines to compress pure, or nearly pure, hydrogen to very high pressures (circa 285 bara). Work has been done recently by a US development firm, in conjunction with Mitsubishi, on compressing hydrogen to around 80 bar in a centrifugal compressor. This technology should regarded as a pre-commercial development, but at the time when a full-scale hydrogen storage project is implemented it may have been developed to point at which it can be used in the new facility.

For Case 4 SMR, producing pure hydrogen from the PSA unit, it therefore appears feasible in the medium term to compress the hydrogen product to around 100 bar in a centrifugal compressor, but a reciprocating compressor will probably be needed for the final stage of compression from 100 to 285 bara.

For the gasifier-based cases (Cases 1 & 2), in which the produced hydrogen is impure and has a molecular weight around 5.5, it can be assumed that compression to 285 bara can be achieved in a centrifugal compressor using existing technology. For reference there are several ammonia plants in Russia designed around 1970 in which the synthesis gas with molecular weight of 9 is compressed to over 300 bar in centrifugal compressors.

From an operational stand point, the purity of hydrogen produced by each technology case discussed should not be significant for the correct function of downstream equipment, as the storage and combustion processes are flexible (discussed below), in fact the volume of dilution gas can generally be varied during operation depending on hydrogen purity to achieve the same input composition to the gas turbine. However, if the hydrogen syngas were to be used for applications other than for combustion for power generation (as discussed in WP3), complications arise.

Other alternative end-uses of hydrogen highlighted in WP3 include: domestic use as fuel in homes (supplied via the National Gas Grid); for transportation; and as a chemical feedstock.

For combustion in a domestic environment, it is likely that high purity hydrogen would be required in order to minimise capacity losses in the National Grid system (as outlined in WP3, the capacity of the grid would be reduced by the addition of hydrogen gas). However, a very small amount of impurities could potentially be tolerated.

The Polymer Electrolyte Membrane (PEM) fuel cells used in hydrogen powered modes of transportation are poisoned by very low concentrations of CO, so require feeds of pure hydrogen to be economically viable.

Similarly, for use as an industrial feedstock a pure feed of hydrogen is required in order to reduce downstream processing of entrained impurities. In general, companies looking to procure a source of hydrogen for use as an industrial feedstock will not accept a gas containing lower than a very high purity level, for example BOC supply bulk hydrogen at a purity of >99.98%(5).

Consequently, the high hydrogen syngas produced from every design case excepting Case 4 would require further purification before it could be used for any of the highlighted alternative technologies. Further purification of the product, e.g. by pressure swing adsorption (PSA) can be provided to meet the purity requirement of a particular user.

For the gasification cases, where it is assumed that the cavern would store hydrogen-rich gas separately, the PSA Unit equipment cost would be in the order of £7m. For the ATR case, where the nitrogen and hydrogen are stored together, the PSA Unit equipment cost would be in the order of £10m.

8.2.2 Hydrogen Storage

Each of the different hydrogen production technology cases generates a different purity of hydrogen and thus in each case, a different gas composition is stored in the salt cavern.

All simulation cases with the exception of the SMR case have assumed that a small fraction of water is present in the input and output streams of the salt cavern (about 0.15 mol%) and that the input and output streams contain the same fraction of water.

In reality, the salt cavern will contain residual brine from the solution mining process, which for an initial period of operation will saturate the stored gas and exit along with the gas when the cavern is evacuated along with traces of salt. Chloride is the usual problem with the brine/ salt caverns as it will corrode the pipe work, so suitable provision needs to be made for this.

When the gas contains CO_2, it will dissolve in the water to form carbonic acid. Carbonic acid will not affect the salt cavern as it is formed from a strong acid and strong base (HCl and NaOH) to give NaCl. The weak acid cannot displace the chloride ions within the salt cavern. However, carbonic acid will affect the material of construction of the well head, downstream pipework and associated processing equipment. This saturated, acidic, corrosive gas output stream can cause downstream complications unless protective measures are put in place (see following section on combustion).

Brine containing dissolved oxygen is more corrosive than brine alone, so measures should be put in place to prevent oxygen entrainment, however under normal operation the high pressure syngas will not contain oxygen. Water should only be present in the outlet gas for an initial period whilst the newly excavated cavern is drying out after the solution mining process. Eventually, the cavern will dry out and the level of water in the outlet gas will be small.

Further to the impacts already mentioned, if the high hydrogen syngas contains a high percentage of gaseous impurities, the calorific value of the gas decreases (see following section on combustion) which means that in order to charge the salt cavern with a gas of equal calorific value to pure hydrogen, a larger amount of energy must be applied during gas compression upon entry to the cavern. This impact is realised through increased operating cost of the cavern.

8.2.3 Hydrogen Combustion

Generally, as long as a gas turbine inlet hydrogen purity level is defined, the actual level of purity is not a major issue for the processing of hydrogen in the flexible gas turbines outlined in this study. The currently available high-hydrogen GTs are designed to burn a high hydrogen syngas/dilution gas mixture covering a wide composition range, and are also capable of both starting up and running on natural gas whenever syngas is unavailable. Start-up of the GT using hydrogen is in principle possible, but it is understood that the suppliers currently expect the GT to be started up with natural gas, with switch-over to diluted hydrogen firing taking place after the combined cycle operation has stabilised.

Depending on the hydrogen purity contained in the syngas plant outlet stream, the volume of dilution gas can be altered to achieve the optimum syngas/dilution gas mixture, ensuring efficient combustion. In all simulation cases, approximately 50 mole% nitrogen dilution gas is modelled as a base case.

It is probable that the power output and efficiency of a gas turbine designed primarily for hydrogen firing will be reduced when firing natural gas, due to the lower mass flow of the natural gas fuel compared with hydrogen plus diluent. If frequent switching under load between hydrogen fuel and natural gas fuel is expected, it will be desirable to initiate a partial switch-over to the other fuel in advance of the anticipated switch-over time.

As mentioned in the hydrogen storage section above, as the mole fraction of syngas impurities increases, the calorific value of the gas mixture decreases. The presence of certain impurities also causes other complications.

A decrease in syngas calorific value will upset the ratio of hydrogen to dilution gas, which is set to achieve 25 ppmvd NO_x at the maximum allowable gas turbine firing rate.

The composition of the hydrogen-rich fuel itself pre-dilution can vary to a limited extent, as the gas turbine inlet conditions can be controlled by varying the flow of dilution gas. The currently available high-hydrogen GTs are designed to burn a high hydrogen syngas/dilution gas mixture covering a wide composition range, and are also capable of starting up and running on natural gas whenever syngas is unavailable. Fuel gas compositions should be closely controlled to achieve the optimal system efficiency and power output.

If the high hydrogen syngas extracted from the salt cavern is highly water/brine-saturated, traces of carbonic acid and salt can over time degrade the working surfaces of the gas burners and turbine blades, resulting in a decrease in system efficiency and the requirement for more regular maintenance/replacement. This makes material selection a highly important activity.

During the design of such a system, the working limits in terms of hydrogen purity and impurity levels should have a clearly defined range and operation conditions should closely resemble these conditions to achieve the optimum system efficiency and power output.

8.3 HSE Impact

8.3.1 Hydrogen Production

When considering the four technological variations covered in this study, hydrogen purity variation generally does not introduce any additional HSE issues in terms of hydrogen production as this part of the process is a closed system.

However, there are complications associated with the alternative end-uses of hydrogen as discussed in WP3, in particular with open system use in a domestic environment. End-use applications in the home, safely modified for pure hydrogen fired operation are perfectly acceptable from a HSE stand point, however if certain impurities such as CO are present in the gas stream, an HSE risk is introduced. When an impurity-laden gas stream is used in a typical domestic end-use application such as a gas stove burner, it is almost impossible to totally prevent some escape of gas into the atmosphere and could be inhaled by the user. CO is a toxic gas and should not be introduced to such as system if it can be avoided.

8.3.2 Hydrogen Storage

The presence of impurities in the hydrogen mainly causes trivial changes to the technical operation of the plant, and should have no significant impact on the storage system, as discussed above.

8.3.3 Hydrogen Combustion

The main HSE impact possible when considering high hydrogen syngas combustion is the processing of toxic impurities, leading to release of toxic gas to atmosphere and potential health risk. This is unlikely under normal operation conditions and indeed many typical impurities will be oxidised to a harmless form in the turbines, however, certain toxic contaminants when accidently entrained into the fuel gas stream (for example H_2S) can be oxidised to more harmful forms (SO2) and expelled to atmosphere. A good place to identify and discuss such issues further would be in an appropriately scoped project HAZID/HAZOP study.

9. REFERENCES

1) National Grid electricity demand data: http://www.nationalgrid.com/uk/Electricity/Data/Demand+Data/.

2) “Impact of Intermittency: How wind variability could change the shape of the British and Irish Electricity Markets”, Poyry, July 2009.

3) "underground gas storage: why and how", H. Plaat, from Evans, D.J. & Chadwick, R.A. (eds) "Underground Gas Storage: Worldwide Experiences and Development in the UK and Europe", The Geological Society, London, Special Publications, 313, 25-37.

4) "Gas Storage: An onshore operator’s perspective", A. Fernando & A. Raman, from Evans, D.J. & Chadwick, R.A. (eds) "Underground Gas Storage: Worldwide Experiences and Development in the UK and Europe", The Geological Society, London, Special Publications, 313, 25-37.

5) "The commercial future of the Ultimate Fuel", D. Hart, Hydrogen Power, UK: Financial Times Energy Publishing.

6) "A Technical and Economic Assessment of Methods for the Storage of Large Quantities of Hydrogen", J. B. Taylor, J. E. Alderson, K. M. kalyanam, International Journal of Hydrogen Energy. (11:1); pp. 5-22.

7) BOConline UK website: http://www.boconline.co.uk/en/index.html

8) http://www.hydrogen.energy.gov/pdfs/review12/pd016_heshmat_2012_o.pdf

ATTACHMENT 1

BASIS OF DESIGN

Contract Number:	1.17.13058
Client's Name:	The Energy Technologies Institute
Project Title:	Hydrogen Storage and Flexible Turbine Systems
Project Location:	Generic UK

REVISION	**0**	**1**	**Signature**
DATE	8th Mar 13	5th July 13	
ORIG. BY	A Price	T. Abbott	
CHKD. BY	S. Ferguson	S. Ferguson	
APP. BY	T. Abbott	T. Abbott	

Hydrogen Storage and Flexible Turbine Systems
Basis of Design

CONTENTS

1.0 INTRODUCTION

The Energy Technologies Institute has employed the services of Foster Wheeler to undertake a study titled “Hydrogen Storage and Flexible Turbine Systems”.

This purpose of this Basis of Design document is to provide a clear and consistent basis on which to evaluate each hydrogen storage and flexible turbine technology / configuration option in support of the study.

2.0 PLANT LOCATION

A specific site location is not defined, rather a generic coastal location in the UK is considered. Where applicable, the site is also assumed to be close to an existing harbour equipped with a suitable pier and coal bay to allow coal transport by large ships and associated ease of coal handling.

3.0 SITE CONDITION

An assumed clear level obstruction (both under and above ground) free site is considered, without the need for any required special civil works.

4.0 PLANT CAPACITY

Each case considered as part of WP1 will be designed to produce electric energy (350 MWe nominal gross capacity with pre-combustion CO_2 capture) to be delivered to the UK National grid. For each of the cases considered, power generation will be intermittent and will vary according to variation in energy demand, through the use of hydrogen storage and flexible turbine systems.

5.0 PLANT OPERATING CONDITIONS

The following climatic conditions marked (*) shall be considered reference conditions for plant performance evaluation across all cases. Individual case deliverables will be produced at reference conditions only.

Atmospheric pressure: 1013 mbar (*)

Relative humidity: average: average 60% (*)

maximum: 95%

minimum: 40%

Ambient temperatures: average 10°C

summer 32°C (*)

minimum -10°C

6.0 CARBON DIOXIDE CAPTURE RATE

Each carbon dioxide abated case will be designed to achieve a target carbon capture level of at least 90%, defined as:

$$CO_2 \text{ Capture Rate (\%)} = 100 \times \frac{\text{Moles carbon contained in the } CO_2 \text{ product}}{\text{Moles carbon contained in the fuel feed}}$$

7.0 FEEDSTOCK, PRODUCT AND UTILITY SUPPLIES

The streams available at plant battery limits are the following:

Coal;

Biomass;

Natural Gas;

CO_2 product;

Sea water supply;

Sea water Return;

Plant/Raw/Potable water;

Chemicals (including amine);

Sulphur product;

Limestone.

Other utilities, including demineralised water, boiler feedwater, instrument and plant air, oxygen, nitrogen will be generated within the complex where necessary and will be available for use at the required conditions.

8.0 FEEDSTOCK SPECIFICATIONS

8.1 Coal

The coal specification is based on an open-cut coal from Eastern Australia:

Proximate analysis	(wt%)
Inherent moisture:	9.50
Ash:	12.20
Coal (dry, ash free):	78.30

Ultimate analysis	(wt%) (dry ash free)
Carbon:	82.50
Hydrogen:	5.60
Nitrogen:	1.77
Oxygen:	9.00
Sulphur:	1.10
Chlorine:	0.03

Gross CV:	27.06 MJ/kg
Net CV:	25.87 MJ/kg
Hardgrove Index:	45
Ash fusion point (reducing temperature)	1350°C

8.2 Biomass

The biomass used is wood pellets of the following specification:

Proximate analysis	(wt%)
Inherent moisture:	7.00
Volatile matter:	79.00
Fixed carbon:	13.80
Ash:	0.20
Gross CV:	18.70 MJ/kg

Ultimate analysis	(wt%)
Moisture:	7.00
Carbon:	43.50
Hydrogen:	4.50
Nitrogen:	0.20
Oxygen:	42.60
Sulphur:	0.01
Chlorine:	0.01
Ash:	0.2

Ash analysis	(wt%)
SiO_2:	13.70
Al_2O_3:	3.30
Fe_2O_3:	4.90
CaO:	34.40
MgO:	6.70
TiO_2:	0.40

Na_2O: 0.30

K_2O: 24.00

P_2O_5: 5.40

SO_3: 6.80

8.3 Natural Gas

Natural gas NTS connection is available.

Natural gas feedstock specification (as NTS spec):

H_2S Content	Not more than 5 mg/m^3
Total Sulphur Content	Not more than 50 mg/m^3
Hydrogen Content	Not more than 0.1% (molar)
Oxygen Content	Not more than 0.001% (molar)
Hydrocarbon Dewpoint	Not more than -2^{o}C, at any pressure up to 85 bar(g)
Water Dewpoint	Not more than -10^{o}C, at 85 bar(g) (or the actual delivery pressure)
Wobbe Number (real gross dry)	Between 48.14 MJ/m^3 and 51.41 MJ/m^3 (at standard temperature and pressure) and in compliance with ICF and SI limits as listed below
Incomplete Combustion Factor	Not more than 0.48
Soot Index	Not more than 0.60
Gross Calorific Value (real gross dry)	Between 36.9 MJ/m^3 and 42.3 MJ/m^3 (at standard temperature and pressure) and in compliance with ICF and SI limits described above, subject to a 1 MJ/m^3 variation.
Inerts	Not more than 7.0mol%, subject to: Carbon Dioxide content – not more than 2.0mol% Nitrogen content – not more than 5.0mol%
Contaminants	Gas shall not contain solid or liquid material which may interfere with the integrity or operation of pipes or any gas appliance within the meaning of the Regulation 2(1) of the Gas Safety (Use of) Regulations 1998 that a consumer could reasonably be expected to operate.
Delivery Temperature	Between 1^{o}C and 38^{o}C
Odour	Gas delivered shall have no odour that might contravene the statutory obligation “not to transmit or distribute any gas at a pressure below 7 bar(g) which does not possess a distinctive and characteristic odour”.

8.4 Back up fuel/power

Natural gas (as detailed in section 8.3) is available for back-up fuel.

National Grid electrical grid connection is available for “black start” power requirement scenarios.

9.0 PRODUCT SPECIFICATIONS

9.1 Carbon Dioxide

Carbon dioxide produced from the plant will be dried and compressed to 150 bar(g) for export from the facility. Product carbon dioxide conditions will be:

Pressure:	150 bar(g)
Temperature:	≤ 30°C

The target carbon dioxide export specification is based on the requirements for EOR.

H_2O	< 50 ppmv
CO_2	> 97 vol%
SO_2	< 50 ppm
H_2S	< 50 ppm
CO	< 3 vol%
Ar	< 3 vol%
O_2	100 ppmv
N_2	< 3 vol%
H_2	< 3 vol%
CH_4	< 2 vol%
COS	< 50ppm

9.2 Power

Power will be generated from the complex at 275 kV and will be transmitted to an assumed existing HV substation for connection onto the UK National Grid. It is assumed that National Grid electrical grid connection is available.

Electric Power

Net Power Output	350 MWe nominal capacity
Voltage	275 kV
Frequency	50 Hz

9.3 Solid by-products

The power plant cases will produce saleable solid by-products, in particular:

IGCC Cases: slag, flyash

10.0 UTILITY SUPPLIES

10.1 Seawater cooling system

The primary cooling system is sea water in a once through system. Services will include the steam turbine condenser and the seawater/closed loop interchanger. Seawater supply assumed to be clear filtered and chlorinated, without suspended solids and organic matter. Seawater supply from a new intake and a seawater outfall will be required as part of the complex.

n.b. Costs of seawater pipelines, intake and outfall are not included within comparative capital cost estimates for WP1.

The following seawater conditions marked (*) shall be considered reference conditions for plant performance evaluation across all WP1 cases. Individual case deliverables will be produced at reference conditions only.

Seawater conditions:

Average supply temperature:	17°C (*)
Average return temperature:	25°C (*)
Operating pressure at Condenser inlet:	3 bar(g)
Maximum allowable ΔP for Condenser:	0.7 bar

10.2 Closed loop water cooling system

The secondary cooling system is a closed loop, seawater cooled cooling water system. All cooling services, with the exception of the steam turbine vacuum condenser, will be placed on this system. This system cools the closed loop water against seawater. The make-up water to the system shall be demineralised water stabilized and conditioned.

The following closed loop water conditions marked (*) shall be considered reference conditions for plant performance evaluation across all WP1 cases. Individual case deliverables will be produced at reference conditions only.

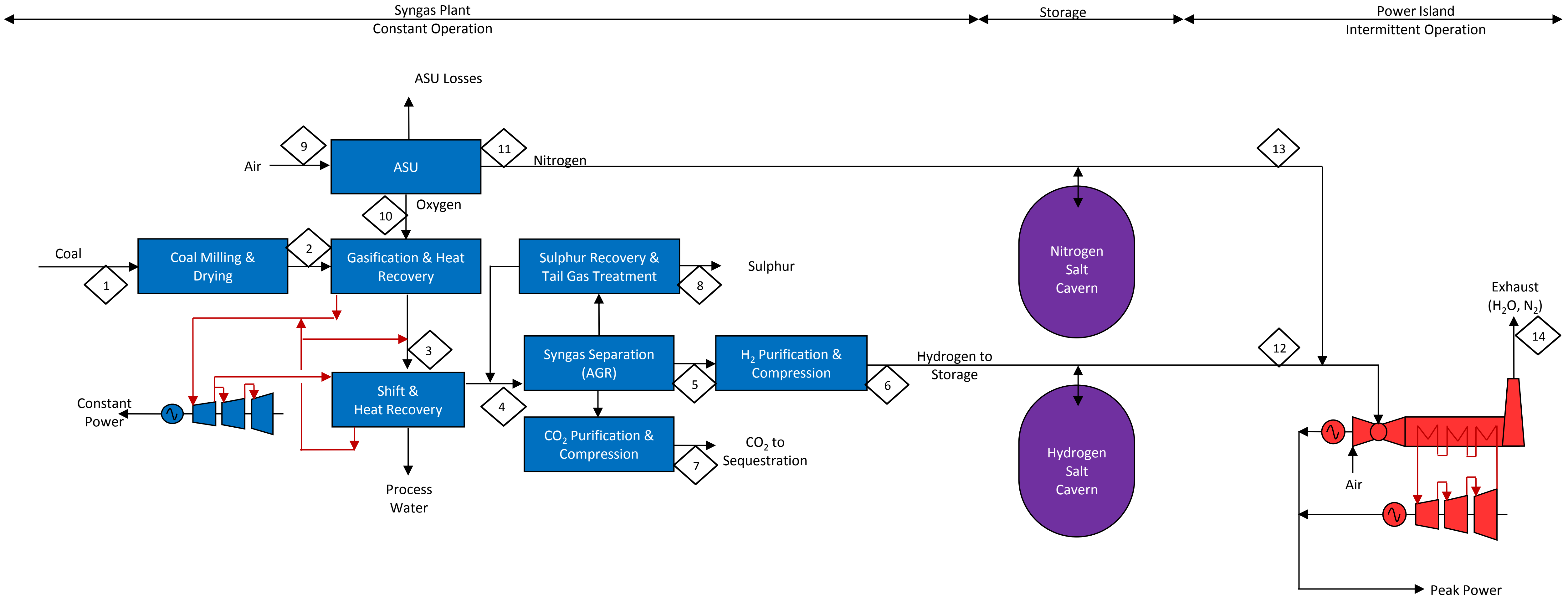
Closed loop cooling water conditions:

Average supply temperature:	21°C (*)

Average return temperature:	31°C (*)
Seawater/closed loop water interchanger ΔT:	4°C (*)
Operating pressure at users:	3.0 bar(g)
Maximum allowable ΔP for users:	1.5 bar

ATTACHMENT 2

OUTLINE HEAT & MATERIAL BALANCE – CASE 1



Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Temperature [C]	20.0	74.3	270.1	34.0	34.0	30.0	29.9	24.0	32.0	25.0	30.0	29.9	29.3	106.5
Pressure [kPa]	101.3	5000.0	3850.0	3420.0	3340.0	3340.0	15100.0	101.3	101.3	4950.0	3340.0	3030.0	3030.0	104.0
Mass Flow [kg/h]	136521	130059	491720	364739	59818	59818	292236	1170	424745	102746	272562	59795	272562	2634809
Molar Flow [kgmole/h]	11712	11356	25983	18772	10870	10870	6755	36	14719	3213	9730	10869	9730	95932
Mole% Hydrogen	25.33	26.15	14.39	56.90	89.31	89.31	1.74	0.00	0.00	0.00	0.00	89.32	0.00	0.00
Mole% CO	0.00	0.00	27.44	1.00	1.57	1.57	0.03	0.00	0.00	0.00	0.00	1.57	0.00	0.00
Mole% CO2	0.00	0.00	0.78	38.68	4.23	4.23	98.14	0.00	0.03	0.00	0.00	4.23	0.00	0.68
Mole% Nitrogen	0.58	0.59	1.99	2.75	4.34	4.34	0.05	0.00	77.31	3.50	100.00	4.34	100.00	74.29
Mole% Oxygen	2.57	2.65	0.00	0.00	0.00	0.00	0.00	0.00	20.74	95.00	0.00	0.00	0.00	11.92
Mole% Argon	0.00	0.00	0.19	0.26	0.39	0.39	0.04	0.00	0.93	1.50	0.00	0.39	0.00	0.81
Mole% H2S	0.00	0.00	0.13	0.19	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% COS	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Carbon	62.70	64.66	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Sulphur	0.31	0.32	0.00	0.00	0.00	0.00	0.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Ash	2.37	2.44	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% H2O	6.15	3.18	55.06	0.21	0.16	0.16	0.00	0.00	0.99	0.00	0.00	0.15	0.00	12.29

01	08/03/13	FIRST ISSUE	SEF	GS	TA
REV	DATE	TITLE	BY	CHK	APP

REVISIONS

HYDROGEN STORAGE AND FLEXIBLE TURBINES PROJECT

BLOCK FLOW DIAGRAM

CASE 1: COAL FED IGCC

ATTACHMENT 3

OUTLINE UTILITY SUMMARY – CASE 1

FOSTER WHEELER ENERGY LIMITED
UTILITIES BALANCE SUMMARY

Case 1 - Coal with CO_2 Capture (90%)

CLIENT: The Energy Technologies Institute
CONTRACT: 13058
NAME: Hydrogen Storage and Flexible Turbine System

						SHEET
REV	O1					
DATE	04/03/2013					1 OF 1
ORIG. BY	AK					
APP. BY	SEF					

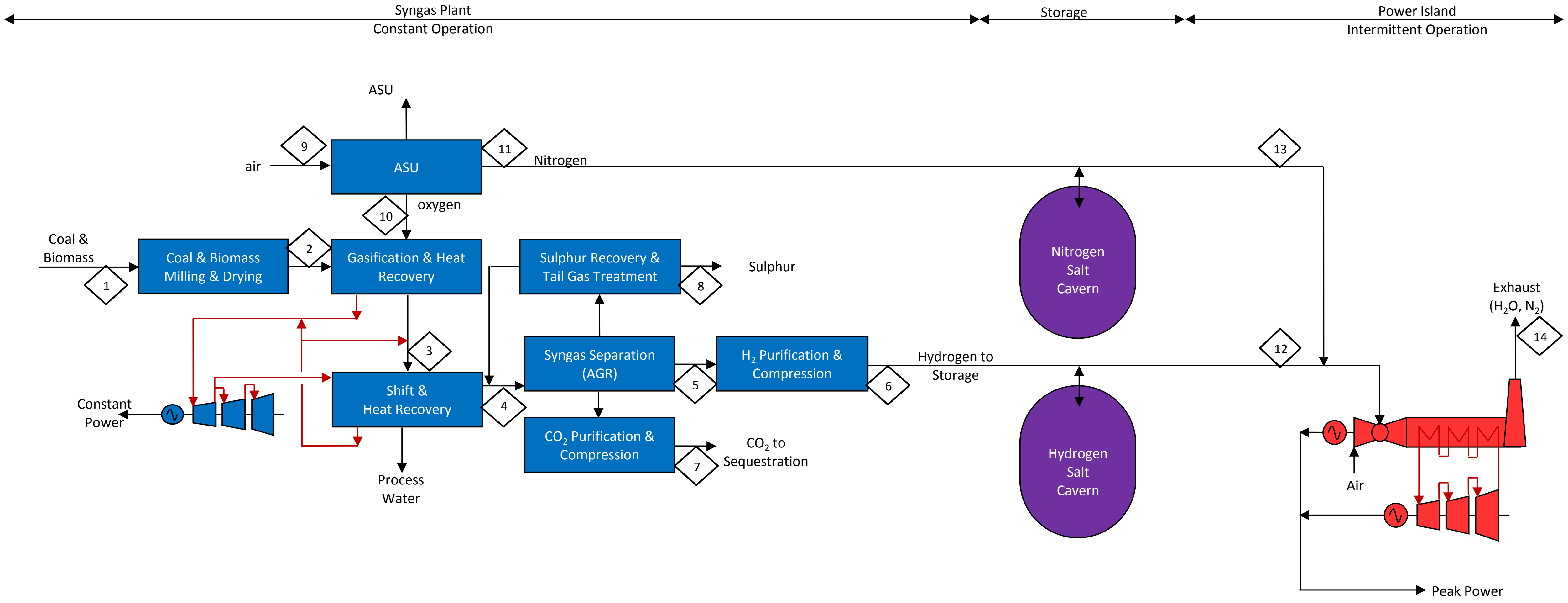
DESCRIPTION	Electric Power (Note 1)			Steam (t/h)			Condensate (Note 2)	Sea Water	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
	Steady State Averaged Power kW	Actual Continuous Power kW	Actual Peak Power kW	124 bar	45 bar	9 bar	t/h	t/h	t/h	t/h	t/h		
Process Units													
Coal Milling and Drying	-6849	-3424	-3424										
Gasification, Quenching, Syngas Cooling	-239	-120	-120										
Heat Recovery and Steam Turbine	59047	29524	29524			44.2		-7157.4	-35.2	-182.7	9.1		
Syngas Cooling and Shift Reactors	-1353	-676	-676						-1676.5		39.1		
Acid Gas Recovery Unit	-7808	-3904	-3904			-30.6	30.6		-4060.7				
Sour Water Stripper	-500	-250	-250			-5.5	5.5		0.0				
Air Separation Unit	-34295	-17148	-17148			-8.9	8.9		-3461.9				
Sulphur Recovery/Tailgas Treatment Unit	-1729	-864	-864			0.8	-0.8		-104.3				
CO2 Compression and Drying	-25120	-12560	-12560						-3629.9		0.2		
N2 & H2 Storage	-27499	-13749	-13749						-2268.1				
Fresh Cooling Water	-954	-477	-477					-19247.4	15236.5				
Sea Cooling Water	-2416	-1208	-1208					26404.8					
Process Units Total	-49713	-24856	-24856	0.0	0.0	0.0	44.2	0.0	0.0	-182.7	48.4		
Power Island													
Gas Turbine Generators	268153		268153										
Heat Recovery Steam Generators	-2297		-2297	261.2	44.4					-26.1	1.9		
Steam Turbine and Condenser Package	127460		127460	-261.2	-44.4		329.5	21890.0					
Sea Cooling Water	-2123		-2123					-21890.0					
Power Island Total	391193	0	391193	0.0	0.0	0.0	329.5	0.0	0.0	-26.1	1.9		
Offsites & Utilities													
Demin Plant	-255	-128	-128							208.8			
Utility water	-71	-35	-35										
Fire Water System	-20	-10	-10										
Condensate Treatment	-255	-128	-128				-336.3				-50.3		
Waste Water Treatment	-262	-131	-131				-37.4						
Storage	-100	-50	-50										
Buildings	-1300	-650	-650										
Offsites & Utilities Total	-2264	-1132	-1132	0.0	0.0	0.0	-373.7	0.0	0.0	208.8	-50.3		
Grand Total	**339216**	**-25988**	**365204**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

NOTES

1. Continuous and peak power figures are based on the syngas production plant running 24 hours a day and the power island running 12 hours a day. All other figures represent the steady state average flow at the capacity of 1 GT.
2. An overall loss of 10% is considered for condensate, vents and blow-downs.

ATTACHMENT 4

OUTLINE HEAT & MATERIAL BALANCE – CASE 2



Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Temperature [C]	20.0	73.9	270.1	34.0	34.0	30.0	29.9	24.0	32.0	25.0	30.0	29.9	29.3	106.6
Pressure [kPa]	101.3	5000.0	3850.0	3420.0	3340.0	3340.0	15100.0	101.3	101.3	4950.0	3340.0	3030.0	3030.0	104.0
Mass Flow [kg/h]	152737	150584	502181	370654	60700	60700	297244	973	401798	97195	271680	60676	271680	2634808
Molar Flow [kgmole/h]	12731	12393	26379	18914	10899	10899	6869	30	13924	3039	9698	10897	9698	95929
Mole% Hydrogen	26.42	26.80	14.34	56.50	89.12	89.12	1.71	0.00	0.00	0.00	0.00	89.13	0.00	0.00
Mole% CO	0.00	0.00	26.87	0.97	1.53	1.53	0.03	0.00	0.00	0.00	0.00	1.53	0.00	0.00
Mole% CO2	0.00	0.00	1.42	39.09	4.36	4.36	98.17	0.00	0.03	0.00	0.00	4.36	0.00	0.69
Mole% Nitrogen	0.45	0.47	2.02	2.81	4.46	4.46	0.05	0.00	77.31	3.50	100.00	4.46	100.00	74.28
Mole% Oxygen	6.59	6.23	0.00	0.00	0.00	0.00	0.00	0.00	20.74	95.00	0.00	0.00	0.00	11.92
Mole% Argon	0.00	0.00	0.17	0.24	0.37	0.37	0.03	0.00	0.93	1.50	0.00	0.37	0.00	0.81
Mole% H2S	0.00	0.00	0.10	0.16	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% COS	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Carbon	58.69	60.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Sulphur	0.23	0.25	0.00	0.00	0.00	0.00	0.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Ash	1.75	2.59	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% H2O	5.87	3.37	55.06	0.21	0.16	0.16	0.00	0.00	0.99	0.00	0.00	0.15	0.00	12.30

01	08/03/13	FIRST ISSUE	SEF	GS	TA
REV	DATE	TITLE	BY	CHK	APP

REVISIONS

HYDROGEN STORAGE AND FLEXIBLE TURBINES PROJECT

BLOCK FLOW DIAGRAM

CASE 2: COAL & BIOMASS FED IGCC

ATTACHMENT 5

OUTLINE UTILITY SUMMARY – CASE 2

FOSTER WHEELER ENERGY LIMITED
UTILITIES BALANCE SUMMARY

Case 2 - Coal+Biomass with CO_2 Capture (90%)

CLIENT: The Energy Technologies Institute
CONTRACT: 13058
NAME: Hydrogen Storage and Flexible Turbine System

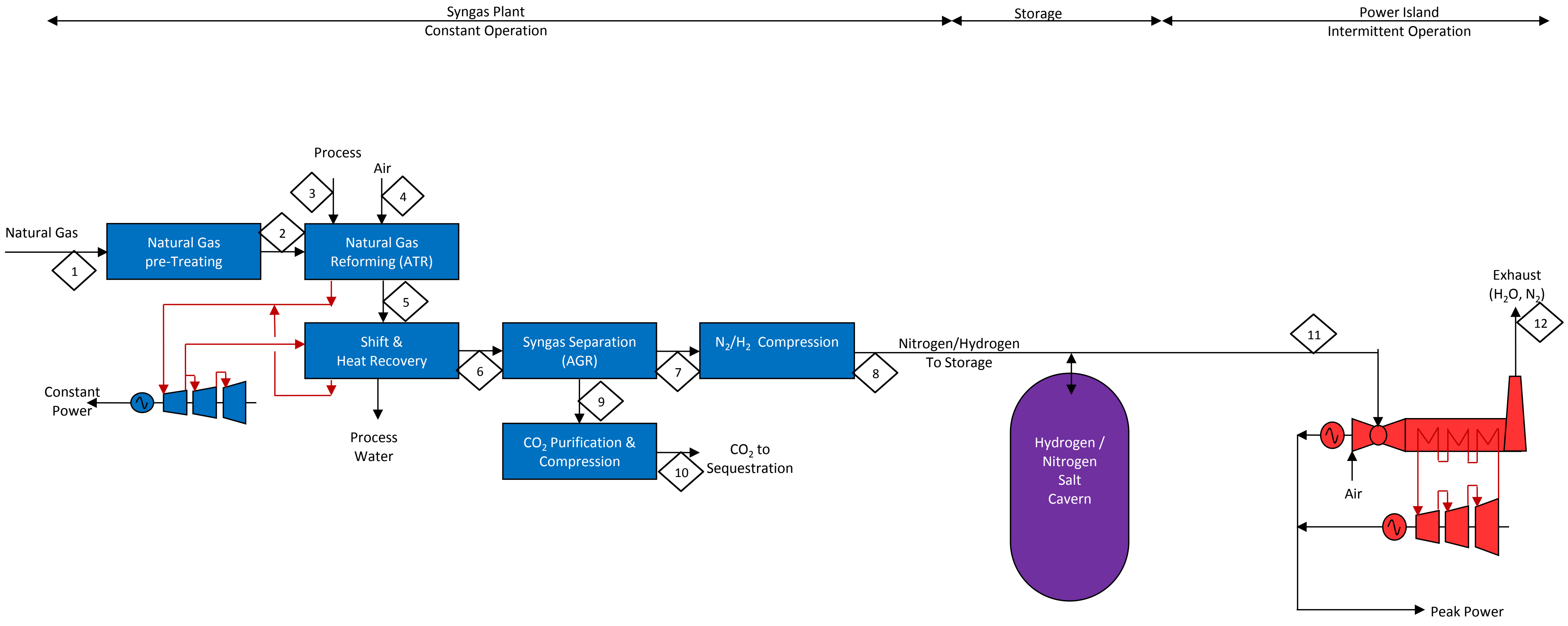
REV	O1					SHEET
DATE	05/03/2013					1 OF 1
ORIG. BY	AK					
APP. BY	SEF					

	DESCRIPTION	Electric Power (Note 1)			Steam (t/h)			Condensate (Note 2)	Sea Water	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
		Annual Averaged Power kW	Actual Continuous Power kW	Actual Peak Power kW	124 bar	45 bar	9 bar	t/h	t/h	t/h	t/h	t/h		
	Process Units													
	Coal Milling and Drying	-7662	-3831	-3831										
	Gasification, Quenching, Syngas Cooling	-283	-142	-142										
	Heat Recovery and Steam Turbine	60334	30167	30167			44.2		-7331.9	-35.8	-186.8	9.2		
	Syngas Cooling and Shift Reactors	-1354	-677	-677						-2416.0		50.3		
	Acid Gas Recovery Unit	-7808	-3904	-3904			-30.6	30.6		-4060.7				
	Sour Water Stripper	-500	-250	-250			-5.5	5.5		0.0				
	Air Separation Unit	-32462	-16231	-16231			-8.9	8.9		-3303.8				
	Sulphur Recovery/Tailgas Treatment Unit	-1882	-941	-941			0.8	-0.8		-116.7				
	CO2 Compression and Drying	-25543	-12771	-12771						-3692.1		0.2		
	N2 & H2 Storage	-27409	-13705	-13705						-2261.3				
	Fresh Cooling Water	-995	-497	-497					-20066.8	15886.5				
	Sea Cooling Water	-2506	-1253	-1253					27398.7					
	Process Units Total	-48069	-24035	-24035	0.0	0.0	0.0	44.2	0.0	0.0	-186.8	59.8		
	Power Island													
	Gas Turbine Generators	268153		268153										
	Heat Recovery Steam Generators	-2296		-2296	261.0	44.4					-26.1	1.9		
	Steam Turbine and Condenser Package	127417		127417	-261.0	-44.4		329.4	21883.4					
	Sea Cooling Water	-2122		-2122					-21883.4					
	Power Island Total	391152	0	391152	0.0	0.0	0.0	329.4	0.0	0.0	-26.1	1.9		
	Offsites & Utilities													
	Demin Plant	-261	-131	-131							212.9			
	Utility water	-72	-36	-36										
	Fire Water System	-20	-10	-10										
	Condensate Treatment	-261	-131	-131				-336.2				-61.7		
	Waste Water Treatment	-271	-136	-136				-37.4						
	Storage	-100	-50	-50										
	Buildings	-1300	-650	-650										
	Offsites & Utilities Total	-2286	-1143	-1143	0.0	0.0	0.0	-373.6	0.0	0.0	212.9	-61.7		
	Grand Total	**340797**	**-25178**	**365974**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

NOTES
1. Continuous and peak power figures are based on the syngas production plant running 24 hours a day and the power island running 12 hours a day. All other figures represent the steady state average flow at the capacity of 1 GT.
2. An overall loss of 10% is considered for condensate, vents and blow-downs.

ATTACHMENT 6

OUTLINE HEAT & MATERIAL BALANCE – CASE 3



Stream	1	2	3	4	5	6	7	8	9	10	11	12
Temperature [C]	1.0	339.8	278.2	32.0	350.0	60.0	30.0	70.4	40.0	29.8	29.7	109.6
Pressure [kPa]	3500.0	3090.0	2970.0	101.3	2760.0	2460.0	2400.0	3340.0	110.0	15100.0	3030.0	104.2
Mass Flow [kg/h]	65079	65079	107106	322697	494491	444219	262686	262686	163173	158435	262530	2510530
Molar Flow [kgmole/h]	3550	3550	5941	11183	26105	23318	18133	18133	3898	3636	18124	91089
Mole% Hydrogen	0.00	0.00	0.00	0.00	29.70	44.18	52.68	52.68	0.81	0.87	52.70	0.00
Mole% CO	0.00	0.00	0.00	0.00	10.21	0.50	0.59	0.59	0.01	0.01	0.59	0.00
Mole% CO2	0.68	0.68	0.03	0.03	4.79	16.29	1.07	1.07	92.05	98.69	1.07	0.66
Mole% Nitrogen	1.47	1.47	0.00	77.31	33.32	37.30	44.53	44.53	0.38	0.41	44.56	74.72
Mole% Oxygen	0.00	0.00	0.00	20.74	0.00	0.00	0.00	0.00	0.00	0.00	0.00	12.21
Mole% Argon	0.00	0.00	0.00	0.93	0.40	0.45	0.53	0.53	0.00	0.00	0.53	0.90
Mole% Methane	87.08	87.08	0.00	0.00	0.29	0.32	0.38	0.38	0.01	0.01	0.38	0.00
Mole% Ethane	7.83	7.83	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Propane	2.94	2.94	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% C3+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% H2O	0.00	0.00	99.93	0.99	21.29	0.97	0.21	0.21	6.71	0.00	0.17	11.51
Mole% Methanol	0.00	0.00	0.02	0.00	0.01	0.00	0.00	0.00	0.02	0.00	0.00	0.00

01	08/03/13	FIRST ISSUE	SEF	GS	TA
REV	DATE	TITLE	BY	CHK	APP

REVISIONS

HYDROGEN STORAGE AND FLEXIBLE TURBINES PROJECT

BLOCK FLOW DIAGRAM

CASE 3: ATR

ATTACHMENT 7

OUTLINE UTILITY SUMMARY – CASE 3

FOSTER WHEELER ENERGY LIMITED
UTILITIES BALANCE SUMMARY

Case 3 - ATR with CO_2 Capture (90%)

CLIENT: The Energy Technologies Institute
CONTRACT: 13058
NAME: Hydrogen Storage and Flexible Turbine System

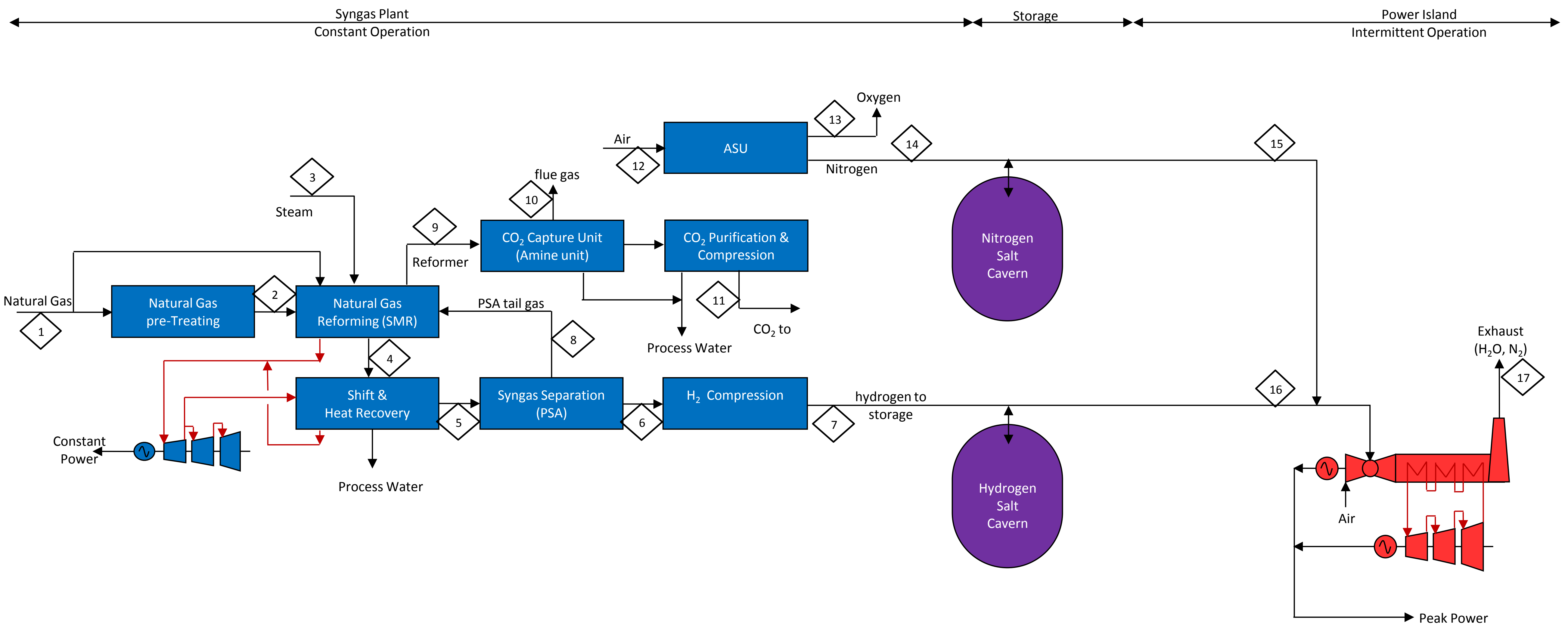
						SHEET
REV	O1					1 OF 1
DATE	04/03/2013					
ORIG. BY	AK					
APP. BY	SEF					

DESCRIPTION	Electric Power (Note 1)			Steam (t/h)			Condensate (Note 2)	Sea Water	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
	Annual Averaged Power kW	Actual Continuous Power kW	Actual Peak Power kW	125 bar	43 bar	7 bar	t/h	t/h	t/h	t/h	t/h		
Process Units													
Natural Gas Pre-Treating & Reforming	-45475	-22738	-22738						-5853.8	-108.9	51.5		
Heat Recovery and Steam Turbine	42598	21299	21299			53.1		-4733.7			0.9		
Acid Gas Recovery Unit	-4665	-2333	-2333			-52.8	52.8		-3279.9		2.0		
CO2 Compression and Drying	-17543	-8771	-8771			-0.4	0.4		-2752.7				
N2/H2 Compression	-6208	-3104	-3104						-528.2				
Fresh Cooling Water	-775	-387	-387					-15633.5	12414.6				
Sea Cooling Water	-1863	-932	-932					20367.2					
Process Units Total	-33932	-16966	-16966	0	0	0	53	0	0	-109	54		
Power Island													
Gas Turbine Generators	268053	0	268053										
Heat Recovery Steam Generators	-2136	0	-2136	-246.4	-57.7	-25.2				-2.9	2.0		
Steam Turbine and Condenser Package	125868	0	125868	246.4	57.7	25.2	329.3	-21876.0					
Sea Cooling Water	-2135	0	-2135					21876.0					
Power Island Total	389650	0	389650	0	0	0	329	0	0	-3	2		
Offsites & Utilities													
Demin Plant	-287	-144	-144							111.8			
Utility water	-80	-40	-40										
Fire Water System	-20	-10	-10										
Condensate Treatment	-287	-144	-144				-344.2				-56.4		
Waste Water Treatment	-264	-132	-132				-38.2						
Storage	-100	-50	-50										
Buildings	-1300	-650	-650										
Offsites & Utilities Total	-2338	-1169	-1169	0.0	0.0	0.0	-382.4	0.0	0.0	111.8	-56.4		
Grand Total	**353380**	**-18135**	**371515**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

NOTES
1. Continuous and peak power figures are based on the syngas production plant running 24 hours a day and the power island running 12 hours a day. All other figures represent the steady state average flow at the capacity of 1 GT.
2. An overall loss of 10% is considered for condensate, vents and blow-downs.

ATTACHMENT 8

OUTLINE HEAT & MATERIAL BALANCE – CASE 4



Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
Temperature [C]	1.0	339.9	321.8	860.0	40.0	40.0	81.0	40.0	143.6	102.0	29.8	32.0	20.0	53.6	29.5	24.0	111.4
Pressure [kPa]	3500	3170	3170	2770	2440	2420	3340	130	102.00	110	15100	101	200	3340	3030	3030	104.2
Mass Flow [kg/h]	76628	62329	185262	247594	168406	19981	19981	148425	930040	732698	186764	416236	103837	312399	312399	19981	2625295
Molar Flow [kgmole/h]	4180	3400	10277	19571	15182	9911	9911	5271	31990	27159	4244	14424	3272	11152	11152	9911	95851
Mole% Hydrogen	0.00	0.00	0.01	48.14	74.18	100.00	100.00	25.64	0.00	0.00	0.00	0.00	0.00	0.00	0.00	100.00	0.00
Mole% CO	0.00	0.00	0.00	9.72	0.39	0.00	0.00	1.12	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% CO2	0.68	0.68	0.04	5.48	19.18	0.00	0.00	55.23	14.69	1.68	99.95	0.03	0.13	0.00	0.00	0.00	0.02
Mole% Nitrogen	1.47	1.47	0.00	0.26	0.33	0.00	0.00	0.95	64.46	75.92	0.02	77.31	0.00	100.00	100.00	0.00	75.35
Mole% Oxygen	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.47	5.26	0.01	20.74	91.40	0.00	0.00	0.00	11.92
Mole% Argon	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.77	0.91	0.01	0.93	4.10	0.00	0.00	0.00	0.77
Mole% Methane	87.08	87.08	0.00	4.32	5.57	0.00	0.00	16.05	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Ethane	7.83	7.83	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Propane	2.94	2.94	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% C3+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% H2O	0.00	0.00	99.94	32.08	0.35	0.00	0.00	1.01	15.61	16.23	0.00	0.99	4.36	0.00	0.00	0.00	11.94
Mole% Methanol	0.00	0.00	0.02	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00

01	08/03/13	FIRST ISSUE	AK	GS	TA
REV	DATE	TITLE	BY	CHK	APP

REVISIONS

HYDROGEN STORAGE AND FLEXIBLE TURBINES PROJECT

BLOCK FLOW DIAGRAM

CASE 4: SMR

ATTACHMENT 9

OUTLINE UTILITY SUMMARY – CASE 4

FOSTER WHEELER ENERGY LIMITED
UTILITIES BALANCE SUMMARY

Case 4 - SMR with CO_2 Capture (90%)

CLIENT: The Energy Technologies Institute
CONTRACT: 13058
NAME: Hydrogen Storage and Flexible Turbine System

						SHEET
REV	O1					
DATE	05/03/2013					1 OF 1
ORIG. BY	AK					
APP. BY	SEF					

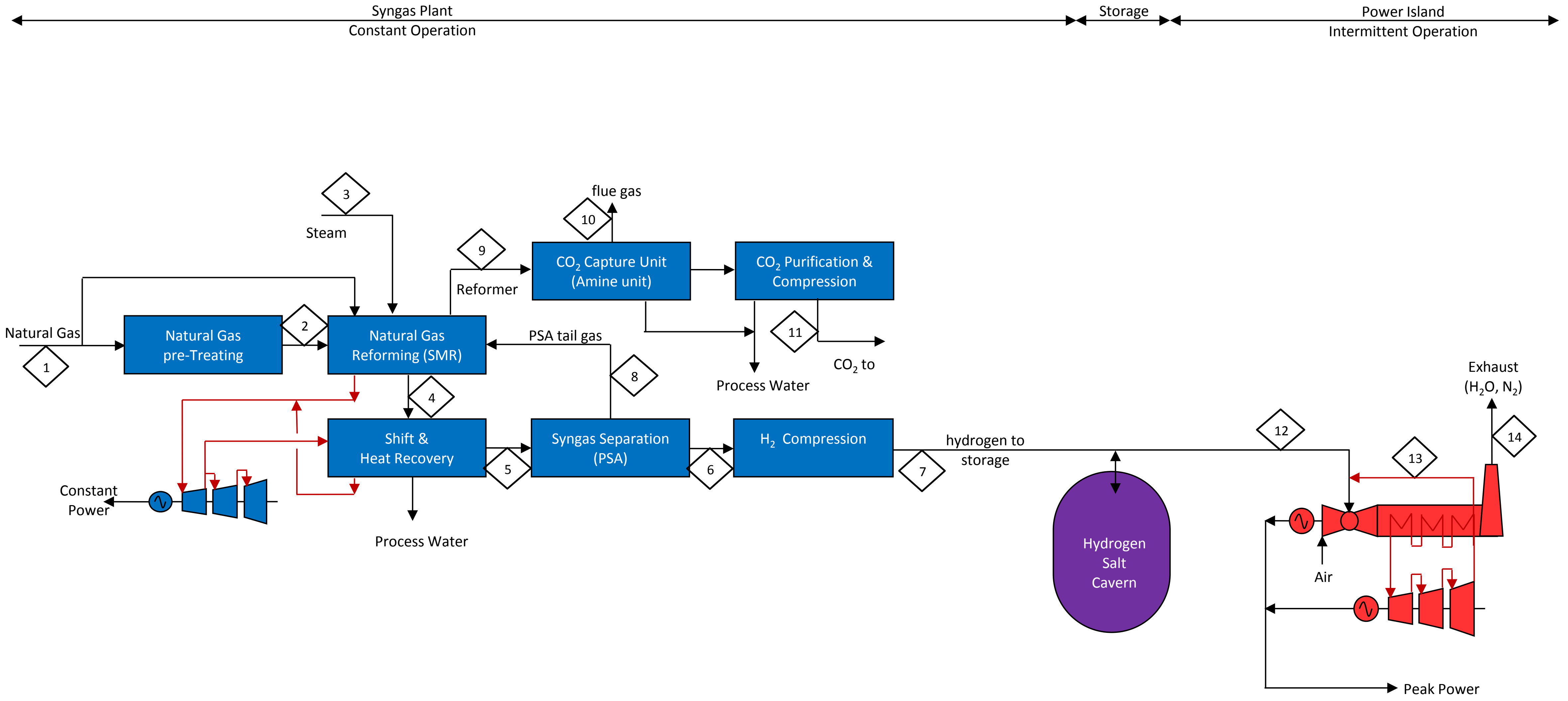
	DESCRIPTION	Electric Power (Note 1)			Steam (t/h)			Condensate (Note 2)	Sea Water	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
		Annual Averaged Power kW	Actual Continuous Power kW	Actual Peak Power kW	148 bar	35 bar	4 bar	t/h	t/h	t/h	t/h	t/h		
	Process Units													
	Natural Gas Pre-Treating & Reforming	-1522	-761	-761						-2454.8	-182.5			
	Heat Recovery Steam Generators	27599	13800	13800			265.2			-47.9		71.6		
	Air Separation Unit	-19148	-9574	-9574			-1.9	1.9		-1746.6				
	Acid Gas Removal Unit	-3078	-1539	-1539			-263.3	263.3		-14708.1	-42.7	59.3		
	CO2 Compression and Drying	-20399	-10199	-10199						-3274.5		10.6		
	N2/H2 Purification & Compression	-34638	-17319	-17319						-2536.3				
	Fresh Cooling Water	-1546	-773	-773					-31198.1	24768.2				
	Sea Cooling Water	-2854	-1427	-1427					31198.1					
	Process Units Total	-55585	-27793	-27793	0.0	0.0	0.0	265.2	0.0	0.0	-225.2	141.5		
	Power Island													
	Gas Turbine Generators	268153		268153										
	Heat Recovery Steam Generators	-2357		-2357	-255.7	-69.1					-16.7			
	Steam Turbine and Condenser Package	129113		129113	255.7	69.1		324.8	-21735.0					
	Sea Cooling Water	-2121		-2121					21735.0					
	Power Island Total	392788	0	392788	0.0	0.0	0.0	324.8	0.0	0.0	-16.7	0.0		
	Offsites & Utilities													
	Demin Plant	-287	-144	-144							241.9			
	Utility water	-80	-40	-40										
	Fire Water System	-20	-10	-10										
	Condensate Treatment	-287	-144	-144				-531.0				-71.6		
	Waste Water Treatment	-264	-132	-132				-59.0				-69.9		
	Storage	-100	-50	-50										
	Buildings	-1300	-650	-650										
	Offsites & Utilities Total	-2338	-1169	-1169	0.0	0.0	0.0	-590.0	0.0	0.0	241.9	-141.5		
	Grand Total	**334865**	**-28961**	**363826**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

NOTES

1. Continuous and peak power figures are based on the syngas production plant running 24 hours a day and the power island running 12 hours a day. All other figures represent the steady state average flow at the capacity of 1 GT.
2. An overall loss of 10% is considered for condensate, vents and blow-downs.

ATTACHMENT 10

OUTLINE HEAT & MATERIAL BALANCE – CASE 4B



Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Temperature [C]	1.0	339.9	321.8	860.0	40.0	40.0	81.0	40.0	136.6	102.0	29.8	25.1	173.6	120.0
Pressure [kPa]	3500	3170	3170	2770	2440	2420	3340	130	102.00	110	15100	3030	15630	104
Mass Flow [kg/h]	73857	59246	176094	235342	160078	18993	18993	141085	894659	704344	180007	18993	119000	2528142
Molar Flow [kgmole/h]	4029	3232	9768	18602	14431	9421	9421	5010	30800	26137	4091	9421	6606	95316
Mole% Hydrogen	0.00	0.00	0.01	48.14	74.18	100.00	100.00	25.64	0.00	0.00	0.00	100.00	0.00	0.00
Mole% CO	0.00	0.00	0.00	9.72	0.39	0.00	0.00	1.11	0.00	0.00	0.00	0.00	0.00	0.00
Mole% CO2	0.68	0.68	0.04	5.48	19.18	0.00	0.00	55.23	14.71	1.68	99.95	0.00	0.00	0.66
Mole% Nitrogen	1.47	1.47	0.00	0.26	0.33	0.00	0.00	0.95	64.47	75.97	0.02	0.00	0.00	64.55
Mole% Oxygen	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.25	5.00	0.01	0.00	0.00	12.20
Mole% Argon	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.77	0.91	0.01	0.00	0.00	0.81
Mole% Methane	87.08	87.08	0.00	4.32	5.57	0.00	0.00	16.05	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Ethane	7.83	7.83	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% Propane	2.94	2.94	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% C3+	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Mole% H2O	0.00	0.00	99.94	32.08	0.35	0.00	0.00	1.01	15.80	16.43	0.00	0.00	1.00	21.79
Mole% Methanol	0.00	0.00	0.02	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00

2	########	FINAL	RR	TA	TA
01	08/03/13	FIRST ISSUE	SEF	TA	TA
REV	DATE	TITLE	BY	CHK	APP
REVISIONS					
HYDROGEN STORAGE AND FLEXIBLE TURBINES PROJECT			BLOCK FLOW DIAGRAM		
energy technologies institute			CASE 4b: SMR with Steam to GT		

ATTACHMENT 11

OUTLINE UTILITY SUMMARY – CASE 4B

FOSTER WHEELER ENERGY LIMITED
UTILITIES BALANCE SUMMARY

Case 4b - SMR with Steam to GT with CO_2 Capture (90%)

CLIENT: The Energy Technologies Institute
CONTRACT: 13058
NAME: Hydrogen Storage and Flexible Turbine System

REV	O1	O2				SHEET
DATE	08/03/2013	11/07/2013				1 OF 1
ORIG. BY	SEF	RR				
APP. BY	TA	TA				

DESCRIPTION	Electric Power (Note 1)			Steam (t/h)			Condensate (Note 2)	Sea Water	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
	Annual Averaged Power kW	Actual Continuous Power kW	Actual Peak Power kW	148 bar	35 bar	4 bar	t/h	t/h	t/h	t/h	t/h		
Hydrogen Production Units													
Natural Gas Pre-Treating & Reforming	-1445	-723	-723						-2392.2	-173.7			
Heat Recovery Steam Generators	28303	14151	14151			255.6			-973.8		68.0		
Air Separation Unit	0	0	0			0.0	0.0		0.0				
Acid Gas Removal Unit	-2966	-1483	-1483			-255.6	255.6		-14176.0	-42.7	59.3		
CO2 Compression and Drying	-19661	-9830	-9830						-3156.0		10.3		
N2/H2 Purification & Compression	-2905	-1453	-1453						0.0				
Fresh Cooling Water	-1313	-656	-656					-28998.1	20698.0				
Sea Cooling Water	-2423	-1212	-1212					28998.1					
Process Units Total	-2410	-1205	-1205	0.0	0.0	0.0	255.6	0.0	0.0	-216.4	137.7		
Storage Cavern													
Hydrogen Compression to Cavern	0	0	0										
Nitrogen Compression to Cavern	0	0	0										
Storage Cavern Total (Note 3)	0	0	0										
Power Island													
Gas Turbine Generators (Note 4)	251138		251138										
Heat Recovery Steam Generators	-2625		-2625	-266.2						-201.3			
Steam Turbine and Condenser Package	106669		106669	266.2			185.1	-12382.6					
Sea Cooling Water	-1609		-1609					12382.6					
Power Island Total	353572	0	353572	0.0	0.0	0.0	185.1	0.0	0.0	-201.3	0.0		
Offsites & Utilities													
Demin Plant	-287	-144	-144							417.7			
Utility water	-80	-40	-40										
Fire Water System	-20	-10	-10										
Condensate Treatment	-287	-144	-144				-396.7				-68.0		
Waste Water Treatment	-264	-132	-132				-44.1				-69.6		
Storage	-100	-50	-50										
Buildings	-1300	-650	-650										
Offsites & Utilities Total	-2338	-1169	-1169	0.0	0.0	0.0	-440.7	0.0	0.0	417.7	-137.7		
Grand Total	**348825**	**-2374**	**351198**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

NOTES
1. Continuous and peak power figures are based on the syngas production plant running 24 hours a day and the power island running 12 hours a day. All other figures represent the steady state average flow at the capacity of 1 GT.
2. An overall loss of 10% is considered for condensate, vents and blow-downs.
3. Cavern pressure is currently unknown, therefore pressure at outlet of hydrogen production units is designed to equal that required for the GT inlet.
4. Gas turbine performance has not been down-rated for this case due to lack of firm vendor data. It is likely that a 1% to 2% loss of efficiency will be experienced when substituting nitrogen dilution for steam dilution.

ATTACHMENT 12

CAPITAL COST SUMMARY – CASE 1

Project: No 13058
Client : ETI
Project: H2 STORAGE STUDY
Location : UK

Rev : 0
Date :
By : KSW
Printed: 08 March 2013

ORDER OF MAGNITUDE ESTIMATE SUMMARY
Carbon Capture Units

COST CODE	DESCRIPTION	Case 1: IGCC Coal power plant with pre-combustion capture 1 TRAIN (50% capacity)								
		Reforming / Gasification (coal handling, milling, storage, gasification etc.) Million's GBP	Air Separation / Oxidant Supply Million's GBP	Acid Gas Removal = Separate H2S & CO2 Removal Million's GBP	Sulphur Plant Million's GBP	Syngas Treatment Unit (including CO Shift & Cooling) Million's GBP	CO2 Compression (to 150 Bar) Million's GBP	Power Block Million's GBP	U&O Million's GBP	Total Million's GBP
	MAJOR EQUIPMENT	52.9	34.1	29.1	8.0	12.7	16.5	27.4	13.0	193.7
	DIRECT BULK MATERIALS	38.3	2.0	10.1	2.8	6.9	4.2	8.3	16.8	89.3
	DIRECT MATERIAL & LABOUR CONTRACTS	3.3	1.0	3.4	0.9	1.5	0.7	3.1	26.7	40.5
	LABOUR ONLY CONTRACTS	18.3	15.8	13.7	3.7	9.8	7.3	4.6	22.4	95.7
	INDIRECTS	8.2	3.4	4.2	1.2	2.2	2.2	3.1	5.1	29.6
	EPC CONTRACTS	16.0	2.6	8.8	2.4	4.6	4.4	2.1	9.9	50.8
	INSTALLED COST	136.9	58.9	69.4	18.9	37.8	35.4	48.6	93.8	499.7
	LAND COSTS 5%	6.8	2.9	3.5	0.9	1.9	1.8	2.4	4.7	25.0
	OWNERS COSTS 10%	13.7	5.9	6.9	1.9	3.8	3.5	4.9	9.4	50.0
	CONTINGENCY 25%	34.2	14.7	17.3	4.7	9.4	8.8	12.1	23.5	124.9
	TOTAL PROJECT COST	**191.7**	**83**	**97**	**27**	**53**	**49**	**68**	**131**	**700**

Notes
1) Major Equipment is inclusive of costs up to FOB
2) Direct Bulk Materials includes Piping, Instrumentation, Electrical, Catalyst & Chemicals, Spares and Shipping costs
3) Direct Material & Labour Contracts includes Civil, Steelwork, Building and Protective Cover
4) Labour Only Contracts includes Mechanical, Electrical & Instrumentation, Pre-commisioning Trade Labour Support and Scaffolding Labour costs
5) Indirects includes Temporary Facilities, Heavy Lifts, Commissioning Services and Vendors Engineers
6) EPC Contracts covers Engineering, Procurement and Construction Management
7) Costs are instantaneous 1 Q 2010

ATTACHMENT 13

CAPITAL COST SUMMARY – CASE 2

Project: No 13058
Client : ETI
Project: H2 STORAGE STUDY
Location : UK

Rev : 0
Date :
By : KSW
Printed: 08 March 2013

ORDER OF MAGNITUDE ESTIMATE SUMMARY

Carbon Capture Units

COST CODE	DESCRIPTION	Case 2: IGCC Coal & Biomass Fed power plant with pre-combustion capture 1 TRAIN (50% capacity)								
		Reforming / Gasification (coal handling, milling, storage, gasification etc.) Million's GBP	Air Separation / Oxidant Supply Million's GBP	Acid Gas Removal = Separate H2S & CO2 Removal Million's GBP	Sulphur Plant Million's GBP	Syngas Treatment Unit (including CO Shift & Cooling) Million's GBP	CO2 Compression (to 150 Bar) Million's GBP	Power Block Million's GBP	U&O Million's GBP	Total Million's GBP
	MAJOR EQUIPMENT	56.5	33.0	29.2	7.1	13.1	16.7	27.4	13.0	196.0
	DIRECT BULK MATERIALS	40.9	1.9	10.1	2.5	7.1	4.3	8.3	16.8	91.9
	DIRECT MATERIAL & LABOUR CONTRACTS	3.5	1.0	3.4	0.8	1.5	0.7	3.1	26.7	40.7
	LABOUR ONLY CONTRACTS	19.5	15.3	13.8	3.4	10.1	7.4	4.6	22.4	96.5
	INDIRECTS	8.8	3.3	4.2	1.0	2.3	2.2	3.1	5.1	30.0
	EPC CONTRACTS	16.3	2.6	8.8	2.3	4.7	4.4	2.1	9.9	51.1
	INSTALLED COST	145.6	57.1	69.5	17.2	38.9	35.7	48.6	93.8	506.2
	LAND COSTS 5%	7.3	2.9	3.5	0.9	1.9	1.8	2.4	4.7	25.3
	OWNERS COSTS 10%	14.6	5.7	6.9	1.7	3.9	3.6	4.9	9.4	50.6
	CONTINGENCY 25%	36.4	14.3	17.4	4.3	9.7	8.9	12.1	23.5	126.6
	TOTAL PROJECT COST	**203.8**	**80**	**97**	**24**	**54**	**50**	**68**	**131**	**709**

Notes
1) Major Equipment is inclusive of costs up to FOB
2) Direct Bulk Materials includes Piping, Instrumentation, Electrical, Catalyst & Chemicals, Spares and Shipping costs
3) Direct Material & Labour Contracts includes Civil, Steelwork, Building and Protective Cover
4) Labour Only Contracts includes Mechanical, Electrical & Instrumentation, Pre-commisioning Trade Labour Support and Scaffolding Labour costs
5) Indirects includes Temporary Facilities, Heavy Lifts, Commissioning Services and Vendors Engineers
6) EPC Contracts covers Engineering, Procurement and Construction Management
7) Costs are instantaneous 1 Q 2010

ATTACHMENT 14

CAPITAL COST SUMMARY – CASE 3

Project: No 13058
Client : ETI
Project: H2 STORAGE STUDY
Location : UK

Rev : 0
Date : 28-Feb-13
By : KSW
Printed: 11 March 2013

ORDER OF MAGNITUDE ESTIMATE SUMMARY
Carbon Capture Units

		CASE 3 ATR OPTION							
COST CODE	**DESCRIPTION**	Air Compression **Million's GBP**	Acid Gas Removal = Separate H2S & CO2 Removal **Million's GBP**	Sulphur Plant **Million's GBP**	Syngas Treatment Unit (including CO Shift & Cooling) **Million's GBP**	CO2 Transfer **Million's GBP**	Power Block **Million's GBP**	U&O **Million's GBP**	Total **Million's GBP**
	MAJOR EQUIPMENT	8.9	9.5	0.4	18.6	9.8	20.3	25.3	92.7
	DIRECT BULK MATERIALS	2.4	11.7	0.5	31.3	4.4	7.0	53.5	110.7
	DIRECT MATERIAL & LABOUR CONTRACTS	1.7	3.1	0.1	7.5	0.8	6.2	77.3	96.6
	LABOUR ONLY CONTRACTS	2.4	10.3	0.5	20.8	3.8	4.0	38.8	80.6
	INDIRECTS	0.7	4.3	0.1	5.6	1.4	2.1	20.5	34.7
	EPC CONTRACTS	1.1	5.2	0.5	11.3	1.8	1.8	40.6	62.3
	INSTALLED COST	17.2	44.1	2.0	95.1	21.9	41.5	255.9	477.7
	LAND COSTS 5%	0.9	2.2	0.1	4.8	1.1	2.1	12.8	23.9
	OWNERS COSTS 10%	1.7	4.4	0.2	9.5	2.2	4.1	25.6	47.8
	CONTINGENCY 25%	4.3	11.0	0.5	23.8	5.5	10.4	64.0	119.4
	TOTAL PROJECT COST	**24**	**62**	**3**	**133**	**31**	**58**	**358**	**669**

Notes
1) Major Equipment is inclusive of costs up to FOB
2) Direct Bulk Materials includes Piping, Instrumentation, Electrical, Catalyst & Chemicals, Spares and Shipping costs
3) Direct Material & Labour Contracts includes Civil, Steelwork, Building and Protective Cover
4) Labour Only Contracts includes Mechanical, Electrical & Instrumentation, Pre-commisioning Trade Labour Support and Scaffolding Labour costs
5) Indirects includes Temporary Facilities, Heavy Lifts, Commissioning Services and Vendors Engineers
6) EPC Contracts covers Engineering, Procurement and Construction Management
7) Costs are instantaneous 1 Q 2010

ATTACHMENT 15

CAPITAL COST SUMMARY – CASE 4

Project: No 13058
Client : ETI
Project: H2 STORAGE STUDY
Location : UK

Rev : 0
Date : 28/02/2013
By : KSW
Printed: 08 March 2013

ORDER OF MAGNITUDE ESTIMATE SUMMARY

Carbon Capture Units

COST CODE	DESCRIPTION		Case 4: SMR OPTION								
			Reforming	Air Separation / Oxidant Supply	Acid Gas Removal = Separate H2S & CO2 Removal	Sulphur Plant	Syngas Treatment Unit (including CO Shift & Cooling)	CO2 Compression (to 150 Bar)	Power Block	U&O	Total
			Million's GBP	**Million's GBP**	**Million's GBP**	**Million's GBP**	**Million's GBP**	**Million's GBP**	**Million's GBP**	**Million's GBP**	**Million's GBP**
	MAJOR EQUIPMENT		23.3	76.4	41.6	0.0	0.0	16.0	18.6	12.5	188.4
	DIRECT BULK MATERIALS		16.9	4.4	14.4	0.0	0.0	4.1	5.6	16.1	61.5
	DIRECT MATERIAL & LABOUR CONTRACTS		1.4	2.3	4.8	0.0	0.0	0.7	2.1	25.6	36.9
	LABOUR ONLY CONTRACTS		8.1	35.9	19.6	0.0	0.0	7.1	3.1	21.5	95.3
	INDIRECTS		3.5	8.1	5.8	0.0	0.0	2.1	2.0	4.6	26.1
	EPC CONTRACTS		5.9	5.4	10.6	0.0	0.0	3.6	1.4	9.5	36.4
	INSTALLED COST		59.2	132.5	96.8	0.0	0.0	33.5	32.9	89.8	444.6
	LAND COSTS	5%	3.0	6.6	4.8	0.0	0.0	1.7	1.6	4.5	22.2
	OWNERS COSTS	10%	5.9	13.3	9.7	0.0	0.0	3.3	3.3	9.0	44.5
	CONTINGENCY	25%	14.8	33.1	24.2	0.0	0.0	8.4	8.2	22.5	111.2
	TOTAL PROJECT COST		**83**	**186**	**135**	**0**	**0**	**47**	**46**	**126**	**622**

Notes
1) Major Equipment is inclusive of costs up to FOB
2) Direct Bulk Materials includes Piping, Instrumentation, Electrical, Catalyst & Chemicals, Spares and Shipping costs
3) Direct Material & Labour Contracts includes Civil, Steelwork, Building and Protective Cover
4) Labour Only Contracts includes Mechanical, Electrical & Instrumentation, Pre-commisioning Trade Labour Support and Scaffolding Labour costs
5) Indirects includes Temporary Facilities, Heavy Lifts, Commissioning Services and Vendors Engineers
6) EPC Contracts covers Engineering, Procurement and Construction Management
7) Costs are instantaneous 1 Q 2010

ATTACHMENT 16

CAPITAL COST SUMMARY – CASE 4B

Project: No 13058
Client : ETI
Project: H2 STORAGE STUDY
Location : UK

Rev : 0
Date : 28/02/2013
By : KSW
Printed: 08 March 2013

ORDER OF MAGNITUDE ESTIMATE SUMMARY
Carbon Capture Units

COST CODE	DESCRIPTION	Case 4B: SMR OPTION								
		Reforming Million's GBP	Air Separation / Oxidant Supply Million's GBP	Acid Gas Removal = Separate H2S & CO2 Removal Million's GBP	Sulphur Plant Million's GBP	Syngas Treatment Unit (including CO Shift & Cooling) Million's GBP	CO2 Compression (to 150 Bar) Million's GBP	Power Block Million's GBP	U&O Million's GBP	Total Million's GBP
	MAJOR EQUIPMENT	23.3		41.6	0.0	0.0	16.0	18.6	12.5	112.0
	DIRECT BULK MATERIALS	16.9		14.4	0.0	0.0	4.1	5.6	16.1	57.1
	DIRECT MATERIAL & LABOUR CONTRACTS	1.4		4.8	0.0	0.0	0.7	2.1	25.6	34.6
	LABOUR ONLY CONTRACTS	8.1		19.6	0.0	0.0	7.1	3.1	21.5	59.4
	INDIRECTS	3.5		5.8	0.0	0.0	2.1	2.0	4.6	18.0
	EPC CONTRACTS	5.9		10.6	0.0	0.0	3.6	1.4	9.5	31.0
	INSTALLED COST	59.2	0.0	96.8	0.0	0.0	33.5	32.9	89.8	312.1
	LAND COSTS 5%	3.0		4.8	0.0	0.0	1.7	1.6	4.5	15.6
	OWNERS COSTS 10%	5.9		9.7	0.0	0.0	3.3	3.3	9.0	31.2
	CONTINGENCY 25%	14.8		24.2	0.0	0.0	8.4	8.2	22.5	78.0
	TOTAL PROJECT COST	**83**	0	**135**	**0**	**0**	**47**	**46**	**126**	**437**

Notes
1) Major Equipment is inclusive of costs up to FOB
2) Direct Bulk Materials includes Piping, Instrumentation, Electrical, Catalyst & Chemicals, Spares and Shipping costs
3) Direct Material & Labour Contracts includes Civil, Steelwork, Building and Protective Cover
4) Labour Only Contracts includes Mechanical, Electrical & Instrumentation, Pre-commisioning Trade Labour Support and Scaffolding Labour costs
5) Indirects includes Temporary Facilities, Heavy Lifts, Commissioning Services and Vendors Engineers
6) EPC Contracts covers Engineering, Procurement and Construction Management
7) Costs are instantaneous 1 Q 2010
8).Option 4B excludes ASU

ATTACHMENT 17

CAPITAL COST SUMMARY – POWER ISLAND

Project: No 13058
Client : ETI
Project: H2 STORAGE STUDY

Rev : 0
Date : 12 March 2013
By : KSW
Printed: 14 March 2013

ORDER OF MAGNITUDE ESTIMATE SUMMARY
POWER ISLAND ONLY

COST CODE	DESCRIPTION		Power Island								
			Reforming / Gasification (coal handling, milling, storage, gasification etc.) Million's GBP	Air Separation / Oxidant Supply Million's GBP	Acid Gas Removal = Separate H2S & CO2 Removal Million's GBP	Sulphur Plant Million's GBP	Syngas Treatment Unit (including CO Shift & Cooling) Million's GBP	CO2 Compression (to 150 Bar) Million's GBP	Power Block Million's GBP	U&O Million's GBP	Total Million's GBP
	MAJOR EQUIPMENT								108.2		108.2
	DIRECT BULK MATERIALS								32.6		32.6
	DIRECT MATERIAL & LABOUR CONTRACTS								12.3		12.3
	LABOUR ONLY CONTRACTS								18.2		18.2
	INDIRECTS								12.1		12.1
	EPC CONTRACTS								8.3		8.3
	INSTALLED COST		0	0	0	0	0	0	191.7	0	191.7
	LAND COSTS	5%							9.6		9.6
	OWNERS COSTS	10%							19.2		19.2
	CONTINGENCY	25%							47.9		47.9
	TOTAL PROJECT COST		**0**	**0**	**0**	**0**	**0**	**0**	**268**	**0**	**268**

Notes
1) Major Equipment is inclusive of costs up to FOB
2) Direct Bulk Materials includes Piping, Instrumentation, Electrical, Catalyst & Chemicals, Spares and Shipping costs
3) Direct Material & Labour Contracts includes Civil, Steelwork, Building and Protective Cover
4) Labour Only Contracts includes Mechanical, Electrical & Instrumentation, Pre-commisioning Trade Labour Support and Scaffolding Labour costs
5) Indirects includes Temporary Facilities, Heavy Lifts, Commissioning Services and Vendors Engineers
6) EPC Contracts covers Engineering, Procurement and Construction Management
7) Costs are instantaneous 1 Q 2010

ATTACHMENT 18

COMPARATIVE PROJECT EXECUTION SCHEDULE – CASE 1

FOSTER WHEELER

Page 1 of 1

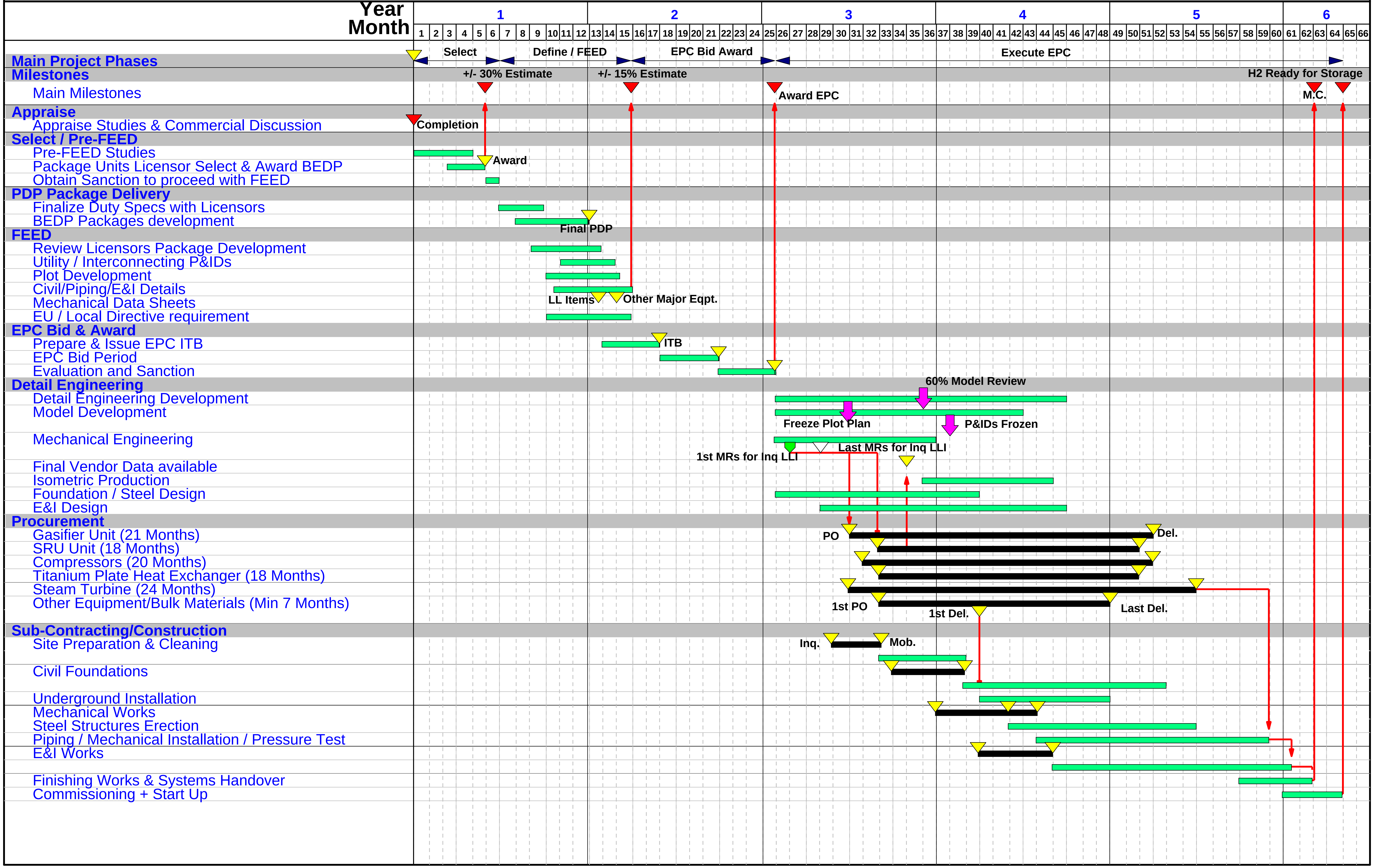
HYDROGEN PRODUCTION OPTIONS

Case 1 - COAL IGCC

Comparative Project Execution Schedule

Rev 0

CLIENT: ETI
PROJECT Nr.: 13058
LOCATION: UK



ATTACHMENT 19

COMPARATIVE PROJECT EXECUTION SCHEDULE – CASE 2

Page 1 of 1

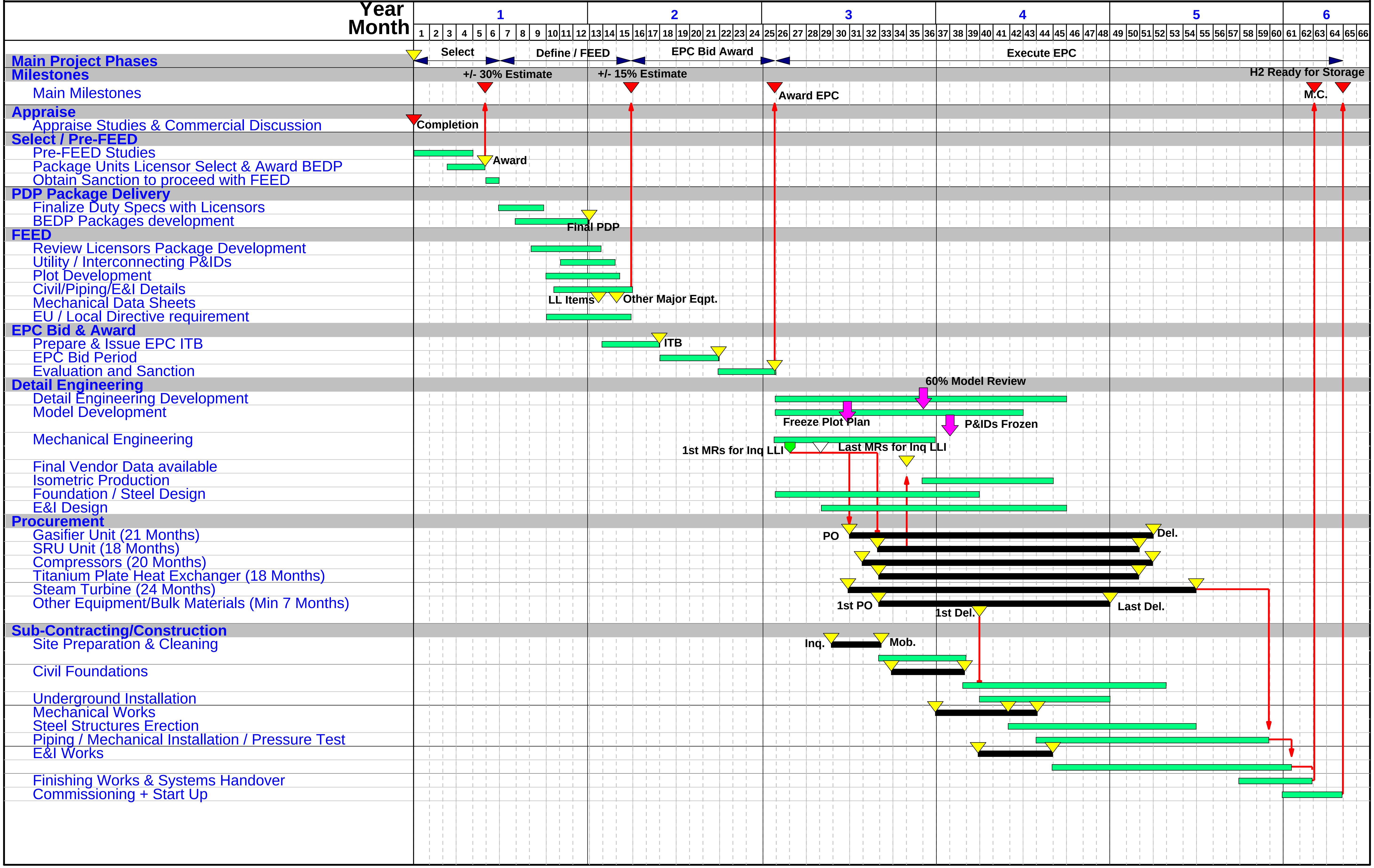
HYDROGEN PRODUCTION OPTIONS

Case 2 - COAL/BIOMASS IGCC

Comparative Project Execution Schedule

Rev 0

CLIENT: ETI
PROJECT Nr.: 13058
LOCATION: UK



Printed 4/3/13

W:\Department\D162\4Peoples Files\Mauro Zito\ETI H2 Storage Study\WP1\Project Schedule\Case 2 - Coal Biomass IGCC\Case 2 2nd draft.mlb

ATTACHMENT 20

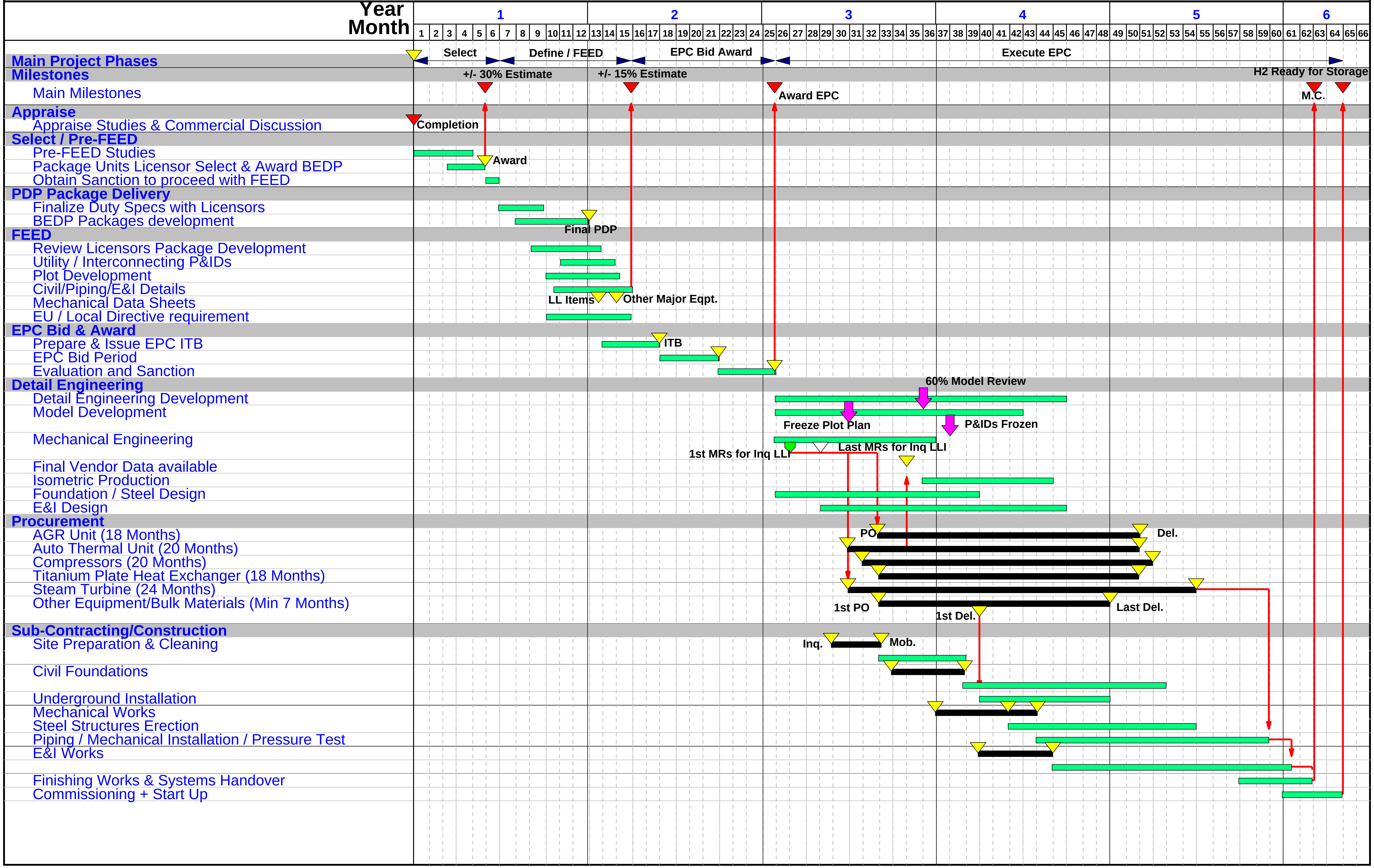
COMPARATIVE PROJECT EXECUTION SCHEDULE – CASE 3

Page 1 of 1

HYDROGEN PRODUCTION OPTIONS
Case 3 - AUTOTHERMAL REFORMING OF NATURAL GAS
Comparative Project Execution Schedule

Rev 0

CLIENT:	ETI
PROJECT Nr.:	13058
LOCATION:	UK



ATTACHMENT 21

COMPARATIVE PROJECT EXECUTION SCHEDULE – CASE 4

FOSTER WHEELER

Page 1 of 1

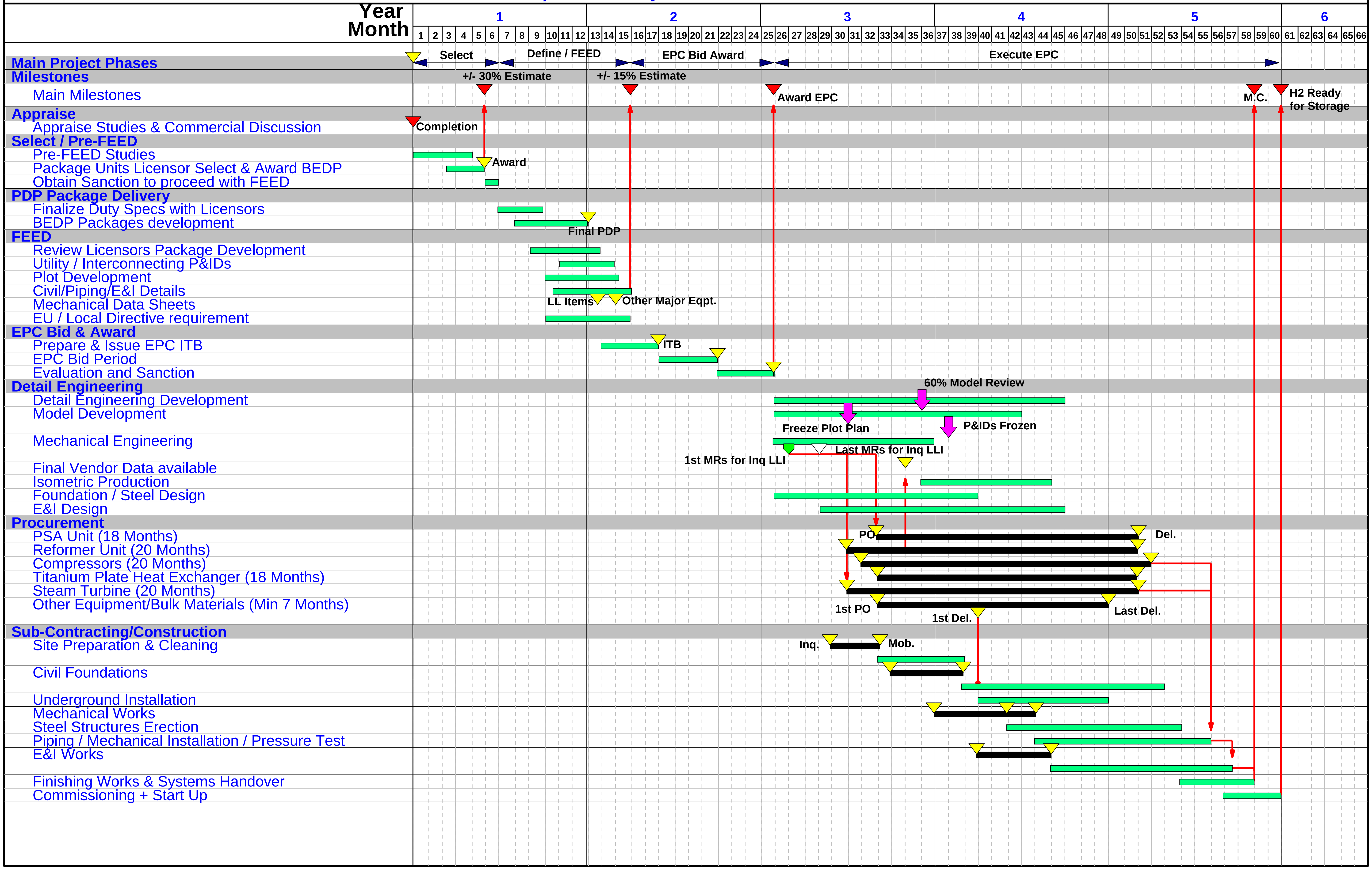
HYDROGEN PRODUCTION OPTIONS

Case 4 - STEAM METHANE REFORMING OF NATURAL GAS

Comparative Project Execution Schedule

Rev 0

CLIENT: ETI
PROJECT Nr.: 13058
LOCATION: UK



ATTACHMENT 22

CAPITAL COST SUMMARY – HYDROGEN PIPELINES

FOSTER WHEELER

Project No: 13058
Client: ETI
Project: H2 Storage & Flexible Turbines
Location: Generic NE England

GAS PIPELINES SUMMARY

B31.8 CODE, GAS AMB & 10BAR
DESIGN FACTOR 0.5, 0.6 or 0.72

Rev : 01
Date : 14-Mar-13
By : KSW

Nom. Dia in.	MATERIAL OF CONSTRUCTION	Pipeline Size OD in.	Wall Thickness mm	PIPELINE MATERIALS GBP/km	PIPELINE CONSTRUCTION GBP/km	PIPELINE CROSSINGS GBP/km	PIPELINE SERVICES GBP/km	OVERALL TOTAL GBP/km
6	API 5L GRB	6.63	7.11	77,900	126,000		22,700	227,000
8	API 5L GRB	8.63	8.18	104,700	164,000		29,900	299,000
10	API 5L GRB	10.75	9.27	132,600	204,000		37,400	374,000
12	API 5L GRB	12.75	9.53	156,400	242,000		44,300	443,000
14	API 5L GRB	14.00	7.92	160,100	266,000		47,300	473,000
16	API 5L GRB	16.00	7.92	186,100	304,000		54,500	545,000
18	API 5L X65	18.00	7.92	222,100	342,000		62,700	627,000
20	API 5L X65	20.00	9.53	262,200	380,000		71,400	714,000
22	API 5L X65	22.00	9.53	296,100	418,000		79,300	793,000
24	API 5L X65	24.00	9.53	334,000	455,000		87,700	877,000
30	API 5L X65	30.00	9.53	468,800	569,000		115,300	1,153,000
32	API 5L X65	32.00	9.53	516,700	607,000		124,900	1,249,000
36	API 5L X65	36.00	9.53	602,500	683,000		142,800	1,428,000
42	API 5L X65	42.00	12.70	862,300	797,000		184,400	1,844,000
48	API 5L X65	48.00	12.70	1,029,500	911,000		215,600	2,156,000
60	API 5L X65	60.00	12.70	1,262,800	1,138,000		266,800	2,668,000

FOSTER WHEELER

Project No: 13058
Client: ETI
Project: H2 Storage & Flexible Turbines
Location: Generic NE England

GAS PIPELINES SUMMARY

B31.8 CODE, GAS 120C & 200BAR
DESIGN FACTOR 0.5

Rev : 0
Date : 08-Mar-13
By : KSW

Nom. Dia in.	MATERIAL OF CONSTRUCTION	Pipeline Size OD in.	Wall Thickness mm	PIPELINE MATERIALS GBP/km	PIPELINE CONSTRUCTION GBP/km	PIPELINE CROSSINGS GBP/km	PIPELINE SERVICES GBP/km	OVERALL TOTAL GBP/km
6	API 5L GRB	6.63	16.94	115,100	126,000		26,800	268,000
8	API 5L GRB	8.63	21.15	168,700	164,000		37,000	370,000
10	API 5L GRB	10.75	25.62	234,000	204,000		48,700	487,000
12	API 5L GRB	12.75	29.83	306,600	242,000		61,000	610,000
14	API 5L GRB	14.00	32.46	360,900	266,000		69,700	697,000
16	API 5L GRB	16.00	36.67	455,800	304,000		84,400	844,000
18	API 5L X65	18.00	23.40	392,400	342,000		81,600	816,000
20	API 5L X65	20.00	25.66	460,200	380,000		93,400	934,000
22	API 5L X65	22.00	27.93	544,700	418,000		107,000	1,070,000
24	API 5L X65	24.00	30.20	640,000	455,000		121,700	1,217,000
30	API 5L X65	30.00	37.00	978,500	569,000		171,900	1,719,000
32	API 5L X65	32.00	39.26	1,105,800	607,000		190,300	1,903,000
36	API 5L X65	36.00	43.80	1,367,500	683,000		227,800	2,278,000
42	API 5L X65	42.00	50.60	1,849,200	797,000		294,000	2,940,000
48	API 5L X65	48.00	57.40	2,361,500	911,000		363,600	3,636,000
60	API 5L X65	60.00	70.99	3,440,200	1,138,000		508,700	5,087,000

FOSTER WHEELER

Project No: 13058
Client: ETI
Project: H2 Storage & Flexible Turbines
Location: Generic NE England

GAS PIPELINES SUMMARY

B31.8 CODE, GAS 120C & 200BAR DESIGN FACTOR 0.6

Rev : 0
Date : 08-Mar-13
By : KSW

Nom. Dia in.	MATERIAL OF CONSTRUCTION	Pipeline Size OD in.	Wall Thickness mm	PIPELINE MATERIALS GBP/km	PIPELINE CONSTRUCTION GBP/km	PIPELINE CROSSINGS GBP/km	PIPELINE SERVICES GBP/km	OVERALL TOTAL GBP/km
6	API 5L GRB	6.63	14.62	106,500	126,000	▨	25,800	258,000
8	API 5L GRB	8.63	18.13	154,400	164,000	▨	35,400	354,000
10	API 5L GRB	10.75	21.85	212,000	204,000	▨	46,200	462,000
12	API 5L GRB	12.75	25.36	275,000	242,000	▨	57,400	574,000
14	API 5L GRB	14.00	27.55	322,700	266,000	▨	65,400	654,000
16	API 5L GRB	16.00	31.06	406,100	304,000	▨	78,900	789,000
18	API 5L X65	18.00	20.00	356,000	342,000	▨	77,600	776,000
20	API 5L X65	20.00	21.89	415,200	380,000	▨	88,400	884,000
22	API 5L X65	22.00	23.78	490,200	418,000	▨	100,900	1,009,000
24	API 5L X65	24.00	25.66	574,000	455,000	▨	114,300	1,143,000
30	API 5L X65	30.00	31.33	876,100	569,000	▨	160,600	1,606,000
32	API 5L X65	32.00	33.22	990,100	607,000	▨	177,500	1,775,000
36	API 5L X65	36.00	37.00	1,221,100	683,000	▨	211,600	2,116,000
42	API 5L X65	42.00	42.66	1,648,300	797,000	▨	271,700	2,717,000
48	API 5L X65	48.00	48.33	2,099,600	911,000	▨	334,500	3,345,000
60	API 5L X65	60.00	59.66	3,030,900	1,138,000	▨	463,200	4,632,000

FOSTER WHEELER

Project No: 13058
Client: ETI
Project: H2 Storage & Flexible Turbines
Location: Generic NE England

GAS PIPELINES SUMMARY

B31.8 CODE, GAS 120C & 200BAR DESIGN FACTOR 0.72

Rev : 0
Date : 08-Mar-13
By : KSW

Nom. Dia in.	MATERIAL OF CONSTRUCTION	Pipeline Size OD in.	Wall Thickness mm	PIPELINE MATERIALS GBP/km	PIPELINE CONSTRUCTION GBP/km	PIPELINE CROSSINGS GBP/km	PIPELINE SERVICES GBP/km	OVERALL TOTAL GBP/km
6	API 5L GRB	6.63	12.68	99,800	126,000		25,100	250,900
8	API 5L GRB	8.63	15.61	141,900	164,000		34,000	339,900
10	API 5L GRB	10.75	18.71	192,800	204,000		44,100	440,900
12	API 5L GRB	12.75	21.63	248,200	242,000		54,500	544,700
14	API 5L GRB	14.00	23.46	290,200	266,000		61,800	618,000
16	API 5L GRB	16.00	26.38	364,000	304,000		74,200	742,200
18	API 5L X65	18.00	17.17	325,400	342,000		74,200	741,600
20	API 5L X65	20.00	18.74	377,000	380,000		84,100	841,100
22	API 5L X65	22.00	20.31	444,300	418,000		95,800	958,100
24	API 5L X65	24.00	21.89	519,500	455,000		108,300	1,082,800
30	API 5L X65	30.00	26.61	790,100	569,000		151,000	1,510,100
32	API 5L X65	32.00	28.18	891,600	607,000		166,500	1,665,100
36	API 5L X65	36.00	31.33	1,096,900	683,000		197,800	1,977,700
42	API 5L X65	42.00	36.05	1,479,100	797,000		252,900	2,529,000
48	API 5L X65	48.00	40.77	1,877,700	911,000		309,900	3,098,600
60	API 5L X65	60.00	50.22	2,684,800	1,138,000		424,800	4,247,600

ATTACHMENT 23

CAPITAL COST SUMMARY – BRINE PIPELINES

FOSTER WHEELER

Project No: 13058
Client: ETI
Project: H2 Storage & Flexible Turbines
Location: Generic NE England

BRINE PIPELINES SUMMARY

B31.4 CODE, BRINE 40C & 20BAR
DESIGN FACTOR 0.5, 0.6 or 0.72

Rev : 01
Date : 14-Mar-13
By : KSW

Nom. Dia in.	MATERIAL OF CONSTRUCTION	Pipeline Size OD in.	Wall Thickness mm	PIPELINE MATERIALS GBP/km	PIPELINE CONSTRUCTION GBP/km	PIPELINE CROSSINGS GBP/km	PIPELINE SERVICES GBP/km	OVERALL TOTAL GBP/km
6	API 5L GRB	6.63	7.11	77,900	126,000	▨	22,700	227,000
8	API 5L GRB	8.63	8.18	104,700	164,000	▨	29,900	299,000
10	API 5L GRB	10.75	9.27	132,600	204,000	▨	37,400	374,000
12	API 5L GRB	12.75	9.53	156,400	242,000	▨	44,300	443,000
14	API 5L GRB	14.00	7.92	160,100	266,000	▨	47,300	473,000
16	API 5L GRB	16.00	7.92	186,100	304,000	▨	54,500	545,000
18	API 5L X65	18.00	7.92	222,100	342,000	▨	62,700	627,000
20	API 5L X65	20.00	9.53	262,200	380,000	▨	71,400	714,000
22	API 5L X65	22.00	9.53	296,100	418,000	▨	79,300	793,000
24	API 5L X65	24.00	9.53	334,000	455,000	▨	87,700	877,000
30	API 5L X65	30.00	9.53	468,800	569,000	▨	115,300	1,153,000
32	API 5L X65	32.00	9.53	516,700	607,000	▨	124,900	1,249,000
36	API 5L X65	36.00	9.53	602,500	683,000	▨	142,800	1,428,000
42	API 5L X65	42.00	12.70	862,300	797,000	▨	184,400	1,844,000
48	API 5L X65	48.00	12.70	1,029,500	911,000	▨	215,600	2,156,000
60	API 5L X65	60.00	12.70	1,262,800	1,138,000	▨	266,800	2,668,000

FOSTER WHEELER

Project No: 13058
Client:ETI
Project: H2 Storage Study
Location: Generic NE England

ETI H2 STORAGE STUDY

OVERALL HDPE PIPELINES SUMMARY - BRINE 20BAR

Rev : '0'
Date : 13-May-13
By : KSW
Printed: 15-May-13

Nom Dia in.	MATERIAL OF CONSTRUCTION	Pipeline Size OD in.	Wall thkness mm	PIPELINE MATERIALS GBP	PIPELINE CONSTRUCTION GBP	PIPELINE CROSSINGS GBP	PIPELINE SERVICES GBP	OVERALL TOTAL GBP
6	HDPE	6.63	17.90	42,200	104,000	▨	16,200	162,000
8	HDPE	8.63	25.20	60,300	135,000	▨	21,700	217,000
10	HDPE	10.75	27.90	69,800	167,000	▨	26,300	263,000
12	HDPE	12.75	35.20	91,400	198,000	▨	32,200	322,000
14	HDPE	14.00	39.70	110,200	218,000	▨	36,500	365,000
16	HDPE	16.00	44.70	136,300	248,000	▨	42,700	427,000
18	HDPE	18.00	50.30	176,600	280,000	▨	50,700	507,000
20	HDPE	20.00	55.80	203,100	311,000	▨	57,100	571,000
22	HDPE	22.00	62.20	248,800	342,000	▨	65,600	656,000

FOSTER WHEELER

Project No: 13058
Client:ETI
Project: H2 Storage Study
Location: Generic NE England

ETI H2 STORAGE STUDY

OVERALL HDPE PIPELINES SUMMARY - BRINE 5BAR

Rev : '0'
Date : 13-May-13
By : KSW
Printed: 15-May-13

Nom Dia in.	MATERIAL OF CONSTRUCTION	Pipeline Size OD in.	Wall thkness mm	PIPELINE MATERIALS GBP	PIPELINE CONSTRUCTION GBP	PIPELINE CROSSINGS GBP	PIPELINE SERVICES GBP	OVERALL TOTAL GBP
6	**HDPE**	**6.63**	**4.90**	**33,700**	**104,000**		**15,300**	**153,000**
8	**HDPE**	**8.63**	**6.90**	**43,300**	**135,000**		**19,800**	**198,000**
10	**HDPE**	**10.75**	**7.70**	**48,800**	**167,000**		**24,000**	**240,000**
12	**HDPE**	**12.75**	**9.70**	**58,400**	**198,000**		**28,500**	**285,000**
14	**HDPE**	**14.00**	**10.90**	**68,200**	**218,000**		**31,800**	**318,000**
16	**HDPE**	**16.00**	**12.30**	**83,200**	**248,000**		**36,800**	**368,000**
18	**HDPE**	**18.00**	**13.80**	**106,100**	**280,000**		**42,900**	**429,000**
20	**HDPE**	**20.00**	**15.30**	**120,000**	**311,000**		**47,900**	**479,000**
22	**HDPE**	**22.00**	**17.20**	**145,300**	**342,000**		**54,100**	**541,000**
24	**HDPE**	**24.00**	**18.00**	**170,500**	**373,000**		**60,400**	**604,000**
30	**HDPE**	**30.00**	**23.10**	**281,400**	**467,000**		**83,200**	**832,000**
32	**HDPE**	**32.00**	**24.50**	**322,600**	**498,000**		**91,200**	**912,000**
36	**HDPE**	**36.00**	**27.60**	**396,400**	**560,000**		**106,300**	**1,063,000**
42	**HDPE**	**42.00**	**32.00**	**573,900**	**654,000**		**136,400**	**1,364,000**
48	**HDPE**	**48.00**	**36.70**	**704,800**	**746,000**		**161,200**	**1,612,000**
60	**HDPE**	**60.00**	**45.00**	**966,100**	**933,000**		**211,000**	**2,110,000**