Department for Business, Energy & Industrial Strategy

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Benchmarking State-of-the-art and Next Generation Technologies

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Report for

Department for Business, Energy & Industrial Strategy

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Benchmarking State-of-the-art and Next Generation Technologies

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1 Acknowledgements

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2 Executive Summary

The Department for Business, Energy and Industrial Strategy (BEIS) has commissioned Wood to execute a study assessing the most promising CO_2 capture technologies in order to inform future innovation spending programmes and to shape future policy direction for carbon capture technologies in the Power and Energy Intensive Industries. The study also aims to evaluate the cost reduction potential and competitiveness of novel UK carbon capture technologies that may be implemented over the next thirty years.

In order to evaluate promising new technologies against current state-of-the-art technologies, power plant benchmark cases were developed for eight current carbon capture processes and two leading next generation carbon capture technologies. All benchmark cases were set to capture at least 90% of the CO_2 arising within the process. In addition to the ten benchmarks, an unabated natural gas CCGT case was also included, referred to as Case 0, which is used as a comparator for a typical state of the art UK power plant. Natural gas and coal fired cases were developed with a target net electrical power output in the range 800 – 1200 MW. However, due to the current limitations on scale of biomass boilers, the last three power benchmarks were set with a smaller capacity in the range 300 – 400 MW.

The benchmark power generation cases presented in this report are:

- Case 0 – Reference Case – Unabated natural gas CCGT
- Case 1 – Natural gas CCGT with post-combustion carbon capture
- Case 2 – Natural gas reformation with pre-combustion carbon capture
- Case 3 – Coal SCPC with post-combustion carbon capture
- Case 4 – Coal SCPC with oxy-combustion carbon capture
- Case 5 – Coal IGCC with pre-combustion carbon capture
- Case 6 – Oxy-fired supercritical gas power generation with carbon capture
- Case 7 – Natural Gas CCGT with Molten Carbonate Fuel Cell carbon capture
- Case 8 – Biomass CFB boiler with post-combustion carbon capture
- Case 9 – Biomass CFB boiler with oxy-combustion carbon capture
- Case 10 – Biomass IGCC with pre-combustion carbon capture

Assessment of novel technologies in the Energy Intensive Industries has been limited to hydrogen production for industrial uses such as refineries and petrochemical facilities. The benchmark case for hydrogen production is based on a conventional steam methane reformer with post-combustion capture of carbon dioxide on the reformer flue gas. This has been scaled for a capacity of 100,000 Nm^3/h, which is a typical size for a single reformer. As with the power generation cases, an unabated hydrogen reference case has also been assessed. Hence, the following hydrogen benchmark cases are presented in this report:

- Case H – Reference Case – Unabated natural gas steam methane reformer
- Case 11 – Natural gas steam methane reformer with post-combustion carbon capture

Results for the hydrogen benchmark case are presented in Section 16 and are not expanded on in this Executive Summary.

Power Generation Case Results Comparison

The technical performance for the Reference Case and ten benchmark cases are shown in Table 2-2. The parasitic demands for the benchmark cases reflect the addition of a CO_2 capture process, which also results in a reduced net efficiency when compared to Case 0.

The plant performance for the natural gas fired CCGT cases (Case 0, Case 1 and Case 7) is based on one of the largest and most efficient natural gas turbines (GE 9HA.01) with power output of over 400 MWe per turbine and a combined cycle net efficiency[1] of over 62%[2]. The fossil-fired pre-combustion cases (Case 2 and Case 5) feature plant performance based on a GE Frame 9 syngas variant gas turbine fired on syngas produced from either natural gas reforming (Case 2) or coal gasification (Case 5). The syngas variant gas turbine has a power output of about 300 MWe per turbine and a combined cycle net efficiency of approximately 47%.

It is evident that although all of the benchmark cases capture around 90% of the CO_2 generated by the process, the overall carbon dioxide emissions for the coal and biomass cases are much higher than the natural gas cases due to inherent higher carbon fraction in coal and biomass.

The biomass cases suffer from reduced efficiency compared to coal cases due to the smaller scale of the units, the lower inherent efficiency of subcritical boilers and the low energy density of the feedstock. For example, the IEAGHG Biomass CCS Study 2009/09 reported a 3% delta in net efficiency between a supercritical pulverised coal boiler co-firing 10% biomass and a smaller subcritical CFB boiler firing 100% biomass (44.8% vs 41.7%). Supercritical pulverised coal boilers are not suitable for all biomass feeds due to the difficulties of milling the biomass feed to a suitable particle size, which restricts the potential range of feedstocks. Drax in Yorkshire uses imported wood pellets, which can be milled to the same powder consistency as pulverised coal, but wood pellets are more expensive.

For a new build state-of-the-art dedicated biomass fired power plant in the range of 250-300 MWe, subcritical circulating fluidised bed (CFB) technology is preferred and accepted in the industry, as opposed to a pulverised biomass boiler. This allows the use of variable quality biomass, allowing a broader range of potential fuel suppliers, reducing the risk associated with biomass supply chain and logistics. For example, a recent award for a 100 MWe biomass-fired power plant in Dangjin[3], the CFB boiler will be designed to run on a range of feedstocks including wood pellets, wood chips and palm kernel shells. The biomass post-combustion and oxy-combustion benchmarks (Case 8 and 9) used Sumitomo Foster Wheeler's 'Compact' tower subcritical circulating fluidised bed (CFB) boiler using virgin biomass wood chip fuel as the basis of design.

Pre-conditioning of biomass fuel is an option to increase the hardness and density of the fuel to make it suitable for pulverised boiler mills. In this study, biomass IGCC pre-combustion (Case 10) assumed torrefied biomass, as this has been demonstrated on an IGCC on biomass fuel field trials.

Table 2-3 shows the economic performance for the reference case and ten benchmark power cases. All costs in the table are presented in British pounds on a Q1 2017 basis. The project cost varies greatly between the cases with coal cases having higher costs than gas cases, partly due to feedstock handling and the more complex process steps required to produce power cleanly. It is evident that Coal-IGCC is the most expensive case, as it involves several process steps to produce power including gasification, carbon monoxide (CO) shift, carbon dioxide / hydrogen sulphide (CO_2 / H_2S) capture and combined cycle. Overall, this makes the IGCC power plant cost 3.5 times as much as the reference Case 0. The biomass cases are even more expensive on a specific cost

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[1] Power net efficiencies provided in this report are calculated using Lower Heating Value (LHV).

[2] H-class gas turbine efficiencies are expected to improve as the product range matures. For example, Bouchain in France has recorded a net efficiency of 62.22% (http://www.powermag.com/worlds-most-efficient-combined-cycle-plant-edf-bouchain/?printmode=1), and GE are advertising the 9HA.02 at near 64% (https://www.genewsroom.com/press-releases/ha-technology-now-available-industry-first-64-percent-efficiency-284144).

[3] https://www.iea-coal.org/sumitomo-shi-fw-wins-contract-for-biomass-cfb-boiler-island/

basis, considering the cost per unit of installed power export capacity, due to the higher volumes of feedstock required and low energy density.

Operating costs for the coal cases, with the exception of fuel costs, are higher than in the natural gas cases. This is mainly due to the higher capital and labour cost of coal cases, higher carbon dioxide emissions and higher CO_2 storage / transportation cost due to the larger volume of CO_2 to be transported. This is balanced to some extent by the lower fuel costs for bulk coal purchase. By contrast, the biomass feed costs are relatively high, although the market for farming and marketing of biomass crops in the UK is likely to develop significantly in the future, which should reduce the costs for large-scale purchase. It should be noted that a large proportion of the fixed operating cost estimates in this study are taken as being proportional to the capital cost estimate, without further differentiation between cases.

Levelised Cost of Electricity (LCOE) is provided as a means to compare the overall costs of building and operating a plant for the duration of its anticipated lifetime on a consistent basis, thus allowing the cost / benefit of a high capital cost, high efficiency plant to be compared with that of a lower capital cost, lower efficiency plant. The resultant figure is an indication of the mean electricity price that would be needed by a power project in order to break even over the life of the plant (i.e. Net Present Value = 0). The LCOE calculation takes into account the capital and operating costs and also reflects the different net power output from the different cases. Both LCOE and net electrical power output are shown in Figure 2-1 below.

In order to avoid potential differences in financing models distorting the cost differentials that arise for technological reasons, a constant hurdle rate is used across all of Cases 1 – 10. This means that the LCOE figures for these cases should not be interpreted as a best estimate of the price that a typical project might need to deploy in 2025, but as comparative benchmarks between the cases.

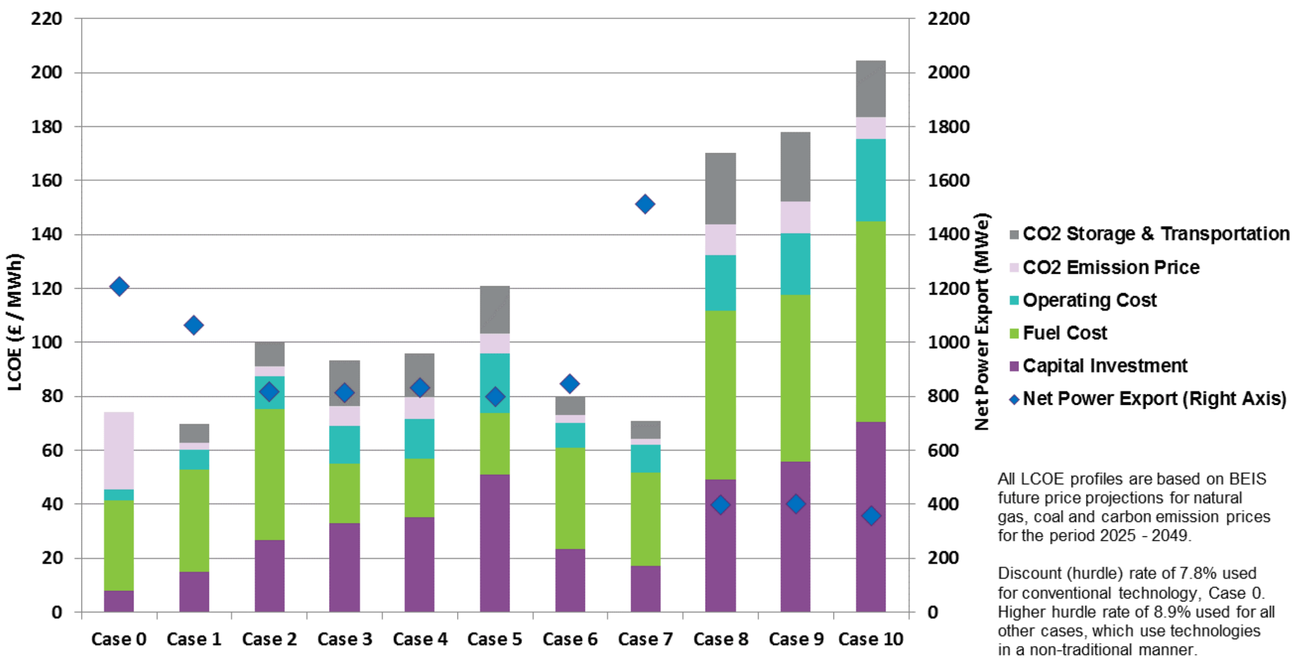

Figure 2-1: LCOE (£/MWh) Contribution for all Power Cases

Table 2-1: Split of LCOE Contribution for Power Cases

LCOE Contribution (£ / MWh)	Capital Cost	Fuel	Operating Cost	Emissions Price	Storage & Transport	Total
Case 0	8.0	33.5	4.0	28.7	0	74.2
Case 1	14.9	37.9	7.2	2.9	7.0	69.9
Case 2	26.6	48.5	12.2	3.8	8.9	100.0
Case 3	32.9	22.2	13.8	7.5	16.9	93.3
Case 4	35.3	21.7	14.7	8.0	16.3	96.0
Case 5	51.1	22.8	22.0	7.5	17.4	120.8
Case 6	23.2	37.7	9.2	3.1	6.9	80.1
Case 7	17.1	34.7	10.1	2.3	6.5	70.7
Case 8	49.0	62.6	20.6	11.7	26.2	170.1
Case 9	55.9	61.8	22.8	11.6	25.8	177.9
Case 10	70.6	74.1	30.5	8.4	20.7	204.3

The stacked bars represent the aggregated contributions of different cost elements towards an overall LCOE for each case. If all of the cases had roughly the same net power export, then the same comparison could be performed on the actual costs for each element. However, when considering the LCOE, it must be recognised that a plant with greater power export will see a reduced LCOE for an equal cost in any category. Diamonds representing the net power export refer to the right-hand axis.

Although the unabated CCGT case, Case 0, has the lowest overall investment cost, it does not result in the lowest overall LCOE. The lowest overall LCOE is provided by Case 1, the CCGT plant with state-of-the-art post-combustion carbon capture. Case 0 features a significant proportion of LCOE arising from the penalty paid for emitting CO_2, which is included in the financial analysis for this study, demonstrating the importance of the carbon price as a potential tool for encouraging low carbon investments in power plant. Please note, in Table 2-3 there are two cost of avoided CO_2 metrics: one that includes the effect of a carbon price, and one that doesn't include a carbon price as this allows the later metric to be compared to the methodology used by other international benchmarking studies. The cost of avoided CO_2 metric that is of relevance to UK (and other countries / regions with a price on CO_2) is the one that includes the effect of a carbon price.

The natural gas fired power plant with integrated reforming and pre-combustion carbon capture, Case 2, does not compete well against the more straight-forward post-combustion case (Case 1). This case has higher capital and operating costs than Case 1 and the power output available for electricity sales is also significantly lower, despite having approximately the same gas feed rate. This result has been seen in other comparative studies. The approach of using natural gas reformation with pre-combustion capture appears to lack promise as a basis for standalone power plant developments. Its strength lies in facilities that require reformed hydrogen as part of a larger refining or petrochemical facility and which can produce and store excess hydrogen for peak-shaving power plants. In the more flexible operation approach, the capacity of the front end of the plant, the reforming and hydrogen production, can be reduced in size relative to the power island if storage is used to meet intermittent power generation requirements. This benefit is not accounted for within the scope of this study. Natural gas reforming processes may also provide a route to decarbonisation of the gas distribution vector, which is used widely for domestic and commercial heating in the UK.

All of the coal-based cases suffer from higher LCOE, which is partly due to their much lower thermal efficiency and partly the result of the higher capex and operating costs associated with these cases. It should be noted that the natural gas and coal UK price forecast sets used for the study reflect a market which has seen a significant impact from global shale gas production, with a long term low price anticipated for natural gas. A different outcome may arise for other countries where coal is abundant, but gas is more difficult to source.

Within the coal cases, it can be seen that the post-combustion and oxy-combustion cases generate electricity at a similar cost. The main difference between these cases is that the oxy-combustion route has not been demonstrated at commercial scale, primarily because the focus of CCUS demonstration in North America (at Boundary Dam and Petra Nova) has been a retrofit to an existing coal boiler. For a future power plant with carbon capture based on a coal feed, both options should be considered. The IGCC case has a much higher capital cost, which makes this an expensive route to generate electricity.

Both of the novel gas-fired international benchmark cases perform well in comparison to the coal cases but demonstrate very similar performance to the CCGT post-combustion benchmark, which is significantly more mature than either gas-fired international benchmark. The oxy-fired supercritical gas turbine (Case 6) achieves a LCOE about 15% higher than Case 1, whilst the LCOE for the CCGT combined with Molten Carbonate Fuel Cells (Case 7) is only marginally higher than Case 1. Both of these technologies have yet to prove themselves at demonstration level before full commercialisation and both cases contain a degree of uncertainty with regard to costs and performance. We have made what we believe to be reasonably balanced assumptions in these areas, assuming an Nth-of-a-kind project philosophy that may be applicable for plants in a generation's time.

Both of the international novel technologies will need to demonstrate their potential improvements in cost and performance before they can be reasonably expected to compete with or outperform the state-of-the-art technology. It will be interesting to assess these technologies again in a few years' time to see how the technology has developed.

An initial view of Figure 2-1 implies that the biomass cases 8-10 are unable to compete against either coal or natural gas fired power plant. However, there are several important elements to take into consideration when viewing these results. Firstly, the biomass cases do not benefit from the same economy of scale as the other cases. The study has considered plant with an exportable power output of around 400 MWe, which is half the capacity of the coal cases and one third of the unabated natural gas case. To illustrate the relative size of the biomass boilers, case 8 & 9 are a quarter the size of their respective coal boilers. This scale was selected to ensure reliable results for existing commercial boilers. Secondly, the biomass cases suffer from high feedstock prices. The prices used for this study are based on delivery to small-scale users such as local CHP facilities. As a larger market develops, it would be fair to assume that real-terms prices would fall. Finally, the figure cannot represent the most important benefit of biomass fired power: the overall life-cycle analysis should result in a net reduction in atmospheric carbon dioxide. It is difficult to put a value on absorbing and capturing CO_2 from the air, but it is hoped that this study and future studies will support the UK Government in developing appropriate incentives to drive this development.

Within the biomass cases, the results are similar to those for the coal cases. Post-combustion and oxy-combustion both provide viable routes with a similar price for electricity production, whilst the IGCC approach is hampered by greater capital costs. The biomass pre-combustion (bio-IGCC) case could be used as a source of renewable hydrogen that can be injected into the gas grid or can be used as a green feedstock for other chemicals. However, there is no way of accounting for those potential economic benefits which would make the bio-IGCC case attractive within the scope of this study.

Table 2-2: Summary of Power Benchmark Case Key Technical Performance

	Units	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7	Case 8	Case 9	Case 10
Total Gross Installed Capacity	MWe	1229	1144	919	953	1113	1063	1264	1645	498	598	493
Gas Turbine (s)	MWe	823	823	554	0	0	671	1264	823	0	0	303
Steam Turbine	MWe	406	321	365	953	1098	392	0	381	498	598	190
Others	MWe	0	0	0	0	15	0	0	440	0	0	0
Total Auxiliary Loads	MWe	21	80	101	139	280	263	416	136	102	196	137
Net Power Export	MWe	1208	1065	818	814	833	800	848	1509	396	402	356
Fuel Flow Rate	kg/h	150,296	150,296	147,539	325,000	325,000	314,899	118,940	195,722	635,178	635,178	225,417
Fuel Flow Rate (LHV)	MWth	1940	1940	1907	2335	2335	2263	1536	2527	1288	1288	1052
Net Efficiency (LHV) - As New	%	62.3	54.9	42.9	34.9	35.7	35.3	55.2	59.7	30.8	31.2	33.9
Net Efficiency (LHV) - Average	%	59.0	52.0	40.7	34.7	35.5	33.5	52.3	56.6	30.6	31.1	32.1
Plant Availability	%	93	90	85	90	90	85	90	90	90	90	85
Total Carbon in Feeds	kg/h	108,640	108,640	106,647	209,950	209,950	203,425	85,975	141,476	158,795	158,795	107,095
Total Carbon Captured	kg/h	0	98,661	96,418	188,926	187,176	183,697	77,378	130,333	142,954	142,748	97,194
Total CO_2 Captured	kg/h	0	361,539	353,319	692,310	685,896	673,147	283,546	477,597	523,849	523,093	356,162
Total CO_2 Emissions	kg/h	398,105	36,566	37,483	77,040	83,455	72,292	31,503	40,934	58,045	58,801	36,283
CO_2 Capture Rate	%	0	90.8	90.4	90.0	89.2	90.3	90.0	92.1	90.0	89.9	90.8
Carbon Footprint	kg CO_2/MWh	329.4	34.3	45.8	94.6	100.2	90.4	37.1	27.1	146.5	146.2	101.9

Table 2-3: Summary of Power Benchmark Case Key Economic Performance

	Units	Case 0	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7	Case 8	Case 9	Case 10
Total Project Cost	£M	672	968	1256	1732	1902	2396	1213	1570	1248	1450	1465
Pre-Licensing, Tech & Design	£M	6	8	11	15	17	22	11	14	11	13	13
Regulatory & Public Enquiry	£M	13	18	24	32	35	44	23	29	23	27	27
EPC Contract Cost	£M	584	845	1107	1547	1702	2151	1068	1392	1107	1290	1305
Infrastructure Connections	£M	29	37	37	29	29	29	37	37	29	29	29
Owner's Costs	£M	41	59	77	108	119	151	75	97	77	90	91
Overall CAPEX Impact (vs Ref Case)		-	44%	87%	158%	183%	256%	80%	134%	86%	116%	118%
Estimate Accuracy		± 30%	± 30%	± 30%	± 30%	± 35%	± 35%	± 45%	± 40%	± 40%	± 40%	± 40%
Total Fixed OPEX	£M pa	36	47	60	81	87	112	55	72	58	66	70
Total Variable OPEX (excl. Fuel & C)	£M pa	0	62	58	108	108	103	44	108	82	82	54
Average Fuel Cost (1)	£M pa	315	305	283	143	143	131	242	398	190	190	183
Typical CO_2 Emission Cost (1)	£M pa	369	33	32	69	75	61	28	37	52	53	31
Discount Rate	% / year	7.8 (2)	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9	8.9
Levelised Cost of Electricity	**£/MWh**	**74.2**	**69.9**	**100.0**	**93.3**	**96.0**	**120.8**	**80.1**	**70.7**	**170.1**	**177.9**	**204.3**
Capital Investment	£/MWh	8.0	14.9	26.2	32.9	35.3	51.1	23.2	17.1	49.0	55.9	70.6
Fuel Cost	£/MWh	33.5	37.9	48.5	22.2	21.7	22.8	37.7	34.7	62.6	61.8	74.1
Operating Cost	£/MWh	4.0	7.2	12.2	13.8	14.7	22.0	9.2	10.1	20.6	22.8	30.5
CO_2 Emissions Price	£/MWh	28.7	2.9	3.8	7.5	8.0	7.5	3.1	2.3	11.7	11.6	8.4
CO_2 Storage & Transportation	£/MWh	0	7.0	8.9	16.9	16.3	17.4	6.9	6.5	26.2	25.8	20.7
Cost of CO_2 Avoided (incl. Carbon Price)	**£/tCO_2**	**-**	**-14.5**	**91.1**	**81.3**	**95.1**	**195.1**	**20.0**	**-11.7**	**524.1**	**566.1**	**571.7**
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	67.1	96.2	85.8	88.0	113.3	77.0	68.4	158.4	166.3	195.8
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	73.1	178.9	171.4	185.5	283.8	107.7	75.8	617.2	659.3	660.7

Note 1: Fuel and Carbon Price profiles per Table 5-7 used for the analysis. Average values across 2025-2049 shown for comparison purposes only.

Note 2: Discount rate for proven conventional technology set at BEIS standard rate of 7.8%. Other technologies with an element of technological risk set at an illustrative higher discount rate of 8.9%.

3 Abbreviations

AACE	American Association of Cost Estimators
Abs.	Absorber
ACCE	Aspen Capital Cost Estimator
ASU	Air Separation Unit
ATR	Auto-Thermal Reformer
BEIS	Department for Business, Energy & Industrial Strategy
BFW	Boiler Feed Water
CAPEX	Capital Expenditure
CCGT	Combined Cycle Gas Turbine
CFB	Circulating Fluidised Bed
CO_2	Carbon Dioxide
Comp	Compressor
CPU	Cryogenic Purification Unit
C&I	Control and Instrumentation
DCC	Direct Contact Cooler
DCO	Development Consent Order
DECC	Department for Energy and Climate Change (now part of BEIS)
EIA	Environmental Impact Assessment
EPC	Engineering, Procurement & Construction
EPCCI	European Power Capital Cost Index
ESP	Electrostatic Precipitator
FEED	Front End Engineering Design
FGD	Flue Gas Desulphurisation
FID	Final Investment Decision
FOAK	First of a Kind
GBP	British Pounds Sterling
GGH	Gas-Gas Heat exchanger
GTG	Gas Turbine Generator
GTW	Gas Turbine World
HP	High Pressure
HRSG	Heat Recovery Steam Generation
HSS	Heat Stable Salts
HHV	Higher Heating Value
IEAGHG	International Energy Agency Greenhouse Gas R&D Programme
IGCC	Integrated Gasification Combined Cycle
IRCC	Integrated Reforming Combined Cycle
LCOE	Levelised Cost of Electricity
LCOH	Levelised Cost of Hydrogen
LHV	Lower Heating Value
LP	Low Pressure
MCFC	Molten Carbonate Fuel Cell

MHE	Main Heat Exchanger
MP	Medium Pressure
MVR	Mechanical Vapour Recovery
MWe	Mega Watt (electrical output)
MWth	Mega Watt (thermal input – typically referring to LHV)
NA	Not Applicable
NG	Natural Gas
NOAK	Nth of a Kind
NOx	Oxides of Nitrogen
OHTL	Overhead Transmission Line
ONS	Office for National Statistics
OPEX	Operating Expenditure
O&U	Offsites and Utilities
P/L	Pipeline
ppmv	Parts Per Million Volume basis
PSA	Pressure Swing Adsorption
RPI	Retail Price Index
SCGP	Shell Coal Gasification Process
SCPC	Supercritical Pulverised Coal
SMR	Steam Methane Reformer
SOx	Oxides of Sulphur
STG	Steam Turbine Generator
Str.	Stripper
SRU	Sulphur Recovery Unit
TEG	Triethylene Glycol
TSA	Temperature Swing Adsorption

4 Introduction

The Department for Business, Energy and Industrial Strategy (BEIS) has commissioned Wood[4] to carry out a study to assess the cost reduction potential and competitiveness of novel (Next Generation) UK carbon capture technologies.

The aim of the study is to assist BEIS in evaluating the most promising CO_2 capture technologies in order to inform future innovation spending programmes and to shape future policy direction for carbon capture technologies in the Power and Energy Intensive Industries (EIIs) respectively.

In order to evaluate promising new technologies, it is necessary to define a basis for technologies to be compared against each other, but also crucial to be able to compare the technologies with the current state-of-the-art technologies. This will make it possible to understand whether or not the new technologies being proposed have the potential to exceed the performance of technologies already available, including consideration that the current technologies will also be anticipated to realise minor performance improvements over the time it would take to bring a new technology to market.

The first stage of the study was to conduct a literature review covering the full range of applicable carbon capture technologies. The aim was to provide a sound background to the range of technologies that have been considered in recent years to help UK-led developers to source background information in support of their processes. This is documented in a separate report, 13333-8820-RP-003.

The next stage of the study was to develop benchmark cases for natural gas, coal and biomass-fired power plants built in the UK with current state-of-the-art carbon capture. Two alternative benchmark cases were added to reflect leading international next generation carbon capture technologies. A hydrogen production benchmark case was also created, to allow assessment of next generation technologies for industrial hydrogen production. Benchmarks for other Energy Intensive Industries have not been generated as part of the current study.

The cost estimating approach used for this study is aligned with AACE Class IV estimates, using a mixture of sized equipment lists costed using past project data and / or budget quotations, plus some packaged units scaled from reliable benchmarks. For well-defined processes using conventional technologies, this produces capital cost estimates of +/- 30%. For less developed technologies, where the scope of supply is less defined, our confidence in the cost estimates is lower, as reported in the results for each benchmark case.

This report documents the benchmarking methodology, the data that has been used, key assumptions and decisions that have been made in order to define the benchmark cases and the results for each case.

Note that in order to protect the intellectual property of UK-led developers, the assessment of technologies that were submitted to BEIS as part of this study is not included in this public-domain report.

[4] The contract was awarded to Amec Foster Wheeler Group Limited prior to the takeover by the John Wood Group plc in October 2017.

5 Assessment Methodology

5.1 Case Selection

The purpose of the benchmark cases is to provide a set of consistently based designs of power and hydrogen plants with carbon capture for comparison with the UK-led technologies. This comparison will enable BEIS to determine which of the UK-led technologies have the potential to be better than the current state-of-the-art and most promising international novel technologies, and which do not appear to have any advantage over technologies already available to the market.

Ten benchmark power cases have been selected covering natural gas, coal and biomass firing options. The flow schemes include post combustion, pre-combustion and oxy-fired schemes which feature capture of 90% of the carbon dioxide in the plant feedstocks. The two leading international novel technology benchmarks can capture at least 90% of the carbon in their feed streams. This range of cases is anticipated to capture all of the best performing technologies against which a new technology would need to compete.

The single benchmark hydrogen case reflects a Steam Methane Reformer, typical of units found in refineries and chemical facilities across the globe, but with post-combustion carbon capture fitted to the exhaust from the reformer.

The benchmark cases selected are:

- Case 1 – Natural Gas CCGT with post-combustion carbon capture
- Case 2 – Natural Gas reforming with pre-combustion carbon capture
- Case 3 – Coal SCPC with post-combustion carbon capture
- Case 4 – Coal SCPC with oxy-combustion carbon capture
- Case 5 – Coal IGCC with pre-combustion carbon capture
- Case 6 – Oxy-fired supercritical gas power generation with carbon capture
- Case 7 – Natural Gas CCGT with Molten Carbonate Fuel Cell carbon capture
- Case 8 – Biomass CFB boiler with post-combustion carbon capture
- Case 9 – Biomass CFB boiler with oxy-combustion carbon capture
- Case 10 – Biomass IGCC with pre-combustion carbon capture
- Case 11 – Steam Methane Reformer with post-combustion carbon capture

Combined cycle gas turbine plants are leading the way in terms of both high efficiency and capital cost effective capacity provision as well as flexibility to respond to grid demand. They are also by far the largest source of fossil fuel based power in the UK at the time of writing. Therefore, it is sensible to include benchmark cases which cover the various potential pathways to decarbonise this type of plant (Cases 1, 2, 6 & 7).

The leading technologies for the post-combustion route are all proprietary amine based solvent systems, with Cansolv currently viewed as representative of an advanced and demonstrated technology provider. This is due to operation of the Boundary Dam coal power plant since 2014 and design development learning from the UK CCS Commercialisation Programme, Peterhead gas power plant project. Therefore, a CCGT with Shell Cansolv post-combustion capture was selected as the first benchmark case, Case 1 (this case represents a scale up of approximately 2.5 times the Peterhead CO_2 production rate). This technology application can be considered TRL-7 as it has been demonstrated at smaller scale for fertiliser manufacture but has yet to be deployed at a scale comparable to Case 1. Shell Cansolv has supported this study, providing process design packages for each of the post-combustion cases (Cases 1, 3, 8 and 11).

Pre-combustion capture on a natural gas based plant is understood to lag significantly behind the post-combustion scheme both in terms of technical efficiency and cost, when considered for purely baseload electricity generation. However, to ensure that this previously observed finding is still correct, it is prudent to include an up to date natural gas reforming combined cycle scheme for pre-combustion CO_2 capture, included as Case 2. This technology application can be considered TRL-5-6 as the main components of the system have all been demonstrated at full scale, but the process arrangement specific to the IRCC scheme has not been demonstrated.

The UK uses both natural gas and coal for utility scale power production, although the use of coal is reducing as fuel-switching is employed as an initial method of achieving CO_2 emissions reductions in line with national targets. Oil and other liquid fuels are not used for utility scale base load power generation to any significant degree used in the UK. Thus, no oil fired cases are included.

The highest efficiency, largest scale coal plants are ultra-supercritical pulverised coal steam generator plants, also referred to as supercritical pulverised coal plants (SCPC). These plants feature steam conditions equal to or greater than 600°C and 220 bar (abs). The use of the term "ultra" is variable, commonly but not exclusively used for conditions above 620°C and 260 bar (abs). This type of plant is suitable for application of post-combustion CO_2 capture technology in the same way as it can be applied to a CCGT plant although additional flue gas pre-treatment, a slightly different solvent and some modifications to the heat integration scheme may be recommended by the licensor. A supercritical pulverised coal plant with Shell Cansolv CO_2 capture is included as Case 3 (this case represents a scale up of approximately 5 times the Boundary Dam CO_2 production rate). This technology application can be considered TRL-9 as there are already two operating plants employing proprietary amine solvents for post combustion CO_2 capture on coal plants globally, including Boundary Dam (1 MTPA), and the larger Petra Nova plant (1.6 MTPA CO_2), which uses MHI technology.

Oxy-combustion carbon capture can also be applied to a SCPC steam generator, which is expected to result in similar technical and economic performance compared with the post-combustion case. Oxy-combustion carbon capture was the technology approach for the White Rose project studied up to Front End Engineering Design (FEED) as part of the UK CCS Commercialisation Programme. While an oxy-fired power plant is not yet demonstrated at utility scale, because there would be no incentive to do so without CO_2 capture, this case has the potential to exceed the performance of the coal post-combustion case. It is therefore important to ensure an up to date benchmark of this technology is included: this is Case 4 in this study. This technology application can be considered TRL-7 as it has been demonstrated at 30 to 50MWth scale.

Pre-combustion carbon capture from coal is achieved using an integrated gasification combined cycle (IGCC) flow scheme, which gasifies the coal and removes the CO_2 to create a hydrogen rich, carbon depleted syngas which is fired in a combined cycle gas turbine to emit an exhaust gas comprised almost entirely of nitrogen, water and oxygen. This flow scheme is often close to the overall performance of the other two coal cases with CO_2 capture, and has been demonstrated at scale without capture. It has been included as Case 5. The IGCC scheme without capture can be considered TRL-9, but the additional effort to capture the CO_2 is small compared to post-combustion. The IGCC scheme with capture can be considered TRL-7 as it has been demonstrated using reduced flow streams from existing IGCC plants without capture, such as the ISAB plant in Sicily.

A UK led CO_2 capture technology will need to compete with state-of-the-art technologies that are either already demonstrated at scale, or have been demonstrated at one tenth of utility scale over a number of years. They will also need to compete against other novel technologies which are already under development outside of the UK. For this reason, two of the most promising novel technologies, have been included as novel international benchmarks.

It is possible to apply oxy-combustion CO_2 capture to a natural gas turbine scheme, although it requires much more adaptation of the underlying technology than applying oxy-combustion to a coal fired boiler and is thus considered a novel technology. While some proponents of such a system have maintained both the gas cycle and the steam cycle, the most advanced technology uses a novel thermodynamic cycle, the Allam cycle, named for its inventor. The Allam cycle technology, developed by NET Power, currently has a 50 MWth scale plant under construction in

Texas. The system can be fired either with natural gas, or via gasification of a solid fuel. For this report a natural gas based case has been adopted because it is both more advanced in development and expected to be intrinsically more efficient than the solid fuel based variant. This case has been included as Case 6. This technology is currently considered as TRL-4 or 5 but once the 50MWth demonstration unit in Texas has been commissioned (scheduled to be completed in late 2017) it will reach TRL-7.

A second promising novel technology for low carbon power generation at utility scale involves the utilisation of fuel cells. While solid oxide fuel cells facilitate CO_2 capture by keeping the fuel stream and the oxidant streams separate, molten carbonate fuel cells (MCFCs) go one step further by also transferring CO_2 from the oxidant side of the cell to the fuel side. A CCGT flue gas stream contains sufficient oxygen to act as the fuel cells' oxidant stream. Therefore, combining MCFCs with a CCGT plant means that 90% of the CO_2 from the GT exhaust can be captured while generating additional electricity instead of increasing the plant's parasitic load. While the technology is capital intensive, its efficiency is expected to be high enough to make the scheme worthy of consideration. This technology is considered to currently be TRL-5, although pilot-testing at the James M. Barry Electric Generating Station in Alabama, announced in 2016, should increase this to TRL-6.

The three biomass cases (Cases 8, 9 and 10) cover three different routes for power generation from biomass, demonstrating maximum greenhouse gas reduction from power plants. Since biomass is counted as 'CO_2 neutral', these biomass cases produce net negative carbon emissions from a renewable and carbon-negative fuel source. Hence, addition of these three cases will be highly relevant for next generation UK-led technologies and are consistent with UK policy encouraging increased use of renewable energy sources. Assessing these cases as part of this study ensures consistency of methodology and consistency in results, which will assist BEIS in transference of results to other future studies.

Each of the three biomass benchmark cases is provided to the same level of definition as the coal and natural gas based state-of-the-art Benchmark Cases. However, due to the lack of large-scale reference plant and the challenges associated with the supply chain, logistics and cost of the biomass feed, the gross electrical output of these cases is smaller than the fossil fuel cases, in the range 500-600 MWe. Keeping within this smaller size range means that the TRL for the post-combustion and oxy-combustion biomass cases may be considered as TRL-7. However, the Bio-IGCC concept (Case 10) is at the pilot testing level of TRL-5 to 6.

Biomass to biogas via fermentation has been discussed as an option for an additional benchmark case. However, a fermentation case would pre-suppose that the gas network would be fully decarbonised in this manner, which is not current policy. Hence, fermentation options have not been included within this study.

In addition to the ten power benchmarks described above, a single unabated natural gas CCGT case was also included, referred to as Case 0, which is used to compare all the cases with CO_2 capture with a typical new build state-of-the-art UK power plant. This reference case allows the technical and economic impact of abating CO_2 emissions from the various benchmark plant flow schemes to be calculated versus an unabated CO_2 emissions reference point.

Although decarbonisation of power will be necessary to achieve the UK's climate change commitments, it will not be enough on its own. The final benchmark case presented in this report recognises that industrial sources of carbon dioxide emissions will also need to be tackled. Hydrogen is generated from natural gas as a feed to oil refining, petrochemical processes, ammonia and methanol production. The steam methane reforming process generates carbon dioxide which must be removed to produce a high-purity hydrogen product. This process therefore presents an obvious target for early decarbonisation. Case 11 presents a Steam Methane Reformer (SMR) benchmark case using state-of-the-art post-combustion technology. As with the other post-combustion benchmark cases, Shell Cansolv have provided data for the amine-based removal process[5].

5.2 Key Benchmark References and Development

In order to arrive at benchmark designs which are adequate for comparing against new technologies, it was not necessary to perform new, detailed design calculations or approach multiple vendors. Much of the information which is required has already been made publicly available by organisations such as the International Energy Agency Greenhouse Gas research and development programme (IEAGHG). These reports provide an excellent reference for the plant performance and capital and operating costs.

It is not possible to use the IEAGHG data directly, however, as it is important to ensure that performance and cost figures are as up to date as possible, they provide a very useful starting point for the design, saving much time and reducing the potential for errors.

The following table shows which references were used as a starting point for which benchmark case.

Table 5-1: Summary of Benchmark Case Key References

Benchmark Case	Key Reference
Case 1 – Natural Gas CCGT with post-combustion carbon capture	IEAGHG, "CO_2 Capture at Gas Fired Power Plants", 2012/8, July 2012, Scenario 3b
Case 2 – Natural Gas IRCC with pre-combustion carbon capture	IEAGHG, "CO_2 Capture at Gas Fired Power Plants", 2012/8, July 2012, Scenario 5
Case 3 – Coal SCPC with post-combustion carbon capture	IEAGHG, "CO_2 Capture at Coal Based Power and Hydrogen Plants", 2014/3, June 2014, Case 2
Case 4 – Coal SCPC with oxy-combustion carbon capture	IEAGHG, "CO_2 Capture at Coal Based Power and Hydrogen Plants", 2014/3, June 2014, Case 3
Case 5 – Coal IGCC with pre-combustion carbon capture	IEAGHG, "CO_2 Capture at Coal Based Power and Hydrogen Plants", 2014/3, June 2014, Case 4.1
Case 6 – Oxy-fired supercritical gas power generation with carbon capture	IEAGHG, "Oxy-Combustion Turbine Power Plants", 2015/05, August 2015, Case 2
Case 7 – Natural Gas CCGT with Molten Carbonate Fuel Cell carbon capture	No key reference
Case 8 – Biomass Fired CFB Boiler with post combustion carbon capture	IEAGHG, "Biomass CCS Study", 2009/09, November 2009, Case 3b
Case 9 – Biomass Fired CFB Boiler with oxy combustion carbon capture	No key reference
Case 10 – Biomass IGCC with pre-combustion carbon capture	No key reference
Case 11 – Natural Gas SMR with post-combustion carbon capture	No key reference [6]

The modelling methodology is slightly different for each of the benchmark cases, so for each technology, we have included a process description and an explanation of how the model was developed. In general, the approach taken has been to select the most relevant case from the key

[5] No assessment of other acid gas removal processes was undertaken as part of this study. Other amine solutions may be more cost effective for specific applications.
[6] The IEAGHG has recently issued report no. 2017-02, "Techno-Economic Evaluation of SMR Based H2 Plant with CCS", which may provide some useful data for comparison. However, it was not referenced by this study.

reference above, and to cross-check the results and costs against previous Wood and public domain references. The next step is to ensure that the technical basis of the design is aligned with the basis of this study, i.e. to ensure that the fuel feedstocks and ambient conditions are aligned and that recent technical advancements have been incorporated. For the two CCGT based cases (Case 1 and Case 7), this involved updating the gas turbine model from a 9F to a 9H class machine, which had a significant impact on the size of the downstream units and the overall efficiency of the scheme.

Once a heat and material balance has been developed for a scheme, a utility balance can be performed. These two deliverables are required to determine the efficiency and carbon balance of the scheme. The material balance is also used to provide the basis for a high level equipment list from which the capital cost estimate can be developed. At this level of study this is done partly on the basis of costing individual equipment items, and partly using vendor or public domain data for packaged units such as the boiler island in the SCPC case. The material balance is also combined with the capital cost estimate and an estimate of manpower requirements to determine the variable portion of the plant operating costs. Once the capital and operating costs have been determined it is possible to calculate illustrative overall project economics such as the Levelised Cost of Electricity or Levelised Cost of Hydrogen.

5.2.1 Biomass Benchmark Case Development

Conversions of existing post-combustion coal fired power plants to biomass fuels have typically been in stoker / fluid bed plants because the biomass fuel is generally cheap wood chips. However, for a pulverised coal plant like Drax in Yorkshire, wood chips cannot easily be used due to high moisture content and the difficulty of producing pulverised wood dust of consistent quality. For a pulverised coal plant retrofit, wood pellets can more easily be used, as these can be milled to the same powder consistency as pulverised coal, but wood pellets are more expensive.

For a new build state-of-the-art dedicated biomass fired power plant in the range of 250-300 MWe, subcritical circulating fluidised bed (CFB) technology is preferred and accepted in the industry, as opposed to a pulverised biomass boiler. Fuel flexibility and fuel switching, as well as co-firing capability give a significant economic advantage to the CFB boiler compared to a pulverised coal boiler. This allows the use of variable quality biomass, allowing a broader range of potential fuel suppliers. This reduces the risk associated with biomass supply chain and logistics, leading to operational stability.

Other advantages of CFB technology can be summarised as follows:

- Addition of limestone in the CFB boiler leads to low SO_2 emissions;
- Due to low combustion temperature and staged combustion, NOx emissions are reduced;
- The lower combustion temperature limits ash fouling and hence reduces the corrosion of heat transfer surfaces;
- The CFB boiler provides good mixing of air and fuel, leading to low carbon monoxide and unburnt hydrocarbons;
- Fuel preparation (e.g. milling or pulverising) is not needed.

Supercritical CFB boilers are proven technology for combustion of coal. In 2009, a 460 MWe coal fired supercritical CFB power generating unit was successfully commissioned in Lagisza, Poland. More coal-fired supercritical CFB power plants with unit sizes of 550 and 600 MWe are under construction in South Korea and China. However, the costs of the exotic materials required for supercritical systems are generally prohibitive at the smaller scales currently proven for biomass fired plant. Therefore, this study has focused on subcritical CFB boiler technology.

Pre-combustion of biomass with carbon capture is achieved using an integrated gasification combined cycle (IGCC) flow scheme. This process produces a hydrogen rich, carbon depleted syngas after removing the CO_2 from the raw gas post gasification. The hydrogen rich gas can be fired in a combined cycle gas turbine which is regarded as the most efficient thermal cycle for power generation. This system acts as a bridge between low quality fuel and highly efficient gas turbine power plant to maximise power generation and minimise emissions. Case 10 is the bio-IGCC case which uses Shell entrained flow gasification technology.

To develop the techno-economic basis for biomass pre-combustion, four different biomass-fired IGCC concepts have been analysed based on the available public domain information. Table 5-2 below compares the reference data available for the four concepts, along with pros and cons for each of the gasifier configurations.

Whilst no life-cycle analysis of the options has been performed, it can be summarised from the information stated in the comparison table that the Shell entrained flow gasifier for the Bio-IGCC option using 100% torrefied biomass pellets is the best option to be considered for this study. This option provides:

- Reference plant with successful trial run;
- Well proven Shell technology and expertise on entrained flow gasification for large scale IGCC plant of 300-600 MWe output;
- Net exportable power comparable with power suppliers for industrial areas;
- Maximum negative emissions on a large-scale Bio-IGCC plant;
- Torrefication of biomass to maximise the energy input to the gasifier in large scale;
- Data available on the torrefied feed composition and energy density.

It should be noted that there is a significant philosophical difference in the choice of the biomass options. The CFB-based cases use a well-proven industry-standard piece of equipment which can burn a very wide range of fuels. Biomass will vary, depending on the type of plant material, how it is processed and stored and the seasons when it is harvested. The CFB boiler will be able to accept a wide range of biomass feedstocks, including imported and indigenous biomass materials and waste materials. This should mean that the plant owner will have good confidence that they can get a competitive supply of fuel. The Shell gasifier, being an entrained flow type of gasifier designed to burn coal, represents a less flexible type of biomass option in terms of fuel. Torrefied wood chips are needed plus an additional fluxant. In practice, little fuel flexibility will be possible. At present, there is no commercial availability for torrefied biomass and so the cost of this material is speculative.

Table 5-2: Bio-IGCC Scheme Options Comparison

Shell Entrained Flow Gasifier – 10% biomass co firing (LHV basis) with coal	Shell Entrained Flow Gasifier – 100% biomass gasification using torrefied biomass pellets	Sumitomo Foster Wheeler CFB Gasifier – 100% biomass gasification	BGL Fixed Bed Slagging Gasifier – 75% waste gasification with coal
Overview: NUON operated 253 MW (net) integrated gasification combined cycle (IGCC) power plant in Buggenum using Shell entrained flow gasification technology with biomass input up to 30 wt% with coal on a continuous basis in early 2004. New biomass storage and feeding systems were put into operation in spring 2006. Since 2007, the plant has been operated with approximately 10% (energy) biomass mixed with coal.	**Overview**: In 2011, maximum 70% co-gasification on energy basis using torrefied pellets was achieved in NUON / Vattenfall Buggenum IGCC plant using Shell entrained flow gasification technology during a 24-hour trial. Public domain information is available providing data related to the torrefied feed composition and energy density. Shell Global Solutions provided information which confirms that 100% biomass gasification is technically achievable using torrefied wood pellets. Shell advise that the gasifier is running reliably on low-ash lignite coals, which are 'young' coals with properties not too dissimilar from those of torrefied biomass pellets.	**Overview**: Sumitomo Foster Wheeler has experience and technical expertise to supply 100% biomass fired Circulating Fluidised Bed (CFB). Pilot testing of 18MWth pressurised air blown CFB gasifier (15-20 bar) in Varnamo demonstration plant, Sweden. 950°C Gasifier temperature destroys most of the tar in the syngas. This gasifier and syngas is suitable for Bio-IGCC after cleaning the syngas suitable to be used in Gas Turbine. Sydkraft AB has built world's first Bio-IGCC pilot plant using wood as feed which is located in Varnamo, Sweden. The plant produced 6 MWe to grid and 9 MWth heat for district heating.	**Overview**: The improved fixed bed slagging version of the existing Lurgi Gasifier. First commercial plant operated at Schwarze Pumpe, Germany, from 2000 until 2007. The gasified fuel mix consisted of 25 % hard coal, 45 % RDF pellets, 10 % plastic waste, 10 % wood and 10 % tar sludge pellets. Gasifier capacity was 27 t/h. 75 % waste material was successfully converted into syngas. Syngas from the BGL gasifier contains a high amount of methane (~8-10%) compared to other type of gasifiers. This gas is suitable for downstream synthetic natural gas (SNG) production.
Configuration: Using two Shell entrained flow gasifiers with 10% biomass co-firing on thermal basis will produce net exportable power for the overall Bio-IGCC scheme which is comparable with the Coal-based IGCC and other benchmark cases. This will generate a small amount of negative emissions.	**Configuration**: Using two Shell entrained flow gasifiers feeding 100% torrefied pellets will produce net exportable power for the overall Bio-IGCC scheme which is comparable with the Coal-based IGCC and other benchmark cases. This scheme will generate maximum negative emissions on a large-scale Bio-IGCC plant.	**Configuration**: Much smaller than the Coal-based IGCC. The offering of air blown atmospheric CFB gasification is in the range of 150 MWth (fuel input) per line. Studies / budgetary proposals have been performed for the pressurized oxygen-steam blown units in the range of 250 MWth to 450 MWth (fuel input) per line. Hence, net power of 50-100 MWe.	**Configuration**: Can be used for Bio-IGCC application using up to 80% waste co-firing with coal; however, the scale will be much smaller than the Coal-based IGCC.
Advantages: • Well proven technology • Several successful trials performed using 30 wt% biomass (demolition wood) co-gasification with coal in Buggenum IGCC plant • During trial run, the gasifier operated at 25 bar and 1600°C resulting in a carbon conversion rate of over 99% • Performance and overall efficiency comparable with the coal-fired gasifier • Net exportable power comparable with the benchmark cases	**Advantages:** • Successful 24-hour trial performed using 70% co-gasification on energy basis using torrefied pellets in Buggenum IGCC plant • Public domain information on torrefied feed composition and energy density available • Maximises the negative emission on a large-scale Bio-IGCC plant • Net exportable power comparable with the benchmark cases	**Advantages:** • Well proven 100% biomass-fired Circulating Fluidised Bed (CFB) Gasifiers • Sumitomo Foster Wheeler's in-house licensed technology and expertise • Several air blown atmospheric CFB gasifier in operation • Pressurised CFB gasifier successfully operated • Highest possible negative emission from a bio-IGCC plant	**Advantages:** • 75% waste gasification trial run in SVZ Schwarze Pumpe Plant • Proven technology for coal • Maximising negative emission from a bio-IGCC plant

Shell Entrained Flow Gasifier – 10% biomass co firing (LHV basis) with coal	Shell Entrained Flow Gasifier – 100% biomass gasification using torrefied biomass pellets	Sumitomo Foster Wheeler CFB Gasifier – 100% biomass gasification	BGL Fixed Bed Slagging Gasifier – 75% waste gasification with coal
Disadvantages: • No detailed technical or performance data available in public domain • Nominal negative emission using 10% biomass on energy basis	**Disadvantages:** • Lower gasification temperature than coal-based case, with higher CO_2 and H_2O in syngas • Dust formation and challenges of storing torrified biomass. • Less steam generation from gasification island reducing overall efficiency • No detailed technical or performance data available in public domain	**Disadvantages:** • Small scale application compared to Coal-based IGCC • Net exportable power much lower than the benchmark cases	**Disadvantages:** • Mixed waste feed including plastics and tar • Small scale application compared to Coal-based IGCC • Net exportable power much lower compared to the benchmark cases • Lower overall carbon capture efficiency due to the high methane content in the syngas to gas turbine • No detailed technical or performance data available in public domain
Reference: Nuon Power Buggenum IGCC Plant: Shell Entrained Flow Gasifier https://www.netl.doe.gov/research/coal/energy-systems/gasification/gasifipedia/nuon	**Reference:** 'First experiences from large scale co- gasification tests with Refined biomass fuels', Nader Padban, Central European Biomass Conference International workshop: Torrefaction of biomass, 17th January 2014, Graz, Austria. 'Biomass torrefaction achieves increased co-gasification shares in entrained flow gasifiers', Carbo et al, IChemE Gasification Conference, Rotterdam, 2014	**Reference:** https://www.amecfw.com/documents/brochures-publications/brochures/pioneering-cfb-technology.pdf 'Biomass IGCC at Vernamo, Sweden – Past and Future', Stahl et al, GCEP Energy Workshop, CA, USA, 2004	**Reference:** 'Operational results from Gasification of waste material and biomass in fixed bed and circulating fluidised bed gasifiers', Schwarze Pumpe 'Further Developments and Commercial Progress of the BGL Gasification Technology', Hansjobst Hirschfelder, Gasification Technologies Conference, Washington, 2010

5.3 Common Basis Assumptions

The following table summarises the key assumptions which are common to all cases. Further details can be found in the Basis of Design document included as Attachment 1.

Table 5-3: Summary of Technical Basis

Parameter	Basis
Location	Coastal, North East England, Greenfield
Fuel Source – Natural Gas	Natural Gas, composition per IEAGHG 2012-15 + 3 ppmv H_2S
Fuel Source – Bituminous Coal	Bituminous Coal, composition per IEAGHG 2014
Fuel Source – Biomass	Wood chips of clean virgin biomass, composition as per IEAGHG 2009-9
Fuel Source – Torrefied Biomass	Torrefied wood pellet of clean virgin wood, specification developed from reference paper (1 & 2) and discussion with Shell Global Solutions
Ambient Conditions	Typical for NE England, sufficiently similar to costal Netherlands location as per IEAGHG 2012/8
Cooling Approach	Water Cooling, inlet / outlet temperatures per IEAGHG 2012-15
Power Configuration – Natural Gas	2 x GE9HA gas turbines + 2 x steam turbines or 2 x GE9FB syngas turbines + 2 x steam turbines (Case 2)
Power Configuration – Coal	Ultra-supercritical PC boiler, single reheat, 1 x steam turbine or 2 x GE9FB syngas turbines + 2 x steam turbines (Case 5)
Power Configuration - Biomass	Sub-critical CFB boiler, single reheat, 1 x steam turbine or 1 x GE9FB syngas turbine + 1 x steam turbine (Case 10)
CO_2 Capture & Compression	2 x trains
CO_2 Export Conditions	110 bar, 30°C, 50 ppm H_2O, 100 ppm O_2
CO_2 Transport and Storage	Costs assumed to be aligned with Leigh Fisher Report, “Electricity Generation Costs and Hurdle Rates, Lot 3: Non-Renewable Technologies”, August 2016
Cost Basis	GB£, Q1 2017

5.4 Capital Cost Estimating Methodology

Capital cost estimates were prepared for each case, using an approach aligned to AACE Class 4. This will typically give an estimate accuracy of around ±30%, although the actual accuracy for each case varies depending on the level of definition for that technology and the availability of public-domain cost information.

Through current and recent projects undertaken, Wood has access to market positions in respect of the global equipment and labour markets, including references for UK-based projects, aligned with the study basis. The cost estimates reflect our best assessment of the selected market.

The CAPEX estimates are largely based on a Wood ‘Indexed’ version of the Aspen Capital Cost Estimator (ACCE) computer programme. The ACCE programme includes ‘in-built’ P&ID (Piping & Instrumentation Diagram) models and is used to generate the base equipment & bulk material costs and direct labour manhours. The prime inputs to the cost estimate are the process definition, sized equipment lists and overall execution strategy. The ACCE output was checked against in-house costs and statistical data from a variety of sources.

All costs are provided in British pounds (GBP), fixed at the end of Q1 2017. Future price profiles for fuel and carbon price are provided on a year-by-year, 2017 real cost basis. Reference costs provided in other currencies have been converted to GBP at the annual average spot rate for 2016, as published by the Bank of England, as follows:

- £1 = $ 1.3542
- £1 = € 1.2233

All capital costs are assumed to be incurred during the four years prior to first start-up, with costs allocated in the following percentages:

- 2021 15%
- 2022 35%
- 2023 40%
- 2024 10%

Plant commissioning is assumed to occur at the start of the first year of operation and thus the availability of the plant is reduced during 2025. However, the model assumes that the EPC costs are due at the end of 2024.

5.4.1 EPC Contract Cost

The EPC contract costs provided in this report refer to a new-build integrated power plant with carbon dioxide capture and compression. It includes all facilities located at the site itself, including process, utilities, storage and administration facilities. It is assumed to be awarded as a single lump sum contract through a competitive tendering process.

It should be noted that the pre-combustion benchmark cases (Case 2, Case 5 and Case 10) and the two international benchmark cases (Case 6 and Case 7) represent concepts that have not been built at this scale to date. For novel process technology, it is standard practice to allow for higher equipment prices, longer construction / commissioning periods and significant contingency to be built into the EPC price. The approach taken for this study is to assume that none of the projects reflect the first-of-a-kind (FOAK) contract costs, but that the technology is considered to be commercial proven with a number of operating units. Hence, EPC costs presented in this report reflect an nth-of-a-kind (NOAK) approach. It is recognised that, within the context of a 2025 start-up date, these concepts may not have actually reached NOAK status, but the study aims to project the most likely competing technologies over the next generation, and so it is considered that NOAK provides the most appropriate comparison. The biomass-fired post-combustion and oxy-combustion cases (Case 8 and Case 9) have been considered at a smaller power export capacity than for the natural gas and coal-fired cases. This reflects the current proven scale for Circulating Fluidised Bed biomass boilers.

Direct Material costs have generally been built-up from equipment costs estimated via Wood's indexed version of ACCE. Factors for Piping, Control & Instrumentation and Electrical bulks are built into our version of ACCE, based on configurations for each type of equipment item. First-fill quantities for solvents, catalysts and other consumables have been estimated and costed based on past project experience.

Typical power project factors for Shipping & Freight, Third Party Inspection and Spare Parts have been applied to the Direct Materials cost.

Materials and Labour Contract costs were developed using in-house factors on the total Materials cost. The factors for these elements vary greatly from unit to unit, depending on the relative quantities of rotating equipment, static equipment, piping elements, control elements and analysers. These factors cover contracts for Civils, Steelwork & Buildings, Mechanical, Electrical & Instrumentation, and Scaffolding, Lagging & Rigging.

The EPC Contractor cost for services includes engineering design, project management, procurement, construction management and commissioning. It also includes for the EPC Contractor's recovery for corporate overheads, project contingency and profit. Naturally, the cost for services and profit margin may vary greatly in different locations and from year to year depending on the level of activity in the region and the degree of competitiveness between contractors. For this study, a flat rate of 17% Materials & Labour has been used to cover all of these elements.

5.4.2 Infrastructure Connection EPC Costs

Power Plant Cases

Offsite connections to the natural gas grid, future CO_2 export pipeline network and overhead power transmission lines are assumed to be provided in one or more EPC contracts, separate from the power plant contract. Clearly, the length of the interconnecting pipelines and power lines are highly dependent on the proposed location for the plant. For the purposes of this study, it has been assumed that all three connections are 10km in length, running to different locations along separate corridors, without major obstructions.

Costs have been developed based on typical material costs and installation factors for buried pipelines and overhead transmission lines. Both the import gas pipeline and the export CO_2 pipeline have been estimated at 14" nominal bore using L360 (X52) seamless pipe.

Table 5-4: Infrastructure Connections for Power Benchmark Cases

No.	Case	Gas Import	CO_2 Export	Power Export
0	Unabated CCGT	Yes	-	Yes
1	NG CCGT + post-combustion CCS	Yes	Yes	Yes
2	NG IRCC, pre-combustion CCS	Yes	Yes	Yes
3	Coal + post-combustion CCS	-	Yes	Yes
4	Coal oxy-combustion with CCS	-	Yes	Yes
5	Coal IGCC, pre-combustion CCS	-	Yes	Yes
6	NG oxy-fired supercritical turbine + CCS	Yes	Yes	Yes
7	NG CCGT + MCFC + CCS	Yes	Yes	Yes
8	Biomass + post-combustion CCS	-	Yes	Yes
9	Biomass oxy-combustion with CCS	-	Yes	Yes
10	Biomass IGCC, pre-combustion CCS	-	Yes	Yes

Hydrogen Plant Case

Like the power cases, new hydrogen units will require offsite connections for the natural gas feed, hydrogen export, CO_2 export and electrical power import, and these are assumed to be provided in one or more EPC contracts. Since hydrogen units are typically built in brownfield locations adjacent to the refinery / chemical plants that they supply, shorter interconnecting pipelines and power lines, at 1 km in length have been assumed. The exception is the CO_2 export pipeline, which has been kept at 10 km in length, due to the relative shortage of anchor projects with existing export pipelines.

Costs have been developed based on typical material costs and installation factors for buried pipelines and buried 11 kV power transmission lines. All pipeline costs have been specified assuming L360 (X52) seamless pipe. Natural gas feed and CO_2 export pipelines have been assumed at 8" nominal bore, whilst the hydrogen export line is estimated at 10".

Table 5-5: Infrastructure Connections for Hydrogen Cases

No.	Case	Gas Import	CO_2 Export	Hydrogen Export	Power Import
H	Unabated Hydrogen Reference Case	Yes	-	Yes	Yes
11	NG SMR + post-combustion CCS	Yes	Yes	Yes	Yes

5.4.3 Pre-Licensing, Technical & Design Costs

The EPC costs discussed in Section 5.4.1 above, reflect the contractor costs that occur from the point of the project developer making its Final Investment Decision (FID). However, most projects proceed through a series of stages from early conceptual design or feasibility studies, through pre-FEED, and Front End Engineering Design (FEED) that demonstrate the bankability of the project at increasing levels of detail.

A rule of thumb, is that each design stage requires an order of magnitude more effort than the previous phase, culminating in the cost for engineering services in the EPC phase. For this study, the developer's costs for this phase of the work has been estimated at 1% of the EPC contract value.

5.4.4 Regulatory, Licensing & Public Enquiry Costs

In order to construct the power station, an application would be made to the Secretary of State for Business, Energy and Industrial Strategy under the Planning Act 2008, as the proposed development would be considered a Nationally Significant Infrastructure Project. Consent would take the form of a Development Consent Order (DCO).

In determining whether to grant consent, the Secretary of State would consider the project's compatibility with national policy and in particular the Overarching National Policy Statement for Energy Infrastructure EN-1 and potentially the National Policy Statement for Fossil Fuel Electricity Generating Infrastructure EN-2.

The process of applying for consent would include the following activities:

- Screening and scoping of environmental information to be supplied;
- Preparation of a Preliminary Environmental Information Report;
- Preparation and agreement with the relevant local planning authority of a Statement of Community Consultation;
- One or two rounds of consultation with statutory stakeholders, landowners and members of the public;
- Preparation of the Environmental Impact Assessment;
- Preparation of design drawings and a number of plans as specified under regulation 5(2), including the application site, landownership, access and public rights of way, environmental designations, etc.;
- Preparation of documents specified by the regulations, including the draft DCO, the Consultation Report, the Book of Reference, the Funding Statement, and the Statement of Reasons;
- Development of other documents in support of the application, such as Habitats Regulation Assessment, Climate Change Resilience Report, Flood Risk Assessment, draft S106 agreement, and lists of other permit and licences to be obtained.

Once submitted and accepted the Secretary of State, the examining authority has six months to examine the application and will call a series of hearings with the applicant. In addition, up to three rounds of written questions can be posed by the examining authority requiring rapid responses.

Following close of the examination the examining authority has three months in which to make a recommendation to the Secretary of State, following which they have a further three months in which to make a decision.

The tasks set out above require the following technical expertise on behalf of the developer:

- Planning and consenting specialists to manage the process and guide the preparation of the application;
- Legal advice to draft the DCO and Book of Reference and to prepare agreements with landowners;
- Surveyors to identify landownerships and other land rights;
- Consultants to provide the strategy for engagement and to run the consultation activities;
- An environmental consultancy to prepare the preliminary EIA;
- Technical support to liaise between the project designers and the above.

Wood has significant experience of both site specific and linear DCO projects and we have provided planning and environmental consultancy support to some of the largest projects to pass through the system to date. Our approach to costing is based upon a review of nationally significant infrastructure projects, their known capital costs and the total consultancy fees required to take the scheme through to the grant of consent.

This review considered a range of projects, ranging from single site schemes to long distance overhead or underground connections. The long distance connection projects incur the greatest consultancy costs, due to the increased workload required to identify alternatives and significant additional work involved in the identification of landownership, land negotiations, consents, wayleaves and potential compulsory purchase orders. Similarly schemes with a smaller capital costs tend to incur higher percentage of consultancy costs given that much of consent work is similar irrespective of project scale.

In view of our experience and the nature of the project, total consultancy fees in the range of 1.5 – 2% of the power plant's capital cost and 4 – 4.5% of the project's infrastructure connection cost are appropriate. For all cases, we have assumed costs for Regulatory requirements and public enquiries of 2% of the onsite EPC cost, plus 4% of the infrastructure connection EPC cost.

5.4.5 Owner's Costs

This element covers the Project Developer's internal costs for developing the project concept through to start-up, including direct-hire personnel, taxes, insurances and costs for land purchase. Clearly, there may be huge variability in these elements, particularly in the land cost. The approach for this study is to assume a NOAK development (as noted in Section 5.4.1) on a greenfield site in a moderately industrial area, with a supportive local council and other stakeholders.

Owner's costs are assumed to be 7% of the EPC Contract Cost.

Technology licence fees are typically included within Owner's Costs for major project cost estimates. License fees have been included for this study, however, due to commercial sensitivity, they have been included within the EPC contract costs, rather than in Owner's Costs.

5.4.6 Start-up Costs

Start-up Costs and Working Capital are assumed to be expended as a single cost during the last few months before first start-up. Start-up Costs are related to having a trained operation and maintenance team on the facility during the commissioning and start-up process and the consumables that are used during the same period. The Working Capital Allowance assumes that a proportion of consumables must be held in stores to facilitate maintenance activities.

- Start-up Labour is set at between 3 – 4 month's Direct Labour, depending on the complexity of the process;
- Start-up Maintenance costs are set at 3 month's overall Maintenance cost;

- Start-up Consumables are set at 3 month's annualised Consumables cost;
- Start-up Fuel is set at the cost of 1 month's fuel, assuming that this is on a take-or-pay agreement.
- Working Capital Allowance is set at 1 month's fuel plus Consumables;

Working Capital Allowance is applied in the final year before start-up and increases year-on-year, based on the fuel price set. It is released in the final year of operation.

5.5 Operating Cost Estimating

Operating and Maintenance costs are generally allocated as fixed and variable costs.

Fixed costs are made up from the following categories:

- Direct labour
- Administrative and general overheads
- Maintenance

Variable costs assessed for this study are:

- Fuel
- CO_2 storage costs
- CO_2 emissions penalties or credits for avoided emissions
- Replacement catalysts, chemicals and equipment

5.5.1 Direct Labour

Specific data for power plant employees was not available to the study team, but the IEAGHG Report 2012/08 based personnel costs on an annual average salary of € 60,000 pa, which provides a base point for consideration. The UK Office for National Statistics (ONS) indicates that between February 2012 and February 2017, UK average weekly earnings rose from £465 to £509 (Ref. https://www.ons.gov.uk/employmentandlabourmarket/peopleinwork/earningsandworkinghours). Applying a similar rise in salary and converting to GBP at the annual average exchange rate for 2016 of £1 : €1.2233 leads to an equivalent 2017 salary of £53,700 pa.

A social burden equivalent to 30% has been added to the salary cost to account for social security payments, pension contributions, medical insurance and other in-company benefits.

The number of staff required to operate and maintain the plant has been assessed for each of the cases by the Asset Operations team at Wood and varied according to the complexity of the process and likely degree of interaction needed on a daily basis. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staff are taken to be in daily positions, working regular hours.

Table 5-6: Operations and Maintenance Staff Manning for Reference Cases

	Reference Case 0 Unabated CCGT	Reference Case H Unabated SMR	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Deputy Plant Manager	1		Daily Position
CO_2 Removal Area Manager			Not required for this case
Process Engineer			Not required for this case
Shift Supervisor	5		3-shift Position
Electrical Assistant	5		3-shift Position

	Reference Case 0 Unabated CCGT	Reference Case H Unabated SMR	Remarks
Control Room Operator	10	5	3-shift Position
Field Operator	10	5	3-shift Position
Sub-Total	**32**	**11**	
Maintenance Staff			
Mechanical Group	3	1	Daily Position
Instrument Group	3		Daily Position
Electrical Group	2		Daily Position
Sub-Total	**8**	**1**	
Laboratory Staff			
Superintendent	1		Daily Position
Analysts	3	1	Daily Position
Sub-Total	**4**	**1**	
Plant Total Staff	**44 ***	**13**	* See note below

* Note that the IEAGHG 2012/08 report estimated 50 permanent roles in Operations and Maintenance for an unabated CCGT plant.

5.5.2 Administrative, General Overheads, Insurance and Local Taxes

These costs include all other Company services not directly involved in the operation of the Complex, such as Management, Personnel Services and Clerical staff. These services vary widely from company to company and are also dependent on the type and complexity of the operation.

For this study, an allowance equivalent to 0.5% of the EPC Contract Cost for annual Administrative and General Overheads.

A further 2.0% of EPC Contract Cost is included to account for insurance and local taxes.

5.5.3 Maintenance

Maintenance costs have been assessed as a percentage of the EPC Contract Cost for different elements of the plant. Rotating machinery typically has a higher maintenance cost than static equipment and therefore a higher maintenance burden has been assumed for the power and feedstock handling sections of the plant, with annual costs assessed as follows:

Power Island Maintenance	2.5% of area EPC Contract Cost
Feedstock Handling	2.5% of area EPC Contract Cost
Steam Methane Reformer	3.0% of area EPC Contract Cost
CO_2 Capture & Compression	1.5% of area EPC Contract Cost
Other Units	1.5% of area EPC Contract Cost

5.5.4 Power System Connection and Use of Service Charges

Costs for connection of power export facilities into the National Grid and related Use of Service charges have been included at the unweighted average cost of £ 3280 / MW export capacity per annum, in line with the 2016 DECC report into Electricity Generation Costs and Hurdle Rates for Non-Renewable Technologies.

5.5.5 Fuel & Carbon Emissions

Fuel and Carbon Emission price sets have been provided by BEIS over the project life-span[7]. Prices are given in 2017 real terms. Coal costs have been converted to GBP using the annual average spot exchange rate for 2016, published by the Bank of England, i.e. £1 = $1.3542. Results presented in this report are generally aligned with the 'Central' price projection, although some sensitivity cases refer to 'High' or 'Low' fuel price sets.

Table 5-7: Fossil Fuel and Carbon Price Sets

	Natural Gas Price (pence / therm)			Coal Price (USD / tonne)			Carbon Price (GBP / tonne CO_2)
	Low	Central	High	Low	Central	High	Central
2017	22.7	31.5	43.4	34.5	39.4	50.2	21.6
2018	22.7	31.5	45.3	34.5	39.4	55.2	21.3
2019	22.7	31.5	47.3	34.5	39.4	61.1	21.1
2020	22.7	31.5	49.3	34.5	39.4	66.0	21.5
2021	23.6	34.5	51.2	36.5	43.4	70.9	21.7
2022	25.6	37.4	54.2	39.4	47.3	76.8	21.9
2023	26.6	40.4	56.2	42.4	51.2	81.8	22.1
2024	28.6	43.4	58.1	45.3	55.2	87.7	22.3
2025	29.6	46.3	60.1	47.3	59.1	92.6	22.6
2026	31.5	49.3	62.1	50.2	63.1	97.5	23.2
2027	32.5	52.2	64.0	53.2	67.0	103.5	30.8
2028	34.5	55.2	67.0	56.2	70.9	108.4	32.6
2029	35.5	58.1	69.0	58.1	74.9	114.3	33.2
2030	37.4	61.1	70.9	61.1	78.8	119.2	36.1
2031	37.4	61.1	70.9	61.1	78.8	119.2	47.5
2032	37.4	61.1	70.9	61.1	78.8	119.2	59.0
2033	37.4	61.1	70.9	61.1	78.8	119.2	70.5
2034	37.4	61.1	70.9	61.1	78.8	119.2	81.9
2035	37.4	61.1	70.9	61.1	78.8	119.2	93.4
2036	37.4	61.1	70.9	61.1	78.8	119.2	104.9
2037	37.4	61.1	70.9	61.1	78.8	119.2	116.3
2038	37.4	61.1	70.9	61.1	78.8	119.2	127.8
2039	37.4	61.1	70.9	61.1	78.8	119.2	139.2
2040	37.4	61.1	70.9	61.1	78.8	119.2	150.7
2041	37.0	60.9	70.8	61.1	78.6	119.3	158.0
2042	37.0	60.9	70.8	61.1	78.6	119.3	165.2
2043	37.0	60.9	70.8	61.1	78.6	119.3	172.5
2044	37.0	60.9	70.8	61.1	78.6	119.3	179.7
2045	37.0	60.9	70.8	61.1	78.6	119.3	187.0
2046	37.0	60.9	70.8	61.1	78.6	119.3	194.2
2047	37.0	60.9	70.8	61.1	78.6	119.3	201.5
2048	37.0	60.9	70.8	61.1	78.6	119.3	208.8
2049	37.0	60.9	70.8	61.1	78.6	119.3	216.0
2050	37.0	60.9	70.8	61.1	78.6	119.3	223.3

[7] Price sets used in this analysis were those published before the start of the assessment (15th March 2017). The latest price sets can be found here: https://www.gov.uk/government/publications/valuation-of-energy-use-and-greenhouse-gas-emissions-for-appraisal

Torrefied biomass cost has been taken from an ECN paper (Carbo et al, 'Torrefied biomass pellets key to establish dense-phase flow feed to entrained flow gasifiers', 8th International Freiberg Conference on IGCC & XTL, Germany, 2016). The paper reports the cost as €29 / MWh which is equivalent to £23.7 / MWh. Torrgas, a Bioenergy Product supplier based in the Netherlands, has provided some useful information about torrefied biomass cost as ~$9 / GJ which is equivalent to £24 / MWh. This cost is in line with the ECN data.

Torrgas also suggested that the cost difference between woody biomass chips and torrefied biomass should be ~ 20%. In the absence of any other cost data, this information has been used to determine the wood chip biomass cost. Using the ECN cost of £23.7 / MWh for torrefied biomass and applying 20% reduction for the cost of wood chips, the cost has been calculated as £19.0 / MWh in 2017 values. A constant real price has been used, assuming that the prices for biomass and torrefied biomass are pegged to inflation. If the market for renewable power production using biomass becomes established, economies of scale would normally result in real-terms reductions in price. However, no credit for future cost reductions has been taken in this report.

Results presented in this report are generally aligned with the 'Central' price projection, although some sensitivity cases refer to 'High' or 'Low' fuel price sets. The 'High' and 'Low' fuel price sets are based on 150% and 80% of the central case, respectively.

5.5.6 CO_2 Transportation and Storage Costs

Since there are no commercial carbon dioxide storage facilities within the UK, storage costs are highly uncertain. The 2016 DECC report into Electricity Generation Costs and Hurdle Rates for Non-Renewable Technologies provides a median cost of £19 / tCO_2 for transportation and storage. The same figure has been used in this report for the main analysis. The same report provides high and low sensitivity costs of £31 / tCO_2 and £8 / tCO_2, respectively.

5.5.7 Catalysts, Chemicals and Equipment Replacement

Costs for upgrading or replacement of solvents, catalysts and equipment with a service life of less than 25 years have been annualised and included in the analysis as variable costs.

This category also covers other variable costs associated with make-up water, refrigerant, waste water disposal and replacement of consumables such as filter elements.

5.5.8 Income from Electricity Sales

The primary income from the facility will be the sale of electricity. Rather than assign a sales price for electricity and then calculate the Net Present Value for each project, this analysis determines the average price of electricity that would be needed for the project to achieve a Net Present Value of zero across the life of the plant: this is the Levelised Cost of Electricity (LCOE).

Likewise, for the hydrogen plant cases, all other variables are fixed and the Levelised Cost of Hydrogen (LCOH) required to achieve a Net Present Value of zero is determined.

5.6 Economic Modelling Factors

The levelised cost of electricity (LCOE) for each of the power production cases has been calculated using a simple, spreadsheet-based economic model. All cases assume full equity-financing by the project development company, hence interest payments, capital loan phasing and contingency release are all excluded. Since the LCOE concept calculates an electricity price that assumes the plant breaks even over its entire life, the model indicates that the company never earns enough to pay corporation tax, which also simplifies the model.

The levelised cost of hydrogen (LCOH) production for the hydrogen benchmark case is calculated in much the same way as the levelised cost of electricity. However, some of the cases require import of electrical power. The LCOE from our Benchmark 1 power case (CCGT with post-combustion carbon capture) has been assumed as the cost for purchasing power from an external supplier. CO_2 emissions per MW of power export for Benchmark Case 1 have also been included within the specific emissions for the hydrogen cases including power import.

5.6.1 Price Escalation and Discount Factors

Fuel price sets have been provided by BEIS, as discussed above in Section 5.5.5. Other costs are generally provided on a Q1 2017 real basis, with zero price escalation to future years (i.e. other costs are assumed to rise in line with inflation).

Over the seventeen years since the turn of the century, the European Power Capital Cost Index provided by IHS, has shown an average rise of 3.5% per annum, although the rate was significantly higher in the period to 2008 and has averaged zero growth since the recession. Over the same period, the UK Retail Price Index (RPI) has averaged 2.8% per annum. Within the accuracy of this analysis, capital costs are rising in line with inflation and hence the assumption of zero real terms escalation has been applied to capital costs.

Future cost and income have mostly been discounted back to 2017 values using a discount factor of 8.9%, which is the hurdle rate assumed by BEIS (at the time of the study) for offshore wind. This is an illustrative figure that reflects a higher level of risk than an unabated CCGT (Case 0), but is held constant across Cases 1 – 10 in order to isolate cost differences that occur due to technological differences rather than choice of financing model. As a result, the LCOEs should not be interpreted as a best estimate of the price needed for a typical project to deploy in 2025, but as benchmarks for which to consider the relative costs of the cases. A similar approach has been taken for the LCOH of Case 11. Finally, a discount factor of 7.8% has been applied to the Unabated Reference Cases (Case 0 and H), since unabated CCGT and hydrogen plant designs reflects a low level of complexity.

5.6.2 Plant Availability

Plant availability represents the proportion of an average year that the plant is available to export power / hydrogen at its rated capacity. This value takes account of both scheduled maintenance and downtime due to equipment failure / emergency repairs.

For a benchmarking exercise, typical average availabilities for existing plant are appropriate. We have assumed a Plant Availability of 93% for the unabated CCGT benchmark to provide consistency with the IEAGHG 2012/08 report.

The addition of the more complex plant required to capture, dehydrate and compress the carbon dioxide for export results in more points of failure. Hence, a Plant Availability of 90% has been used for the post-combustion and oxy-combustion cases, in line with the IEAGHG 2012/08, 2014/03 and 2015/05 reports. The integrated gasification (or reforming) benchmarks are more difficult to operate and so a lower availability of 85% has been used for these cases.

For consistency, the Reference unabated hydrogen plant has been assumed to have a Plant Availability of 93%, whilst the different technologies for an integrated hydrogen plant with carbon capture are assumed to have Plant Availability of 90%.

The economic model assumes that Plant Availability is reduced by 40% during the first year of operation, due to commissioning activities and more frequent unplanned shutdowns.

The Plant Availability does not reflect whether a market is available for the operating company to sell electricity / hydrogen to. Our economic model has been set-up with an additional factor for Plant Load Factor, which can be used to model reduced income in cases where a plant is intended to (or is forced to) operate during a more restricted period of time. The Plant Load Factor has been set to 100% for all of the analysis covered in this report.

5.6.3 Power Degradation

As gas turbines and other power plant are used, the turbine blades are gradually eroded and engrained with dirt, reducing the efficiency of the system. Gas turbines are subject to frequent blade-cleaning campaigns, which maintain operating efficiency to a degree, and major overhauls of gas turbines are performed every five to six years, but some permanent loss in performance is always experienced.

Given that different equipment suppliers may provide different degradation profiles and will claim different levels of performance recovery following overhaul, it was decided that this study should use BEIS standard profiles for the expected performance degradation of gas turbines throughout the project lifecycle. The table below records two degradation profiles: one for an H-class gas turbine within a combined cycle power configuration that has been applied to cases using gas turbines, and one for an oxy-coal process, which has been applied to the other cases, which use steam turbines only. These profiles are consistent with other work performed for BEIS / DECC in recent years, such as the Coal and Gas Assumptions Report, issued in March 2014 by Parsons Brinckerhoff. No attempt has been made to confirm or update these profiles as part of this study.

Within the results for each Benchmark case, two overall process efficiencies are generally provided: one "As New" efficiency, assuming zero degradation in performance, and one "Average" efficiency, which uses the lifetime average efficiency across the 25-year project life, in accordance with the degradation profiles below.

Table 5-8: Power System Degradation Profiles

	CCGT H-Class	Oxy-Coal
Year 1	100.0%	100.0%
Year 2	98.2%	100.0%
Year 3	96.5%	99.5%
Year 4	94.8%	99.5%
Year 5	93.1%	98.5%
Year 6	99.2%	100.0%
Year 7	97.5%	100.0%
Year 8	95.7%	99.5%
Year 9	94.1%	99.5%
Year 10	92.3%	98.5%
Year 11	90.6%	100.0%
Year 12	98.4%	100.0%
Year 13	96.7%	99.5%
Year 14	95.0%	99.5%
Year 15	93.3%	98.5%
Year 16	91.6%	100.0%
Year 17	89.9%	100.0%
Year 18	97.6%	99.5%
Year 19	95.9%	99.5%
Year 20	94.2%	98.5%
Year 21	92.6%	100.0%
Year 22	90.9%	100.0%
Year 23	89.2%	99.5%
Year 24	96.8%	99.5%
Year 25	95.1%	98.5%
Lifetime Average	**94.8%**	**99.5%**

6 Case 1 – Natural Gas CCGT with Post-Combustion Carbon Capture

6.1 Overview

This case consists of a natural gas fired combined cycle power plant based upon 2 GE Frame 9HA.01 gas turbines each with a dedicated heat recovery steam generator (HRSG) and steam turbine in a 2 x 2 configuration. The flue gas from both HRSGs is routed to a single train Shell Cansolv proprietary post combustion CO_2 capture unit, where it is boosted in pressure using a flue gas fan, then cooled in a gas/gas heat exchanger before entering a direct contact cooler. CO_2 is captured from the cooled flue gas using an amine based solvent in an absorption column and is released from the solvent in the stripper. The captured CO_2 is then compressed in 4 stages, dehydrated and then compressed in a further stage to the required export pressure of 110 bar (abs).

Table 6-1 describes the process units with trains which are also shown in Figure 6-1.

Table 6-1: CCGT Process Units with Trains

Unit Description	Trains
Gas Turbine & Generator Package	2 x 50%
Heat Recovery Steam Generation	2 x 50%
Steam Turbine & Generator Package	2 x 50%
CO_2 Capture Unit	1 x 100%
CO_2 Compression & Dehydration	2 x 50%
Offsite & Utilities	

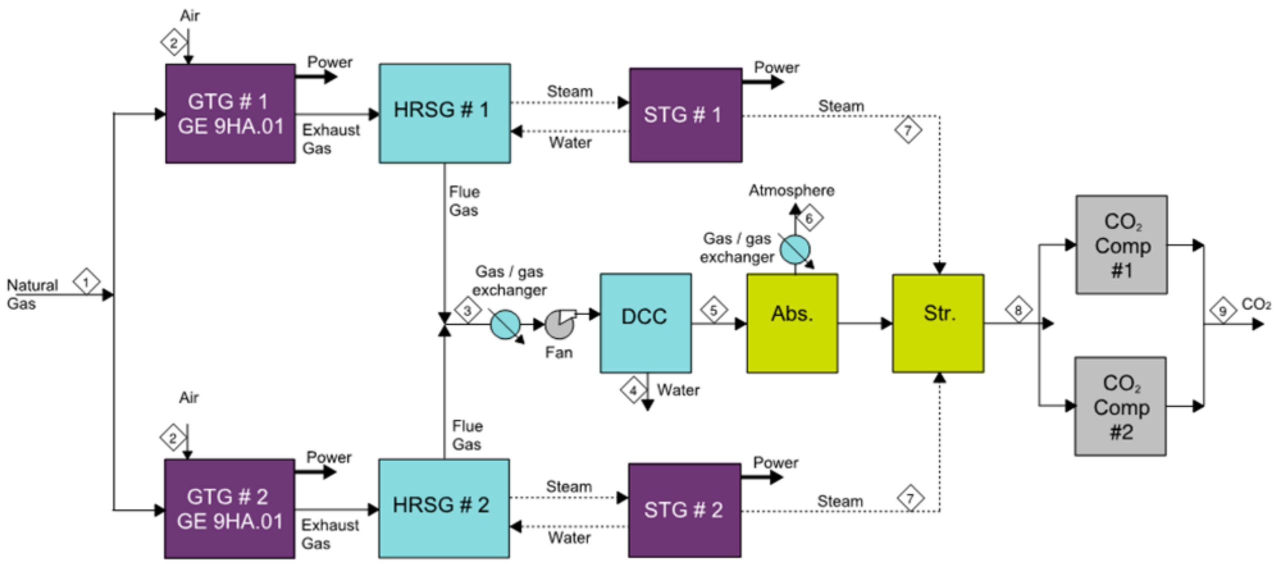

Figure 6-1: Case 1 Block Flow Diagram

6.2 Model Development

Usually gas turbine performance would be determined within Gatecycle using data from the extensive library, however, the library did not contain the latest and best performing gas turbine selected for this study, the GE 9HA.01. Performance and cost data for a combined cycle plant based upon GE 9HA.01 gas turbines was taken from Gas Turbine World 2014-2015 Handbook and up-rated, as recommended in the handbook, for the cooler than ISO site ambient conditions.

Using Gatecycle, it is possible to model both the gas and steam cycles in the power island allowing the impact of LP steam extraction for the post-combustion solvent reboiler heat load to be modelled directly.

The Gatecycle model takes natural gas at the conditions specified in the basis of design, models its combustion and the resulting flue gas as it progresses through the HRSG. It is not suitable for modelling any other sections of the plant however, and the flue gas at the point where it would enter the stack provides the interface between the power island model and any downstream process model.

Cansolv Technologies Inc. (Shell Cansolv) was contracted to provide a Process Design Package for absorption and regeneration system using its proprietary CANSOLV Absorbent DC-201, which is generally recognised as an industry leader in post-combustion CO_2 capture. All modelling of the proprietary amine absorption and stripping systems was conducted by Shell Cansolv. Detailed results cannot be published in this report due to confidentiality restrictions and some details have been redacted from the attached deliverables, but the overall results presented in this section reflect the latest performance results achieved by Shell Cansolv.

The CO_2 compression system can be modelled with accuracy, including high level key heat integration, and this has been undertaken in Hysys for this study.

6.3 Process Description

6.3.1 Power Island

Natural gas is received from the grid and metered before being routed to the power island where it is preheated and fed to two parallel gas turbines. The compressors of the gas turbines draw air from the atmosphere and compress it before mixing it with the natural gas fuel in the combustion chamber. The hot combusted gas is then expanded through the turbine which turns a generator (and the compressor) to generate electrical power. The exhaust from each turbine is directed into a heat recovery steam generator where the residual heat energy contained in the flue gas is recovered, as much as possible, by generating steam.

Large natural gas combined cycle power plants are sufficiently large to make it worthwhile to use three pressure levels of steam as well as reheating of the MP steam in order to maximise the heat recovered (on smaller plants fewer steam levels are usually justified).

6.3.2 Proprietary Solvent CO_2 Capture

An outline of the Shell Cansolv process as applied to a natural gas fired combined cycle plant is shown in Figure 6-2 below:

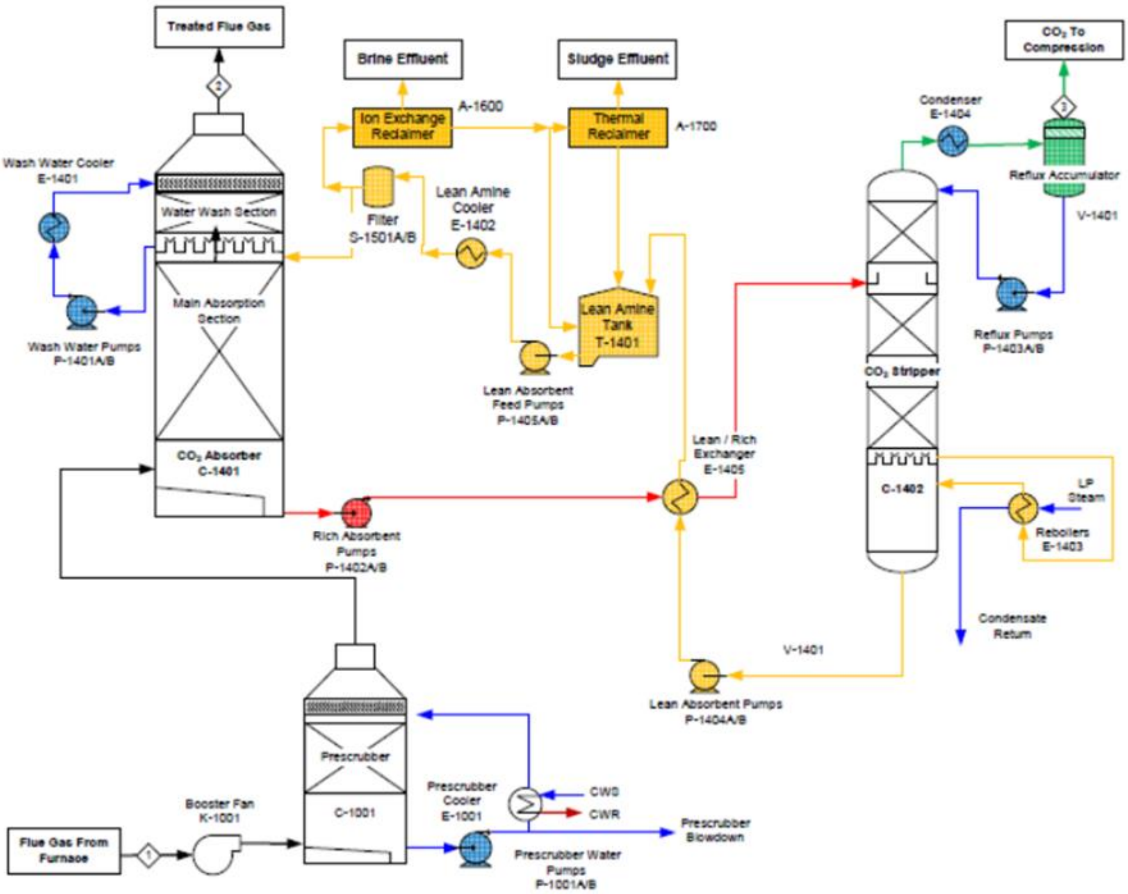

Figure 6-2: Cansolv CCGT Process Configuration
(Image courtesy of Cansolv Technologies Inc.)

Flue gas from the two HRSG's is combined into a single duct, cooled in a gas/gas heat exchanger and boosted in pressure to overcome the pressure drop through the downstream equipment. The boosted flue gas then enters a single direct contact cooler column in order to sub-cool the flue gas to 35°C to maximise performance of the CO_2 absorbent, minimising the required circulation rate and thus the energy consumption and capital cost of the unit.

The cooled flue gas is ducted to the bottom of an absorption column where it is contacted with the proprietary solvent. For a plant of this scale and flue gas type, typical absorber dimensions range from 22m to 32m in square cross section. CO_2 absorption from the flue gas occurs by counter-current contact with CANSOLV Absorbent DC-201 in the CO_2 Absorber which is a vertical multi-level packed-bed tower. CO_2 is absorbed into the solvent by chemical reaction leaving a flue gas depleted in CO_2 at the top of the column. The absorption reaction is exothermic, however, the low concentration of CO_2 in the gas turbine exhaust gas results in only a moderate temperature increase and thus no external cooling is required in this case.

The treated flue gas passes through a water wash section in order to prevent emissions of solvent and any solvent degradation products such as nitrosamines. The treated flue gas is then warmed in the gas/gas heat exchanger and routed to a stack for discharge to the atmosphere.

The CO_2 rich absorbent is collected in the bottom sump of the CO_2 Absorber and is pumped by the CO_2 Rich Absorbent Pumps and heated in the CO_2 Lean/Rich Exchangers to recover heat from the hot lean absorbent discharged from the CO_2 Regenerator. The rich absorbent is piped to the top of the CO_2 Stripper for absorbent regeneration and CO_2 recovery. The rich absorbent enters the column under the CO_2 top packing section and flows onto a gallery tray that allows for disengagement of any vapour from the rich absorbent before it flows down to the two stripping packing sections under the gallery tray. The rich absorbent is depleted of CO_2 by water vapour generated in the Regenerator Reboilers which flows in an upward direction counter-current to the rich absorbent.

Lean absorbent flowing from the bottom packing section of the CO_2 Regenerator is collected on a chimney tray and gravity fed to the Regenerator Reboilers. Water vapour and lean amine flow by thermosyphon effect from the reboilers back to the CO_2 Regenerator sump, underneath the chimney tray. Water vapour flows upwards through the chimney tray to strip the CO_2 while the lean absorbent collects in the bottom sump.

Water vapour in the regenerator, carrying the stripped CO_2, flows up the regenerator column into the top packing section, where a portion of the vapour is condensed by recycled reflux to enrich the overhead CO_2 gas stream. The regenerator overhead gas is partially condensed in the Regenerator Condensers. The partially condensed two phase mixture gravity flows to the CO_2 Reflux Accumulator where the two phases separate. The reflux water is collected and returned via the Reflux Pumps to the regenerator rectification section. The CO_2 product gas is piped to the CO_2 Compression System. The pressure of the Regenerator can either be controlled by a product CO_2 discharge control valve or by the inlet guide vanes of the downstream CO_2 Compressors.

The flow of steam to the reboiler is proportional to the rich absorbent flow sent to the CO_2 Regenerator. The set-point of the low pressure steam flow controller feeding the Regenerator Reboilers is also dependent on the regenerator top temperature controller. The steam to absorbent flow ratio set-point is adjusted by this temperature controller. The temperature at the top of the column is set to maintain the required vapour traffic and stripping efficiency. The steam flow rate is controlled by modulating a steam flow control valve.

All amine based systems require some form of solvent maintenance system as over time the absorbent in the CO_2 Capture System accumulates Heat Stable Salts (HSS), as well as non-ionic amine degradation products, that must be removed from the absorbent. This is achieved through thermal reclamation. An ion exchange package is included for bulk HSS removal upstream of a thermal reclaimer.

The ion exchange package is designed to remove Heat Stable Salts (HSS) from the Cansolv DC Absorbent. These salts are continuously formed within the absorbent, primarily due to residual amounts of NO_2 and SO_2 contained in the flue gas. Once absorbed, NO_2 forms nitric and nitrous acid while SO_2 forms sulphurous acid which oxidizes to sulphuric acid. These acids, and some organic acids formed by the oxidative degradation of the amine, neutralize a portion of the amine, which is then inactivated for further CO_2 absorption.

The purpose of the Thermal Reclaimer Unit is to remove the non-ionic degradation products as well as HSS from the active absorbent. The thermal reclaimer unit distils the absorbent under vacuum conditions to separate the water and amine, leaving the non-ionic degradation products in the bottom. A slipstream is taken from the treated CO_2 lean absorbent exiting the ion exchange package and fed to the Thermal Reclaimer Unit. This stream will essentially consist of water, amine, degradation products, residual CO_2 and small amounts of sodium nitrate and sodium sulphate. The design flow rate of CO_2 lean absorbent sent to the thermal reclaimer is based on the calculated amine degradation rate. To maintain the degradation products below design concentration, the thermal reclaimer must process a specific flowrate of CO_2 lean absorbent. The reclaimed absorbent is sent to the Lean Absorbent Tank. The separated degradation products are stored in a storage tank, where they are diluted and cooled with process water. Diluted residues are periodically disposed of offsite, typically via incineration.

6.3.3 CO_2 Compression and Dehydration

The CO_2 is compressed to 60 barg in 4 stages, each with intercooling and water knock-out. This recovers the vast majority of the water content, but is not sufficient for most pipeline specifications. Numerous studies have compared drying with tri-ethylene glycol (TEG) versus use of molecular sieve adsorption which conclude that there is little to choose between the two methods.

For the purposes of this study we have assumed a TEG dehydration unit is selected, since that was the selection made in the reference IEAGHG 2012 report. In the natural gas fired case the final stage of CO_2 pressurisation to 110 bar (abs) is achieved using a compressor, while in the coal fired post combustion case one further stage of compression followed by a condenser then a stage of pumping is used.

6.4 Technical Performance Evaluation

Table 6-2: Technical Performance for Case 1

	Units	Reference Case (Unabated CCGT)	Natural Gas Post-combustion (CCGT) with CCS
Total Gross Installed Capacity	MWe	1229.4	1144.3
Gas Turbine (s)	MWe	823.5	823.5
Steam Turbine	MWe	405.9	320.8
Others	MWe	0	0
Total Auxiliary Loads	MWe	20.9	79.7
Feedstock Handling	MWe	0	0
Power Island	MWe	14.7	14.7
Air Separation Unit	MWe	0	0
CO_2 Capture & Comp.	MWe	0	52.0
Utilities	MWe	6.2	13.0
Net Power Export	MWe	1208.5	1064.6
Fuel Flow Rate	kg/h	150,296	150,296
Fuel Flow Rate (LHV)	MWth	1940.2	1940.2
Net Efficiency (LHV) - As New	%	62.3	54.9
Net Efficiency (LHV) - Average	%	59.0	52.0
Total Carbon in Feeds	kg/h	108,640	108,640
Total Carbon Captured	kg/h	0	98,661
Total CO_2 Captured	kg/h	0	361,539
Total CO_2 Emissions	kg/h	398,105	36,566
CO_2 Capture Rate	%	0	90.8
Carbon Footprint	kg CO_2/MWh	329.4	34.3

The plant performance of the GE 9HA.01 based CCGT power plant with state of the art Shell Cansolv post-combustion carbon capture is summarised in the above table. The unabated CCGT case, for the same power island configuration is also listed in the table for the purposes of comparison. The Cansolv case captures 90% of the CO_2, while suffering a 7.4% point net efficiency loss.

The following points can be highlighted as basic difference between the two cases:

- The Reference case uses one of the largest and most efficient natural gas fired gas turbines GE 9HA.01 with large power output of > 400 MWe per turbine.
- The CCGT Cansolv case uses the same power island configuration and thus benefits from a very high efficiency underlying power plant before carbon capture is applied.

- The addition of carbon capture results in additional parasitic electrical load of 58.8 MWe, as well as a significant parasitic steam load required for regeneration of the proprietary solvent which can be seen in the reduced electrical generation from the steam turbine in the table above of 85.1 MWe.
- Overall, the Cansolv process and CO_2 compression and dehydration result in a 7.4% point decrease in the LHV efficiency of the power plant.
- The carbon footprint for the CCGT Cansolv case is ~ 10 times lower than the Reference unabated case as this case captures 90% of the process CO_2 for transportation and storage.

Note the carbon footprint stated above covers the impact of the power plant, CO_2 capture, treatment and compression facilities only. Upstream emissions related to natural gas distribution and downstream emissions related to the carbon dioxide storage are not included within this study.

6.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staffs are taken to be in daily positions, working regular hours.

Table 6-3: Operations and Maintenance Staff Manning for Case 1

	Reference Case Unabated CCGT	Natural Gas Post-combustion (CCGT) with CCS	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Deputy Plant Manager	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	Daily Position
Process Engineer	NA	1	Daily Position
Shift Supervisor	5	10	3-shift Position
Electrical Assistant	5	5	3-shift Position
Control Room Operator	10	15	3-shift Position
Field Operator	10	20	3-shift Position
Sub-Total	32	54	
Maintenance Staff			
Mechanical Group	3	3	Daily Position
Instrument Group	3	3	Daily Position
Electrical Group	2	2	Daily Position
Sub-Total	8	8	
Laboratory Staff			
Superintendent	1	1	Daily Position
Analysts	3	4	Daily Position
Sub-Total	4	5	
Plant Total Staff	44	67	* See note below

* Note that the IEAGHG 2012/08 report estimated 79 permanent roles in Operations and Maintenance

Table 6-4: Economic Performance Comparison for Case 1

	Units	Reference Case (Unabated CCGT)	Natural Gas Post-combustion (CCGT) with CCS
Total Project Cost	£M	672.2	968.2
Specific Total Project Cost	£/kW	556	909
Pre-Development Costs			
Pre-Licensing & Design	£M	5.8	8.5
Regulatory & Public Enquiry	£M	12.9	18.4
EPC Contract Cost	£M	583.6	845.2
Other Costs			
Infrastructure Connections		29.0	37.0
Owner's Costs		40.9	59.2
Overall CAPEX Impact (vs Ref Case)	£M	-	296.0
Overall CAPEX Impact (vs Ref Case)	%	-	44
Total Fixed OPEX	£M pa	36.2	47.5
Total Variable OPEX (excl. Fuel & Carbon)	£M pa	0.2	62.3
Average Fuel Cost	£M pa	315	305
Average CO_2 Emission Cost	£M pa	369	32.8
Total Start-up Cost (excl. Fuel)	£M	4.4	7.7
Discount Rate	% / year	7.8	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	69.9
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	-14.5
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	67.1
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	73.1

The economic performance of the CCGT with Cansolv post-combustion carbon capture is summarised in Table 6-4 along with unabated CCGT case for the purposes of comparison. The capital cost estimate for both the Reference Case and Case 1 are assessed to have an accuracy of ± 30%.

The absolute total project cost for this case is ~44% higher than the Reference unabated case while producing 12% less net power output, making it 63% higher on a specific cost basis:

- The Cansolv system adds to the total project capital cost, with the large low pressure absorber tower being a significant individual cost item. However, combining one wall of the direct contact cooler with the absorber and moving from two trains to one results in significant cost savings versus previous designs. The CO_2 compressor is also a significant individual item cost, but the increase in the total plant cost is small compared to other cases.
- The operating costs (excluding fuel and carbon price) are more than double the operating costs of the reference plant, which demonstrates the cost of running the more complex plant and the cost of CO_2 transportation and storage.

- Despite the capital and operating costs (excluding fuel) being higher for this case than the unabated case, the Levelised Cost of Electricity (LCOE) is lower, at £69.9 / MWh compared with £74.2 / MWh in the Reference case. This is due to the very high efficiency of this case and its moderate increase in capital cost combined with its low carbon emission per unit of net power produced.

Sensitivities on fuel cost and carbon price are provided in Section 17.4.

The chart below shows the balance of factors contributing to the overall LCOE. It can be seen that the fuel cost is the major factor but that the capital investment is also significant, with the operating cost and CO_2 transportation and storage somewhat smaller.

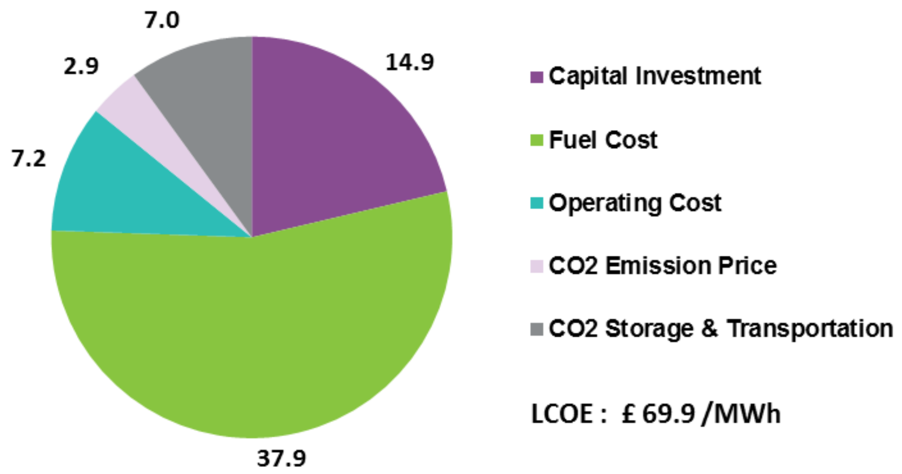

Figure 6-3: LCOE (£/MWh) Contribution for Case 1

By contrast, the figure below shows the LCOE breakdown for the Reference case. It can be seen that the cost penalty for emitting CO_2 is almost as significant in the calculation of LCOE as is the fuel cost.

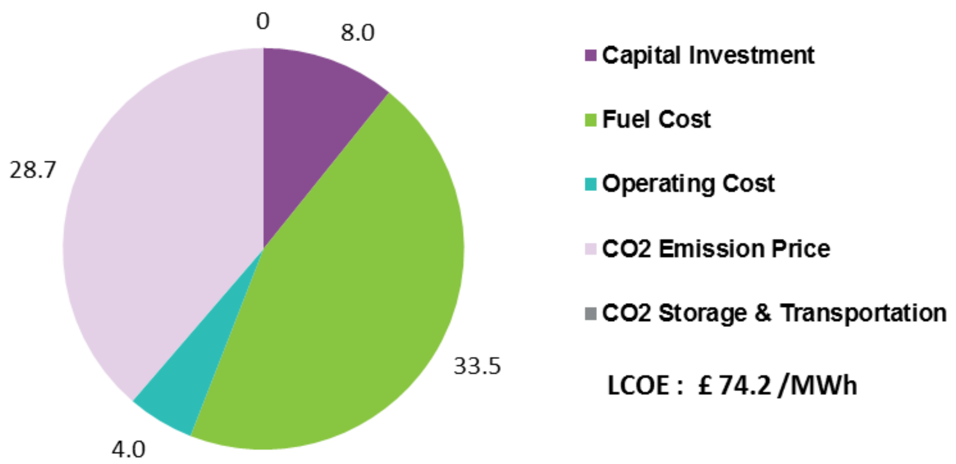

Figure 6-4: LCOE (£/MWh) Contribution for Case 0

6.5.1 Comparison of Results with IEAGHG 2012/08 Report

The value of £ 69.9 / MWh for Levelised Cost of Electricity (LCOE) presented in this report differs significantly from the equivalent result of £ 57.8 / MWh (€ 70.7 / MWh) reported for Scenario 3b in the IEAGHG Report 2012/08. This results from a variety of differing assumptions used for the two studies, as summarised in Figure 6-5 below.

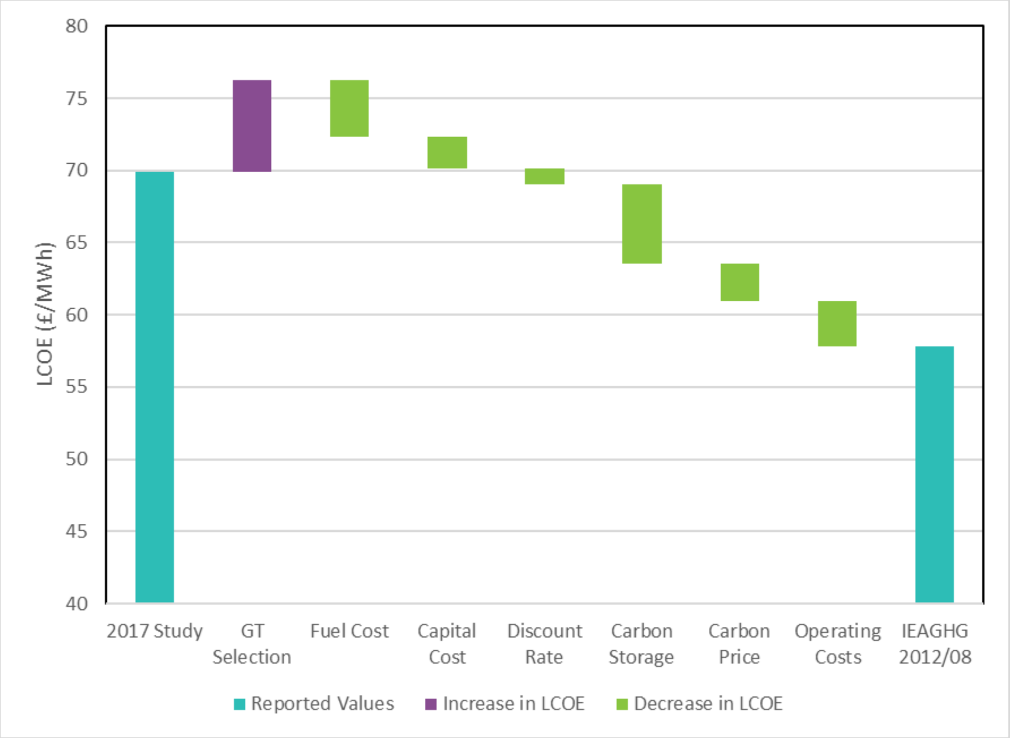

Figure 6-5: Case 1 LCOE Comparison with IEAGHG 2012/08 Report

The plant performance using the GE 9HA gas turbine is superior to that of the GE 9FB turbine used in the 2012 study, with overall plant efficiency increased from 52.0% to 54.9%. By chance, the use of a gas turbine degradation profile in the current study exactly counteracts the increased efficiency, reducing it to an average of 52.0%. However, using the 2012 performance has other effects that would increase the LCOE: lower rates of CO_2 emitted to atmosphere and directed to storage, but more significantly a 25% reduction in power export, which increases the LCOE.

The 2012 study used a base price of € 6 / GJ (£ 17.7 / MWh) for natural gas, which seemed appropriate at that time before the shale gas revolution caused prices to crash. The gas price profile used for this study (Table 5-7) indicates a 2017 gas price that that is about 60% of the 2012 value, but then increases in real terms so that the price in the current study is 18% higher from 2030 onwards. Across the whole lifecycle, the fuel costs are higher for this study.

The EPC costs for the two studies are different, primarily because the current study uses a larger capacity gas turbine, resulting in equivalent increases in size for the other equipment. Due to economies of scale and improvements in the design of the CO_2 removal unit, the increase in cost is not as large as it might otherwise be. Using the 2012 capital cost estimate would result in a small drop in the LCOE.

The 2012 study used a slightly lower discount factor of 8.0% versus 8.9% in the current study. It also assumed lower costs for CO_2 transportation and storage (€ 5 / tCO_2) and for carbon emissions (€ 10 / tCO_2). These all contribute to higher LCOE in this study than in the 2012 study.

A combination of small differences in the estimation of operating costs has an effect upon the LCOE for the two studies. The 2012 study assumed much lower costs for General Administration and Overheads, Maintenance, Insurance and Local Taxes.

7 Case 2 – Natural Gas IRCC with Pre-Combustion Carbon Capture

7.1 Overview

This case consists of a natural gas fed integrated reforming combined cycle (IRCC) power plant based upon two gas reforming trains feeding 2 x GE Frame 9 syngas variant gas turbines each with a dedicated heat recovery steam generator (HRSG) and steam turbine in a 2x2 configuration. The natural gas is reformed in an auto-thermal reforming process, shifted to maximise pre-combustion CO_2 production with CO_2 subsequently captured in a Selexol physical absorption process. The captured CO_2 is then compressed in 4 stages, dehydrated and then compressed further to the required export pressure of 110 bar (abs).

7.1.1 Process Configuration

The main process configuration of the IRCC plant is as follows:

- Auto-thermal Reforming of natural gas with air and steam;
- Process air for reforming extracted from Gas Turbine compressor;
- Two stages water gas shift reaction process;
- Acid gas removal (CO_2) using Selexol physical solvent system;
- CO_2 compression and pumping up to 110 bara;
- Combined cycle based on two GE F-class syngas variant gas turbines each with a dedicated heat recovery steam generator (HRSG) and steam turbine.

Table 7-1 describes the process units with trains which are also shown in Figure 7-1.

Table 7-1: IRCC Process Units with Trains

Unit Number	Unit Description	Trains
100	Fuel Pre-treatment & Pre-reformer	2 x 50%
200	Auto-thermal Reforming & Shift Process	2 x 50%
300	Acid Gas Removal (AGR)	2 x 50%
400	CO_2 Compression & Dehydration	2 x 50%
500	Gas Turbine & Generator Package	2 x 50%
600	Heat Recovery Steam Generation	2 x 50%
700	Steam Turbine & Generator Package	2 x 50%
800	Offsite & Utilities	

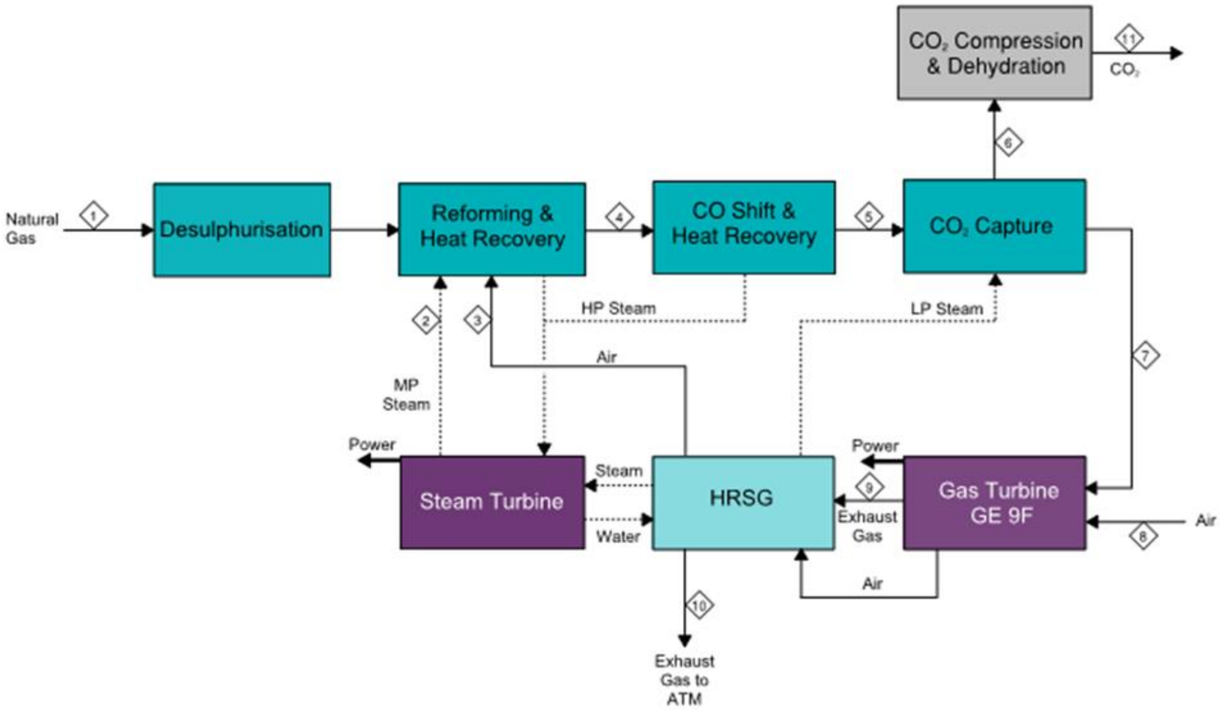

Figure 7-1: Case 2 Block Flow Diagram

7.2 Model Development

This model was built based upon a combination of the design experience gained from the two FEED projects developed by Wood using natural gas fired ATR scheme and the natural gas pre-combustion case in the reference IEAGHG 2012 'CO_2 Capture at Gas Powered Power Plants' report (Case 5). The process has been developed to maximise integration between the reforming section and the CCGT to achieve an overall plant efficiency as high as possible.

The natural gas flow rate to the IRCC complex has been kept consistent with IEAGHG 2012 Case 5. A natural gas specification meeting the UK National Grid specification was used into the model.

The overall integrated reforming combined cycle system has been simulated in Aspen Hysys using a process flow scheme similar to that given in the reference IEAGHG 2012 report. The Peng Robinson property package was used to model the process scheme, whereas the water and steam cycle has been modelled using the NBS Steam property package. However, there are few notable differences between the process parameters of the model for this study and the IEAGHG report.

- The auto-thermal reformer operating pressure used for this study (30 barg) is based on typical information received in from the equipment suppliers, which is lower than the IEAGHG 2012 report. The IEAGHG 2012 report was based on a higher reformer pressure (40 barg) to avoid the recompression of syngas to the gas turbine; however, this design suffered from reduced equilibrium conversion of methane (~94%). Reducing the operating pressure to ~30 barg increases the equilibrium conversion of methane from 94% to >99%.
- Due to the lower equilibrium conversion of ~ 94%, the methane slip in the syngas is higher in the IEAGHG report model leading to lower overall carbon capture of 81.6%; whereas achieving >99% methane conversion for the BEIS model leads to 90% overall carbon capture.

The heat integration between process units and the power plant has been developed to minimise the heat loss from the system and to maximise power output. Process air for reforming is extracted from the gas turbine compressor: this hot air is also used for process heating. The HP, MP and LP steam conditions and the process pressure profile are in line with the design experience gained from two previous FEED projects developed by Wood.

7.3 Process Description

7.3.1 Natural Gas Reforming and Shift

Natural gas is received from the grid and is metered before being routed to the reforming unit. The gas is pre-heated, mixed with some recycled decarbonised fuel gas and passed through a hydrogenator to convert any mercaptans in the feed gas to H_2S. The H_2S is then removed in zinc oxide beds as sulphur species are poisonous to the downstream reformer catalyst. Steam is mixed with the desulphurised feed gas and the combined stream is further heated before passing through a pre-reformer which partially reforms the gas, particularly the heavier components, prolonging the life and reducing the total duty of the main reformer.

The pre-reformed gas is then fed to the main Auto-Thermal Reformer (ATR), using air extracted from the gas turbine air compressor. The ATR process using air as the oxidant is attractive for IRCC schemes because gas turbines running on CO_2 depleted syngas require a high degree of fuel dilution, which is provided inherently by the nitrogen in the air. The ATR converts the methane to a mixture of CO_2, CO and hydrogen while producing a significant amount of high grade heat which is recovered by generation of HP steam in a waste heat boiler.

The reformed gas is then shifted to convert more of the CO to CO_2 while producing additional hydrogen in a two stage, high temperature followed by low temperature shift process with intermediate heat recovery. The shifted syngas is then cooled further prior to CO_2 removal.

7.3.2 CO_2 Capture

Cooled shifted syngas is fed to a physical solvent CO_2 removal unit (Selexol process). CO_2 is removed from the syngas stream in an absorber tower where chilled Selexol solvent is contacted with the gas stream. The CO_2 depleted syngas, essentially decarbonised fuel gas composed of approximately 44 mol% nitrogen and 52 mol% hydrogen, is then fed to the power island.

The CO_2 rich solvent is flashed in a series of successively lower pressure stages to recover most of the CO_2, with the flashed gas from the first flash at about 8 barg being recycled back to the CO_2 capture unit inlet via a recycle compressor. Prior to the final stage of flashing at 0.5 barg, the solvent is heated against the lean regenerated solvent, and then further against hot shifted syngas, or LP steam. The flashed vapour streams are comprised mostly of CO_2 and are sent to the CO_2 compression and dehydration unit.

7.3.3 Power Island

The decarbonised fuel gas is heated to about 80°C against waste heat and then further to about 190°C to maximise GT efficiency. The ratio of nitrogen to hydrogen is controlled by varying the quantity of air fed to the ATR. The preheated fuel gas is then fed to two parallel GE Frame 9 syngas variant gas turbines.

The exhaust gas from each GT is then passed to a dedicated heat recovery steam generator (HRSG) for each machine which cools the exhaust gas to about 80°C before releasing it to the atmosphere via a stack. In the HRSG, three pressure levels of steam are generated which are fed to steam turbines. The HRSG includes reheating of the medium pressure steam from the exhaust of the HP section of the steam turbine as well as pre-heating of air for the ATR.

7.3.4 CO_2 Compression and Dehydration

The water saturated CO_2 from the top of the absorber column is partially condensed, the aqueous phase from which is then returned to the stripper column as reflux, before being fed to the CO_2 compressor. The CO_2 is compressed to 30 to 40 barg in 3 or 4 stages, each with intercooling and water knock-out. This recovers the vast majority of the water content, but is not sufficient for most pipeline specifications. Numerous studies have compared drying with tri-ethylene glycol (TEG) versus use of molecular sieve adsorption, which concluded that there is little to choose between the two methods. For the purposes of this study we have assumed a TEG dehydration unit is selected, since that was the selection made in the reference study, IEAGHG 2012 report.

Following dehydration, the CO_2 passes through a final stage of compression to the export pressure of 110 bar (abs).

7.4 Technical Performance Evaluation

Table 7-2: Technical Performance Comparison for Case 2

	Units	Reference Case (Unabated CCGT)	Natural Gas Pre-combustion (IRCC) with CCS
Total Gross Installed Capacity	MWe	1229.4	919.1
Gas Turbine (s)	MWe	823.5	554.4
Steam Turbine	MWe	405.9	364.7
Others	MWe	0	0
Total Auxiliary Loads	MWe	20.9	101.3
Feedstock Handling	MWe	0	0
Power Island	MWe	14.7	10.4
Air Separation Unit	MWe	0	0
CO_2 Capture	MWe	0	45.8
CO_2 Compression	MWe	0	34.7
Utilities	MWe	6.2	10.2
Net Power Export	MWe	1208.5	817.9
Fuel Flow Rate	kg/h	150,296	147,539
Fuel Flow Rate (LHV)	MWth	1940.2	1906.5
Net Efficiency (LHV) - As new	%	62.3	42.9
Net Efficiency (LHV) - Average	%	59.0	40.7
Total Carbon in Feeds	kg/h	108,640	106,647
Total Carbon Captured	kg/h	0	96,418
Total CO_2 Captured	kg/h	0	353,319
Total CO_2 Emissions	kg/h	398,105	37,483
CO_2 Capture Rate	%	0	90.4
Carbon Footprint	kg CO_2/MWh	329.4	45.8

The plant performance of the full scale pre-combustion system with carbon capture (2 trains of 954 MWth fuel input) using natural gas fuel is summarised in the above table. The overall performance of the system includes CO_2 balance and removal efficiency. The unabated CCGT Reference case is included in the table for the purposes of comparison. The IRCC case captures 90% of the CO_2; however, it suffers from a 19.4% net efficiency loss.

The following points can be highlighted as basic difference between the two cases:

- The Reference case uses one of the largest and most efficient natural gas fired gas turbines GE 9HA.01 with large power output of > 400 MWe per gas turbine.

- The IRCC case uses a GE Frame 9 syngas variant gas turbine fired with syngas produced from natural gas reforming. The gas turbine efficiency is ~ 42% with the gross power output ~ 300 MWe per turbine. The power output from the gas turbine is further reduced to ~ 277 MWe due to parasitic load required for the large gas turbine compressor used to provide the process air for reforming. Overall, due to the combination of two different types of gas turbines fired by different fuel (natural gas and syngas) and process heat integration, the gross power output from the IRCC case is 390 MWe less than the Reference case with similar natural gas fuel input to the process boundary.
- The carbon footprint for the IRCC case is ~ 7 times lower than the Reference unabated case as the IRCC case captures 90% of the process CO_2 for transportation and storage.

7.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staffs are taken to be in daily positions, working regular hours.

Table 7-3: Operations and Maintenance Staff Manning for Case 2

	Reference Case Unabated CCGT	Natural Gas Pre-combustion (IRCC) with CCS	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Deputy Plant Manager	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	Daily Position
Process Engineer	NA	2	Daily Position
Shift Supervisor	5	10	3-shift Position
Electrical Assistant	5	5	3-shift Position
Control Room Operator	10	15	3-shift Position
Field Operator	10	30	3-shift Position
Sub-Total	32	65	
Maintenance Staff			
Mechanical Group	3	4	Daily Position
Instrument Group	3	3	Daily Position
Electrical Group	2	2	Daily Position
Sub-Total	8	9	
Laboratory Staff			
Superintendent	1	1	Daily Position
Analysts	3	4	Daily Position
Sub-Total	4	5	
Plant Total Staff	44	79	* See note below

* Note that the IEAGHG 2012/08 report estimated 101 permanent roles in Operations and Maintenance

Table 7-4: Economic Performance Comparison for Case 2

	Units	Reference Case (Unabated CCGT)	Natural Gas Pre-combustion (IRCC) with CCS
Total Project Cost	£M	672.2	1,256.3
Specific Total Project Cost	£/kW	556	1,536
Pre-Development Costs			
Pre-Licensing & Design	£M	5.8	11.1
Regulatory & Public Enquiry	£M	12.9	23.6
EPC Contract Cost	£M	583.6	1,107.1
Feedstock Handling	£M	0	0
Power Island	£M	583.6	785.5
Air Separation Unit	£M	0	0
CO_2 Capture	£M	0	78.2
CO_2 Compression	£M	0	51.4
Utilities	£M	0	191.9
Other Costs			
Infrastructure Connections		29.0	37.0
Owner's Costs		40.9	77.5
Overall CAPEX Impact (vs Ref Case)	£M	-	584.1
Overall CAPEX Impact (vs Ref Case)	%	-	87
Total Fixed OPEX	£M pa	36.2	60.3
Total Variable OPEX (excl. Fuel & Carbon)	£M pa	0.2	58.2
Average Fuel Cost	£M pa	315	283
Average CO_2 Emission Cost	£M pa	369	31.7
Total Start-up Cost (excl. Fuel)	£M	4.4	10.0
Discount Rate	% / year	7.8	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	100.0
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	91.1
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	96.2
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	178.9

The economic performance of the full scale pre-combustion system with carbon capture using natural gas fuel is summarised in the Table 7-4 along with the unabated CCGT Reference case. The capital cost estimate for Case 2 is assessed to have an accuracy of ± 30%. The total project cost for the IRCC case is 87% higher than the Reference case for reasons explained below:

- The Power Island cost for the IRCC case is 35% higher than the Reference case as the IRCC Power Island includes natural gas pre-treatment (hydrogenation, desulphurisation and pre-reforming), auto-thermal reforming and combined cycle costs.

- Carbon capture and compression cost for the IRCC case contributes to the ~£320 M (including utility costs), which is not required for the unabated case.
- Predevelopment and Owners' costs are higher for the IRCC case as these are calculated as a percentage of the EPC contract value.

Total fixed operating cost for the IRCC system is 66% higher than the Reference case, since a majority of the fixed cost (such as general overhead, taxes, maintenance, etc.) are calculated as a percentage of capital costs. The total variable operating cost excluding fuel for the IRCC case is related to the carbon capture process, which is not relevant for the unabated Reference case. The combination of fixed and variable OPEX leads to ~3.3 times higher for the IRCC case compared to Reference case.

The increased project cost, operating cost and cost related to the CO_2 transportation & storage makes the levelised cost of electricity for the IRCC higher than Reference case. Figure 7-2 shows the list of the different contributing factors and the level of contribution towards the overall LCOE value. It is important to note that the fuel cost is the biggest contributor to the LCOE followed by capital investment and operating cost.

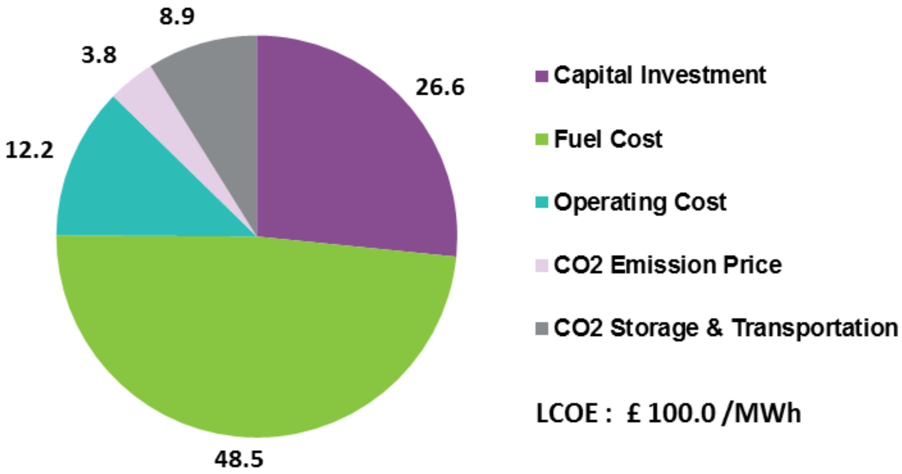

Figure 7-2: LCOE (£/MWh) Contribution for Case 2

7.5.1 Comparison of Results with IEAGHG 2012/08 Report

The value of £ 100.0 / MWh for Levelised Cost of Electricity (LCOE) presented in this report differs significantly from the equivalent result of £ 75.0 / MWh (€ 91.7 / MWh) reported for Scenario 5 in the IEAGHG Report 2012/08. This results from a variety of differing assumptions used for the two studies, as summarised in Figure 7-3 below.

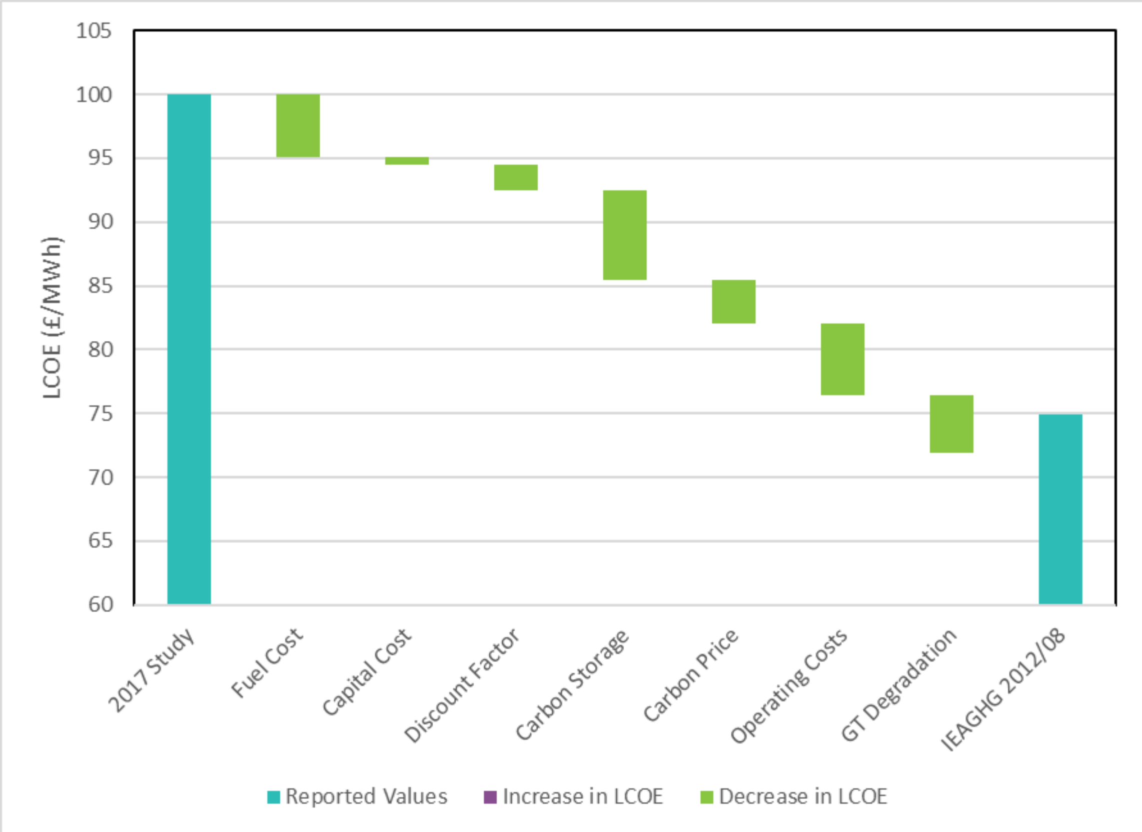

Figure 7-3: Case 2 LCOE Comparison with IEAGHG 2012/08 Report

The 2012 study used a base price of € 6 / GJ (£ 17.7 / MWh) for natural gas, which seemed appropriate at that time before the shale gas revolution caused prices to crash. The gas price profile used for this study (Table 5-7) indicates a 2017 gas price that that is about 60% of the 2012 value, but then increases in real terms so that the price in the current study is 18% higher from 2030 onwards. Across the whole lifecycle, the fuel costs are higher for this study.

The EPC costs for the two studies were very similar (despite the different estimating methodologies) and so this has little impact on the LCOE.

The 2012 study used a slightly lower discount factor of 8.0% versus 8.9% in the current study. It also assumed lower costs for CO_2 transportation and storage (€ 5 / tCO_2) and for carbon emissions (€ 10 / tCO_2). These all contribute to higher LCOE in this study than in the 2012 study.

A combination of small differences in the estimation of operating costs has a significant effect upon the LCOE for the two studies. The 2012 study assumed much lower costs for General Administration and Overheads, Maintenance, Insurance and Local Taxes.

The LCOE is also higher in the current study because the lifetime degradation in gas turbine performance has been accounted for.

8 Case 3 – Coal SCPC with Post-Combustion Carbon Capture

8.1 Overview

This case consists of a pulverised coal fired supercritical power plant in a once through steam generator with superheating and single steam reheating, and a single steam turbine at a 1000 MWe net power production scale. The flue gas from the boiler is routed to a gas/gas heat exchanger, is boosted in pressure using a flue gas fan, then is fed to the flue gas desulphurisation unit. The desulphurised flue gas is fed to a proprietary CO_2 capture unit where it enters a direct contact cooler. CO_2 is captured from the cooled flue gas using an amine based solvent in an absorption column and is released from the solvent in the stripper. The captured CO_2 is compressed in 4 stages, dehydrated, compressed in a further stage and then pumped to the required export pressure of 110 bar (abs).

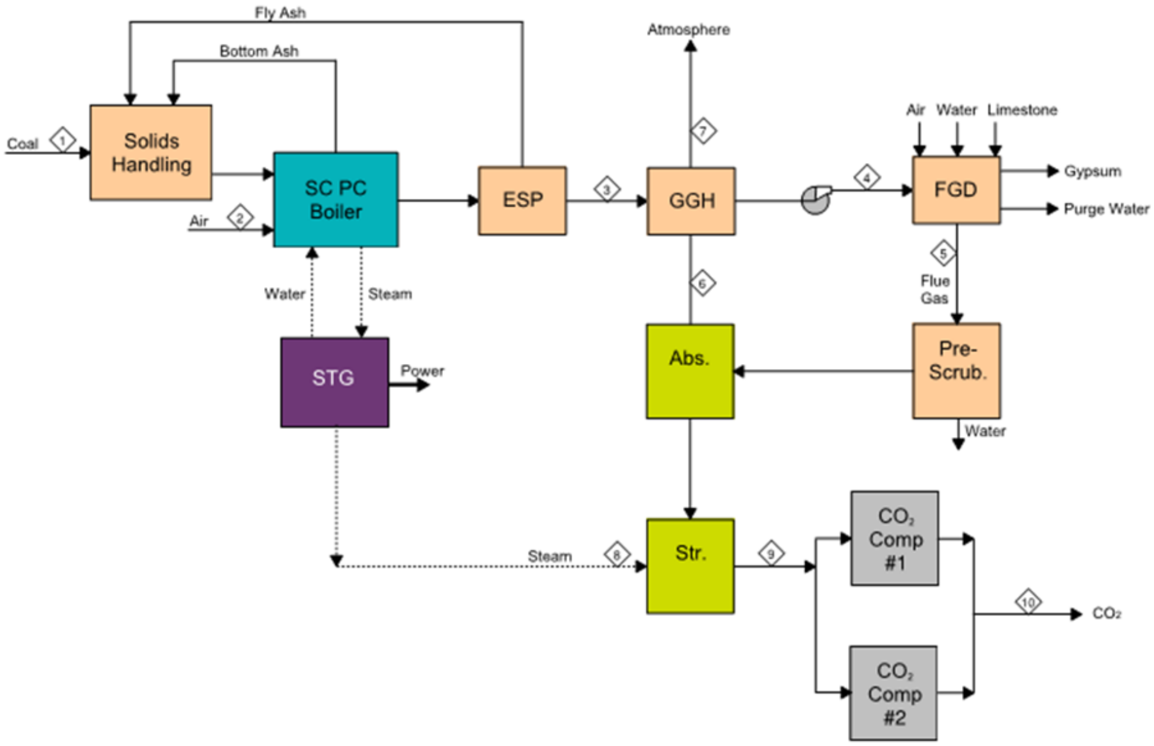

Figure 8-1: Case 3 Block Flow Diagram

8.2 Model Development

The supercritical pulverised coal boiler (SCPC) was modelled using Gatecycle, which models the air preheating, coal combustion, steam generation, steam turbine and boiler feed water (BFW) preheat train. The SCPC Power Island was simulated in order to provide verification of the IEAGHG figures, and so that the impact on the steam turbine power output could be determined for both the Shell Cansolv benchmark and the future novel technology cases.

A Gatecycle model was constructed using the same flow scheme as that given in the reference IEAGHG 2014/03 Coal and Hydrogen with CCS Report. The IEAGHG report feed coal, which is an appropriate coal for a UK study as well, was specified in the model along with the same steam conditions. It was possible to achieve a very high level of agreement with the IEAGHG report results by modifying the residual carbon in the coal ash, the steam conditions and the air preheater.

The BFW pumps are driven by a steam turbine in the IEAGHG flow diagram (reducing the parasitic electrical load). This was modelled in Hysys to estimate the required steam flow and then taken as a steam extraction parallel to the extraction to the CO_2 capture reboilers in the Gatecycle model. The condensate pump is electrically driven and becomes the main contributor to the electrical parasitic load of the power island.

Cansolv Technologies Inc. (Shell Cansolv) was contracted to provide a Process Design Package for absorption and regeneration system using its proprietary CANSOLV Absorbent DC-103. This solvent was used at the SaskPower Boundary Dam facility in Canada and is generally recognised as an industry leader in post-combustion CO_2 capture. All modelling of the proprietary amine absorption and stripping systems was conducted by Shell Cansolv. Detailed results cannot be published in this report due to confidentiality restrictions and some details have been redacted from the attached deliverables, but the overall results presented in this section reflect the latest performance results achieved by Shell Cansolv.

Figure 8-2: SaskPower Boundary Dam Facility
(Image courtesy of Cansolv Technologies Inc.)

The CO_2 compression system can be modelled with accuracy, including high level key heat integration, and this has been undertaken in Hysys for this study.

8.3 Process Description

8.3.1 Solids Storage and Handling

Coal is received at the plant via train, unloaded and conveyed to the coal storage pile which holds an inventory of 30 days of coal feed to the plant. Coal is conveyed to feed hoppers then fed to two parallel crushers which break down lumps of coal to maximum size of 35mm. This coal is then conveyed to day silos. Tramp iron is recovered from the coal using magnetic plate separators.

Limestone is also delivered to site by train and stored with 30 days' inventory in a dedicated storage building. Conveying and crushing systems are similar to those used for the coal.

Fly ash from the Electrostatic Precipitator (ESP) and bottom ash from the boiler itself are collected into storage silos. Bottom ash requires crushing for ease of transportation before both ash types are loaded onto trucks for transportation.

Gypsum is the product of limestone's reaction with sulphur species in the flue gas in the Flue Gas Desulphurisation (FGD) unit. It is discharged from the FGD as a paste and stored in a dedicated storage building. It is also loaded onto trucks for transportation.

8.3.2 Power Island

The supercritical pulverised coal boiler is treated as a specialist package and is a typical commercial single pass tower type boiler.

The boiler features low-NOx burners located in the bottom part of the furnace with staged combustion to also help minimise NOx formation along with over-fired air use. Fans force air from the atmosphere through the preheater where it is heated against the flue gas.

Coal from the day silos is pulverised in mills and conveyed pneumatically by the pre-heated primary air to the burners. The remaining air is supplied via the staged combustion system.

Hot combustion products exit the main furnace and their heat is recovered in first the radiant section then the convection sections before passing through the regenerative air preheaters. As much heat is recovered as possible into the steam cycle.

Boiler feed water (BFW) from a deaerator is pumped using steam driven boiler feed water pumps to over 300 bar and is preheated using successively higher temperature and pressure steam extracted from the steam turbine. The pre-heated BFW then passes through the economiser in the convection section, then the water wall of the furnace, then primary and secondary superheaters to become supercritical steam at 620°C and 270 barg when it is fed to the HP section of the steam turbine.

The MP steam from the exhaust of the HP section of the steam turbine is returned to the radiant section for reheating to 600°C before entering the MP section of the steam turbine. Part of the LP steam is then used to drive the BFW pumps, another part is routed to the reboilers in the CO_2 capture unit and the remainder passes on to the LP section of the steam turbine.

The LP steam turbine exhausts at vacuum conditions of 0.04 barg, or as close to that pressure as can be achieved in the condenser given the cooling water temperature. The condenser is directly below the LP steam turbine and also receives the exhaust from the steam turbine drives of the BFW pumps and the required make-up water.

Steam condensate is pumped to approximately 10 barg and preheated in the CO_2 capture and compression units, then using steam extractions before being returned to the deaerator to complete the circuit.

8.3.3 Flue Gas Treatment

In addition to the low-NOx burners and overfire air, additional NOx removal is also required. To achieve this a Selective Catalytic Reduction (SCR) system is included between the convection section and the air preheater. Ammonia, the reducing agent, is injected immediately upstream of a catalyst surface where the NO and NO_2 are reduced to N2 and water.

Flue gas from the air preheater is passed through the ESP, which removes fly ash: a baghouse may be an alternative option with potential NOx reduction benefits, but was not considered in this case. The flue gas is cooled in a gas-gas heat exchanger (GGH) and drawn through an induced draft fan before entering the flue gas desulphurisation package, which removes the sulphur compounds and some of the NOx. The other side of the gas-gas heat exchanger warms the treated flue gas from the CO_2 capture unit in order to provide sufficient buoyancy for good dispersion at the top of the stack.

8.3.4 Proprietary Solvent CO_2 Capture

An outline of the Shell Cansolv process as applied to a supercritical pulverised coal power plant is shown in the following diagram:

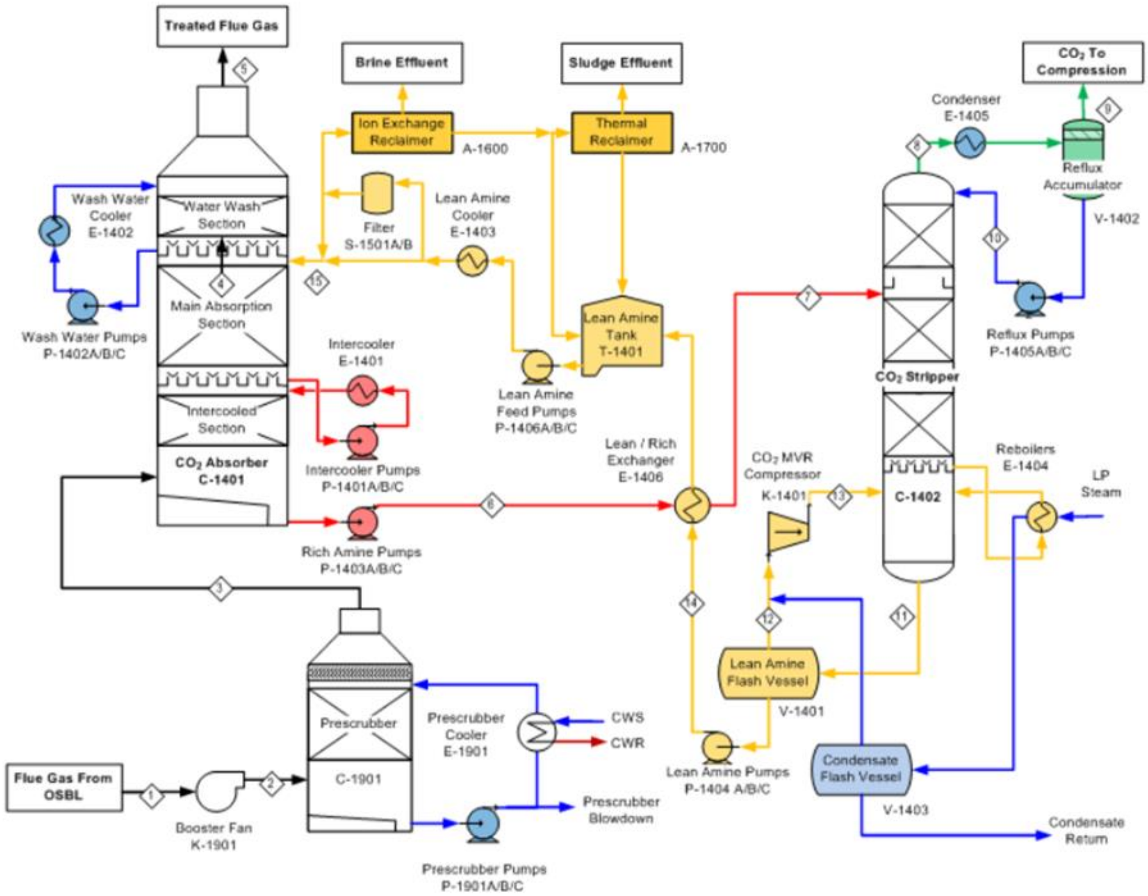

Figure 8-3: Cansolv SCPC Process Configuration
(Image courtesy of Cansolv Technologies Inc.)

Flue gas is ducted to a pre-scrubber column which performs the dual function of sub-cooling the stream to below its water saturation temperature, causing water condensation, and SO_2 removal. The flue gas is sub-cooled to 35°C to reduce the required absorbent circulation rate and thus energy consumption and CAPEX of the CANSOLV unit.

In order to decrease the impact of SO_2 on the absorbent the pre-scrubber will use caustic to reduce the SO_2 content in the flue gas upstream of the CO_2 Absorber. SO_2 removal is controlled by adding caustic on pH control in a caustic polishing section inside the pre-scrubber column. For the purposes of this study it is assumed that the concentration of SO_2 leaving the pre-scrubber polishing section would be 1 ppmv.

The cooled and pre-scrubbed flue gas is ducted to the bottom of the absorption column. For a plant of this scale and flue gas type, typical absorber dimensions range from 15m to 21m in square cross section. CO_2 absorption from the flue gas occurs by counter-current contact with CANSOLV Absorbent DC-103 in the CO_2 Absorber which is a vertical multi-level packed-bed tower. CO_2 is absorbed into the solvent by chemical reaction leaving a flue gas depleted in CO_2 at the top of the column.

The absorption reaction is exothermic, so for high CO_2 inlet concentrations, such as this coal case, the heat generated by the absorption must be removed from the system. This is required to prevent an excessive temperature increase in the absorbent, which would reduce the absorption capacity and increase water evaporation from the absorbent into the heated flue gas. To remove heat from the absorption column, hot absorbent is collected on a chimney tray above the bottom packing section, pumped to the CO_2 Absorber Intercooler via the Absorber Intercooler Pumps, and then returned to the absorber to complete CO_2 absorption in the bottom packing section.

A water wash packed bed section is included at the top of the CO_2 Absorber to capture volatile or entrained absorbent and to condense water to maintain the water balance in the system. Wash water is drawn from a chimney tray and is re-circulated to the top of the packed section, via the Water Wash Cooler, by the Water Wash Pumps. The Wash Water Cooler reduces the temperature of circulating wash water, which minimises water loss and enhances capture efficiency of the volatile absorbent. Water condensed from the flue gas into the wash water section overflows from the chimney tray to the CO_2 absorption section below. The treated flue gas leaving the Water Wash Section flows upwards to the stack and is released to atmosphere. The design flue gas outlet temperature is selected such that the overall required water make-up rate is minimised.

The treated flue gas is then warmed against flue gas exiting the ESP in the gas/gas heat exchanger and routed to a stack for discharge to the atmosphere.

The rich absorbent is collected in the bottom sump of the CO_2 Absorber and is pumped by the CO_2 Rich Absorbent Pumps and heated in the CO_2 Lean/Rich Exchangers to recover heat from the hot lean absorbent discharged from the Lean Absorbent Flash Vessel. The rich absorbent is piped to the top of the CO_2 Stripper for absorbent regeneration and CO_2 recovery. The rich absorbent enters the column under the CO_2 top packing section and flows onto a gallery tray that allows for disengagement of any vapour from the rich absorbent before it flows down to the two stripping packing sections under the gallery tray. The rich absorbent is depleted of CO_2 by water vapour generated in the Regenerator Reboilers which flows in an upward direction counter-current to the rich absorbent.

Lean absorbent flowing from the bottom packing section of the CO_2 Regenerator is collected on a chimney tray and gravity fed to the Regenerator Reboilers. Water vapour and lean amine flow by thermosyphon effect from the reboilers back to the CO_2 Regenerator sump, underneath the chimney tray. Water vapour flows upwards through the chimney tray to strip the CO_2 while the lean absorbent collects in the bottom sump. Lean absorbent flows by gravity from the CO_2 Regenerator sump, through a level control valve to the Lean Absorbent Flash Vessel, where it flashes and releases water vapour for reuse in the CO_2 Regenerator. Water vapour released from the absorbent in the Lean Absorbent Flash Vessel is compressed in the CO_2 mechanical vapour recovery (MVR) Package and introduced at the bottom of the CO_2 Regenerator to contribute to the stripping of the CO_2. This system minimizes the steam and energy consumption of the Cansolv Unit.

Water vapour in the regenerator, carrying the stripped CO_2, flows up the regenerator column into the top packing section, where a portion of the vapour is condensed by recycled reflux to enrich the overhead CO_2 gas stream.

The regenerator overhead gas is partially condensed in the Regenerator Condensers. The partially condensed two phase mixture gravity flows to the CO_2 Reflux Accumulator where the two phases separate. The reflux water is collected and returned via the Reflux Pumps to the regenerator rectification section. The CO_2 product gas is piped to the CO_2 Compression System. The pressure of the Regenerator can either be controlled by a product CO_2 discharge control valve or by the inlet guide vanes of the downstream CO_2 Compressors.

The flow of steam to the reboiler is proportional to the rich absorbent flow sent to the CO_2 Regenerator. The set-point of the low pressure steam flow controller feeding the Regenerator Reboilers is also dependent on the regenerator top temperature controller. The steam to absorbent flow ratio set-point is adjusted by this temperature controller. The temperature at the top of the column is set to maintain the required vapour traffic and stripping efficiency. The steam flow rate is controlled by modulating a steam flow control valve.

All amine based systems require some form of solvent maintenance system as over time the absorbent in the CO_2 Capture System accumulates Heat Stable Salts (HSS), as well as non-ionic amine degradation products, that must be removed from the absorbent. This is achieved through thermal reclamation. An ion exchange package is included for bulk HSS removal upstream of a thermal reclaimer.

The ion exchange package is designed to remove HSS from the Cansolv DC Absorbent. These salts are continuously formed within the absorbent, primarily due to residual amounts of NO_2 and SO_2 contained in the flue gas. Once absorbed, NO_2 forms nitric and nitrous acid while SO_2 forms sulphurous acid which oxidises to sulphuric acid. These acids, and some organic acids formed by the oxidative degradation of the amine, neutralise a portion of the amine, which is then inactivated for further CO_2 absorption.

The purpose of the Thermal Reclaimer Unit is to remove the non-ionic degradation products as well as HSS from the active absorbent. The thermal reclaimer unit distils the absorbent under vacuum conditions to separate the water and amine, leaving the non-ionic degradation products in the bottom. A slipstream is taken from the treated CO_2 lean absorbent exiting the ion exchange package and fed to the Thermal Reclaimer Unit. This stream will essentially consist of water, amine, degradation products, residual CO_2 and small amounts of sodium nitrate and sodium sulphate. The design flow rate of CO_2 lean absorbent sent to the thermal reclaimer is based on the calculated amine degradation rate. To maintain the degradation products below design concentration, the thermal reclaimer must process a specific flowrate of CO_2 lean absorbent. The reclaimed absorbent is sent to the Lean Absorbent Tank. The separated degradation products are stored in a storage tank, where they are diluted and cooled with process water. Diluted residues are periodically disposed of offsite, typically via incineration.

8.3.5 CO_2 Compression and Dehydration

The CO_2 is compressed to 30 barg in 4 stages, each with intercooling and water knock-out. This recovers the vast majority of the water content, but is not sufficient for most pipeline specifications. Numerous studies have compared drying with tri-ethylene glycol (TEG) versus use of molecular sieve adsorption, which conclude that there is little to choose between the two methods. For the purposes of this study we have assumed a TEG dehydration unit is selected, since that was the selection made in the reference study, IEAGHG 2014 report.

In the coal fired case final CO_2 pressurisation to 110 bar (abs) is achieved using one further stage of compression followed by a condenser then a stage of pumping.

8.4 Technical Performance Evaluation

Table 8-1: Technical Performance Comparison for Case 3

	Units	Reference Case (Unabated CCGT)	Coal SCPC Post-Combustion with CCS
Total Gross Installed Capacity	MWe	1229.4	953.5
Gas Turbine (s)	MWe	823.5	0
Steam Turbine	MWe	405.9	953.5
Others	MWe	0	0
Total Auxiliary Loads	MWe	20.9	139.4
Feedstock Handling	MWe	0	3.4
Power Island	MWe	14.7	31.2
Air Separation Unit	MWe	0	0
CO_2 Capture & Comp.	MWe	0	88.3
Utilities	MWe	6.2	16.5
Net Power Export	MWe	1208.5	814.2
Fuel Flow Rate	kg/h	150,296	325,000
Fuel Flow Rate (LHV)	MWth	1940.2	2335.5
Net Efficiency (LHV) - As New	%	62.3	34.9

	Units	Reference Case (Unabated CCGT)	Coal SCPC Post-Combustion with CCS
Net Efficiency (LHV) - Average	%	59.0	34.7
Total Carbon in Feeds	kg/h	108,640	209,950
Total Carbon Captured	kg/h	0	188,926
Total CO_2 Captured	kg/h	0	692,310
Total CO_2 Emissions	kg/h	398,105	77,040
CO_2 Capture Rate	%	0	90.0
Carbon Footprint	kg CO_2/MWh	329.4	94.6

The plant performance of the supercritical pulverised coal power plant with state-of-the-art Shell Cansolv post-combustion carbon capture is summarised in the above table. The unabated CCGT Reference case is also listed in the table for the purposes of comparison. The Cansolv case captures 90% of the CO_2, while suffering a 27.4% point net efficiency loss compared to the unabated gas case.

The following points can be highlighted as basic difference between the two cases:

- The Reference case uses one of the largest and most efficient natural gas fired gas turbines GE 9HA.01 with large power output of > 400 MWe per turbine.
- The supercritical pulverised coal power plant, by comparison, has an efficiency closer to ~ 39% before carbon capture is applied. Thus, the underlying power plant is far less thermally efficient.
- The inclusion of carbon capture results in additional parasitic load associated with the CO_2 capture and compression process of 88 MWe.
- In additional to the parasitic electrical load, there is a significant parasitic steam load required for regeneration of the proprietary solvent.
- This makes the net exportable power from the SCPC Cansolv case 394 MWe less than the Reference case while burning 395 MWth more fuel, thus there is a 27.4% point lower net LHV efficiency.
- The carbon footprint for the SCPC Cansolv case is less than one third that of the Reference unabated case as this case captures 90% of the process CO_2 for transportation and storage.

8.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staffs are taken to be in daily positions, working regular hours.

Table 8-2: Operations and Maintenance Staff Manning for Case 3

	Reference Case Unabated CCGT	Coal SCPC Post-Combustion with CCS	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position

	Reference Case Unabated CCGT	Coal SCPC Post-Combustion with CCS	Remarks
Deputy Plant Manager	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	Daily Position
Process Engineer	NA	3	Daily Position
Shift Supervisor	5	10	3-shift Position
Electrical Assistant	5	5	3-shift Position
Control Room Operator	10	20	3-shift Position
Field Operator	10	35	3-shift Position
Sub-Total	32	76	
Maintenance Staff			
Mechanical Group	3	6	Daily Position
Instrument Group	3	4	Daily Position
Electrical Group	2	3	Daily Position
Sub-Total	8	13	
Laboratory Staff			
Superintendent	1	1	Daily Position
Analysts	3	6	Daily Position
Sub-Total	4	7	
Plant Total Staff	44	96	* See note below

* Note that the IEAGHG 2014/03 report estimated 105 permanent roles in Operations and Maintenance

Table 8-3: Economic Performance Comparison for Case 3

	Units	Reference Case (Unabated CCGT)	Coal SCPC Post-Combustion with CCS
Total Project Cost	£M	672.2	1,732.2
Specific Total Project Cost	£/kW	556	2,128
Pre-Development Costs			
Pre-Licensing & Design	£M	5.8	15.5
Regulatory & Public Enquiry	£M	12.9	32.1
EPC Contract Cost	£M	583.6	1,547.3
Other Costs			
Infrastructure Connections		29.0	29.0
Owner's Costs		40.9	108.3
Overall CAPEX Impact (vs Ref Case)	£M	-	1,060.0
Overall CAPEX Impact (vs Ref Case)	%	-	158

	Units	Reference Case (Unabated CCGT)	Coal SCPC Post-Combustion with CCS
Total Fixed OPEX	£M pa	36.2	80.6
Total Variable OPEX (excl. Fuel & Carbon)	£M pa	0.2	108.0
Average Fuel Cost	£M pa	315	143
Average CO_2 Emission Cost	£M pa	369	69.1
Total Start-up Cost (excl. Fuel)	£M	4.4	10.9
Discount Rate	% / year	7.8	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	93.3
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	81.3
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	85.8
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	171.4

The economic performance of the SCPC with Cansolv post-combustion carbon capture is summarised in Table 8-3 along with unabated CCGT Reference case for the purposes of comparison. The capital cost estimate for Case 3 is assessed to have an accuracy of ± 30%.

The total project cost for this case is ~160% higher than the Reference unabated natural gas case while producing 33% less net power output:

- The power island and utilities capital cost is considerably higher than the power island and utilities cost of the equivalent CCGT plant. The Cansolv system adds to the total project cost with significant cost elements related to the large low pressure absorber tower and CO_2 compressor. The total effect is a much more capital intensive plant.
- The operating costs are also much higher than the operating costs of the reference natural gas plant, although this is partially related to the study's assumptions on how fixed operating costs are related to the capital cost. The higher CO_2 transportation and storage costs resulting from the larger carbon content of coal compared to natural gas also has a substantial effect on the variable operating costs.
- Both the capital and operating costs (excluding fuel and carbon price) are higher for this case, which compounded with a lower efficiency results in a significantly higher LCOE for this case when compared against the Reference case. However, it should be noted that this case provides the lowest LCOE of the coal cases, as can be seen in more detail in Section 13.

The chart below shows the balance of factors contributing to the overall LCOE. It can be seen that capital investment causes a larger portion of the LCOE than the operating costs, CO_2 storage and transportation, carbon price and fuel cost, a marked change from the gas-fired cases.

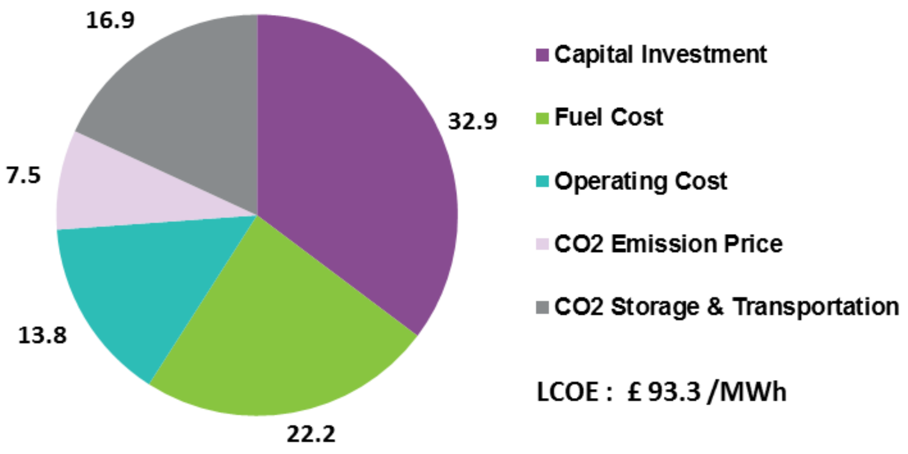

Figure 8-4: LCOE (£/MWh) Contribution for Case 3

8.5.1 Comparison of Results with IEAGHG 2014/03 Report

The value of £ 93.3 / MWh for Levelised Cost of Electricity (LCOE) presented in this report differs marginally from the equivalent result of £ 77.4 / MWh (€ 94.7 / MWh) reported for Case 2 in the IEAGHG Report 2014/03. This results from a variety of differing assumptions used for the two studies, as summarised in Figure 8-5 below.

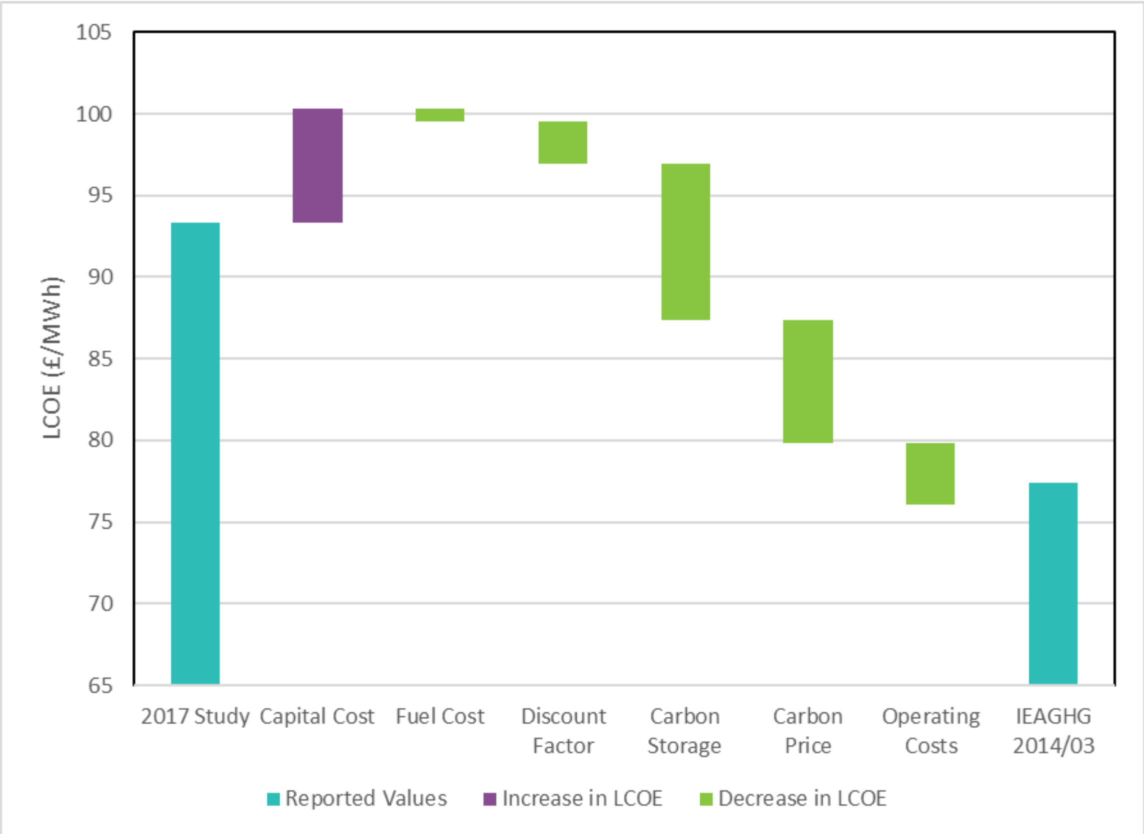

Figure 8-5: Case 3 LCOE Comparison with IEAGHG 2014/03 Report

The capital cost estimating methodology adopted in 2014 resulted in an EPC cost that is about 20% higher than in the current study, mainly due to the improvements in the size and efficiency of the CO_2 capture unit, but also due to changes in the installation factors applied. This is offset to some extent by the addition of costs for infrastructure connections and pre-development costs in the current study. Using the 2014 capital cost estimating methodology within the 2017 analysis would make a sizeable increase in the LCOE.

The coal price used in the 2014 study was € 2.5 / GJ (£ 7.4 / MWh). This is about 20% higher than the coal price for the start-up year of 2025, but is 9% lower than the cost used for 2030 onwards (refer to Table 5-7). Overall the cost impact between the study results is neutral.

The 2014 study used a slightly lower discount factor of 8.0% versus 8.9% in the current study. It also assumed lower costs for CO_2 transportation and storage (€ 10 / tCO_2) and excluded the price of carbon emissions entirely. These all contribute to a higher LCOE in this study than in the 2014 study.

A combination of small differences in the estimation of operating costs has a significant effect upon the LCOE for the two studies. The 2014 study assumed much lower costs for Maintenance, Insurance and Local Taxes, although the allowance for Administration and General Overheads was higher.

9 Case 4 – Coal SCPC with Oxy-Combustion Carbon Capture

9.1 Overview

This case consists of an oxy-fired pulverised coal fed supercritical power plant in a once through steam generator with superheating and single steam reheating, with a single steam turbine at the same thermal input capacity as the power island featured in Case 3. Oxygen for firing in the boiler is supplied by a cryogenic air separation unit (ASU). The flue gas from the boiler is routed to a multi-pass gas/gas heat exchanger (GGH) followed by heat recovery and fly ash removal before a portion of the flow is routed back to the boilers as Secondary Recycle, via the gas/gas heat exchanger. The remaining flue gas passes through further heat recovery before the Primary Recycle is split off and recycled via flue gas desulphurisation (FGD) and the gas/gas heat exchanger to the coal mills where it is employed as conveying gas to feed the pulverised coal to the boiler. The unrecycled flue gas stream is compressed and purified in a cryogenic purification unit (CPU) to the required export pressure of 110 bar (abs).

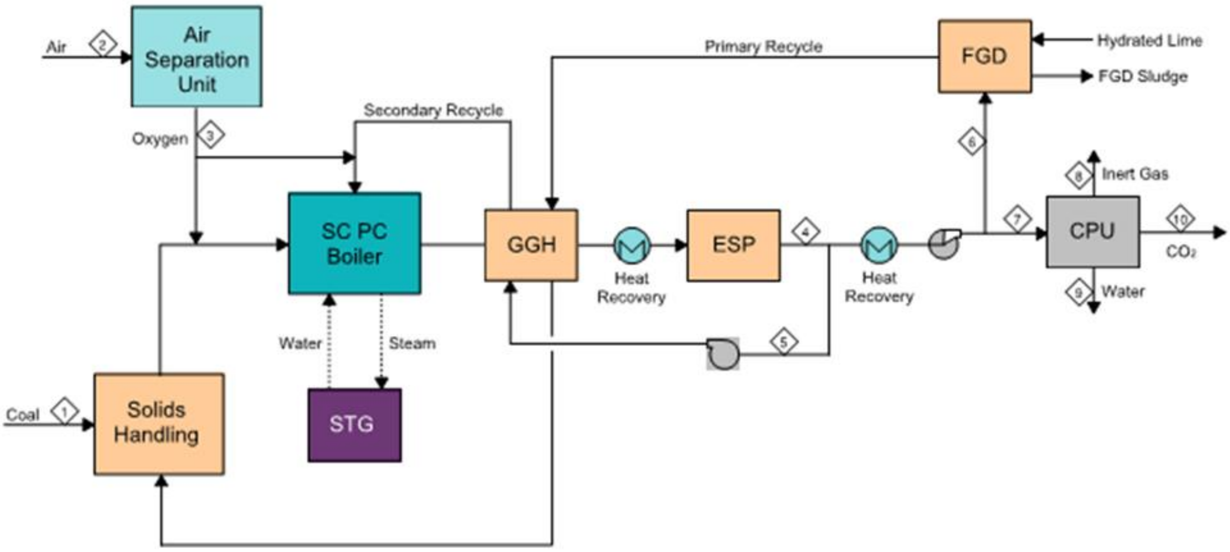

Figure 9-1: Case 4 Block Flow Diagram

9.2 Model Development

In this case, the supercritical pulverised coal boiler (SCPC) was modelled using Hysys. Gatecycle is not ideal for modelling non-air oxidant systems and using Hysys has the benefit of being able to build the entire plant in a single model, thus reducing the likelihood of inconsistencies between the power island and the surrounding plant.

The purpose of the modelling in this study is to verify the heat and material balance and determine the overall efficiency of the plant rather than for equipment sizing, although the model can be used to sense check and fill in blanks in the reference data. It is not intended to model the packaged units other than for high level thermodynamic and material balance checking.

The ASU, FGD and CPU were modelled using component splitters since these are quite specialised and refined package units. If a novel technology assessment requires further investigation inside the box of these units then a greater level of detail model can be developed at that stage. NOx generation and reduction was not modelled.

The boiler was modelled using a conversion reactor, thus allowing the residual carbon in the ash to be controlled. The ash itself was modelled as a user defined hypothetical component. In order to match the reference heat and material balance, in particular the nitrogen content in the recycles, it was necessary to add an air in-leak stream of 90 tph to the boiler.

The steam raising in the boiler and from the flue gas, and downstream heat exchange matched that of the reference without any significant differences in terms of achievable temperatures and flowrates. The contact cooler and recycles converged, matching the compositions and flows in the reference once the air in-leak stream was added.

9.3 Process Description

9.3.1 Air Separation Unit

Air from the atmosphere is filtered and initially compressed to 3.5 bar (abs) in an axial flow compressor. The hot outlet of this compressor could be used for pre-heating condensate in the Power Island, but cooling water is shown on the flow diagrams in the IEAGHG 2014 reference document in that case, which has also been assumed for this study, for consistency. Remaining impurities such as CO_2 and water are removed in adsorbent beds which alternate between adsorption and regeneration modes.

The purified 3.5 bar (abs) air stream is split in two, with one stream divided again into two streams and sent, via the main heat exchanger, to the intermediate pressure distillation column and the expander section of the compander respectively. The remaining 3.5 bar (abs) stream is compressed further, in two stages with intercooling, before being split into two further streams, passing through the main heat exchanger and the first stream being fed to the bottom of the high pressure column and the second stream being further divided to feed the mid-sections of both the intermediate and high pressure columns.

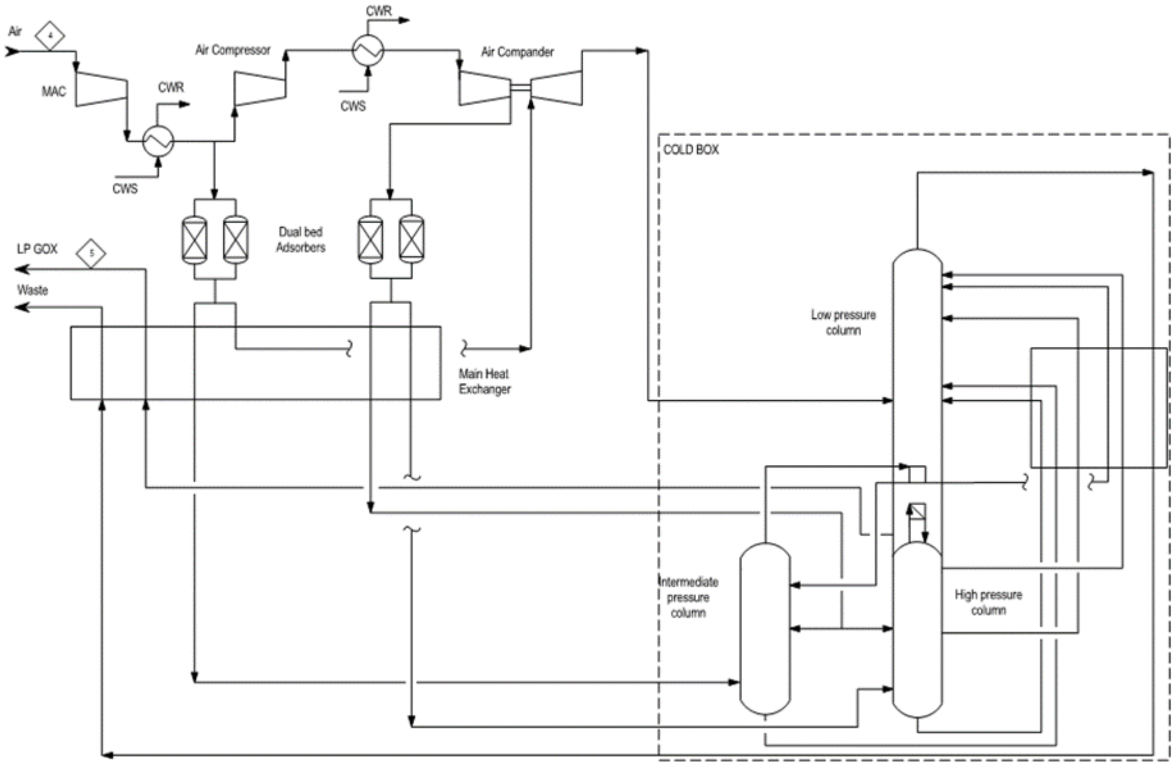

Figure 9-2: ASU Block Flow Diagram (from IEAGHG 2014/03, p388)

The main cryogenic heat exchanger consists of several parallel aluminium plate-fin exchanger blocks manifolded together. The cryogenic distillation columns are contained within a cold box and divided into low pressure, intermediate pressure and high pressure columns. Liquid oxygen product is produced from the bottom of the low pressure column at the required purity of 97%, while the final “waste” nitrogen stream is produced from the top of the low pressure column.

Since the oxy-combustion SCPC requires low pressure oxygen there is no need for further pressurisation of the oxygen stream in this case, therefore this ASU has a significantly lower energy usage per tonne of oxygen produced, compared to the IGCC case ASU (Case 5).

9.3.2 Solids Storage and Handling

Coal is received to the plant via train and unloaded and conveyed to the coal storage pile which holds an inventory of 30 days of coal feed to the plant. Coal is conveyed to feed hoppers then fed to two parallel crushers which break down lumps of coal to a maximum size of 35mm. This coal is then conveyed to day silos. Tramp iron is recovered from the coal using magnetic plate separators.

Limestone is also delivered to site by train and stored with 30 days inventory in a dedicated storage building. Conveying and crushing systems are similar to those used for the coal.

Fly ash from the electrostatic precipitator (ESP) and bottom ash from the boiler itself are collected into storage silos. Bottom ash requires crushing for ease of transportation before both ash types are loaded onto trucks for transportation.

Gypsum is the product of limestone's reaction with sulphur species in the flue gas in the FGD unit. It is discharged from the FGD as a paste and stored in a dedicated storage building. It is also loaded onto trucks for transportation.

9.3.3 Power Island

The oxy-fired supercritical pulverised coal boiler is treated as a specialist package and is a typical commercial single pass tower type boiler. It is expected that although oxy-fired versions of the supercritical pulverised coal boiler are not yet commercial at the scale required for this case, the behaviour and design features will not be significantly different from the air fired plant.

The boiler features low-NOx burners located in the bottom part of the furnace with staged combustion with flue gas and oxygen to help minimise NOx formation along with over-fired flue gas and oxygen use.

Coal from the day silos is pulverised in mills and conveyed pneumatically to the burners using a combined stream of oxygen and primary recycle flue gas. The remaining oxygen and the secondary recycle are supplied via the staged combustion system.

Hot combustion products exit the main furnace and their heat is recovered in first the radiant section then the convection sections before passing through the regenerative air preheaters. As much heat is recovered as possible into the steam cycle.

Boiler feed water (BFW) from a deaerator is pumped using steam driven boiler feed water pumps to over 300 bar and preheated firstly against cooling flue gas, then further using successively higher temperature and pressure steam extracted from the steam turbine. The pre-heated BFW then passes through the economiser in the convection section, then the water wall of the furnace, then primary and secondary superheaters to become supercritical steam at 620°C and 270 barg when it is fed to the HP section of the team turbine.

The MP steam from the exhaust of the HP section of the steam turbine is returned to the radiant section for reheating to 600°C before entering the MP section of the steam turbine. Part of the LP steam is then used to drive the BFW pumps and supply steam to the deaerator while the remainder passes on to the LP section of the steam turbine.

The LP steam turbine exhausts at vacuum conditions of 0.04 barg, or as close to that pressure as can be achieved in the condenser given the cooling water temperature. The condenser is directly below the LP steam turbine and also receives the exhaust from the steam turbine drives of the BFW pumps and the required make-up water.

Steam condensate is pumped to approximately 10 barg and preheated against flue gas and in the CO_2 compression and purification unit, then using steam extractions before being returned to the deaerator to complete the circuit.

9.3.4 Flue Gas Treatment and Recycles

A tubular preheater replaces the gas/gas exchanger used for flue gas cooling and combustion air preheating in the non-oxyfired SCPC Case 3. This exchanger uses the hot flue gas exiting the convection section to preheat both the primary and secondary recycles. Downstream of the tubular preheaters, the flue gas is further cooled against BFW then passes through the ESP, which removes fly ash.

Immediately downstream of the ESP, the flue gas is split into two streams. One becomes the secondary recycle and is returned to the boilers via a forced draft fan and the tubular preheater. The remaining flue gas stream is then cooled further against power island condensate before entering a contact cooler.

In the contact column, flue gas at approximately 110°C is quenched with water from the bottom of the direct contact cooler. In the contact column, a circulating water stream is used to cool the flue gas to about 28°C. The circulating water stream is cooled against cooling water and this system is a net producer of water as can be seen on the overall heat and material balance. This system also removes the more water-soluble acid species present in the flue gas, SO_3 and HCl.

The cooled flue gas is then boosted in the induced draft (ID) fan and further divided into two streams. The first stream becomes the primary recycle stream and is treated in a flue gas desulphurisation (FGD) unit, heated in the tubular pre-heater and routed to the solids handling area as conveying gas for the coal feed. The FGD unit for this application is a circulating fluid bed scrubber, which uses lime injection and a fabric filter to remove sufficient SO_2 to prevent accumulation in the recycle system.

The remaining flue gas from the ID fan is routed to the CO_2 compression and purification unit.

9.3.5 CO_2 Compression and Purification

The cooled flue gas from the ID fan is at approximately 38°C, low pressure, and composed of about 75 mol% CO_2, 14 mol% nitrogen, 6 mol% oxygen and nearly 2 mol% argon, with water making up the balance. Low levels of SO_2 and NOx also requiring removal before the stream is of sufficient purity for transportation and storage.

In this study, the process offered by Air Products has been assumed, to be consistent with the IEAGHG 2014 reference design. The process consists of the following process steps:

- Sour compression
- Temperature swing adsorption (TSA)
- Auto-refrigerated inerts removal
- Final compression to 110 bar (abs)

Sour Compression

The flue gas stream is compressed adiabatically to 15 bar (abs) and 300°C, then cooled against various streams which require heating; the inerts from the downstream cold box, boiler island BFW and condensate, and finally cooling water. The flue gas then undergoes several reaction steps.

The first reaction step is to oxidise any remaining NO to NO_2 in the following reaction:

$$NO + {}^{1}/_{2}\, O_2 \rightarrow NO_2 \qquad (1)$$

This is achieved readily at this pressure and low temperature while there is still plenty of oxygen present in the flue gas. Only a few seconds at 15 bar (abs) is required to achieve conversion of most of the NO.

The second step is to react NO_2 and SO_2 to produce sulphuric acid as follows:

$$NO_2 + SO_2 + H_2O \rightarrow NO + H_2SO_4 \qquad (2)$$

Remaining NO_2 is then converted to nitric acid as follows:

$$2NO_2 + H_2O \rightarrow HNO_2 + HNO_3 \qquad (3)$$

$$HNO_2 \rightarrow HNO_3 + 2NO + H_2O \qquad (4)$$

All of the NO produced in reactions 2 and 4 is reconverted to NO_2 in the first reaction. Any mercury present at this point in the system is simultaneously removed via the formation of mercuric nitrate due to mercury's ready reaction with nitric acid.

Reactions 1 and 2 begin in the final stage of flue gas cooling but contacting columns are needed to ensure the reactions proceed sufficiently to meet the CO_2 product specification. The flue gas travels up the first contacting column against a stream of descending acid water to completely convert all SO_2 present. Part of the contact liquid is cooled and returned while the remaining part is sent to waste water treatment.

The CO_2 from the top of the first contacting column is the further compressed to 30 bar in an integrally geared compressor and cooled against cooling water before being fed to the bottom of the second contacting column. The contact liquid in this second column is a nitric acid solution which converts the remaining NOx content to nitric acid. As in the first contactor, part of the liquid is recycled and part is sent to waste water treatment. This process is said to remove all of the SO_2 and 90% of the NOx from the flue gas / CO_2 stream.

TSA System

The CO_2 is dried in an adsorbent bed to a dew point of -55°C prior to inerts removal. The bed is regenerated thermally using MP steam from the power island in a cyclical process in which two beds alternate between dehydration mode and regeneration mode.

Auto-refrigerated Inerts Removal

The CO_2 at 30 bar is cooled to -54°C in a series of two multiple pass aluminium plate-fin exchangers contained in a cold box, then flashed to remove the bulk of the inerts. The vapour stream, containing mostly inerts is passed back through the exchangers for cold-recovery, warmed against hot CO_2 in the sour compression section then expanded in a power recover turbine before being vented to atmosphere at a safe location.

The flashed liquid CO_2 is warmed in one of the main exchangers, expanded to 16-17 bar and fed to a distillation column. The vapour product from this column is compressed and recycled to the front of the cold box. The liquid CO_2 product from the distillation column is divided into two streams which are expanded to 5.6 bara and 16-17 bara to provide the cooling required in the main heat exchangers before entering the CO_2 compressor. The lower pressure stream is compressed in an integrally geared compressor up to 16-17 bara then cooled. It then joins the 16-17 bar stream and the combined stream is compressed in two intercooled stages up to 110 bara.

9.4 Technical Performance Evaluation

Table 9-1: Technical Performance Comparison for Case 4

	Units	Reference Case (Unabated CCGT)	Coal Oxy-combustion with CCS
Total Gross Installed Capacity	MWe	1229.4	1112.8
Gas Turbine (s)	MWe	823.5	0
Steam Turbine	MWe	405.9	1097.7
Others	MWe	0	15.1
Total Auxiliary Loads	MWe	20.9	280.2
Feedstock Handling	MWe	0	3.3
Power Island	MWe	14.7	20.9

	Units	Reference Case (Unabated CCGT)	Coal Oxy-combustion with CCS
Air Separation Unit	MWe	0	213.5
CO_2 Capture	MWe	0	26.8
CO_2 Compression	MWe	0	0
Utilities	MWe	6.2	15.7
Net Power Export	MWe	1208.5	832.6
Fuel Flow Rate	kg/h	150,296	325,000
Fuel Flow Rate (LHV)	MWth	1940.2	2335.0
Net Efficiency (LHV) - As New	%	62.3	35.7
Net Efficiency (LHV) - Average	%	59.0	35.5
Total Carbon in Feeds	kg/h	108,640	209,950
Total Carbon Captured	kg/h	0	187,176
Total CO_2 Captured	kg/h	0	685,896
Total CO_2 Emissions	kg/h	398,105	83,455
CO_2 Capture Rate	%	0	89.2
Carbon Footprint	kg CO_2/MWh	329.4	100.2

The plant performance of the supercritical pulverised coal power plant with oxy-combustion carbon capture is summarised in the above table. The unabated CCGT case is also listed in the table for the purposes of comparison. The oxy-fired case captures 89.2% of the CO_2, in keeping with the IEAGHG reference used, while suffering a 26.6% net efficiency loss compared to the unabated gas Reference case.

The following points can be highlighted as basic difference between the two cases:

- The Reference case uses one of the largest and most efficient natural gas fired gas turbines GE 9HA.01 with large power output of > 400 MWe per turbine.
- The supercritical pulverised coal power plant, by comparison, has an efficiency closer to ~ 39% before carbon capture is applied. Thus, the underlying power plant is far less thermally efficient.
- The addition of carbon capture results in additional parasitic load associated with the CO_2 capture and compression process of 240 MWe. This makes the net exportable power from the Oxy-SCPC case 375 MWe less than the Reference case while burning 395 MWth more fuel, thus the net LHV efficiency is 26.6% points lower.
- The carbon efficiency for the Oxy-SCPC case is less than one third that of the Reference unabated case as this case captures 89.2% of the process CO_2 for transportation and storage.

9.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staffs are taken to be in daily positions, working regular hours.

Table 9-2: Operations and Maintenance Staff Manning for Case 4

	Reference Case Unabated CCGT	Coal Oxy-combustion with CCS	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Deputy Plant Manager	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	Daily Position
Process Engineer	NA	3	Daily Position
Shift Supervisor	5	10	3-shift Position
Electrical Assistant	5	5	3-shift Position
Control Room Operator	10	20	3-shift Position
Field Operator	10	40	3-shift Position
Sub-Total	32	81	
Maintenance Staff			
Mechanical Group	3	7	Daily Position
Instrument Group	3	4	Daily Position
Electrical Group	2	4	Daily Position
Sub-Total	8	15	
Laboratory Staff			
Superintendent	1	1	Daily Position
Analysts	3	6	Daily Position
Sub-Total	4	7	
Plant Total Staff	44	103	* See note below

* Note that the IEAGHG 2014/03 report estimated 105 permanent roles in Operations and Maintenance

Table 9-3: Economic Performance Comparison for Case 4

	Units	Reference Case (Unabated CCGT)	Coal Oxy-combustion with CCS
Total Project Cost	£M	672.2	1,901.9
Specific Total Project Cost	£/kW	556	2,284
Pre-Development Costs			
Pre-Licensing & Design	£M	5.8	17.0
Regulatory & Public Enquiry	£M	12.9	35.2
EPC Contract Cost	£M	583.6	1,701.5
Feedstock Handling	£M	0	109.1
Power Island	£M	583.6	774.5
Air Separation Unit	£M	0	353.4

	Units	Reference Case (Unabated CCGT)	Coal Oxy-combustion with CCS
CO_2 Capture	£M	0	197.5
CO_2 Compression	£M	0	0
Utilities	£M	0	267.0
Other Costs			
Infrastructure Connections		29.0	29.0
Owner's Costs		40.9	119.1
Overall CAPEX Impact (vs Ref Case)	£M	-	1,229.7
Overall CAPEX Impact (vs Ref Case)	%	-	183
Total Fixed OPEX	£M pa	36.2	86.8
Total Variable OPEX (excl. Fuel & Carbon)	£M pa	0.2	108.3
Average Fuel Cost	£M pa	315	143
Average CO_2 Emission Cost	£M pa	369	74.8
Total Start-up Cost (excl. Fuel)	£M	4.4	11.8
Discount Rate	% / year	7.8	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	96.0
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	95.1
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	88.0
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	185.5

The economic performance of the Oxy-SCPC is summarised in Table 9-3 along with unabated CCGT case for the purposes of comparison. The capital cost estimate for Case 4 is assessed to have an accuracy of ± 35%.

The total project cost for this case is 183% higher than the Reference unabated natural gas case while producing 31% less net power output:

- The Oxy-combustion system adds to the total project cost with the ASU and CPU both significant cost items, where the CPU includes CO_2 compression. But it should also be noted that the power island and utilities capital cost is much higher than the power island and utilities cost of the CCGT plant. The total effect being a much more capital intensive plant.
- The operating costs are also much higher than the operating costs of the reference natural gas plant, although this is partially related to the study's assumptions on how fixed operating costs are related to the capital cost. The higher CO_2 transportation and storage costs resulting from the larger carbon content of coal compared to natural gas also has a substantial effect on the variable operating costs.
- Both the capital and operating costs (excluding fuel and carbon price) are higher for this case, which compounded with a lower efficiency results in a significantly higher LCOE for this case compared to the unabated case. The LCOE for this case is 15% higher than the LCOE of the coal post-combustion case, as can be seen in more detail in Section 13.

The chart below shows the balance of factors contributing to the overall LCOE. It can be seen that the capital investment is a significantly larger portion of the LCOE than the operating costs, CO_2 transportation and storage and fuel cost.

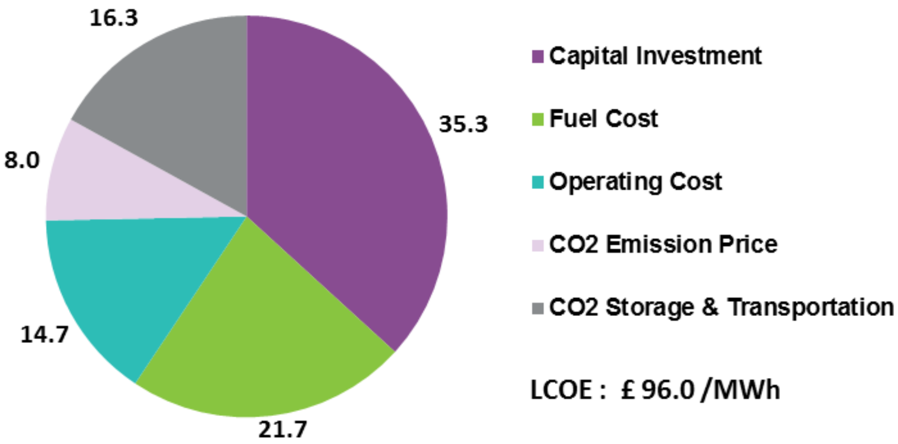

Figure 9-3: LCOE (£/MWh) Contribution for Case 4

9.5.1 Comparison of Results with IEAGHG 2014/03 Report

The value of £ 96.0 / MWh for Levelised Cost of Electricity (LCOE) presented in this report differs marginally from the equivalent result of £ 74.9 / MWh (€ 91.6 / MWh) reported for Case 3 in the IEAGHG Report 2014/03. This results from a variety of differing assumptions used for the two studies, as summarised in Figure 9-4 below.

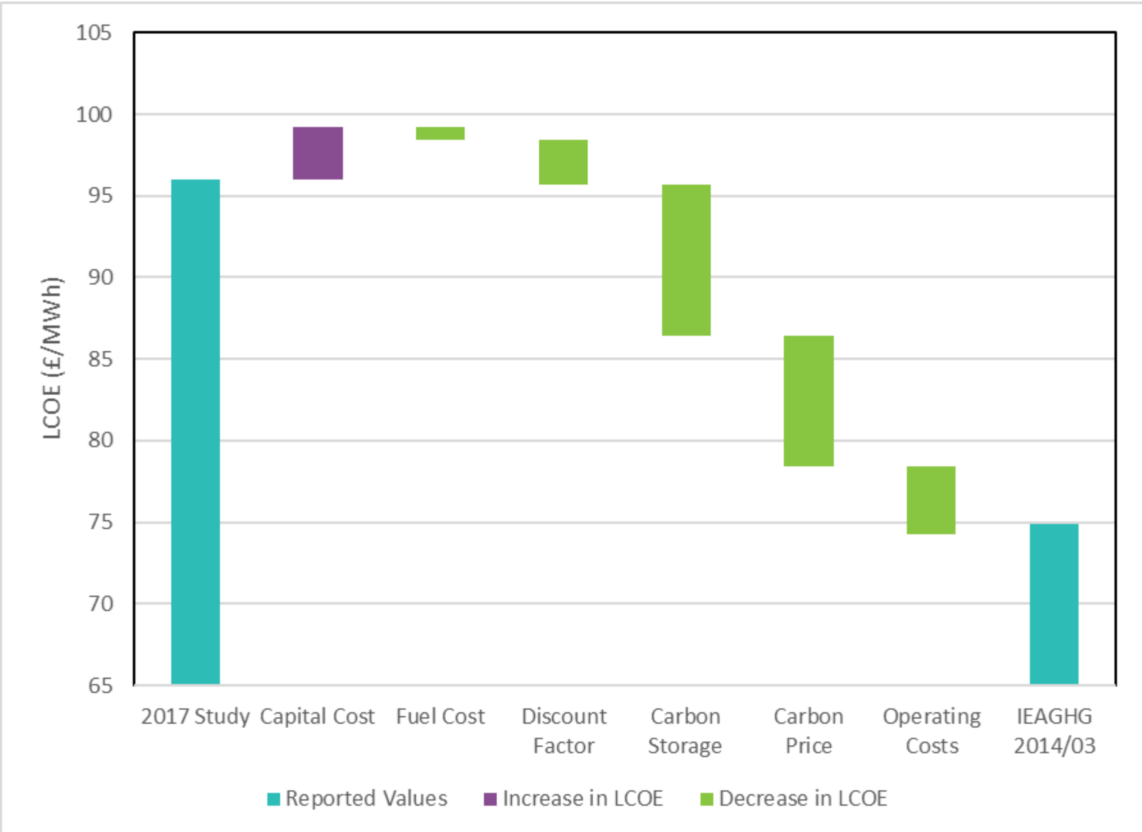

Figure 9-4: Case 4 LCOE Comparison with IEAGHG 2014/03 Report

The capital cost estimating methodology adopted in 2014 resulted in a 10% higher EPC cost, but excluded infrastructure connections and pre-development costs. Taking these into account within the current evaluation would increase the LCOE by a small margin.

The coal price used in the 2014 study was € 2.5 / GJ (£ 7.4 / MWh). This is about 20% higher than the coal price for the start-up year of 2025, but is 9% lower than the cost used for 2030 onwards (refer to Table 5-7). Overall the cost impact between the study results is neutral.

The 2014 study used a slightly lower discount factor of 8.0% versus 8.9% in the current study. It also assumed lower costs for CO_2 transportation and storage (€ 10 / tCO_2) and excluded the price of carbon emissions entirely. These all contribute to a higher LCOE in this study than in the 2014 study.

A combination of small differences in the estimation of operating costs has a significant effect upon the LCOE for the two studies. The 2014 study assumed much lower costs for Maintenance, Insurance and Local Taxes, although the allowance for Administration and General Overheads was higher.

10 Case 5 – Coal IGCC with Pre-Combustion Carbon Capture

10.1 Overview

10.1.1 Gasification Process Selection

This case consists of an integrated gasification combined cycle power plant with carbon capture based on the Shell gasification technology. The plant is designed to process coal to produce electric power for export.

There are other gasification technologies suitable for IGCC application, namely GE oxy-gasification process with radiant syngas cooler and MHI air-blown gasification process. In terms of commercial application, MHI gasification technology has built a pilot plant and one demonstration plant in Japan. Whereas, Shell and GE gasification technology are well established and have been demonstrated at various capacities in different parts of the world.

The Shell Coal Gasification Process (SCGP) uses a dry feed of pulverised coal and offers feed flexibility. The process has been proven on a wide variety of solid fuels, ranging from bituminous coal to lignite, as well as petroleum coke in a coal mix. The process is capable of handling coal feed with biomass / sewage sludge. The reference list for the Shell gasification process includes 22 facilities for various applications such as power, hydrogen, ammonia and methanol plants. This, shows that while there are relatively few operating examples of IGCC plants, Shell gasification plants are much more widespread than IGCC plants and are very well demonstrated.

The GE gasification system uses slurry coal feed. This system is applicable to other solid feeds such as petcoke, asphalt, heavy oil, vacuum residue, etc. The coal grinding and slurry preparation system is dedicated to the preparation of the coal slurry feed for the gasifier. The number of facilities using the GE gasification system for solid fuel application is 21 mainly for ammonia, methanol, and synthetic natural gas production plants. As for the Shell process, this shows that there are many operating examples of GE gasification plants beyond the IGCC application of the technology.

It can be summarised that both Shell and GE gasification system are well placed to be used as the gasification island basis for the study. However, as mentioned above, the Shell gasification process is capable of using biomass in a fuel-mix to the gasifier. Therefore, using the Shell gasifier for the evaluation will provide a consistent basis for both the coal IGCC case and biomass IGCC case. On that basis, Shell coal gasification process (SCGP) has been chosen for this study.

10.1.2 Process Configuration

The main process configuration of the IGCC plant is as follows:

- Medium-pressure (40 barg) Shell Coal Gasification Process (SCGP);
- Hybrid CO shift followed by two stages of sour shift;
- Acid gas removal (H_2S and CO_2) using Selexol physical solvent system;
- Sulphur recovery based on Claus process;
- CO_2 compression and pumping up to 110 bara;
- Combined cycle based on two GE F-class syngas variant gas turbines.

Table 10-1 describes the process units with trains which are also shown in Figure 10-1.

Table 10-1: IGCC Process Units with Trains

Unit Number	Unit Description	Trains
100	Coal / Limestone I Handling & Storage	N/A
200	Shell Coal Gasification Package (SCGP)	2 x 50%
300	Air Separation Unit (ASU)	2 x 50%
400	Syngas Treatment & Conditioning	2 x 50%
500	Acid Gas Removal (AGR)	2 x 50%
600	Sulphur Recovery Unit (SRU) & Tail Gas Treatment	2 x 50%
700	CO_2 Compression & Dehydration	2 x 50%
800	Gas Turbine & Generator Package	2 x 50%
900	Heat Recovery Steam Generation	2 x 50%
1000	Steam Turbine & Generator Package	2 x 50%
1100	Offsite & Utilities	

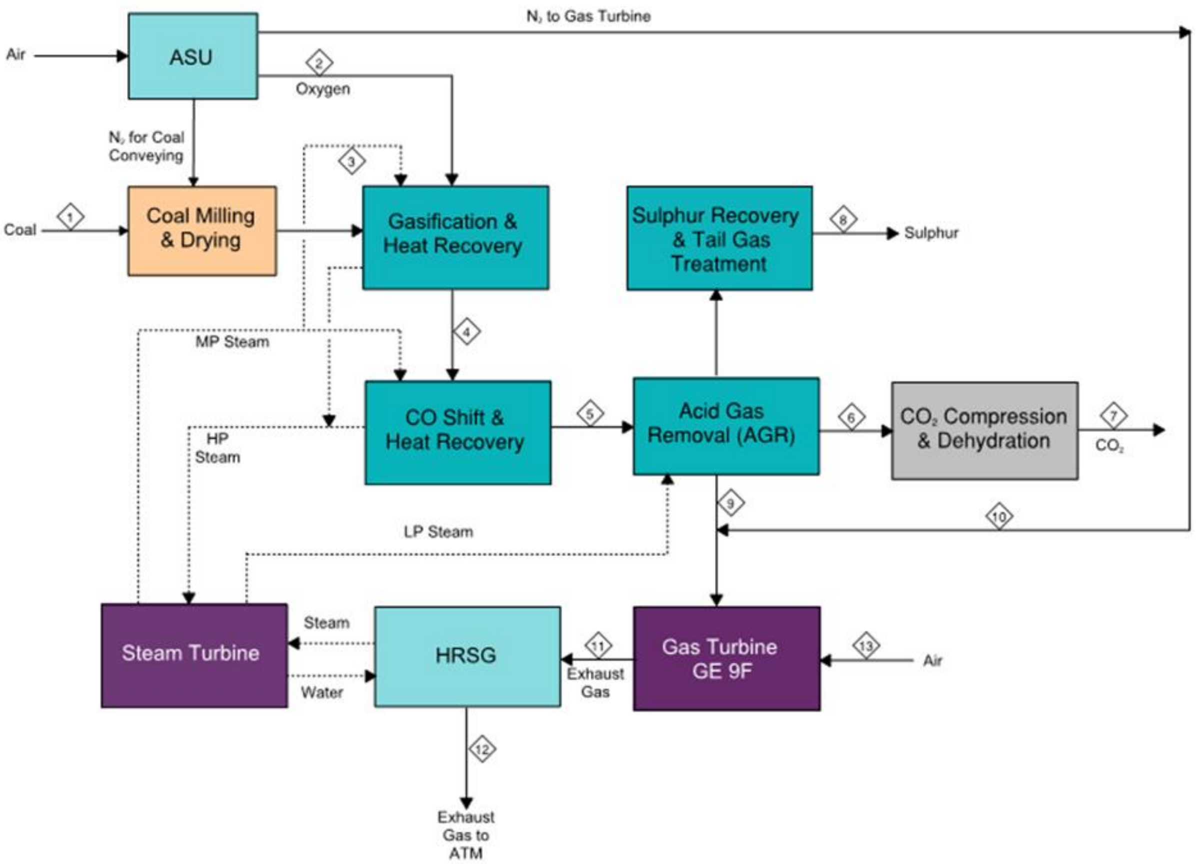

Figure 10-1: Case 5 Block Flow Diagram

10.2 Process Model Development

The coal flow rate to the IGCC Complex is governed by the thermal requirements of two commercially available GE F-class syngas variant gas turbines in the combined cycle, at the reference ambient temperature of the study.

The Coal IGCC using Shell gasification technology has been simulated in Aspen Hysys using a similar process flow scheme to that given in the reference IEAGHG 2014 Coal and Hydrogen with CCS report. The Peng Robinson property package has been used to model the process scheme, whereas the water and steam cycle has been modelled using the NBS Steam property package.

The IEAGHG report's coal specification, which is an appropriate coal for a UK study as well, was used in the model along with the same feed rate. The HP, MP and LP steam conditions and the process pressure profile are in line with the IEAGHG report. The heat integration between the process units and combined cycle has been developed to minimise the heat loss from the system and maximise power output. Syngas variant gas turbine technical data used for this case is based on supplier data for an equivalent gas turbine to that used in the IEAGHG report.

The data from the model are in good agreement with the IEAGHG report in terms of the syngas composition, steam production and parasitic load.

10.3 Process Description

10.3.1 Coal / Limestone Storage and Handling

The unit prepares and stores the coal feed and limestone fluxant delivered to the plant. The coal and limestone storage capacity is designed to hold an inventory of 30 days of design consumption to neutralise any delivery disruptions.

The coal feeding system, from the storage building to the gasification island, consists of conveyors, elevated feed hoppers, crushers, magnetic plate separators, and day silos. The crushers are designed to break down big lumps of coal to a size not exceeding 35 mm. Coal from the crushers is transferred by enclosed belt conveyors to the day silos which are close to the gasification island. Magnetic plate separators remove tramp iron from crushed coal. Sampling systems analyse both the as-received coal and the as-fired coal to ensure the reliable and efficient operation of the plant. To control the plant environmental emissions, all equipment is connected to bag filters and exhaust fans that permit the capture of any coal powder generated in the coal handling area.

The limestone transport system to the gasification island consists of similar equipment to the coal handling. Limestone is added to the coal feed before being fed to the mills for pulverisation.

10.3.2 Shell Coal Gasification Package (SCGP)

This unit is mainly composed of following processes:

- Coal Milling and Drying;
- Coal Pressurisation and Feeding;
- Gasification and Syngas Cooler;
- Slag Removal;
- Dry Solids Removal;
- Wet Scrubbing;
- Primary Waste Water Treatment.

The SCGP scope includes all process units listed above except Coal Milling and Drying. In order to meet the Shell SCGP specification for particle size distribution and moisture content, the coal needs to be milled and dried. The key features of the Shell Coal Gasification Process (SCGP) are the following:

- Pressurised system with compact equipment;
- Entrained flow slagging gasifier;
- Oxy-steam gasification leading to high gasification efficiency;
- Multiple burner design providing good mixing, high conversion, scale-up possibility;

- Dry feed of pulverised coal providing high gasification efficiency, high feed flexibility.

Gasification

The coal feed is gasified in the entrained flow slagging gasifier using a mixture of oxygen and superheated process steam. Due to the entrained flow, high temperature, and ash slagging condition, an almost complete carbon conversion (>99%) is achieved. 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. Slag also forms a protective layer on the gasification membrane wall providing insulation, minimising heat losses and protecting the gasifier wall against high heat load variations during process upsets.

The operating syngas pressure of the gasifier is about 40 bar. Hot syngas from the gasifier is initially quenched with recycle syngas to approximately 800°C. The combined gas stream is then cooled in a syngas cooler to generate HP & MP steam. The syngas cooler is of the water pipe type, typically containing both evaporating and superheating surfaces.

Slag Removal

Depending on the ash content in the coal, about 70-80% of the mineral content of the coal / fresh ash leaves the gasification zone in the form of molten slag. Limestone is used as an additive to the coal feed, which acts as a moderator affecting the ash fusion temperature of coal in order to ensure that the slag flows freely down the membrane wall. The heat from the molten slag is removed in the slag bath. The slag is non-leachable and non-hazardous.

Dry Solids Removal & Wet Scrubbing

The system consists of high pressure high temperature (HPHT) filter system that removes 99.9% of the entrained solids in the syngas stream. The gas leaving the dry solids removal is further processed in a wet scrubbing system which consists of a venturi scrubber and a packed bed wash column. Residual solids, as well as halide contents of the syngas is reduced to <<1 ppmv.

Primary Waste Water Treatment

The primary waste water treatment system contains one slurry stripper and a solid / liquid separation step. The system treats the bleed from slag bath and wet scrubbing systems.

10.3.3 Air Separation Unit (ASU)

The ASU capacity is defined by oxygen requirements for the IGCC Complex, mainly for the gasification process, plus the consumption of the Sulphur Recovery Unit (SRU). It supplies 95 mole% oxygen to the gasification island and the SRU. The total required oxygen flow rate for this case is approximately 250 t/h.

The ASU also produces very high pressure and medium pressure nitrogen. The very high pressure nitrogen is for the gasification system, acting as carrier gas for the pulverised coal feed pneumatic transport system, whereas medium pressure nitrogen is used as diluent for the syngas to the gas turbine for NOx control.

10.3.4 Syngas Treatment & Conditioning

Saturated raw syngas from the gasification wet scrubber unit, at approximately 40 barg, passes through the series of shift reactors where CO is shifted to H_2 and CO_2; and COS is converted to H_2S. A hybrid water gas shift (WGS) scheme has been used for this study. The scheme is used to minimise the steam consumption and amount of condensate flowing to the sour water stripper, achieving an overall CO conversion greater than 98%.

The first WGS reactor is a low steam shift reactor, converting about 35% of CO to CO_2. The catalyst for this reactor is designed to minimise the unwanted methanation reaction. This is followed by a conventional 2-stage sour shift reactor to convert the remaining CO. The hot shifted syngas from both the second and third shift reactors is cooled down in a series of heat exchangers for heat recovery steam generation. Final cooling of the syngas is made against clean syngas coming from the Acid Gas Removal (AGR) unit followed by cooling water, before passing through a sulphur impregnated activated carbon bed to remove approximately 95% of the mercury. Cool, mercury-depleted syngas then enters the AGR unit.

Process condensate from syngas cooling is sent to the Sour Water Stripper in order to avoid accumulation of ammonia and H_2S and other dissolved gases in the water recycle to the gasification section. Part of the condensate from the accumulator is sent to the Gasification Island, while the remaining condensate is sent to the Waste Water Treatment Unit.

10.3.5 Acid Gas Removal (AGR)

The AGR Unit removes the H_2S and CO_2 from the shifted syngas by using Selexol as a physical solvent. Shifted syngas combined with the recycle stream from the Tail Gas Recovery Unit passes through a H_2S Absorber followed by three parallel CO_2 absorbers where H_2S and CO_2 are removed from the gas by Selexol solvent. Solvent regeneration is accomplished in the regenerator, where H_2S and CO_2 are stripped from the liquid phase to the gas phase by steam. The CO_2 removal rate is approximately 92% of the carbon dioxide entering the unit, reaching an overall carbon capture of approximately 90%.

The AGR is designed to meet the following process specifications of the treated gas and of the CO_2 product exiting the unit:

- The H_2S+COS concentration of the treated gas exiting the unit is ~ 1 ppmv;
- The CO_2 product should meet the specification of inerts around 4%, H_2S content lower than 20 ppmv and CO content lower than 0.2% mol;
- The acid gas H_2S concentration is about 35% dry basis, suitable to feed the oxygen blown Claus process.

10.3.6 Sulphur Recovery Unit (SRU) & Tail Gas Treatment

The H_2S-rich acid gas from the AGR is treated in the SRU, where H_2S is converted into elemental sulphur using low pressure oxygen from the ASU. 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 quenched with process water, compressed and recycled back to the inlet of AGR absorber. The overall sulphur production rate is approximately 65 tonnes per day.

10.3.7 CO_2 Compression & Dehydration

This unit is mainly composed of a compression and dehydration package, followed by final stage CO_2 pumps, supplied by specialised vendors. Three different streams of CO_2 from the AGR unit are routed to the CO_2 compression unit, where it is initially compressed to ~30 bara and dried to <50 ppmv water using a molecular sieve adsorption process. After dehydration, the CO_2 stream is compressed to a supercritical condition at 80 bara. The resulting stream of CO_2 is pumped to the required pressure of 110 bara ready for transportation.

10.3.8 Combined Cycle

The combined cycle uses two state-of-the-art GE 9F, 50 Hz syngas variant gas turbines, commercially available for high hydrogen content gas. Due to high flame speed (flash back risk) and lower auto-ignition delay time for hydrogen compared to natural gas. The pre-mix burner which is normally used for the commercially available natural gas fired gas turbine can't be used for the syngas variant gas turbine. Also, the combustion of hydrogen-rich fuel leads to a high flame temperature and consequent high thermal NOx formation. Fuel dilution to the gas turbine is therefore necessary to meet the NOx emission limits. Hence, for gas turbines firing syngas and high hydrogen gas with lower LHV (on volumetric basis), significant design changes have been adopted from the conventional natural gas fired gas turbines.

For syngas and high hydrogen gas, diffusion burners are used in place of the pre-mix burner and control of NOx is achieved by diluting the fuel with large quantity of nitrogen (nearly 5:1 of N_2:H_2

based syngas on a mass basis). In addition, saturated nitrogen is injected directly into the gas turbine combustion chamber for final dilution to moderate the high flame temperature.

The decarbonised fuel gas is preheated in the syngas / syngas exchanger and against LP steam after being mixed with nitrogen, coming from the ASU, up to maximum hydrogen content of 65 mole%. The gas turbine compressors provide combustion air to the burner only, i.e. no air integration with the ASU is foreseen. The exhaust gases from the gas turbine enter the HRSG at 560°C. The HRSG recovers heat available from the exhaust gas, producing steam at three different pressure levels for the steam turbine. The final exhaust gas temperature to the stack of the HRSG is ~80°C. The combined cycle is thermally integrated with the process unit, in order to maximise the net electrical efficiency of the plant.

10.4 Technical Performance Evaluation

Table 10-2: Technical Performance Comparison for Case 5

	Units	Reference Case (Unabated CCGT)	Coal IGCC with CCS
Total Gross Installed Capacity	MWe	1229.4	1062.8
Gas Turbine (s)	MWe	823.5	671.0
Steam Turbine	MWe	405.9	391.7
Others	MWe	0	0
Total Auxiliary Loads	MWe	20.9	263.0
Feedstock Handling	MWe	0	0.4
Power Island	MWe	14.7	12.9
Air Separation Unit	MWe	0	106.4
CO_2 Capture	MWe	0	87.0
CO_2 Compression	MWe	0	45.7
Utilities	MWe	6.2	10.6
Net Power Export	MWe	1208.5	799.8
Fuel Flow Rate	kg/h	150,296	314,899
Fuel Flow Rate (LHV)	MWth	1940.2	2262.9
Net Efficiency (LHV) - As New	%	62.3	35.3
Net Efficiency (LHV) - Average	%	59.0	33.5
Total Carbon in Feeds	kg/h	108,640	203,425
Total Carbon Captured	kg/h	0	183,697
Total CO_2 Captured	kg/h	0	673,147
Total CO_2 Emissions	kg/h	398,105	72,292
CO_2 Capture Rate	%	0	90.3
Carbon Footprint	kg CO_2/MWh	329.4	90.4

The plant performance of the full scale coal IGCC plant with carbon capture is summarised in the above table. The overall performance of the system also includes the CO_2 balance and removal efficiency. The unabated CCGT Reference case is also listed in the table for the purposes of comparison.

The following points can be highlighted as basic difference between the two cases:

- The Reference case uses one of the largest and most efficient natural gas fired gas turbines GE 9HA.01 with large power output of > 400 MWe per turbine.
- IGCC case uses a GE Frame 9 syngas variant gas turbine fired by syngas produced from coal gasification. The gas turbine efficiency is ~ 42% with the gross power output 336 MWe per turbine.
- The IGCC case suffers from a large parasitic energy demand associated with the coal milling / drying, gasification island, CO_2 capture and compression process. This makes the net exportable power from the IGCC plant 408 MWe less than the Reference case leading to 27% lower net LHV efficiency.
- However, the carbon efficiency for the IGCC case is less than one third of the Reference unabated case, as the IGCC case captures over 90% of the process CO_2 for transportation and storage.

10.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staffs are taken to be in daily positions, working regular hours.

Table 10-3: Operations and Maintenance Staff Manning for Case 5

	Reference Case Unabated CCGT	Coal IGCC with CCS	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Deputy Plant Manager	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	Daily Position
Process Engineer	NA	5	Daily Position
Shift Supervisor	5	10	3-shift Position
Electrical Assistant	5	10	3-shift Position
Control Room Operator	10	25	3-shift Position
Field Operator	10	50	3-shift Position
Sub-Total	32	103	
Maintenance Staff			
Mechanical Group	3	8	Daily Position
Instrument Group	3	8	Daily Position
Electrical Group	2	6	Daily Position
Sub-Total	8	22	
Laboratory Staff			
Superintendent	1	1	Daily Position
Analysts	3	7	Daily Position
Sub-Total	4	8	
Plant Total Staff	44	133	* See note below

* Note that the IEAGHG 2014/03 report estimated 133 permanent roles in Operations and Maintenance

Table 10-4: Economic Performance Comparison for Case 5

	Units	Reference Case (Unabated CCGT)	Coal IGCC with CCS
Total Project Cost	£M	672.2	2,396.3
Specific Total Project Cost	£/kW	556	2,996
Pre-Development Costs			
Pre-Licensing & Design	£M	5.8	21.5
Regulatory & Public Enquiry	£M	12.9	44.2
EPC Contract Cost	£M	583.6	2,151.0
Feedstock Handling	£M	0	71.1
Power Island	£M	583.6	1,349.6
Air Separation Unit	£M	0	240.4
CO_2 Capture	£M	0	82.4
CO_2 Compression	£M	0	67.1
Utilities	£M	0	340.4
Other Costs			
Infrastructure Connections		29.0	29.0
Owner's Costs		40.9	150.6
Overall CAPEX Impact (vs Ref Case)	£M	-	1,724.1
Overall CAPEX Impact (vs Ref Case)	%	-	256
Total Fixed OPEX	£M pa	36.2	112.2
Total Variable OPEX (excl. Fuel & Carbon)	£M pa	0.2	103.5
Average Fuel Cost	£M pa	315	131
Average CO_2 Emission Cost	£M pa	369	61.2
Total Start-up Cost (excl. Fuel)	£M	4.4	16.8
Discount Rate	% / year	7.8	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	120.8
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	195.1
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	113.3
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	283.8

The economic performance of the full scale coal IGCC system with carbon capture is summarised in the table above, along with unabated CCGT Reference case for the purposes of comparison. The capital cost estimate for Case 5 is assessed to have an accuracy of ± 35%.

The total project cost for the IGCC case is 3.5 times as high as the Reference unabated case. The following points highlight the difference between the cases:

- The Power Island cost for the IGCC includes coal handling, Shell gasification island, CO shift reactors and combined cycle costs. Overall this makes the IGCC power island cost 2.3 times as higher as the Reference case, which includes only the combined cycle system.
- The carbon capture and compression cost for the IGCC case contributes to the ~£150 M, which is not required for the unabated case.
- Predevelopment costs and Owners' cost are higher for the IGCC case, since these are calculated as a percentage of the EPC contract value.
- The operating costs are also much higher than the operating costs of the reference natural gas plant, although this is partially related to the study's assumptions on how fixed operating costs are related to the capital cost. The higher CO_2 transportation and storage costs resulting from the larger carbon content of coal compared to natural gas also has a substantial effect on the variable operating costs.

The increased project cost, operating cost and cost related to the CO_2 transportation & storage makes the Levelised Cost of Electricity for the IGCC system higher than Reference case. Figure 10-2 shows the different contributing factors and the level of contribution towards the overall LCOE.

The capital investment cost is the biggest contributor to the LCOE followed by operating cost. Fuel costs and the cost for Transportation and storage of carbon dioxide also have significant LCOE contributions.

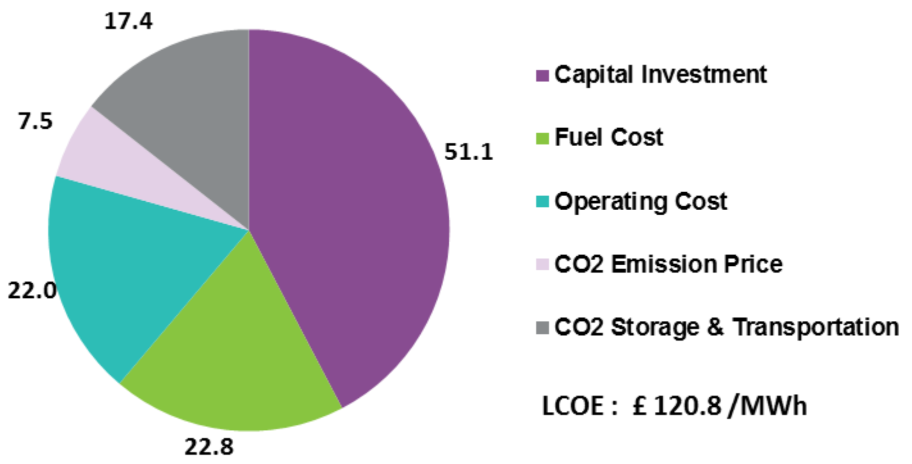

Figure 10-2: LCOE (£/MWh) Contribution for Case 5

10.5.1 Comparison of Results with IEAGHG 2014/03 Report

The value of £ 115.9 / MWh for Levelised Cost of Electricity (LCOE) presented in this report differs from the equivalent result of £ 95.2 / MWh (€ 116.5 / MWh) reported for Case 4.1 in the IEAGHG Report 2014/03. This results from a variety of differing assumptions used for the two studies, as summarised in Figure 10-3 below.

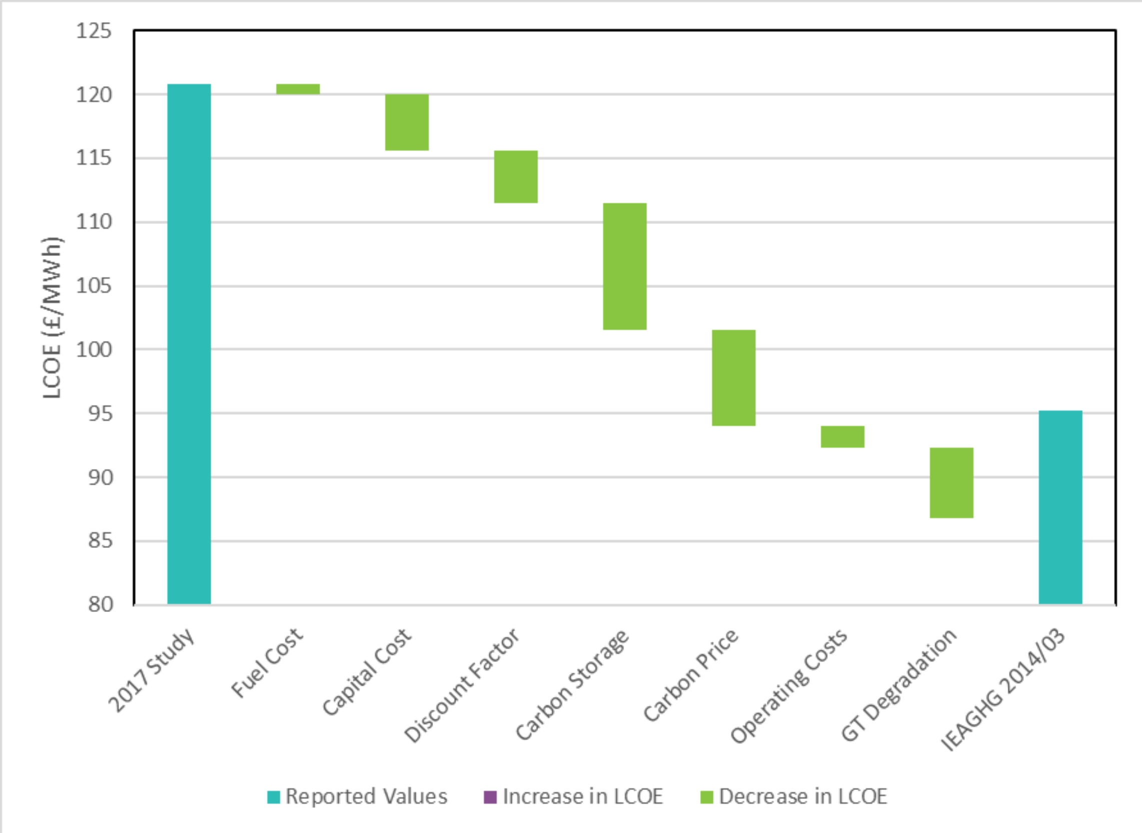

Figure 10-3: Case 5 LCOE Comparison with IEAGHG 2014/03 Report

The coal price used in the 2014 study was € 2.5 / GJ (£ 7.4 / MWh). This is about 20% higher than the coal price for the start-up year of 2025, but is 9% lower than the cost used for 2030 onwards (refer to Table 5-7). Overall the cost impact between the study results is neutral.

The capital cost estimating methodology adopted in 2014 resulted in a slightly lower EPC cost, but also excluded infrastructure connections and pre-development costs. Using the 2014 capital cost estimate in the current analysis would result in a small drop in the LCOE.

The 2014 study used a slightly lower discount factor of 8.0% versus 8.9% in the current study. It also assumed lower costs for CO_2 transportation and storage (€ 10 / tCO_2) and excluded the price of carbon emissions entirely. These all contribute to a higher LCOE in this study than in the 2014 study.

Although the assumptions used for estimating operating costs differed between the 2014 IEAGHG report and the current analysis, the net effect is marginal. Using the 2014 basis for operating costs would result in a small drop in LCOE.

The LCOE is also higher in the current study because the lifetime degradation in gas turbine performance has been accounted for.

This case results in the largest delta between the 2017 LCOE and how the results would appear if the 2014 basis were used. However, the sum of the individual deltas shown above does not equal the compound impact. When all of the individual elements are combined within our study matrix, the overall result is slightly higher than the IEAGHG result at £ 97.6 / MWh.

11 Case 6 – Oxy-fired Supercritical Gas Power Generation with Carbon Capture

11.1 Overview

The process used in this case is a novel oxy-combustion cycle that combusts natural gas with oxygen at high pressure and temperature using the hot combustion products to drive a turbine in a novel thermodynamic cycle called the Allam Cycle. The Allam Cycle uses novel supercritical CO_2 combustion technology. NET Power LLC has developed the proprietary Allam Cycle which produces electricity and high pressure pipeline quality CO_2 by-product from fossil fuel at a cost and efficiency that they claim is comparable with the current power generation systems without CO_2 capture. Toshiba Corporation, Exelon Corporation and CB&I are partnering with NET Power LLC to commercialize the system by developing a 50MWth plant which is under development. Some of the bespoke items used for the cycle have been developed specifically for the application in conjunction with the equipment suppliers like Toshiba and Heatric.

In this study, NET Power technology has been techno-economically assessed. Net Power is very sensitive about their technology and intellectual property. No information has been obtained from the technology developer related to the process for this study. Hence, the technical basis for the development of the Net Power cycle has been derived entirely from public domain data.

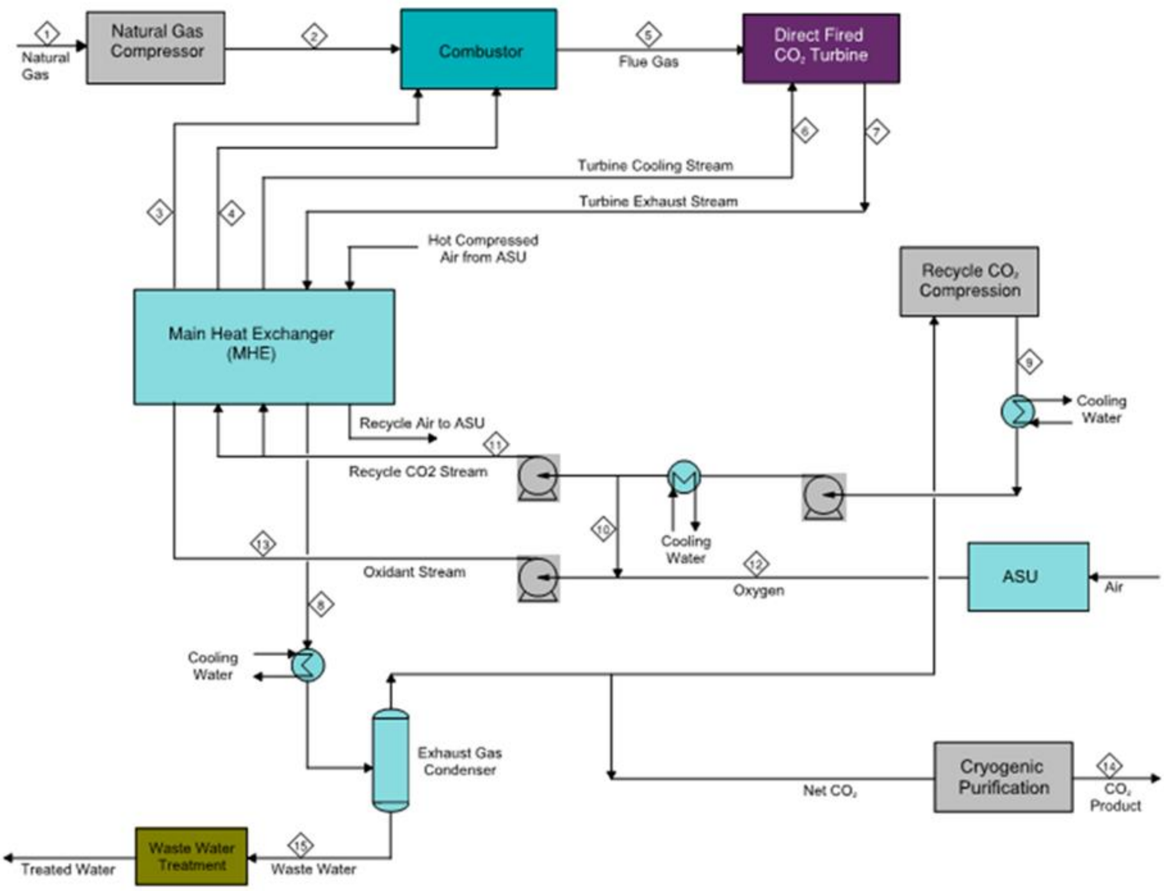

Figure 11-1: Case 6 Block Flow Diagram

11.2 Model Development

IEAGHG 2015/05 report on 'Oxy-Combustion Turbine Power Plants' includes a high-level narrative of the design and novelty of the process from NET Power LLC without disclosing any precise efficiency data and design consideration of the bespoke items. In absence of any input from technology developer, the process modelling of Allam cycle developed for this study is entirely based on the public domain information and design assumptions to achieve a reasonable representation of the process.

Net Power together with EPC partner CB&I performed a design, engineering and costing of a commercial scale plant based on 500 MWth fuel input to the system. This is stated by Net Power in the IEAGHG 2015-05 Oxy-Combustion Turbines Report in Chapter D2-Case 2 Net Power Commentary. For this study, we have used this capacity as a basis for commercial plant. Hence 3 x 500 MWth trains has been considered for the model development.

The process was modelled using Hysys as Gatecycle is not ideal for modelling CO_2-rich oxidant systems. The Peng Robinson property package has been used to model the process scheme whereas the water and steam cycle has been modelled using NBS Steam property package.

The purpose of the modelling in this study is to develop the heat and material balance and determine the overall efficiency of the plant which was verified against reference data in IEAGHG 2015/05 report. The model data also used to develop sized equipment list. It is not intended to model the packaged units other than for high level thermodynamic and material balance checking. The CO_2-rich recycle stream, oxidant stream and main heat exchange streams matched closely with the reference data without any significant differences in terms of achievable temperatures and flowrates. Overall plant technical performances are in good agreement with the reference data.

11.3 Process Description

Allam Cycle is a closed loop oxy-fuel thermodynamic power cycle, which uses supercritical CO_2 as the working fluid and captures almost all of the CO_2 for sequestration at a pressure and quality ready for transportation. It is a high-pressure, low-pressure ratio Brayton cycle, operating with a single turbine. Oxy-fuel combustion offers advantages over post combustion while coupling with carbon capture due to mainly eliminating diluent nitrogen from the process stream and producing a CO_2-enriched flue gas ready for sequestration.

Figure 11-1 shows the simplified block flow diagram of the Allam cycle using natural gas as fossil fuel. The overall scheme can be described as following units:

- Power cycle
- Air separation unit (ASU)
- CO_2 purification unit (CPU)
- Utility system

The overall scheme has three trains of power cycle facilities using 3 x 500MWth of natural gas fuel, with a common cryogenic purification process to meet CO_2 product purity spec.

11.3.1 Power Cycle

Each power cycle unit includes following main process equipment:

- High pressure combustor
- Direct-fired CO_2 turbine expander
- Main heat exchanger (MHE)
- Recycle CO_2 compressor
- CO_2-rich stream pump
- Oxidant pump

Natural gas feed (500MWth) to the process boundary is compressed up to 305 bara before being injected in the high pressure oxy-fuel combustor (~300 bara). Natural gas is combusted with CO_2-rich recycle stream and oxidant streams providing a high pressure feed stream to a direct-fired CO_2

turbine. The recycle CO_2-rich stream to the combustor is required to control the combustor outlet temperature around 1150 °C.

The turbine operates with an inlet pressure of 300 bara and with pressure ratio of ~9. In order to maintain turbine blade metal integrity, a part of the turbine exhaust gas has been recycled via main heat exchanger and used as a cooling medium to lower the metal temperature below 850 °C.

The hot turbine exhaust gas at 34 bara and 740 °C enters in the economiser heat exchanger (MHE). The heat available in the exhaust gas is used to heat three process streams:

- CO_2-rich stream recycled to the combustor
- Oxidant stream to the combustor
- CO_2-rich stream used for turbine blade cooling

There is also heat integration between main heat exchanger and hot compressed air stream from ASU. This improves the overall heat balance and process efficiency.

Most of the turbine exhaust gas is being recycled within the system whereas the net product stream is being transferred to the CO_2 purification system. CO_2-rich process stream is compressed up to 80 bara in a four-stage intercooled compressor before being pumped to 120 bara. The stream is then divided into three streams as described below:

- ~59% of the CO_2-rich gas is pumped up to 305 bara before being split into two streams: recycle stream to the combustor (~48%) and turbine blade cooling stream (~11%)
- Rest is mixed with pure oxygen from ASU resulting a oxidant stream and then pumped in a separate pumping stage up to 305 bar and heated in MHE before being fed to the combustor

11.3.2 Air Separation Unit

Cryogenic air separation has been assumed due to the requirement for high purity O_2 (99.5 mol%) for the process. The system is designed to produce relativity high pressure O_2 at 120 bara. Three ASU units have been assumed which are linked with three power cycles.

The cryogenic ASU has been integrated with the process to provide an external source of heat by eliminating the inter-cooling between air compression stages and exchanging the adiabatic heat of compression with the process.

11.3.3 CO_2 Purification & Separation Unit

In order to reduce the level of oxygen in the CO_2 product to within the specified 100 ppmv limit, a Cryogenic Purification Unit (CPU) has been added in the CO_2 product line. The CPU takes a side-stream of the recirculating CO_2 at 30 bara and purifies and pressurises it separately to produce a product stream at 110 bara.

11.4 Technical Performance Evaluation

Table 11-1: Technical Performance Comparison for Case 6

	Units	Reference Case (Unabated CCGT)	Oxy-Combustion (Allam Cycle)
Total Gross Installed Capacity	MWe	1229.4	1263.9
Gas Turbine (s)	MWe	823.5	1263.9
Steam Turbine	MWe	405.9	0
Others	MWe	0	0
Total Auxiliary Loads	MWe	20.9	415.5
Feedstock Handling	MWe	0	0.0

	Units	Reference Case (Unabated CCGT)	Oxy-Combustion (Allam Cycle)
Power Island	MWe	14.7	13.9
Air Separation Unit	MWe	0	170.9
CO_2 Capture	MWe	0	220.4
CO_2 Compression	MWe	0	Incl. with CO_2 Capture
Utilities	MWe	6.2	10.4
Net Power Export	MWe	1208.5	848.4
Fuel Flow Rate	kg/h	150,296	118,940
Fuel Flow Rate (LHV)	MWth	1940.2	1536.3
Net Efficiency (LHV) - As New	%	62.3	55.2
Net Efficiency (LHV) - Average	%	59.0	52.3
Total Carbon in Feeds	kg/h	108,640	85,975
Total Carbon Captured	kg/h	0	77,378
Total CO_2 Captured	kg/h	0	283,546
Total CO_2 Emissions	kg/h	398,105	31,503
CO_2 Capture Rate	%	0	90.0
Carbon Footprint	kg CO_2/MWh	329.4	37.1

The plant performance of the full scale Net Power system using natural gas fuel and including a CPU is summarised in the above table. The overall performance of the Allam cycle also includes CO_2 balance and removal efficiency. The unabated CCGT Reference case is also listed in the table for the purposes of comparison.

The Allam cycle captures 90% of the CO_2; however, it suffers from a 7.1% point net efficiency loss. The following points can be highlighted as basic difference between the Allam cycle and the Reference case:

- The power island configuration for Reference Case is two of the largest and most efficient natural gas fired gas turbine GE 9HA.01 with large power output of ~ 400 MWe per turbine followed by two steam turbines. The gas turbine efficiency is ~ 43.6% with the combined cycle plant gross efficiency ~65% (GTW 2104-15).
- The Net Power system uses three direct fired CO_2 turbines. This configuration increases the gross power output from Net Power scheme by approx. 3% compared to the CCGT Reference scheme.
- However, the Net Power system suffers from a large parasitic demand associated with the inherent high pressure CO_2 recycling and ASU. This makes the net exportable power approx. 30% lower compared to the Reference Case leading to 7.1% lower net LHV efficiency.
- The carbon footprint for the Net Power case is ~ 9 times lower than the Reference unabated case as Net Power case captures 90% of the process CO_2 for transportation and storage.

11.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-

hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staffs are taken to be in daily positions, working regular hours.

Table 11-2: Operations and Maintenance Staff Manning for Case 6

	Reference Case Unabated CCGT	Oxy-Combustion (Allam Cycle)	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Deputy Plant Manager	1	1	Daily Position
CO_2 Removal Area Manager	NA		Daily Position
Process Engineer	NA	1	Daily Position
Shift Supervisor	5	5	3-shift Position
Electrical Assistant	5	5	3-shift Position
Control Room Operator	10	15	3-shift Position
Field Operator	10	15	3-shift Position
Sub-Total	32	43	
Maintenance Staff			
Mechanical Group	3	3	Daily Position
Instrument Group	3	3	Daily Position
Electrical Group	2	2	Daily Position
Sub-Total	8	8	
Laboratory Staff			
Superintendent	1	1	Daily Position
Analysts	3	3	Daily Position
Sub-Total	4	4	
Plant Total Staff	44	55	* See note below

* Note that the IEAGHG 2015/05 report estimated 82 permanent roles in Operations and Maintenance

Table 11-3: Economic Performance Comparison for Case 6

	Units	Reference Case (Unabated CCGT)	Oxy-Combustion (Allam Cycle)
Total Project Cost	£M	672.2	1,213.2
Specific Total Project Cost	£/kW	556	1,430
Pre-Development Costs			
Pre-Licensing & Design	£M	5.8	10.7
Regulatory & Public Enquiry	£M	12.9	22.9
EPC Contract Cost	£M	583.6	1,067.9
Feedstock Handling	£M	0	0
Power Island	£M	583.6	565.8
Air Separation Unit	£M	0	284.3

	Units	Reference Case (Unabated CCGT)	Oxy-Combustion (Allam Cycle)
CO_2 Capture	£M	0	45.8
CO_2 Compression	£M	0	0
Utilities	£M	0	172.0
Other Costs			
Infrastructure Connections		29.0	37.0
Owner's Costs		40.9	74.7
Overall CAPEX Impact (vs Ref Case)	£M	-	541.0
Overall CAPEX Impact (vs Ref Case)	%	-	80
Total Fixed OPEX	£M pa	36.2	55.0
Total Variable OPEX (excl. Fuel & Carbon)	£M pa	0.2	43.8
Average Fuel Cost	£M pa	315	242
Average CO_2 Emission Cost	£M pa	369	28.2
Total Start-up Cost (excl. Fuel)	£M	4.4	7.0
Discount Rate	% / year	7.8	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	80.1
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	20.1
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	77.0
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	107.7

The economic performance of the 3 x 500 MWth Net Power full scale system using natural gas fuel is summarised in Table 11-3 along with unabated CCGT Reference case for the purposes of comparison. The capital cost estimate for Case 6 is assessed to have an accuracy of ± 45%. Note that this is a lower level of accuracy than for any of the other Benchmark cases, resulting from uncertainty regarding the costs for the novel supercritical gas turbines.

The total project cost for the Net Power case is 80% higher than the Reference unabated case which is explained below:

- The Net Power system uses several proprietary equipment items such as the high pressure combustor, direct fired CO_2 turbine and main heat exchanger, which have been extensively developed by the equipment suppliers Toshiba and Heatric. This is reflected in the Net Power system's cost compared to the Reference case, which uses a Power Island with standard rotating and static equipment.
- The carbon capture and compression cost for the Net Power case contributes ~£50 M which is not required for the unabated case.
- Predevelopment cost and the other costs are higher for the Net Power case as these are being calculated as a percentage of the EPC contract value.

Total fixed operating cost for the Net Power system is 51% higher than the Reference case as a majority of the fixed cost (such as general overhead, taxes, maintenance, etc.) are calculated as a percentage of capex. Total variable operating cost excluding fuel for the Net Power case is related to the carbon capture process which is not relevant for the unabated Reference case. The combination of fixed and variable OPEX is 2.7 times higher for the Net Power case compared to Reference case.

The increased project cost, operating cost and cost related to the CO_2 transportation & storage makes the Levelised Cost of Electricity for the Net Power case higher than the Reference case. Figure 11-2 shows the list of the different contributing factors and the level of contribution towards the overall LCOE value. It is important to note from that the fuel cost is the biggest contributor to the LCOE followed by capital investment and operating cost. CO_2 emission price has lowest impact on the LCOE calculation.

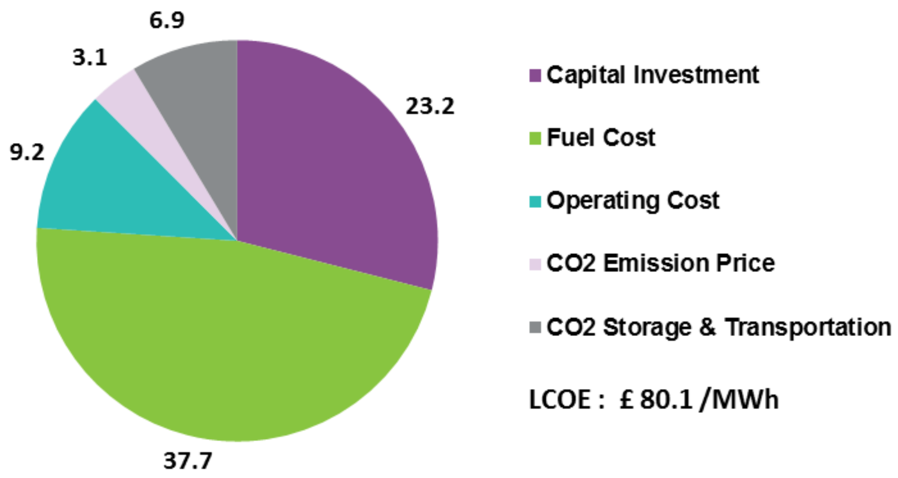

Figure 11-2: LCOE (£/MWh) Contribution for Case 6

11.5.1 Comparison of Results with IEAGHG 2015/05 Report

The value of £ 80.1 / MWh for Levelised Cost of Electricity (LCOE) presented in this report differs significantly from the equivalent result of £ 68.3 / MWh (€ 83.6 / MWh) reported for Case 2 in the IEAGHG Report 2015/05. This results from a variety of differing assumptions used for the two studies, as summarised in Figure 11-3 below.

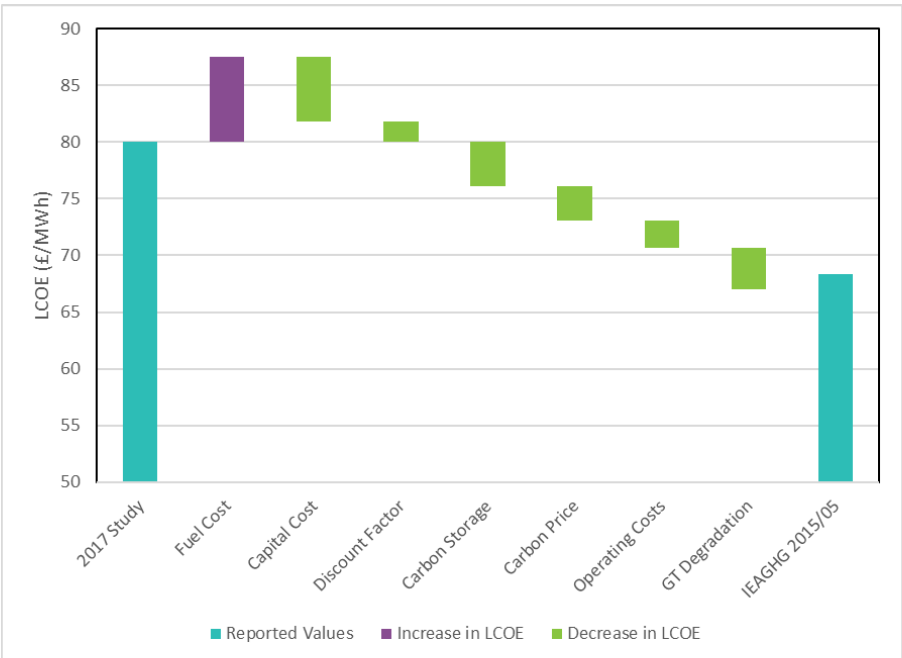

Figure 11-3: Case 6 LCOE Comparison with IEAGHG 2015/05 Report

The 2015 study used a base price of € 8 / GJ (£ 23.5 / MWh) for natural gas, which seemed appropriate at that time before the shale gas revolution caused prices to crash. The gas price profile used for this study (Table 5-7) indicates a 2017 gas price that is about 45% of the 2015 value. This rises in real terms, but is still 12% lower than the 2015 value from 2030 onwards. The higher fuel price used in 2015 would result in a 10% increase in LCOE if applied to this study.

The capital cost estimating methodology adopted in 2015 used smaller factors for mechanical contracts and engineering services; it also excluded infrastructure connections and pre-development costs. Using the 2015 capital cost estimate within the 2017 analysis would reduce the LCOE by a similar amount to the increase due to fuel costs noted above.

The 2015 study used a slightly lower discount factor of 8.0% versus 8.9% in the current study. It also assumed lower costs for CO_2 transportation and storage (€ 10 / tCO_2) and excluded the price of carbon emissions entirely. These all contribute to a higher LCOE in this study than in the 2015 study.

A combination of small differences in operating costs has a noticeable effect upon the LCOE for the two studies. The 2015 study assumed much lower costs for Maintenance, Insurance and Local Taxes, although the allowance for Administration and General Overheads was higher.

The drop in average performance resulting from the use of a gas turbine degradation profile in the current study also has an impact. Using the 'as new' performance throughout the 2017 analysis would result in a small drop in LCOE.

12 Case 7 – Natural Gas CCGT with MCFC Power Generation and Carbon Capture

12.1 Overview

This case consists of a natural gas combined cycle power plant integrated with molten carbonate fuel cells which capture CO_2 from the gas turbine exhaust while using an additional stream of natural gas and generating additional electrical power. The anode exhaust, containing mostly CO_2, water and some unconverted hydrogen and CO is compressed and purified using a cryogenic purification step before recycling the unconverted CO and hydrogen back to the inlet of the fuel cell anode. The CO_2 is then further compressed to the required specification of 110 bar (abs).

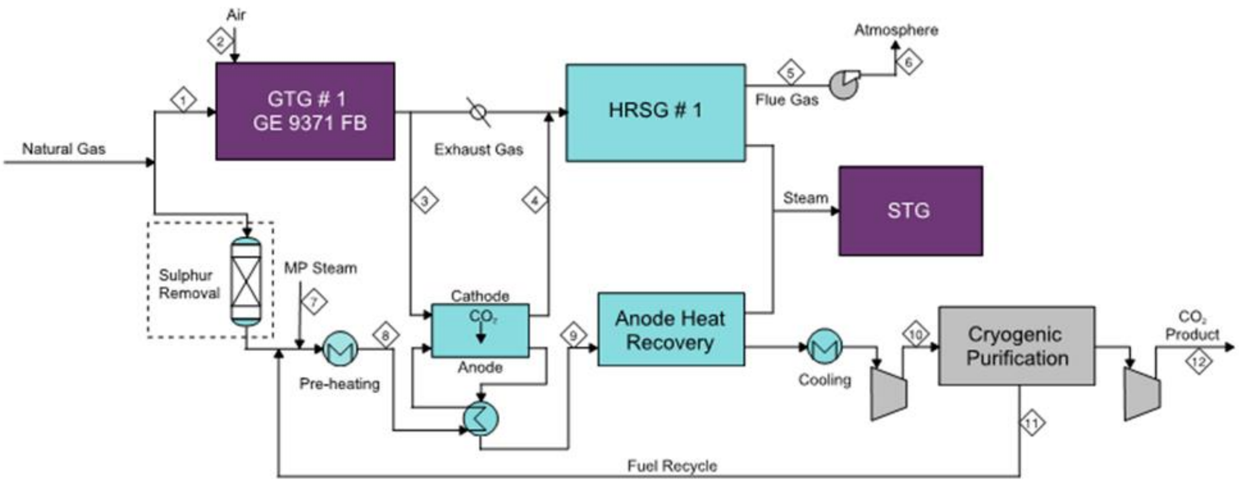

Figure 12-1: Case 7 Preliminary Block Flow Diagram

12.2 Model Development

12.2.1 Material Balance

The CCGT power island was modelled in the same way as for Case 1. Performance and cost data for a combined cycle plant based upon GE 9HA.01 gas turbines were taken from Gas Turbine World 2014-2015 Handbook and up-rated, as recommended in the handbook, for the cooler than ISO site ambient conditions. However, it was necessary to modify the GT exhaust flow rate in order to account for the removal of CO_2 and oxygen in the flue gas flowing through the HRSG.

The chemistry within the cells was modelled in Hysys and cross checked manually, for mass balancing purposes, using the chemical reactions shown in the following diagram. Scrutinising the material balances in the published literature shows that 100% conversion can be assumed for the reforming reactions (only the reaction for methane is shown below, reactions for ethane, propane etc. are similar) and 71% conversion for the shift reaction.

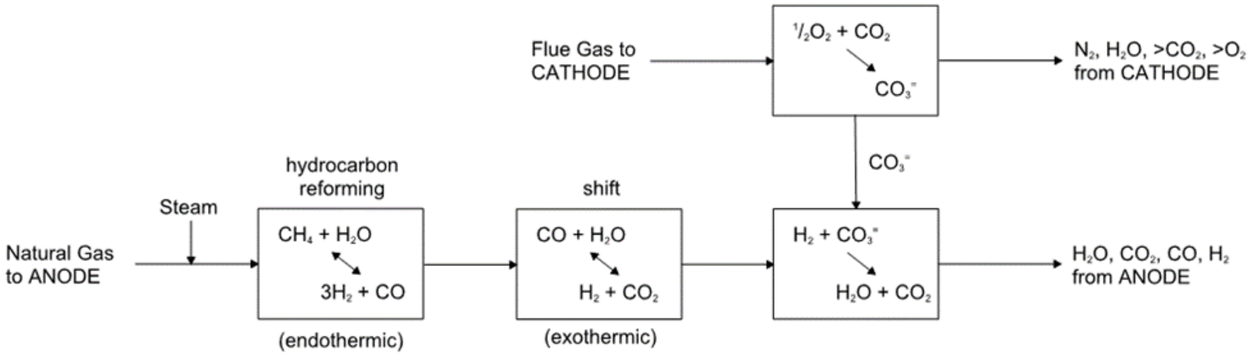

Figure 12-2: Simplified Representation of the Chemistry within a Molten Carbonate Fuel Cell

Combining the reforming and shift reactions, gives the following overall reaction:

$$CH_4 + 2H_2O \longleftrightarrow 4H_2 + CO_2$$

While, combining the CO3= formation with the hydrogen conversion, gives the following overall reaction:

$$2O_2 + 4CO_2 + 4H_2 \longleftrightarrow 4H_2O + 4CO_2$$

Resulting in the overall reaction for all four steps, as follows:

$$CH_4 + 2H_2O + 2O_2 + 4CO_2 \longleftrightarrow 4H_2O + 5CO_2$$

The above theoretical representation shows that; for every mole of methane used by the MCFC, one mole of CO_2 is created and captured, while four further moles of CO_2 are captured from the oxidant stream, or flue gas. The above, however, does not consider either losses, or the fact that two of these reactions are equilibrium reactions. Thus, we cannot expect perfect conversion across the fuel cell.

It is possible to estimate the installed capacity of fuel cells required to capture 90% of the CO_2 from our 2x9HA GTs based upon figures given by FuelCell Energy, such as:

"A 2.8MW DFC3000 fuel cell powerplant during normal power operation is transferring about 3200 kg of CO_2 per hour from the cathode to anode streams in the stack modules. In carbon capture mode, this system could capture and purify about 2300 kg per hour of external CO_2 in addition to the CO_2 exhaust of the DFC powerplant."

Using this statement, we arrive at an installed capacity of 436 MWe of fuel cells to capture 90% of the CO_2 contained in our GTs' fuel gas. This figure is confirmed by the graph of ratio of fuel cell capacity requirement relative to total power plant power for different types of power plant as a function of percent of CO_2 captured.

The thermal efficiency of the fuel cell is quoted by FuelCell Energy Inc. as 47% on an HHV basis for the fuel cell stack itself. This allows the quantity of fuel used by the cells to be calculated in an iterative calculation accounting for the recycle of the unconverted fuel species back to the fuel cell anode (iterative due to the HHV of the fuel at the anode inlet changing each time the composition and quantity of recycle changes).

It is important to be clear that the 47% figure quoted is based upon the HHV at point B below, not point A, the total fuel gas entering the MCFC train of the plant. The efficiency of the fuel cell section of the plant using point A as a basis is in the region of 70 to 75%.

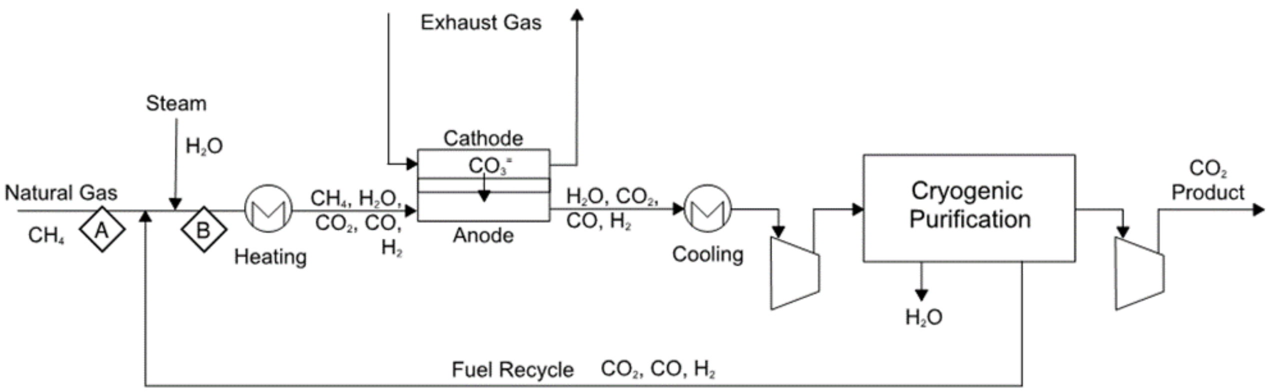

Figure 12-3: Simplified Flow Diagram of a MCFC Train

12.2.2 Fuel Cell Train Layout

The largest commercially installed MCFC fuel cells in operation are the 2.8 MW units supplied by FuelCell Energy Inc. It was therefore anticipated that it may be necessary to assume a design featuring 156 of these units are arranged across a flat space with significant manifold arrangements and hence pressure drop and plot space requirements (plot sizing is outside of the scope of this work). However, significant design development work has been undertaken by FuelCell Energy Inc. since the most recent publicly available references for large scale CO_2 capture units. FuelCell Energy Inc. provided us with the following scheme they envisage for a plant of similar scale.

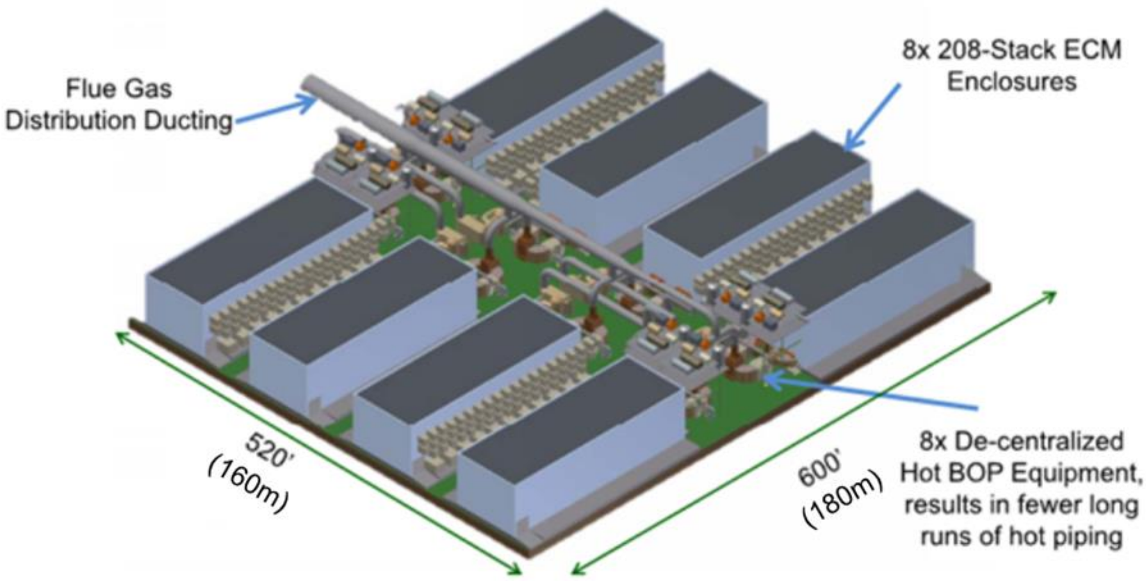

Figure 12-4: Sketch of a 350MWe MCFC Installation
(Image courtesy of FuelCell Energy Inc. based upon work supported by the US Department of Energy under Award Number DE-FE0026580)

The above scheme benefits from much improved space utilisation and manifolding, although we anticipate larger, square ductwork being required compared to those shown above. It also shows how a single gas/gas exchanger per "enclosure" could be arranged. We have assumed a slightly larger arrangement for our design, with 5 enclosures and 5 gas/gas exchangers required per GT train (10 in total). All other equipment is assumed to be one item per GT train (except pumps, which are assumed to be 2 x 100%) for cost estimate development purposes.

The ducting required for distribution of the 550°C GT exhaust gas to the fuel cells and then 660°C CO_2 depleted exhaust gas return to the GT HRSGs is both large and of special material construction. In order to minimise both plot space and cost for this ducting, a single inlet and single outlet duct for the full plant capacity has been specified, or graduated cross sectional area.

12.2.3 Anode Exhaust Heat Recovery

The set of papers presented by Politecnico di Milano show a heat recovery steam generator in which 3 levels of steam appear to be raised from both exhaust streams in parallel without the two streams mixing.

Wood consulted with our Fired Heater Division who have no experience of such a heat recovery steam generator or knowledge of application in industry. The number of steam pressure levels selected in power island conceptual design is generally related to the mass flow rate of exhaust gas the heat is to be recovered from with more steam levels being justified in plants with greater mass flow rate of exhaust gas. Since the anode exhaust is an order of magnitude smaller than the GT exhaust gas stream, we have judged that a single pressure level is appropriate for recovering this heat. Using LP steam is also convenient since the amount of steam which can be generated almost exactly matches that which is required for the steam to carbon ratio of the fuel cell plus the preheating of the fuel mixture.

It is recommended that subsequent development of this design considers a range of options for heat recovery from the anode exhaust in order to identify the optimum cost-effective design.

12.2.4 Materials of Construction

The high temperature gas/gas exchanger will be operating under harsh conditions where both hydrogen embrittlement and metal dusting are likely. While hydrogen embrittlement is relatively well understood (we specify high Co Mo steels for high temperature hydrogen service) metal dusting can be more of a challenge. Our specialist material engineers recommended that:

"No metallic material is immune to metal dusting... Increasing chromium content helps reducing the risk. Nickel alloys such as alloy 625 or 800 H may be used but again not immune. There are some recently developed materials that claimed to be better in metal dusting resistance such as UNS N06696".

We have specified the highest grade recommended for cost estimating purposes, but it may be possible to reduce the cost by using an intermediate heat transfer medium, such as steam. Steam coils have been assumed for heat recovery from the anode exhaust downstream of the gas/gas exchanger.

12.2.5 Stack Lifetime

MCFC stacks are reported to have a current lifespan of about 5 years, however, FuelCell Energy Inc. anticipate achieving a 7-year lifetime by 2020, and hence 7 years has been used as the anticipated lifetime for this study. They also anticipate a 10-year life as being reasonably achievable in the mid-term.

12.3 Process Description

The scheme selected for incorporating MCFCs into a natural case combined cycle power plant for CO_2 capture is based upon the internal reforming schemes presented by the research team at the Politecnico di Milano which places the MCFCs between the gas turbine exit and the heat recovery steam generator. The unconverted fuel species remaining in the anode exhaust are recovered using a cryogenic physical separation stage. The scheme differs from those suggested by the Politecnico di Milano team in that the recovered fuel recycle is returned to the MCFC inlet rather than the GT inlet, as per the Carbon Capture Project's (CCP) findings of higher overall plant efficiency presented at LEAP 2016 ("CCP Novel CO_2 Capture Technology Evaluation: WP1 MCFC package", April 2016, S. Consonni et al.). FuelCell energy have also provided or confirmed key performance data as the leading supplier of megawatt scale fuel cells that can be used for carbon capture.

Our design basis specifies 3 ppm (molar) H_2S in the natural gas, therefore a zinc oxide bed desulphurisation step is required upstream of the MCFCs. Our fuel gas is available at 70 bar (abs) and 9°C, therefore let down to the near atmospheric pressure operating conditions of the fuel cells causes cooling, with little opportunity for energy recovery, requiring additional heating.

12.3.1 CCGT + MCFC Power Island

Natural gas is received from the grid and metered before being routed to the power island. The feed gas is then divided into four streams, two streams are preheated and sent to the two parallel gas turbines and two streams are directed to the MCFC trains.

The air compressors of the gas turbines draw air from the atmosphere and compress it before mixing it with the natural gas fuel in the combustion chamber. The hot combusted gas is then expanded through the turbine which turns a generator (and the compressor) to generate electrical power. The exhaust from each turbine would normally be directed straight to a heat recovery steam generator, but in this case, it is routed instead to the cathode side of the MCFCs.

The natural gas feeding the MCFCs is let down across a let-down station to approximately 24 bar (abs) and mixed with recycled fuel containing 60 wt% CO_2, 30 wt% CO and the balance made up of hydrogen and nitrogen. This mixture is then pre-heated against LP steam before a further pressure let down step. Superheated LP steam is then added to the fuel to achieve a steam to carbon ratio of 2. A final stage of heating is provided in a high temperature inlet/outlet gas/gas exchanger against the anode exhaust gas to achieve 580°C at the MCFC anode inlet.

The MCFC performs several chemical steps; first the hydrocarbon species present in the inlet are reformed with steam to hydrogen and CO, then shifted, using more steam to convert much of the CO to CO_2 while producing additional hydrogen. The hydrogen is then oxidised using oxygen which has been drawn across the fuel cell electrolyte in the form of CO_3= from the flue gas on the cathode side of the cell. The movement of CO_3= across the electrolyte generates electrical current at the same time as capturing CO_2 from the flue gas.

Flue gas from the gas turbine exhaust at 645°C is diverted using a damper to the cathode side of the fuel cell where it is reduced in both CO_2 and oxygen content before being routed back to the inlet of the heat recovery steam generator (HRSG). The HRSG will be reduced in size somewhat compared to that which would be required without the presence of the MCFC because of the reduced mass flow rate of exhaust gas. Thus, the steam which can be generated from that exhaust gas is also less than in a case without the MCFCs present.

In order to overcome the pressure drop in the interconnecting duct work, an induced draft fan is required at the outlet of the HRSG in order to ensure that the treated flue gas has sufficient pressure to disperse from the top of the stack.

The anode exhaust leaving the high temperature gas/gas exchanger is comprised of 44 mol% water, 44 mol% CO_2, 7 mol% unconverted hydrogen and 4 mol% unconverted CO. This stream leaves the exchanger at 580°C and contains enough heat to use boiler feed water from the power island to pre-heat, generate and superheat steam required for both direct addition to the reforming process as well as pre-heating the fuel mixture with a small margin remaining.

The anode exhaust is then cooled against cooling water and compressed to 32 bar (abs), with intermediate cooling and water knock-out, before being fed to a molecular sieve dehydration unit then a cryogenic physical separation stage involving flash at -53°C to recover unconverted CO and hydrogen. This is then recycled to the fuel cell anode. The purified CO_2 stream is then further compressed to 110 bar (abs) ready for export.

The principle impurities remaining in the CO_2 are 0.8 mol% CO, 0.4 mol% hydrogen and 0.2 mol% nitrogen, no water and no oxygen, well within the specification given in the basis of design, with the exception of the CO content. No further treatment steps have been added at this stage, to reduce the CO content to below the specification of 0.2 vol%, but could be considered if the process is taken forward for further development.

The CO content in the captured CO_2 product has not been included in the overall carbon capture figure for the plant.

12.4 Technical Performance Evaluation

Table 12-1: Technical Performance Comparison for Case 7

	Units	Reference Case (Unabated CCGT)	Natural Gas CCGT with MCFC and CCS
Total Gross Installed Capacity	MWe	1229.4	1645.0
Gas Turbine (s)	MWe	823.5	823.5
Steam Turbine	MWe	405.9	381.5
Others	MWe	0	440.0
Total Auxiliary Loads	MWe	20.9	136.4

	Units	Reference Case (Unabated CCGT)	Natural Gas CCGT with MCFC and CCS
Feedstock Handling	MWe	0	0
Power Island	MWe	14.7	14.7
Air Separation Unit	MWe	0	0
CO_2 Capture	MWe	0	48.1
CO_2 Compression	MWe	0	64.5
Utilities	MWe	6.2	9.1
Net Power Export	MWe	1208.5	1508.6
Fuel Flow Rate	kg/h	150,296	195,722
Fuel Flow Rate (LHV)	MWth	1940.2	2526.7
Net Efficiency (LHV) - As New	%	62.3	59.7
Net Efficiency (LHV) - Average	%	59.0	56.6
Total Carbon in Feeds	kg/h	108,640	141,476
Total Carbon Captured	kg/h	0	130,333
Total CO_2 Captured	kg/h	0	477,597
Total CO_2 Emissions	kg/h	398,105	40,834
CO_2 Capture Rate	%	0	92.1
Carbon Footprint	kg CO_2/MWh	329.4	27.1

The plant performance of the GE 9HA.01 based CCGT power plant with molten carbonate fuel cells for post-combustion carbon capture is summarised in the above table. The unabated CCGT Reference case, for the same power island configuration is also listed in the table for the purposes of comparison. The MCFC case captures 90% of the CO_2 from the GT exhausts and burns additional fuel, from which 100% of the CO_2 emitted is captured while producing additional power, the net effect of which is an increase in net power production and only a minimal 2.6% point net efficiency loss.

The following points can be highlighted as basic differences between the two cases:

- The Reference case uses one of the largest and most efficient natural gas fired gas turbines GE 9HA.01 with large power output of > 400 MWe per turbine.
- The CCGT plus MCFC case uses the same high efficiency gas turbine power island configuration plus the fuel cells which have a gross LHV efficiency of ~75%. Thus, this case benefits from a very high efficiency underlying power production before any parasitic loads for carbon capture are applied.
- The parasitic loads associated with the CO_2 capture and compression process result in a net exportable power from the CCGT with MCFC case of 301 MWe more than the Reference case, but with additional fuel fired. These balance each other somewhat such that there is still a reduction in efficiency versus the unabated case, but only of 2.6% lower net LHV efficiency.
- Although the MCFCs require a significant amount of steam for the reforming and shift steps within the cell, this steam can be generated via heat recovery from the fuel cell exhaust.
- CO_2 compression power appears high compared to some schemes because the first compression stages are also compressing the unconverted hydrogen, CO and water vapour prior to the cryogenic purification and fuel recycle step.

- The carbon efficiency for the CCGT with MCFC case is 8% of the Reference unabated case as this case captures 92% of the CO_2 produced in power generation for transportation and storage.

12.5 Economic Performance Evaluation

Table 12-2: Operations and Maintenance Staff Manning for Case 7

	Reference Case Unabated CCGT	Natural Gas CCGT with MCFC and CCS	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Deputy Plant Manager	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	Daily Position
Process Engineer	NA	1	Daily Position
Shift Supervisor	5	10	3-shift Position
Electrical Assistant	5	10	3-shift Position
Control Room Operator	10	15	3-shift Position
Field Operator	10	25	3-shift Position
Sub-Total	32	64	
Maintenance Staff			
Mechanical Group	3	5	Daily Position
Instrument Group	3	3	Daily Position
Electrical Group	2	4	Daily Position
Sub-Total	8	12	
Laboratory Staff			
Superintendent	1	1	Daily Position
Analysts	3	3	Daily Position
Sub-Total	4	4	
Plant Total Staff	44	80	

Table 12-3: Economic Performance Comparison for Case 7

	Units	Reference Case (Unabated CCGT)	Natural Gas CCGT with MCFC and CCS
Total Project Cost	£M	672.2	1,569.6
Specific Total Project Cost	£/kW	556	1,038
Pre-Development Costs			
Pre-Licensing & Design	£M	5.8	13.9
Regulatory & Public Enquiry	£M	12.9	29.3
EPC Contract Cost	£M	583.6	1,392.0
Feedstock Handling	£M	0	0
Power Island	£M	583.6	571.5
Air Separation Unit	£M	0	0

	Units	Reference Case (Unabated CCGT)	Natural Gas CCGT with MCFC and CCS
CO_2 Capture	£M	0	714.5
CO_2 Compression	£M	0	105.9
Utilities	£M	0	0
Other Costs			
Infrastructure Connections		29.0	37.0
Owner's Costs		40.9	97.4
Overall CAPEX Impact (vs Ref Case)	£M	-	897.5
Overall CAPEX Impact (vs Ref Case)	%	-	134
Total Fixed OPEX	£M pa	36.2	71.9
Total Variable OPEX (excl. Fuel & Carbon)	£M pa	0.2	108.5
Average Fuel Cost	£M pa	315	398
Average CO_2 Emission Cost	£M pa	369	36.6
Total Start-up Cost (excl. Fuel)	£M	4.4	17.7
Discount Rate	% / year	7.8	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	70.7
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	-11.7
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	68.4
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	75.8

The economic performance of the CCGT with MCFC system for power generation with carbon capture is summarised in Table 12-3 along with unabated CCGT Reference case for the purposes of comparison. The capital cost estimate for Case 7 is assessed to have an accuracy of ± 40%.

The total project cost for this case is 134% higher than the Reference unabated case while producing 25% more net power output:

- The MCFC system includes a number of high capital cost elements, such as the MCFC cells themselves and exotic materials required for heat exchangers downstream of the fuel cell and high temperature ducting. These elements have been captured within the cost for the CO_2 Capture Unit.
- The CCGT Power Island has slightly lower capital cost in this case, due to the smaller mass flow rate of exhaust gas (due to removal of CO_2 and Oxygen upstream of the HRSG) and consequently smaller steam turbines.
- Operating costs are also high for this case because the MCFC stacks need to be replaced every 7 years, which is the major contributor to this figure.
- Despite the capital and operating costs excluding fuel being substantially higher for this case than the unabated case, the Levelised cost of electricity (LCOE) is lower, at £70.7 / MWh, compared to the unabated cost of £74.2 / MWh. This is due to the very high efficiency of this case combined with its very low carbon emission per unit of net power produced.

The chart below shows the balance of factors contributing to the overall LCOE. It can be seen that the fuel cost is the major factor but that the capital investment and operating cost are also very significant.

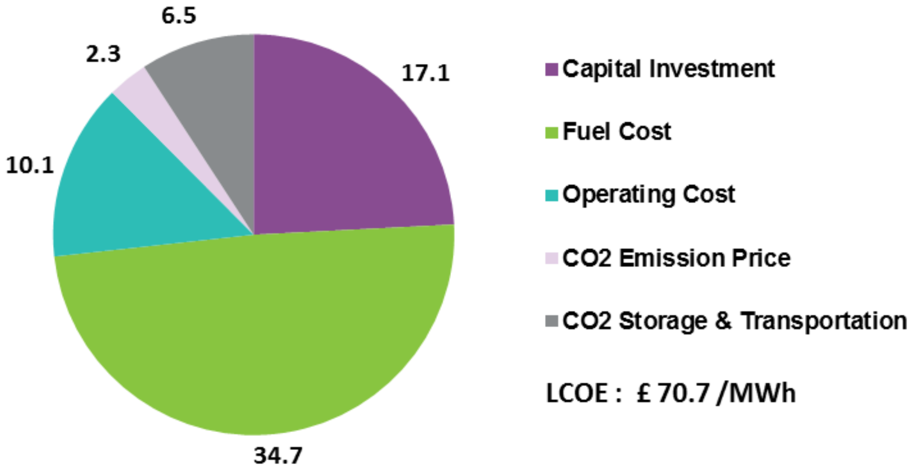

Figure 12-5: LCOE (£/MWh) Contribution for Case 7

13 Case 8 – Biomass Fired CFB Boiler with Post-Combustion Carbon Capture

13.1 Overview

This case consists of two parallel trains of 300 MWe biomass-fired sub-critical circulating fluidised bed (CFB) boiler power plants, each with a once through steam generator with superheating and single steam reheating. The selection of this configuration is outlined in Section 5.2.1. The flue gas from the biomass CFB boiler requires no flue gas desulphurisation before entering the CO_2 capture process due to the minimal sulphur in the biomass feed. CO_2 is captured from the cooled flue gas using an amine based solvent in an absorption column and is released from the solvent in the stripper. The captured CO_2 leaving the proprietary CO_2 capture unit is then compressed in 4 stages, dehydrated and then pumped to the required export pressure of 110 bar (abs). The treated flue gas exiting the CO_2 absorber is heated in a condensate heater (CH) before release to atmosphere.

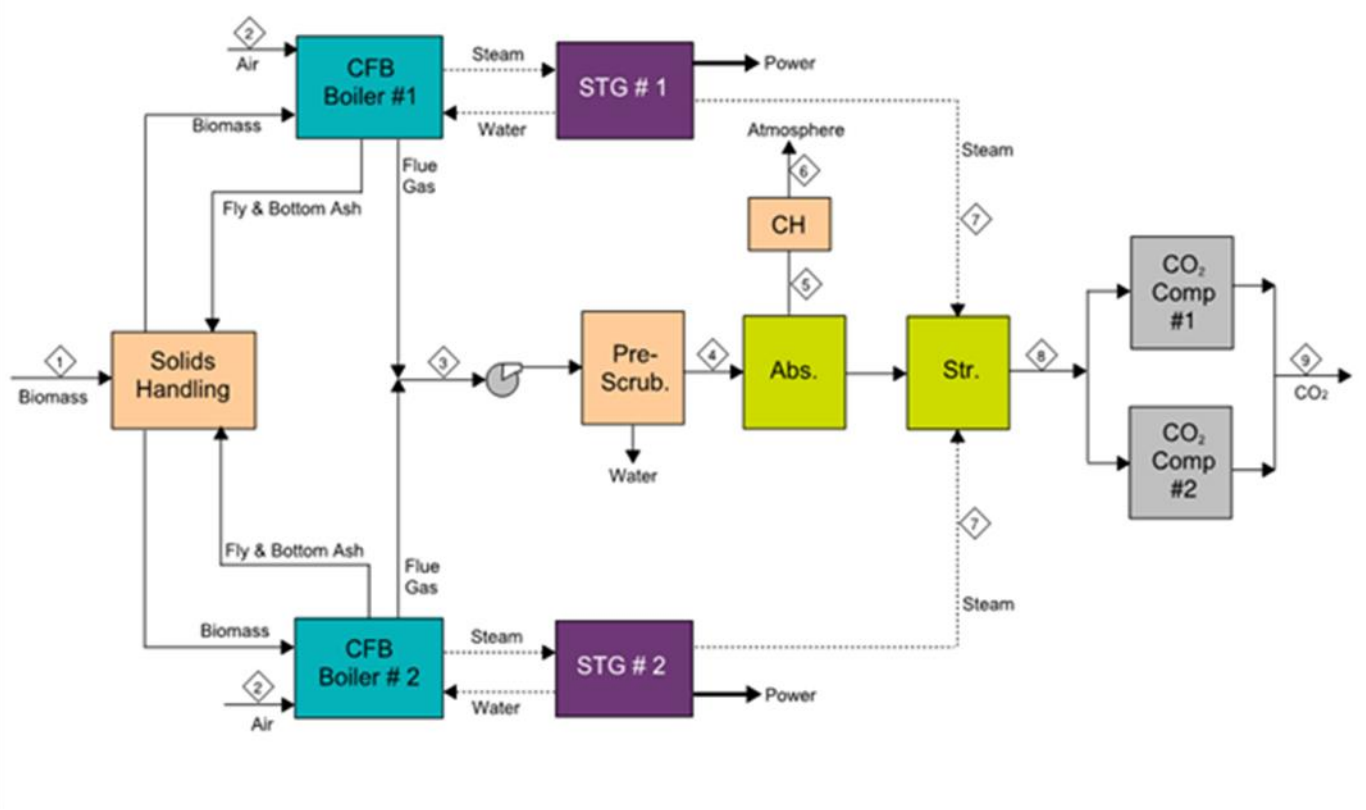

Figure 13-1: Case 8 Block Flow Diagram

13.2 Model Development

The subcritical circulating fluidised bed (CFB) boiler technical data has been developed for this case using the in-house data of a typical 300 MWe biomass-fired CFB boiler. The steam turbine and its integration with the Cansolv CO_2 reboiler have been modelled in Hysys to determine the power output from the power island. The CO_2 compression system has also been modelled in Hysys. The Peng Robinson property package has been used to model the process scheme whereas the water and steam cycle has been modelled using the NBS Steam property package.

Cansolv Technologies Inc. (Shell Cansolv) was contracted to provide a Process Design Package for the CO_2 removal unit. Cansolv submitted the package of absorption and regeneration system which uses its proprietary CANSOLV Absorbent DC-103 for CO_2 absorption. This solvent is suitable for CO_2 capture from the flue gas produced from both coal and biomass-fired boilers. This solvent was used at the SaskPower Boundary Dam facility in Canada which is a coal fired power plant and is generally recognised as an industry leader in post-combustion CO_2 capture. All modelling of the proprietary amine absorption and stripping systems was conducted by Shell Cansolv. Detailed results cannot be published in this report due to confidentiality restrictions and some details have been redacted from the attached deliverables, but the overall results presented in this section reflect the latest performance results achieved by Shell Cansolv.

Figure 13-2: SaskPower Boundary Dam Facility
(Image courtesy of Cansolv Technologies Inc.)

13.3 Process Description

13.3.1 Biomass Storage and Handling

Biomass is received at the plant via trucks and unloaded by cranes and conveyed to the biomass storage building by belt conveyors. The storage building holds an inventory of 30 days of biomass feed to the plant.

The transfer of stored biomass from the storage building is done by means of screws which load the belt conveyors. The biomass conveying system is enclosed to avoid environmental emissions. The conveyors in the transfer towers are fitted with filters and an exhaust system.

13.3.2 Boiler Island

The biomass boiler used for this case is Sumitomo Foster Wheeler's 'Compact' tower subcritical circulating fluidised bed (CFB) boiler with integrated water cooled solid separators. The boiler island includes the fuel feeding system, the furnace, the solid separators with the solid return channels and INTREX superheaters, fans and air heaters.

The biomass feeding system consists of multiple feeders which include a day silo, drag chain feeder, conveyor and discharge system. The discharge is via a dosing screw and wall feeding screw.

The furnace has a single fluidising grid under which primary air is introduced in a controlled manner to achieve uniform fluidisation. The combustion is controlled at a relatively low temperature (~850°C) by introducing secondary air at different elevations within the furnace walls to ensure staged combustion and also to minimise NOx formation. Hot flue gas exiting the boiler economiser is used to preheat the combustion air before entering the CO_2 capture unit.

In addition to the staged combustion with overfire air, additional NOx removal is required to limit the NOx content in the flue gas to ~ 1 ppmv before entering the CO_2 capture unit. This is achieved by using a Selective Catalytic Reduction (SCR) system in between the convection section and the air preheater. Ammonia, the reducing agent, is injected immediately upstream of a catalyst surface where the NO and NO_2 are reduced to N_2 and water.

Flue gas desulphurisation and limestone addition to the combustion chamber to limit the SOx emission is not required for this case because of the low sulphur content in the biomass.

The high pressure pre-heated BFW is heated in the economiser against the flue gas and sent to the steam drum. Dry HP steam is superheated in the furnace roof, convective superheaters and INTREX heat exchangers.

MP steam from the exhaust of the HP section of the steam turbine is reheated in the convective reheaters and INTREX heat exchangers.

Two types of ash are generated by the process: furnace bottom ash and fly ash from the flue gas. A bag house filter is provided to remove the entrained particulates from the flue gas, which is collected as fly ash. Both ashes are collected into storage silos before being disposed. Bottom ash is generally disposed of to landfill whereas fly ash can be used in the cement industry.

13.3.3 Steam Turbine

The Power Island is composed of a single condensing steam turbine and preheating lines. Superheated steam from the boiler at 176 bara and 568°C is sent to the steam turbine which consists of HP, MP and LP sections. The MP steam from the exhaust of the HP section of the steam turbine is reheated in the boiler island to 568°C before entering the MP section of the steam turbine at 39 bara. Part of the LP steam is routed to the reboiler in the CO_2 capture unit and the remainder passes on to the LP section of the steam turbine.

The LP steam turbine exhausts at vacuum conditions of 0.04 barg, or as close to that pressure as can be achieved in the condenser given the cooling water temperature. The condenser is directly below the LP steam turbine and also receives the required make-up water.

Steam condensate is pumped to approximately 10 barg and preheated using steam extractions before being returned to the deaerator to complete the circuit.

Chemical injection in the water circuit is made from dedicated packages to control the water quality.

13.3.4 Proprietary Solvent CO_2 Capture

An outline of the Shell Cansolv process for CO_2 capture as applied to a biomass power plant is shown in Figure 13-3.

Flue gas is transferred to a pre-scrubber column which performs the dual function of cooling (with water knock-out) and SO_2 removal. Cooling the flue gas to 35°C reduces the required absorbent circulation rate and thus energy consumption and CAPEX of the Cansolv unit.

In order to decrease the impact of SO_2 on the absorbent, the pre-scrubber uses caustic to reduce the SO_2 content in the flue gas upstream of the CO_2 Absorber. SO_2 removal is controlled by adding caustic on pH control in a caustic polishing section inside the pre-scrubber column. For the purposes of this study it is assumed that the concentration of SO_2 leaving the pre-scrubber polishing section would be 1 ppmv.

The cooled and pre-scrubbed flue gas is ducted to the bottom of the absorption column. For a plant of this scale and flue gas type, typical absorber dimensions range from 13m to 18m in square cross section. CO_2 absorption from the flue gas occurs by counter-current contact with CANSOLV Absorbent DC-103 in the CO_2 Absorber, which is a vertical multi-level packed-bed tower. CO_2 is absorbed into the solvent by chemical reaction leaving a flue gas depleted in CO_2 at the top of the column.

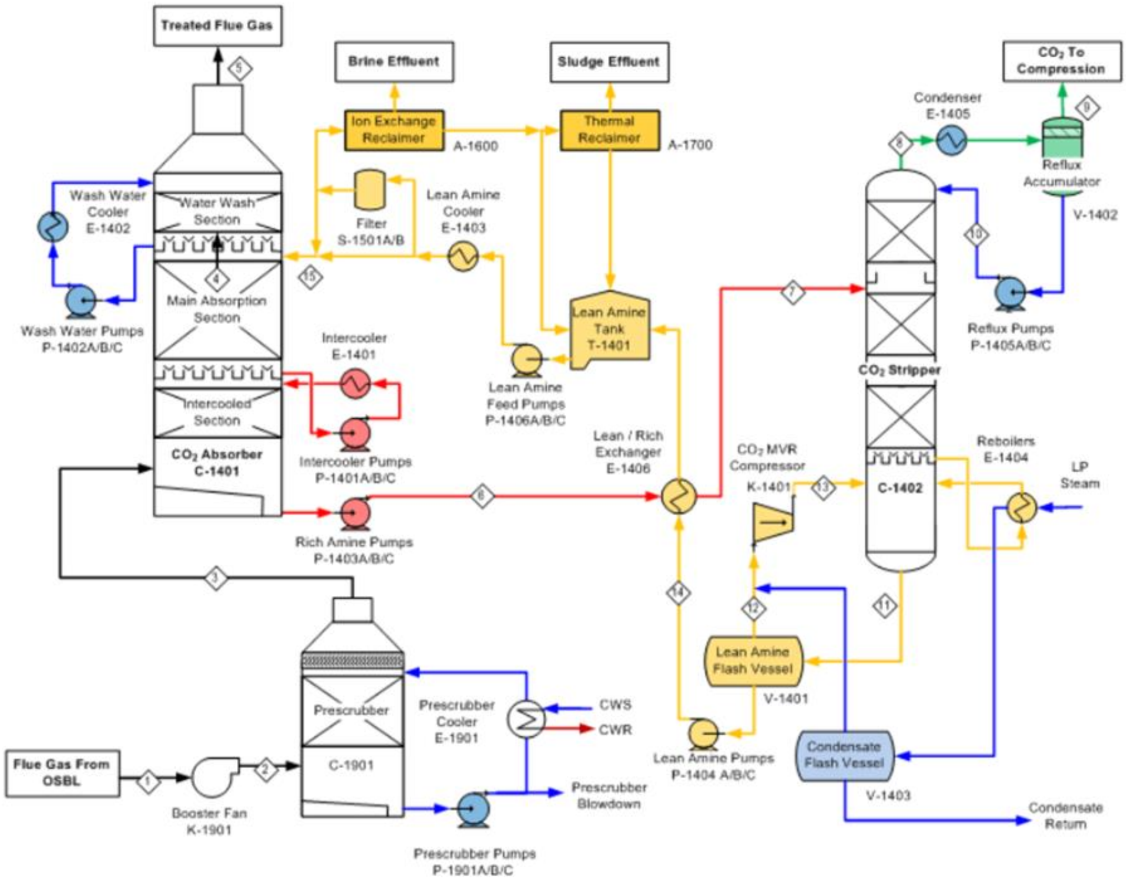

Figure 13-3: Cansolv Process Configuration
(Image courtesy of Cansolv Technologies Inc.)

The CO_2 absorption reaction is exothermic; this increases the temperature of the solvent leading to reduced absorption capacity and increased water evaporation from the absorbent into the heated flue gas. To remove heat from the solvent, an absorber intercooler has been used. A water wash packed bed section is included at the top of the CO_2 Absorber to capture volatile or entrained solvent and to condense water to maintain the water balance in the system. The treated flue gas leaving the Water Wash Section is then warmed up to ~ 80°C against reboiler steam condensate in the condensate heater and routed to a stack for discharge to the atmosphere. The design flue gas outlet temperature is selected such that the overall required water make-up rate is minimised.

The rich solvent from the Absorber column passes through the CO_2 Lean / Rich Exchangers where it is heated by the hot lean absorbent. The hot rich solvent then enters the top of the CO_2 Stripper / Regenerator for solvent regeneration and CO_2 recovery. The rich solvent is depleted of CO_2 by water vapour generated in the reboilers.

Lean solvent from the Stripper is collected in the Stripper Reboilers. Water vapour and lean amine flow from the reboilers back to the Stripper. Lean absorbent then goes to the Lean Absorbent Flash Vessel, where it flashes and releases water vapour which is then compressed in the CO_2 Mechanical Vapour Recovery (MVR) package and is introduced at the bottom of the Stripper to contribute to the stripping of the CO_2. This system minimises the steam and energy consumption of the Cansolv Unit.

The regenerator overhead gas is partially condensed in the condensers producing two phase mixtures of water and CO_2. The reflux water is returned to the regenerator rectification section. The CO_2 product gas is piped to the CO_2 Compression System.

All amine-based systems require some form of solvent maintenance system as over time the absorbent in the CO_2 Capture System accumulates Heat Stable Salts (HSS), as well as non-ionic amine degradation products that must be removed from the solvent. This is achieved through thermal reclamation. An ion exchange package is included for bulk HSS removal upstream of a thermal reclaimer.

The ion exchange package is designed to remove HSS from the CANSOLV DC Absorbent. These salts are continuously formed within the absorbent, primarily due to residual amounts of NO_2 and SO_2 contained in the flue gas. Once absorbed, NO_2 forms nitric and nitrous acid while SO_2 forms sulphurous acid which oxidises to sulphuric acid. These acids, and some organic acids formed by the oxidative degradation of the amine, neutralise a portion of the amine, which is then inactivated for further CO_2 absorption.

The purpose of the Thermal Reclaimer Unit is to remove the non-ionic degradation products as well as HSS from the active absorbent. The thermal reclaimer unit distils the absorbent under vacuum conditions to separate the water and amine, leaving the non-ionic degradation products in the bottom. A slipstream is taken from the treated CO_2 lean absorbent exiting the ion exchange package and fed to the Thermal Reclaimer Unit. This stream will essentially consist of water, amine, degradation products, residual CO_2 and small amounts of sodium nitrate and sodium sulphate. The design flow rate of CO_2 lean absorbent sent to the thermal reclaimer is based on the calculated amine degradation rate. To maintain the degradation products below design concentration, the thermal reclaimer must process a specific flowrate of CO_2 lean absorbent. The reclaimed absorbent is sent to the Lean Absorbent Tank. The separated degradation products are stored in a tank, where they are diluted and cooled with process water. Diluted residues are periodically disposed of offsite, typically via incineration.

13.3.5 CO_2 Compression and Dehydration

The CO_2 is compressed to 30 barg in 4 stages, each with intercooling and water knock-out. This recovers the vast majority of the water content, but is not sufficient for most pipeline specifications. Numerous studies have compared drying with tri-ethylene glycol (TEG) versus use of molecular sieve adsorption which concludes that there is little to choose between the two methods. For this case, a TEG dehydration unit is selected, similar to the coal case 3.

Final CO_2 pressurisation up to 110 bara is achieved using one further stage of compression followed by a condenser then a stage of pumping.

13.4 Technical Performance Evaluation

Table 13-1: Technical Performance Comparison for Case 8

	Units	Reference Case (Unabated CCGT)	Coal Post-Combustion with CCS	Biomass Post-Combustion with CCS
Total Gross Installed Capacity	MWe	1229.4	953.5	498.0
Gas Turbine (s)	MWe	823.5	0	0
Steam Turbine	MWe	405.9	953.5	498.0
Others	MWe	0	0	0
Total Auxiliary Loads	MWe	20.9	139.4	101.8
Feedstock Handling	MWe	0	3.4	1.8
Power Island	MWe	14.7	31.2	25.2
Air Separation Unit	MWe	0	0	0
CO_2 Capture & Comp.	MWe	0	88.3	63.8
Utilities	MWe	6.2	16.5	11.0
Net Power Export	MWe	1208.5	814.2	396.2
Fuel Flow Rate	kg/h	150,296	325,000	635,178

	Units	Reference Case (Unabated CCGT)	Coal Post-Combustion with CCS	Biomass Post-Combustion with CCS
Fuel Flow Rate (LHV)	MWth	1940.2	2335.5	1288.0
Net Efficiency (LHV) - As New	%	62.3	34.9	30.8
Net Efficiency (LHV) - Average	%	59.0	34.7	30.6
Total Carbon in Feeds	kg/h	108,640	209,950	158,795
Total Carbon Captured	kg/h	0	188,926	142,954
Total CO_2 Captured	kg/h	0	692,310	523,849
Total CO_2 Emissions	kg/h	398,105	77,040	58,045
CO_2 Capture Rate	%	0	90.0	90.0
Carbon Footprint	kg CO_2/MWh	329.4	94.6	146.5

The plant performance of the subcritical CFB biomass power plant with state-of-the-art Shell Cansolv post-combustion carbon capture is summarised in the above table. The unabated natural gas Reference case and equivalent post-combustion capture coal case are also listed in the table for comparison.

This biomass-fired case is not expected to compete on a performance or economic basis with an unabated gas-fired power plant. The Reference case is included to allow the calculation of the Cost of CO_2 Avoided. Compared with the Reference case, the biomass post-combustion with Cansolv case captures 90% of the CO_2 while suffering a 31.5% point net efficiency loss. Further comparison of this case versus the Reference case is of little value for this report.

The following points can be highlighted as basic difference between the two post-combustion cases using Cansolv technology for CO_2 capture:

- Fuel thermal energy to the supercritical pulverised coal power plant is 1.8 times higher than biomass power plant whereas the net power output from the coal case is 2.1 times higher than biomass case. This leads to the coal case achieving 4.1% points higher net LHV efficiency.
- This is mainly due to the relatively lower parasitic demand for the coal case leading to more power output for a given thermal input. The larger scale of the coal plant and supercritical operation make it more efficient.
- Parasitic demand for the Power Island for the coal case is only 1.2 times higher even though processing 1.8 times more fuel on thermal basis. This results from the amount of biomass fuel processed by the biomass case, which is nearly double the coal case due to its lower calorific value.
- Parasitic demand for the CO_2 capture and compression train for the coal case is only 1.4 times higher than biomass case. This is due to comparatively large volume of flue gas to be processed for biomass case by the Cansolv which is 82% higher than the coal case even though only processing 55% of the fuel on thermal basis.
- The coal post-combustion plant with 90% carbon capture emits 77 t/hr of CO_2 to the atmosphere whereas carbon emissions. The biomass power plant emits 58 t/hr of carbon dioxide and will need to pay the resultant carbon price. However, if the overall carbon chain is taken into account this process is carbon negative.
- The carbon footprint for the biomass appears relatively high at 146.5 kg CO_2/MWh, but as noted above, the overall process captures carbon dioxide from the atmosphere for long-term storage.

13.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staff are taken to be in daily positions, working regular hours.

Table 13-2: Operations and Maintenance Staff Manning for Case 8

	Reference Case Unabated CCGT	Coal Post-Combustion with CCS	Biomass Post-Combustion with CCS	Remarks
Operations Staff				
Plant Manager	1	1	1	Daily Position
Deputy Plant Manager	1	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	1	Daily Position
Process Engineer	NA	3	2	Daily Position
Shift Supervisor	5	10	10	3-shift Position
Electrical Assistant	5	5	5	3-shift Position
Control Room Operator	10	20	15	3-shift Position
Field Operator	10	35	30	3-shift Position
Sub-Total	32	76	65	
Maintenance Staff				
Mechanical Group	3	6	5	Daily Position
Instrument Group	3	4	4	Daily Position
Electrical Group	2	3	3	Daily Position
Sub-Total	8	13	12	
Laboratory Staff				
Superintendent	1	1	1	Daily Position
Analysts	3	6	5	Daily Position
Sub-Total	4	7	6	
Plant Total Staff	44	96	83	

Table 13-3: Economic Performance Comparison for Case 8

	Units	Reference Case Unabated CCGT	Coal Post-Combustion with CCS	Biomass Post-Combustion with CCS
Total Project Cost	£M	672.2	1,732.2	1,247.6
Specific Total Project Cost	£/kW	556	2,128	3,149
Pre-Development Costs				
Pre-Licensing & Design	£M	5.8	15.5	11.1

	Units	Reference Case Unabated CCGT	Coal Post-Combustion with CCS	Biomass Post-Combustion with CCS
Regulatory & Public Enquiry	£M	12.9	32.1	23.3
EPC Contract Cost	£M	583.6	1,547.3	1106.7
Other Costs				
Infrastructure Connections		29.0	29.0	29.0
Owner's Costs		40.9	108.3	77.5
Overall CAPEX Impact (vs Ref Case)	£M	-	1,060.0	575.4
Overall CAPEX Impact (vs Ref Case)	%	-	158	86
Total Fixed OPEX	£M pa	36.2	80.6	58.0
Total Variable OPEX (excl. Fuel / Carbon)	£M pa	0.2	108.0	82.1
Average Fuel Cost	£M pa	315	143	190
Average CO_2 Emission Cost	£M pa	369	69.1	52.1
Total Start-up Cost (excl. Fuel)	£M	4.4	10.9	8.2
Discount Rate	% / year	7.8	8.9	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	93.3	170.1
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	81.3	524.1
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	85.8	158.4
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	171.4	617.2

The economic performance of the biomass-fired CFB boiler with Cansolv post-combustion carbon capture is summarised in Table 13-3 along with unabated CCGT case and coal post-combustion with CCS case for the purposes of comparison. The capital cost estimate for Case 8 is assessed to have an accuracy of ± 40%.

The total project cost for this case is 186% higher than the Reference unabated natural gas case while producing 67% less net power output.

The total project cost for the biomass case is ~28% lower than the coal case while producing half the net power output. This equates to a higher capex intensity for the biomass case than the coal case.

The capital cost for the Cansolv unit is not proportional to the fuel thermal input for the biomass and coal case; even though the biomass case only processes 55% of the fuel on a thermal basis compared to the coal case, the amount of flue gas entering the capture unit is about 82% of the coal case.

The operating costs (excluding fuel and carbon emission costs) for the biomass power plant with capture plant is only 26% lower than the coal plant, which demonstrates that the cost of running the biomass plant has a substantial effect on the project.

The cost of biomass fuel is 4.7 times higher than coal on a thermal basis. The combination of higher fuel cost, higher capital and operating cost, along with the lower overall efficiency for the biomass plant, has a significant impact on the LCOE for this case. However, no financial benefit is included for the positive impact of negative emissions with 90% capture of the process CO_2 from the 'CO_2 neutral' renewable fuel source.

The chart below shows the balance of factors contributing to the overall LCOE for Case 8. It can be seen that the biomass cost is the major contributing factor to the LCOE, followed by capital costs, CO_2 storage and transportation and operating cost. This is a marked change from the coal-fired post-combustion case (Case 3) where capital investment is the significantly larger portion of the LCOE than other contributing factors (refer to Figure 8-4).

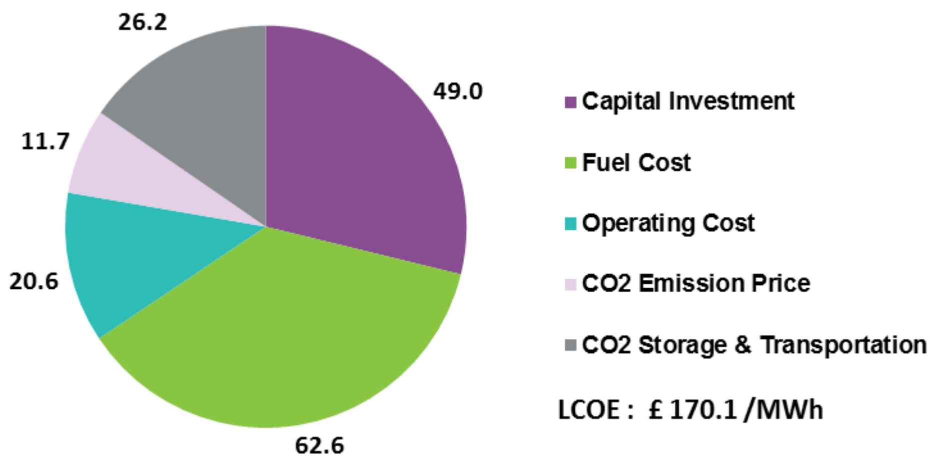

Figure 13-4: LCOE (£/MWh) Contribution for Case 8

13.5.1 Comparison of Results with IEAGHG 2009 Report

The value of 30.8% for net plant efficiency presented in this report differs from the efficiency reported for Case 3b (250 MWe CFB Boiler with CO_2 Capture) in the IEAGHG Report 2009/09. The contributing elements to the higher efficiency for Case 8 are shown in Figure 13-5 below.

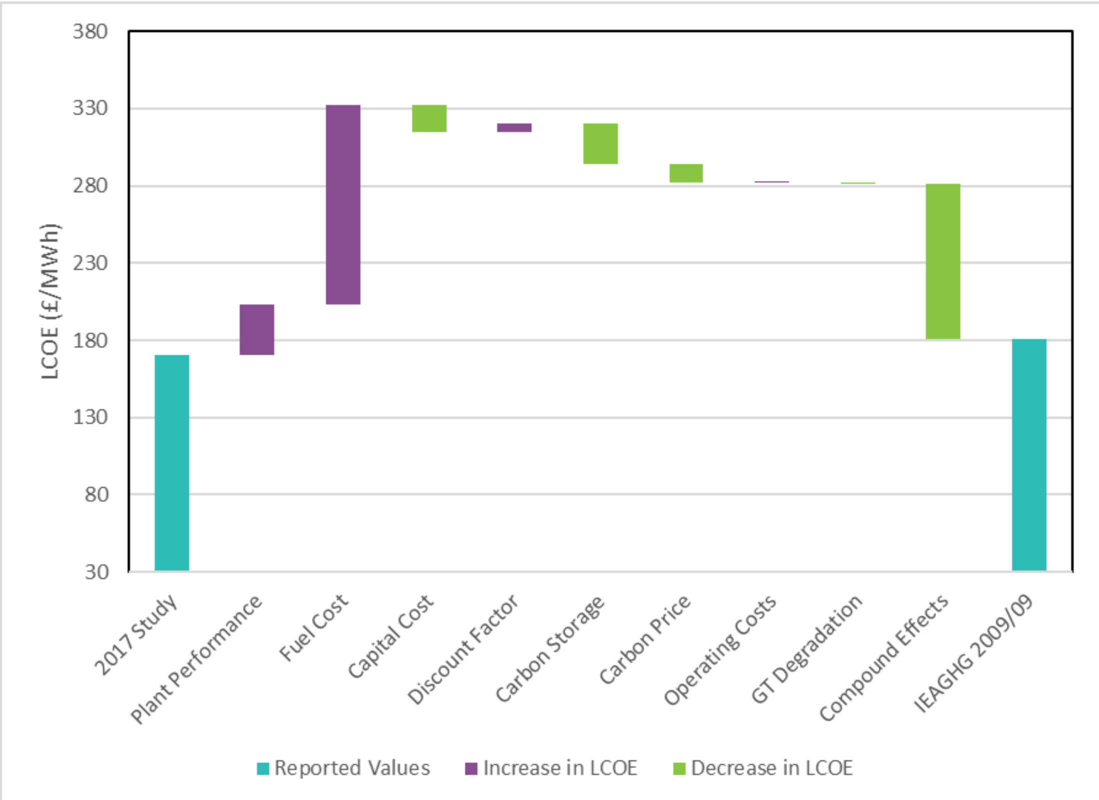

Figure 13-5: Case 8 Net Plant Efficiency Comparison with IEAGHG 2009/09 Report

The CFB boiler considered for the IEAGHG study is 91.2% efficient whereas the state of the art CFB boiler used for this study is 93.3% efficient. Even though both studies use the same feedstock (wood chips), the more efficient modern boiler requires less fuel to produce similar power. The more efficient CFB boiler contributes to 0.4% of the net efficiency improvement.

The parasitic demand for the Case 8 is ~ 9% less than IEAGHG case. The main contributing factor is parasitic demand for the CO_2 compression unit for Case 8, which is significantly less than the IEAGHG study. It is not clear from the IEAGHG report what efficiency had been considered for the compressor, hence it is difficult to quantify the difference. The CO_2 capture unit (Cansolv) parasitic demand for Case 8 is higher than the generic capture plant considered for the IEAGHG report. However, overall demand for electric load for Case 8 is lower, which contributes to 0.7% of the net efficiency improvement.

The power output from the steam turbine for Case 8 is higher than IEAGHG with the main contribution from LP turbine section as less LP steam is used for the Cansolv reboiler than for the generic capture unit for IEAGHG Case 3b. Overall, the higher power output from the steam turbine for Case 8 contributes to ~3.8% of the net efficiency improvement.

14 Case 9 – Biomass Fired CFB Boiler with Oxy-Combustion Carbon Capture

14.1 Overview

This case consists of two parallel trains of oxy-fired sub-critical circulating fluidised bed (CFB) boiler power plants processing biomass feed, each in a steam generator with superheating and single steam reheating. Oxygen for firing in the boiler is supplied by a cryogenic air separation unit (ASU). Electricity is generated from a single steam turbine generator (STG). The flue gas from the boiler is routed to a multi-pass gas/gas heat exchanger (GGH) followed by heat recovery before a portion of the flow is routed back to the boilers as Secondary Recycle, via the gas/gas heat exchanger. The remaining flue gas passes through further heat recovery before the Primary Recycle is split off and recycled via the gas/gas heat exchanger to the biomass mills. The unrecycled flue gas stream is compressed and purified in a cryogenic purification unit (CPU) to the required export pressure of 110 bar (abs).

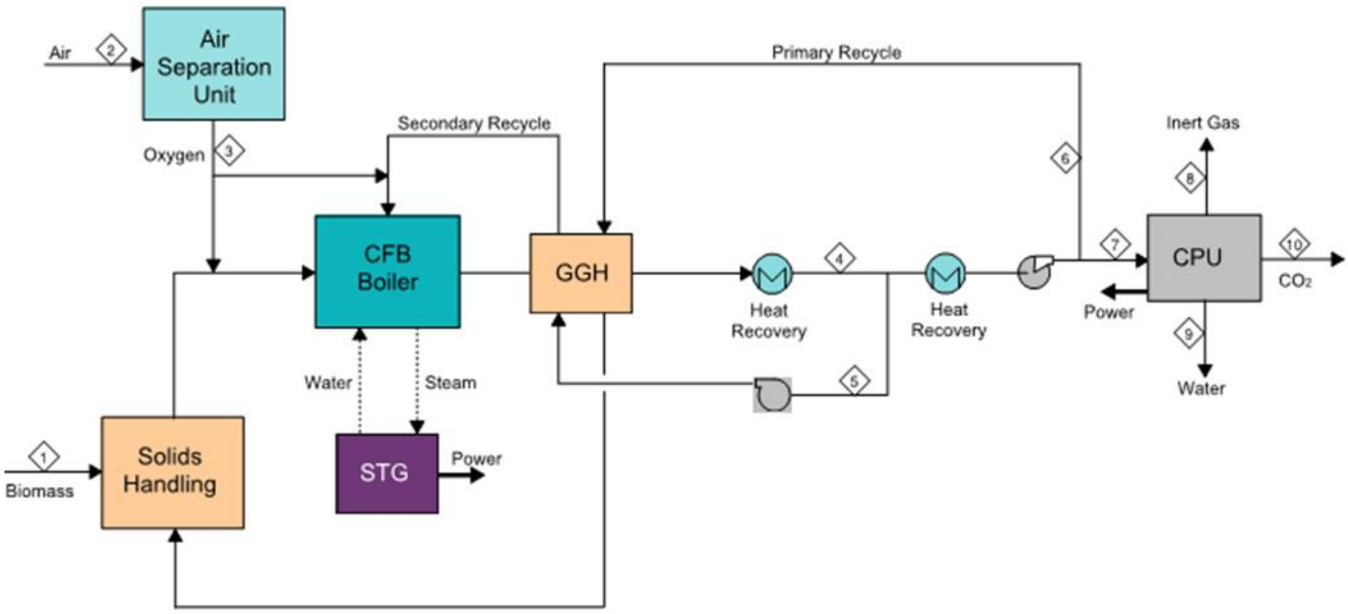

Figure 14-1: Case 9 Block Flow Diagram

14.2 Model Development

The oxy-fired subcritical circulating fluidised bed (CFB) boiler is treated as a specialist package and is a typical commercial single pass tower type boiler. It is expected that although oxy-fired versions of the subcritical circulating fluidised bed (CFB) boiler are not yet commercial at the scale required for this case, the behaviour and design features will not be significantly different from the air fired plant. Hence, the oxy-fired CFB boiler technical data has been kept similar to the air fired CFB boiler used for Case 8. However, because dedicated carbon capture plant is not required for this case, the steam turbine power output is higher than Case 8, as no steam extraction is needed for the carbon capture plant reboiler. The CO_2 compression system has been modelled in Hysys. The Peng Robinson property package has been used to model the process scheme whereas the water and steam cycle has been modelled using the NBS Steam property package.

It is not intended to model the packaged units, other than for high level thermodynamic and material balance checking. The air separation unit (ASU) and cryogenic purification unit (CPU) were modelled using component splitters since these are both specialised package units. These units can be modelled in greater level of detail if technology assessment is required and more data is available.

14.3 Process Description

14.3.1 Air Separation Unit

Air Separation Unit (ASU) capacity is defined by oxygen requirements of the biomass boiler to achieve complete combustion. The required purity is 97mole% pure oxygen to the CFB boiler. The required oxygen flow rate for each oxy-CFB boiler is approximately 240 t/h.

Since the oxy-combustion boiler requires low pressure oxygen, there is no need for further pressurisation of the oxygen stream in this case, therefore this ASU has a significantly lower energy usage per tonne of oxygen produced, compared to the IGCC case ASU (Case 10).

As the required purity, pressure of O_2 and overall process design of this case is similar to the oxy-fired supercritical pulverised coal boiler case (Case 4), it can be assumed that the design features of the ASU will be similar to Case 4 as described below.

Air from the atmosphere is filtered and initially compressed to 3.5 bar (abs) in an axial flow compressor. The hot outlet of this compressor could be used for pre-heating condensate in the Power Island. Remaining impurities such as CO_2 and water are removed in adsorbent beds which alternate between adsorption and regeneration modes.

The purified 3.5 bar (abs) air stream is split in two, with one stream divided again into two streams and sent, via the main heat exchanger, to the intermediate pressure distillation column and the expander section of the compander respectively. The remaining 3.5 bar (abs) stream is compressed further, in two stages with intercooling, before being split into two further streams passing through the main heat exchanger. The first stream is fed to the bottom of the high pressure column and the second stream is further divided to feed the mid-sections of both the intermediate and high pressure columns.

The main cryogenic heat exchanger consists of several parallel aluminium plate-fin exchanger blocks manifolded together. The cryogenic distillation columns are contained within a cold box and divided into low pressure, intermediate pressure and high pressure columns. Liquid oxygen product is produced from the bottom of the low pressure column at the required purity of 97%, while the final “waste” nitrogen stream is produced from the top of the low pressure column.

14.3.2 Biomass Storage and Handling

Biomass is received at the plant via trucks and unloaded by cranes and conveyed to the biomass storage building by belt conveyors. The storage building holds an inventory of 30 days of biomass feed to the plant.

The transfer of stored biomass from the storage building is done by means of screws which load the belt conveyors. The biomass conveying system is enclosed to avoid environmental emissions. The conveyors in the transfer towers are fitted with filters and an exhaust system.

14.3.3 Boiler Island

As discussed in Section 14.2, even though the oxy-fired subcritical CFB boiler is treated as a specialist package, its behaviour and design features will not be significantly different from the air fired CFB boiler described in Case 8. The biomass boiler considered for this case is Sumitomo Foster Wheeler’s ‘Compact’ tower subcritical circulating fluidised bed (CFB) boiler capable of using oxygen as an oxidant instead of air. The other features of the boiler have been considered similar to the Case 8 CFB boiler described in Section 13.3.2.

The biomass feeding system consists of multiple feeders which include a day silo, drag chain feeder, conveyor, and discharge system. The furnace has a single fluidising grid under which the combined stream of oxygen and primary recycle flue gas is introduced in a controlled manner to achieve uniform fluidisation. The remaining oxygen and the secondary recycle are supplied via the staged combustion system to control the overall combustion temperature and also to minimise NOx formation. In addition to the staged combustion, additional NOx removal is also required to limit the NOx content in the flue gas to ~ 1 ppmv before entering the CPU. This is achieved by using a

Selective Catalytic Reduction (SCR) system in between the convection section and the air preheater. Ammonia, the reducing agent, is injected immediately upstream of a catalyst surface where the NO and NO_2 are reduced to N_2 and water.

Hot combustion products exit the boiler economiser and the residual heat is recovered from the flue gas in the regenerative air preheaters before passing through the CPU unit.

Flue gas desulphurisation and limestone addition to the combustion chamber to limit the SOx emission is not required for this case because of low sulphur content in the biomass.

The high pressure pre-heated BFW is heated in the economiser against the flue gas and sent to the steam drum. Dry HP steam is superheated in the furnace roof, convective superheaters and INTREX heat exchangers.

MP steam from the exhaust of the HP section of the steam turbine is reheated in the convective reheaters and INTREX heat exchangers.

Two types of ash are generated by the process; furnace bottom ash and fly ash from the flue gas. A bag house filter is provided to remove the entrained particulates from the flue gas, which is collected as fly ash. Both ashes are collected into storage silos before being disposed of. Bottom ash is generally disposed in the landfill whereas fly ash can be used in the cement industry.

14.3.4 Steam Turbine

The Power Island is composed of a single condensing steam turbine and preheating lines. Superheated steam from the boiler generated at 176 bara and 568°C is sent to the steam turbine which consists of HP, MP and LP sections. The MP steam from the exhaust of the HP section of the steam turbine is reheated in the boiler island to 568°C before entering the MP section of the steam turbine at 39 bara. The LP steam turbine exhausts at vacuum conditions of 0.04 barg, or as close to that pressure as can be achieved in the condenser given the cooling water temperature.

Steam condensate is pumped to approximately 10 barg and preheated using steam extractions before being returned to the deaerator to complete the circuit. The condenser is directly below the LP steam turbine and also receives the required make-up water.

Chemical injection in the water circuit comes from dedicated packages to control the water quality.

14.3.5 Flue Gas Cooling and Recycles

A gas/gas heat exchanger (GGH) is used to cool the hot flue gas exiting the convection section which preheats both the primary and secondary recycles. After the GGH, the flue gas stream is cooled and split into two streams. One becomes the secondary recycle and is returned to the boilers via a forced draft fan. The remaining flue gas stream is then cooled further against power island condensate before entering a contact cooler. In the contact column, flue gas at approximately 110°C is quenched with water from the bottom of the direct contact cooler. In the contact column, a circulating water stream is used to cool the flue gas to 28°C. The circulating water stream is cooled against cooling water and this system is a net producer of water.

The cooled flue gas is then boosted in the induced draft fan and further divided into two streams. The first stream becomes the primary recycle stream and is heated in the gas/gas exchanger before being routed to the solids handling area as conveying gas for the biomass feed. The remaining flue gas from the ID fan is routed to the CO_2 compression and purification unit.

14.3.6 CO_2 Compression and Purification

The cooled flue gas from the ID fan is at approximately 38°C and low pressure. It requires purification to remove oxygen, inerts and moisture and is then compressed up to 110 bara for transport and storage.

In this study, the process offered by Air Products has been assumed, to be consistent with the Case 4 coal oxy-combustion process design. The process consists of the following process steps:

- Compression up to 30 bara
- Temperature Swing Adsorption (TSA)
- Auto-refrigerated inerts removal
- Final compression to 110 bar (abs)

Compression

The flue gas stream is compressed adiabatically to 15 bar (abs) and 300°C, then cooled against various streams which require heating: the inerts from the downstream cold box, boiler island BFW, condensate and finally cooling water. The flue gas then undergoes several reaction steps.

The first reaction step is to oxidise any remaining NO to NO_2 in the following reaction:

$$NO + {}^{1}/_{2}\, O_2 \rightarrow NO_2 \qquad (1)$$

The second step is to react NO_2 and SO_2 to produce sulphuric acid as follows:

$$NO_2 + SO_2 + H_2O \rightarrow NO + H_2SO_4 \qquad (2)$$

Remaining NO_2 is then converted to nitric acid as follows:

$$2NO_2 + H_2O \rightarrow HNO_2 + HNO_3 \qquad (3)$$

$$HNO_2 \rightarrow HNO_3 + 2NO + H_2O \qquad (4)$$

All of the NO produced in reactions 2 and 4 is reconverted to NO_2 in the first reaction. Any mercury present at this point in the system is simultaneously removed via the formation of mercuric nitrate due to mercury's ready reaction with nitric acid.

Reactions 1 and 2 begin in the final stage of flue gas cooling but contacting columns are needed to ensure the reactions proceed sufficiently to meet the CO_2 product specification. The flue gas travels up the first contacting column against a stream of descending acid water to completely convert all SO_2 present. Part of the contact liquid is cooled and returned while the remaining part is sent to waste water treatment.

The CO_2 from the contacting column is further compressed to 30 bar in an integrally geared compressor and cooled against cooling water before being fed to the bottom of the second contacting column. The contact liquid in this second column is a nitric acid solution which converts the remaining NOx content to nitric acid. As in the first contactor, part of the liquid is recycled and part is sent to waste water treatment. This process is said to remove all of the SO_2 and 90% of the NOx from the flue gas / CO_2 stream.

TSA System

The CO_2 is dried in an adsorbent bed to a dew point of -55°C prior to inerts removal. The bed is regenerated thermally using MP steam from the power island in a cyclical process in which two beds alternate between dehydration mode and regeneration mode.

Auto-refrigerated Inerts Removal

The CO_2 at 30 bara is cooled to -54°C in a series of two multiple pass aluminium plate-fin exchangers contained in a cold box, then flashed to remove the bulk of the inerts. The vapour stream, containing mostly inerts is passed back through the exchangers for cold-recovery, warmed against hot CO_2 in the sour compression section then expanded in a power recovery turbine before being vented to atmosphere at a safe location.

The flashed liquid CO_2 is warmed in one of the main exchangers, expanded to 16-17 bara and fed to a distillation column. The vapour product from this column is compressed and recycled to the front of the cold box. The liquid CO_2 product from the distillation column is divided into two streams which are expanded to 5.6 bara and 16-17 bara to provide the cooling required in the main heat exchangers before entering the CO_2 compressor. The lower pressure stream is compressed in an integrally geared compressor up to 16-17 bara, then cooled. It joins the 16-17 bara stream and the combined stream is compressed in two intercooled stages up to 110 bara.

14.4 Technical Performance Evaluation

Table 14-1: Technical Performance Comparison for Case 9

	Units	Reference Case (Unabated CCGT)	Coal Oxy-combustion with CCS	Biomass Oxy-combustion with CCS
Total Gross Installed Capacity	MWe	1229.4	1112.8	598.0
Gas Turbine (s)	MWe	823.5	0	0
Steam Turbine	MWe	405.9	1097.7	598.0
Others	MWe	0	15	0
Total Auxiliary Loads	MWe	20.9	280.2	195.9
Feedstock Handling	MWe	0	3.3	1.8
Power Island	MWe	14.7	20.9	26.3
Air Separation Unit	MWe	0	213.5	88.9
CO_2 Capture	MWe	0	26.8	68.0
CO_2 Compression	MWe	0	0	0
Utilities	MWe	6.2	15.7	11.0
Net Power Export	MWe	1208.5	832.6	402.1
Fuel Flow Rate	kg/h	150,296	325,000	635,178
Fuel Flow Rate (LHV)	MWth	1940.2	2335.0	1288.0
Net Efficiency (LHV) - As New	%	62.3	35.7	31.2
Net Efficiency (LHV) - Average	%	59.0	35.5	31.1
Total Carbon in Feeds	kg/h	108,640	209,950	158,795
Total Carbon Captured	kg/h	0	187,176	142,748
Total CO_2 Captured	kg/h	0	685,896	523,093
Total CO_2 Emissions	kg/h	398,105	83,455	58,801
CO_2 Capture Rate	%	0	89.2	89.9
Carbon Footprint	kg CO_2/MWh	329.4	100.2	146.2

The plant performance of the subcritical CFB boiler with oxy-combustion carbon capture is summarised in the table above. Both the Reference case (unabated CCGT) and coal oxy-combustion with CCS case (Case 4) are also listed in the table for the purposes of comparison.

Similar to the post-combustion case, biomass-fired oxy-combustion case is not expected to compete on a performance or economic basis with an unabated gas-fired power plant. The Reference case is included to allow the calculation of the Cost of CO_2 Avoided. Compared with the Reference case, the biomass oxy-combustion with 90% CO_2 capture suffers a 32.1% point net efficiency loss. Further comparison of this case versus the Reference case is of little value for this report.

The following points can be highlighted as basic difference between the two oxy-combustion cases with 90% CO_2 capture:

- Fuel thermal energy to the supercritical pulverised oxy-fired coal power plant is 1.8 times higher than for the biomass oxy-fired power plant, whereas the net power output from the coal case is 2.1 times higher than biomass case. This leads to 4.5% points higher net LHV efficiency for the coal case.
- This is mainly due to the relatively lower parasitic demand for the coal case leading to more power output for a given thermal input. The larger scale of the coal plant and supercritical operation make it more efficient.
- Parasitic demand for the Power Island (including the ASU load for oxygen production) for the coal case is only 1.3 times higher even though processing 1.8 times more fuel on a thermal basis. This results from the amount of biomass fuel processed by the biomass case which is nearly double the coal case, due to its lower calorific value.
- Parasitic demand for the CO_2 capture / compression train for the coal case is only 1.6 times higher than biomass case. This is due to comparatively large volume of flue gas to be processed for the biomass case, which is ~ 77% higher than the coal case even though only processing 55% of the fuel on thermal basis.
- The coal oxy-combustion plant with 90% carbon capture emits ~83 t/hr of CO_2 to the atmosphere. The biomass power plant emits 59 t/hr of carbon dioxide and will need to pay the resultant carbon price. However, if the overall carbon chain is taken into account this process is carbon negative.
- The carbon footprint for the biomass appears relatively high at 146.2 kg CO_2/MWh, but as noted above, the overall process captures carbon dioxide from the atmosphere for long-term storage.

14.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant is listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staff are taken to be in daily positions, working regular hours.

Table 14-2: Operations and Maintenance Staff Manning for Case 9

	Reference Case Unabated CCGT	Coal Oxy-combustion with CCS	Biomass Oxy-combustion with CCS	Remarks
Operations Staff				
Plant Manager	1	1	1	Daily Position
Deputy Plant Manager	1	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	1	Daily Position
Process Engineer	NA	3	2	Daily Position
Shift Supervisor	5	10	10	3-shift Position
Electrical Assistant	5	5	5	3-shift Position
Control Room Operator	10	20	15	3-shift Position
Field Operator	10	40	35	3-shift Position
Sub-Total	32	81	70	
Maintenance Staff				
Mechanical Group	3	7	6	Daily Position
Instrument Group	3	4	4	Daily Position

	Reference Case Unabated CCGT	Coal Oxy-combustion with CCS	Biomass Oxy-combustion with CCS	Remarks
Electrical Group	2	4	4	Daily Position
Sub-Total	8	15	14	
Laboratory Staff				
Superintendent	1	1	1	Daily Position
Analysts	3	6	5	Daily Position
Sub-Total	4	7	6	
Plant Total Staff	44	103	90	

Table 14-3: Economic Performance Comparison for Case 9

	Units	Reference Case Unabated CCGT	Coal Oxy-Combustion with CCS	Biomass Oxy-Combustion with CCS
Total Project Cost	£M	672.2	1,901.9	1,449.5
Specific Total Project Cost	£/kW	556	2,284	3,605
Pre-Development Costs				
Pre-Licensing & Design	£M	5.8	17.0	12.9
Regulatory & Public Enquiry	£M	12.9	35.2	27.0
EPC Contract Cost	£M	583.6	1,701.5	1,290.3
Feedstock Handling	£M	5.8	109.1	86.3
Power Island	£M	12.9	774.5	594.2
Air Separation Unit	£M	5.8	353.4	240.3
CO_2 Capture	£M	12.9	197.5	167.3
CO_2 Compression	£M	5.8	0	0
Utilities	£M	12.9	267.0	202.2
Other Costs				
Infrastructure Connections		29.0	29.0	29.0
Owner's Costs		40.9	119.1	90.3
Overall CAPEX Impact (vs Ref Case)	£M	-	1,229.7	777.3
Overall CAPEX Impact (vs Ref Case)	%	-	183	116
Total Fixed OPEX	£M pa	36.2	86.8	66.0
Total Variable OPEX (excl. Fuel / Carbon)	£M pa	0.2	108.3	81.5
Average Fuel Cost	£M pa	315	143	190
Average CO_2 Emission Cost	£M pa	369	74.8	52.7
Total Start-up Cost (excl. Fuel)	£M	4.4	11.8	8.9

	Units	Reference Case Unabated CCGT	Coal Oxy-Combustion with CCS	Biomass Oxy-Combustion with CCS
Discount Rate	% / year	7.8	8.9	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	96.0	177.9
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	95.1	566.1
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	88.0	166.3
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	185.5	659.3

The economic performance of the biomass fired oxy-CFB boiler with carbon capture is summarised in Table 14-3 along with the unabated CCGT Reference case and coal oxy-combustion with CCS case for the purposes of comparison. The capital cost estimate for Case 9 is assessed to have an accuracy of ± 40%.

The total project cost for this case is 116% higher than the Reference unabated natural gas case, while producing 66% less net power output. The total project cost for the biomass case is 24% lower than the coal case while producing less than half of the net power output. This equates to a higher capex intensity for the biomass case than the coal case.

The operating costs (excluding fuel and carbon emissions costs) for the biomass power plant with capture plant is only 24% lower than the coal plant, demonstrating that the cost of running the oxy-fired biomass plant has a significant impact on the LCOE.

The cost of biomass fuel is 4.7 times higher than coal on a thermal basis. The combination of higher fuel cost, higher capital and operating cost, along with the lower overall efficiency for the oxy-fired biomass power plant, has a significant impact on the LCOE. However, no financial benefit is included for the positive impact of negative emissions with 90% capture of the process CO_2 from the 'CO_2 neutral' renewable fuel source.

The chart below shows the balance of factors contributing to the overall LCOE. It can be seen that, as with the biomass post combustion case, the biomass cost is the major contributing factor to the LCOE followed by capital investment, CO_2 storage and transportation and operating cost. This is marked change from the coal-fired oxy-combustion case (Case 4) where capital investment is a significantly larger portion of the LCOE than other contributing factors.

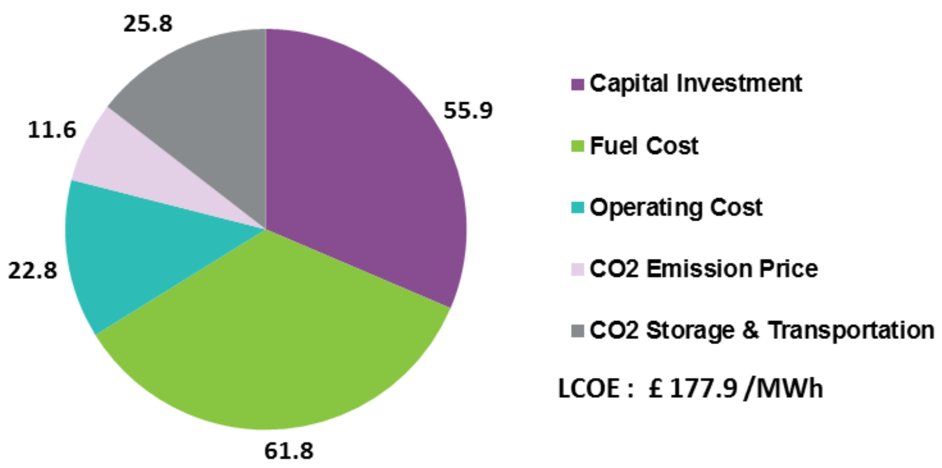

Figure 14-2: LCOE (£/MWh) Contribution for Case 9

15 Case 10 – Biomass IGCC with Pre-Combustion Carbon Capture

15.1 Overview

This case consists of two parallel trains of torrefied wood pellet fired syngas production units: feeding syngas to a single train of combined cycle power island consisting of one GE 9F syngas variant gas turbine followed by heat recovery steam generation (HRSG) and single steam turbine. Syngas production units are based upon the medium pressure Shell gasification process with a dry feed system and syngas cooler. A hybrid carbon monoxide (CO) shift stage is followed by two stages of sour shift and Selexol physical solvent system for removal of acid gases. Captured CO_2 from the Selexol unit is compressed in 4 stages, dehydrated and then pumped to the required export pressure of 110 bar (abs). Oxygen for the gasification process is supplied by a cryogenic air separation unit (ASU). The syngas production unit produces sweetened and decarbonised syngas which is then fed to the single train combined cycle power island.

Support for the development of this case has been provided by Rob van den Berg, Principal Licensing Technology Manager Gasification at Shell Global Solutions International.

15.1.1 Process Configuration

The main process configuration of the IGCC plant is as follows:

- Medium-pressure (~40 barg) Shell Biomass Gasification Process;
- Hybrid CO shift followed by two stages of sour shift;
- Acid gas removal (CO_2) using Selexol physical solvent system;
- CO_2 compression and pumping up to 110 bara;
- Combined cycle based on GE F-class syngas variant gas turbines.

Table 15-1 describes the process units with trains which are also shown in Figure 15-1.

Table 15-1: IGCC Process Units with Trains

Unit Number	Unit Description	Trains
100	Biomass Handling & Storage	N/A
200	Shell Biomass Gasification Package	2 x 50%
300	Air Separation Unit (ASU)	2 x 50%
400	Syngas Treatment & Conditioning	2 x 50%
500	Acid Gas Removal (AGR)	2 x 50%
700	CO_2 Compression & Dehydration	2 x 50%
800	Gas Turbine & Generator Package	1 x 100%
900	Heat Recovery Steam Generation	1 x 100%
1000	Steam Turbine & Generator Package	1 x 100%
1100	Offsite & Utilities	

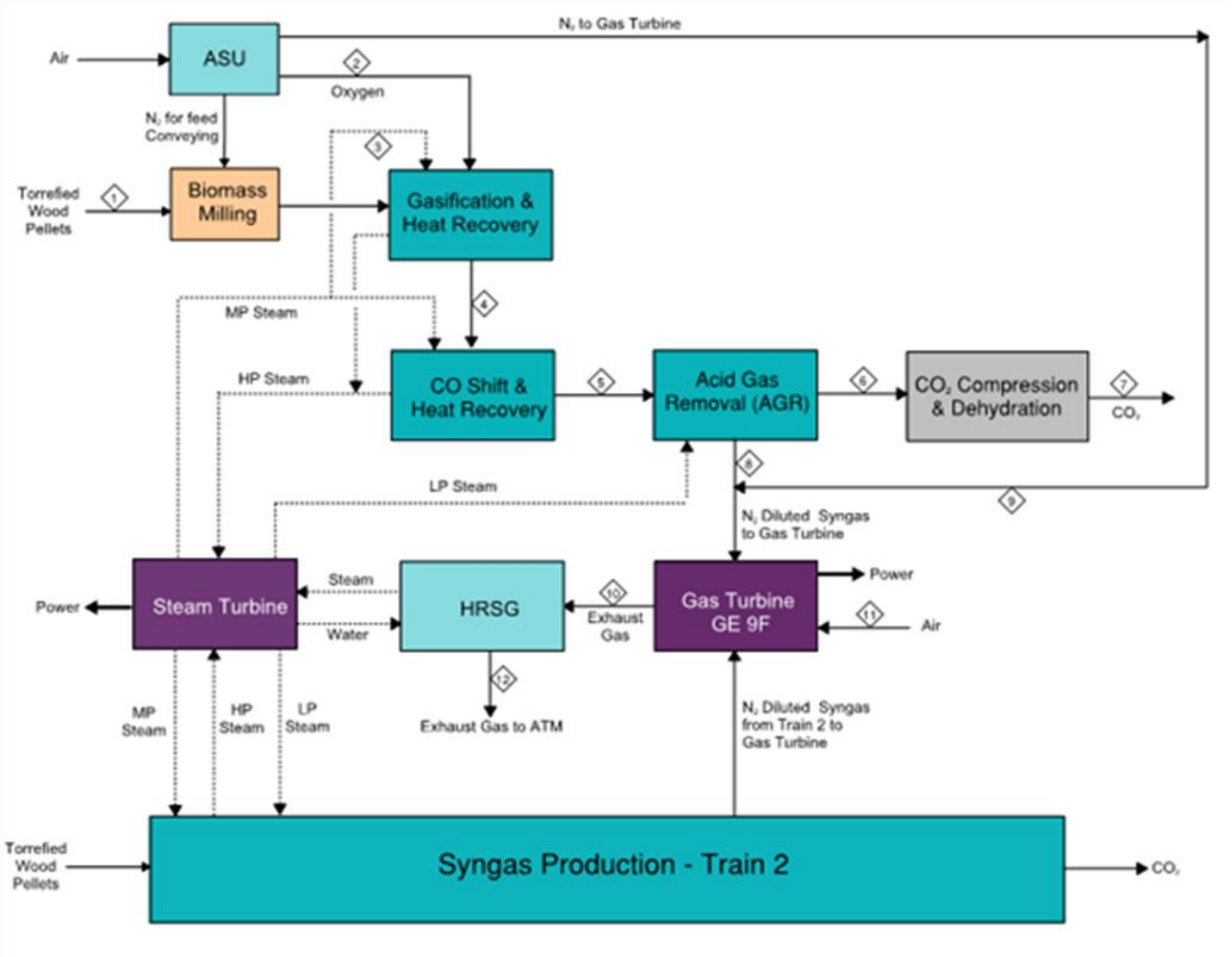

Figure 15-1: Case 10 Block Flow Diagram

15.2 Process Model Development

The biomass feed considered for this process is energy-dense torrefied biomass. Torrefication of biomass is a natural prolonged heat treatment process that makes the biomass stable and durable. This process increases the energy density of the biomass by equalising the material and lowering its moisture level to nearly 0%. The specification of the torrefied biomass (virgin wood chips) for this study has been developed from a paper presented by ECN (Carbo et al, 'Torrefied Biomass Pellets Key to Establish Dense-Phase Flow Feed to Entrained Flow Gasifiers', 8th International Freiberg Conference on IGCC & XTL, Germany, 2016) and discussion with Shell Global Solutions.

The bio-IGCC case has been developed using 100% torrified wood pellet fired to the Shell gasification process. Shell advised that in order to be able to gasify 100% torrefied biomass using a Shell entrained flow gasifier, an external source of mineral matter needs to be added to the feed to increase the ash content from ~4%, for typical torrefied biomass, up to 6-8%. This is required to make sure that the combined 'ash' in the feedstock would have the right viscosity under the actual gasification conditions and also to meet the required criteria of producing a protective slag layer on the gasifier wall. Shell advised to use kaolin clay for the torrefied biomass instead of commonly applied fluxant limestone, since the CaO content of the biomass is already fairly high.

The Shell gasification process model for 100% torrefied biomass has been developed using the following information provided by Shell Global Solutions:

- The torrefied biomass feed rate to each entrained flow gasifier is 112.7 t/h;

- Addition of 1.5% kaolin clay is required to increase the ash content of the feed in order to obtain the proper slag viscosity and protective slag layer on the gasifier wall;
- The required gasification temperature for biomass is slightly below 1500°C, which is generally similar to that required for lignite as feedstock;
- In this case, the quenched gas temperature is around 750°C, determined by the properties of the molten ash;
- Syngas production rate per gasifier is ~122,000 Nm^3/hr ($CO+H_2$);
- The cold gas efficiency of the gasification island is 77.6% (LHV basis).

The overall bio-IGCC flow scheme using Shell gasification technology has been simulated in Aspen Hysys using the similar process flow scheme as for the coal-IGCC case (Case 5) - that given in the reference IEAGHG 2014 report (Coal and Hydrogen with CCS report). The Peng Robinson property package has been used to model the process scheme, whereas the water and steam cycle has been modelled using the NBS Steam property package. The simulation data are in good agreement with the information provided by Shell Global Solutions in terms of syngas production rate and cold gas efficiency of the gasification island.

The HP, MP and LP steam conditions and the process pressure profile are kept in line with Case 5 (coal-IGCC process). The heat integration between the process units and combined cycle has been developed to minimise the heat loss from the system and maximise power output. The single train of combined cycle Power Island consists of one GE 9F syngas variant gas turbine followed by HRSG and single steam turbine. Syngas variant gas turbine technical data used for this case is based on typical supplier data.

15.3 Process Description

15.3.1 Torrefied Biomass / Kaolin Clay Storage and Handling

Torrefied biomass and fluxant are stored on site, with an inventory of 30 days of design consumption to counteract any delivery disruptions.

The biomass feeding system, from the storage building to the gasification island, consists of conveyors, elevated feed hoppers, crushers and day silos. The crushers are designed to break down torrefied biomass pellets to a size not exceeding 35 mm. Biomass from the crushers is transferred by enclosed belt conveyors to the day silos, which are close to the gasification island. Sampling systems analyse both the as-received and the as-fired biomass to ensure the reliable and efficient operation of the plant. To control plant environmental emissions, all equipment is connected to bag filters and exhaust fans that permit the capture of any powder generated in the feed handling area.

The kaolin clay transport system to the gasification island consists of similar equipment to the biomass handling. Kaolin clay is added to the biomass feed before being fed to the mills for pulverisation.

15.3.2 Shell Biomass Gasification Package

This unit is mainly composed of following processes:

- Biomass Milling
- Biomass Pressurization and Feeding
- Gasification and Syngas Cooler
- Slag Removal
- Dry Solids Removal
- Wet Scrubbing
- Primary Waste Water Treatment

The Shell gasification package scope includes all process units listed above except biomass milling. In order to meet the Shell specification for particle size distribution, the biomass needs to be milled. Drying is not required as the moisture content of the torrefied biomass is as low as 3%. The key features of the Shell Gasification Process are the following:

- Pressurised system with compact equipment;
- Entrained flow slagging gasifier;
- Oxy-steam gasification leading to high gasification efficiency;
- Multiple burner design providing good mixing, high conversion, scale-up possibility;
- Dry feed of pulverised biomass providing high gasification efficiency, high feed flexibility.

Gasification

The biomass feed is gasified in the entrained flow slagging gasifier using a mixture of oxygen and superheated process steam. Due to the entrained flow, high temperature, and ash slagging condition, an almost complete carbon conversion (>99%) is achieved. The operating temperature of the gasifier zone is slightly below 1500°C. At this temperature, ash from the biomass is converted into molten slag, which runs down the gasifier walls to the slag removal zone, where it is contacted with water and solidifies. Slag forms a protective layer on the gasification membrane wall providing insulation, minimising heat losses and protecting the gasifier wall against high heat load variations during process upsets.

The operating syngas pressure of the gasifier is about 40 bar. Hot syngas from the gasifier is initially quenched with recycle syngas to approximately 750°C. The combined gas stream is then cooled in a syngas cooler to generate HP & MP steam. The syngas cooler is of the water pipe type, typically containing both evaporating and superheating surfaces.

Slag Removal

About 70-80% of the mineral content of the biomass ash leaves the gasification zone in the form of molten slag. Kaolin clay is used as an additive to the biomass feed because it acts as a moderator, affecting the ash fusion temperature of the biomass to ensure that the slag flows freely down the membrane wall. The heat from the molten slag is removed in the slag bath. The slag is non-leachable and non-hazardous.

Dry Solids Removal & Wet Scrubbing

The system consists of a high pressure / high temperature (HPHT) filter system that will remove 99.9% of the entrained solids in the syngas stream. The gas leaving the dry solids removal is further processed in a wet scrubbing system that consists of a venturi scrubber and a packed bed wash column. Residual solids, as well as the halide content of the syngas reduces to <<1 ppmv.

Primary Waste Water Treatment

The primary waste water treatment system contains one slurry stripper and a solid / liquid separation step. The system treats the bleed from slag bath and wet scrubbing systems.

15.3.3 Air Separation Unit (ASU)

The ASU capacity is defined by oxygen requirements of the gasification process. It supplies 95 mol% oxygen to the gasification island. The total required oxygen flow rate for the case is approximately 115 t/h.

The ASU also produces very high pressure and medium pressure nitrogen. The very high pressure nitrogen is for the gasification system, acting as carrier gas for the biomass feed pneumatic transport system whereas medium pressure nitrogen is used as a diluent for the syngas to the gas turbine for NOx control. The carrier nitrogen enters the gasifier with the biomass.

15.3.4 Syngas Treatment & Conditioning

Saturated raw syngas from the gasification wet scrubber unit, at approximately 40 barg, passes through the series of shift reactors where CO is shifted to H_2 and CO_2, and COS is converted to H_2S. A hybrid water gas shift (WGS) scheme has been used for this study. The scheme is used to minimise the steam consumption and amount of condensate flowing to the sour water stripper, achieving an overall CO conversion greater than 98%.

The first WGS reactor is a low steam shift reactor, converting about 35% of CO to CO_2. The catalyst for this reactor is designed to minimise the unwanted methanation reaction. This is followed by a conventional 2-stage sour shift reactor to convert the remaining CO. The hot shifted syngas from both the second and third shift reactors is cooled down in a series of heat exchangers for heat recovery steam generation. Final cooling of the syngas is made against clean syngas coming from the Acid Gas Removal (AGR) unit followed by cooling water before passing through a sulphur impregnated activated carbon bed to remove approximately 95% of the mercury. Cool, mercury-depleted syngas then enters the AGR unit.

Process condensate from syngas cooling is sent to the Sour Water Stripper in order to avoid accumulation of ammonia and other dissolved gases in the water recycle to the gasification section. Part of the condensate from the accumulator is sent to the Gasification Island, while the remaining condensate is sent to the Waste Water Treatment Unit.

15.3.5 Acid Gas Removal (AGR)

The AGR Unit removes CO_2 from the shifted syngas, using Selexol as physical solvent. Shifted syngas combined with the recycle stream from the Tail Gas Recovery Unit passes through three parallel CO_2 absorbers where CO_2 is removed from the gas by Selexol. Solvent regeneration is accomplished in the regenerator, where CO_2 is stripped from the liquid phase to the gas phase by steam. The CO_2 removal rate is approximately 92% of the carbon dioxide entering the unit, reaching an overall carbon capture of approximately 90%.

15.3.6 CO_2 Compression & Dehydration

This unit is mainly composed of a compression and dehydration package, followed by last stage CO_2 pumps, supplied by specialised vendors. Three different streams of CO_2 from the AGR unit are routed to the CO_2 compression unit where it is initially compressed to ~30 bara and dried to <50 ppmv water using a molecular sieve adsorption process. After dehydration, the CO_2 stream is finally compressed to a supercritical condition at 80 bara. The resulting stream of CO_2 is pumped to the required pressure of 110 bara ready for transportation.

15.3.7 Combined Cycle Power Generation

The combined cycle uses a state-of-the-art GE 9F, 50 Hz syngas variant gas turbine, commercially available for high hydrogen content gas. Due to high flame speed (flash back risk) and lower auto ignition delay time for hydrogen compared to natural gas, the pre-mix burner which is normally used for the natural gas-fired turbine can't be used for the syngas variant gas turbine. Also, the combustion of hydrogen-rich fuel leads to a high flame temperature and consequent high thermal NOx formation. Fuel dilution to the gas turbine is therefore necessary to meet the NOx emission limits. Hence, for gas turbines firing syngas and high hydrogen gas with lower LHV (on a volumetric basis), significant design changes have been adopted from the conventional natural gas fired gas turbines.

For syngas and high hydrogen gas, diffusion burners are used in place of a pre-mix burner and control of NOx is achieved by diluting the fuel with large quantity of nitrogen (nearly 5:1 of N_2:H_2 based on syngas mass basis). In addition, saturated nitrogen is injected directly into the gas turbine combustion chamber for final dilution to moderate the high flame temperature.

The decarbonised fuel gas is preheated in the syngas/syngas exchanger and against LP steam after being mixed with nitrogen coming from the ASU, up to maximum hydrogen content of 65 mole%. The gas turbine compressors provide combustion air to the burner only, i.e. no air integration with the ASU is foreseen. The exhaust gases from the gas turbine enter the HRSG at

560°C. The HRSG recovers heat available from the exhaust gas, producing steam at three different pressure levels for the steam turbine. The final exhaust gas temperature to the stack of the HRSG is ~80°C. The combined cycle is thermally integrated with the process unit, in order to maximise the net electrical efficiency of the plant.

15.4 Technical Performance Evaluation

Table 15-2: Technical Performance Comparison for Case 10

	Units	Reference Case (Unabated CCGT)	Coal IGCC with CCS	Biomass IGCC with CCS
Total Gross Installed Capacity	MWe	1229.4	1062.8	493.3
Gas Turbine (s)	MWe	823.5	671.0	302.9
Steam Turbine	MWe	405.9	391.7	190.4
Others	MWe	0	0	0
Total Auxiliary Loads	MWe	20.9	263.0	137.2
Feedstock Handling	MWe	0	0.4	0.3
Power Island	MWe	14.7	12.9	6.3
Air Separation Unit	MWe	0	106.4	48.9
CO_2 Capture	MWe	0	87.0	51.3
CO_2 Compression	MWe	0	45.7	24.0
Utilities	MWe	6.2	10.6	6.3
Net Power Export	MWe	1208.5	799.8	356.1
Fuel Flow Rate	kg/h	150,296	314,899	225,417
Fuel Flow Rate (LHV)	MWth	1940.2	2262.9	1051.9
Net Efficiency (LHV) - As New	%	62.3	35.3	33.9
Net Efficiency (LHV) - Average	%	59.0	33.5	32.1
Total Carbon in Feeds	kg/h	108,640	203,425	107,095
Total Carbon Captured	kg/h	0	183,697	97,194
Total CO_2 Captured	kg/h	0	673,147	356,162
Total CO_2 Emissions	kg/h	398,105	72,292	36,283
CO_2 Capture Rate	%	0	90.3	90.8
Carbon Footprint	kg CO_2/MWh	329.4	90.4	101.9

The plant performance of the torrefied biomass fired IGCC plant with carbon capture is summarised in the above table. The overall performance of the system includes the CO_2 balance and removal efficiency. The unabated CCGT Reference case along with the coal-IGCC case (Case 5) are also listed in the table for comparison purposes.

The bio-IGCC case is not expected to compete on a performance or economic basis with an unabated gas-fired power plant. The Reference case is included to allow the calculation of the Cost of CO_2 Avoided. Compared with the Reference case, the bio-IGCC case captures ~91% of the CO_2 while suffering a 28.4% point net efficiency loss. Further comparison of this case versus the Reference case is of little value for this report.

Both coal and biomass IGCC cases use Shell entrained flow gasification technology for the syngas production. They also both use the GE Frame 9 syngas variant gas turbine for the power plant. The process configurations are similar for both cases apart from the scale and that the bio-IGCC case does not require any sulphur treatment facility as there is negligible sulphur in the biomass feed. It is evident from Table 15-2 that the net plant efficiency for the two IGCC cases is comparable whereas carbon footprint is very different. The following points can be highlighted as basic differences between the two IGCC cases:

- Fuel thermal energy to the coal-IGCC power plant is 2.15 times higher than bio-IGCC power plant whereas the net power output from the coal case is 2.25 times higher than biomass case. This small difference leads to only 1.4% points higher net LHV efficiency for the coal case.
- This is mainly due to the relatively lower parasitic demand for the coal case leading to more power output for a given thermal input. The larger scale of the coal plant and supercritical operation also increase its efficiency.
- The energy density of the torrefied biomass feed to the gasifier is substantially higher than the untreated biomass used for Cases 8 & 9 due to the additional processing step of torrefication. The supply cost is also higher due to this additional step.
- The coal-IGCC plant with 90% carbon capture emits 72 t/hr of CO_2 to the atmosphere. The biomass power plant emits 36 t/hr of carbon dioxide and will need to pay the resultant carbon price. However, if the overall carbon chain is taken into account this process is carbon negative.
- The carbon footprint for the biomass appears is slightly higher than for the coal IGCC case at 101.9 kg CO_2/MWh, but as noted above, the overall process captures carbon dioxide from the atmosphere for long-term storage.

15.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staff are taken to be in daily positions, working regular hours.

Table 15-3: Operations and Maintenance Staff Manning for Case 10

	Reference Case Unabated CCGT	Coal IGCC with CCS	Bio-IGCC with CCS	Remarks
Operations Staff				
Plant Manager	1	1	1	Daily Position
Deputy Plant Manager	1	1	1	Daily Position
CO_2 Removal Area Manager	NA	1	1	Daily Position
Process Engineer	NA	5	4	Daily Position
Shift Supervisor	5	10	10	3-shift Position
Electrical Assistant	5	10	10	3-shift Position
Control Room Operator	10	25	20	3-shift Position
Field Operator	10	50	45	3-shift Position
Sub-Total	32	103	92	
Maintenance Staff				
Mechanical Group	3	8	7	Daily Position

	Reference Case Unabated CCGT	Coal IGCC with CCS	Bio-IGCC with CCS	Remarks
Instrument Group	3	8	8	Daily Position
Electrical Group	2	6	6	Daily Position
Sub-Total	8	22	21	
Laboratory Staff				
Superintendent	1	1	1	Daily Position
Analysts	3	7	6	Daily Position
Sub-Total	4	8	7	
Plant Total Staff	44	133	120	

Table 15-4: Economic Performance Comparison for Case 10

	Units	Reference Case Unabated CCGT	Coal IGCC with CCS	Biomass IGCC with CCS
Total Project Cost	£M	672.2	2,396.3	1,465.4
Specific Total Project Cost	£/kW	556	2,996	4,115
Pre-Development Costs				
Pre-Licensing & Design	£M	5.8	21.5	13.0
Regulatory & Public Enquiry	£M	12.9	44.2	27.3
EPC Contract Cost	£M	583.6	2,151.0	1,304.7
Feedstock Handling	£M	5.8	71.1	43.8
Power Island	£M	12.9	1,349.6	829.2
Air Separation Unit	£M	5.8	240.4	125.1
CO_2 Capture	£M	12.9	82.4	54.4
CO_2 Compression	£M	5.8	67.1	45.5
Utilities	£M	12.9	340.4	206.6
Other Costs				
Infrastructure Connections		29.0	29.0	29.0
Owner's Costs		40.9	150.6	91.3
Overall CAPEX Impact (vs Ref Case)	£M	-	1,724.1	793.2
Overall CAPEX Impact (vs Ref Case)	%	-	256	118
Total Fixed OPEX	£M pa	36.2	112.2	70.5
Total Variable OPEX (excl. Fuel / Carbon)	£M pa	0.2	103.5	54.2
Average Fuel Cost	£M pa	315	131	183
Average CO_2 Emission Cost	£M pa	369	61.2	30.7
Total Start-up Cost (excl. Fuel)	£M	4.4	16.8	10.8

	Units	Reference Case Unabated CCGT	Coal IGCC with CCS	Biomass IGCC with CCS
Discount Rate	% / year	7.8	8.9	8.9
Levelised Cost of Electricity (incl. Carbon Price)	£/MWh	74.2	120.8	204.3
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	195.1	571.7
Levelised Cost of Electricity (zero Carbon Price)	£/MWh	45.5	113.3	195.8
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	283.8	660.7

The economic performance of the commercial scale bio-IGCC system with carbon capture is summarised in the Table 15-4 along with unabated CCGT Reference case and coal-IGCC case (Case 5) for the purposes of comparison. The capital cost estimate for Case 10 is assessed to have an accuracy of ± 40%.

The total project cost for this case is more than double the Reference unabated natural gas case while producing 70% less net power output. The total project cost for the biomass case is 39% lower than the coal case while producing 55% less net power. This equates to a higher capex intensity for the biomass case than the coal case.

The operating costs (excluding fuel and carbon emissions costs) for the biomass power plant with capture plant is only 42% lower than the coal plant, demonstrating that the cost of running the biomass plant has more profound impact on LCOE than for the coal case.

The cost of torrefied biomass fuel is 5.8 times higher than coal on a thermal basis. The combination of higher fuel cost, higher capital and operating cost, along with the lower overall efficiency for the bio-IGCC make a significant impact on the LCOE. However, no financial benefit is included for the positive impact of negative emissions with 90% capture of the process CO_2 from the ‘CO_2 neutral’ renewable fuel source.

Figure 15-2 shows the list of the different contributing factors and the level of contribution towards the overall LCOE value. It is important to note that the torrefied biomass cost is the biggest contributor to the LCOE, closely followed by capital investment. CO_2 storage and transportation, and operating cost are also significant cost elements. As with the other biomass cases, the proportion of costs associated with capital costs is lower than for the coal IGCC case.

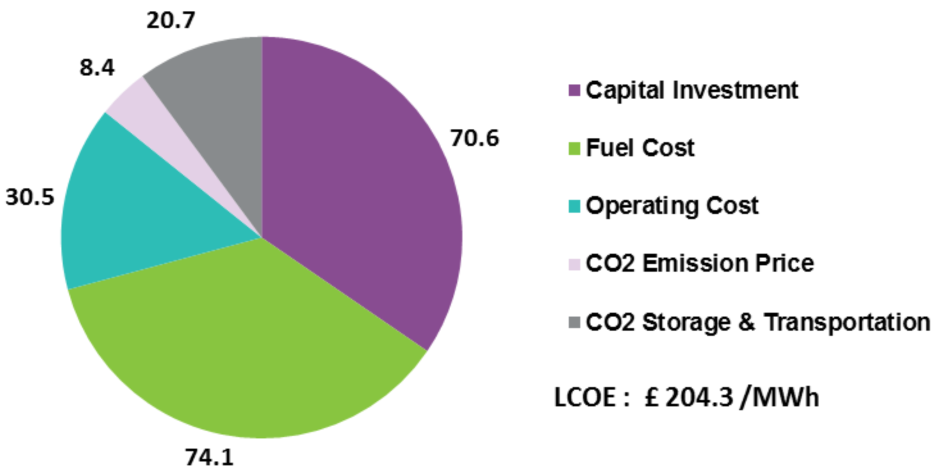

Figure 15-2: LCOE (£/MWh) Contribution for Case 10

16 Case 11 – Natural Gas Steam Methane Reformer with Post-Combustion Carbon Capture

16.1 Overview

This case consists of a natural gas fed steam methane reformer (SMR) followed by two stage shift then pressure swing adsorption (PSA) for production of 100,000 Nm3/h of 99.99% purity hydrogen. The PSA tail gas is combined with additional natural gas to fire the reformer. Heat recovery from the hot syngas is used to generate the steam required for the reforming and shift reactions as well as to provide steam and power for the CO_2 capture unit and compressors via a steam turbine generator. Post combustion carbon capture is applied to the reformer flue gas. The captured CO_2 is then compressed in 4 stages, dehydrated, condensed and then pumped to the required export pressure of 110 bar (abs).

The capacity of 100,000 Nm3/h fits well with a single train Wood terraced-wall steam methane reformer (SMR). Larger units can easily be created through use of multiple reformer trains and may achieve an economy of scale in common units such as the PSA.

The product purity of 99.99% hydrogen is targetted at industrial users within the oil refining and petrochemical industries. A hydrogen plant designed to feed a decarbonised gas network for domestic and commercial users would likely use a lower degree of purity with some associated cost savings. However, that was not selected as the focus for this study.

16.1.1 Process Configuration

The main process configuration of the SMR plant is as follows:

- Steam Methane Reforming of natural gas with steam;
- Two stages water gas shift reaction process;
- Pressure Swing Adsorption (PSA) for carbon removal and hydrogen purification;
- CO_2 capture using Cansolv chemical solvent system;
- CO_2 compression and pumping up to 110 bara;
- Heat recovery and steam turbine generator to provide electrical power.

Table 16-1 describes the process units with trains which are also shown in Figure 16-1.

Table 16-1: SMR Process Units with Trains

Unit Number	Unit Description	Trains
100	Fuel Pre-treatment & Pre-reformer	1 x 100%
200	Auto-thermal Reforming & Shift Process	1 x 100%
300	Pressure Swing Adsorption	1 x 100%
400	CO_2 Compression & Dehydration	2 x 50%
500	Cansolv CO_2 Capture Unit	1 x 100%
600	Heat Recovery Steam Generation	1 x 100%
700	Steam Turbine & Generator Package	1 x 100%
800	Offsites & Utilities	

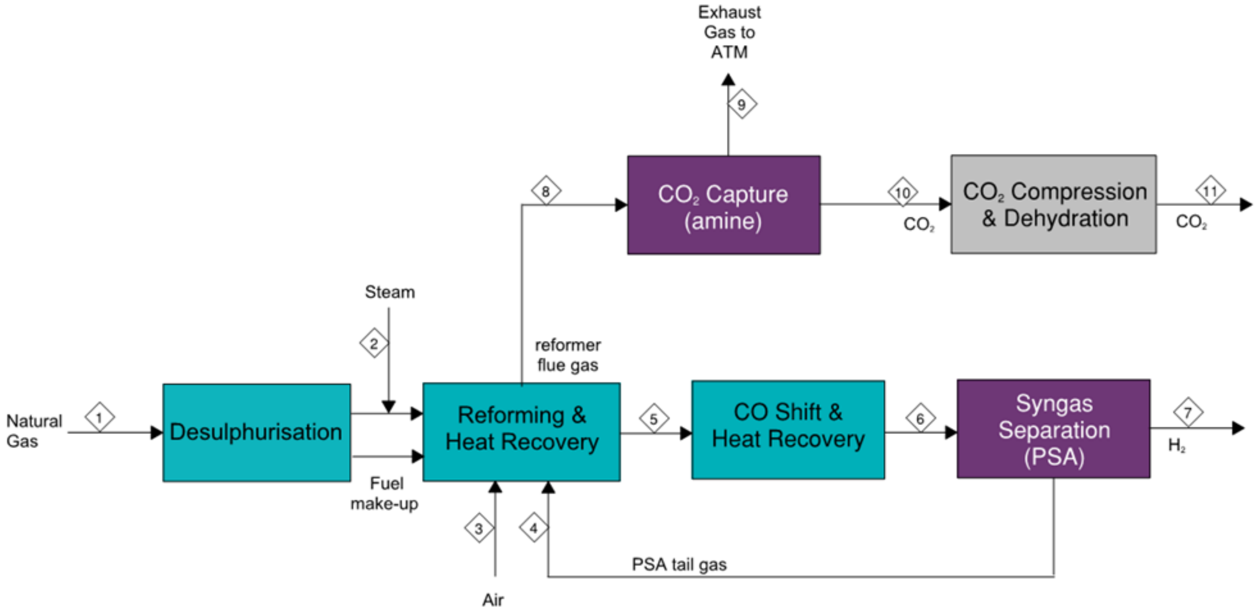

Figure 16-1: Case 11 Block Flow Diagram

16.2 Model Development

The SMR and PSA units were modelled by adapting an existing Hysys model of an SMR hydrogen plant, which only required modification of the feed gas and the capacity (100,000 Nm³/h = 9000 kg/h). Since Wood licenses an SMR-based process, this was available in house.

The flue gas flowrate, composition and conditions were then supplied to Shell Cansolv who provided a process design package using its process to capture 90% of the CO_2 from the reformer flue gas. A black-box model of the Cansolv process was then added to the Hysys model, allowing the CO_2 compression and dehydration, heat integration, steam and power requirements to be modelled.

Due to the large amount of high grade heat required and produced by an SMR hydrogen unit, such units are usually key sources of high pressure steam (30-40 bar) on an oil refinery. Thus, it was anticipated that it could be possible to utilise this excess steam generation capability to produce the steam and power required to run the CO_2 capture and compression units (and the other smaller utility loads of the SMR and PSA). It was found that it was possible to meet the steam and power requirements of the whole plant in this way, by using two stages of steam turbine generator, the first letting down steam to the pressure required for the CO_2 capture unit reboiler, and the second, much smaller machine, to produce a small amount of extra power by letting the remaining steam down to atmospheric pressure.

16.3 Process Description

16.3.1 Natural Gas Reforming

Natural Gas is received from the grid and metered on entry to the plant. It is then 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 any organic sulphur into H2S and then through a zinc oxide H_2S adsorbent. Reforming catalysts are severely poisoned by sulphur and so it must be completely removed upstream.

Desulphurised feed from the pre-treating section is heated then mixed with superheated steam at a steam to carbon ratio of 2.7, then heated further 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 reformer outlet temperature is 860°C. Process gas (raw syngas) at the outlet of the reformer is cooled down in a waste heat boiler to generate steam and then routed to the downstream Shift Conversion Unit.

The overall effect of reactions is endothermic and the required 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 reformer is cooled down in a series of coils in the reformer convection section. The convection section heating duties comprise reformer mixed feed heater, MP steam superheater, MP steam generator and combustion air preheater.

16.3.2 Shift and Heat Recovery

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 at 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, operating 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. Heat is recovered down to a syngas temperature of 110ºC, with the final cooling stage provided by cooling water down to 25ºC.

16.3.3 Syngas Separation

Much of the water contained in the syngas is knocked out, removed and recycled back into the process as the syngas is cooled. The cooled, shifted syngas is routed to a PSA unit where pure H_2 is produced by separation from methane, water, 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 99.99% pure hydrogen product is produced at 24 bar (abs) from the PSA and is ready for export.

16.3.4 Proprietary Solvent CO_2 Capture

The reformer is equipped with an induced draft fan which provides the pressure required to exhaust the flue gas. In this case, it is assumed that the discharge pressure of this fan is increased to ensure that the pressure drop through the CO_2 capture unit is overcome. This increases the anticipated load on the fan by a factor of five.

NOx removal is required upstream of the CO_2 capture process. To achieve this a Selective Catalytic Reduction (SCR) system is included between the reformer and the CO_2 capture unit. Ammonia, the reducing agent, is injected immediately upstream of a catalyst surface where the NO and NO_2 are reduced to N_2 and water.

An outline of the Shell Cansolv process as applied to a natural gas fired combined cycle plant is shown in Figure 16-2 below. The same process (at a smaller scale) is applicable for CO_2 removal from the reformer flue gas.

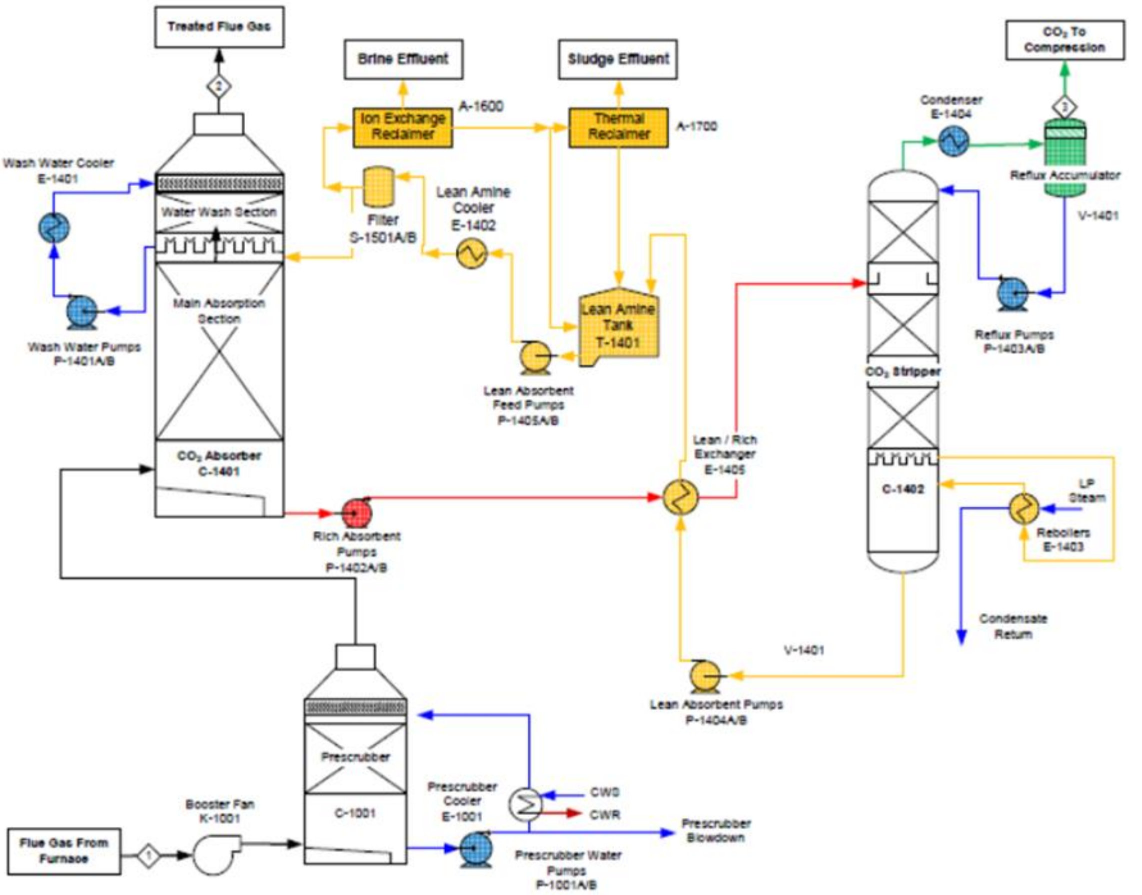

Figure 16-2: Cansolv CCGT Process Configuration
(Image courtesy of Cansolv Technologies Inc.)

Flue gas from the SCR is ducted to a pre-scrubber column which performs the dual function of sub-cooling (with water knock-out) and SO_2 removal. The flue gas is sub-cooled to 35°C to reduce the required absorbent circulation rate and thus energy consumption and CAPEX of the Cansolv unit.

In order to decrease the impact of SO_2 on the absorbent, the pre-scrubber will use caustic to reduce the SO_2 content in the flue gas upstream of the CO_2 Absorber. SO_2 removal is controlled by adding caustic on pH control in a caustic polishing section inside the pre-scrubber column. For the purposes of this study, it is assumed that the concentration of SO_2 leaving the pre-scrubber polishing section would be 1 ppmv.

The cooled and pre-scrubbed flue gas is ducted to the bottom of the absorption column where it is contacted with the proprietary solvent. For a plant of this scale and flue gas type, typical absorber dimensions range from 5m to 10m in square cross section. CO_2 absorption from the flue gas occurs by counter-current contact with Cansolv Absorbent DC-201 in the CO_2 Absorber which is a vertical multi-level packed-bed tower. CO_2 is absorbed into the solvent by chemical reaction leaving a flue gas depleted in CO_2 at the top of the column. The absorption reaction is exothermic, and the high concentration of CO_2 in the reformer flue gas (16 mol%) means that an intercooler is required to maintain a reasonable temperature profile in the absorber in this case.

The treated flue gas passes through a water wash section in order to prevent emissions of solvent and any solvent degradation products such as nitrosamines. The treated flue gas is then warmed against hot condensate from the Regenerator Reboilers and routed to a stack for discharge to the atmosphere.

The CO_2 rich absorbent is collected in the bottom sump of the CO_2 Absorber and is pumped by the CO_2 Rich Absorbent Pumps and heated in the CO_2 Lean / Rich Exchangers to recover heat from the hot lean absorbent discharged from the CO_2 Regenerator. The rich absorbent is piped to the top of the CO_2 Stripper for absorbent regeneration and CO_2 recovery. The rich absorbent enters the column under the CO_2 top packing section and flows onto a gallery tray that allows for disengagement of any vapour from the rich absorbent before it flows down to the two stripping packing sections under the gallery tray. The rich absorbent is depleted of CO_2 by water vapour generated in the Regenerator Reboilers which flows in an upward direction counter-current to the rich absorbent.

Lean absorbent flowing from the bottom packing section of the CO_2 Regenerator is collected on a chimney tray and gravity fed to the Regenerator Reboilers. Water vapour and lean amine flow by thermosyphon effect from the reboilers back to the CO_2 Regenerator sump, underneath the chimney tray. Water vapour flows upwards through the chimney tray to strip the CO_2 while the lean absorbent collects in the bottom sump.

Water vapour in the regenerator, carrying the stripped CO_2, flows up the regenerator column into the top packing section, where a portion of the vapour is condensed by recycled reflux to enrich the overhead CO_2 gas stream. The regenerator overhead gas is partially condensed in the Regenerator Condensers. The partially condensed two phase mixture gravity flows to the CO_2 Reflux Accumulator where the two phases separate. The reflux water is collected and returned via the Reflux Pumps to the regenerator rectification section. The CO_2 product gas is piped to the CO_2 Compression System. The pressure of the Regenerator can either be controlled by a product CO_2 discharge control valve or by the inlet guide vanes of the downstream CO_2 Compressors.

The flow of steam to the reboiler is proportional to the rich absorbent flow sent to the CO_2 Regenerator. The set-point of the low pressure steam flow controller feeding the Regenerator Reboilers is also dependent on the regenerator top temperature controller. The steam to absorbent flow ratio set-point is adjusted by this temperature controller. The temperature at the top of the column is set to maintain the required vapour traffic and stripping efficiency. The steam flow rate is controlled by modulating a steam flow control valve.

All amine based systems require some form of solvent maintenance system as over time the absorbent in the CO_2 Capture System accumulates Heat Stable Salts (HSS), as well as non-ionic amine degradation products, that must be removed from the absorbent. This is achieved through thermal reclamation. An ion exchange package is included for bulk HSS removal upstream of a thermal reclaimer.

The ion exchange package is designed to remove Heat Stable Salts (HSS) from the Cansolv DC Absorbent. These salts are continuously formed within the absorbent, primarily due to residual amounts of NO_2 and SO_2 contained in the flue gas. Once absorbed, NO_2 forms nitric and nitrous acid while SO_2 forms sulphurous acid which oxidizes to sulphuric acid. These acids, and some organic acids formed by the oxidative degradation of the amine, neutralise a portion of the amine, which is then inactivated for further CO_2 absorption.

The purpose of the Thermal Reclaimer Unit is to remove the non-ionic degradation products as well as HSS from the active absorbent. The thermal reclaimer unit distils the absorbent under vacuum conditions to separate the water and amine, leaving the non-ionic degradation products in the bottom. A slipstream is taken from the treated CO_2 lean absorbent exiting the ion exchange package and fed to the Thermal Reclaimer Unit. This stream will essentially consist of water, amine, degradation products, residual CO_2 and small amounts of sodium nitrate and sodium sulphate. The design flow rate of CO_2 lean absorbent sent to the thermal reclaimer is based on the calculated amine degradation rate. To maintain the degradation products below design concentration, the thermal reclaimer must process a specific flowrate of CO_2 lean absorbent. The reclaimed absorbent is sent to the Lean Absorbent Tank. The separated degradation products are stored in a storage tank, where they are diluted and cooled with process water. Diluted residues are periodically disposed of offsite, typically via incineration.

16.3.5 CO_2 Compression and Dehydration

The CO_2 is compressed to 30 barg in 3 stages, each with intercooling and water knock-out. This recovers the vast majority of the water content, but is not sufficient for most pipeline specifications. Numerous studies have compared drying with tri-ethylene glycol (TEG) versus use of molecular sieve adsorption which conclude that there is little to choose between the two methods. For the purposes of this study we have assumed a TEG dehydration unit is selected, consistent with the natural gas CCGT benchmark cases.

Final CO_2 pressurisation to 110 bar (abs) is achieved using one further stage of compression, followed by a condenser then a stage of pumping.

16.3.6 Heat Recovery, Steam and Power Generation

The reforming reaction requires a significant quantity of superheated steam at around 32 bar (abs) and 320ºC. This is mostly generated via water heating downstream of the shift, further heating between the two shift stages and steam generation taking place upstream of the shift in the waste heat boiler. Some further steam is also generated in the reformer convection section and all of the steam is then superheated in the reformer convection section.

Roughly 42% of the 35 bar (abs) superheated steam is required for the reforming process and is added to the process stream upstream of the reformer feed heater in the reformer convection section.

The remaining superheated steam is fed to a steam turbine generator which lets the steam down to LP steam at 5 bar (abs) while generating electrical power. About 90% of the LP steam is then used in the CO_2 stripper reboiler with the remaining steam let down to near atmospheric pressure in a further steam turbine stage. These two turbines provide the total power required for steady state operation of the plant.

Alternative heat integration and power provision options would include potential use of cooling syngas as the heat source in the stripper reboiler and treated flue gas reheater, and direct use of a steam turbine to drive the CO_2 compressor instead of using electricity as an intermediate energy vector. These options should be considered when undertaking a more detailed design to ensure that the optimum configuration is selected.

16.4 Technical Performance Evaluation

The following table compares the performance of the Benchmark 11 hydrogen plant with an unabated hydrogen plant in which the excess steam is used to generate a small amount of electricity which is exported to the grid.

Table 16-2: Technical Performance Comparison for Case 11

	Units	Reference Case (Unabated SMR)	Natural Gas SMR with Cansolv CCS
Electrical Power Production	MWe	18.5	12.3
Steam Turbine	MWe	18.5	12.3
Others	MWe	0	0
Total Auxiliary Loads	MWe	2.8	12.3
Syngas Production	MWe	0.6	2.2
CO_2 Capture	MWe	0	7.5
CO_2 Compression	MWe	0	Incl. in CO_2 Capture
Utilities	MWe	2.2	2.6
Net Power Import (Export)	MWe	(15.7)	0.0
Feed Flow Rate	kg/h	34,578	34,578
Feed Flow Rate (LHV)	MWth	446	446
Feed Flow Rate (HHV)	MWth	494	494
Product Flow Rate	kg/h	8,994	8,994

	Units	Reference Case (Unabated SMR)	Natural Gas SMR with Cansolv CCS
Product Flow Rate	Nm3/h	100,000	100,000
Product Flow Rate (LHV)	MWth	300	300
Product Flow Rate (HHV)	MWth	351	351
Hydrogen Yield	%	66.6	66.6
Net Efficiency (LHV)	%	71.8	67.2
Net Efficiency (HHV)	%	75.7	70.9
Total Carbon in Feeds	kg/h	24,994	24,994
Total Carbon Captured	kg/h	0	22,527
Total CO_2 Captured	kg/h	0	82,550
Total CO_2 Emissions	kg/h	91,590	9,040
CO_2 Capture Rate	%	0	90.1
Carbon Footprint	kg CO_2/kNm^3	916	90.4

The following points can be highlighted as differences between the two cases:

- The abated CO_2 case shows a much higher electrical power requirement, about half of which is due to the CO_2 compressor, with a further significant load required to provide for the increased load on the reformer flue gas fan to overcome the additional pressure drop due to the post-combustion CO_2 capture system.
- Adding the CO_2 capture system also adds a significant steam load to the plant, which can be seen in the reduced output from the steam turbine in Case 11. The unabated reference case uses the additional steam to generate exportable power, but could also be exported to local users depending on the location.
- The two cases use the same quantity of feedstock to make the same quantity and quality of product, thus have the same hydrogen yield (moles of hydrogen produced per mole of hydrogen contained in the feed gas and net water consumed). However, since the unabated case also exports power, this results in higher thermal efficiency for the unabated case at 71.8% compared to 67.2% in Case 11.
- The high carbon footprint of hydrogen production from natural gas, 916 kgCO_2/kNm^3 can be reduced very effectively by adding the Cansolv post-combustion CO_2 capture unit, to achieve a footprint of 90 kgCO_2/Nm^3.

The benchmark SMR process with post-combustion CCS requires a water make up stream of approximately 195 tph to generate steam for reforming and shift and for make-up to the cooling water system.

16.5 Economic Performance Evaluation

The capital and operating cost methodology used for the cost estimation, economic modelling and calculation for this case has been described in Sections 5.4 and 5.5. The number of staff required to operate and maintain the plant has been listed in the following table. A daily pattern of three 8-hour shifts has been assumed, with two shift teams on leave at any time, resulting in five shift teams. Other staff are taken to be in daily positions, working regular hours.

Table 16-3: Operations and Maintenance Staffing for Case 11

	Reference Case (Unabated SMR)	Natural Gas SMR with Cansolv CCS	Remarks
Operations Staff			
Plant Manager	1	1	Daily Position
Process Engineer	NA	1	Daily Position
Control Room Operator	5	5	3-shift Position
Field Operator	5	5	3-shift Position
Sub-Total	11	12	
Maintenance Staff			
Mechanical Group	1	1	Daily Position
Sub-Total	1	1	
Laboratory Staff			
Analysts	1	2	Daily Position
Sub-Total	1	2	
Plant Total Staff	13	15	

Table 16-4: Economic Performance Comparison for Case 11

	Units	Reference Case (Unabated SMR)	Natural Gas SMR with Cansolv CCS
Total Project Cost	£M	144.1	237.3
Pre-Development Costs			
Pre-Licensing & Design	£M	1.3	2.1
Regulatory & Public Enquiry	£M	2.7	4.5
EPC Contract Cost	£M	127.4	207.2
Other Costs			
Infrastructure Connections		3.8	9.0
Owner's Costs		8.9	14.5
Overall CAPEX Impact (vs Ref Case)	£M	-	93.2
Overall CAPEX Impact (vs Ref Case)	%	-	65
Total Fixed OPEX	£M pa	7.4	10.7
Total Variable OPEX (excl. Feed & Carbon)	£M pa	0.7	14.3
Average Feed Cost	£M pa	72.6	70.2
Average CO_2 Emission Cost	£M pa	84.9	8.1
Total Start-up Cost (excl. Fuel)	£M	0.5	1.2
Discount Rate	% / year	7.8	8.9

	Units	Reference Case (Unabated SMR)	Natural Gas SMR with Cansolv CCS
Levelised Cost of Hydrogen (incl. Carbon Price)	£/kNm^3	194.9	165.8
Levelised Cost of Hydrogen (incl. Carbon Price)	£/kW$_{th}$	65.0	55.3
Cost of CO_2 Avoided (incl. Carbon Price)	£/tCO_2	-	-35.2
Levelised Cost of Hydrogen (zero Carbon Price)	£/kNm^3	118.8	158.6
Levelised Cost of Hydrogen (zero Carbon Price)	£/kW$_{th}$	39.6	52.9
Cost of CO_2 Avoided (zero Carbon Price)	£/tCO_2	-	48.3

The economic performance of the natural gas fed SMR plant with Cansolv post-combustion CCS is summarised in Table 16-4 along with an equivalent unabated case for comparison. The capital cost estimate for Case 11 is assessed to have an accuracy of ± 30%.

The total project cost for the Case 11 abated case is ~65% higher than the Reference unabated case, with other key economic results highlighted below:

- The Cansolv system adds to the total project capital cost, with the large low pressure absorber tower being a significant individual cost item. However, combining one wall of the direct contact cooler with the absorber results in significant cost savings versus previous designs. The CO_2 compressor is also a significant individual item cost.
- The operating costs (excluding feedstock) are more than three times the operating costs of the reference plant, which demonstrates the cost of running the more complex plant and the cost of CO_2 transportation and storage for a process with a high intrinsic carbon footprint per unit of product.
- Despite the capital and operating costs (excluding feeds) being higher for this case than the unabated case, the Levelised Cost of Hydrogen (LCOH) is lower, at £165.8 / kNm³ compared with £194.9 / kNm³ in the reference case. This is due to the emissions penalty being applied in the unabated case.

It is important to note from the figure below that the feed gas cost is by far the biggest contributor to the LCOH followed by capital investment. Operating cost and cost of CO_2 transportation and storage are somewhat less significant contributors by comparison. CO_2 emission price has lowest impact on the LCOH calculation.

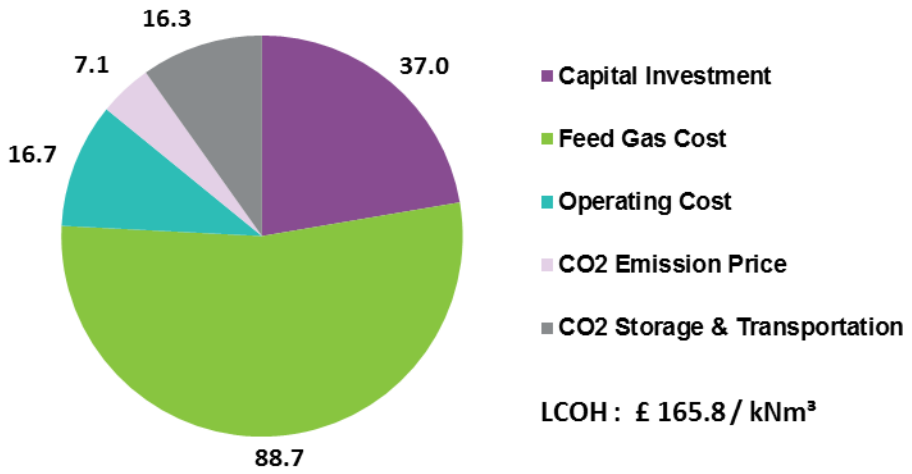

Figure 16-3: LCOH (£/kNm3) Contribution for Case 11

By contrast, the figure below shows the LCOH breakdown for the unabated hydrogen production SMR case, where it can be seen that the cost penalty for emitting CO_2 is almost as significant in the calculation of LCOH as is the fuel cost.

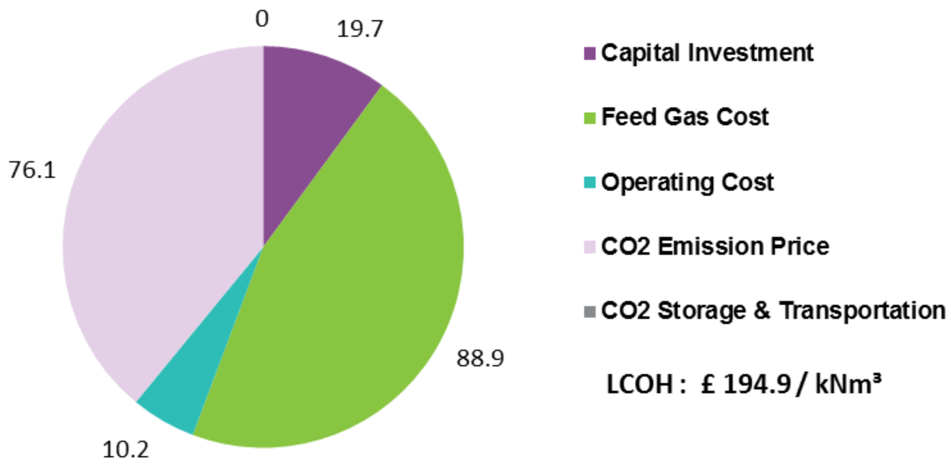

Figure 16-4: LCOH (£/kNm3) Contribution for Reference Hydrogen Case H

17 Power Case Comparison

Each case has been compared with the technical and economic performance of an unabated combined cycle gas turbine power plant, since this is the type of plant currently most widely used for power generation in the UK. In this section, the economic performance of all cases presented in this report will be compared with each other.

The following graph shows how the main components of the Levelised Cost of Electricity (LCOE) compare for all ten cases along with the net power output for each case, since these vary considerably from case to case.

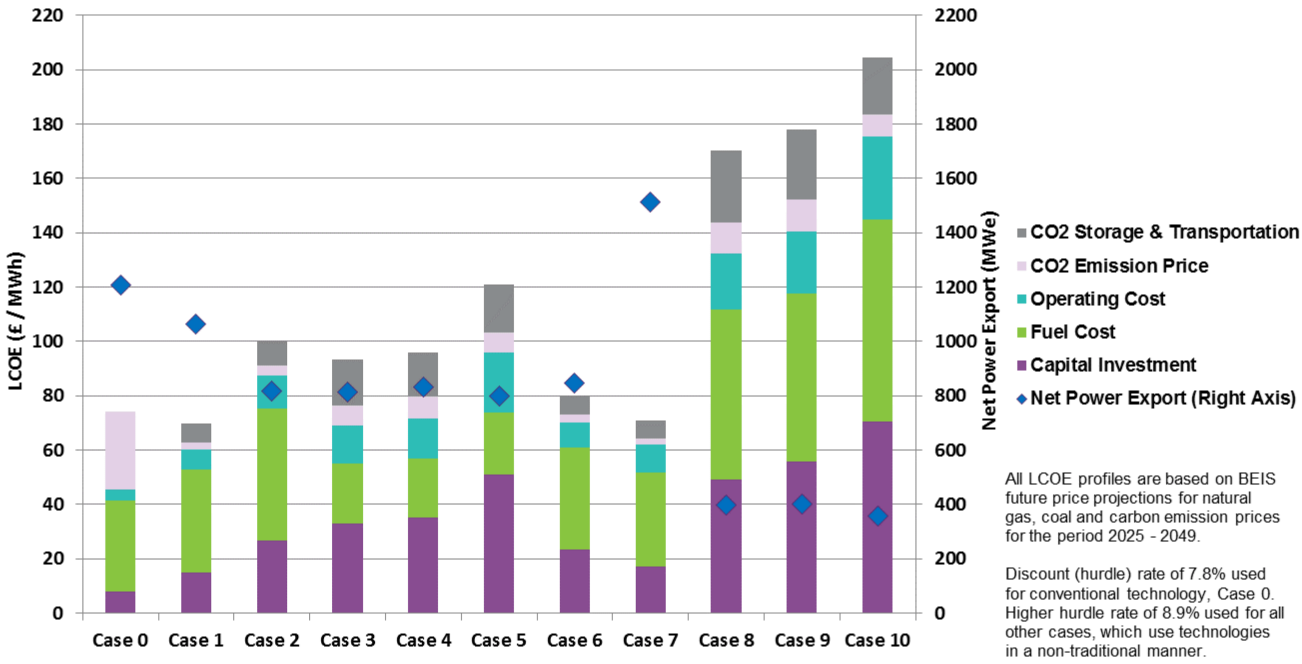

Figure 17-1: LCOE (£/MWh) Contribution for all Power Cases

LCOE Contribution (£ / MWh)	Capital Cost	Fuel	Operating Cost	Emissions Price	Storage & Transport	Total
Case 0	8.0	33.5	4.0	28.7	0	74.2
Case 1	14.9	37.9	7.2	2.9	7.0	69.9
Case 2	26.6	48.5	12.2	3.8	8.9	100.0
Case 3	32.9	22.2	13.8	7.5	16.9	93.3
Case 4	35.3	21.7	14.7	8.0	16.3	96.0
Case 5	51.1	22.8	22.0	7.5	17.4	120.8
Case 6	23.2	37.7	9.2	3.1	6.9	80.1
Case 7	17.1	34.7	10.1	2.3	6.5	70.7
Case 8	49.0	62.6	20.6	11.7	26.2	170.1
Case 9	55.9	61.8	22.8	11.6	25.8	177.9
Case 10	70.6	74.1	30.5	8.4	20.7	204.3

The stacked bars in Figure 17-1 represent the aggregated contributions of different costs towards an overall Levelised Cost of Electricity (LCOE) for each case. If all of the cases had roughly the same net power export, then the same comparison could be performed on the actual costs for each element. However, when considering the LCOE, it must be recognised that a plant with greater power export will see a reduced LCOE for an equal cost in any category. Thus, the figure also shows a diamond representing the net power export, as shown against the right-hand axis.

Note that the fuel and carbon price assumed for all cases is based on BEIS future projections as shown in Table 5-7. Thus, in the first year of operation (2025), the carbon price is taken as £22.60 / tonne CO_2 (real terms, 2017), with significant real terms growth throughout the 25-year life of plant.

17.1 Natural Gas Cases

Although the unabated CCGT case, Case 0, has the lowest overall investment cost, it does not result in the lowest overall LCOE. The lowest overall LCOE is provided by Case 1, the CCGT plant with state-of-the-art post-combustion carbon capture. Case 0 features a significant proportion of LCOE arising from the penalty paid for emitting CO_2, which is included in the financial analysis for this study, demonstrating the importance of the carbon price as a potential tool for encouraging low carbon investments in power plant. Please note, in Table 2-3 there are two cost of avoided CO_2 metrics: one that includes the effect of a carbon price, and one that doesn't include a carbon price as this allows the later metric to be compared to the methodology used by other international benchmarking studies. The cost of avoided CO_2 metric that is of relevance to UK (and other countries/regions with a price on CO_2) is the one that includes the effect of a carbon price.

The natural gas fired power plant with integrated reforming and pre-combustion carbon capture, Case 2, does not compete well against the more straight-forward post-combustion case (Case 1). This case has higher capital and operating costs than Case 1, and the power output available for electricity sales is also significantly lower, despite having approximately the same gas feed rate. This result has been seen in other comparative studies. The approach of using natural gas reformation with pre-combustion carbon capture appears to lack promise as a basis for standalone power plant developments. Its strength lies in facilities that require reformed hydrogen as part of a larger refining or petrochemical facility (refer to Benchmark Case 11) and which can produce and store excess hydrogen for peak-shaving power plants. Natural gas reforming processes may also provide a route to decarbonisation of the gas distribution vector, which is used widely for domestic and commercial heating in the UK. There is no way of accounting for those potential economic benefits which would make the IRCC case attractive within the scope of analysis included in this study.

17.2 Natural Gas versus Coal

The three coal cases, Cases 3-5, all show significantly higher LCOE than the post-combustion natural gas case. Comparing the two post-combustion cases (Case 1 vs Case 3) or the two integrated reformation / gasification cases (Case 2 vs Case 5) it can be seen that the capital and operating costs of the coal cases are much higher: this trend is accentuated in the LCOE figures by the lower power output of the coal cases. The cost contributions for CO_2 storage and carbon emissions are also significantly greater for the coal cases, simply because coal delivers far more of its energy in the form of carbon, with less contribution of hydrogen than is present in natural gas. Carbon dioxide emissions for the coal cases are approximately double for the coal cases, even though all of the cases capture 90% of the CO_2 generated.

One benefit that has historically been enjoyed by coal is a very low price base compared to natural gas, and it is noticeable the contribution of fuel cost to LCOE for the three coal cases is lower than for the natural gas cases. Prior to the shale gas revolution, natural gas prices commanded a significant premium and thus the fuel costs shown in this report are lower than in the IEAGHG 2012/08 report. However, some may suggest that this analysis may have been skewed by an over-optimistic gas price forecast versus an overly pessimistic coal price. Figure 17-2 below considers the impact on LCOE for each of the cases if the high or low price forecast sets shown in Table 5-7 were used, in place of the central forecasts used for the results above.

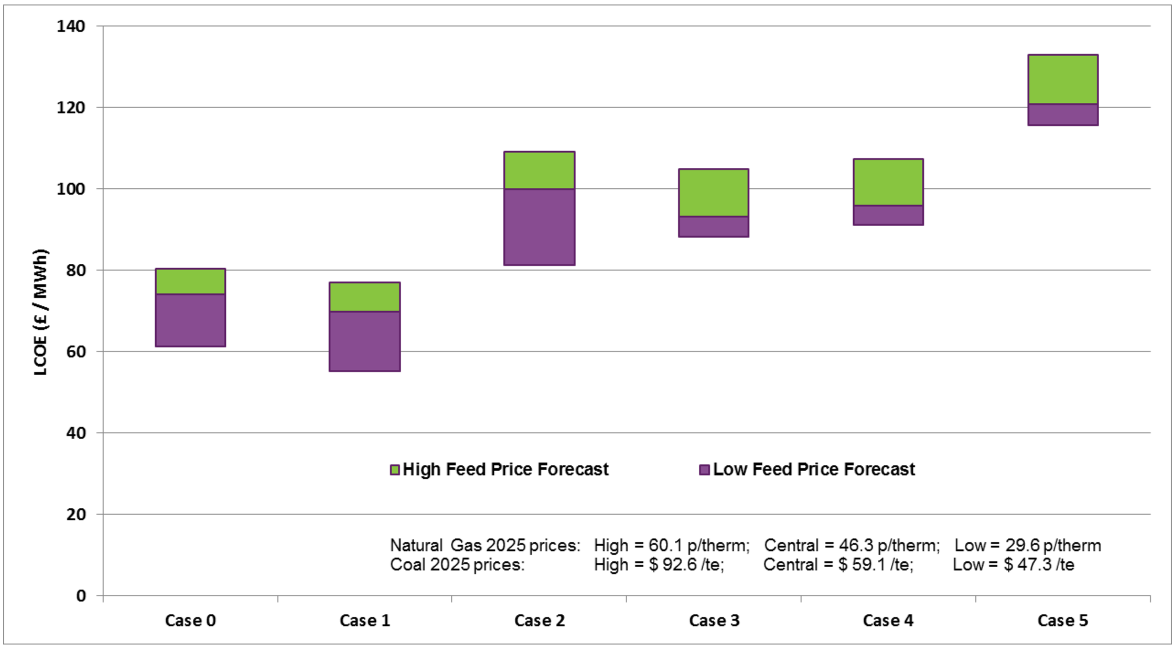

Figure 17-2: Sensitivity of LCOE to Fuel Price Forecast

In Figure 17-2, the lowest end of the bars represents the low fuel price forecast and the uppermost extent of the bars represents the high fuel price forecast. The middle point, where the colour of the bar changes represents the central fuel price forecast: the same values as the aggregate LCOE in Figure 17-1. This indicates that even with the highest gas price and lowest coal price, the cost of electricity for the coal cases would still need to much higher for these projects to break even.

It should be noted that these findings might be different for other countries where the relative prices of the two fuels are very different, and for financial scenarios in which the flexibility to be able to produce hydrogen when power is less in demand is of value.

The results for the coal oxy-combustion case (Case 4) are close to those for the conventional post-combustion technology (Case 3) indicating that for a country where coal is considerably cheaper than gas, oxy-combustion technology remains a viable alternative that should be considered during the feasibility stage of any power project. The IGCC case (Case 5) should also be included in the feasibility analysis if there is likely to be a significant benefit from the ability to produce hydrogen.

17.3 International Benchmarks

When compared to the state of the art post-combustion case, the two novel international technologies appear to show promise, as their designs are less mature, potentially allowing future cost reductions.

Case 6, the oxy-fired supercritical gas turbine (or ‘Allam Cycle’) was evaluated to have an aggregate LCOE approximately 15% higher than for the post-combustion case. This case was assessed to have higher capital costs than Case 1 and similar operating costs, although the power output was lower. Whilst the philosophy adopted for this study was to develop costs on the basis of an Nth-of-a-kind development, it should be acknowledged that no demonstration plant has been operated yet, although a 50 MWth scale plant is under construction, therefore the capital and operating costs for this case are more uncertain than for the more conventional cases. It will be interesting to see how this technology advances to commercial scale over the next decade.

Molten Carbonate Fuel Cells (MCFC) are already operating as standalone power units, although not incorporating CO_2 capture or in combination with a CCGT and not at the scale presented in Case 7. Based on this analysis, this concept shows tremendous promise for commercial development in a political environment where the cost of emitting carbon dioxide is costed through a carbon price: the resultant LCOE for this concept was almost identical to that of the post-combustion benchmark.

One thing to note with the Case 7 results is that although the case features higher capital and operating costs for the CCGT with fuel cells, this concept also delivers significantly more power than Case 1 due to the additional power output from the MCFCs themselves. Thus, the costs of generating each unit of electricity are roughly equivalent.

Key uncertainties in the model development for the MCFC case are the cost of commercial-scale manufacture, the practicality of installing and maintaining large numbers of MCFC units and the operating lifetime of the fuel cells. We have made what we believe to be reasonably balanced assumptions in these areas, again assuming an Nth-of-a-kind philosophy. As with the Allam Cycle, we look forward to assessing this benchmark again in a few years' time to see how the technology has developed.

Both of these international novel technologies will need to demonstrate their potential improvements in cost and performance before they can be reasonably expected to compete with or outperform the proven state-of-the-art technology.

17.4 Sensitivity to Carbon Price and Storage / Transportation Costs

Two elements of the economic analysis that are potentially more controversial than others are the projected carbon price and the price that would be paid for transportation and storage of carbon dioxide: both of these price forecasts relate to markets that do not exist at the moment and which are therefore highly uncertain. The figures below show how each of the eight natural gas and coal benchmark case results would vary if different assumptions were made for these variables.

The central carbon price forecast used in this study was developed by the UK Department for Business, Energy and Industrial Strategy as shown in Table 5-7. This approach would not be effective if implemented unilaterally by the UK, without international co-operation. Figure 17-3 below shows how the results for LCOE would vary if the carbon price forecast were doubled (upper extent of the bars) or if no carbon price was implemented for UK projects (lower extent of the bars). In all cases, the central part of the bars, where the colour changes, represents the base assumption as reported elsewhere in this report.

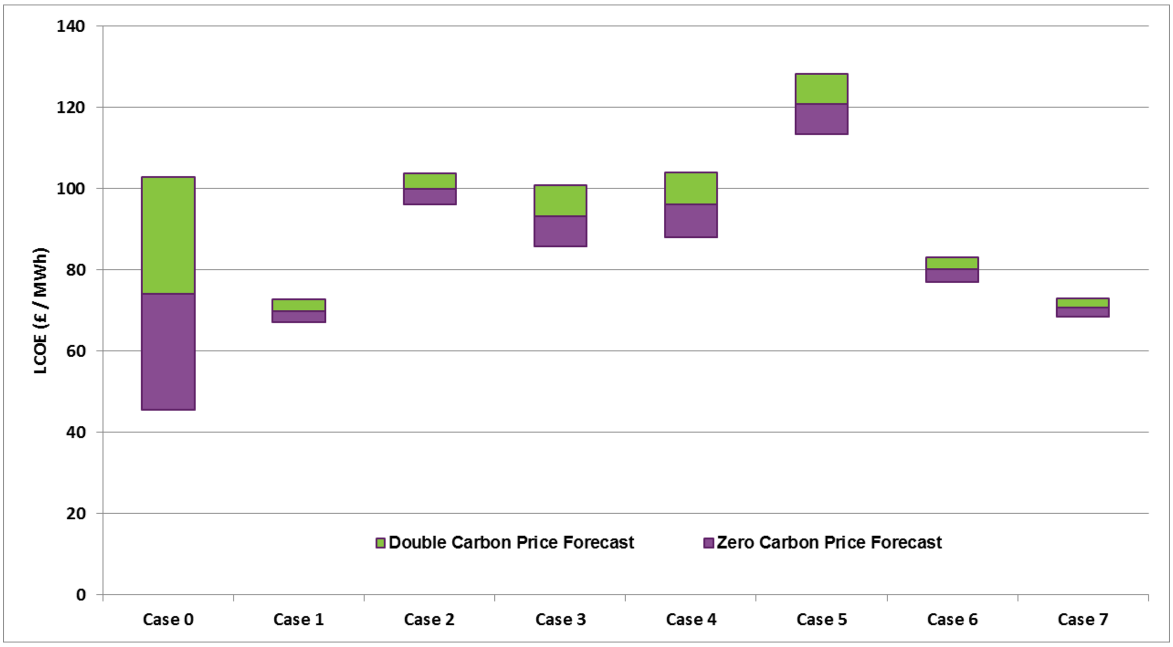

Figure 17-3: Sensitivity of LCOE to Carbon Price Forecast

Not surprisingly, the case that is affected the most by any change to carbon price policy is the Reference case (Case 0). This is because the unabated CCGT emits 100% of the CO_2 generated, whilst the other cases capture 90% or more of the carbon dioxide for geological sequestration. The lower extent of the bars indicates a situation where there is no cost associated with emitting carbon dioxide, and any power plant with carbon capture would be unable to compete economically against an unabated plant. The results for the other benchmark cases are not significantly affected by the carbon price. It is significant to note that all of the natural gas fired cases with carbon capture and the post-combustion coal plant would feature a lower LCOE than the unabated plant if the carbon price was doubled.

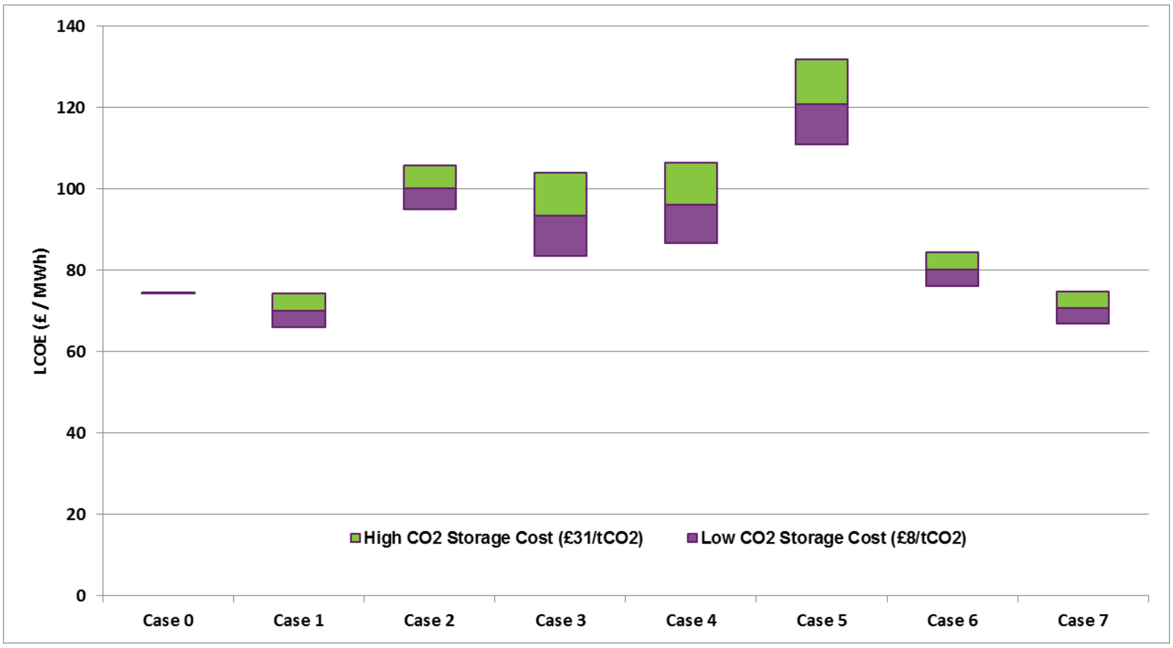

Figure 17-4: Sensitivity of LCOE to CO_2 Storage and Transportation Cost

Figure 17-4 assesses the impact of the cost of CO_2 transportation and storage on the overall economics for each case. Transportation and storage costs for this study have been taken from the BEIS-sponsored study into non-renewable technology energy hurdle rates performed in 2016 by Leigh-Fisher. The base assumption used for this study and the cost represented by the middle point of each bar, where the colour changes, is the central cost of £19 / tCO_2 reported by Leigh-Fisher. The upper extent of each bar shows the LCOE that would result from using the high storage cost of £31 / tCO_2 and the lower extent of each bar represents the equivalent cost using the low storage cost of £8 / tCO_2. Case 0 does not capture and store any carbon dioxide and so its LCOE does not vary in this sensitivity.

Whilst a realistic cost for storage and transportation is the subject of much debate and will have clear consequences for the economic viability of future decarbonised power plants, the figure above demonstrates that it is not a significant factor in selecting between competing technologies or fossil-fuel sources. The cost affects each of the gas cases and each of the coal cases to much the same extent because they export similar amounts of carbon dioxide.

For reference, it is worth noting that the above-mentioned Leigh-Fisher figures are somewhat higher than those used by the IEAGHG who used the figure of €10 / tCO_2 stored in the 2014 coal benchmarking report and €5 / tCO_2 stored in the 2012 natural gas benchmarking report.

17.5 Biomass Case Comparisons

An initial view of Figure 17-1 implies that the biomass cases 8-10 are unable to compete against either coal or natural gas fired power plant. However, there are several important elements to take into consideration when viewing these results. Firstly, the biomass cases do not benefit from the same economy of scale as the other cases. The study has considered plant with an exportable power output of around 400 MWe, which is half the size of the coal cases and one third of the unabated natural gas case. This scale was selected to ensure reliable results for existing commercial boilers. To illustrate the relative size of the biomass boilers, case 8 & 9 are a quarter the size of their respective coal boilers. Future technology development may allow larger scale units to operate on a comparable basis. For example, use of pulverised torrefied biomass may allow use of large-scale supercritical pulverised fuel boilers, as used in Cases 3 & 4. Secondly, the biomass cases suffer from high feedstock prices. The prices used for this study are based on delivery to small-scale users such as local CHP facilities. As a larger market develops, it would be fair to assume that real-terms prices would fall, but that assumption has not been used for this study. Finally, Figure 17-1 cannot represent the most important benefit of biomass fired power: the overall life-cycle analysis should result in a net reduction in atmospheric carbon dioxide. It is difficult to put a value on absorbing and capturing CO_2 from the air, but it is hoped that this study and future studies will support the UK Government in developing appropriate incentives to drive this development.

When comparing the three biomass cases against each other, the pre-combustion carbon capture, Bio-IGCC Case 10, does not compete well against the post-combustion and oxy-combustion cases (Cases 8 & 9). Case 10 uses energy-dense torrefied wood pellets instead of lower calorific biomass in the form of wood chips. This equates to 20% less fuel on a thermal basis compared to Case 8, which is equivalent to ~2.8 times lower fuel flow on a mass basis. Even though it uses much less fuel, the feedstock cost contribution to LCOE is significantly higher. Also, Case 10 has a higher capital cost, with 10% lower net power output. The oxy-combustion process (Case 9) performs better than Case 10, but is disadvantaged when compared to the simpler post-combustion process (Case 8), primarily due to the higher capital investment cost.

The biomass pre-combustion (bio-IGCC) case could be used as a source of renewable hydrogen that can be injected in the gas grid or can be used as a green feedstock for other chemicals. However, there is no way of accounting for those potential economic benefits which would make the bio-IGCC case attractive within the scope of analysis included in this study.

17.5.1 Sensitivity on Biomass Fuel Subsidies

As noted above, the three biomass fired cases suffer from reduced scale and reduced efficiency, combined with higher feedstock prices, whilst taking no credit for the key benefit – negative CO_2 emissions.

Figure 17-5 examines the impact of different biomass feed subsidy levels on the LCOE for the three biomass power cases, compared against the best performing state-of-the-art Benchmark Case, the natural gas fired CCGT with post-combustion carbon capture, Case 1.

The figure shows how subsidy levels of £4, £6 or £10 per GJ would reduce the LCOE for the biomass concepts. These are relatively arbitrary levels that were defined by Ricardo for assessment of competing non-bioenergy feedstock uses in the UK and Global Bioenergy Resource Model (ref. https://www.gov.uk/government/publications/uk-and-global-bioenergy-resource-model), and relate to existing low value feedstock prices. Note that the Central biomass feed price used in this study is £19.0 /MWh (£5.28 /GJ) and hence the higher two levels of subsidy would imply that the project developer would be paid to take the biomass, in a manner similar to the way existing waste incinerators are paid a gate-fee to receive municipal waste.

The chart indicates that a subsidy of between £ 8-9 /GJ would be needed to make Case 8 competitive with a large-scale natural gas fired power plant with post-combustion carbon capture.

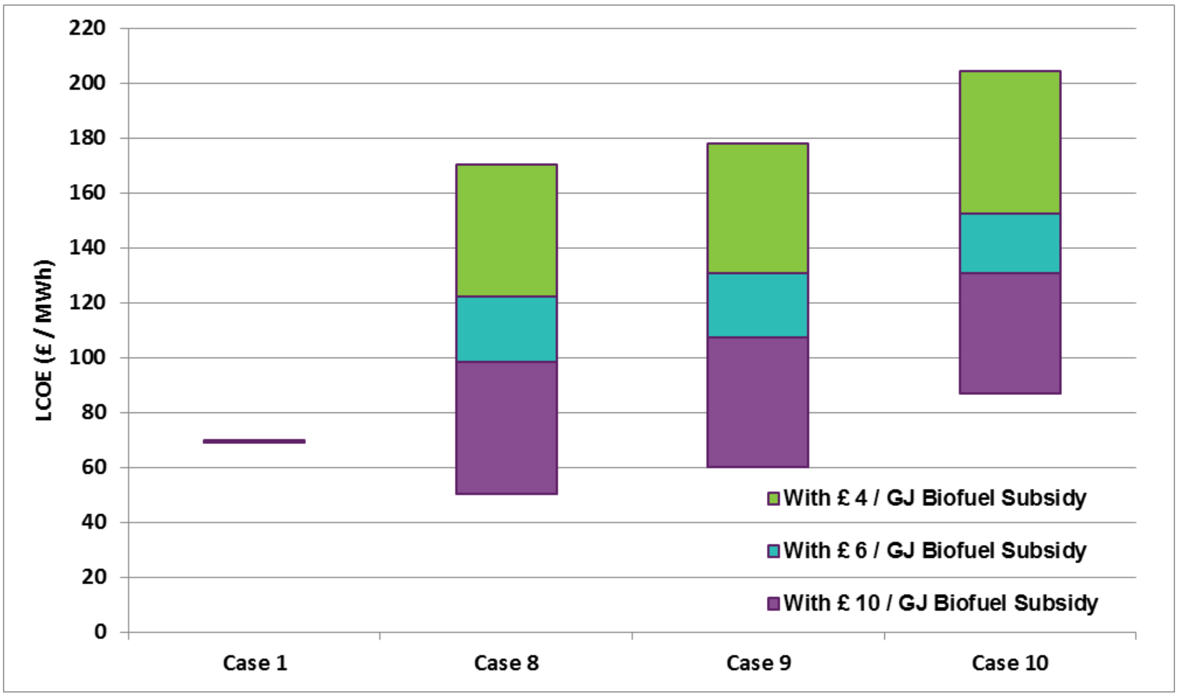

Figure 17-5: Sensitivity of LCOE to Biomass Feedstock Subsidies

It is difficult to objectively assess how the performance of the biomass cases may improve with improvements in scale and technology, but if it follows a similar path to development of other technologies, then one should expect the LCOE for these cases to trend significantly downwards as the technology matures. It is recommended that future studies review these potential performance improvements, especially in relation to how future innovation funding might encourage these improvements and the subsidy levels that would be needed to incentivise project developers.

ATTACHMENT 1: Study Basis of Design

Department for
Business, Energy
& Industrial Strategy

amec foster wheeler

Document Title:	Basis of Design
Document Number:	13333-8110-PD-001
Contract:	13333
Client's Name:	Department for Business, Energy & Industrial Strategy
Project Title:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology
Project Location:	Generic NE England

REVISION	**O1**	**Signature**	**1A**	**Signature**	**2A**	**Signature**
DATE	21 Feb 17		05 May 17		26 June 17	
ORIG. BY	S Ferguson		S Ferguson		R Ray	
CHKD BY	A Tarrant		A Tarrant		S Ferguson	
APP. BY	A Tarrant		A Tarrant		A Tarrant	

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Basis of Design

13333-8110-PD-001, Rev. 2A

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1 Introduction

The Department for Business, Energy & Industrial Strategy (BEIS) is evaluating the most promising CO_2 capture technologies in order to inform future innovation spending programmes and to shape future policy direction for carbon capture technologies in the Power and Energy Intensive Industries (EIIs) respectively.

The aim of the study is to investigate UK led technologies which will be the focus of the future innovation programme as providers of opportunities for UK economic growth. The study objective is to perform techno-economic assessment of the potential cost reduction and competitiveness of these technologies.

The study will involve:

- Technical assessment of performance, materials, processes, capital and operating costs
- Benchmarking novel CO_2 capture processes against state of the art technologies
- Undertaking detailed engineering studies of the two best performing novel technologies applied to four industry applications (which may include: post combustion carbon capture on gas fired power, carbon capture on steam methane reforming, carbon capture on gas processing and one of iron & steel, chemicals, refining or cement production.)

The primary outcomes from the study will be project planning deliverables, interim reports for work packages 2, 3 and 4, a matrix spreadsheet allowing comparison of technology and potential implementation cases across a number of key metrics and a final report summarising the technologies reviewed and the assessment results. Four Advisory Project Board meetings will also take place including the presentation of final results.

2 Purpose & Scope of Document

The purpose of this document is to record the key basis parameters which will define the design of the process schemes presented in this study.

The fossil fuel based state of the art technologies to be included will be:

- Post-combustion capture for power generation on gas
- Pre-combustion capture for power generation on gas
- Post-combustion capture for power generation on coal
- Oxy-combustion capture for power generation on coal
- Pre-combustion capture for power generation on coal

Two promising novel international processes that UK-led developers might need to compete against will be included, as follows:

- Oxy-fired supercritical power generation on gas
- Molten Carbonate Fuel Cell capture for power generation on gas

Three biomass cases will also be included:

- Post-combustion capture for power generation on biomass
- Oxy-combustion capture for power generation on biomass
- Pre-combustion capture for power generation on biomass

3 Benchmark Case Definition

3.1 Case 1: Post-combustion Capture for Power Generation on Gas

This case consists of a natural gas fired combined cycle power plant based upon 2 GE Frame 9HA.01 gas turbines each with a dedicated heat recovery steam generator (HRSG) and steam turbine in a 2x2 configuration. The flue gas from each HRSG is routed to a proprietary CO_2 capture unit where it is boosted in pressure using a flue gas fan, then cooled in a gas/gas heat exchanger before entering a direct contact cooler. CO_2 is captured from the cooled flue gas using an amine based solvent in an absorption column and is released from the solvent in the stripper. The captured CO_2 leaving the proprietary CO_2 capture unit is then compressed in 4 stages, dehydrated and then pumped to the required export pressure of 110 bar (abs).

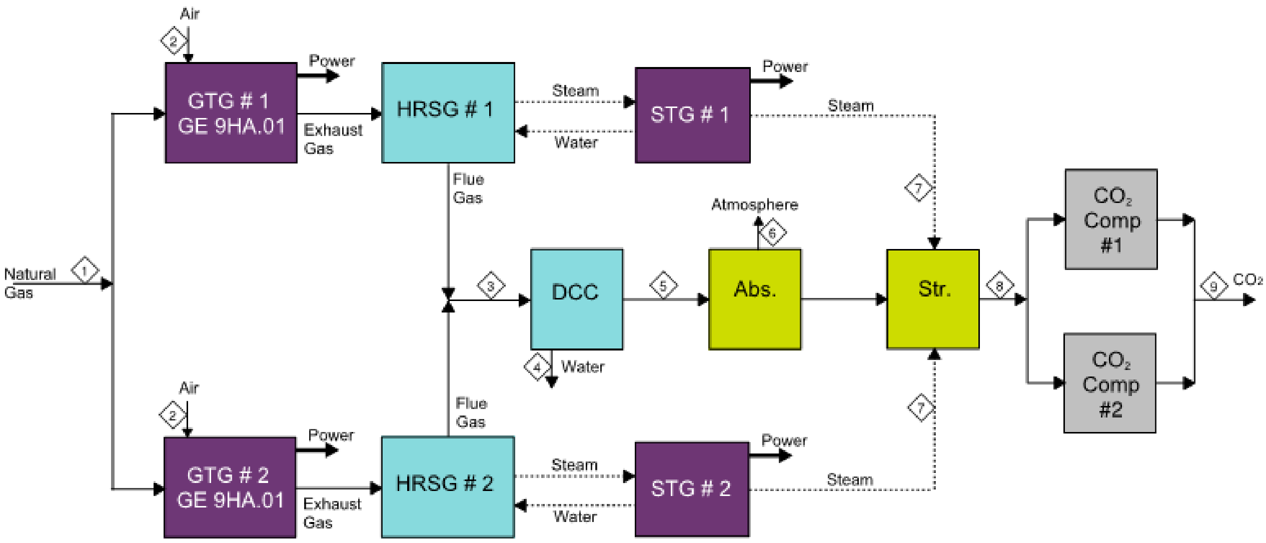


Figure 3-1: Case 1 Preliminary Block Flow Diagram

3.2 Case 2: Natural Gas IRCC with Pre Combustion Carbon Capture

This case consists of a natural gas fed integrated reforming combined cycle power plant based upon 2 gasification trains feeding 2 x GE Frame 9 syngas variant gas turbines each with a dedicated heat recovery steam generator (HRSG) and steam turbine in a 2x2 configuration. The natural gas is reformed in an autothermal reforming process, shifted to maximise pre-combustion CO_2 production with CO_2 subsequently captured in a Selexol physical absorption process. The captured CO_2 is then compressed in 4 stages, dehydrated and then compressed further to the required export pressure of 110 bar (abs).

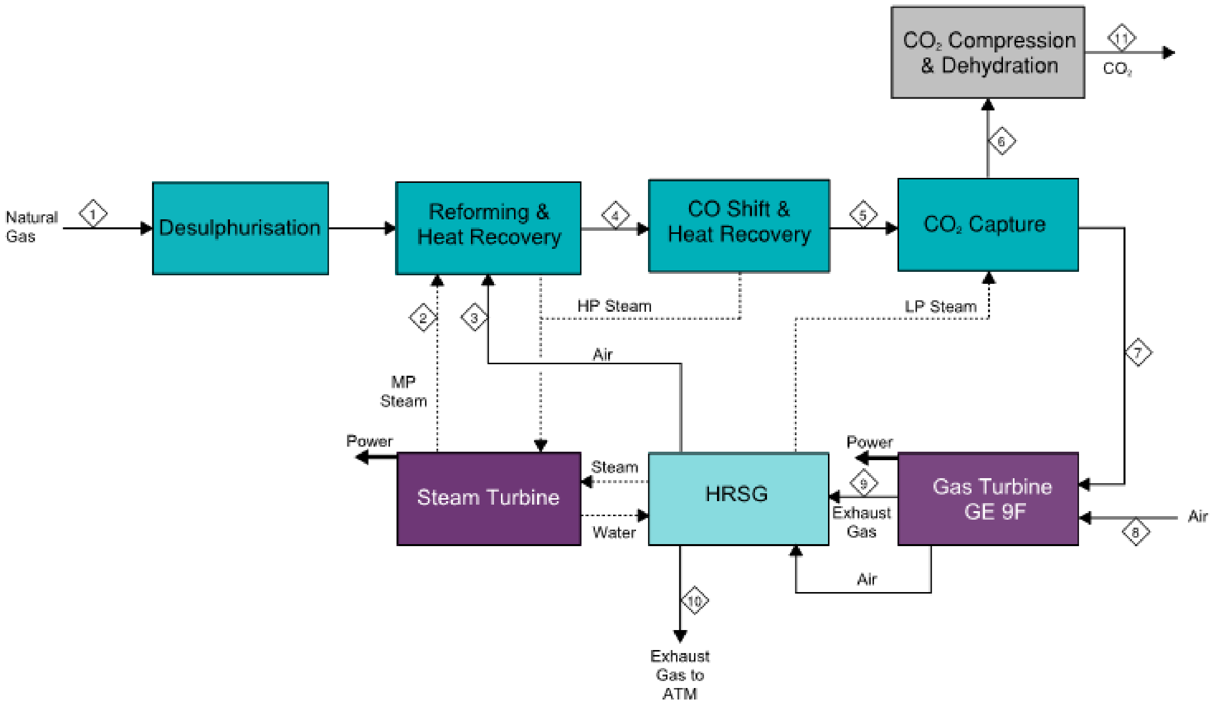


Figure 3-2: Case 2 Preliminary Block Flow Diagram

3.3 Case 3: Post-combustion Capture for Power Generation on Coal

This case consists of a pulverised coal fired supercritical power plant in a steam generator with superheating and single steam reheating, with a single steam turbine at a 1000 MWe net power production scale. The flue gas from the boiler is routed to a gas/gas heat exchanger then boosted in pressure using a flue gas fan, then is fed to the flue gas desulphurisation unit. The desulphurised flue gas is then fed to a proprietary CO_2 capture unit where it enters a direct contact cooler. CO_2 is captured from the cooled flue gas using an amine based solvent in an absorption column and is released from the solvent in the stripper. The captured CO_2 leaving the proprietary CO_2 capture unit is then compressed in 4 stages, dehydrated and then pumped to the required export pressure of 110 bar (abs).

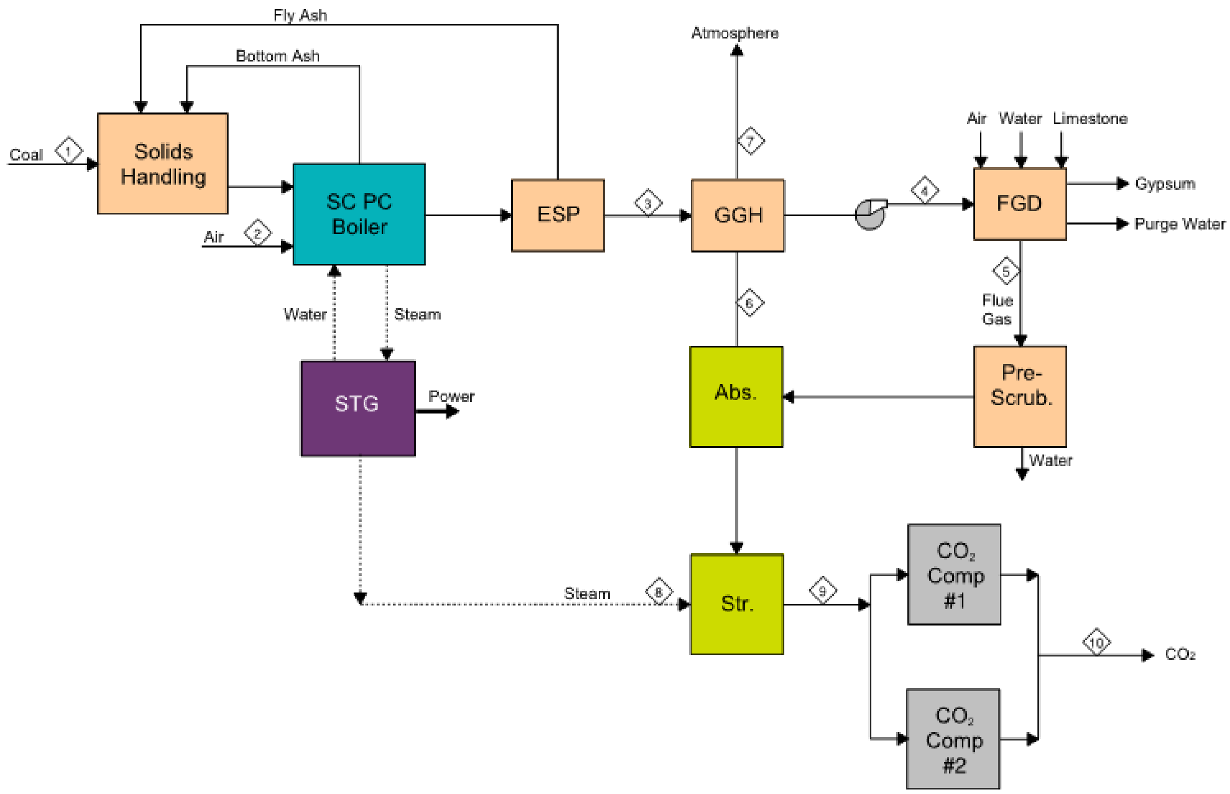


Figure 3-3: Case 3 Preliminary Block Flow Diagram

3.4 Case 4: Oxy-combustion Capture for Power Generation on Coal

This case consists of an oxy-fired pulverised coal fed supercritical power plant in a once through steam generator with superheating and single steam reheating, with a single steam turbine at the same thermal input capacity as the power island featured in Case 3. Oxygen for firing in the boiler is supplied by a cryogenic air separation unit (ASU). The flue gas from the boiler is routed to a multi-pass gas/gas heat exchanger followed by heat recovery and fly ash removal before a portion of the flow is routed back to the boilers as Secondary Recycle, via the gas/gas heat exchanger. The remaining flue gas passes through further heat recovery before the Primary Recycle is split off and recycled via flue gas desulphurisation and the gas/gas heat exchanger to the coal mills. The unrecycled flue gas stream is compressed and purified in a cryogenic purification system to the required export pressure of 110 bar (abs).

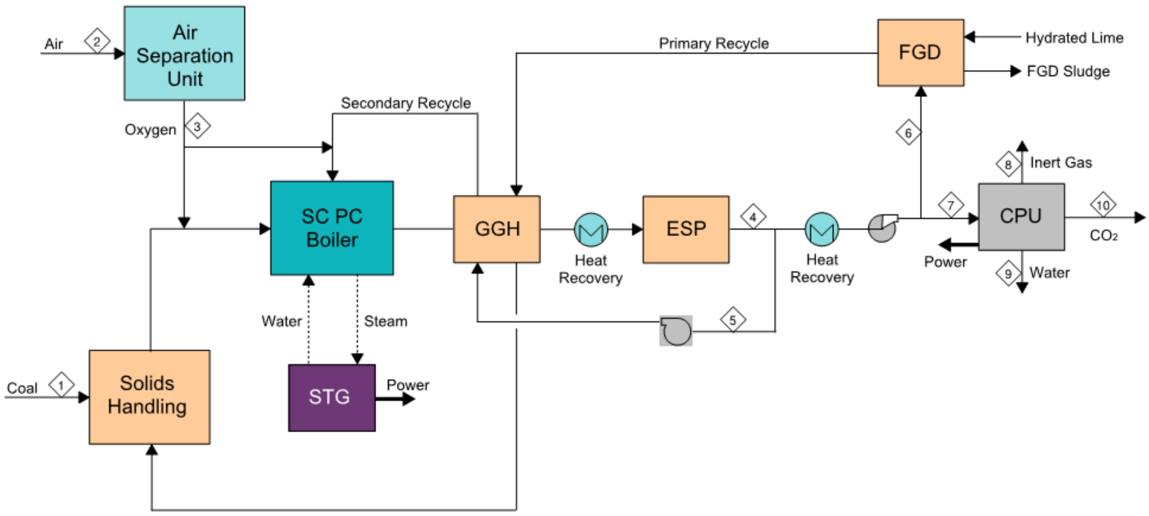


Figure 3-4: Case 4 Preliminary Block Flow Diagram

3.5 Case 5: Pre-combustion Capture for Power Generation on Coal

This case consists of an integrated gasification combined cycle power plant based upon the medium pressure Shell gasification process with a dry feed system and syngas cooler. A hybrid CO shift stage is followed by two stages of sour shift and Selexol physical solvent system for removal of acid gases. Sweetened, decarbonised syngas is then fed to two F-class gas turbines operating in a combined cycle power island. Sour gases are routed to an oxygen blown Claus unit with tail gas treatment and recycle to the Selexol unit. Captured CO_2 from the Selexol unit is compressed in 4 stages, dehydrated and then pumped to the required export pressure of 110 bar (abs). Oxygen for the gasification and Claus processes is supplied by a cryogenic air separation unit (ASU).

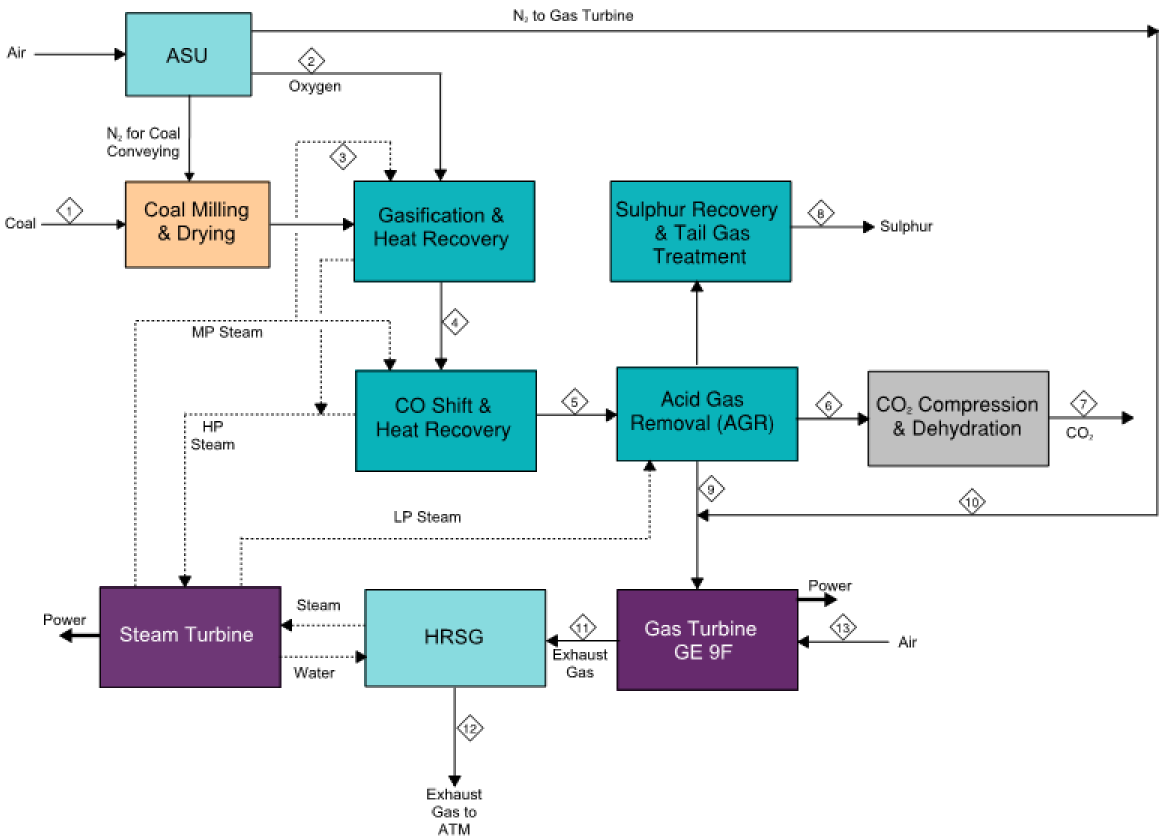


Figure 3-5: Case 5 Preliminary Block Flow Diagram

3.6 Case 6: Oxy-fired Supercritical Power Generation on Gas

The process used in this case combusts natural gas with oxygen at high pressure and temperature using the hot combustion products to drive a turbine in a novel thermodynamic cycle called the Allam Cycle. High temperature exhaust from the turbine enters an economiser heat exchanger which recovers heat to the turbine inlet streams before being cooled further to knock out most of the water of combustion. The CO_2 is then compressed and divided into the recycle stream and the product stream which is further purified using a cryogenic purification unit and pumped to the required export pressure of 110 bar (abs). Oxygen for the process is supplied by a cryogenic air separation unit (ASU) which is partially integrated with the main process.

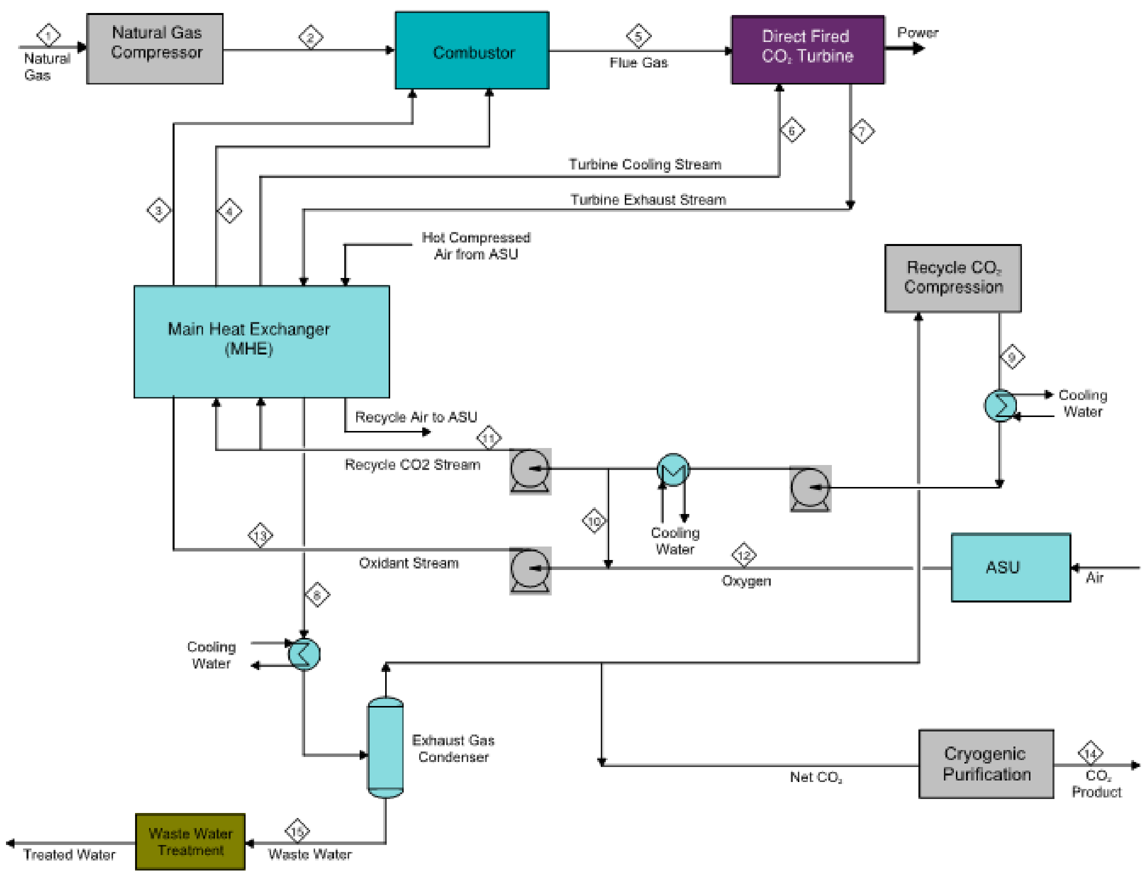


Figure 3-6: Case 6 Preliminary Block Flow Diagram

3.7 Case 7: Molten Carbonate Fuel Cell Capture for Power Generation on Gas

This case consists of a natural gas combined cycle power plant integrated with molten carbonate fuel cells (MCFCs) which capture CO_2 from the gas turbine exhaust while using an additional stream of natural gas and generating additional electrical power. The anode exhaust, containing mostly CO_2, water and some unconverted hydrogen and CO is compressed and purified using a cryogenic purification step before recycling the unconverted CO and hydrogen back to the inlet of the fuel cell anode. The CO_2 is then further compressed to the required specification of 110 bar (abs).

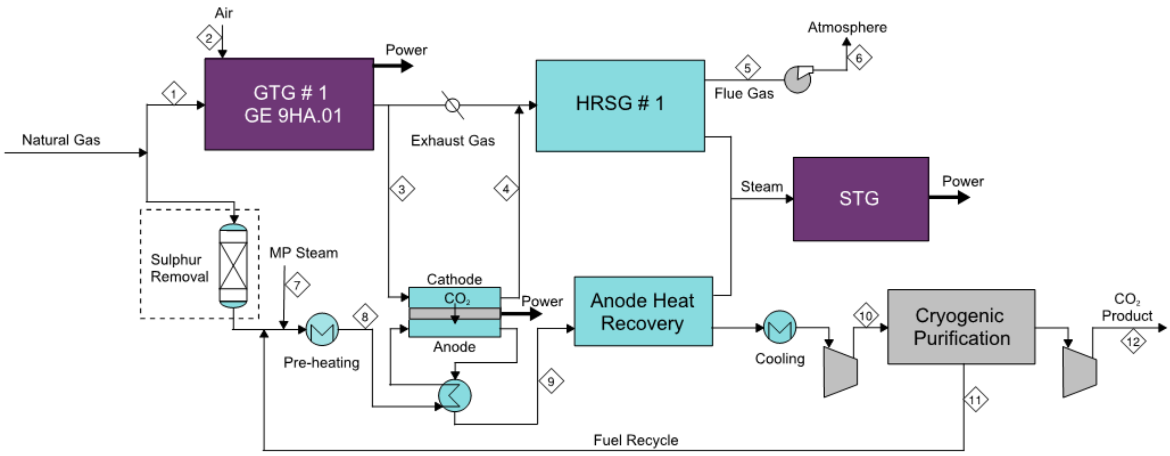


Figure 3-7: Case 7 Preliminary Block Flow Diagram

3.8 Case 8: Post Combustion Capture for Power Generation on Biomass

This case consists of two parallel trains of 300 MWe biomass fired sub-critical circulating fluidised bed (CFB) boiler power plants, each in a steam generator with superheating and single steam reheating. The flue gas from the biomass CFB boiler requires no SCR or FGD for denitrification and desulphurisation before entering the CO_2 capture process. CO_2 is captured from the cooled flue gas using an amine based solvent in an absorption column and is released from the solvent in the stripper. The captured CO_2 leaving the proprietary CO_2 capture unit is then compressed in 4 stages, dehydrated and then pumped to the required export pressure of 110 bar (abs). The treated flue gas exiting the CO_2 absorber is heated in a condensate heater (CH) before releasing to the atmosphere.

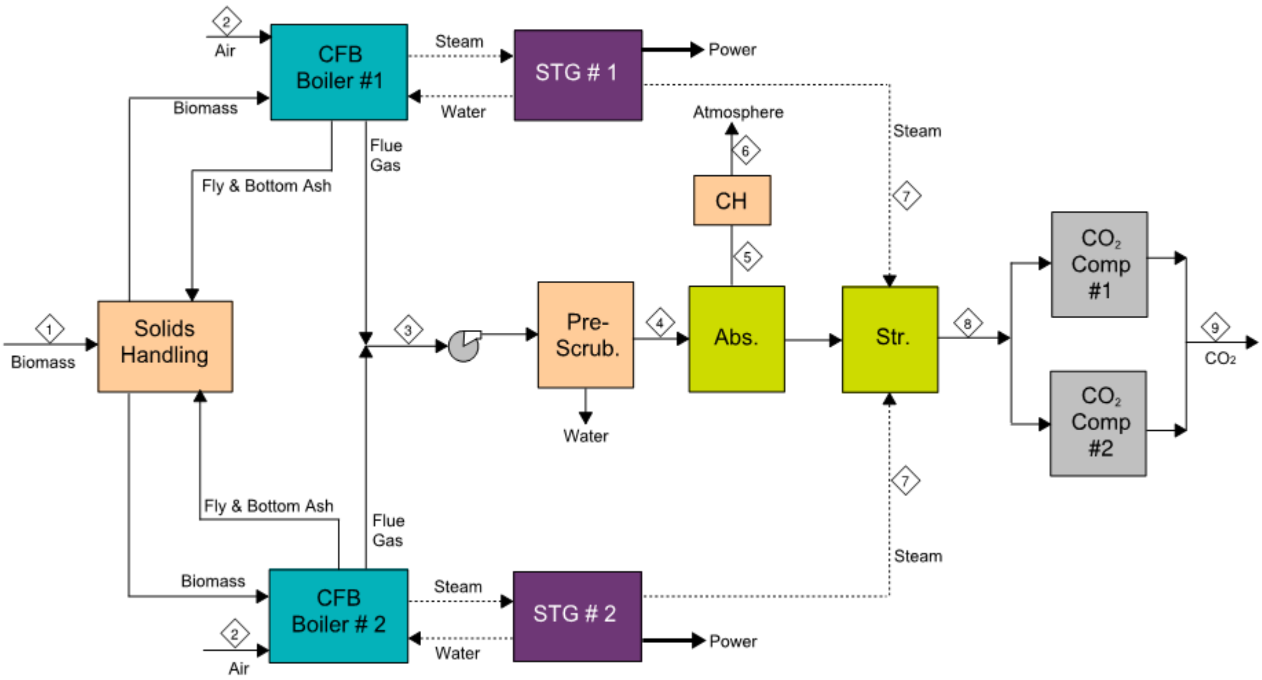


Figure 3-8: Case 8 Preliminary Block Flow Diagram

3.9 Case 9: Oxy Combustion Capture for Power Generation on Biomass

This case consists of two parallel trains of an oxy-fired sub-critical circulating fluidised bed (CFB) boiler power plants processing biomass feed, each in a steam generator with superheating and single steam reheating. Oxygen for firing in the boiler is supplied by a cryogenic air separation unit (ASU). The flue gas from the boiler is routed to a multi-pass gas/gas heat exchanger followed by heat recovery and fly ash removal before a portion of the flow is routed back to the boilers as Secondary Recycle, via the gas/gas heat exchanger. The remaining flue gas passes through further heat recovery before the Primary Recycle is split off and recycled via the gas/gas heat exchanger to the biomass mills. The unrecycled flue gas stream is compressed and purified in a cryogenic purification system to the required export pressure of 110 bar (abs).

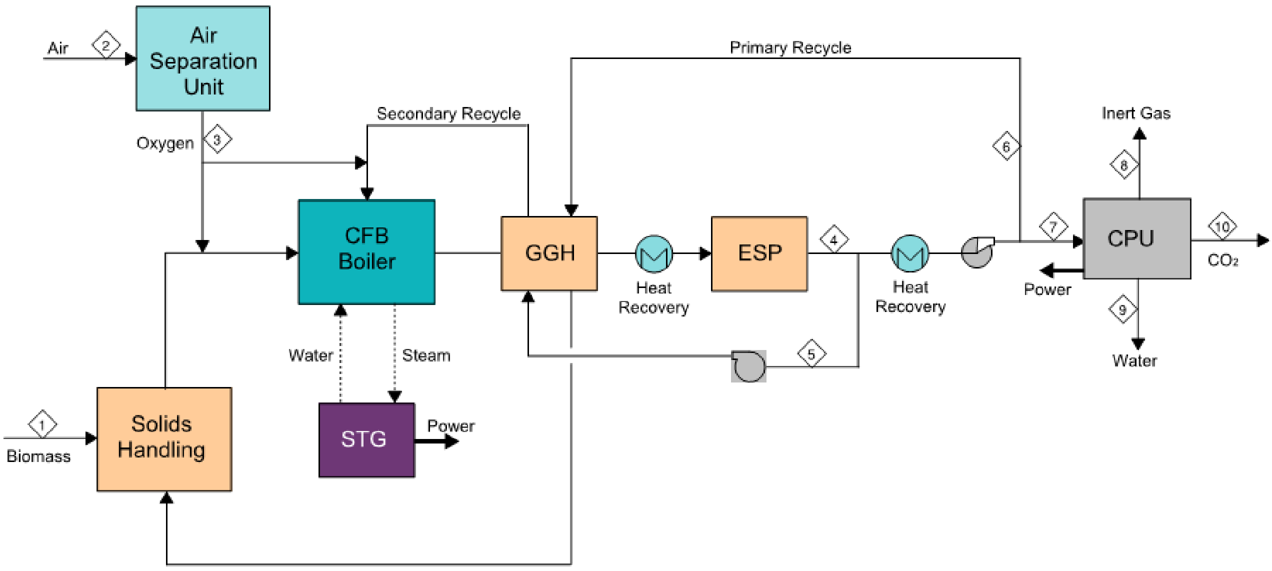


Figure 3-9: Case 9 Preliminary Block Flow Diagram

3.10 Case 10: Pre Combustion Capture for Power Generation on Biomass

This case consists of two parallel trains of torrefied wood pellet fired syngas production units; feeding syngas to a single train of combined cycle power island consisting of one GE 9F syngas variant gas turbine followed by HRSG and single steam turbine. Syngas production units are based upon the medium pressure Shell gasification process with a dry feed system and syngas cooler. A hybrid CO shift stage is followed by two stages of sour shift and Selexol physical solvent system for removal of acid gases. Captured CO_2 from the Selexol unit is compressed in 4 stages, dehydrated and then pumped to the required export pressure of 110 bar (abs). Oxygen for the gasification process is supplied by a cryogenic air separation unit (ASU). Syngas production unit produces sweetened and decarbonised syngas which is then fed to the single train combined cycle power island.

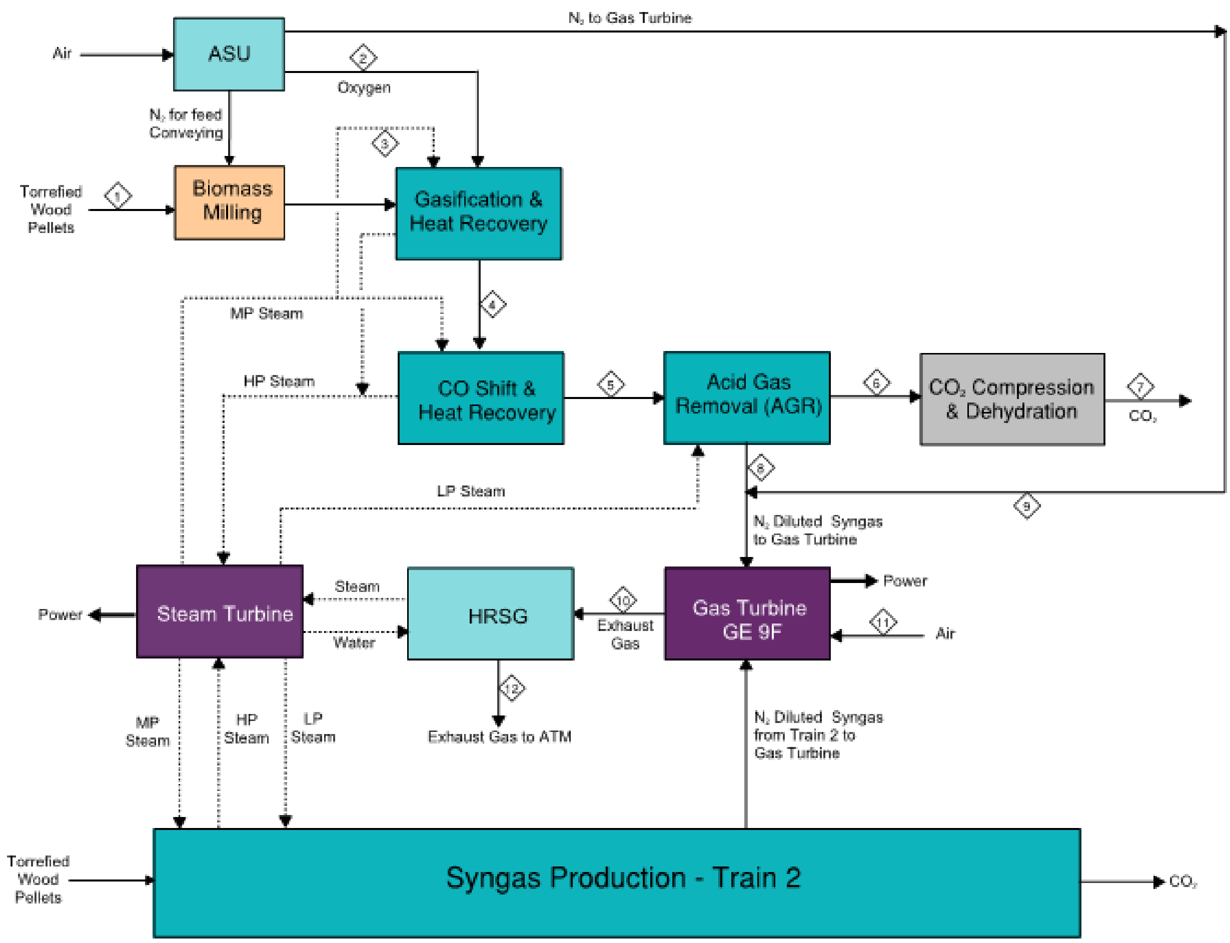


Figure 3-10: Case 10 Preliminary Block Flow Diagram

4 Design Conditions

4.1 Plant Location and Site Condition

The site is assumed to be a green-field, coastal location in the North East of England. A clear, level, obstruction (both under and above ground) free site, without the need for any required special civil works is assumed.

4.2 Plant Operating Conditions

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

Atmospheric pressure:	1013 mbar (*)
Relative humidity:	average 80% (*) maximum 100% minimum 10%
Ambient temperatures:	average 9°C (*) maximum 30°C minimum -10°C

The following air composition will be used:

Table 4-1: Ambient Air Composition

Species	Composition (mole fraction)
CO_2	0.0004
N_2	0.7721
O_2	0.2071
Ar	0.0092
H_2O	0.0112
Total	1.00

4.3 Battery Limits

Streams assumed to cross the plant battery limits are the following:

- Natural Gas
- Coal
- Biomass
- Torrefied wood pellet
- CO_2 product
- Closed loop cooling water supply and return
- Town's water
- Chemicals (including amine)
- Instrument/Plant air
- Nitrogen

4.4 Feedstock Specifications

4.4.1 Natural Gas Specifications

The following fuel gas specification has been used for natural gas, which meets the UK National Grid specification:

Table 4-2: Natural Gas Specification

Characteristic	UK National Grid Specification	Value Used
H_2S Content	Not more than 5 mg/m^3	3 ppm (molar)
Total Sulphur Content	Not more than 50 mg/m^3	40 mg/m^3
Hydrogen Content	Not more than 0.1% (molar)	0.1% (molar)
Oxygen Content	Not more than 0.001% (molar)	0.001% (molar)
Hydrocarbon Dewpoint	Not more than -2°C, at any pressure up to 85 bar(g)	<-2°C
Water Dewpoint	Not more than -10^{o}C, at 85 bar(g) (or the actual delivery pressure)	<-10°C
Temperature	Between 1^{o}C and 38^{o}C	9.0°C
Pressure	Not specified	70.0 Bar (abs)
Lower Heating Value (LHV), @9 °C	Not specified	46474 kJ/kg
Lower Heating Value (LHV), @25 °C	Not specified	46506 kJ/kg
Higher Heating Value (HHV), @25 °C	Between 36.9 MJ/m^3 and 42.3 MJ/m^3 (at standard temperature and pressure)	51477 kJ/kg
Wobbe Index	Between 48.14 MJ/m^3 and 51.41 MJ/m^3 (at standard temperature and pressure)	49.68 MJ/Sm^3
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.	
Composition		
Nitrogen, N_2	0.89	Vol %
Carbon Dioxide, CO_2	2.00	Vol %
Methane, CH_4	89.00	Vol %
Ethane, C_2H_6	7.00	Vol %
Propane, C_3H_8	1.00	Vol %
n-Butane, C_4H_{10}	0.10	Vol %
n-Pentane, C_5H_{12}	0.01	Vol %
Total	100.00	Vol %

4.5 Coal Specification

The coal specification considered for the benchmark cases in this study is Australian bituminous coal (shown in Table 4-3 below) in order to maintain consistency with IEAGHG 2014/3 report.

Other coal specifications are also acceptable if used by technology provider. However, the detailed composition should be specified in their documents.

Table 4-3: Coal Specification

Proximate Analysis	wt% - As Received
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
Oxygen	8.97
Nitrogen	1.80
Sulphur	1.10
Chlorine	0.03
Heating Value	MJ/kg – As Received
HHV	27.06
LHV	25.87
Fusion Temperature	°C
Ash fusion temperature at reduced atmosphere	1350

The following ash analysis should be used for the pulverised coal case.

Table 4-4: Pulverised Coal Ash Analysis

Coal Ash Analysis	Value	Units
SiO_2	50.0	Wt %
Al_2O_3	30.0	Wt %
TiO_2	2.0	Wt %
Fe_2O_3	9.7	Wt %
CaO	3.9	Wt %
MgO	0.4	Wt %
Na_2O	0.1	Wt %
K_2O	0.1	Wt %

P_2O_5	1.7	Wt %
SO_3	1.7	Wt %

4.6 Biomass Specifications

The biomass specification considered for benchmark cases 8 & 9 in this study is wood chips of clean virgin biomass (shown in Table 4-5 below) in order to maintain consistency with IEAGHG 2009-9 report.

Other biomass specifications are also acceptable if used by technology provider. However, the detail composition should be specified in their documents.

Table 4-5: Virgin Wood Chips Specification

Proximate Analysis	wt% - As Received
Moisture content	50
Volatile matter (dry ash free basis)	80
Ultimate Analysis	wt% - Dry basis
Carbon	50
Hydrogen	5.4
Oxygen	42.2
Nitrogen	0.3
Sulphur	0.05
Chlorine	0.02
Ash	2.0
Alkaline in Ash (Na+K)	≤4.5
Heating Value	MJ/kg – As Received
LHV	7.3
Fusion Temperature	°C
Ash fusion temperature at reduced atmosphere	>1100
Bulk Density	Kg/m3
Bulk Density	300

The following ash analysis should be used for the biomass case:

Table 4-6: Biomass Ash Analysis

Biomass Ash Analysis	Value	Units
SiO_2	15 - 50.0	Wt %
TiO_2	0.1 – 0.4	Wt %
Al_2O_3	4.0 – 10	Wt %
Fe_2O_3	1.0 – 4.0	Wt %
MgO	1.0 – 5.0	Wt %
CaO	20 – 30	Wt %
Na_2O	0.5 – 2.3	Wt %
K_2O	1.0 – 6.5	Wt %
P_2O_5	0.5 – 2.5	Wt %
MnO	1.0 – 3.0	Wt %
SO_3	0.5 – 2.0	Wt %

The torrefied wood pellet specification considered for the Bio-IGCC benchmark case 10 is shown in Table 4-7. This specification has been developed from reference paper (1) and discussion with Shell Global Solutions. The ash composition of the torrefied wood pellet is similar to biomass ash as listed in Table 4-6.

Table 4-7: Torrefied Wood Pellet Specification

Proximate Analysis	wt% - As Received
Moisture content	3.0
Volatile matter	65.5
Fixed Carbon	27.6
Ash	3.9
Ultimate Analysis	**wt% - As Received**
Carbon	47.5
Hydrogen	5.1
Oxygen	40.1
Nitrogen	0.3
Sulphur	0.05
Chlorine	0.02
Ash	3.9
Moisture	3.0
Heating Value	**MJ/kg – As Received**

LHV	16.8
Density	
Bulk Density, Kg/l	0.65-0.8
Energy Density, GJ/m3	12-19

4.7 CO_2 Product Specification

The following CO_2 product specifications has been used for design:

Table 4-8: CO_2 Product Specification

Characteristic	Value	Units
Temperature	30	°C
Pressure	110	Bar (abs)
Maximum Impurities		
Total Inerts (N_2 + Ar)	4.00	Vol %
Methane + other hydrocarbons	4.00	Vol %
Water, H_2O	500	ppmv
Hydrogen Sulphide, H_2S	200	ppmv
Oxygen, O_2	100	ppmv
Sulphur Dioxide, SO_2	100	ppmv
Nitrogen Dioxide, NO_2	100	ppmv

4.8 Flue Gas Characteristics

4.8.1 Natural Gas CCGT Flue Gas Characteristics

For those cases where the CO_2 capture process does not affect the flue gas characteristics exiting the heat recovery steam generator, the flue gas will have the following characteristics:

Table 4-9: Natural Gas CCGT Flue Gas Characteristics

Characteristic	Value	Units
Mass Flow Rate	5,975,000 *	kg/h
Pressure	0.00	barg
Temperature	97	°C
Composition		
Nitrogen, N_2	74.07	Vol %
Carbon Dioxide, CO_2	4.30	Vol %
Carbon Monoxide, CO	0.00	Vol %
Argon, Ar	0.88	Vol %

Oxygen, O_2	11.68	Vol %
Water, H_2O	9.07	Vol %
Sulphur Dioxide, SO_2	0.13**	ppmv
Nitrogen Oxide, NO	25**	ppmv

*Combined flow rate for both trains

**Dry basis at 15% O_2.

4.8.2 Pulverised Coal Flue Gas Characteristics

Flue gas conditions for the coal case are provisional and dependent on temperature of treated flue gas and pressure drop through the CO_2 capture unit, as we have proposed that the gas/gas heat exchanger and fan are outside of the CO_2 capture unit scope for this case due to the requirement for an FGD downstream of the fan.

Table 4-10: Pulverised Coal Flue Gas Characteristics

Characteristic	Value	Units
Mass Flow Rate	3,741,000	kg/h
Pressure	0.00	barg
Temperature	47.0	°C
Composition		
Nitrogen, N_2	71.40	Vol %
Carbon Dioxide, CO_2	13.68	Vol %
Carbon Monoxide, CO	0.00	Vol %
Argon, Ar	0.84	Vol %
Oxygen, O_2	3.2	Vol %
Water, H_2O	10.88	Vol %
Particulates (ash)	6*	ppmv
Sulphur Dioxide, SO_2	10*	ppmv
Sulphur Trioxide, SO_3	13*	ppmv
Nitrogen Oxide, NO	142.5*	ppmv
Nitrogen Dioxide, NO_2	7.5*	ppmv

* Dry basis.

4.8.3 Biomass Flue Gas Characteristics

The flue gas leaving the CFB boiler will have the following characteristics:

Table 4-11: Biomass CFB Flue Gas Characteristics

Characteristic	Value	Units
Mass Flow Rate	3,075,000	kg/h
Pressure	0.00	barg
Temperature	90.0	°C
Composition		
Nitrogen, N_2 + Argon, Ar	60.00	Vol %
Carbon Dioxide, CO_2	11.90	Vol %
Oxygen, O_2	3.9	Vol %
Water, H_2O	24.10	Vol %
Particulates (dry)	5*	mg/Nm^3
Carbon Monoxide, CO	50*	mg/Nm^3
Sulphur Dioxide, SO_2	35*	mg/Nm^3
Nitrogen Oxide, NO	112*	ppmv
Nitrogen Dioxide, NO_2	28*	ppmv

* Dry basis.

4.9 Utilities

4.9.1 Cooling Water

Although once through cooling water will usually result in the highest possible thermal efficiency of a power plant, the majority of CCGT power plants world-wide do not use once through cooling, even some that are coastally located. Cooling will be provided by mechanical draught wet cooling towers using an approach temperature of 7°C to the wet bulb temperature and a temperature rise of 11°C.

Cooling Water Supply = 14°C

Cooling Water Return = 25°C

4.9.2 Steam and Condensate

Where the process either requires steam for heating purposes, or generates steam in order to utilise excess heat, then these streams will be included on the heat & material balance. Steam pressure and temperature levels will vary according to the power plant feed and configuration. Appropriate levels will be selected based on the power demand and temperature limitations. The base power plant design will be modified to receive or deliver steam at suitable conditions.

4.9.3 Power

It is assumed that power is available for start-up and continuous operation. A typical grid connection at 275kV will be available and higher voltage levels will be specified, if required.

4.9.4 Other Utilities

Standard utilities will also be required by the plant, on the following bases:

- Purge nitrogen from cylinders,
- Instrument air, from two parallel package units (2 x 100% configuration),
- Raw water, demineralised water and potable water,
- Auxiliary boiler for start-up
- Diesel generator for emergency power supply and black-start if required.
- Water treatment plant
- Fire water

5 Reference

1. Carbo et al, ‘Biomass torrefaction achieves increased co-gasification shares in entrained flow gasifiers’, IChemE Gasification Conference, Rotterdam, 2014.

ATTACHMENT 2: Reference Case – Unabated CCGT

- Capital Cost Estimate Summary

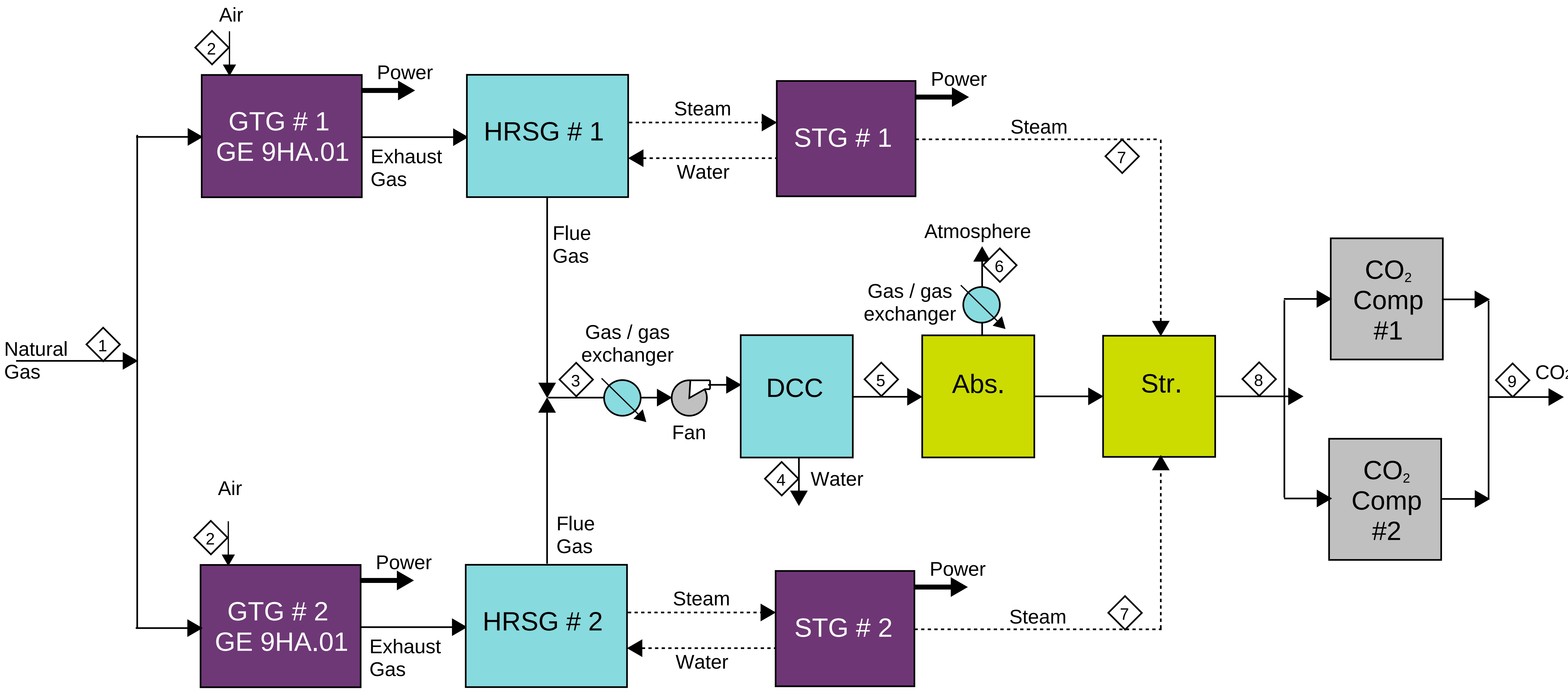
Amec Foster Wheeler		**Prepared By :**	**K.D. Nelson**
Client : BEIS		**Base Date :**	**1Q2017**
Project : Novel Carbon Capture Technology Study		**Rev. No. :**	**'6'**
Contract No.: 13333		**Print Date :**	**14-Dec-17**
Case 0 : Baseline CCPP - No Abatement			

Description	Case 0 : Baseline CCPP - No Abatement Total Cost GBP
Sub-Total Direct Materials	335,300,000
Other Material Costs	
Shipping / Freight	16,765,000
Third Party Inspection	3,353,000
Spare Parts (Comm/2yrs Op)	6,706,000
Sub-Total Materials	362,124,000
Material & Labour Contracts	
Civils/Steelwork & Buildings	50,295,000
Sub-Total Material & Labour Contracts	50,295,000
Labour Only Contracts	
Mechanical	60,354,000
Electrical/Instrumentation	16,765,000
Scaffolding/Lagging/Rigging	9,254,000
Sub-Total Material & Labour Contracts	86,373,000
Sub-Total Materials & Labour	498,792,000
EPCm Cost	
Engineering Services/Construction Management	74,819,000
Commissioning	9,976,000
Sub-Total EPCm Cost	84,795,000
Total EPC Cost	583,587,000
Other Costs	
Pre-Licensing, Technical and Design etc	5,836,000
Regulatory, Licensing and Public Enquiry etc	12,900,000
Infrastructure Connection Costs	29,000,000
Owners Costs	40,851,000
Sub-Total Other Costs	88,587,000
Total Project Cost	**672,174,000**

ATTACHMENT 3: Case 1 – CCGT with Post-Combustion Capture[8]

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List

[8] Please note, as the basis of design for Benchmark 1 is based on Shell Cansolv's proprietary design, a capital cost estimate is not included to maintain confidentiality.



NOTES

Department for Business, Energy & Industrial Strategy

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REV.	DATE	ORIG.	CHK'D	APPR.
01	24/07/17	SF	RR	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram - Case 1 - CCGT with Cansolv Post Combustion CO_2 Capture

Drawing No

Rev

amec foster wheeler

HEAT & MATERIAL BALANCE - Case 1 - Post-Combustion Capture for Power Generation on Gas

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**REV - O2**	**SHEET**
CONTRACT:	13333	**DATE**	15/05/2017	11/09/2017	1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S. Ferguson	S. Ferguson	
		APP. BY	T. Tarrant	T. Tarrant	

Stream Number	1	2	3	4	5	6	7	8	9
Stream Name	B/L Natural Gas	GT Air Intake	Flue Gas to CO2 Capture	DCC Water Purge	Flue Gas To Absorber	Treated Gas to Atm.	LP Steam from ST	CO2 to Compression	Product CO2
Pressure (bar abs)	70.00	1.01	1.01	6.00	1.05	1.01	4.41	1.99	110.00
Temperature (°C)	9.0	9.0	97.1	51.2	35.0	80.0	148.0	49.0	30.0
Mass rate (kg/h)	150296	5824704	5975000	328365	5716403	5355289	490000	367730	352274
Molar rate (kmol/h)	8341	197751	210647	18232	198470	189977	27192	8649	8151
Volume rate (m3/h)	2239	4575259	5591512	328	4728952	5507591	211424	153361	472
Component									
Nitrogen (mol%)	0.89	77.21	74.07	0.07	77.14	80.59	0.00	0.00	0.00
CO2 (mol%)	2.00	0.04	4.30	0.00	4.48	0.47	0.00	94.25	100.00
Methane (mol%)	89.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ethane (mol%)	7.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Propane (mol%)	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-Butane (mol%)	0.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-Pentane (mol%)	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen (mol%)	0.00	20.71	11.68	0.03	12.17	12.71	0.00	0.00	0.00
Argon (mol%)	0.00	0.92	0.88	0.00	0.92	0.96	0.00	0.00	0.00
CO (mol%)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Water (mol%)	0.00	1.12	9.07	99.90	5.29	5.27	100.00	5.75	0.00
Proprietary Solvent (mol%)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Other (mol%)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

UTILITIES BALANCE SUMMARY - Case 1 - Post-Combustion Capture for Power Generation on Gas

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**REV - O2**			**SHEET**
CONTRACT:	13333	**DATE**	17/05/2017	11/09/2017			1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S Ferguson	S Ferguson			
		APP. BY	T Tarrant	T Tarrant			

UNIT No.	**DESCRIPTION**	**Electric Oper. Load**	**Fuel Gas LHV**	**Steam (t/h)**			**Condensate (t/h)**			**Cooling water**	**Raw Water**	**REV**
		MW	**MW**	**HPS 171 barg**	**MPS 40 barg**	**LPS 4.8 barg**	**HP T/h**	**MP T/h**	**LP T/h**	**T/h**	**T/h**	
Power Island												
	GT Gross Electrical Power	823.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		
	GT Auxiliaries & BOP	-10.3	-1940.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0		02
	ST Gross Electrical Power	320.8	0.0	745.6	853.6	952.1	0.0	0.0	-952.1	0.0		
	ST Auxiliaries (inc BFW pumps, condenser)	-4.4	0.0	-745.6	-853.6	-460.1	0.0	0.0	460.1	-34486.0		02
	Power Island Net	1129.6	-1940.2						460.1	-34486.0		
CO2 Capture												
	Flue Gas Blower	-18.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		
	Pumps	-5.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		
	Thermal Reclaimer Reboiler	-1.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0			
	Solvent Reboiler	0.0	0.0	0.0	0.0	-492.0	0.0	0.0	492.0	0.0		
	CO2 Capture Net	-24.8				-492.0			492.0	-34920.0		
CO2 Compression												
	CO2 Compressor	-26.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		
	CO2 Compressor Inter/After Coolers	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-4430.0		
	CO2 Capture Net	-26.8								-4430.0		
Utilities & Offistes Systems												
	Cooling Water System	-13.0								73836.0	-1073.2	02
	Demin Plant	0									1073.2	02
	Buildings and Misc extra loads	0.0										
	Effluent Treatment	0.0										
	U&O Net	0.0								73836.0		
Overall Performance												
	Gross Production	1144.3	0	0.0	0.0	0.0	0.0	0.0	0.0	73836.0	1073.2	
	Total Parasitic Loads	-66.2	0	0.0	0.0	0.0	0.0	0.0	0.0	-73836.0	-1073.2	
	Net Produced	1078.0	-1940	0.0	0.0	0.0	0.0	0.0	460.1	0.0	0.0	
	Net LHV Efficiency	55.56%	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	

	EQUIPMENT LIST FOR COMPRESSORS	Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	Client: Department for Business, Energy & Idustrial Strategy Contract No: 13333	Originated	SF	SF		SHEET 1 of 7
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR	RR		
		Approved	TT	TT		
	Case: Natural Gas Combined Cycle with Post Combustion CO2 Capture	Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
FA-001	Flue Gas Fan	Axial	1			1.385 / 1.384	0.1	1.0 / 1.1		0.999 / 0.999			28.36	CO2 Capture Unit Scope	
K-001	CO2 Compressor Package (5 stages)	Multi-Stage Integrally Geared	2 x 50%		57,591	1.284 / 4.925	108.0	2.0 / 110.0		0.991 / 0.297	13382	CrNi alloy	42.52	180 t/h CO2	
						/		/		/					
						/		/		/					
						/		/		/					
						/		/		/					
						/		/		/					

Notes:

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EQUIPMENT LIST FOR HEAT EXCHANGERS

Client: Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333			
Rev.	**REV 01**	**REV 02**	**REV 03**	
Originated	SF	SF		**SHEET** 2 **of** 7
Checked	RR	RR		
Approved	TT	TT		
Date	08/05/2017	15/05/2017		

Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Case: Natural Gas Combined Cycle with Post Combustion CO2 Capture

EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	MATERIAL TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
E-001	DCC Cooler	Gasketed Plate and frame	1						51.0 / 10.0	80.0 / 7						**CO2 Capture Unit Scope**	02
E-002	CO2 Wash Water cooler	Gasketed Plate and frame	2						80.0 / 3/FV	80.0 / 7.0						**CO2 Capture Unit Scope**	02
E-003	Lean amine cooler	Gasketed Plate and frame	1						51.0 / 10.0	140.0 / 9.5						**CO2 Capture Unit Scope**	02
E-004	Lean/Rich exchangers	Gasketed Plate and frame	7						51.0 / 10.3	140.0 / 6						**CO2 Capture Unit Scope**	02
E-005	Stripper condenser	Welded Plate and frame	2						51.0 / 10.0	150.0 2/FV						**CO2 Capture Unit Scope**	02
E-006	Reboilers	Welded Plate and frame	8						150.0 / 2/FV	295.0 / 5/FV (tubeside)						**CO2 Capture Unit Scope**	02
E-007	Gas / Gas Exchanger		1						105.0 / 2.000	125 / 2.0	CS	CS					
E-101	CO2 Compressor Cooler - Stage 1	Shell & Tube	2				6.29	1246	tubeside 51.0 / 10.0	160.0 / 10	SS304	SS304					
E-102	CO2 Compressor Cooler - Stage 2	Shell & Tube	2				3.89	662	tubeside 51.0 / 10.0	160.0 / 12	SS304	SS304					
E-103	CO2 Compressor Cooler - Stage 3	Shell & Tube	2				4.21	607	tubeside 51.0 / 10.0	160.0 / 29	SS304	SS304					
E-104	CO2 Compressor Cooler - Stage 4	Shell & Tube	2				6.02	771	tubeside 51.0 / 10.0	160.0 / 70	SS304	SS304					
E-105	CO2 Compressor Cooler - Stage 5	Shell & Tube	2				8.83	1166	tubeside 51.0 / 10.0	100.0 / 115	SS304	SS304					

Notes:
1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler	EQUIPMENT LIST FOR PUMPS		Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy — Contract No: 13333	Originated	SF	SF		SHEET 3 of 7
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR	RR		
			Approved	TT	TT		
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture	Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP / SG / VISC'Y °C cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
P-001	DCC Pump	Centrifugal	3 x 50%					6.0		30.0 1.004 0.797	Design Temp: 80/-10 Design Press.: 7.5			**CO2 Capture Unit Scope**	02
P-002	Absorber WW pump	Centrifugal	3 x 50%					3.6		30.0 1.004 0.797	Design Temp: 80/-10 Design Press.: 7			**CO2 Capture Unit Scope**	02
P-003	Rich amine pumps	Centrifugal	3 x 50%					12.9		36.0 0.983	Design Temp: 80/-10 Design Press.: 11			**CO2 Capture Unit Scope**	02
P-008	Stripper reflux pump	Centrifugal	3 x 50%					6.0		137.0 0.919 0.205	Design Temp: 295/-10 Design Press.: 12.4/FV			**CO2 Capture Unit Scope**	02
P-006	Lean amine pumps	Centrifugal	3 x 50%					3.9		119.0 0.95	Design Temp: 140/-10 Design Press.: 8/FV			**CO2 Capture Unit Scope**	02
P-007	Lean amine feed pumps	Centrifugal	3 x 50%					6.9		50.0 0.95	Design Temp: 140/-10 Design Press.: 8/FV			**CO2 Capture Unit Scope**	02
P-004	Amine Drain Pump	Centrifugal	1 x 100%					4		60.0 0.983	Design Temp: 100/-10 Design Press.:2.40			**CO2 Capture Unit Scope**	
	Demin Water Pump	Pump	3					1		20.0	Design Temp- 80 / -10 Design Pressure: 5.20		Stainless Steel casing and impeller	Sizing: 1.7 x 0.7 x 1.4	
	Towns Water Transfer Pump	Centrifugal	3					1		20.0	Design Temp- 80 / -10 Design Pressure: 6.0		Cast iron casing with Stainless steel impeller	Size: 1.8 X 1.8 X 1.8	
	Firewater Pump Package	Pump	3							20.0	Design Temp- 80 / -10 Design Pressure: 19		CS	Sizing: 8 x 4 x 3.2 1 x diesel, 1 x electric and 1 x jockey	
	CO2 Compressor Condensate Return Pump	Pump	2 x 100%		4.5	4.4		3		25.0 0.996	Design Temp- 80 / -10 Design Pressure: 5.0	1	CS		

Notes:

amec foster wheeler	EQUIPMENT LIST FOR VESSELS		Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy — Contract No: 13333	Originated	SF	SF		SHEET 4 of 7
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR	RR		
			Approved	TT	TT		
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture	Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	DIMENSIONS HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	DESIGN CONDITIONS PRESS barg	DESIGN CONDITIONS VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	MATERIALS OF CONST'N INTERNALS MAT./LINING/ CA	REMARKS	REV
C-001	DCC (Direct Contact Cooler)	Rectangular column	1				V	150	0.30	1.013				**CO2 Capture Unit Scope**	02
C-002	Absorber (absorber section)	Rectangular column	1				V	80/-10	0.10	1.013				**CO2 Capture Unit Scope**	02
C-003	Absorber (water wash section)	Rectangular column	1				V	80/-10	0.10	1.013				**CO2 Capture Unit Scope**	02
C-004	Stripper	Vertical cylinder	1				V	250/-10	3.50	FV				**CO2 Capture Unit Scope**	
V-002	CO2 Reflux Accumulator	Horizontal	1				H	295/-10	5.00	FV				**CO2 Capture Unit Scope**	02
V-101	CO2 Compressor Stage 1 KO drum	Vertical drum	2	3.60	7.20	73	V	80	3.50	1.013	demister	SS304	SS304	180 t/h CO2	
V-102	CO2 Compressor Stage 2 KO drum	Vertical drum	2	3.00	6.00	42	V	80	6.00	1.013	demister	SS304	SS304	180 t/h CO2	
V-103	CO2 Compressor Stage 3 KO drum	Vertical drum	2	2.40	4.80	22	V	80	10.00	1.013	demister	SS304	SS304	180 t/h CO2	
V-104	CO2 Compressor Stage 4 KO drum	Vertical drum	2	1.80	3.60	9	V	80	25.00	1.013	demister	SS304	SS304	180 t/h CO2	
V-105	CO2 Compressor Stage 5 KO drum	Vertical drum	2	1.40	2.80	4	V	80	65.00	1.013	demister	SS304	SS304	180 t/h CO2	

Notes: **1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank**
2. V - Vertical H - Horizontal

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EQUIPMENT LIST FOR TANKS			**Rev.**	**REV 01**	**REV 02**	**REV 03**	
Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Originated**	SF	SF		**SHEET** 5 **of** 7
Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology		**Checked**	RR	RR		
			Approved	TT	TT		
Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture		**Date**	03/02/2017	15/05/2017		

EQUIPMENT NUMBER	**DESCRIPTION**	**NO REQ'D**	**DIMENSIONS**		**TOTAL VOLUME**	**ROOF TYPE**	**BOTTOM TYPE**	**STORAGE**		**SG**	**CORRO-SION**	**'MATERIALS**	**MANWAYS**		**TYPE**	**HEATING COIL**	**INSUL-ATION**	**REMARKS**	**REV**
			ID	**HEIGHT**				**PRESS**	**TEMP**				**SHELL**	**ROOF**					
			m	**m**	**m^3**														
T-001	Lean Amine Tank	1				Storage tank		0.02	30						Vertical, sized for full inventory			**CO2 Capure Unit Scope**	
T-002	Amine Drain Tank	1													Horizontal, underground.			**CO2 Capure Unit Scope**	02
T-003	Absorbent Make-up tank	1				Storage tank		0.02	30						Vertical			**CO2 Capure Unit Scope**	02
T-006	Towns Water Storage Tank	1	10.00	10.00	780.00	Vertical cylindrical		0.0075 / -0.0025	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. 0.0075 /	
T-007	Demin Water Tank	1	10.00	9.00	700.00	Tank		0.90	20			Lined CS						Design Temp: 80/-10 deg C	
T-008	Firewater Storage Tank	1	13.0	7.8		Tank			20			Lined CS						Design Pres.: 0.0075 / -0.0025 Design Temp: 80/-10	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	SF	SF		SHEET 6 of 7
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR	RR		
		UK Carbon Capture Technology			Approved	TT	TT		
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture			Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
CCGT Power Island	**2 x GE 9HA.01 + 2 x ST (50 Hz), total power output = 1,232,019 kW at site conditions**										**Total GT Power = 823452 kW Total ST Power = 320800 kW**			**Gas Turbine World 2015-2016**	
S-001	Ion Exchange Package		6											**CO2 Capture Unit Scope**	
S-002	Thermal Reclaimer Package	Vacuum Distillation Column	1											**CO2 Capture Unit Scope**	
S-003	CO2 Absorbent Filtration Unit	Cartridge Type Filter												**CO2 Capture Unit Scope**	
S-004	Filtration Unit	Fixed Bed Filter												**CO2 Capture Unit Scope**	
S-005	Nitrogen Package	Tank/pump	1						7	Design Temp: 80/-10 deg C Design Press.: 14 barg Oper. Temp.: 20 deg C		CS			
S-006	Compressed Air Package	Compressor	1							Design Temp- 80 / -10 deg C Design Pressure: 8.70 barg					
TEG-101	TEG Dehydration Package		2							Design Temp: 150/-10 deg C Design Press.: 76 barg				Dehydrates 180 t/h CO2, removes 142 kg/h water to get to spec of 50 ppm	
P-001	Flow metering and analyser package	Metering	1					300000		Design Temp: 150/-10 deg C Design Press.: 114 barg				Fiscal metering package	

Notes:

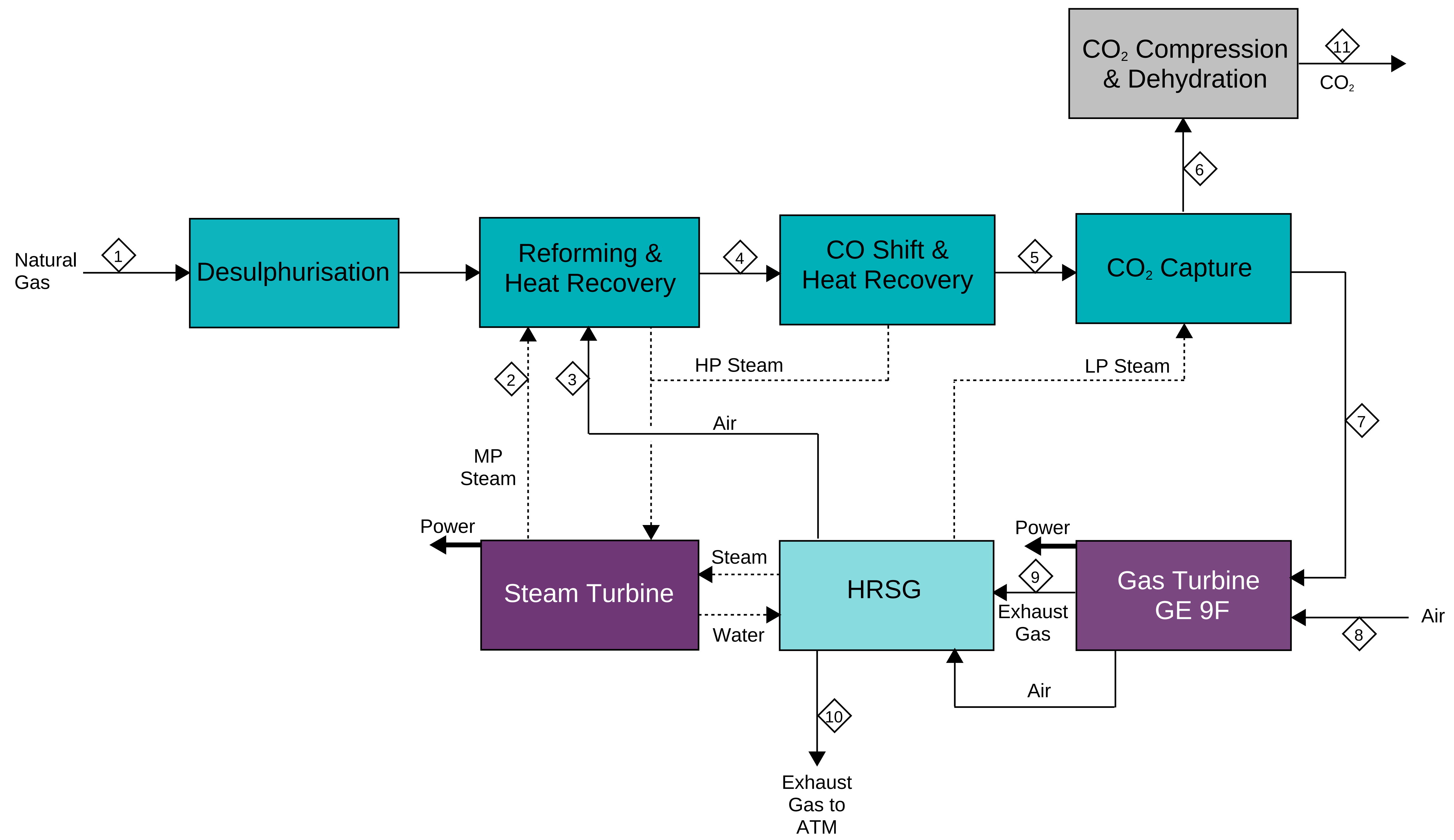
amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	SF	SF		SHEET 7 of 7
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR	RR		
	Unit No:	UK Carbon Capture Technology			Approved	TT	TT		
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture			Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Flue gas ducting from 2 HRSGs to single GGH		6 m X6 m Estimated Length 73 m	Square		CS	Flow Rate: 5968 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from single GGH to single fan		6 m X6 m Estimated Length 5 m	Square		CS	Flow Rate: 5968 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from single fan to single DCC		6 m X6 m Estimated Length 5 m	Square		CS	Flow Rate: 5968 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from single DCC to single absorber		5.5 m X 5.5 m Estimated Length 30 m	Square		CS	Flow Rate: 5814 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from single absorber to single GGH		5.5 m X 5.5 m Estimated Length 70 m	Square		CS	Flow Rate: 5447 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from single GGH to two HRSG stacks (stacks included in Power Island Package)		5.5 m X 5.5 m Estimated Length 10 m	Square		CS	Flow Rate: 5447 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02

Notes:

ATTACHMENT 4: Case 2 – Integrated Reforming & Combined Cycle

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List
- Capital Cost Estimate Summary



NOTES

1. Two Trains of all equipment

Department for Business, Energy & Industrial Strategy

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REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	SF	RR	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram
Case 2 - Natural Gas IRCC Case

Drawing No | Rev

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HEAT & MATERIAL BALANCE - Case 2 NG IRCC

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**SHEET**
CONTRACT:	13333	**DATE**	12/04/2017	1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	R. Ray	
		APP. BY	T. Tarrant	

Stream Number	1	2	3	4	5	6	7	8	9	10
Stream Name	B/L NG	MP Steam to ATR	GT Air to Process	Syngas to Shift Reactor	Shifted Syngas to AGR	CO2 to Compression	Syngas to Gas Turbine	Air to Gas Turbine	Flue Gas from Gas Turbine	Exhaust Gas to ATM
Temperature (°C)	9	315	400	350	60	40	190	9	565	80
Pressure (bar abs)	70.00	30.00	18.00	27.60	24.70	1.10	29.90	1.01	1.04	0.04
Mass rate (kg/h)	147539	237082	712704	1097538	986694	365040	569569	4120598	4690147	4690147
Molar rate (kmol/h)	8188	13160	24609	57900	51755	8721	39392	142796	171681	171681
Volume rate (m3/h)	2198	19915	76952	108933	57788	205234	51202	3303868.5	11506658	125648273
Molecular Weight	18.0	18.0	29.0	19.0	19.1	41.9	14.5	28.9	27.3	27.3
Component										
Hydrogen (mol%)	0.00	0.00	0.00	29.65	44.13	0.81	52.74	0.00	0.00	0.00
Nitrogen (mol%)	0.89	0.00	78.10	33.35	37.31	0.38	44.59	77.31	74.54	74.54
CO2 (mol%)	2.00	0.00	0.00	4.81	16.33	92.06	0.98	0.03	0.48	0.48
Methane (mol%)	89.00	0.00	0.00	0.28	0.32	0.01	0.39	0.00	0.00	0.00
Ethane (mol%)	7.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Propane (mol%)	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-Butane (mol%)	0.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-Pentane (mol%)	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen (mol%)		0.00	20.96	0.00	0.00	0.00	0.00	20.74	10.95	10.95
Argon (mol%)		0.00	0.94	0.40	0.45	0.00	0.53	0.93	0.90	0.90
CO (mol%)		0.00	0.00	10.25	0.50	0.01	0.60	0.00	0.00	0.00
Water (mol%)		100.00	0.00	21.26	0.97	6.71	0.17	0.99	13.14	13.14
Total	100	100	100	100	100	100	100	100	100	100

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Case 2 - Pre-Combustion Capture for Power Generation on Gas

CLIENT: Department for Business, Energy & Industrial Strategy
CONTRACT: 13333
NAME: Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

CHANGE	**REV - O1**				SHEET
DATE	03/04/2017				1 OF 1
ORIG. BY	R. Ray				
APP. BY	T. Tarrant				

DESCRIPTION	Electric Oper. Load	Steam (t/h)			Condensate (Note 2)	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
	KW	127 bar	35 bar	8 bar	LP (t/h)	t/h	t/h	t/h		
Syngas Production Units					-19.3	-6674.1				
Process Air Compressor	-14755									
Acid Gas Recovery Unit	-18581									
N2/H2 Compression	-12000									
CO2 Compression and Drying	-34740							4.5		
Other	-495							113.5		
Process Steam Production/Consumption		530.6	-237.1	50.1						
Process Units Total	-80571	530.6	-237.1	50.1	-19.3	-6674.1	0.0	118.0		
Power Island										
GT Gross Electrical Power	554444									
GT Auxiliaries	-2000									
HRSG & Auxiliaries	-8440	225.3	329.9	26.7			-286.7	4.3		
ST Electrical Power	364688	-755.9	-848.7	-925.5	925.5	-43791.7				
Power Island Total	908691	-530.6	237.1	-50.1	925.5	-43791.7	-286.7	4.3		
Offsites & Utilities										
Fresh Cooling Water	-7731					50465.8				
Demin Plant/Condensate Plant	-1240				-815.5		286.7			
Utility water/Fire Water	-248									
Waste Water Treatment	-528				-90.6			-122		
Buildings and Misc loads	-500									
Offsites & Utilities Total	-10247	0.0	0.0	0.0	-906.2	50465.8	286.7	-122.2		
Grand Total	**817873**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

amec foster wheeler	EQUIPMENT LIST FOR COMPRESSORS			Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Originated**	RR			**SHEET** 1 **of** 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		**Checked**	SF			
		UK Carbon Capture Technology		**Approved**	TT			
	Case:	Natural Gas IRCC with CO2 Capture		**Date**	27/04/2017			

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
400	K-401	CO2 Compressor Package (6 stages)	Multi-Stage Integrally Geared	2x50% (1/train)			1.284 / 1.910	111.9	1.1 / 113.0		0.994 / 0.745	16930	CrNi alloy	42.00	182.5 t/h CO2 per train	
							/		/		/					
							/		/		/					
							/		/		/					
							/		/		/					

Notes:

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EQUIPMENT LIST FOR HEAT EXCHANGERS	**Rev.**	**REV 01**	**REV 02**	**REV 03**
Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	**Originated**	RR		
Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**Checked**	SF		
	Approved	TT		
Case: Natural Gas IRCC with CO2 Capture	**Date**	27/04/2017		

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
200	E-201	CO Shift HP Steam Generator	Kettle	2x50% (1/train)			H:549 t/h C: 285 t/h	85.28	1172	355.0 / 134	975.0 / 42	SS316	SS316					
100	E-101	LP Steam Generator	Kettle	2x50% (1/train)			H:356.4 t/h C: 37 t/h	20.86	1328	195.0 / 10	425.0 42	SS316	SS316				Included in Unit 100 Package	
200	E-202	CO Shift Gas-Gas exchanger	Shell & Tube	2x50% (1/train)			H:236 t/h C: 192.4 t/h	35.66	10668	575.0 / 43	675.0 / 42	1.25CR-0.5Mo	1.25CR-0.5Mo					
200	E-203	CO Shift HP Steam Heater	Shell & Tube	2x50% (1/train)			H:550 t/h C: 265.3 t/h	12.68	3234	380.0 / 46	452.0 / 42	SS316	SS316					
200	E-204	CO Shift HP Steam Superheater	Shell & Tube	2x50% (1/train)			H:313 t/h C: 265.3 t/h	39.03	7356	535.0 /	675.0 / 42	SS316	SS316					
200	E-205	HP BFW Heater1	Shell & Tube	2x50% (1/train)			H:550 t/h C:285 t/h	24.38	5063	219.0 / 35	260.5 / 42	SS316	SS316					
200	E-206	HP BFW Heater2	Shell & Tube	2x50% (1/train)			H:550 t/h C:285 t/h	47.64	12152	341.0 / 6	405.0 / 42	CS	CS					
200	E-207	HP BFW Heater3	Shell & Tube	2x50% (1/train)			H:550 t/h C:285 t/h	6.19	975	353.0 / 6	399.0 / 42	CS	CS					
200	E-208	Syngas Heater	Shell & Tube	2x50% (1/train)			H:547 t/h C:285 t/h	12.14	5013	130.0 / 6	153.0 / 42	CS	CS					
200	E-209	Syngas Cooler	Shell & Tube	2x50% (1/train)			H:547 t/h C:312 t/h	39.81	3726	50.0 / 6	143.0 / 42	CS	CS					
200	E-210	Blowdown cooler	Shell & Tube	2x50% (1/train)			H:11.7 t/h C: 152.3 t/h	1.94	29	50.0 / 6	195.0 / 9.3	CS	CS					
200	E-211	Process Water cooler	Shell & Tube	2x50% (1/train)			H:55.3 t/h C: 553 t/h	6.95	125	50.0 / 6	159.0 / 37	CS	CS					
200	E-212	Process Water - water exchanger	Shell & Tube	2x50% (1/train)			H:553 t/h C: 553 t/h	7.43	3095	225.0 / 6	259.0 / 37	CS	CS					
400	E-401	CO2 Compressor Cooler - Stage 1	Shell & Tube	2x50% (1/train)			C:501.8 t/h H: 182.5 t/h	6.40	978	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
400	E-402	CO2 Compressor Cooler - Stage 2	Shell & Tube	2x50% (1/train)			C:282 t/h H: 176.6 t/h	3.60	551	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
400	E-403	CO2 Compressor Cooler - Stage 3	Shell & Tube	2x50% (1/train)			C:365.7 t/h H: 195.5 t/h	4.67	545	tubeside 50.0 / 10.0	160.0 / 15	SS304	SS304					
400	E-404	CO2 Compressor Cooler - Stage 4	Shell & Tube	2x50% (1/train)			C:322.9 t/h H: 197.2 t/h	4.12	480	tubeside 50.0 / 10.0	160.0 / 50	SS304	SS304					
400	E-405	CO2 Compressor Cooler - Stage 5	Shell & Tube	2x50% (1/train)			C:211.1 t/h H: 172.2 t/h	2.69	292	tubeside 50.0 / 10.0	100.0 / 70	SS304	SS304					
400	E-406	CO2 Compressor Cooler - Stage 6	Shell & Tube	2x50% (1/train)			C:860 t/h H: 172.2 t/h	10.97	1095	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-1001	Steam Condenser (Water Cooled)	Shell & Tube	2x50% (1/train)			H:462.7 t/h C: 21880 t/h	278.90	22962	tubeside 50.0 / 10.0	100.0 / 5	SS304	SS304					

Notes: **Notes:**

1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler	EQUIPMENT LIST FOR PUMPS				Rev.	REV 01	REV 02	REV 03	
	Client	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333	**Originated**	RR			SHEET 3 **of** 8
	Descriptior	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	SF			
		UK Carbon Capture Technology			**Approved**	TT			
	Case:	Natural Gas IRCC with CO2 Capture			**Date**	27/04/2017			

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP / SG / VISC'Y °C		cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
600	P-601	Vacumn Condensate return pump	Centrifugal	2/train		463	465	75	12		31.0	1	0.200	Design Temp: 80/-10 Design Press.: 14/FV	206	CS	One operating; One spare	
800	P-801	Process Cooling Water Pumps (closed loop)	Centrifugal	3/train	2x50% duty, 1 spare	5907	5924	75	3		25.0	1	0.900	Design Temp: 80/-10 Design Press.:9.5	617	CS	Two operating; One spare Per train	
600	P-602	Steam Condenser Cooling Water Pumps	Centrifugal	3/train	2x50% duty, 1 spare	10938	10969	75	3		25.0	1	0.900	Design Temp: 80/-10 Design Press.:9.5	2438	CS	Two operating; One spare Per train	
200	P-201	Process Water Pump	Centrifugal	1/train		108	113	75	10		170.0	1	0.900	Design Temp: 200/-10 Design Press.:15	43	CS	One operating; One spare	
600	P-603	Demin Water Pump	Centrifugal	2/train		143	144	75	7		20.0	1	1.000	Design Temp- 80 / -10 Design Pressure: 5.20	37	Stainless Steel casing and impeller	One operating; One spare	
800	P-802	Firewater Pump Package	Centrifugal	3							20.0			Design Temp- 80 / -10 Design Pressure: 19		CS	Sizing: 8 x 4 x 3.2 1 x diesel, 1 x electric and	
600	P-604	HP BFW Pump	Centrifugal	2/train		398	444	75	120.00		170	0.9	0.2		2094	CS	One operating; One spare	
600	P-605	MP BFW Pump	Centrifugal	2/train		167	186	75	28.00		170	0.9	0.2		187	CS	One operating; One spare	
600	P-606	LP BFW Pump	Centrifugal	2/train		64	71	75	2.00		170	0.9	0.2		31.0	CS	One operating; One spare	

Notes:

amec foster wheeler	**EQUIPMENT LIST FOR VESSELS**		**Rev.**	**REV 01**	**REV 02**	**REV 03**
	Client:	Department for Business, Energy & Industrial Strategy **Contract No:** 13333	**Originated**	RR		
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**Checked**	SF		
			Approved	TT		
	Case:	Natural Gas IRCC with CO2 Capture	**Date**	27/04/2017		

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	DIMENSIONS HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	DESIGN CONDITIONS PRESS barg	DESIGN CONDITIONS VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	MATERIALS OF CONST'N INTERNALS MAT./LINING/ CA	REMARKS	REV
200	**R-201**	Auto-thermal Reformer	Reactor	2x50% (1/train)	6.00	65.00		V	1500	31	1.013	Packing height 10 m	CS	Refractory Brickwork at combustion zone	550 t/hr of feed to reactor	
200	**R-202**	High Temperature Shift Reactor	Reactor	2x50% (1/train)	8.40	7.75		V	500	31	1.013		1.0 Cr - 0.5 Mo	1.0 Cr - 0.5 Mo	550 t/hr of feed to reactor	
200	**R-203**	Low Temperature Shift Reactor	Reactor	2x50% (1/train)	11.20	6.50		V	400	31	1.013		1.25 Cr - 0.5 Mo	1.25 Cr - 0.5 Mo	550 t/hr of feed to reactor	
200	**V-201**	HP Steam drum	Drum	2x50% (1/train)	2.30	4.60		V	401	31	1.014		1.25 Cr - 0.5 Mo	1.25 Cr - 0.5 Mo	551 t/hr of feed to reactor	
200	**V-202**	Condensate Seperator	Vertical drum	2x50% (1/train)	4.33	8.66		V	402	31	1.015		1.25 Cr - 0.5 Mo	1.25 Cr - 0.5 Mo	552 t/hr of feed to reactor	
200	**V-203**	Condensate Seperator	Vertical drum	2x50% (1/train)	3.90	7.80		V	403	31	1.016		1.25 Cr - 0.5 Mo	1.25 Cr - 0.5 Mo	553 t/hr of feed to reactor	
200	**V-204**	Blowdown Drum	Vertical drum	2x50% (1/train)	0.73	1.46		V	403	31	1.016		1.25 Cr - 0.5 Mo	1.25 Cr - 0.5 Mo	553 t/hr of feed to reactor	
400	**V-401**	CO2 Compressor Stage 1 KO drum	Vertical drum	2x50% (1/train)	4.15	8.30	131	V	80	1.8	1.013	Wire Mesh Pad 1.35 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
400	**V-402**	CO2 Compressor Stage 2 KO drum	Vertical drum	2x50% (1/train)	3.30	6.60	66	V	80	3.2	1.013	Wire Mesh Pad 0.86 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
400	**V-403**	CO2 Compressor Stage 3 KO drum	Vertical drum	2x50% (1/train)	2.70	5.40	36	V	80	6.2	1.013	Wire Mesh Pad 0.57 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
400	**V-404**	CO2 Compressor Stage 4 KO drum	Vertical drum	2x50% (1/train)	2.26	4.52	21	V	80	13.2	1.013	Wire Mesh Pad 0.40 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
400	**V-405**	CO2 Compressor Stage 5 KO drum	Vertical drum	2x50% (1/train)	1.85	3.70	12	V	80	31	1.013	Wire Mesh Pad 0.27 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		

Notes:

1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank

2. V - Vertical H - Horizontal

amec foster wheeler	EQUIPMENT LIST FOR TANKS				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	SF			SHEET 5 of 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology			Checked	RR			
					Approved	TT			
	Case:	Natural Gas IRCC with CO2 Capture			Date	27/04/2017			

UNIT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS ID m	HEIGHT m	TOTAL VOLUME m^3	ROOF TYPE	BOTTOM TYPE	STORAGE PRESS Bar	TEMP	SG	CORRO-SION	'MATERIALS	MANWAYS SHELL	ROOF	CONNECT-IONS	HEATING COIL	INSUL-ATION	REMARKS	REV
	Raw Water tank	1	24.0	9.6	13320	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 24 hr Storage	
	Demin Water Tank	1	23.9	9.6	3590	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 24 hr Storage	
	Firewater Storage Tank	1	10.0	6.0	471	Storage tank		Atm	20			Lined CS						Design Pres.: 0.0075 / -0.0025 Design Temp: 80/-10 deg C	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333	**Originated**	RR			**SHEET** 6 **of** 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology			**Checked**	SF			
					Approved	TT			
	Case:	Natural Gas IRCC with CO2 Capture			**Date**	27/04/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
100	**Feedstock Pre-treatment Block (2 x 50%)** Furnace Pre-heater Pressure reducing stations Booster air compressor Process air coolers Natural gas pre-heaters Syngas to Hydrogenetor compressor Hydrogenetor Desulphuriser Pre-reformer					Two package Flowrate to Package: 73.77 t/hr	
300	Acid Gas Removal Unit (Selexol)	2 x 50% Feed gas : 580000 Nm3/hr per train Operating Pressure: 25 barg	Selexol Process	11 MW Each Train		CO2 removal Total CO2 removal: 4310 t/d; Total Carbon Capture : 90%	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT			Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy	Contract No: 13333	Originated	RR			SHEET 7 of 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology		Checked	SF			
				Approved	TT			
	Case:	Natural Gas IRCC with CO2 Capture		Date	27/04/2017			

UNIT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
500	Gas Turbine & Generator Package	GE 9F Syngas Variant Gas Turbine	2 x 50% (1/train)								303.4 MW Output Turbine generator				
600	HRSG		2 x 50% (1/train)											Horizontal, Natural Draft 3 Pressure Level	
600	Phosphate Injection		2 x 50% - 1/Train												
600	Oxygen Scavanger Injection Package		2 x 50% - 1/Train												
700	Steam Turbine & Generator Package		2 x 50% (1/train)								182.6 MW Output Turbine generator				
800	Cooling Tower	Evaporative, Natural Drive Cooling Tower									Total Heat duty 950 MWth			Diameter: 145 m Height: 210 m	
800	Cooling Tower packages						14090							Filtration Package; Hypochlorite Dosing Package; Antiscalant Package	
800	Nitrogen Package	Tank/pump	1						7	Design Temp: 80/-10 deg C Design Press.: 14 barg Oper. Temp.: 20 deg C		CS			
800	Compressed Air Package	Compressor	1							Design Temp- 80 / -10 deg C Design Pressure: 8.70 barg					
400	TEG Dehydration Package		2 x 50% (1/train)							Design Temp: 150/-10 deg C Design Press.: 75 barg				Each package dehydrates 196.3 t/h CO2, removes 190.8 kg/h water to get to spec of 50 ppm	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 8 of 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Natural Gas IRCC with CO2 Capture			Date	27/04/2017			

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
600	Stack	2 x 50% - 1/Train					2345 tph flue gas @ 80 oC and 1.02 bara per train	

Notes:

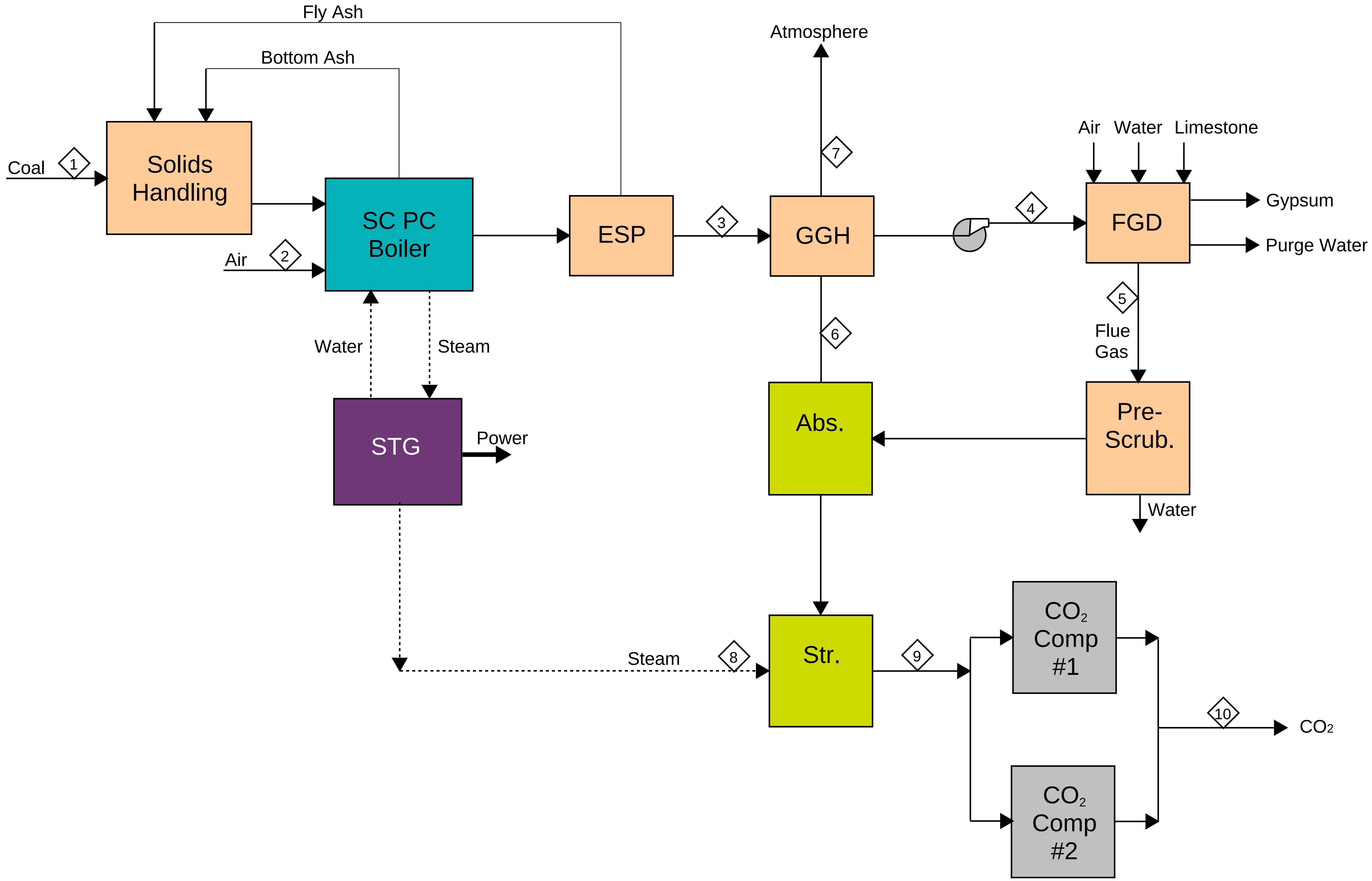
Amec Foster Wheeler | **Prepared By :** K.D. Nelson
Client : BEIS | **Base Date :** 1Q2017
Project : Novel Carbon Capture Technology Study | **Rev. No. :** '6'
Contract No.: 13333 | **Print Date :** 14-Dec-17
Case 2 : IRCC Pre-combustion Capture for Power Generation on Gas

Description	Case 2 : IRCC Pre-combustion Capture for Power Generation on Gas						
	Unit 100	Unit 200	Unit 300	Unit 400	Unit 500-700	Unit 800	
	Feedstock Pre-Treatment	ATR & Shift	Acid Gas Removal	CO_2 Compression Block	Combined Cycle Block	Utility Units	Total
	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP
Sub-Total Direct Materials	31,010,000	144,514,000	48,858,000	29,547,000	287,445,000	108,270,000	649,644,000
Other Material Costs							
Shipping / Freight	1,551,000	7,226,000	2,443,000	1,477,000	14,372,000	5,414,000	32,483,000
Third Party Inspection	310,000	1,445,000	489,000	295,000	2,874,000	1,083,000	6,496,000
Spare Parts (Comm/2yrs Op)	620,000	2,890,000	977,000	591,000	5,749,000	2,165,000	12,992,000
Sub-Total Materials	33,491,000	156,075,000	52,767,000	31,910,000	310,440,000	116,932,000	701,615,000
Material & Labour Contracts							
Civils/Steelwork & Buildings	7,753,000	17,342,000	5,863,000	4,432,000	43,117,000	21,654,000	100,161,000
Sub-Total Material & Labour Contracts	7,753,000	17,342,000	5,863,000	4,432,000	43,117,000	21,654,000	100,161,000
Labour Only Contracts							-
Mechanical	3,101,000	17,342,000	5,863,000	5,318,000	51,740,000	17,323,000	100,687,000
Electrical/Instrumentation	1,240,000	4,335,000	1,466,000	1,477,000	14,372,000	5,414,000	28,304,000
Scaffolding/Lagging/Rigging	521,000	2,601,000	879,000	815,000	7,933,000	2,728,000	15,477,000
Sub-Total Labour Only Contracts	4,862,000	24,278,000	8,208,000	7,610,000	74,045,000	25,465,000	144,468,000
Sub-Total Materials & Labour	46,106,000	197,695,000	66,838,000	43,952,000	427,602,000	164,051,000	946,244,000
EPCm Costs							-
Engineering Services/Construction Management	6,916,000	29,654,000	10,026,000	6,593,000	64,140,000	24,608,000	141,937,000
Commissioning	922,000	3,954,000	1,337,000	879,000	8,552,000	3,281,000	18,925,000
Sub-Total EPCm Costs	7,838,000	33,608,000	11,363,000	7,472,000	72,692,000	27,889,000	160,862,000
Total EPC Cost	53,944,000	231,303,000	78,201,000	51,424,000	500,294,000	191,940,000	1,107,106,000
Other Costs							-
Pre-Licensing, Technical and Design etc	539,000	2,313,000	782,000	514,000	5,003,000	1,919,000	11,070,000
Regulatory, Licensing and Public Enquiry etc	1,150,000	4,931,000	1,667,000	1,096,000	10,665,000	4,092,000	23,601,000
Infrastructure Connection Costs						37,000,000	37,000,000
Owners Costs	3,776,000	16,191,000	5,474,000	3,600,000	35,021,000	13,436,000	77,498,000
Sub-Total Other Costs	5,465,000	23,435,000	7,923,000	5,210,000	50,689,000	56,447,000	149,169,000
Total Project Costs	**59,409,000**	**254,738,000**	**86,124,000**	**56,634,000**	**550,983,000**	**248,387,000**	**1,256,275,000**

ATTACHMENT 5: Case 3 – SCPC with Post-Combustion Capture[9]

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List

[9] Please note, as the basis of design for Benchmark 3 is based on Shell Cansolv's proprietary design, a capital cost estimate is not included to maintain confidentiality.



NOTES

Department for Business, Energy & Industrial Strategy

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REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	SF	RR	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram - Case 3 - SCPC with Post Combustion CO_2 Capture

Drawing No

Rev

HEAT & MATERIAL BALANCE - Case 3 - Post-Combustion Capture for Power Generation on Coal

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**SHEET**
CONTRACT:	13333	**DATE**	08/05/2017	1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S. Ferguson	
		APP. BY	T. Tarrant	

Stream Number	1	2	3	4	5	6	7	8	9	10
Stream Name	B/L Coal Feed	Total Air Intake	Flue Gas from ESP	Flue Gas to FGD	Flue Gas To CO2 Capture	Flue Gas from Absorber	Flue Gas to Atmosphere	LP Steam to Reboilers	CO2 to Compression	Product CO2
Pressure (bar abs)	1.01	1.01	1.01	1.01	1.01	1.02	1.01	4.41	2.00	110.00
Temperature (°C)	9.0	9.0	132.0	90.0	47.0	33.0	90.0	148.0	49.0	30.0
Mass rate (kg/h)	325000	3383000	3667000	3667000	3741000	2887739	2887739	775000	709690	692310
Molar rate (kmol/h)		117198	123412	123412	127470	103366	103366	43008	16693	15732
Volume rate (m3/h)		2712911	4102972	3799916	3329942	2592200	3079852	340676	224142	928
Component										
Hydrogen (mol%)	%wt AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO (mol%)	C: 64.6%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2 (mol%)	H: 4.38%	0.03	14.06	14.06	13.68	1.69	1.69	0.00	94.25	100.00
Nitrogen (mol%)	O: 7.02%	77.27	73.56	73.56	71.40	88.30	88.30	0.00	0.00	0.00
Oxygen (mol%)	S: 0.86%	20.73	3.28	3.28	3.20	3.96	3.96	0.00	0.00	0.00
Argon (mol%)	N: 1.41%	0.92	0.87	0.87	0.84	1.00	1.00	0.00	0.00	0.00
Methane (mol%)	Cl: 0.03%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Water (mol%)	Moisture: 9.5%	1.05	8.16	8.16	10.88	5.05	5.05	100.00	5.75	0.00
SO2 (mol%)	Ash: 12.2%	0.00	0.07	0.07	0.00	0.00	0.00	0.00	0.00	0.00
Total		100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

amec foster wheeler

UTILITIES BALANCE SUMMARY - Case 3 - Post-Combustion Capture for Power Generation on Coal

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**			**SHEET**
CONTRACT:	13333	**DATE**	18/05/2017			1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S Ferguson			
		APP. BY	T Tarrant			

UNIT No.	**DESCRIPTION**	**Electric Oper. Load**	**Coal Feed LHV**	**Steam (t/h)**			**Condensate (t/h)**			**Cooling water**	**REV**
		MW	**MW**	**HPS 171 barg**	**MPS 40 barg**	**LPS 4.8 barg**	**HP T/h**	**MP T/h**	**LP T/h**	**T/h**	
Solids Handling and Boiler Island											
	Feedstock and Solids Handling	-3.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Boiler & Auxilliaries (inc ID Fan)	-22.4	-2335.5	2924.0	2362.0	2145.8	0.0	0.0	0.0	0.0	
	Flue Gas Desulphurisation	-4.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Power Island Auxilliaries (ST)	-4.8	0.0	0.0	0.0	-186.8	0.0	0.0	186.8	0.0	
	Power Island (ST)	953.5	0.0	-2924.0	-2362.0	-1189.0	0.0	0.0	880.5	-58157.8	
	Power Island Net	919.0	-2335.5	0.0	0.0	0.0			1067.3	-58157.8	
CO2 Capture											
	Flue Gas Blower	-16.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Pumps	-6.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Thermal Reclaimer Reboiler	-0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	MVR Compressor	-14.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Solvent Reboiler	0.0	0.0	0.0	0.0	-770.0	0.0	0.0	770.0	0.0	
	CO2 Capture Net	-37.2				-770.0			770.0	-50170.0	
CO2 Compression & Pumping											
	CO2 Compressor	-49.7		0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Supercritical CO2 Pump	-1.5		0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Inter/After Coolers & Condenser	0.0		0.0	0.0	0.0	0.0	0.0	0.0	-4253.0	
	CO2 Capture Net	-51.1								-4253.0	
Utilities & Offistes Systems											
	Cooling Water System	-15.0								112680.8	
	Demin Plant	0.0									
	Buildings and Misc extra loads	-1.4								-100.0	
	Effluent Treatment	0.0									
	U&O Net	-16.5									
Overall Performance											
	Gross Production	953.5	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Total Parasitic Loads	-139.3	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Net Produced	814.2	-2335	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Net LHV Efficiency	34.86%	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	

amec foster wheeler	EQUIPMENT LIST FOR COMPRESSORS		Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy **Contract No:** 13333	**Originated**	SF	SF		**SHEET** 1 **of** 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**Checked**	RR	RR		
			Approved	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture	**Date**	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
K-201	ID Fan	Axial	1		4,031,071	1.364 / 1.363	0.13	0.9 / 1.0		0.999 / 0.999	16350	CS	29.71		02
K-202	MVR Compressor	Centrifugal	1			1.326 / 1.317	1.1	1.1 / 2.1		0.993 / 0.992			18.01	**CO2 Capture Unit Scope**	02
K-001	CO2 Compressor Package (4 stages)	Multi-Stage Integrally Geared	2 x 50%		109,990	1.285 / 1.592	70.0	2.0 / 72.0		0.990 / 0.829	24835	CrNi alloy	42.52	346.5 t/h CO2	
						/		/		/					
						/		/		/					
						/		/		/					

Notes:

amec foster wheeler	EQUIPMENT LIST FOR HEAT EXCHANGERS				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy	Contract No:	13333	Originated	SF	SF		SHEET 2 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR	RR		
		UK Carbon Capture Technology			Approved	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture			Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS /TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
E-001	Pre-Scrubber Cooler	Gasketed Plate and frame	1						51.0 / 10.0	80.0 / 7						CO2 Capture Unit Scope	
E-002	CO2 Absorber Intercooler	Gasketed Plate and frame	3						51.0 / 10.0	80.0 / 7						CO2 Capture Unit Scope	
E-003	CO2 Wash Water cooler	Gasketed Plate and frame	1						51.0 / 10.0	80.0 / 7						CO2 Capture Unit Scope	
E-004	Lean amine cooler	Gasketed Plate and frame	2						51.0 / 10.0	140.0 / 9.5						CO2 Capture Unit Scope	
E-005	Lean/Rich exchangers	Gasketed Plate and frame	15						80.0 / 10.3	140.0 / 6						CO2 Capture Unit Scope	
E-006	Stripper condenser	Welded Plate and frame	2						51.0 / 10.0	150.0 2/FV						CO2 Capture Unit Scope	
E-007	Reboilers	Welded Plate and frame	14						150.0 / 2/FV	295.0 / 5/FV (tubeside)						CO2 Capture Unit Scope	
E-008	Gas / Gas Exchanger		1						75.0 / 2.000	157 / 2.0	CS	CS					
E-101	CO2 Compressor Cooler - Stage 1	Shell & Tube	2				13.06	2415	tubeside 51.0 / 10.0	150.0 / 8.2	SS304	SS304					
E-102	CO2 Compressor Cooler - Stage 2	Shell & Tube	2				8.23	1318	tubeside 51.0 / 10.0	150.0 / 12.3	SS304	SS304					
E-103	CO2 Compressor Cooler - Stage 3	Shell & Tube	2				8.87	1233	tubeside 51.0 / 10.0	150.0 / 33	SS304	SS304					
E-104	CO2 Condenser	Shell & Tube	2				24.46	1263	tubeside 51.0 / 10.0	150.0 / 76	SS304	SS304					
E-105	CO2 Product Cooler	Shell & Tube	2				1.35	207	tubeside 51.0 / 10.0	100.0 / 115	SS304	SS304					

Notes: **1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type**
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler	EQUIPMENT LIST FOR PUMPS				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy	**Contract No:**	13333	**Originated**	SF	SF		SHEET 3 **of** 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	RR	RR		
		UK Carbon Capture Technology			**Approved**	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture			**Date**	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP °C	SG	VISC'Y cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
P-001	Prescrubber Pumps	Centrifugal	3 x 50%					3.1		30.0	1.004	0.797	Design Temp: 80/-10 Design Press.: 7.5			**CO2 Capture Unit Scope**	
P-002	Absorber WW pump	Centrifugal	3 x 50%					3.1		30.0	1.004	0.797	Design Temp: 80/-10 Design Press.: 7			**CO2 Capture Unit Scope**	
P-003	CO2 Intercooler Pumps	Centrifugal	3 x 50%					1.7		30.0	1.004	0.797	Design Temp: 80/-10 Design Press.: 7			**CO2 Capture Unit Scope**	
P-004	Rich amine pumps	Centrifugal	3 x 50%					4.5		36.0	0.983		Design Temp: 80/-10 Design Press.: 11			**CO2 Capture Unit Scope**	
P-005	Stripper reflux pump	Centrifugal	3 x 50%					4.8		137.0	0.919	0.205	Design Temp: 295/-10 Design Press.: 12.4/FV			**CO2 Capture Unit Scope**	
P-006	Lean amine pumps	Centrifugal	3 x 50%					2.1		119.0	0.95		Design Temp: 140/-10 Design Press.: 8/FV			**CO2 Capture Unit Scope**	
P-007	Lean amine feed pumps	Centrifugal	3 x 50%					5.9		50.0	0.95		Design Temp: 140/-10 Design Press.: 8/FV			**CO2 Capture Unit Scope**	
P-008	Amine Drain Pump	Centrifugal	1 x 100%					2.1		60.0	0.983		Design Temp: 100/-10 Design Press.:2.40			**CO2 Capture Unit Scope**	
P-101	Supercritical CO2 Pump	Centrifugal	2 x 50%		346.2	493.5		40.0		25.0	0.702	0.061	Design Temp: 140/-10 Design Press.:140.0	731.1	304L SS		
	Demin Water Pump	Pump	3					1		20.0			Design Temp- 80 / -10 Design Pressure: 5.20		Stainless Steel casing and impeller	Sizing: 1.7 x 0.7 x 1.4	
	Towns Water Transfer Pump	Centrifugal	3					1		20.0			Design Temp- 80 / -10 Design Pressure: 6.0		Cast iron casing with Stainless steel impeller	Size: 1.8 X 1.8 X 1.8	
	Firewater Pump Package	Pump	3							20.0			Design Temp- 80 / -10 Design Pressure: 19		CS	Sizing: 8 x 4 x 3.2 1 x diesel, 1 x electric and 1 x jockey	
	CO2 Compressor Condensate Return Pump	Pump	2 x 100%		8.6	8.6		3		25.0	0.996	0.900	Design Temp- 80 / -10 Design Pressure: 6.0	1	CS		

Notes:

amec foster wheeler	EQUIPMENT LIST FOR VESSELS				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy	Contract No:	13333	Originated	SF	SF		SHEET 4 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology			Checked	RR	RR		
					Approved	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture			Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	PRESS barg	VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	INTERNALS MAT./LINING/ CA	REMARKS	REV
C-001	Prescrubber	Rectangular column	1				V	150	0.30	1.013				**CO2 Capture Unit Scope**	02
C-002	Absorber (absorber section)	Rectangular column	1				V	80/-10	0.10	1.013				**CO2 Capture Unit Scope**	02
C-003	Absorber (water wash section)	Rectangular column	1				V	80/-10	0.10	1.013				**CO2 Capture Unit Scope**	02
C-005	Stripper	Vertical cylinder	1				V	250/-10	3.50	FV				**CO2 Capture Unit Scope**	
V-001	Condensate Flash Vessel	Horizontal Drum	1				H	250/-10	3.50	FV				**CO2 Capture Unit Scope**	02
V-002	CO2 Reflux Accumulator	Horizontal Drum	1				H	295/-10	5.00	FV				**CO2 Capture Unit Scope**	02
V-003	Lean Flash Vessel	Horizontal Drum	1				H	295/-10	5.00	FV				**CO2 Capture Unit Scope**	02
V-101	CO2 Compressor Stage 1 KO drum	Vertical drum	2	5.00	10.00	229	V	80	3.50	1.013	demister	SS304	SS304	346.5 t/h CO2	
V-102	CO2 Compressor Stage 2 KO drum	Vertical drum	2	4.00	8.00	117	V	80	6.00	1.013	demister	SS304	SS304	346.5 t/h CO2	
V-103	CO2 Compressor Stage 3 KO drum	Vertical drum	2	3.10	6.20	55	V	80	12.00	1.013	demister	SS304	SS304	346.5 t/h CO2	
V-104	CO2 Compressor Stage 4 KO drum	Vertical drum	2	2.40	4.80	25	V	80	33.00	1.013	demister	SS304	SS304	346.5 t/h CO2	

Notes:
1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

amec foster wheeler	EQUIPMENT LIST FOR TANKS		Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy Contract No: 13333	Originated	SF	SF		SHEET 5 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR	RR		
			Approved	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture	Date	03/02/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS ID m	DIMENSIONS HEIGHT m	TOTAL VOLUME m^3	ROOF TYPE	BOTTOM TYPE	STORAGE PRESS	STORAGE TEMP	SG	CORRO-SION	'MATERIALS	MANWAYS SHELL	MANWAYS ROOF	CONNECT-IONS	HEATING COIL	INSUL-ATION	REMARKS	REV
T-001	Lean Amine Tank	1				Storage tank		0.02	30						Vertical, sized for full inventory			**CO2 Capure Unit Scope**	
T-002	Amine Drain Tank	1													Horizontal, underground.			**CO2 Capure Unit Scope**	
T-003	Absorbent Make-up tank	1				Storage tank		0.02	30						Vertical			**CO2 Capure Unit Scope**	02
T-006	Towns Water Storage Tank	1	10.00	10.00	780.00	Vertical cylindrical		0.0075 / -0.0025	20			Lined CS							
T-007	Demin Water Tank	1	10.00	9.00	700.00	Tank		0.90	20			Lined CS							
T-008	Firewater Storage Tank	1	13.0	7.8		Tank			20			Lined CS							

Notes:

	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	Client:	Department for Business, Energy & Industrial Strategy	Contract No:	13333	Originated	SF	SF		SHEET 6 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR	RR		
		UK Carbon Capture Technology			Approved	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture			Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
200	Supercritical Boiler: Coal Mill Fuel feeding system One fired boiler furnace Low NOx burners system including main burners and pilots Economiser/superheater coils and water wall circuit Reheating coils Air pre-heater Ash Collection hoppers Combustion air fan with electric motor	2868 t/h main steam				Thermal Input: 2435 MWth (HHV), 2335 MWth (LHV) Main steam conditions: 270 bar (abs), 600 °C Reheat steam conditions: 60 bar (abs), 620 °C 2 x 60% primary air, 2 x 60% secondary air	
200	Electrostatic Precipitator	Removal efficiency 99.9%					
200	Stack						
200	Continuous Emissions Monitoring System						
200	SCR System: Reactor Casing Catalyst Bypass System Ammonia Injection System Handling Equipment Control System						
200	Wet Flue Gas Desuplhurisation Unit Limestone Feeder Absorber tower Oxydation air blower Make-up water system Limestone slurry preparations Reagent feed pump Gypsum dewatering Miscellaneous Equipment	Flue gas inlet flowrate = 2766,000 Nm3/h Removal Efficiency = 98.5 %					

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy	**Contract No:**	13333	**Originated**	SF	SF		SHEET 7 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	RR	RR		
		UK Carbon Capture Technology			**Approved**	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture			**Date**	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
S-001	Ion Exchange Package		1											**CO2 Capture Unit Scope**	
S-002	Thermal Reclaimer Package	Vacuum Distillation Column	1											**CO2 Capture Unit Scope**	
S-003	CO2 Absorbent Filtration Unit	Cartridge Type Filter												**CO2 Capture Unit Scope**	
S-004	Filtration Unit	Cartridge Type Filter												**CO2 Capture Unit Scope**	
S-003	Nitrogen Package	Tank/pump	1						7	Design Temp: 80/-10 deg C Design Press.: 14 barg Oper. Temp.: 20 deg C		CS			
S-004	Compressed Air Package	Compressor	1							Design Temp- 80 / -10 deg C Design Pressure: 8.70 barg					
TEG-101	TEG Dehydration Package		2							Design Temp: 150/-10 deg C Design Press.: 76 barg				Dehydrates 346.4 t/h CO2, removes 245 kg/h water to get to spec of 50 ppm	
P-001	Flow metering and analyser package	Metering	1					692400 kg/h CO2		Design Temp: 150/-10 deg C Design Press.: 114 barg				Fiscal metering package	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR SOLIDS HANDLING EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy	Contract No:	13333	Originated	SF	SF		SHEET 8 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR	RR		
		UK Carbon Capture Technology			Approved	TT	TT		
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture			Date	08/05/2017	15/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Coal Handling System:	325 t/h coal feed				30 Days storage:	
	Wagon tipper					Storage piles = 2 x 128,000 tonnes	
	Receiving Hopper, vibratory feeder and belt extractor						
	Conveyors		Belt type				
	Transfer Towers		Enclosed type				
	As-Received Coal Sampling System		Two-stage type				
	As-Recived Magnetic separator system		Magnetic plate type				
	Crusher Towers		Impactor reduction				
	As-Fired Coal sampling system		Swing hammer type				
	As-Fired Magnetic separator system		Magnetic plate type				
	Coal silos	2 x 4900 m3				for daily storage	
	Filters						
	Fan						
	Limestone Handling System:	9.2 t/h limestone feed				30 Days storage:	
	Wagon tipper					Limestone Storage Volume = 6000 m3	
	Receiving Hopper, vibratory feeder and belt extractor						
	Conveyors		Belt type				
	Transfer Towers		Enclosed type				
	Limestone sampling system		Swing hammer type				
	Separator system		Magnetic plate type				
	Limestone Mills						
	Limestone Silo	1 x 200 m3				for daily storage	
	Filters						
	Fan						
	Ash Handling System:	12.5 t/h bottom ash				14 Days Storage	
	Ash Storage Silos					Bottom ash storage volume = 6000 m3	
	Ash Conveyors						
	Bottom ash crusher	29.2 t/h fly ash				14 Days Storage	
	Pneumatic conveying system					Fly ash storage volume = 14000 m3	
	Compressors						
	Filters						
	Fan						
	Gypsum Handling System:	16.9 t/h gypsum				30 Days storage:	
	Storage Unit					Gypsum Storage Volume = 9360 m3	
	Conveyors					1 operating, 1 spare	

Notes:

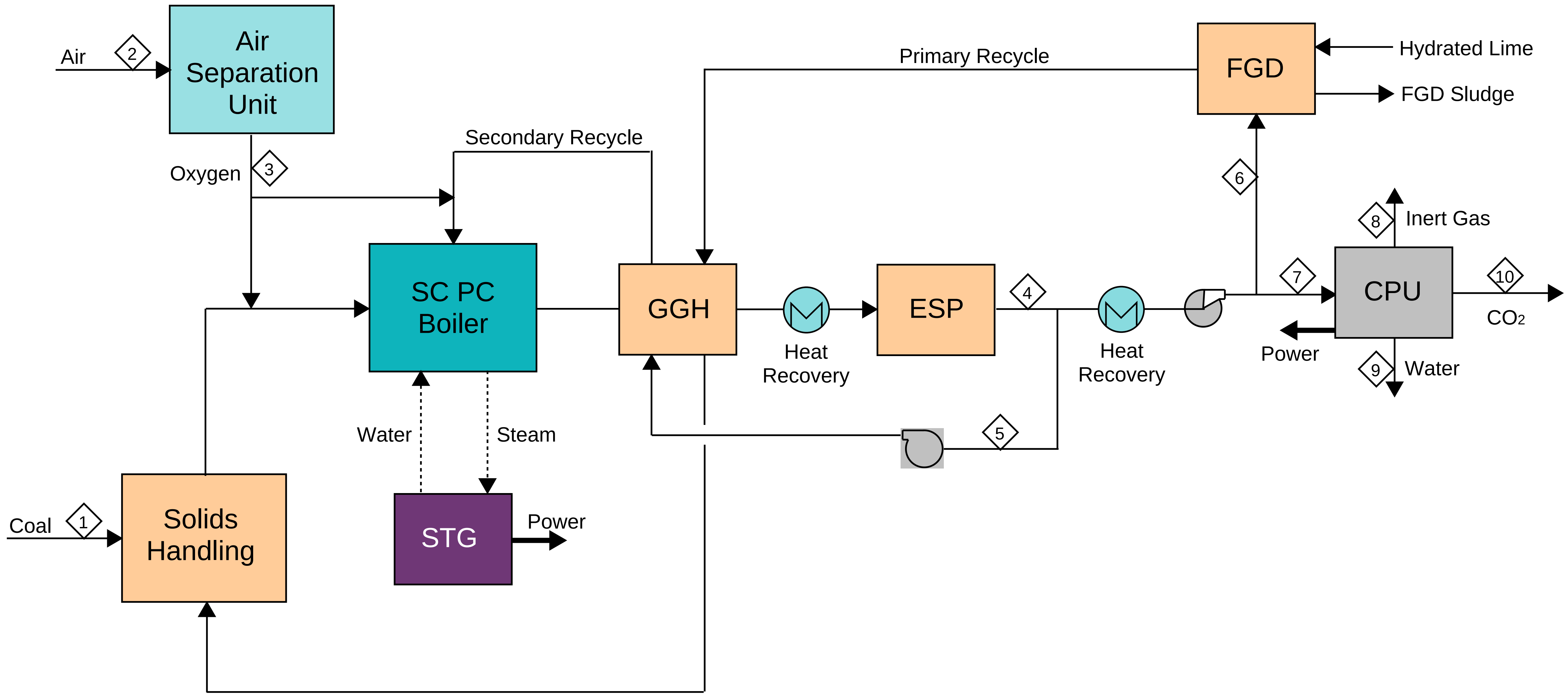
amec foster wheeler	**EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT**		**Rev.**	**REV 01**	**REV 02**	**REV 03**
	Client:	Department for Business, Energy & Industrial Strategy **Contract No:** 13333	**Originated**	SF	SF	
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	**Checked**	RR	RR	
	Unit No:	UK Carbon Capture Technology	**Approved**	TT	TT	
	Case:	Supercritical Pulverised Coal with Post Combustion CO2 Capture	**Date**	08/05/2017	15/05/2017	

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Flue gas ducting from ESP to GGH		5 m X 5 m Estimated Length 30 m	Square		CS	Flow Rate: 3667 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from GGH to fan		5 m X 5 m Estimated Length 5 m	Square		CS	Flow Rate: 3667 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from fan to FGD		5 m X 5 m Estimated Length 5 m	Square		CS	Flow Rate: 3667 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from FGD to Prescrubber		4.5 m X 4.5 m Estimated Length 15 m	Square		CS	Flow Rate: 3741 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from prescrubber to absorber		4 m X 4 m Estimated Length 27 m	Square		CS	Flow Rate: 3598 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from absorber to GGH		4 m X 4 m Estimated Length 75 m	Square		CS	Flow Rate: 2888 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02
	Flue gas ducting from GGH to Power Island stack		4.5 m X 4.5 m Estimated Length 10 m	Square		CS	Flow Rate: 2888 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	02

Notes:

ATTACHMENT 6: Case 4 – SCPC with Oxy-Combustion

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List
- Capital Cost Estimate Summary



NOTES

Department for Business, Energy & Industrial Strategy

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REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	SF	RR	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram - Case 4 - SCPC with Oxy Combustion CO_2 Capture

Drawing No

Rev

amec foster wheeler

HEAT & MATERIAL BALANCE - Case 4 - Oxy-Combustion Capture for Power Generation on Coal

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**
CONTRACT:	13333	**DATE**	16/02/2017
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S. Ferguson
		APP. BY	T. Tarrant

SHEET 1 OF 1

Stream Number	1	2	3	4	5	6	7	8	9	10
Stream Name	B/L Coal Feed	Air to ASU	Oxygen to Boiler	Flue Gas from ESP	Secondary Recycle	Primary Recycle to FGD	Flue gas to CPU	Inert gas from CPU	Water from CPU	Product CO2
Pressure (bar abs)	1.01	1.01	1.01	1.01	1.01	1.01	1.01	1.60	2.00	110.00
Temperature (°C)	9.0	9.0	9.0	160.0	170.0	38.0	38.0	210.0	25.0	30.0
Mass rate (kg/h)	325000	2920000	694000	3400000	1490000	850000	932880	222000	20470	685985
Molar rate (kmol/h)		101200	21635	92700	40630	21210	23290	6630	930	15587
Volume rate (m3/h)										
Component										
Hydrogen (mol%)	%wt AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO (mol%)	C: 64.6%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2 (mol%)	H: 4.38%	0.03	0.00	62.96	62.96	74.41	74.41	24.50	0.03	99.99
Nitrogen (mol%)	O: 7.02%	77.27	1.20	11.44	11.44	13.54	13.54	47.24	0.00	0.00
Oxygen (mol%)	S: 0.86%	20.73	97.00	5.29	5.29	6.25	6.25	21.41	0.00	0.01
Argon (mol%)	N: 1.41%	0.92	1.80	1.53	1.53	1.82	1.82	6.33	0.00	0.00
Water (mol%)	Cl: 0.03%	1.05	0.00	18.60	18.60	3.76	3.76	0.52	94.55	0.00
SO2 (mol%)	Moisture: 9.5%	0.00	0.00	0.17	0.17	0.18	0.18	0.00	0.02	0.00
NO / NO2 (mol%)	Ash: 12.2%	0.00	0.00	0.01	0.01	0.04	0.04	0.00	0.00	0.00
HNO3 (mol%)		0.00	0.00	0.17	0.17	0.18	0.18	0.00	0.99	0.00
H2SO4 (mol%)		0.00	0.00	0.01	0.01	0.04	0.04	0.00	4.41	0.00
Total		100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

UTILITIES BALANCE SUMMARY - Case 4 - Oxy-Combustion Capture for Power Generation on Coal

CLIENT:	Department for Business, Energy & Industrial Strategy			**CHANGE**			**REV - O1**				**SHEET**
CONTRACT:	13333			**DATE**			15/03/2017				1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology			**ORIG. BY**			S Ferguson				
				APP. BY			T Tarrant				
UNIT No.	**DESCRIPTION**	**Electric Oper. Load**	**Coal LHV**	**Steam (t/h)**			**Condensate / BFW (t/h)**			**Cooling water**	**REV**
		MW	**MW**	**HPS 171 barg**	**MPS 40 barg**	**LPS 4.8 barg**	**HP T/h 326 barg**	**MP T/h 8.5 barg**	**LP T/h 0.04 bar (abs)**	**T/h**	
Air Separation Unit											
	Main Air Compressors	-111.1									
	Booster Air Compressor & Pumps	-18.3									
	Air Separation Unit Net	-129.3								-1250.0	
Solids Handling and Boiler Island											
	Feedstock and Solids Handling	-3.26									
	Boiler & Auxilliaries (inc FGD & ID&SR Fans)	-14.37	-2335.0	2900.0	2900.0	2195.0	-2900.0	824.0	2076.0	-11080.0	
	Steam Cycle Auxilliaries	-6.55			-664.0	-185.0	2900.0	-824.0	-2076.0	-5080.0	
	Power Island (ST)	1097.7		-2900.0	-2195.0	-2010.0				-94745.0	
	Solids Handling and Boiler Island Net	1073.5	-2335.0	0.0	41.0	0.0	0.0	0.0	0.0	-110905.0	
CO2 Purification and Compression											
	Temperature Swing Adsorption	0.0			-41.0						
	Flue gas compression	-83.4									
	CO2 Compression	-26.8									
	Overhead Recycle	-0.8									
	Expander	15.1									
	CO2 Purification & Compression Net	-95.9			-41.0					-9015.0	
Utilities & Offistes Systems											
	Cooling Water System	-14.3								121270	
	Buildings and Misc extra loads	-1.4								-100.0	
	U&O Net	-15.7								121170.0	
Overall Performance											
	Gross Production	1112.8	0	0.0	0.0	0.0	0.0	0.0	0.0	121270.0	
	Total Parasitic Loads	-280.2	0	0.0	0.0	0.0	0.0	0.0	0.0	-121270.0	
	Net Produced	832.6	-2335	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Net LHV Efficiency	35.7%	0								

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EQUIPMENT LIST FOR COMPRESSORS	Rev.	REV 01	REV 02	REV 03	
Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	Originated	SF			SHEET 1 of 10
Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR			
	Approved	TT			
Case: Coal Oxy SCPC with CO2 Capture	Date	26/04/2017			

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
400	**K-4001 A/B**	ID Fan	Axial ID Fan	2 x 50%	Motor	575,000	1.304 / 1.302	0.1	0.93 1.01		0.996 / 0.996	2880	CS	40.20		
200	**K-2002 A/B**	Secondary Recycle Fan	Forced Draft Fan	2 x 50%	Motor	788,000	1.278 / 1.275	0.1	0.93 1.01		0.998 / 0.998	3786	CS	36.74	Included in PK-201 Oxycombustion Boiler	
700	**K-7001 A/B**	CO2 compressor Stage 1-3	Centrifugal	2 x 50%	Motor	283,450	1.302 / 1.253	14.4	1.1 / 15.5		0.996 / 0.998	48000	SS	40.09	Included in PK-701 Sour Compression	
700	**K-7002 A/B**	CO2 compressor (1 Stage)	Centrifugal	2 x 50%	Motor	18,735	1.381 / 1.379	18.0	14.0 / 32.0		0.942 / 0.942	15000	SS	40.84	Included in PK-703 Cold Section	
700	**K-7003 A/B**	Recycle Overheads Compressors	Centrifugal	2 x 50%	Motor							1000			Included in PK-703 Cold Section	
700	**K-7004 A/B**	Inerts Expander	Centrifugal	1 x 100%	Generator	11,760	1.329 / 1.361	28.8	30.0 / 1.2		1.006 / 0.999	16600	CS	33.63	Included in PK-703 Cold Section	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR HEAT EXCHANGERS	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Industrial Strategy **Contract No:** 13333	Originated	SF			SHEET 2 of 10
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR			
		Approved	TT			
	Case: Coal Oxy SCPC with CO2 Capture	Date	26/04/2017			

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	MATERIAL TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
200	**E-2003**	BFW Economiser #1	Coil	1 x 100%			H:3405 t/h C: 2900 t/h	66.50	3594	207.0 / 406	573.0 / 0	N/A	SS304					
200	**E-2004**	BFW Economiser #2	Coil	1 x 100%			H:3405 t/h C: 2900 t/h	360.00	40903	388.0 / 406	547.2 / 0	N/A	SS304					
200	**E-2005**	Condensate Heater #1	Coil	1 x 100%			H:3405 t/h C: 2076 t/h	101.00	47506	163.0 / 406	185.0 / 0	N/A	SS304					

Notes: **Notes:**

1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler	EQUIPMENT LIST FOR PUMPS	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	Originated	SF			SHEET 3 **of** 10
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	Checked	RR			
	UK Carbon Capture Technology	Approved	TT			
	Case: Coal Oxy SCPC with CO2 Capture	Date	26/04/2017			

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP °C	SG	VISC'Y cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
1100	P-1101 A-F	Cooling Water Pumps (primary system)	Centrifugal	6 x 16.7%	Motor	16000	16000	75	2		25.0	1	1.170	Design Temp: 80/-10 Design Press.:9.5	1900	CS	Six Operating	
1100	P-1102 A/B/C	Cooling Water Pumps (secondary system)	Centrifugal	3 x 50%	Motor	13500	13500	75	2		25.0	1	1.170	Design Temp: 80/-10 Design Press.:9.5	2200	CS	Two operating; One spare	
1100	P-1103 A/B	Cooling Tower make-up pumps	Centrifugal	2 x 100%	Motor	2400	2400	75	3		25.0	1	1.170	Design Temp: 80/-10 Design Press.:9.5	300	CS	One operating; One spare	
1100	P-1104 A/B	Raw Water Pumps	Centrifugal	2 x 100%	Motor	10	10	75	4		9.0	1	1.170	Design Temp: 80/-10 Design Press.:9.5	5.5	CS	One operating; One spare	
1100	P-1105 A/B	Demin Water Pump	Centrifugal	2 x 100%	Motor	5	5	75	7		20.0	1	1.000	Design Temp- 80 / -10 Design Pressure: 5.20	3.5	Stainless Steel casing and impeller	One operating; One spare	
1100	P-1106 A/B	Firewater Pump Package	Centrifugal	3							20.0	1	0.900	Design Temp- 80 / -10 Design Pressure: 19		CS	Sizing: 8 x 4 x 3.2 1 x diesel, 1 x electric and	
200	P-2001 A/B	Main BFW Pumps	Centrifugal	2 x 50%	Steam Turbine	1600	1778	81	319.00	6.00	148	0.9	0.2		16900	CS	Two operating	
200	P-2002	Start-up BFW Pumps	Centrifugal	1 x 100%	Motor	1600	1778	81	319.00		148	0.9	0.2		16900	CS	Two operating	
200	P-2003 A/B	Condensate Pumps	Centrifugal	2 x 100%	Motor	2500	2500	75	9.50		29	1	0.82		1120	CS	Two operating	

Notes:

		EQUIPMENT LIST FOR VESSELS	Rev.	REV 01	REV 02	REV 03	
amec foster wheeler		**Client:** Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	Originated	SF			SHEET 4 of 10
		Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	Checked	RR			
		UK Carbon Capture Technology	Approved	TT			
		Case: Coal Oxy SCPC with CO2 Capture	Date	26/04/2017			

EQUIPMENT NUMBER		DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	DIMENSIONS HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	DESIGN CONDITIONS PRESS barg	DESIGN CONDITIONS VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	MATERIALS OF CONST'N INTERNALS MAT./LINING/ CA	REMARKS	REV
200	**V-2001**	Deaerator	Horizontal	1 x 100%				H	173	7	1.013		CS	CS	Includes vertical section with 3 trays.	

Notes:

1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

amec foster wheeler	EQUIPMENT LIST FOR TANKS	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	Originated	SF			SHEET 5 of 10
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR			
		Approved	TT			
	Case: Coal Oxy SCPC with CO2 Capture	Date	26/04/2017			

UNIT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS ID m	DIMENSIONS HEIGHT m	TOTAL VOLUME m^3	ROOF TYPE	BOTTOM TYPE	STORAGE PRESS Bar	STORAGE TEMP °C	SG	CORRO-SION	'MATERIALS	MANWAYS SHELL	MANWAYS ROOF	CONNECT-IONS	HEATING COIL	INSUL-ATION	REMARKS	REV
300	Liquid Oxygen Storage Tank	1 common tank	5.0	9.8	193	Refrig. Storage tank		Atm	-184									Design Temp: 80/-184 deg C Design Pres. Atm 200 tonnes liquid Oxygen Storage	
1100	Raw Water storage tank	1	4.0	9.5	120	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 24 hr Storage	
1100	Demin Water Storage Tank	1	4.0	4.8	60	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 24 hr Storage	
1100	Firewater Storage Tank	1	13.0	7.8	1035	Storage tank		Atm	20			Lined CS						Design Pres.: 0.0075 / -0.0025 Design Temp: 80/-10 deg C	

Notes:

	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	SF			SHEET 6 of 10
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Coal Oxy SCPC with CO2 Capture			Date	26/04/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Supercritical Boiler:	2900 t/h main steam				Thermal Input:	
200	Coal Mill					2443 MWth (HHV), 2335 MWth (LHV)	
	Fuel feeding system						
	One fired boiler furnace					Main steam conditions: 270 bar (abs), 600 °C	
	Low Nox burners system including main burners and pilots					Reheat steam conditions: 60 bar (abs), 620 °C	
	Economiser/superheater coils and water wall circuit						
	Reheating coils						
	Tubular gas/gas heater	117 MWth				Approximate Duty	
	Secondary flue gas recycle fans with electric motor (2 x 60%)						
	Flue Gas ducts						
	Start-up system						
	Ash Collection hoppers						
	Bottom Ash Cooling devices						
	Flue Gas / BFW heat exchanger	108 MWth				Approximate Duty	
	Flue Gas / Condensate heat exchanger	28 MWth				Approximate Duty	
300	Air Separation Unit (ASU) (3 x 33%)	3 x 231 t/h 97% pure O2, delivered at atmospheric pressure.	Cryogenic	43.1 MW each		Three ASU packages, including main air compressors, booster compressors, compander, air purification unit, main heat exchanger, ASU compander, ASU column system, pumps, ASU chiller	
200	Electrostatic Precipitator	Removal efficiency 99.9%					
200	Contact Cooler						
200	Stack					For start up and upset conditions.	
	Flue Gas Desuplhurisation Unit	476000 Nm3/h					
	Flue gas ducts with dampers	Removal efficiency = 95%					
200	One Circulating fluid bed scrubber (CFBS)						
	Internal scrubber equipment						
	One fabric filter with recirculating system						
	Air Blowers						
	Silo for absorbent						
	Product silo for residue						
	Water storage tanks						
	Water injection system for CFBS						
	Lime hydration system						

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT			Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Industrial Strategy	**Contract No:**	13333	**Originated**	SF			**SHEET** 7 **of** 10
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	RR			
	UK Carbon Capture Technology			**Approved**	TT			
	Case: Coal Oxy SCPC with CO2 Capture			**Date**	26/04/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Steam Turbine Generator Package:	1100 Mwe					
1000	Steam Turbine						
	Lube Oil System	HP Admission: 2900 t/h @ 270 bar					
	Cooling System	Hot Reheat Sdmission;					
	Hydraulic Control System	2200 t/h @ 60 bar					
	Drainage System	LP admission: 2020 t/h @ 5.9 bar					
	Seals System						
	Electrical generator and auxiliaries						
	Steam Condenser Package:	1210 MWth					
1000	Steam Condenser						
	Hot Well						
	Vacuum pump (or ejectors)						
	Start-up ejector (if required)						
	Steam Turbine Bypass System:						
1000	MP dump tube						
	LP dump tube						
	HP/MP letdown station						
	MP letdown station						
	LP letdown station						
1000	Phosphate injection package						
1000	Oxygen scavenger injection package						
1000	Amines injection package						
	Sour Compression Section	923.146 t/h flue gas feed					
701	Two stage raw flue gas compressors	2 x 48 Mwe					
	Contacting column with liquid pump-around for sulphuric & nitric acid removal						
	Flue gas cooling downstream compressor						
	BFW Heater	14 MWth					
	Condensate Heater	26 MWth					

Notes:

amec foster wheeler	**EQUIPMENT LIST FOR PACKAGE EQUIPMENT**		**Rev.**	**REV 01**	**REV 02**	**REV 03**	
	Client: Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Originated**	SF			**SHEET** 8 **of** 10
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		**Checked**	RR			
	UK Carbon Capture Technology		**Approved**	TT			
	Case: Coal Oxy SCPC with CO2 Capture		**Date**	26/04/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Dehydration Package						
702	Dual Bed dessicant system						
	Cold Box for Inerts Removal						
703	Main heat exchangers						
	CO_2 Liquid separator						
	CO_2 distillation column						
	CO_2 compressors and coolers	2 x 15 MWe					
	Inerts heater						
	Inerts expander	16.6 MWe					
	Overhead recycle compressors	1.0 MWe					
	Cooling Water Filtration Package						
1100	Cooling Water sidestream filters	12300 m3/h					
	Sodium Hyperchloride Dosing Package						
1100	Sodium Hypochloride storage tank						
	Sodium Hypochloride dosage pumps						
	Antiscalant Package						
1100	Dispersant storage tank						
	Dispersant dosage pumps						
	Demineralised Water Package						
1100	Multimedia Filter						
	Reverse Osmosis Cartridge Filter						
	Electro-deionisation System						
1100	Condensate Polishing Package						
1100	Plant and Instrument Air Package						
1100	Emergency Diesel Generator System						
1100	Waste Water Treatment Unit						
1100	Auxiliary Boiler						

Notes:

amec foster wheeler	EQUIPMENT LIST FOR SOLIDS HANDLING EQUIPMENT	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	**Originated**	SF			**SHEET** 9 **of** 10
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	**Checked**	RR			
	UK Carbon Capture Technology	**Approved**	TT			
	Case: Coal Oxy SCPC with CO2 Capture	**Date**	26/04/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Coal Handling System:	325 t/h coal feed				30 Days storage:	
	Wagon tipper					Storage piles = 2 x 128,000 tonnes	
100	Receiving Hopper, vibratory feeder and belt extractor						
	Conveyors		Belt type				
	Transfer Towers		Enclosed type				
	As-Received Coal Sampling System		Two-stage type				
	As-Recived Magnetic separator system		Magnetic plate type				
	Crusher Towers		Impactor reduction				
	As-Fired Coal sampling system		Swing hammer type				
	As-Fired Magnetic separator system		Magnetic plate type				
	Coal silos	2 x 4900 m3				for daily storage	
	Filters						
	Fans						
	Limestone Handling System:	7.0 t/h limestone feed				30 Days storage:	
	Wagon tipper					Limestone Storage Volume = 4600 m3	
100	Receiving Hopper, vibratory feeder and belt extractor						
	Conveyors		Belt type				
	Transfer Towers		Enclosed type				
	Limestone sampling system		Swing hammer type				
	Separator system		Magnetic plate type				
	Limestone Mills						
	Limestone Silo	2 x 155 m3				for daily storage	
	Filters						
	Fans						
	Ash Handling System:	12.5 t/h bottom ash				14 Days Storage	
	Ash Storage Silos					Bottom ash storage volume = 6000 m3	
100	Ash Conveyors						
	Bottom ash crusher	29.2 t/h fly ash				14 Days Storage	
	Pneumatic conveying system					Fly ash storage volume = 14000 m3	
	Compressors						
	Filters						
	Fan						
	FGD By-Product Handling System:	13.3 t/h				30 Days storage:	
100	Storage Unit					Storage Volume = 7367 m3	
	Conveyors					1 operating, 1 spare	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	**Originated**	SF			**SHEET** 10 **of** 10
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	**Checked**	RR			
	UK Carbon Capture Technology	**Approved**	TT			
	Case: Coal Oxy SCPC with CO2 Capture	**Date**	26/04/2017			

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
CT-1101 A/B	Cooling Tower - Natural Draft		2 x 775 MWth Diameter: 120 m each Height: 210 m Water Inlet: 17 m			Concrete	Including Cooling Tower Basin	
900	Stack	Single stack for upset and start up				Concrete	For start-up and upset conditions only. 3415 tph flue gas @ 160 °C and 1.02 bara per train	

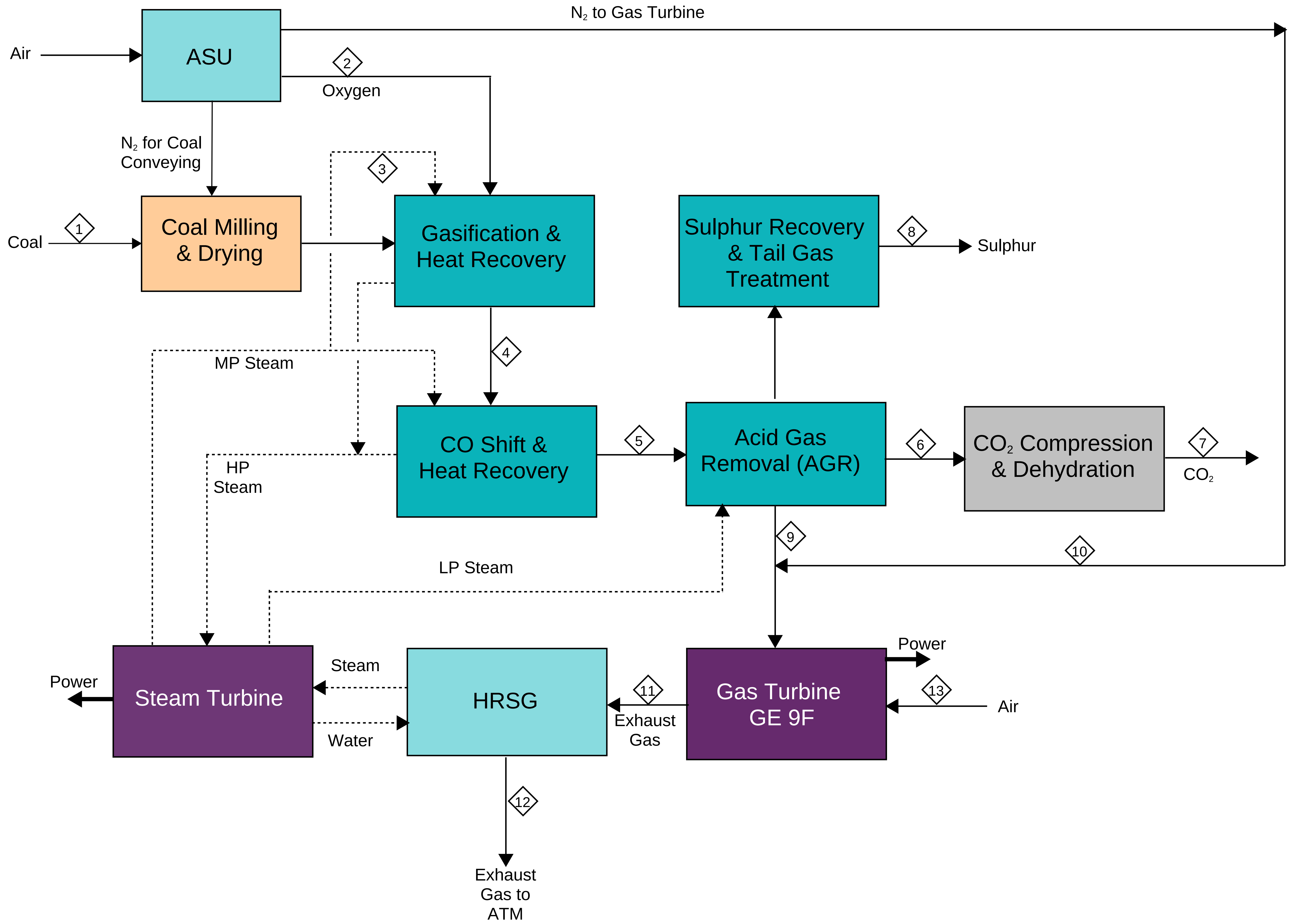
Notes:

Amec Foster Wheeler		**Prepared By :**	**K.D. Nelson**
Client : BEIS		**Base Date :**	**1Q2017**
Project : Novel Carbon Capture Technology Study		**Rev. No. :**	**'6'**
Contract No.: 13333		**Print Date :**	**14-Dec-17**
Case 4 : Oxy-combustion Capture for Power Generation on Coal			

Description	Case 4 : Oxy-combustion Capture for Power Generation on Coal							
	Unit 900	Unit 1000	Unit 2000	Unit 2100	Unit 3000	Unit 4000	Unit 6000	
	ASU	Feedstock & Solids Handling	Boiler Island	Flue Gas Desulphurisation	Steam Cycle	CPU	Utility Units	Total
	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP
Sub-Total Direct Materials	164,700,000	62,700,000	232,500,000	20,400,000	149,400,000	123,400,000	150,620,000	903,720,000
Other Material Costs								-
Shipping / Freight	8,235,000	3,135,000	11,625,000	1,020,000	7,470,000	6,170,000	7,531,000	45,186,000
Third Party Inspection	1,647,000	627,000	2,325,000	204,000	1,494,000	1,234,000	1,506,000	9,037,000
Spare Parts (Comm/2yrs Op)	3,294,000	1,254,000	4,650,000	408,000	2,988,000	2,468,000	3,012,000	18,074,000
Sub-Total Materials	177,876,000	67,716,000	251,100,000	22,032,000	161,352,000	133,272,000	162,669,000	976,017,000
Material & Labour Contracts								
Civils/Steelwork & Buildings	41,175,000	15,675,000	74,400,000	4,080,000	22,410,000	14,808,000	30,124,000	202,672,000
Sub-Total Material & Labour Contracts	41,175,000	15,675,000	74,400,000	4,080,000	22,410,000	14,808,000	30,124,000	202,672,000
Labour Only Contracts								
Mechanical	49,410,000	6,270,000	58,125,000	3,264,000	26,892,000	14,808,000	24,099,000	182,868,000
Electrical/Instrumentation	24,705,000	2,508,000	16,275,000	1,020,000	7,470,000	3,702,000	7,531,000	63,211,000
Scaffolding/Lagging/Rigging	8,894,000	1,053,000	8,928,000	514,000	4,123,000	2,221,000	3,796,000	29,529,000
Sub-Total Material & Labour Contracts	83,009,000	9,831,000	83,328,000	4,798,000	38,485,000	20,731,000	35,426,000	275,608,000
Sub-Total Materials & Labour	302,060,000	93,222,000	408,828,000	30,910,000	222,247,000	168,811,000	228,219,000	1,454,297,000
EPCm Costs								
Engineering Services/Construction Management	45,309,000	13,983,000	61,324,000	4,637,000	33,337,000	25,322,000	34,233,000	218,145,000
Commissioning	6,041,000	1,864,000	8,177,000	618,000	4,445,000	3,376,000	4,564,000	29,085,000
Sub-Total EPCm Costs	51,350,000	15,847,000	69,501,000	5,255,000	37,782,000	28,698,000	38,797,000	247,230,000
Total EPC Cost	353,410,000	109,069,000	478,329,000	36,165,000	260,029,000	197,509,000	267,016,000	1,701,527,000
Other Costs								
Pre-Licensing, Technical and Design etc	3,534,000	1,091,000	4,783,000	362,000	2,600,000	1,975,000	2,670,000	17,015,000
Regulatory, Licensing and Public Enquiry etc	7,311,000	2,256,000	9,896,000	748,000	5,379,000	4,086,000	5,524,000	35,200,000
Infrastructure Connection Costs							29,000,000	29,000,000
Owners Costs	24,739,000	7,635,000	33,483,000	2,532,000	18,202,000	13,826,000	18,691,000	119,108,000
Sub-Total Other Costs	35,584,000	10,982,000	48,162,000	3,642,000	26,181,000	19,887,000	55,885,000	200,323,000
Total Project Costs	**388,994,000**	**120,051,000**	**526,491,000**	**39,807,000**	**286,210,000**	**217,396,000**	**322,901,000**	**1,901,850,000**

ATTACHMENT 7: Case 5 – Integrated Gasification with Combined Cycle

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List
- Capital Cost Estimate Summary



NOTES

1. Two Trains of all equipments

Department for Business, Energy & Industrial Strategy

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REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	RR	SF	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram
Case 5 - Coal IGCC Case

Drawing No | Rev

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HEAT & MATERIAL BALANCE - Case 5 COAL IGCC

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**
CONTRACT:	13333	**DATE**	21/02/2017
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	R. Ray
		APP. BY	T. Tarrant

SHEET 1 OF 1

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13
Stream Name	B/L Coal	Oxygen to Gasifier	Steam to Gasifier	Syngas to Shift Reactor	Shifted Syngas to AGR	CO2 to Compression	Product CO2	Sulphur	H2-rich Gas to Gas Turbine	N2 to GT as Diluent	Flue Gas from Gas Turbine	Exhaust Gas	Air to Gas Turbine
Temperature (°C)	AMB	25.00	302.40	260.00	34.00	-	30.00	-	163.50	30.00	565.00	83.00	9.00
Pressure (bar abs)	ATM	49.50	44.00	38.50	34.20	-	110.00	-	29.90	33.40	1.04	ATM	9.00
Mass rate (kg/h)	314899	250286	21707	782849	868608	675993	674183	2692	176057	900000	6563993	6563993	5361392
Molar rate (kmol/h)		7784	1205	39003	43399	15655	15575	84	27338	32128	240392	240392	185795
Volume rate (m3/h)		3746	1155	44360	31579	44532	961	-	33493	25099	16111849	178179182	4650562
Molecular Weight		32.0	18.0	20.1	20.0	43.2	43.3	32.1	6.4	28.0	27.3	27.3	28.86
Component													
H2 (mol%)		0.00	0.00	19.78	54.73	1.67	1.68		85.93	0.00	0.00	0.00	0.00
CO (mol%)		0.00	0.00	41.90	0.69	0.02	0.02		1.08	0.00	0.00	0.00	0.00
CO2 (mol%)		0.00	0.00	1.18	38.63	97.80	98.21		4.22	0.00	0.59	0.59	0.03
Methane (mol%)		0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00	0.00	0.00	0.00
Ethane (mol%)		0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00	0.00	0.00	0.00
Propane (mol%)		0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00	0.00	0.00	0.00
n-Butane (mol%)		0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00	0.00	0.00	0.00
N2 (mol%)		1.50	0.00	5.60	5.03	0.10	0.10		7.93	100.00	74.32	74.32	77.31
Oxygen (mol%)		95.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00	10.92	10.92	20.74
Argon (mol%)		3.50	0.00	0.56	0.51	0.00	0.00		0.80	0.00	0.79	0.79	0.93
H2S (mol%)		0.00	0.00	0.20	0.00	0.00	0.00		0.00	0.00	0.00	0.00	0.00
COS (mol%)		0.00	0.00	0.02	0.00	0.00	0.00		0.00	0.00	0.00	0.00	0.00
Water (mol%)		0.00	100.00	30.75	0.21	0.42	0.00		0.03	0.00	13.38	13.38	0.99
Total		100	100	100	100	100	100		100.0	100.00	100.0	100.0	100.0

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Case 5 - Pre-Combustion Capture for Power Generation on Coal

CLIENT: Department for Business, Energy & Industrial Strategy
CONTRACT: 13333
NAME: Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

CHANGE	REV - O1				SHEET
DATE	03/03/2017				1 OF 1
ORIG. BY	R. Ray				
APP. BY	T. Tarrant				

DESCRIPTION	Electric Oper. Load	Steam (t/h)			Condensate (Note 2)	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
	KW	127 bar	47 bar	8 bar	LP (t/h)	t/h	t/h	t/h		
Syngas Production Units					265.0					
Coal Handling & Storage	-370									
Gasification Island	-25400		-22			-2720				
Syngas Cooling and Shift Reactors	-469		-479.2			-7400		208.8		
Acid Gas Recovery Unit	-20970					-7387				
Air Separation Unit	-106440			-265.0		-7763				
N2 Compression & Others	-39533					-3003				
Sulphur Recovery/Tailgas Treatment Unit	-580					-126				
CO2 Compression and Drying	-45673					-6455		0.7		
Process Heat Recovery Steam Production		480.2	408.9	33.5						
Process Units Total	-239435	480.2	-92.0	-231.5	265.0	-34854.2	0.0	209.5		
Power Island										
GT Gross Electrical Power	671027									
GT Auxiliaries	-2000									
HRSG & Auxiliaries	-10915	516.6	123.3	102.7			-640.7	4.2		
ST Electrical Power	391723	-996.9	-1028.2	-899.4	899.4	-41592.2				
Power Island Total	1049835	-480.2	92.0	231.5	899.4	-41592.2	-640.7	4.2		
Offsites & Utilities										
Fresh Cooling Water	-8518					76446.5				
Demin Plant/Condensate Plant	-933				-1048.0		640.7			
Utility water/Fire Water	-248									
Waste Water Treatment	-401				-116.4			-214		
Buildings and Misc loads	-500									
Offsites & Utilities Total	-10601	0.0	0.0	0.0	-1164.5	76446.5	640.7	-213.7		
Grand Total	**799800**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

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EQUIPMENT LIST FOR COMPRESSORS

Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333

Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Case: Coal IGCC with CO2 Capture

Rev.	REV 01	REV 02	REV 03
Originated	RR		
Checked	SF		
Approved	TT		
Date	13/03/2017		

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
700	K-001	CO2 Compressor Package (8 stages)	Multi-Stage Integrally Geared	2x50% (1/train)			1.287 / 4.925	78.9	1.1 / 80.0		0.992 / 0.297	22071	CrNi alloy	44.01	336.6 t/h CO2	
							/		/		/					
							/		/		/					
							/		/		/					
							/		/		/					
							/		/		/					

Notes:

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	Client:	Department for Business, Energy & Idustrial Strategy	Contract No: 13333	Originated	RR			SHEET 2 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		Checked	SF			
		UK Carbon Capture Technology		Approved	TT			
	Case:	Coal IGCC with CO2 Capture		Date	13/03/2017			

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	MATERIAL TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
400	E-401	CO Shift HP Steam Generator	Kettle	2x50% (1/train)			H:631 t/h C: 77.4 t/h	41.18	1443	355.0 / 134	516.0 / 42	SS316	SS316					
400	E-402	CO Shift MP Steam Generator 1	Kettle	2x50% (1/train)			H:631 t/h C: 41.2 t/h	23.48	1242	285.3 / 51	395.0 / 42	SS316	SS316					
400	E-403	CO Shift LP Steam Generator	Kettle	2x50% (1/train)			H:631 t/h C: 15.5 t/h	23.48	1259	195.0 / 10	325.0 42	SS316	SS316					
400	E-404	CO Shift MP Steam Generator 2	Kettle	2x50% (1/train)			H:631 t/h C: 24.1 t/h	13.73	1277	285.0 / 51	331.0 / 42	SS316	SS316					
400	E-405	CO Shift Gas-Gas exchanger	Shell & Tube	2x50% (1/train)			H:631 t/h C: 391.4 t/h	13.89	12592	285.0 / 43	294.0 / 42	1.25CR-0.5Mo	1.25CR-0.5Mo					
400	E-406	Satrator Heater	Shell & Tube	2x50% (1/train)			H:631 t/h C: 721 t/h	51.95	30164	225.0 / 46	252.0 / 42	CS	CS					
400	E-407	AGR Process Reboiler	Shell & Tube	2x50% (1/train)			H:569.2 t/h	64.65	9561	150.0 /	198.0 / 42	1.25CR-0.5Mo	1.25CR-0.5Mo					
400	E-408	N2 Heater	Shell & Tube	2x50% (1/train)			H:475.7 t/h C: 450 t/h	12.00	8557	144.0 / 35	162.5 / 42	SS316	SS316					
400	E-409	Syngas Heater	Shell & Tube	2x50% (1/train)			H:475.7 t/h C:88 t/h	9.08	9064	140.0 / 35	150.0 / 42	SS316	SS316					
400	E-410	Syngas Cooler	Shell & Tube	2x50% (1/train)			H:475.7 t/h C: 2666.4 t/h	34.02	4633	50.0 / 6	138.0 / 42	CS	CS					
400	E-411	Blowdown cooler	Shell & Tube	2x50% (1/train)			H:3.34 t/h C: 43.2 t/h	0.55	8	50.0 / 6	195.0 / 9.3	CS	CS					
400	E-412	Process Water cooler	Shell & Tube	2x50% (1/train)			H:91.7 t/h C: 990.3 t/h	12.64	215	50.0 / 6	168.0 / 37	CS	CS					
700	E-701	CO2 Compressor Cooler - Stage 1	Shell & Tube	2x50% (1/train)			H:74.3 t/h C: 32.8 t/h	0.42	123	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
700	E-702	CO2 Compressor Cooler - Stage 2	Shell & Tube	2x50% (1/train)			H:74.3 t/h C: 71.3 t/h	0.91	168	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
700	E-703	CO2 Compressor Cooler - Stage 3	Shell & Tube	2x50% (1/train)			H:260 t/h C: 112 t/h	1.43	342	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
700	E-704	CO2 Compressor Cooler - Stage 4	Shell & Tube	2x50% (1/train)			H:260 t/h C: 166 t/h	2.12	378	tubeside 50.0 / 10.0	160.0 / 70	SS304	SS304					
700	E-705	CO2 Compressor Cooler - Stage 5	Shell & Tube	2x50% (1/train)			H:375 t/h C: 470 t/h	5.90	695	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-706	CO2 Compressor Cooler - Stage 6	Shell & Tube	2x50% (1/train)			H:375 t/h C: 548 t/h	7.00	798	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-707	CO2 Compressor Cooler - Stage 7	Shell & Tube	2x50% (1/train)			H:337 t/h C: 340 t/h	4.33	659	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-708	CO2 Compressor Cooler - Stage 8	Shell & Tube	2x50% (1/train)			H:337 t/h C: 1094 t/h	13.96	1948	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-709	CO2 Pump Cooler	Shell & Tube	2x50% (1/train)			H:337 t/h C: 391 t/h	5.12	1233	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
1000	E-1001	Steam Condenser (Water Cooled)	Shell & Tube	2x50% (1/train)			H:456.3 t/h C: 21600 t/h	275.34	22668	tubeside 50.0 / 10.0	100.0 / 5	SS304	SS304					

Notes: Notes:

1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

EQUIPMENT LIST FOR PUMPS

Rev.	REV 01	REV 02	REV 03	SHEET 3 of 9
Originated	RR			
Checked	SF			
Approved	TT			
Date	13/03/2017			

Client Department for Business, Energy & Idustrial Strategy **Contract No:** 13333

Descriptior Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Case: Coal IGCC with CO2 Capture

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP / SG / VISC'Y °C		cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
1000		Vacumn Condensate return pump	Centrifugal	2/train		450	452	75	12		31.0	1	0.200	Design Temp: 80/-10 Design Press.: 14/FV	200	CS	One operating; One spare	
1100		Process Cooling Water Pumps (closed loop)	Centrifugal	3/train	2x50% duty, 1 spare	8714	8739	75	3		25.0	1	0.900	Design Temp: 80/-10 Design Press.:9.5	971	CS	Two operating; One spare Per train	
1100		Steam Condenser Cooling Water Pumps	Centrifugal	3/train	2x50% duty, 1 spare	10398	10428	75	3		25.0	1	0.900	Design Temp: 80/-10 Design Press.:9.5	1159	CS	Two operating; One spare Per train	
1100		Process Water Pump	Centrifugal	1/train		113	126	75	4		170.0	1	0.900	Design Temp: 200/-10 Design Press.:15	19	CS	Baseplate length = 1.401m x width = 0.4924m	
1100		Demin Water Pump	Centrifugal	2/train	2x50% duty, 1 spare	320	321	75	7		20.0	1	1.000	Design Temp- 80 / -10 Design Pressure: 5.20	84	Stainless Steel casing and impeller		
1100		Firewater Pump Package	Centrifugal	3							20.0			Design Temp- 80 / -10 Design Pressure: 19		CS	Sizing: 8 x 4 x 3.2 1 x diesel, 1 x electric and	
1100		Cooling Tower Makeup Pump	Centrifugal	2			1465	75							185		One operating; One spare; 30 m head	
1100		Raw Water Pump Pump	Centrifugal	2			555	75							110		One operating; One spare; 50 m head	
900		HP BFW Pump	Centrifugal	2/train	Centrifugal	567	632	75	127.00		170	0.9	0.2		2980	CS	One operating; One spare	
900		MP BFW Pump	Centrifugal	2/train	Centrifugal	269	299	75	39.00		170	0.9	0.2		434	CS	One operating; One spare	
900		LP BFW Pump	Centrifugal	2/train	Centrifugal	155	173	75	2.00		170	0.9	0.2		75	CS	One operating; One spare	
700		CO2 Pump	Centrifugal	1/train	Centrifugal	337		75	30.00		30	0.5	0.04		766	SS316L	Liquid CO2 flowrate : 337 t/h	
400		Saturator Circulating Water Pump	Centrifugal	1/train	Centrifugal	721	791.8	75	5.00		145	0.9	0.2		121	CS		

Notes:

EQUIPMENT LIST FOR VESSELS		**Rev.**	**REV 01**	**REV 02**	**REV 03**
Client:	Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	**Originated**	RR		
Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**Checked**	SF		
		Approved	TT		
Case:	Coal IGCC with CO2 Capture	**Date**	13/03/2017		

EQUIPMENT NUMBER		DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	DIMENSIONS HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	DESIGN CONDITIONS PRESS barg	DESIGN CONDITIONS VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	MATERIALS OF CONST'N INTERNALS MAT./LINING/ CA	REMARKS	REV
400	V-401	Condensate Seperator	Vertical drum	2x50% (1/train)	3.10	6.20	54.59	V	80	40	1.013	Wire Mesh Pad 0.75 100	CS	CS		
400	V-402	Condensate Seperator	Vertical drum	2x50% (1/train)	4.40	8.80	156.11	V	80	40	1.013	Wire Mesh Pad 1.52 100	CS	CS		
400	V-403	Condensate Seperator	Vertical drum	2x50% (1/train)	3.90	7.80	108.71	V	80	40	1.013	Wire Mesh Pad 1.19 100	CS	CS		
401	V-404	Condensate Seperator	Vertical drum	2x50% (1/train)	3.40	6.80	72.03	V	80	40	1.013	Wire Mesh Pad 0.91 100	CS	CS		
400	V-405	Condensate Accumulator	Drum	2x50% (1/train)	1.00	2.00	1.83	H	80	40	1.013	Wire Mesh Pad 0.08 100	CS	CS		
401	V-406	Blowdown Drum	Vertical drum	2x50% (1/train)	1.10	2.20	2.44	V	80	10	1.013	Wire Mesh Pad 0.10 100	CS	CS		
400	T-400	Syngas Saturator	Vertical Column	2x50% (1/train)	3.50	7.00	78.57	V	80	40	1.013	Wire Mesh Pad 0.96 100	CS	CS		
400	R-401	1st Shift Catalyst Reactor	Reactor	2x50% (1/train)												
400	R-402	2nd Shift Catalyst Reactor	Reactor	2x50% (1/train)												
400	R-403	3rd Shift Catalyst Reactor	Reactor	2x50% (1/train)												
700	V-701	CO2 Compressor Stage 1 KO drum	Vertical drum	2x50% (1/train)	3.70	7.40	93	V	80	4.19	1.013	Wire Mesh Pad 1.08 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-702	CO2 Compressor Stage 2 KO drum	Vertical drum	2x50% (1/train)	3.70	7.40	93	V	80	6.19	1.013	Wire Mesh Pad 1.08 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-703	CO2 Compressor Stage 3 KO drum	Vertical drum	2x50% (1/train)	3.50	4.80	57	V	80	8.69	1.013	Wire Mesh Pad 0.96 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-704	CO2 Compressor Stage 4 KO drum	Vertical drum	2x50% (1/train)	3.02	4.00	36	V	80	17.39	1.013	Wire Mesh Pad 0.72 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-705	CO2 Compressor Stage 5 KO drum	Vertical drum	2x50% (1/train)	2.50	3.20	20	V	80	34.69	1.013	Wire Mesh Pad 0.49 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		

Notes:

1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

amec foster wheeler	**EQUIPMENT LIST FOR TANKS**			
	Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology		
	Case:	Coal IGCC with CO2 Capture		

Rev.	**REV 01**	**REV 02**	**REV 03**
Originated	SF		
Checked	RR		
Approved	TT		
Date	03/02/2017		

UNIT NUMBER	**DESCRIPTION**	**NO REQ'D**	**DIMENSIONS** ID m	HEIGHT m	**TOTAL VOLUME** m^3	**ROOF TYPE**	**BOTTOM TYPE**	**STORAGE** PRESS Bar	TEMP	**SG**	**CORRO-SION**	**'MATERIALS**	**MANWAYS** SHELL	ROOF	**CONNECT-IONS**	**HEATING COIL**	**INSUL-ATION**	**REMARKS**	**REV**
1100	Raw Water tank	1	36.0	14.4	13320	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 24 hr Storage	
1100	Demin Water Tank	1	37.1	14.8	13370	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 25 hr Storage	
1100	Firewater Storage Tank	1	13.0	7.8	1035	Storage tank		Atm	20			Lined CS						Design Pres.: 0.0075 / -0.0025 Design Temp: 80/-10 deg C	

Notes:

	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 6 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Coal IGCC with CO2 Capture			Date	13/03/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
200	Coal Milling & Drying (4 x 33%)	4 x 2500 t/d raw coal, As Received basis Water content after dryer outlet : 1.3%				3 operating, one spare	
200	**Shell Coal Gasification Package (2 x 50%)** Coal pressurization & feeding Shell Gasifiers Syngas coolers Slag removal system Dry Fly Ash removal system Wet Scrubbing Primary water treatment Nitrogen + Blowback systems Flare headers and fuel distribution systems cooling water systems Process water systems Steam/Condensate systema Plant/Instrument air systems NaOH/HCl distribution systems	2 x 3780 t/d coal to Gasifier, As Received basis				Two Gasification package	
300	Air Separation Unit (ASU) (2 x 50%)	2 x 125 t/h 95% pure O2	Cryogenic	53.2 MW each		Two ASU package	
300	N2 Compressors package (2 x 50%)	Multi-Stage with intercoolers 2 x 450 t/hr of N2 360057.82 Nm3/hr 12549.61 Am3/hr	Electric motor driven, Centrifugal	16.62 MW for each trains		Two packages Outlet Pressure : 32 bara	
300	LOX (Liquid Oxygen) Storage Tank with Vaporiser	1 x 100% (1 for both trains) 953 Am3 of O2 1001 t of O2	Fixed Roof Storage Tank			To enhance the ASU Reliability Operating Pressure : 5 bara Operating Temp: - 165 oC 8 hour of storage for 1 Gasification Train	
300	LIN (Liquid Nitrogen) Storage Tank with Vaporiser	1 x 100% (1 for both trains) 333 Am3 of N2 243 t of N2	Fixed Roof Storage Tank			Operating Pressure : 5 bara Operating Temp: - 180 oC 8 hour of storage for 1 Gasification Train & 4 min of N2 for GT	
500	Acid Gas Removal Unit (Selexol)	2 x 50% Feed gas : 437110 Nm3/hr per train Operating Pressure: 37 barg	Selexol Process	10.5 MW Each Train		CO2 & H2S removal Total CO2 removal: 8085 t/d; Total Carbon Capture : 90%	
600	Sulphur Recovery Unit (SRU), Tail Gas Treatment & Sour Water Package	2 x 50% Sulphur Product : 32.3 t/d per train Feed gas : 6000 Nm3/hr per train Treated Tail gas : 1520 Nm3/hr per train				Sulphur Content >99.9 mole% min (dry basis)	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT		Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy Contract No: 13333	Originated	RR			SHEET 7 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	Checked	SF			
		UK Carbon Capture Technology	Approved	TT			
	Case:	Coal IGCC with CO2 Capture	Date	13/03/2017			

UNIT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
800	Gas Turbine & Generator Package	GE 9F Syngas Variant Gas Turbine	2 x 50% (1/train)								336 MW Output Turbine generator				
900	HRSG		2 x 50% (1/train)											Horizontal, Natural Draft 3 Pressure Level	
900	Phosphate Injection		2 x 50% - 1/Train												
900	Oxygen Scavanger Injection Package		2 x 50% - 1/Train												
1000	Steam Turbine & Generator Package		2 x 50% (1/train)								200 MW Output Turbine generator				
1100	Cooling Tower	Evaporative, Natural Drive Cooling Tower									Total Heat duty 950 MWth			Diameter: 145 m Height: 210 m	
1100	Cooling Tower packages						7400							Filtration Package; Hypochlorite Dosing Package; Antiscalant Package	
1100	Nitrogen Package	Tank/pump	1						7	Design Temp: 80/-10 deg C Design Press.: 14 barg Oper. Temp.: 20 deg C		CS			
1100	Compressed Air Package	Compressor	1							Design Temp- 80 / -10 deg C Design Pressure: 8.70 barg					
700	TEG Dehydration Package		2 x 50% (1/train)							Design Temp: 150/-10 deg C Design Press.: 76 barg				Each package dehydrates 375 t/h CO2, removes 341 kg/h water to get to spec of 500 ppm	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR SOLIDS HANDLING EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 8 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Coal IGCC with CO2 Capture			Date	13/03/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
100	**Coal Handling System:**	315 t/h coal feed				30 Days storage:	
	Wagon tipper					Storage piles = 2 x 124,000 tonnes	
	Receiving Hopper, vibratory feeder and belt extractor						
	Conveyors		Belt type				
	Transfer Towers		Enclosed type				
	As-Received Coal Sampling System		Two-stage type				
	As-Recived Magnetic separator system		Magnetic plate type				
	Crusher Towers		Impactor reduction				
	As-Fired Coal sampling system		Swing hammer type				
	As-Fired Magnetic separator system		Magnetic plate type				
	Coal silos	2 x 4800 m3				for daily storage	
	Filters						
	Fan						
100	**Limestone Handling System:**	Limestone feed, t/hr				30 Days storage:	
	Wagon tipper	9.1				Limestone Storage Volume = 6000 m3	
	Receiving Hopper, vibratory feeder and belt extractor						
	Conveyors		Belt type				
	Transfer Towers		Enclosed type				
	Limestone sampling system		Swing hammer type				
	Separator system		Magnetic plate type				
	Limestone Mills						
	Limestone Silo					for daily storage	
	Filters	1 x 200 m3					
	Fan						
1100	Sulphur Storage & Handling	65 t/h Sulphur storage				30 days storage	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 9 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Coal IGCC with CO2 Capture			Date	13/03/2017			

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
400	Mercury Adsorber			Sulfur-impregnated activated carbon beds				
900	Stack	2 x 50% - 1/Train					3476.2 tph flue gas @ 84 oC and 1.02 bara per trin	

Notes:

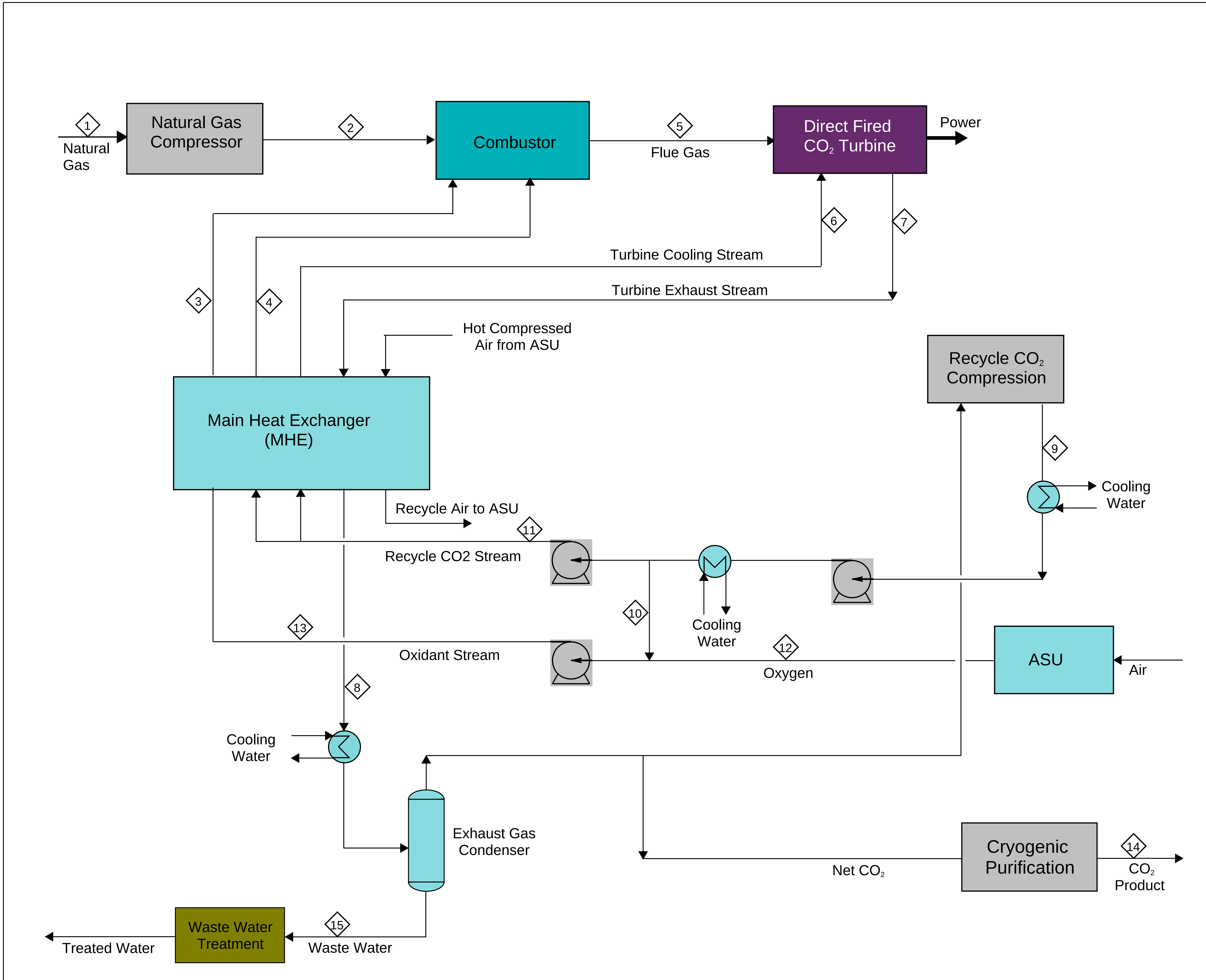
Amec Foster Wheeler
Client : BEIS
Project : Novel Carbon Capture Technology Study
Contract No.: 13333
Case 5 : IGCC Pre-combustion Capture for Power Generation on Coal

Prepared By : K.D. Nelson
Base Date : 1Q2017
Rev. No. : '6'
Print Date : 14-Dec-17

Description	Case 5 : IGCC Pre-combustion Capture for Power Generation on Coal									
	Unit 100	Unit 200	Unit 300	Unit 400	Unit 500	Unit 600	Unit 700	Unit 800-1000	Unit 1100	
	Solids Handling	Gasification Island	ASU	Syngas Treatment & Sour Water System	AGR	SRU & TGTU	CO_2 Compression Block	Combined Cycle Block	Utility Units	Total
	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP
Sub-Total Direct Materials	40,900,000	273,900,000	112,028,800	69,090,600	51,500,000	39,600,000	38,536,200	334,607,000	192,030,000	1,152,192,600
Other Material Costs										
Shipping / Freight	2,045,000	13,695,000	5,601,000	3,455,000	2,575,000	1,980,000	1,927,000	16,730,000	9,602,000	57,610,000
Third Party Inspection	409,000	2,739,000	1,120,000	691,000	515,000	396,000	385,000	3,346,000	1,920,000	11,521,000
Spare Parts (Comm/2yrs Op)	818,000	5,478,000	2,241,000	1,382,000	1,030,000	792,000	771,000	6,692,000	3,841,000	23,045,000
Sub-Total Materials	44,172,000	295,812,000	120,990,800	74,618,600	55,620,000	42,768,000	41,619,200	361,375,000	207,393,000	1,244,368,600
Material & Labour Contracts										
Civils/Steelwork & Buildings	10,225,000	87,648,000	28,007,000	10,364,000	6,180,000	9,900,000	5,780,000	50,191,000	38,406,000	246,701,000
Sub-Total Material & Labour Contracts	10,225,000	87,648,000	28,007,000	10,364,000	6,180,000	9,900,000	5,780,000	50,191,000	38,406,000	246,701,000
Labour Only Contracts										
Mechanical	4,090,000	68,475,000	33,609,000	12,436,000	6,180,000	12,672,000	6,937,000	60,229,000	30,725,000	235,353,000
Electrical/Instrumentation	1,636,000	19,173,000	16,804,000	3,455,000	1,545,000	3,960,000	1,927,000	16,730,000	9,602,000	74,832,000
Scaffolding/Lagging/Rigging	687,000	10,518,000	6,050,000	1,907,000	927,000	1,996,000	1,064,000	9,235,000	4,839,000	37,223,000
Sub-Total Labour Only Contracts	6,413,000	98,166,000	56,463,000	17,798,000	8,652,000	18,628,000	9,928,000	86,194,000	45,166,000	347,408,000
Sub-Total Materials & Labour	60,810,000	481,626,000	205,460,800	102,780,600	70,452,000	71,296,000	57,327,200	497,760,000	290,965,000	1,838,477,600
EPCm Costs										
Engineering Services/Construction Management	9,122,000	72,244,000	30,819,000	15,417,000	10,568,000	10,694,000	8,599,000	74,664,000	43,645,000	275,772,000
Commissioning	1,216,000	9,633,000	4,109,000	2,056,000	1,409,000	1,426,000	1,147,000	9,955,000	5,819,000	36,770,000
Sub-Total EPCm Costs	10,338,000	81,877,000	34,928,000	17,473,000	11,977,000	12,120,000	9,746,000	84,619,000	49,464,000	312,542,000
Total EPC Cost	71,148,000	563,503,000	240,388,800	120,253,600	82,429,000	83,416,000	67,073,200	582,379,000	340,429,000	2,151,019,600
Other Costs										
Pre-Licensing, Technical and Design etc	711,000	5,635,000	2,404,000	1,203,000	824,000	834,000	671,000	5,824,000	3,404,000	21,510,000
Regulatory, Licensing and Public Enquiry etc	1,462,000	11,580,000	4,940,000	2,471,000	1,694,000	1,714,000	1,378,000	11,968,000	6,997,000	44,204,000
Infrastructure Connection Costs									29,000,000	29,000,000
Owners Costs	4,980,000	39,445,000	16,827,000	8,418,000	5,770,000	5,839,000	4,695,000	40,767,000	23,830,000	150,571,000
Sub-Total Other Costs	7,153,000	56,660,000	24,171,000	12,092,000	8,288,000	8,387,000	6,744,000	58,559,000	63,231,000	245,285,000
Total Project Costs	**78,301,000**	**620,163,000**	**264,559,800**	**132,345,600**	**90,717,000**	**91,803,000**	**73,817,200**	**640,938,000**	**403,660,000**	**2,396,304,600**

ATTACHMENT 8: Case 6 – Oxy-fired Supercritical Gas Turbine

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List
- Capital Cost Estimate Summary



NOTES

1. Turbine/Generator and CO2 Compressor on same shaft
2. 3 x 500 MWth Trains of all equipments except CPU
3. One CPU common to three Trains

Department for Business, Energy & Industrial Strategy

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REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	RR	SF	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram
Case 6 - Oxy-combustion (Allam Cycle)

Drawing No
Rev

amec foster wheeler

HEAT & MATERIAL BALANCE - Case 6 (1500 MW_{TH}) ALLAM CYCLE

CLIENT:	OGCI	**CHANGE**	**REV - O1**						**SHEET**
CONTRACT:	13279	**DATE**	06/03/2017						1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation)	**ORIG. BY**	R. Ray						
		APP. BY	T. Tarrant						

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Stream Name	Natural Gas from BL	Natural Gas to Combustor	Oxidant to Combustor	Recycle CO2 to Combustor	Flue Gas to Turbine	Turbine Cooling Gas	Turbine Exhaust Gas to MHE	Turbine Exhaust Gas to Condenser	Compressed Recycle CO2 Stream	Stream to Oxygen Mixer	Recycle CO2 Stream to MHE	Oxygen from ASU	Oxidant to MHE	CO2 to BL	Waste Water
Pressure (bar abs)	70.0	305.0	303.0	303.0	300.0	303.0	34.0	33.0	80.0	120.0	305.0	120.0	305.0	110.0	2.5
Temperature (°C)	9.0	145.0	720.0	720.0	1150.0	<400	740.0	55.0	43.0	26.0	50.0	9.0	45.0	30.0	25.0
Mass rate (kg/h)	118940	118940	4412340	4574360	9105640	1043500	10149140	10149140	9587420	3969560	5617860	442780	4412340	284020	241200
Molar rate (kmol/h)	6600	6600	104600	104600	216110	23860	239970	239970	219230	90770	128460	13830	104600	12910	26780
Component															
Nitrogen (mol%)	0.89	0.89	1.05	1.18	1.11	1.18	1.12	1.12	1.18	1.18	1.18	0.20	1.05	0.16	0.00
CO2 (mol%)	2.00	2.00	84.94	97.88	91.80	97.88	92.41	92.41	97.88	97.88	97.88	0.00	84.94	99.83	0.00
Methane (mol%)	89.00	89.00	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
Ethane (mol%)	7.00	7.00	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
Propane (mol%)	1.00	1.00	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
n-Butane (mol%)	0.10	0.10	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
n-Pentane (mol%)	0.01	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
Oxygen (mol%)	0.00	0.00	13.34	0.21	0.20	0.21	0.20	0.20	0.21	0.21	0.21	99.50	13.34	0.01	0.00
Argon (mol%)	0.00	0.00	0.53	0.57	0.53	0.57	0.54	0.54	0.57	0.57	0.57	0.30	0.53	0.00	0.00
CO (mol%)	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
Water (mol%)	0.00	0.00	0.13	0.15	6.36	0.15	5.74	5.74	0.15	0.15	0.15	0.00	0.13	0.00	100.00
Total	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0

amec foster wheeler

CASE 6 -UTILITIES BALANCE SUMMARY - Oxy-combustion (Allam Cycle)

CLIENT:	Department for Business, Energy & Idustrial Strategy	**CHANGE**	**REV - O1**					**SHEET**
CONTRACT:	13333	**DATE**	23/03/2017					1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	R. Ray					
		APP. BY	T Tarrant					

UNIT No.	**DESCRIPTION**	**Electric Oper. Load**	**Fuel Gas LHV**	**Steam (t/h)**			**Condensate (t/h)**			**Cooling water**	**Raw Water**	**REV**
		MW	**MW**	**HPS 171 barg**	**MPS 40 barg**	**LPS 4.8 barg**	**HP T/h**	**MP T/h**	**LP T/h**	**T/h**	**T/h**	
Power Island												
	GT Gross Electrical Power	1263.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	GT Auxiliaries	-5.3	-1536.3	0.0	0.0	0.0	0.0	0.0	0.0	-4360.0	0.0	
	Natural Gas Compression	-8.6										
	ST Gross Electrical Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	ST Auxiliaries (inc BFW pumps, condenser)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Power Island Net	1250.0	-1536.3							-4360.0		
Air Separation Unit (ASU)												
	Main Air Compressors	-150.0								-2800.0		
	Booster Air Compressor & other load	-20.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-1440.0	0.0	
	ASU Net	-170.9								-4240.0		
Recycle Gas Compression												
	Recycle Gas Compression upto 80 bara	-109.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
	Recycle Gas + Coolant Compression upto 305 bara	-62.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-60120.0	0.0	
	O2 + Recycle Gas Compression upto 305 bara	-37.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
	CO2 Capture Net	-208.0								-60120.0		
CO2 Purification Unit (CPU)												
	Autorefrigerated Purification Unit compressor load	-14.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-2380.0	0.0	
	Autorefrigerated Purification Unit expander production	1.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	CPU Net	-12.4								-2380.0		
Utilities & Offistes Systems												
	Cooling Water System	-8.9									-1280	
	Balance of Plant including WWT	-1.5									230	
	U&O Net	-10.4								-71100.0	-1050.0	
Overall Performance												
	Gross Production	1263.9	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Total Parasitic Loads	-415.5	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Net Produced	848.4	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Net LHV Efficiency	55.2%	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	

amec foster wheeler	EQUIPMENT LIST FOR COMPRESSORS		Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	**Originated**	RR			**SHEET** 1 **of** 6
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	**Checked**	SF			
	Unit No:	Allam Cycle	**Approved**	TT			
	Case:	6	**Date**	22/03/2017			

EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
	Natural Gas Compressor	Compressor	3x 33%	Electrical	591	1.690 / 1.500	235.0	70.0 / 305.0		0.8000 / 0.900	2867	SS316L	18.02		
	Recycle CO2 Compressor	Compressor	3x 33%	Electrical	45,582	1.610 / 3.799	47.0	33.0 / 80.0		0.813 / 0.540	37333	SS316L	44.01	4 stage compressor with intercoolers	
	Direct Fired CO2 Turbine	Turbine	3x 33%	Electrical	30,245	1.175 / 1.189	266.0	300.0 / 34.0		1.063 / 1.007	423333	Nickel based super alloy	42.10		

amec foster wheeler	EQUIPMENT LIST FOR HEAT EXCHANGERS	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Idustrial Strategy Contract No: 13333	Originated	RR			SHEET 2 of 6
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	Checked	SF			
	Unit No: Allam Cycle	Approved	TT			
	Case: 6	Date	22/03/2017			

EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	MATERIAL TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
	Exhaust Gas Cooler	HE	3x 33%				33.00	9358	Tube side 50 / 8	Shell side 80 / 35		SS316L					
	Main Heat Exchanger	HE	3x 33%				836.0		50.0 / 320	765 / 35		Cobalt or Nickel based super alloy				Compact multi-channel plate-fin type	
	Cooler after Recycle CO2 Compressor	HE	3x 33%				118.00	48415	Shell side 50 / 8	Tube side 60 / 84		SS316L					
	Cooler after Recycle CO2 Pump	HE	3x 33%				28.40	2084	Tube side 50 / 8	Shell side 52 / 125		SS316L					

Notes:
1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler	EQUIPMENT LIST FOR PUMPS			Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No: 13333	Originated	RR			SHEET 3 of 6
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		Checked	SF			
	Unit No:	Allam Cycle		Approved	TT			
	Case:	6		Date	22/03/2017			

EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP / SG / VISC'Y °C cP	DESIGN TEMPERATURE/PRESSURE °C/barg	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
	Recycle Pump upto 120 bar	Centrifugal	3x 33%	Electrical	3196 t/hr x 1.1	4850 m3/hr x 1.1	85	40.00		'35 / 0.7 / 0.05	Design Temp: 60 Design Press.: 125	6200	SS316L		
	Recycle Pump upto 305 bar	Centrifugal	3x 33%	Electrical	1873 t/hr x 1.1	2404 m3/hr x 1.1	85	185.00		'50 / 0.8 / 0.06	Design Temp: 75 Design Press.: 320	14300	SS316L		
	Oxidant Pump	Centrifugal	3x 33%	Electrical	1471 t/hr x 1.1	2095 m3/hr x 1.1	85	185.00		25 / 0.7 / 0.05	Design Temp: 50 Design Press.: 320	0	SS316L		

amec foster wheeler	EQUIPMENT LIST FOR VESSELS				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 4 of 6
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
	Unit No:	Allam Cycle			Approved	TT			
	Case:	6			Date	22/03/2017			

EQUIPMENT NUMBER	DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	PRESS barg	VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	INTERNALS MAT./LINING/ CA	REMARKS	REV
	Flue Gas Water Separator	V	3x 33%	13.9	27.8	4922	V	50	3.20	N/A		SS316L			

Notes:
1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

amec foster wheeler	EQUIPMENT LIST FOR TANKS				REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13279	RR			SHEET 5 of 6
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			SF			
	Unit No:	Allam Cycle			TT			
	Case:	6			22/03/2017			

EQUIPMENT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS		TOTAL VOLUME	ROOF TYPE	BOTTOM TYPE	STORAGE		SG	CORRO-SION	'MATERIALS	MANWAYS		CONNECT-IONS	HEATING COIL	INSUL-ATION	REMARKS	REV
			ID m	HEIGHT m	m^3			PRESS	TEMP				SHELL	ROOF					
	Firewater Storage Tank	1	13.0	7.8		Tank			20			Lined CS						Design Pres.: 0.0075 / - 0.0025	

Notes:

	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	Client:	Department for Business, Energy & Idustrial Stra	Contract No:	13333	Originated	RR			SHEET 6 of 6
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
	Unit No:	Allam Cycle			Approved	TT			
	Case:	6			Date	22/03/2017			

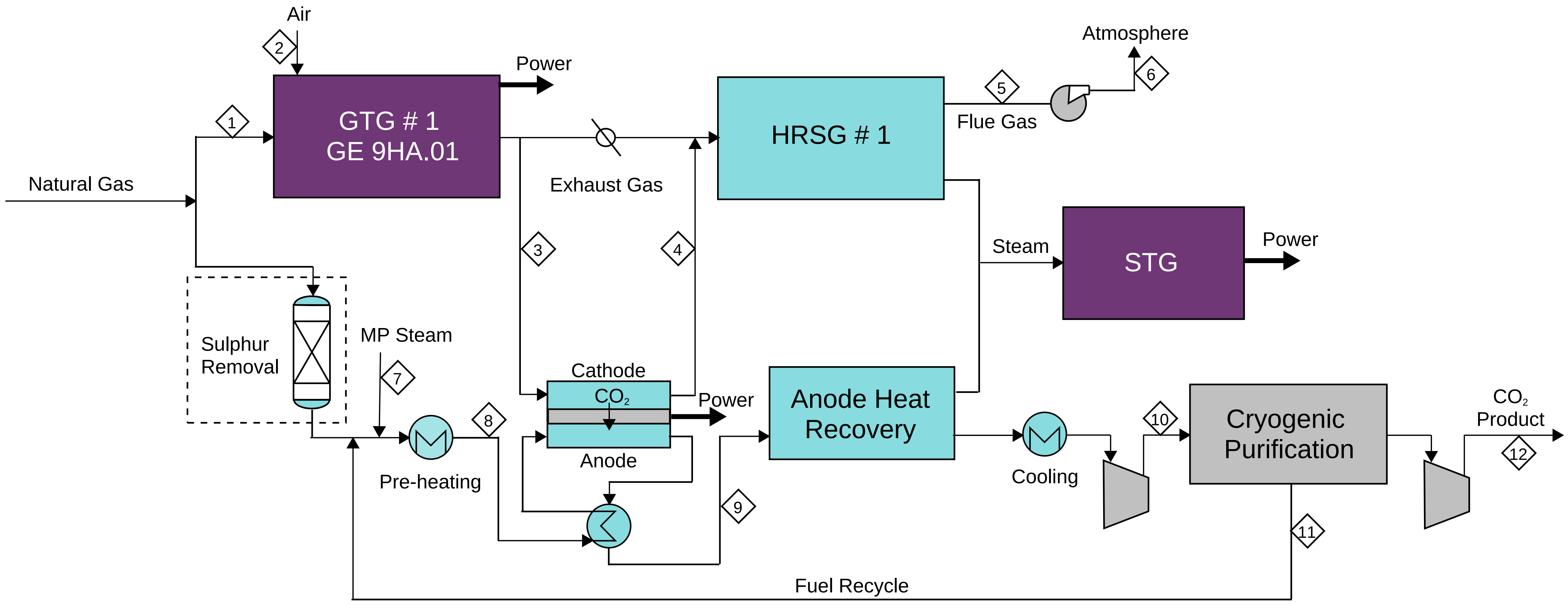
EQUIPMENT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
	CPU		1x 100%				4525	319860						Cryogenic CO2 removal Unit; Common to 3 trains	
	ASU		3x 33%					147593						3542 TPD of O2; Purity: 99.5 mole% O2 Pressure: 120 bar	
	Combustor		3x 33%					3035213	Operating Pressure 300 bar		513333	Cobalt or Nickel based super alloy			
	Waste Water Treatment		1x 100%					80400						Common to 3 trains	
	LOX Storage Tank		1x 100%											700t (Toenhance the ASU Reliability)	
	Naturak Gas Flow metering and analyser package	Metering	1x 100%					118941		Design Temp: 80 deg C Design Press.: 114 barg				Fiscal metering package	
										Design Temp: 80/-10 deg					

Amec Foster Wheeler		**Prepared By :**	**K.D. Nelson**
Client : BEIS		**Base Date :**	**1Q2017**
Project : Novel Carbon Capture Technology Study		**Rev. No. :**	**'6'**
Contract No.: 13333		**Print Date :**	**14-Dec-17**
Case 6 : Oxy-fired Supercritical Power Generation on Gas			

Description	Case 6 : Oxy-fired Supercritical Power Generation on Gas (Allam Cycle)				
	Power & CO_2 Cycle	CO_2 Purification Block	ASU	Utility Units	Total
	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP
Sub-Total Direct Materials	325,063,000	28,625,000	132,495,000	97,000,000	583,183,000
Other Material Costs					
Shipping / Freight	16,253,000	1,431,000	6,625,000	4,850,000	29,159,000
Third Party Inspection	3,251,000	286,000	1,325,000	970,000	5,832,000
Spare Parts (Comm/2yrs Op)	6,501,000	573,000	2,650,000	1,940,000	11,664,000
Sub-Total Materials	351,068,000	30,915,000	143,095,000	104,760,000	629,838,000
Material & Labour Contracts					
Civils/Steelwork & Buildings	48,759,000	3,435,000	33,124,000	19,400,000	104,718,000
Sub-Total Material & Labour Contracts	48,759,000	3,435,000	33,124,000	19,400,000	104,718,000
Labour Only Contracts					
Mechanical	58,511,000	3,435,000	39,749,000	15,520,000	117,215,000
Electrical/Instrumentation	16,253,000	859,000	19,874,000	4,850,000	41,836,000
Scaffolding/Lagging/Rigging	8,972,000	515,000	7,155,000	2,444,000	19,086,000
Sub-Total Labour Only Contracts	83,736,000	4,809,000	66,778,000	22,814,000	178,137,000
Sub-Total Materials & Labour	483,563,000	39,159,000	242,997,000	146,974,000	912,693,000
EPCm Costs					
Engineering Services/Construction Management	72,534,000	5,874,000	36,450,000	22,046,000	136,904,000
Commissioning	9,671,000	783,000	4,860,000	2,939,000	18,253,000
Sub-Total EPCm Costs	82,205,000	6,657,000	41,310,000	24,985,000	155,157,000
Total EPC Cost	565,768,000	45,816,000	284,307,000	171,959,000	1,067,850,000
Other Costs					
Pre-Licensing, Technical and Design etc	5,658,000	458,000	2,843,000	1,720,000	10,679,000
Regulatory, Licensing and Public Enquiry etc	12,133,000	983,000	6,097,000	3,687,000	22,900,000
Infrastructure Connection Costs				37,000,000	37,000,000
Owners Costs	39,604,000	3,207,000	19,901,000	12,037,000	74,749,000
Sub-Total Other Costs	57,395,000	4,648,000	28,841,000	54,444,000	145,328,000
Total Project Costs	**623,163,000**	**50,464,000**	**313,148,000**	**226,403,000**	**1,213,178,000**

ATTACHMENT 9: Case 7 – CCGT with Molten Carbonate Fuel Cells

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List
- Capital Cost Estimate Summary



NOTES

1. Two Trains of all equipments

Department for Business, Energy & Industrial Strategy

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REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	SF	RR	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram
Case 7: Molten Carbonate Fuel Cell Case

Drawing No

Rev

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HEAT & MATERIAL BALANCE - Case 7 - Molten Carbonate Fuel Cell Capture for Power Generation on Gas

CLIENT:	Department for Business, Energy & Industrial Strategy					**CHANGE**	**REV - O1**	**REV - O2**		**SHEET**		
CONTRACT:	13333					**DATE**	12/05/2017	15/08/2017		1	OF	1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel					**ORIG. BY**	S. Ferguson	S. Ferguson				
CASE:	(Next Generation) UK Carbon Capture Technology					**APP. BY**	T. Tarrant	T. Tarrant				
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Name	Natural Gas to GT	GT Air Intake	Flue Gas to Cathode	Cathode Exhaust	Flue Gas To ID Fan	Treated Gas to Atm.	MP Steam to Anode	Fuels to Anode	Anode Exhaust	to CO2 Purification	Fuel Recycle	Product CO2
Pressure (bar abs)	70.00	1.01	1.04	0.83	0.80	1.01	6.00	3.00	1.08	32.00	76.40	110.00
Temperature (°C)	9.0	9.0	645.0	662.0	87.0	115.7	560.0	410.2	581.1	106.6	23.2	30.0
Mass rate (kg/h)	73221	2909824	2984091	2739578	2739578	2739578	64211	132648	392686	392686	46306	237412
Molar rate (kmol/h)	4064	100805	105204	99097	99097	99097	3563	6980	13666	13666	2189	5432
Volume rate (m3/h)	892	1900632	6290199	7520689	3002751	2560491	32306	103396	711863	5680	2087	311
Component												
Nitrogen (mol%)	0.89	77.21	74.07	78.63	78.63	78.63	0.00	2.22	1.14	1.14	6.59	0.22
Hydrogen (mol%)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	13.69	7.13	7.13	43.66	0.35
CO2 (mol%)	2.00	0.04	4.30	0.46	0.46	0.46	0.00	9.16	43.72	43.72	28.08	98.65
Methane (mol%)	89.00	0.00	0.00	0.00	0.00	0.00	0.00	15.66	0.00	0.00	0.00	0.00
Ethane (mol%)	7.00	0.00	0.00	0.00	0.00	0.00	0.00	1.23	0.00	0.00	0.00	0.00
Propane (mol%)	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.18	0.00	0.00	0.00	0.00
n-Butane (mol%)	0.10	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00
n-Pentane (mol%)	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen (mol%)	0.00	20.71	11.68	10.35	10.35	10.35	0.00	0.00	0.00	0.00	0.00	0.00
Argon (mol%)	0.00	0.92	0.88	0.93	0.93	0.93	0.00	0.00	0.00	0.00	0.00	0.00
CO (mol%)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.79	3.78	3.78	21.67	0.78
Water (mol%)	0.00	1.12	9.07	9.63	9.63	9.63	100.00	51.05	44.23	44.23	0.00	0.00
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

UTILITIES BALANCE SUMMARY - Case 7 - Molten Carbonate Fuel Cell Capture for Power Generation on Gas

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**REV - O2**		**SHEET**
CONTRACT:	13333	**DATE**	17/05/2017	11/09/2017		1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S Ferguson	S Ferguson		
		APP. BY	T Tarrant	T Tarrant		

UNIT No.	**DESCRIPTION**	**Electric Oper. Load**	**Fuel Gas LHV**	**Steam (t/h)**			**Condensate (t/h)**			**Cooling water**	**REV**
		MW	**MW**	**HPS 171 barg**	**MPS 40 barg**	**LPS 4.8 barg**	**HP T/h**	**MP T/h**	**LP T/h**	**T/h**	
Power Island											
	GT Gross Electrical Power	823.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	GT Auxiliaries	-10.3	-1940.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	ST Gross Electrical Power	381.5	0.0	700.8	802.3	894.9	0.0	0.0	0.0	0.0	
	ST Auxiliaries (inc BFW pumps, condenser)	-4.4	0.0	-700.8	-802.3	-894.9	0.0	0.0	894.9	-32412.9	
	Power Island Net	1190.3	-1940.2	0.0	0.0	0.0	0.0	0.0	894.9	-32412.9	
CO2 Capture											
	MCFC Fuel Pre-heater	0.0	0.0	0.0	0.0	-21.0	0.0	0.0	21.0	0.0	
	MCFC Stack	440.0	-586.5	0.0	0.0	-128.4	0.0	0.0	128.4	0.0	
	Anode Exhaust Cooling	0.0	0.0	0.0	0.0	158.5	0.0	0.0	-158.5	-12416.0	
	Flue Gas ID Fan	-46.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	CO2 Booster Compressor	-2.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-148.6	
	CO2 Capture Net	391.9	-586.5			9.2			-9.2	-12564.6	
CO2 Compression											
	CO2 Compressor	-64.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	-7512.0	
	CO2 Compression Net	-64.5								-7512.0	
Utilities & Offistes Systems											
	Cooling Water System	-9.1								52489.5	02
	Demin Plant	0									
	Buildings and Misc extra loads	0.0									
	Effluent Treatment	0.0									
	U&O Net **(assume included above)**	-9.1								52489.5	
Overall Performance											
	Gross Production	1645.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Total Parasitic Loads	-136.4	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Net Produced	1508.6	-2526.7	0.0	0.0	9.2	0.0	0.0	885.7	0.0	
	Net LHV Efficiency	59.71%	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	

	EQUIPMENT LIST FOR COMPRESSORS	Rev.	REV 01			
amec foster wheeler	Client: Department for Business, Energy & Industrial Strategy Contract No: 13333	Originated	SF			SHEET 1 of 8
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	Checked	RR			
	Unit No: UK Carbon Capture Technology	Approved	TT			
	Case: Natural Gas Combined Cycle with Post Combustion CO2 Capture	Date	12/05/2017			

EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT INLET / OUTLET	REMARKS	REV
FA-001	Flue Gas Induced Draft Fan (1/train)	Axial	1/train		3,684,280	1.387 / 1.384	0.213	0.800 / 1.013		0.999 / 1.000	23500	CS	27.65	Design Temp:150; Normal Op. Temp: 120 deg C	01
K-001	CO2 Compressor Package (6 stages)	Multi-Stage Integrally Geared	1/train		184,699	1.304 / 1.798	110.0	1.1 / 111.0		0.996 / 0.775	36120	SS316	36.65 / 43.70	237 t/h CO2	01
K-002	CO2 Re-booster	Centrifugal	1/train		4,579	1.322 / 1.317	8.3	5.8 / 14.1		0.965 / 0.959	1728	CS	43.66	59.8 t/h CO2	01

Notes:

EQUIPMENT LIST FOR HEAT EXCHANGERS

Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333
Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		
Unit No:	UK Carbon Capture Technology		
Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture		

Rev.	**REV 01**			**SHEET** 2 **of** 8
Originated	SF			
Checked	RR			
Approved	TT			
Date	17/05/2017			

EQUIPMENT NUMBER	**DESCRIPTION**	**EXCHANGER TYPE(1)/ SUB-TYPE**	**No.off x DUTY %**	**No.OF SHELLS/TUBES (ST)**	**TEMA TYPE(ST)/ HEADER CONST(AC) (2)**	**RATE(3)**	**DUTY MW**	**HEAT T'FER AREA(6) m^2**	**DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg**	**DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg**	**MATERIAL PLATE/ SHELL**	**MATERIAL TUBE(ST/AC) HEAD(AC)**	**No.OF BAYS/FANS (AC)**	**FAN TYPE (5)**	**TOTAL FAN POWER kW**	**REMARKS**	**REV**
E-001	MCFC Fuel Pre-heater	Shell & Tube	(1/train) X100%				8.1	2118	295.0 / 25.5	320.0 / 7	1.25 Cr 0.5 Mo Steel	1.25 Cr 0.5 Mo Steel					01
E-002	Anode Inlet/Outlet Exchanger	Gas / Gas Exchanger	5/train				3.0	486	605.0 / 2.0	687.0 / 2.0	Nickel Alloy 625 or 800H	Nickel Alloy 625 or 800H				Recently developed material UNS N06696 if available	01
E-003	Anode Heat Recovery Steam Superheater 1	Coil	(1/train) X100%				7.5	960	tubeside 585.0 / 10.0	ducting 600.0 / 2.0	6 Cr 0.5 Mo Steel	6 Cr 0.5 Mo Steel				Water, CO2, H2, CO	01
E-004	Anode Heat Recovery Steam Superheater 2	Coil	(1/train) X100%				5.9	283	tubeside 320.0 / 10.0	ducting 555.0 / 2.0	6 Cr 0.5 Mo Steel	6 Cr 0.5 Mo Steel				Water, CO2, H2, CO	01
E-005	Anode Heat Recovery Steam Generater	Coil	(1/train) X100%				47.4	5001	tubeside 180.0 / 10.0	ducting 505.0 / 2	3 Cr 0.5 Mo Steel	3 Cr 0.5 Mo Steel				Water, CO2, H2, CO	01
E-006	Anode Heat Recovery Economiser	Coil	(1/train) X100%				11.5	3229	tubeside 175.0 / 10.0	ducting 205.0 / 2	SS304	SS304				Water, CO2, H2, CO	01
E-007	Anode Exhaust Cooler	Shell & Tube	(1/train) X100%				81.87	32741	tubeside 51.0 / 10.0	shell 125.0 / 2.0	SS304	CS				Water, CO2, H2, CO	01
E-008	CO2 Purification Recuperator	Shell & Tube	1/train				1.0	150	100.0 / 25.0	130.0 / 35.0 (tubeside)	SS304	SS304					01
E-009	CO2 Purification Cooler	Shell & Tube	1/train				6.2	986	tubeside 51.0 / 10.0	135.0 / 35.0	SS304	CS					01
E-010	CO2 Purification Exchanger 1	Low Temperature Multipass Plate Exchanger	1/train				20.18	12361	Coldest -70.0 / 33.0	Warmest 55.0 / 33.0	SS304	SS304					01

Notes: **1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type**
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

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EQUIPMENT LIST FOR HEAT EXCHANGERS

Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Rev.**	**REV 01**	**SHEET** 3 **of** 8
Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		**Originated**	SF	
Unit No:	UK Carbon Capture Technology		**Checked**	RR	
Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture		**Approved**	TT	
			Date	17/05/2017	

EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	MATERIAL TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
E-011	CO2 Purification Exchanger 2	Low Temperature Multipass Plate Exchanger	1/train				0.30	78	Coldest -80.0 / 28.0	Warmest -65.0 / 28.0	SS304	SS304					01
E-012	CO2 Purification Exchanger 3	Shell & Tube	1/train				0.26	72	tubeside -78.0 / 28.0	-66.0 / 32.3	SS304	SS304					01
E-013	CO2 Purification Exchanger 4	Shell & Tube	1/train				5.50	5062	tubeside -81.0 / 28.0	-67.0 / 32.3	SS304	SS304					01
E-101	CO2 Compressor Cooler - Stage 1	Shell & Tube	1/train				7.61	1166	tubeside 51.0 / 10.0	145.0 / 6.5	SS304	CS					01
E-102	CO2 Compressor Cooler - Stage 2	Shell & Tube	1/train				6.55	1024	tubeside 51.0 / 10.0	145.0 / 9.4	SS304	CS					01
E-103	CO2 Compressor Cooler - Stage 3	Shell & Tube	1/train				6.88	900	tubeside 51.0 / 10.0	145.0 / 14.4	SS304	CS					01
E-105	CO2 Rebooster Cooler	Shell & Tube	1/train				0.98	135	tubeside 51.0 / 10.0	110.0 / 18.0	CS	CS					01
E-106	CO2 Compressor Cooler - Stage 5	Shell & Tube	1/train				6.60	722	tubeside 51.0 / 10.0	145.0 / 45.0	CS	CS					01
E-107	CO2 Compressor Cooler - Stage 6	Shell & Tube	1/train				15.61	1682	tubeside 51.0 / 10.0	145.0 / 115.0	CS	CS					01

Notes: **1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type**
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler	EQUIPMENT LIST FOR PUMPS				Rev.	REV 01			
	Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333	**Originated**	SF			SHEET 4 **of** 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	RR			
	Unit No:	UK Carbon Capture Technology			**Approved**	TT			
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture			**Date**	12/05/2017			

EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP / SG / VISC'Y °C cP	TEMPERATURE/PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
P-002	CO2 Compressor Condensate Return Pump	Pump	2 x 100% per train		120	119		3		25.0 1.00 0.798	Design Temp 60 Design Pressure: 5	12	SS304		01

Notes:

amec foster wheeler

EQUIPMENT LIST FOR VESSELS		Rev.	REV 01			
Client:	Department for Business, Energy & Idustrial Strategy Contract No: 13333	Originated	SF			SHEET 5 of 8
Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	Checked	RR			
Unit No:	UK Carbon Capture Technology	Approved	TT			
Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture	Date	12/05/2017			

EQUIPMENT NUMBER	DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	PRESS barg	VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	INTERNALS MAT./LINING/ CA	REMARKS	REV
V-001	CO2 Purification Water KO Drum	Vertical drum	1/train	2.2	4.4	19.51	V	50	33.00	1.013	demister	SS304	SS304		01
V-002	CO2 Purification KO Drum 1	Vertical drum	1/train	1.00	2.00	1.8	V	-70	33.00	1.013	demister	SS304	SS304		01
V-003	CO2 Purification KO Drum 2	Vertical drum	1/train	0.90	1.80	1.3	V	-80	30.00	1.013	demister	SS304	SS304		01
V-101	CO2 Compressor Stage 1 KO drum	Vertical drum	1/train	5.40	10.80	289	V	50	3.50	1.013	demister	SS304	SS304		01
V-102	CO2 Compressor Stage 2 KO drum	Vertical drum	1/train	4.40	3.80	156	V	50	3.50	1.013	demister	SS304	SS304		01
V-103	CO2 Compressor Stage 3 KO drum	Vertical drum	1/train	3.60	7.20	86	V	50	7.00	1.013	demister	SS304	SS304		01
V-104	CO2 Compressor Stage 4 KO drum	Vertical drum	1/train	2.80	5.60	40	V	50	27.00	1.013	demister	SS304	SS304		01

Notes:
1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

amec foster wheeler	EQUIPMENT LIST FOR TANKS				REV 01			
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	SF		SHEET 6 of 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR		
	Unit No:	UK Carbon Capture Technology			Approved	TT		
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture			Date	12/05/2017		

EQUIPMENT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS ID m	HEIGHT m	TOTAL VOLUME m^3	ROOF TYPE	BOTTOM TYPE	STORAGE PRESS	TEMP	SG	CORRO-SION	'MATERIALS	MANWAYS SHELL	ROOF	CONNECT-IONS	HEATING COIL	INSUL-ATION	REMARKS	REV
T-008	Firewater Storage Tank	1	14.0	8.2	1262.3	Tank			20			Lined CS						Design Pres.: 0.0075 / - 0.0025	01

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01			SHEET 7 of 8
	Client:	Department for Business, Energy & Industrial Strategy	Contract No:	13333	Originated	SF			
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR			
	Unit No:	UK Carbon Capture Technology			Approved	TT			
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture			Date	12/05/2017			

EQUIPMENT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
CCGT Power Island	**2 x GE 9HA.01 + 2 x ST (50 Hz).**										**Total GT Power = 823452 kW Total ST Power = 370823 kW**			**Gas Turbine World 2015-2016**	**01**
Z-001 A/B/C/D/E	**Molten Carbonate Fuel Cell 208-Stack ECM Enclosures**	**MCFC Fuel Cells**	**5/train**								**436000**			**Installed cost of 1200 €/kW, from FuelCell Energy**	**01**
Z-002	Mol Sieve Dehydration Package		1/train							Design Temp: 80 deg C Design Press.: 38 barg				dehydrates 284 t/h CO2, removes 222 kg/h water to get to spec of 100 ppm	01
Z-003	Flow metering and analyser package	Metering	1					474821						Fiscal metering package	01
	Nitrogen Package	Tank/pump	1						7			CS			01
	Compressed Air Package	Compressor	1											Design Temp- 80 / -10 deg C Opt. Temp: 20 deg C Design Pressure: 8.70 barg	01

Notes:

amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT		Rev.	REV 01			
	Client:	Department for Business, Energy & Idustrial Strategy Contract No: 13333	Originated	SF			SHEET 8 of 8
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	Checked	RR			
	Unit No:	UK Carbon Capture Technology	Approved	TT			
	Case:	Natural Gas Combined Cycle with Post Combustion CO2 Capture	Date	17/05/2017			

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Flue gas ducting to Fuel Cell Andoes 001 / DCC (1/train)	1/train	10 m X 10 m (Note 1) Estimated Length 200 m	Square		CS outer casing, Ni alloy lining plates, ceramic insultation in between.	Flow Rate: 2984 t/h Design Temp: 670/-10 deg C; Normal Op. Temp: 645 deg C Design/Operating Pressure: 0.30/0.10 barg	01
	Flue gas ducting to GT HRSG 001 / DCC (1/train)	1/train	10 m X 10 m (Note 1) Estimated Length 200 m	Square		CS outer casing, Ni alloy lining plates, ceramic insultation in between.	Flow Rate: 2720 t/h Design Temp: 687/-10 deg C; Normal Op. Temp: 662 deg C Design/Operating Pressure: 0.30/0.10 barg	01
	Cathode Ducting to Cathode heat recovery	1/train	2.3 m X 2.3 m Estimated Length 200 m	Square		CS outer casing, Ni alloy lining plates, ceramic insultation in between.	Flow Rate: 392.7 t/h Design Temp: 600 deg C; Normal Op. Temp: 575 deg C Design/Operating Pressure: 0.10/0.0 barg	01
	CO2 Analyser House							

Notes:

1) **Exhaust gas from the two GT trains is combined and distributed via a single large duct to each of the 10 fuel cell enclosures. The total length is 200m. The ducting to the first pair of enclosures is 40m long and 10 x 10m, this then reduces to a futher 40m section which is 9 x 9m, followed by another 40m section 8 x 8m, the penultimate 40m section is 6.5 x 6.5m and the final 40m section is 4.5 x 4.5m. The ducting returning the treated flue gas to the two GT HRSGs is of the same size with reversed gradual increase in cross sectional area.**

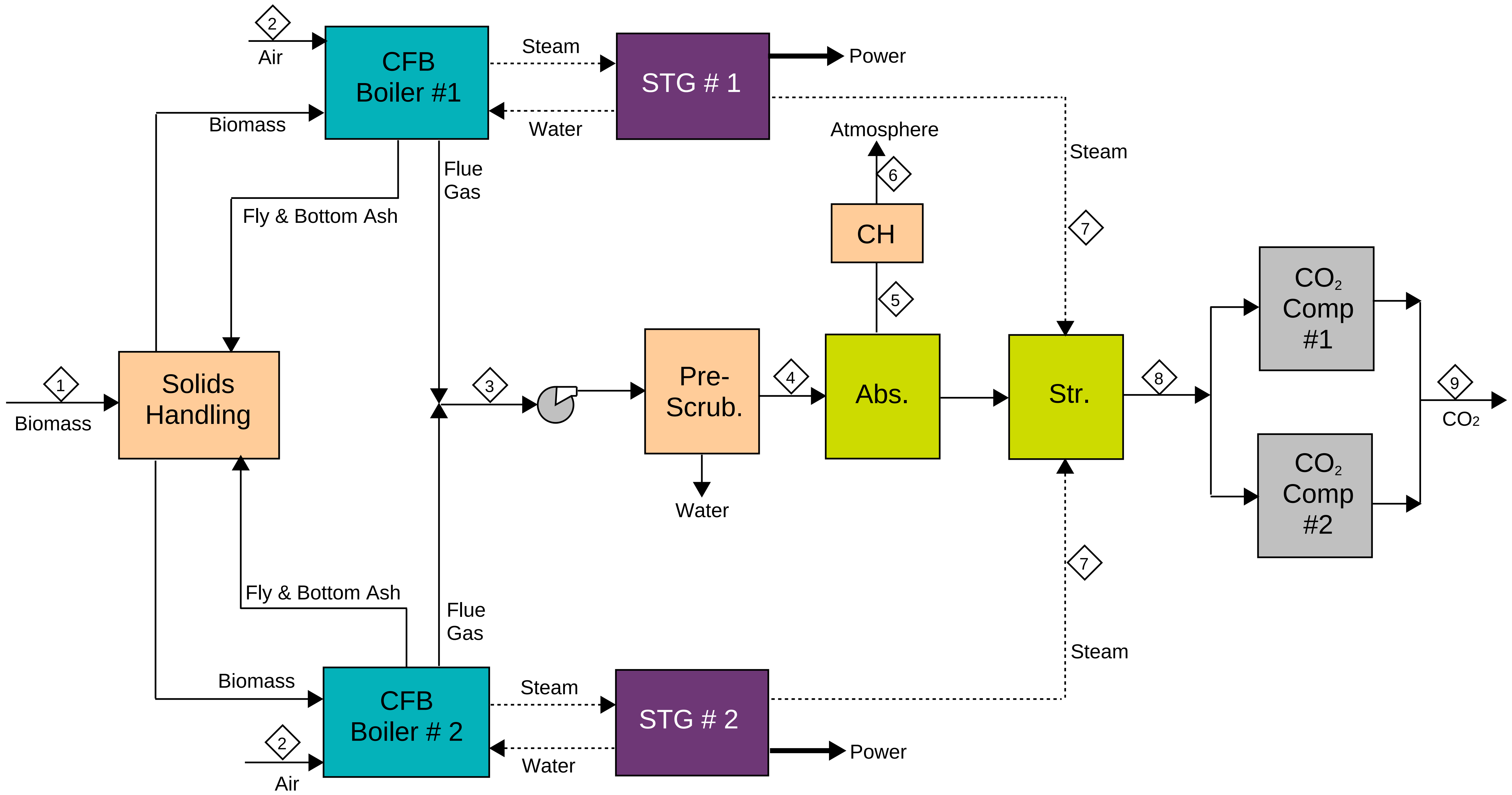
Amec Foster Wheeler	**Prepared By :**	**K.D. Nelson**
Client : BEIS	**Base Date :**	**1Q2017**
Project : Novel Carbon Capture Technology Study	**Rev. No. :**	**'6'**
Contract No.: 13333	**Print Date :**	**14-Dec-17**
Case 7 : Molten Carbonate Fuel Cell Capture for Power Generation on Gas		

Description	Case 7 : Molten Carbonate Fuel Cell Capture for Power Generation on Gas			
	Combined Cycle Block	CO_2 Capture Block	CO_2 Compression Block	Total
	Cost GBP	Cost GBP	Cost GBP	Cost GBP
Sub-Total Direct Materials	328,352,000	426,598,000	60,861,000	815,811,000
Other Material Costs				
Shipping / Freight	16,418,000	21,330,000	3,043,000	40,791,000
Third Party Inspection	3,284,000	4,266,000	609,000	8,159,000
Spare Parts (Comm/2yrs Op)	6,567,000	8,532,000	1,217,000	16,316,000
Sub-Total Materials	354,621,000	460,726,000	65,730,000	881,077,000
Material & Labour Contracts				
Civils/Steelwork & Buildings	49,253,000	63,990,000	9,129,000	122,372,000
Sub-Total Material & Labour Contracts	49,253,000	63,990,000	9,129,000	122,372,000
Labour Only Contracts				
Mechanical	59,103,000	63,990,000	10,955,000	134,048,000
Electrical/Instrumentation	16,418,000	12,798,000	3,043,000	32,259,000
Scaffolding/Lagging/Rigging	9,063,000	9,215,000	1,680,000	19,958,000
Sub-Total Material & Labour Contracts	84,584,000	86,003,000	15,678,000	186,265,000
Sub-Total Materials & Labour	488,458,000	610,719,000	90,537,000	1,189,714,000
EPCm Costs				
Engineering Services/Construction Management	73,269,000	91,608,000	13,581,000	178,458,000
Commissioning	9,769,000	12,214,000	1,811,000	23,794,000
Sub-Total EPCm Costs	83,038,000	103,822,000	15,392,000	202,252,000
Total EPC Cost	571,496,000	714,541,000	105,929,000	1,391,966,000
Other Costs				
Pre-Licensing, Technical and Design etc	5,715,000	7,145,000	1,059,000	13,919,000
Regulatory, Licensing and Public Enquiry etc	12,030,000	15,041,000	2,230,000	29,301,000
Infrastructure Connection Costs	37,000,000			37,000,000
Owners Costs	40,005,000	50,018,000	7,415,000	97,438,000
Sub-Total Other Costs	94,750,000	72,204,000	10,704,000	177,658,000
Total Project Costs	**666,246,000**	**786,745,000**	**116,633,000**	**1,569,624,000**

ATTACHMENT 10: Case 8 – CFB Boiler with Post-Combustion Capture[10]

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List

[10] Please note, as the basis of design for Benchmark 8 is based on Shell Cansolv's proprietary design, a capital cost estimate is not included to maintain confidentiality.



NOTES

Two parallel trains of all the equipment shown.

Department for Business, Energy & Industrial Strategy

amec foster wheeler

REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	SF	RR	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram - Case 8 - Biomass CFB with Post Combustion CO_2 Capture

Drawing No

Rev

HEAT & MATERIAL BALANCE - Case 8 - Biomass Post-Combustion Capture for Power Generation

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**SHEET**
CONTRACT:	13333	**DATE**	22/05/2017	1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	R. Ray	
		APP. BY	T. Tarrant	

Stream Number	1	2	3	4	5	6	7	8	9
Stream Name	B/L Biomass Feed	Total Air Intake	Flue Gas from Boiler	Flue Gas from Prescrubber	Flue Gas from Absorber	Flue Gas to Atmosphere	LP Steam to Reboilers	CO2 to Compression	Product CO2
Pressure (bar abs)	1.01	1.01	1.01	1.06	1.02	1.01	4.41	2.01	110.00
Temperature (°C)	9.0	9.0	90.0	35.0	34.0	80.0	148.0	49.0	30.0
Mass rate (kg/h)	635178	2456787	3075000	2677654	2140549	2140549	580000	537002	523849
Molar rate (kmol/h)			111155	89101	76470	76470	32186	12631	11903
Volume rate (m3/h)			3305035	2158375	1918483	2200234	245230	169602	1377
Component									
Hydrogen (mol%)	%wt AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO (mol%)	C: 25%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2 (mol%)	H: 2.7%	0.03	11.90	14.85	1.73	1.73	0.00	94.25	100.00
Nitrogen (mol%)	O: 21.1%	77.27	60.10	74.97	87.36	87.36	0.00	0.00	0.00
Oxygen (mol%)	S: 0.03%	20.73	3.90	4.87	5.67	5.67	0.00	0.00	0.00
Argon (mol%)	N: 0.15%	0.92	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methane (mol%)	Cl: 0.01%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Water (mol%)	Moisture: 50%	1.05	24.10	5.31	5.24	5.24	100.00	5.75	0.00
SO2 (mol%)	Ash: 1%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total		100.00	0.00	100.00	100.00	100.00	100.00	100.00	100.00

UTILITIES BALANCE SUMMARY - Case 8 - Biomass fired CFB Boiler with CO2 Capture

CLIENT:	Department for Business, Energy & Industrial Strategy			CHANGE			REV - O1				SHEET
CONTRACT:	13333			DATE			07/06/2017				1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology			ORIG. BY			R Ray				
				APP. BY			T Tarrant				
UNIT No.	DESCRIPTION	Electric Oper. Load	Coal Feed LHV	Steam (t/h)			Condensate (t/h)			Cooling water	REV
		MW	MW	HPS 175 barg	MPS 38 barg	LPS 3.4 barg	HP T/h	MP T/h	LP T/h	T/h	
Solids Handling and Boiler Island											
	Solid Handilng & Storage	-1.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Boiler Island & Flur Gas Treatment	-22.6	-1288.0	1648.8	1476.0	1255.2	0.0	0.0	0.0	0.0	
	Power Island Auxilliaries (ST)	-2.7	0.0	0.0	-220.8		0.0	220.8		0.0	
	Power Island (ST)	498.0	0.0	-1648.8	-1255.2	-675.2	0.0	0.0	675.2	-20967	
	Power Island Net	471.0	-1288.0	0.0	0.0	580.0		#REF!	675.2	-20967.1	
CO2 Capture											
	CO2 Capture Net	-24.7	0.0	0.0	0.0	-580.0			580.0	-59370.0	
CO2 Compression & Pumping											
	CO2 Compressor	-38.1		0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Supercritical CO2 Pump	-1.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Inter/After Coolers & Condenser	0.0		0.0	0.0	0.0	0.0	0.0	0.0	-6620	
	CO2 Compression Net	-39.1	0.0							-6620.5	
Utilities & Offistes Systems											
	Cooling Water System	-9.6								87057.6	
	Buildings and Misc extra loads	-1.4								-100.0	
	U&O Net	-11.0	0.0								
Overall Performance											
	Gross Production	498.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Total Parasitic Loads	-101.8	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Net Produced	396.2	-1288	0.0	0.0	0.0	0.0	0.0	#REF!	#REF!	
	Net LHV Efficiency	30.8%	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	

amec foster wheeler	EQUIPMENT LIST FOR COMPRESSORS	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	**Originated**	RR			**SHEET** 1 **of** 9
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)	**Checked**	SF			
	UK Carbon Capture Technology	**Approved**	TT			
	Case: Biomass fired CFB Boiler with CO2 Capture	**Date**	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
K-401	Flue Gas Fan	Axial	1			1.364 / 1.363				0.999 0.999			27.66	**CO2 Capture Unit Scope**	
K-402	MVR Compressor	Centrifugal	1			1.326 / 1.317				0.993 / 0.992			18.01	**CO2 Capture Unit Scope**	
K-501	CO2 Compressor Package (4 stages)	Multi-Stage Integrally Geared	2 x 50%		83,228	1.285 / 1.592	70.0	2.0 / 72.0		0.990 0.829	19028	CrNi alloy	42.52	262 t/h CO2	
						/		/							

Notes:

amec foster wheeler	EQUIPMENT LIST FOR HEAT EXCHANGERS	Rev.	REV 01	REV 02	REV 03	
	Client: Department for Business, Energy & Idustrial Strategy Contract No: 13333	Originated	RR			SHEET 2 of 9
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	SF			
		Approved	TT			
	Case: Biomass fired CFB Boiler with CO2 Capture	Date	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS /TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	MATERIAL TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
E-401	Pre-Scrubber Cooler	Gasketed Plate and frame	2						51.0 / 10.0	80.0 / 7						CO2 Capture Unit Scope	
E-402	CO2 Absorber Intercooler	Gasketed Plate and frame	2						51.0 / 10.0	80.0 / 7						CO2 Capture Unit Scope	
E-403	CO2 Wash Water cooler	Gasketed Plate and frame	1						51.0 / 10.0	80.0 / 7						CO2 Capture Unit Scope	
E-404	Lean amine cooler	Gasketed Plate and frame	2						51.0 / 10.0	140.0 / 9.5						CO2 Capture Unit Scope	
E-405	Lean/Rich exchangers	Gasketed Plate and frame	12						80.0 / 10.3	140.0 / 6						CO2 Capture Unit Scope	
E-406	Stripper condenser	Welded Plate and frame	2						51.0 / 10.0	150.0 2/FV						CO2 Capture Unit Scope	
E-407	Stripper Reboilers	Welded Plate and frame	11						150.0 / 2/FV	295.0 / 5/FV (tubeside)						CO2 Capture Unit Scope	
E-408	Condensate Heater		1						105.0 / 2.000	125 / 2.0						CO2 Capture Unit Scope	
E-501	CO2 Compressor Cooler - Stage 1	Shell & Tube	2				9.51	1402	tubeside 51.0 / 10.0	150.0 / 8.2	SS304	SS304					
E-502	CO2 Compressor Cooler - Stage 2	Shell & Tube	2				6.50	855	tubeside 51.0 / 10.0	150.0 / 12.3	SS304	SS304					
E-503	CO2 Compressor Cooler - Stage 3	Shell & Tube	2				6.88	786	tubeside 51.0 / 10.0	150.0 / 33	SS304	SS304					
E-504	CO2 Condenser	Shell & Tube	2				18.42	2317	tubeside 51.0 / 10.0	150.0 / 76	SS304	SS304					
E-505	CO2 Product Cooler	Shell & Tube	2				0.92	309	tubeside 51.0 / 10.0	100.0 / 115	SS304	SS304					

Notes: **1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type**
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler	EQUIPMENT LIST FOR PUMPS				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 3 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Biomass fired CFB Boiler with CO2 Capture			Date	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP / SG / VISC'Y °C		cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
P-401	Prescrubber Pumps	Centrifugal	3 x 50%							30.0	1.004	0.797	Design Temp: 80/-10 Design Press.: 7.5			**CO2 Capture Unit Scope**	
P-402	Absorber WW pump	Centrifugal	3 x 50%							30.0	1.004	0.797	Design Temp: 80/-10 Design Press.: 7			**CO2 Capture Unit Scope**	
P-403	CO2 Intercooler Pumps	Centrifugal	3 x 50%							30.0	1.004	0.797	Design Temp: 80/-10 Design Press.: 7			**CO2 Capture Unit Scope**	
P-404	Rich amine pumps	Centrifugal	3 x 50%							36.0	0.983		Design Temp: 80/-10 Design Press.: 11			**CO2 Capture Unit Scope**	
P-405	Stripper reflux pump	Centrifugal	3 x 50%							137.0	0.919	0.205	Design Temp: 295/-10 Design Press.: 12.4/FV			**CO2 Capture Unit Scope**	
P-406	Lean amine pumps	Centrifugal	3 x 50%							119.0	0.95		Design Temp: 140/-10 Design Press.: 8/FV			**CO2 Capture Unit Scope**	
P-407	Lean amine feed pumps	Centrifugal	3 x 50%							50.0	0.95		Design Temp: 140/-10 Design Press.: 8/FV			**CO2 Capture Unit Scope**	
P-408	Amine Drain Pump	Centrifugal	1 x 100%							60.0	0.983		Design Temp: 100/-10 Design Press.:2.40			**CO2 Capture Unit Scope**	
P-501	Supercritical CO2 Pump	Centrifugal	2 x 50%		262.0	354.0		40.0		25.0	0.702	0.061	Design Temp: 140/-10 Design Press.:140.0	514.0	304L SS		
P-601	Demin Water Pump	Pump	3					1		20.0			Design Temp- 80 / -10 Design Pressure: 5.20		Stainless Steel casing and impeller	Sizing: 1.7 x 0.7 x 1.4	
P-602	Towns Water Transfer Pump	Centrifugal	3					1		20.0			Design Temp- 80 / -10 Design Pressure: 6.0		Cast iron casing with Stainless steel impeller	Size: 1.8 X 1.8 X 1.8	
P-603	Firewater Pump Package	Pump	3							20.0			Design Temp- 80 / -10 Design Pressure: 19		CS	Sizing: 8 x 4 x 3.2 1 x diesel, 1 x electric and 1 x jockey	
P-502	CO2 Compressor Condensate Return Pump	Pump	2 x 100%		6.6	6.6		3		25.0	0.996	0.900	Design Temp- 80 / -10 Design Pressure: 6.0	1	CS		

Notes:

	EQUIPMENT LIST FOR VESSELS			Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	Client:	Department for Business, Energy & Idustrial Strategy	Contract No: 13333	Originated	RR			SHEET 4 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		Checked	SF			
		UK Carbon Capture Technology		Approved	TT			
	Case:	Biomass fired CFB Boiler with CO2 Capture		Date	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	DIMENSIONS HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	DESIGN CONDITIONS PRESS barg	DESIGN CONDITIONS VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	MATERIALS OF CONST'N INTERNALS MAT./LINING/ CA	REMARKS	REV
C-401	Prescrubber	Rectangular column	1				V	150	0.30	1.013				**CO2 Capture Unit Scope**	
C-402	Absorber (absorber section)	Rectangular column	1				V	80/-10	0.10	1.013				**CO2 Capture Unit Scope**	
C-403	Absorber (water wash section)	Rectangular column	1				V	80/-10	0.10	1.013				**CO2 Capture Unit Scope**	
C-405	CO2 Stripper	Vertical cylinder	1				V	250/-10	3.50	FV				**CO2 Capture Unit Scope**	
V-401	Condensate Flash Vessel	Horizontal Drum	1				H	250/-10	3.50	FV				**CO2 Capture Unit Scope**	
V-402	CO2 Reflux Accumulator	Horizontal Drum	1				H	295/-10	5.00	FV				**CO2 Capture Unit Scope**	
V-403	Lean Flash Vessel	Horizontal Drum	1				H	295/-10	5.00	FV				**CO2 Capture Unit Scope**	
V-501	CO2 Compressor Stage 1 KO drum	Vertical drum	2	4.50	9.00	143	V	80	3.50	1.013	demister	SS304	SS304	262 t/h CO2	
V-502	CO2 Compressor Stage 2 KO drum	Vertical drum	2	3.50	7.00	79	V	80	6.00	1.013	demister	SS304	SS304	262 t/h CO2	
V-503	CO2 Compressor Stage 3 KO drum	Vertical drum	2	2.70	5.40	36	V	80	12.00	1.013	demister	SS304	SS304	262 t/h CO2	
V-504	CO2 Compressor Stage 4 KO drum	Vertical drum	2	2.10	4.20	17	V	80	33.00	1.013	demister	SS304	SS304	262 t/h CO2	

Notes:
1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

amec foster wheeler	EQUIPMENT LIST FOR TANKS			Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Originated**	RR			**SHEET** 5 **of** 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		**Checked**	SF			
		UK Carbon Capture Technology		**Approved**	TT			
	Case:	Biomass fired CFB Boiler with CO2 Capture		**Date**	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS ID m	DIMENSIONS HEIGHT m	TOTAL VOLUME m^3	ROOF TYPE	BOTTOM TYPE	STORAGE PRESS	STORAGE TEMP	SG	CORRO-SION	'MATERIALS	MANWAYS SHELL	MANWAYS ROOF	CONNECT-IONS	HEATING COIL	INSUL-ATION	REMARKS	REV
T-401	Lean Amine Tank	1				Storage tank		0.02	30						Vertial, sized for full inventory			**CO2 Capure Unit Scope**	
T-402	Amine Drain Tank	1													Horizontal, underground.			**CO2 Capure Unit Scope**	
T-403	Absorbent Make-up tank	1				Storage tank		0.02	30						Vertical			**CO2 Capure Unit Scope**	
T-601	Towns Water Storage Tank	1	10.00	10.00	780.00	Vertical cylindrical		0.0075 / -0.0025	20			Lined CS							
T-602	Demin Water Tank	1	10.00	9.00	700.00	Tank		0.90	20			Lined CS							
T-603	Firewater Storage Tank	1	13.0	7.8		Tank			20			Lined CS							

Notes:

	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	**Client:**	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333	**Originated**	RR			**SHEET** 6 **of** 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	SF			
		UK Carbon Capture Technology			**Approved**	TT			
	Case:	Biomass fired CFB Boiler with CO2 Capture			**Date**	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
200	Subcritical Circulated Fluidised Bed (CFB) Boiler Includes: Fuel Feeding System The furnace The solid seperators with the solid return channels INTREX superheaters, back pass, fans and air heater	318 t/h biomass feed to each CFB Boiler 2 trains				Thermal Input: 644 MWth (LHV) to each Boiler; 2 x Boiler in Operation Main steam conditions: 176 bar (abs), 568 °C Reheat steam conditions: 39 bar (abs), 568 °C	
200	Stack						
200	Continuous Emissions Monitoring System						
300	Steam Turbine & Generator Package	2 trains		249 MW Output			
600	Cooling Tower and Packages		Evaporative, Natural Drive Cooling Tower				

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 7 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Biomass fired CFB Boiler with CO2 Capture			Date	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3 /h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
S-401	Ion Exchange Package		1											**CO2 Capture Unit Scope**	
S-402	Thermal Reclaimer Package	Vacuum Distillation Column	1											**CO2 Capture Unit Scope**	
S-403	CO2 Absorbent Filtration Unit	Cartridge Type Filter	1											**CO2 Capture Unit Scope**	
S-404	Activated Carbon Filtration Unit	Cartridge Type Filter	1											**CO2 Capture Unit Scope**	
S-601	Nitrogen Package	Tank/pump	1						7	Design Temp: 80/-10 deg C Design Press.: 14 barg Oper. Temp.: 20 deg C		CS			
S-602	Compressed Air Package	Compressor	1							Design Temp- 80 / -10 deg C Design Pressure: 8.70 barg					
S-501	TEG Dehydration Package		2							Design Temp: 150/-10 deg C Design Press.: 76 barg				Dehydrates 262 t/h CO2, removes 245 kg/h water to get to spec of 50 ppm	
S-603	Flow metering and analyser package	Metering	1					524000 kg/h CO2		Design Temp: 150/-10 deg C Design Press.: 114 barg				Fiscal metering package	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR SOLIDS HANDLING EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 8 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
		UK Carbon Capture Technology			Approved	TT			
	Case:	Biomass fired CFB Boiler with CO2 Capture			Date	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
100	Biomass Handling and Storage	318 t/h biomass feed 2 trains					

Notes:

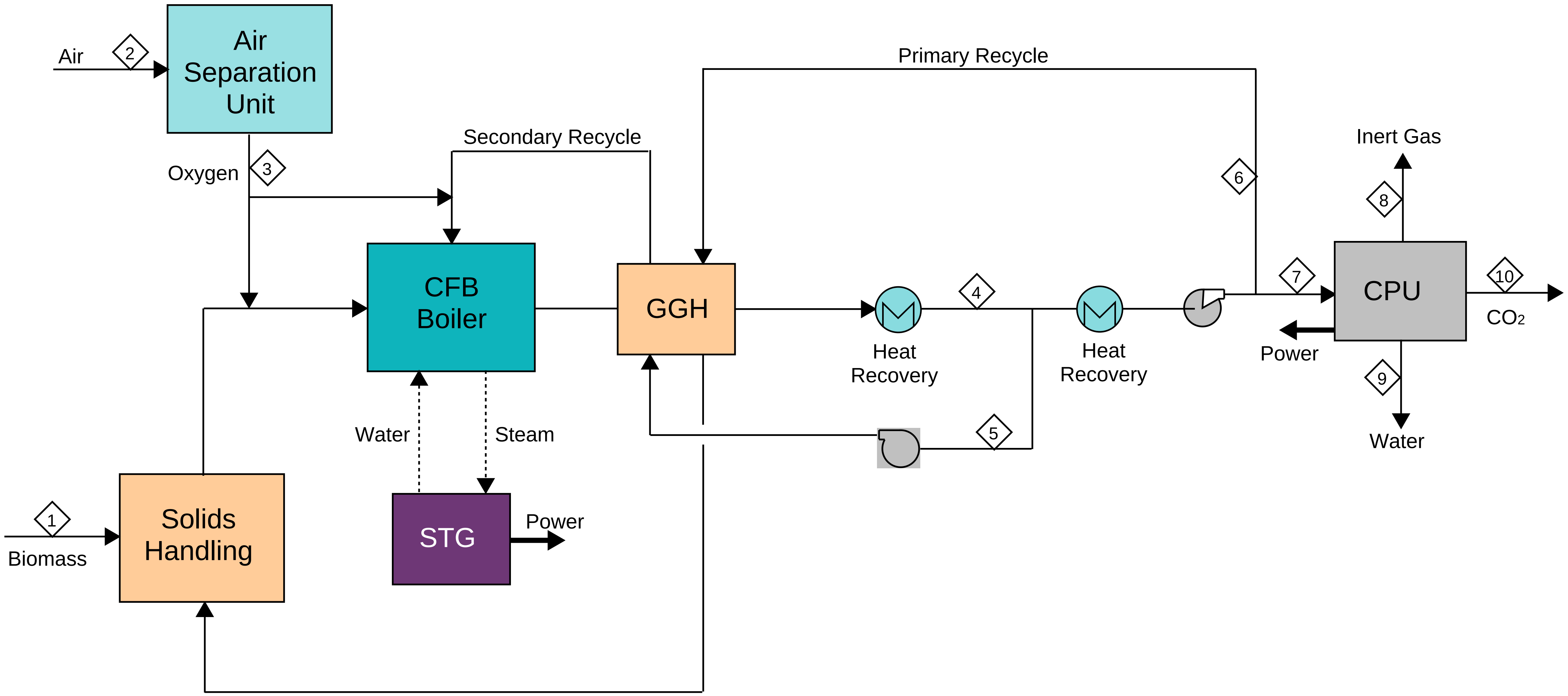
amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET 9 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF			
	Unit No:	UK Carbon Capture Technology			Approved	TT			
	Case:	Biomass fired CFB Boiler with CO2 Capture			Date	05/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Flue gas ducting from Boiler to fan		5 m X 5 m Estimated Length 5 m	Square		CS	Flow Rate: 3075 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 90 deg C Design/Operating Pressure: 0.30/0.10 barg	
	Flue gas ducting from fan to Prescrubber		5 m X 5 m Estimated Length 5 m	Square		CS	Flow Rate: 3075 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 100 deg C Design/Operating Pressure: 0.30/0.10 barg	
	Flue gas ducting from prescrubber to absorber		4 m X 4 m Estimated Length 27 m	Square		CS	Flow Rate: 2678 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 35 deg C Design/Operating Pressure: 0.30/0.10 barg	
	Flue gas ducting from absorber to Condensate Heater		4 m X 4 m Estimated Length 75 m	Square		CS	Flow Rate: 2141 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 34 deg C Design/Operating Pressure: 0.30/0.10 barg	
	Flue gas ducting from Condensate Heater to Power Island stack		4.5 m X 4.5 m Estimated Length 10 m	Square		CS	Flow Rate: 2141 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 80 deg C Design/Operating Pressure: 0.30/0.10 barg	

Notes:

ATTACHMENT 11: Case 9 – CFB Boiler with Oxy-Combustion

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Capital Cost Estimate Summary



NOTES

Two parallel trains of all the equipment shown.

Department for Business, Energy & Industrial Strategy

amec foster wheeler

REV.	DATE	ORIG.	CHK'D	APPR.
01	27/04/17	SF	RR	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram - Case 8 - Biomass CFB with Post Combustion CO_2 Capture

Drawing No

Rev

HEAT & MATERIAL BALANCE - Case 9 - Oxy-Combustion Capture for Power Generation on Biomass

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**
CONTRACT:	13333	**DATE**	19/06/2017
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	R. Ray
		APP. BY	T. Tarrant

SHEET 1 OF 1

Stream Number	1	2	3	4	5	6	7	8	9	10
Stream Name	B/L Biomass Feed	Air to ASU	Oxygen to Boiler	Flue Gas from ESP	Secondary Recycle	Primary Recycle to FGD	Flue gas to CPU	Inert gas from CPU	Water from CPU	Product CO2
Pressure (bar abs)	1.01	1.01	1.01	1.01	1.01	1.01	1.01	1.60	2.00	110.00
Temperature (°C)	9.0	9.0	9.0	160.0	160.0	38.6	38.6	210.0	25.0	30.0
Mass rate (kg/h)	635178	2007000	477186	2955624	1126960	642894	721504	187004	11407	523093
Molar rate (kmol/h)		69569	14868	96890	36943	16106	18074.93	5570	618	11886
Volume rate (m3/h)		1609538	37230	3762504	1434618	391434	439296			762
Component										
Hydrogen (mol%)	%wt AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO (mol%)	C: 25%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2 (mol%)	H: 2.7%	0.03	0.00	41.62	41.62	72.99	72.99	23.47	0.04	99.99
Nitrogen (mol%)	O: 21.1%	77.27	1.20	7.05	7.05	12.37	12.37	40.13	0.00	0.00
Oxygen (mol%)	S: 0.03%	20.73	97.00	5.40	5.40	9.47	9.47	30.71	0.00	0.01
Argon (mol%)	N: 0.15%	0.92	1.80	0.92	0.92	1.62	1.62	5.24	0.00	0.00
Water (mol%)	Cl: 0.01%	1.05	0.00	44.98	44.98	3.52	3.52	0.46	98.86	0.00
SO2 (mol%)	Moisture: 50%	0.00	0.00	0.02	0.02	0.03	0.03	0.00	0.81	0.00
NO / NO2 (mol%)	Ash: 1%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HNO3 (mol%)		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2SO4 (mol%)		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total		100	100	100	100	100	100	100	100	100

UTILITIES BALANCE SUMMARY - Case 9 - Biomass fired Oxy-CFB Boiler with CO2 Capture

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**				**SHEET**
CONTRACT:	13333	**DATE**	14/06/2017				1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	R Ray				
		APP. BY	T Tarrant				

UNIT No.	**DESCRIPTION**	**Electric Oper. Load**	**Coal Feed LHV**	**Steam (t/h)**			**Condensate (t/h)**			**Cooling water**	**REV**
		MW	**MW**	**HPS 175 barg**	**MPS 38 barg**	**LPS 3.4 barg**	**HP T/h**	**MP T/h**	**LP T/h**	**T/h**	
Solids Handling and Boiler Island											
	Solid Handilng & Storage	-1.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
	Boiler Island & Flur Gas Treatment	-22.6	-1288.0	1648.8	1476.0	1255.2					
	Power Island Auxilliaries (ST)	-3.7	0.0		-220.8			220.8			
	Power Island (ST)	598.0	0.0	-1648.8	-1255.2	-1255.2			1255.2	-39984	
	Power Island Net	570.0	-1288.0	0.0	0.0	0.0				-39983.6	
CO2 Purification and Compression											
	CO2 Capture Net	-68.0								-10931.1	
Utilities & Offistes Systems											
	U&O Net	-11.0								51774.2	
Air Separation Unit											
	ASU Net	-88.9								-859.5	
Overall Performance											
	Gross Production	598.0								0.0	
	Total Parasitic Loads	-195.9									
	Net Produced	402.1	-1288								
	Net LHV Efficiency	31.2%	0								

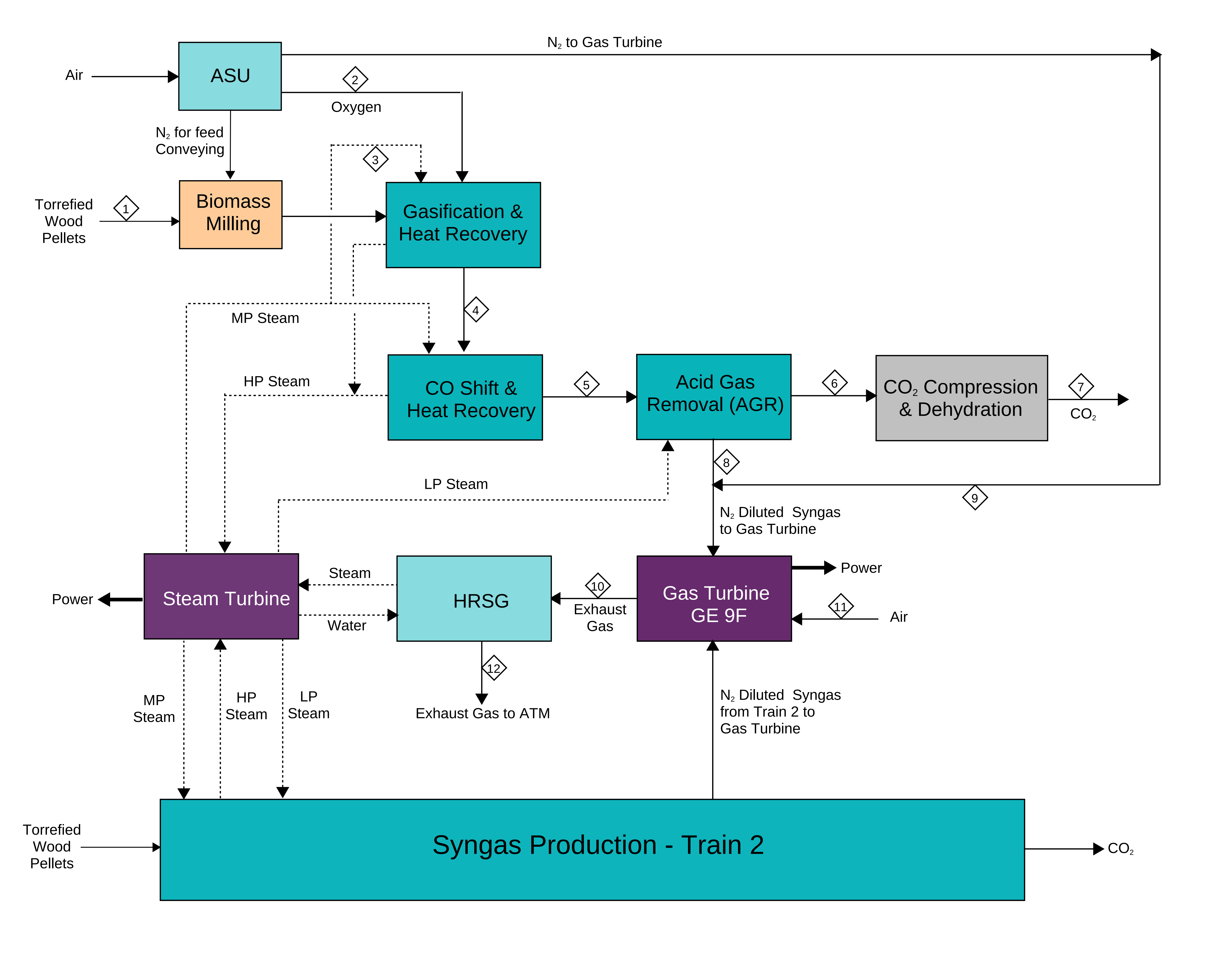
Amec Foster Wheeler
Client : BEIS
Project : Novel Carbon Capture Technology Study
Contract No.: 13333
Case 9 : Oxy-combustion Capture for Power Generation on Biomass

Prepared By : K.D. Nelson
Base Date : 1Q2017
Rev. No. : '6'
Print Date : 14-Dec-17

Description	Case 9 : Oxy-combustion Capture for Power Generation on Biomass						
	Unit 100	Unit 200	Unit 300	Unit 400	Unit 500	Unit 600	
	Feedstock & Solids Handling	Boiler Island	Steam Cycle	CPU	ASU	Utility Units	Total
	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP
Sub-Total Direct Materials	49,600,000	204,400,000	99,800,000	104,500,000	111,996,000	114,060,000	684,356,000
Other Material Costs							-
Shipping / Freight	2,480,000	10,220,000	4,990,000	5,225,000	5,600,000	5,703,000	34,218,000
Third Party Inspection	496,000	2,044,000	998,000	1,045,000	1,120,000	1,141,000	6,844,000
Spare Parts (Comm/2yrs Op)	992,000	4,088,000	1,996,000	2,090,000	2,240,000	2,281,000	13,687,000
Sub-Total Materials	53,568,000	220,752,000	107,784,000	112,860,000	120,956,000	123,185,000	739,105,000
Material & Labour Contracts							
Civils/Steelwork & Buildings	12,400,000	65,408,000	14,970,000	12,540,000	27,999,000	22,812,000	156,129,000
Sub-Total Material & Labour Contracts	12,400,000	65,408,000	14,970,000	12,540,000	27,999,000	22,812,000	156,129,000
Labour Only Contracts							
Mechanical	4,960,000	51,100,000	17,964,000	12,540,000	33,599,000	18,250,000	138,413,000
Electrical/Instrumentation	1,984,000	14,308,000	4,990,000	3,135,000	16,799,000	5,703,000	46,919,000
Scaffolding/Lagging/Rigging	833,000	7,849,000	2,754,000	1,881,000	6,048,000	2,874,000	22,239,000
Sub-Total Material & Labour Contracts	7,777,000	73,257,000	25,708,000	17,556,000	56,446,000	26,827,000	207,571,000
Sub-Total Materials & Labour	73,745,000	359,417,000	148,462,000	142,956,000	205,401,000	172,824,000	1,102,805,000
EPCm Costs							
Engineering Services/Construction Management	11,062,000	53,913,000	22,269,000	21,443,000	30,810,000	25,924,000	165,421,000
Commissioning	1,475,000	7,188,000	2,969,000	2,859,000	4,108,000	3,456,000	22,055,000
Sub-Total EPCm Costs	12,537,000	61,101,000	25,238,000	24,302,000	34,918,000	29,380,000	187,476,000
Total EPC Cost	86,282,000	420,518,000	173,700,000	167,258,000	240,319,000	202,204,000	1,290,281,000
Other Costs							
Pre-Licensing, Technical and Design etc	863,000	4,205,000	1,737,000	1,673,000	2,403,000	2,022,000	12,903,000
Regulatory, Licensing and Public Enquiry etc	1,806,000	8,800,000	3,635,000	3,500,000	5,029,000	4,229,000	26,999,000
Infrastructure Connection Costs						29,000,000	29,000,000
Owners Costs	6,040,000	29,436,000	12,159,000	11,708,000	16,822,000	14,154,000	90,319,000
Sub-Total Other Costs	8,709,000	42,441,000	17,531,000	16,881,000	24,254,000	49,405,000	159,221,000
Total Project Costs	**94,991,000**	**462,959,000**	**191,231,000**	**184,139,000**	**264,573,000**	**251,609,000**	**1,449,502,000**

ATTACHMENT 12: Case 10 – Biomass IGCC

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List
- Capital Cost Estimate Summary



NOTES

1. Two Trains of Syngas Production Unit

2. Syngas Production Unit includes ASU, Biomass Milling, Gasification Island, CO Shift, AGR and CO2 Compression & Dehydration

3. One train of Power Island including Gas Turbine, HRSG & Steam Turbine

Department for Business, Energy & Industrial Strategy

amec foster wheeler

REV.	DATE	ORIG.	CHK'D	APPR.
01	08/06/17	RR	SF	TT

Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram
Case 10 - Biomass IGCC Case

Drawing No

Rev

amec foster wheeler

HEAT & MATERIAL BALANCE - Case 10 Torrefied Biomass IGCC

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**SHEET**	
CONTRACT:	13333	**DATE**	28/06/2017	1 OF 1	
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	R. Ray		
		APP. BY	T. Tarrant		

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12
Stream Name	B/L Torrefied Biomass	Oxygen to Gasifier	Steam to Gasifier	Syngas to Shift Reactor	Shifted Syngas to AGR	CO2 to Compression	Product CO2	H2-rich Gas to Gas Turbine	N2 to GT as Diluent	Flue Gas from Gas Turbine	Air to Gas Turbine	Exhaust Gas
Temperature (°C)	AMB	25.00	302.40	260.00	34.00		30.00	160.00	30.00	565.00	9.00	80.00
Pressure (bar abs)	ATM	49.50	44.00	38.50	34.20		110.00	29.90	33.40	1.04	9.00	ATM
Mass rate (kg/h)	225417	115100	12342	478566	458087	357266	356699	93479	478400	3178412	2539210	3178412
Molar rate (kmol/h)		3580	685	22582	21264	8254	8222	12836	17078	116273	87994	116273
Volume rate (m3/h)		1722	657	25394	15389	18529	504	13616	17497	7793006	2202554	84768381
Molecular Weight		32.2	18.0	21.2	21.5	43.3	43.4	7.3	28.0	27.3	28.86	27.3
Component												
H2 (mol%)	%wt AR	0.00	0.00	15.37	50.67	1.44	1.44	83.01	0.00	0.00	0.00	0.00
CO (mol%)	C: 47.51%	0.00	0.00	32.77	0.43	0.01	0.01	0.71	0.00	0.00	0.00	0.00
CO2 (mol%)	H: 5.13%	0.00	0.00	6.33	41.67	98.04	98.43	4.75	0.00	0.62	0.03	0.62
Methane (mol%)	O: 40.1%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N2 (mol%)	S: 0.05%	1.50	0.00	6.14	6.52	0.12	0.12	10.72	100.00	74.38	77.31	74.38
Oxygen (mol%)	N: 0.31%	95.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	11.07	20.74	11.07
Argon (mol%)	Cl: 0.02%	3.50	0.00	0.45	0.47	0.00	0.00	0.78	0.00	0.79	0.93	0.79
H2S (mol%)	Moisture: 3.0%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
COS (mol%)	Ash: 3.88%	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Water (mol%)		0.00	100.00	38.93	0.22	0.39	0.00	0.03	0.00	13.13	0.99	13.13
Total		100	100	100	100	100	100	100.0	100.00	100.0	100.0	100.0

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Case 10 - Pre-Combustion Capture for Power Generation on Torrefied Wood

CLIENT: Department for Business, Energy & Industrial Strategy
CONTRACT: 13333
NAME: Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

CHANGE	REV - O1				SHEET
DATE	28/06/2017				1 OF 1
ORIG. BY	R. Ray				
APP. BY	T. Tarrant				

DESCRIPTION	Electric Oper. Load	Steam (t/h)			Condensate (Note 2)	Fresh Cooling Water	Demin Water	Process Water	REMARKS	REV
	KW	127 bar	47 bar	8 bar	LP (t/h)	t/h	t/h	t/h		
Syngas Production Units					139.9					
Coal Handling & Storage	-265									
Gasification Island	-18800		-12			-2013				
Syngas Cooling and Shift Reactors	-421		-277.4			-4883		183.7		
Acid Gas Recovery Unit	-11059					-7387				
Air Separation Unit	-48949			-139.9		-4057				
N2 Compression & Others	-21014					-1596				
Sulphur Recovery/Tailgas Treatment Unit	0					0				
CO2 Compression and Drying	-24049					-3418		0.3		
Process Heat Recovery Steam Production		209.3	276.4	19.6						
Process Units Total	-124557	209.3	-13.4	-120.3	139.9	-23354.5	0.0	184.0		
Power Island										
GT Gross Electrical Power	302896									
GT Auxiliaries	-1000									
HRSG & Auxiliaries (inc BFW pumps, condense	-5305	253.9	55.9	41.1			-365.3	1.9		
ST Electrical Power	190398	-463.3	-505.8	-426.7	426.7	-19745.4				
Power Island Total	486989	-209.3	13.4	120.3	426.7	-19745.4	-365.3	1.9		
Offsites & Utilities										
Fresh Cooling Water	-4802					43099.8				
Demin Plant/Condensate Plant	-533				-509.9		365.3			
Utility water/Fire Water	-182									
Waste Water Treatment	-301				-56.7			-186		
Buildings and Misc loads	-500									
Offsites & Utilities Total	-6318	0.0	0.0	0.0	-566.6	43100	365.3	-185.9		
Grand Total	**356114**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**	**0.0**		

amec foster wheeler	**EQUIPMENT LIST FOR COMPRESSORS**	**Rev.**	**REV 01** / **REV 02** / **REV 03**	
	Client: Department for Business, Energy & Industrial Strategy **Contract No:** 13333	**Originated**	RR	**SHEET** 1 **of** 9
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**Checked**	SF	
		Approved	TT	
	Case: Biomass IGCC with CO2 Capture	**Date**	29/06/2017	

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
700	K-001	CO2 Compressor Package (8 stages)	Multi-Stage Integrally Geared	2x50% (1/train)			1.287 / 4.925	78.9	1.1 / 80.0		0.992 / 0.297	11637	CrNi alloy	44.01	178.1 t/h CO2	
							/		/		/					
							/		/		/					
							/		/		/					
							/		/		/					
							/		/		/					

Notes:

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EQUIPMENT LIST FOR HEAT EXCHANGERS

Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333

Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Case: Biomass IGCC with CO2 Capture

Rev.	REV 01	REV 02	REV 03
Originated	RR		
Checked	SF		
Approved	TT		
Date	29/06/2017		

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
400	E-401	CO Shift HP Steam Generator	Kettle	2x50% (1/train)			H:378 t/h C: 31 t/h	16.33	657	355.0 / 134	475.0 / 42	SS316	SS316					
400	E-402	CO Shift MP Steam Generator 1	Kettle	2x50% (1/train)			H:378 t/h C: 25 t/h	13.89	732	285.3 / 51	395.0 / 42	SS316	SS316					
400	E-403	CO Shift LP Steam Generator	Kettle	2x50% (1/train)			H:378 t/h C: 9.2 t/h	5.18	278	195.0 / 10	325.0 42	SS316	SS316					
400	E-404	CO Shift MP Steam Generator 2	Kettle	2x50% (1/train)			H:378 t/h C: 8.3 t/h	4.73	528	285.0 / 51	315.0 / 42	SS316	SS316					
400	E-405	CO Shift Gas-Gas exchanger	Shell & Tube	2x50% (1/train)			H:378 t/h C: 239.4 t/h	7.24	8969	285.0 / 43	291.0 / 42	1.25CR-0.5Mo	1.25CR-0.5Mo					
400	E-406	Satrator Heater	Shell & Tube	2x50% (1/train)			H:378 t/h C: 721 t/h	27.27	18192	225.0 / 46	254.0 / 42	CS	CS					
400	E-407	AGR Process Reboiler	Shell & Tube	2x50% (1/train)			H:342 t/h	64.65	9561	150.0 /	198.0 / 42	1.25CR-0.5Mo	1.25CR-0.5Mo					
400	E-408	N2 Heater	Shell & Tube	2x50% (1/train)			H:247 t/h C: 239.2 t/h	6.23	4738	142.0 / 35	159.0 / 42	SS316	SS316					
400	E-409	Syngas Heater	Shell & Tube	2x50% (1/train)			H:247 t/h C:47 t/h	9.08	9421	135.0 / 35	145.0 / 42	SS316	SS316					
400	E-410	Syngas Cooler	Shell & Tube	2x50% (1/train)			H:247 t/h C: 1174 t/h	14.98	2100	50.0 / 6	134.0 / 42	CS	CS					
400	E-411	Blowdown cooler	Shell & Tube	2x50% (1/train)			H:1.9 t/h C: 24.3 t/h	0.31	5	50.0 / 6	195.0 / 9.3	CS	CS					
400	E-412	Process Water cooler	Shell & Tube	2x50% (1/train)			H:90 t/h C: 1217 t/h	15.53	231	50.0 / 6	195.0 / 37	CS	CS					
700	E-701	CO2 Compressor Cooler - Stage 1	Shell & Tube	2x50% (1/train)			H:39.3 t/h C: 17.3 t/h	0.22	65	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
700	E-702	CO2 Compressor Cooler - Stage 2	Shell & Tube	2x50% (1/train)			H:39.3 t/h C: 37.6 t/h	0.48	89	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
700	E-703	CO2 Compressor Cooler - Stage 3	Shell & Tube	2x50% (1/train)			H:137 t/h C: 59.1 t/h	0.75	180	tubeside 50.0 / 10.0	160.0 / 10	SS304	SS304					
700	E-704	CO2 Compressor Cooler - Stage 4	Shell & Tube	2x50% (1/train)			H:137 t/h C: 87.7 t/h	1.12	199	tubeside 50.0 / 10.0	160.0 / 70	SS304	SS304					
700	E-705	CO2 Compressor Cooler - Stage 5	Shell & Tube	2x50% (1/train)			H:198.5 t/h C: 224 t/h	3.12	367	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-706	CO2 Compressor Cooler - Stage 6	Shell & Tube	2x50% (1/train)			H:198.4 t/h C: 290 t/h	3.70	422	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-707	CO2 Compressor Cooler - Stage 7	Shell & Tube	2x50% (1/train)			H:178.3 t/h C: 179.8 t/h	2.29	349	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-708	CO2 Compressor Cooler - Stage 8	Shell & Tube	2x50% (1/train)			H:178.3 t/h C: 601.6 t/h	7.68	1071	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
700	E-709	CO2 Pump Cooler	Shell & Tube	2x50% (1/train)			H:178.3 t/h C: 192 t/h	2.45	590	tubeside 50.0 / 10.0	100.0 / 115	SS304	SS304					
1000	E-1001	Steam Condenser (Water Cooled)	Shell & Tube	2x50% (1/train)			H:213.3 t/h C: 9873 t/h	125.87	10363	tubeside 50.0 / 10.0	100.0 / 5	SS304	SS304					

Notes: **Notes:**

1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

amec foster wheeler

EQUIPMENT LIST FOR PUMPS		**Rev.**	**REV 01**	**REV 02**	**REV 03**
Client Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Originated**	RR		
Description Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology		**Checked**	SF		
		Approved	TT		
Case: Biomass IGCC with CO2 Capture		**Date**	29/06/2017		

UNIT NUMBER	EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP °C	SG	VISC'Y cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
1000		Vacumn Condensate return pump	Centrifugal	2x100%		213	214	75	12		31.0	1	0.200	Design Temp: 80/-10 Design Press.: 14/FV	95	CS	One operating; One spare	
1100		Process Cooling Water Pumps (closed loop)	Centrifugal	3/train	2x50% duty, 1 spare	5839	5855	75	3		25.0	1	0.900	Design Temp: 80/-10 Design Press.:9.5	651	CS	Two operating; One spare Per train	
1100		Steam Condenser Cooling Water Pumps	Centrifugal	3x50%	2x50% duty, 1 spare	4936	4950	75	3		25.0	1	0.900	Design Temp: 80/-10 Design Press.:9.5	550	CS	Two operating; One spare	
1100		Process Water Pump	Centrifugal	2x50%		60	67	75	4		170.0	1	0.900	Design Temp: 200/-10 Design Press.:15	10	CS		
1100		Demin Water Pump	Centrifugal	2x100%		183	183	75	7		20.0	1	1.000	Design Temp- 80 / -10 Design Pressure: 5.20	48	Stainless Steel casing and impeller	One operating; One spare	
1100		Firewater Pump Package	Centrifugal	3							20.0			Design Temp- 80 / -10 Design Pressure: 19		CS	Sizing: 8 x 4 x 3.2 1 x diesel, 1 x electric and	
1100		Cooling Tower Makeup Pump	Centrifugal	2			825	75							104		One operating; One spare;	
1100		Raw Water Pump Pump	Centrifugal	2			313	75							62		One operating; One spare;	
900		HP BFW Pump	Centrifugal	3x50%	Centrifugal	536	596	75	127.00		170	0.9	0.2		2814	CS	Two operating; One spare	
900		MP BFW Pump	Centrifugal	2x100%	Centrifugal	338	374	75	39.00		170	0.9	0.2		542	CS	One operating; One spare	
900		LP BFW Pump	Centrifugal	2x100%	Centrifugal	155	173	75	2.00		170	0.9	0.2		75	CS	One operating; One spare	
700		CO2 Pump	Centrifugal	2x100%	Centrifugal	178		75	30.00		30	0.5	0.04		387	SS316L	Liquid CO2	
400		Saturator Circulating Water Pump	Centrifugal	2x100%	Centrifugal	721	792	75	5.00		145	0.9	0.2		125	CS		

Notes:

		EQUIPMENT LIST FOR VESSELS		Rev.	REV 01	REV 02	REV 03	
amec foster wheeler	Client:	Department for Business, Energy & Idustrial Strategy	Contract No: 13333	Originated	RR			SHEET 4 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		Checked	SF			
		UK Carbon Capture Technology		Approved	TT			
	Case:	Biomass IGCC with CO2 Capture		Date	29/06/2017			

EQUIPMENT NUMBER		DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	PRESS barg	VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	INTERNALS MAT./LINING/ CA	REMARKS	REV
400	V-401	Condensate Seperator	Vertical drum	2x50% (1/train)	2.46	4.92	27.28	V	80	40	1.013	Wire Mesh Pad 0.48 100	CS	CS		
400	V-402	Condensate Seperator	Vertical drum	2x50% (1/train)	3.42	6.84	73.31	V	80	40	1.013	Wire Mesh Pad 0.92 100	CS	CS		
400	V-403	Condensate Seperator	Vertical drum	2x50% (1/train)	2.78	5.56	39.37	V	80	40	1.013	Wire Mesh Pad 0.61 100	CS	CS		
400	V-404	Condensate Seperator	Vertical drum	2x50% (1/train)	2.45	4.90	26.95	V	80	40	1.013	Wire Mesh Pad 0.47 100	CS	CS		
400	V-405	Condensate Accumulator	Drum	2x50% (1/train)	0.50	1.00	0.23	H	80	40	1.013	Wire Mesh Pad 0.02 100	CS	CS		
400	V-406	Blowdown Drum	Vertical drum	2x50% (1/train)	0.90	1.80	1.34	V	80	10	1.013	Wire Mesh Pad 0.06 100	CS	CS		
400	T-400	Syngas Saturator	Vertical Column	2x50% (1/train)	2.86	5.72	42.87	V	80	40	1.013	Wire Mesh Pad 0.64 100	CS	CS		
400	R-401	1st Shift Catalyst Reactor	Reactor	2x50% (1/train)												
400	R-402	2nd Shift Catalyst Reactor	Reactor	2x50% (1/train)												
400	R-403	3rd Shift Catalyst Reactor	Reactor	2x50% (1/train)												
700	V-701	CO2 Compressor Stage 1 KO drum	Vertical drum	2x50% (1/train)	2.75	5.50	38	V	80	4.19	1.013	Wire Mesh Pad 0.59 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-702	CO2 Compressor Stage 2 KO drum	Vertical drum	2x50% (1/train)	2.50	5.00	29	V	80	6.19	1.013	Wire Mesh Pad 0.49 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-703	CO2 Compressor Stage 3 KO drum	Vertical drum	2x50% (1/train)	2.55	4.80	29	V	80	8.69	1.013	Wire Mesh Pad 0.51 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-704	CO2 Compressor Stage 4 KO drum	Vertical drum	2x50% (1/train)	2.25	4.00	19	V	80	17.39	1.013	Wire Mesh Pad 0.40 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		
700	V-705	CO2 Compressor Stage 5 KO drum	Vertical drum	2x50% (1/train)	1.85	3.20	10	V	80	34.69	1.013	Wire Mesh Pad 0.27 100	CS with 3mm min 304L cladding	CS with 3mm min 304L cladding		

Notes:

1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

amec foster wheeler	EQUIPMENT LIST FOR TANKS				Rev.	REV 01	REV 02	REV 03				
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	SF			SHEET	5	of	9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR						
		UK Carbon Capture Technology			Approved	TT						
	Case:	Biomass IGCC with CO2 Capture			Date	03/02/2017						

UNIT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS ID m	HEIGHT m	TOTAL VOLUME m^3	ROOF TYPE	BOTTOM TYPE	STORAGE PRESS Bar	TEMP	SG	CORRO-SION	'MATERIALS	MANWAYS SHELL	ROOF	CONNECT-IONS	HEATING COIL	INSUL-ATION	REMARKS	REV
1100	Raw Water tank	1	29.7	11.9	7504	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 24 hr Storage	
1100	Demin Water Tank	1	30.8	12.3	7622	Storage tank		Atm	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. Atm 25 hr Storage	
1100	Firewater Storage Tank	1	13.0	7.8	1035	Storage tank		Atm	20			Lined CS						Design Pres.: 0.0075 / -0.0025 Design Temp: 80/-10 deg C	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT				Rev.	REV 01	REV 02	REV 03				
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	RR			SHEET	6	of	9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	SF						
		UK Carbon Capture Technology			Approved	TT						
	Case:	Biomass IGCC with CO2 Capture			Date	29/06/2017						

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
200	Biomass Milling (3 x 50%)	3 x 2700 t/d Torrefied Biomass, As Received basis				2 operating, one spare	
200	**Shell Torrefied Biomass Gasification Package (2 x 50%)** Biomass pressurization & feeding Shell Gasifiers Syngas coolers Slag removal system Dry Fly Ash removal system Wet Scrubbing Primary water treatment Nitrogen + Blowback systems Flare headers and fuel distribution systems cooling water systems Process water systems Steam/Condensate systema Plant/Instrument air systems NaOH/HCl distribution systems	2 x 2700 t/d coal to each Gasifier, As Received basis				Two Gasification package	
300	Air Separation Unit (ASU) (2 x 50%)	2 x 58 t/h 95% pure O2	Cryogenic	25 MW each		Two ASU package	
300	N2 Compressors package (2 x 50%)	Multi-Stage with intercoolers 2 x 240 t/hr of N2 191391 Nm3/hr 8749 Am3/hr	Electric motor driven, Centrifugal	10.51 MW for each trains		Two packages Outlet Pressure : 32 bara	
300	LOX (Liquid Oxygen) Storage Tank with Vaporiser	1 x 100% (1 for both trains) 438 Am3 of O2 460 t of O2	Fixed Roof Storage Tank			To enhance the ASU Reliability Operating Pressure : 5 bara Operating Temp: - 165 oC 8 hour of storage for 1 Gasification Train	
300	LIN (Liquid Nitrogen) Storage Tank with Vaporiser	1 x 100% (1 for both trains) 231 Am3 of N2 168 t of N2	Fixed Roof Storage Tank			Operating Pressure : 5 bara Operating Temp: - 180 oC 8 hour of storage for 1 Gasification Train & 4 min of N2 for GT	
500	Acid Gas Removal Unit (Selexol)	2 x 50% Feed gas : 238310 Nm3/hr per train Operating Pressure: 37 barg	Selexol Process	5.53 MW Each Train		CO2 & H2S removal Total CO2 removal: 4274 t/d; Total Carbon Capture : 90%	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR PACKAGE EQUIPMENT			Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Industrial Strategy	Contract No: 13333	Originated	RR			SHEET 7 of 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)		Checked	SF			
		UK Carbon Capture Technology		Approved	TT			
	Case:	Biomass IGCC with CO2 Capture		Date	29/06/2017			

UNIT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
800	Gas Turbine & Generator Package	GE 9F Syngas Variant Gas Turbine	1 x 100%								303 MW Output Turbine generator				
900	HRSG		1 x 100%											Horizontal, Natural Draft 3 Pressure Level	
900	Phosphate Injection		1 x 100%												
900	Oxygen Scavanger Injection Package		1 x 100%												
1000	Steam Turbine & Generator Package		1 x 100%								200 MW Output Turbine generator				
1100	Cooling Tower	Evaporative, Natural Drive Cooling Tower	1 x 100%								Total Heat duty 540 MWth			Diameter: 100 m Height:120 m	
1100	Cooling Tower packages						4169							Filtration Package; Hypochlorite Dosing Package; Antiscalant Package	
1100	Nitrogen Package	Tank/pump	1						7	Design Temp: 80/-10 deg C Design Press.: 14 barg Oper. Temp.: 20 deg C		CS			
1100	Compressed Air Package	Compressor	1							Design Temp- 80 / -10 deg C Design Pressure: 8.70 barg					
700	TEG Dehydration Package		1 x 100%							Design Temp: 150/-10 deg C Design Press.: 76 barg				Package dehydrates 178 t/h CO2, removes 180 kg/h water to get to spec of 500 ppm	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR SOLIDS HANDLING EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333	**Originated**	RR			**SHEET** 8 **of** 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	SF			
		UK Carbon Capture Technology			**Approved**	TT			
	Case:	Biomass IGCC with CO2 Capture			**Date**	29/06/2017			

UNIT NUMBER	DESCRIPTION	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
100	**Torrefied Biomass Handling System**	113 t/h Biomass feed per train				30 Days storage: Storage piles = 2 x 81200 tonnes	
100	**kaolin clay Handling System:**	kaolin clay feed 2 t/hr per train				30 Days storage: Storage piles = 2 x 1500 tonnes	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:**	13333	**Originated**	RR			**SHEET** 9 **of** 9
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			**Checked**	SF			
		UK Carbon Capture Technology			**Approved**	TT			
	Case:	Biomass IGCC with CO2 Capture			**Date**	29/06/2017			

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
400	Mercury Adsorber			Sulfur-impregnated activated carbon beds				
900	Stack	1 x 100%					1590 tph flue gas @ 80 oC and 1.02 bara per trin	

Notes:

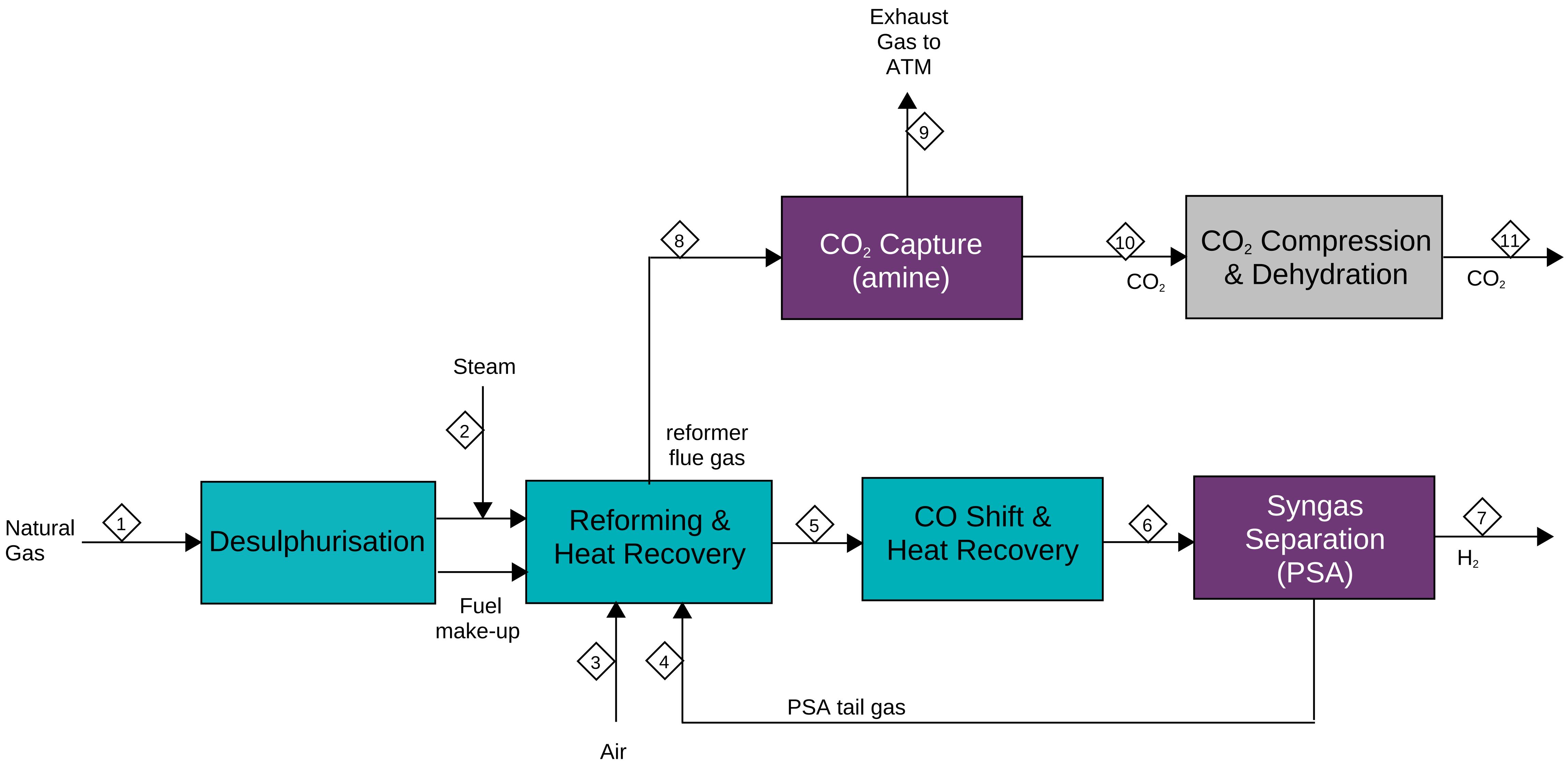
Amec Foster Wheeler	**Prepared By :**	**K.D. Nelson**
Client : BEIS	**Base Date :**	**1Q2017**
Project : Novel Carbon Capture Technology Study	**Rev. No. :**	**'6'**
Contract No.: 13333	**Print Date :**	**14-Dec-17**
Case 10 : IGCC Pre-combustion Capture for Power Generation on Biomass		

Description	Case 10 : IGCC Pre-combustion Capture for Power Generation on Biomass								
	Unit 100	Unit 200	Unit 300	Unit 400	Unit 500	Unit 700	Unit 800-1000	Unit 1100	
	Biomass Handling & Storage	Shell Torrefied Biomass Gasification Island	ASU	Syngas Treatment & Sour Water System	AGR	CO_2 Compression Block	Combined Cycle Block	Utility Units	Total
	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP	Cost GBP
Sub-Total Direct Materials	25,200,000	205,152,000	58,300,600	49,587,400	34,000,000	26,157,000	184,353,000	116,550,000	699,300,000
Other Material Costs									
Shipping / Freight	1,260,000	10,258,000	2,915,000	2,479,000	1,700,000	1,308,000	9,218,000	5,828,000	34,966,000
Third Party Inspection	252,000	2,052,000	583,000	496,000	340,000	262,000	1,844,000	1,166,000	6,995,000
Spare Parts (Comm/2yrs Op)	504,000	4,103,000	1,166,000	992,000	680,000	523,000	3,687,000	2,331,000	13,986,000
Sub-Total Materials	27,216,000	221,565,000	62,964,600	53,554,400	36,720,000	28,250,000	199,102,000	125,875,000	755,247,000
Material & Labour Contracts									
Civils/Steelwork & Buildings	6,300,000	65,649,000	14,575,000	7,438,000	4,080,000	3,924,000	27,653,000	23,310,000	152,929,000
Sub-Total Material & Labour Contracts	6,300,000	65,649,000	14,575,000	7,438,000	4,080,000	3,924,000	27,653,000	23,310,000	152,929,000
Labour Only Contracts									
Mechanical	2,520,000	51,288,000	17,490,000	8,926,000	4,080,000	4,708,000	33,184,000	18,648,000	140,844,000
Electrical/Instrumentation	1,008,000	14,361,000	8,745,000	2,479,000	1,020,000	1,308,000	9,218,000	5,828,000	43,967,000
Scaffolding/Lagging/Rigging	423,000	7,878,000	3,148,000	1,369,000	612,000	722,000	5,088,000	2,937,000	22,177,000
Sub-Total Labour Only Contracts	3,951,000	73,527,000	29,383,000	12,774,000	5,712,000	6,738,000	47,490,000	27,413,000	206,988,000
Sub-Total Materials & Labour	37,467,000	360,741,000	106,922,600	73,766,400	46,512,000	38,912,000	274,245,000	176,598,000	1,115,164,000
EPCm Costs									
Engineering Services/Construction Management	5,620,000	54,111,000	16,038,000	11,065,000	6,977,000	5,837,000	41,137,000	26,490,000	167,275,000
Commissioning	749,000	7,215,000	2,138,000	1,475,000	930,000	778,000	5,485,000	3,532,000	22,302,000
Sub-Total EPCm Costs	6,369,000	61,326,000	18,176,000	12,540,000	7,907,000	6,615,000	46,622,000	30,022,000	189,577,000
Total EPC Cost	43,836,000	422,067,000	125,098,600	86,306,400	54,419,000	45,527,000	320,867,000	206,620,000	1,304,741,000
Other Costs									
Pre-Licensing, Technical and Design etc	438,000	4,221,000	1,251,000	863,000	544,000	455,000	3,209,000	2,066,000	13,047,000
Regulatory, Licensing and Public Enquiry etc	917,000	8,830,000	2,617,000	1,806,000	1,138,000	952,000	6,713,000	4,323,000	27,296,000
Infrastructure Connection Costs								29,000,000	29,000,000
Owners Costs	3,069,000	29,545,000	8,757,000	6,041,000	3,809,000	3,187,000	22,461,000	14,463,000	91,332,000
Sub-Total Other Costs	4,424,000	42,596,000	12,625,000	8,710,000	5,491,000	4,594,000	32,383,000	49,852,000	160,675,000
Total Project Costs	**48,260,000**	**464,663,000**	**137,723,600**	**95,016,400**	**59,910,000**	**50,121,000**	**353,250,000**	**256,472,000**	**1,465,416,000**

ATTACHMENT 13: Case 11 – SMR with Post-Combustion Capture[11]

- Block Flow Diagram
- Heat & Material Balance
- Utility Summary
- Equipment List

[11] Please note, as the basis of design for Benchmark 11 is based on Shell Cansolv's proprietary design, a capital cost estimate is not included to maintain confidentiality.



NOTES

REV.	DATE	ORIG.	CHK'D	APPR.
01	18/07/17	SF	RR	TT

Department for Business, Energy & Industrial Strategy

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Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Contract No. 13333

Block Flow Diagram
Case 11 - Natural Gas SMR Case

Drawing No

Rev

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HEAT & MATERIAL BALANCE - Case 11 Natural Gas SMR

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**	**SHEET**
CONTRACT:	13333	**DATE**	05/10/2017	1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S. Ferguson	
		APP. BY	T. Tarrant	

Stream Number	1	2	3	4	5	6	7	8	9	10	11
Stream Name	B/L NG	MP Steam to SMR	Air to Reformer	PSA Tail Gas to Reformer	Raw Syngas from Reformer	Shifted Syngas to PSA	Product Hydrogen	Reformer Flue Gas from Fan	Exhaust to Atmosphere	CO2 to compression	Product CO2
Temperature (°C)	9	322	9	40	860	40	40	142	80	49	30
Pressure (bar abs)	70.00	31.70	1.01	1.30	27.70	24.40	24.20	1.15	1.01	1.99	110.00
Mass rate (kg/h)	34578	82708	307579	66869	111377	75863	8994	380355	266690	84599	82550
Molar rate (kmol/h)	1919	4588	10659	2372	8801	6833	4461	13057	9453	1990	1876
Component											
Hydrogen (mol%)	0.00	0.00	0.00	25.65	48.13	74.19	100.00	0.00	0.00	0.00	0.00
Nitrogen (mol%)	0.89	0.00	77.31	0.60	0.16	0.21	0.00	63.25	87.36	0.00	0.00
CO2 (mol%)	2.00	0.00	0.03	55.50	5.51	19.27	0.00	15.96	2.20	94.25	100.00
Methane (mol%)	89.00	0.00	0.00	16.13	4.35	5.60	0.00	0.00	0.00	0.00	0.00
Ethane (mol%)	7.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Propane (mol%)	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-Butane (mol%)	0.10	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
n-Pentane (mol%)	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen (mol%)	0.00	0.00	20.74	0.00	0.00	0.00	0.00	3.41	4.71	0.00	0.00
Argon (mol%)	0.00	0.00	0.93	0.00	0.00	0.00	0.00	0.76	1.05	0.00	0.00
CO (mol%)	0.00	0.00	0.00	1.11	9.77	0.39	0.00	0.00	0.00	0.00	0.00
Water (mol%)	0.00	100.00	0.99	1.01	32.07	0.35	0.00	16.62	4.68	5.75	0.00
Total	100	100	100	100	100	100	100	100	100	100	100

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UTILITIES BALANCE SUMMARY - Case 11 - Natural Gas SMR

CLIENT:	Department for Business, Energy & Industrial Strategy	**CHANGE**	**REV - O1**					**SHEET**
CONTRACT:	13333	**DATE**	05/10/2017					1 OF 1
NAME:	Assessing the Cost Reduction Potential & Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	**ORIG. BY**	S Ferguson					
		APP. BY	T Tarrant					

UNIT No.	**DESCRIPTION**	**Electric Oper. Load**	**Natural Gas LHV**	**Steam (t/h)**			**Condensate (t/h)**			**Cooling water**	**Raw water**	**REV**
		MW	**MW**	**HPS 171 barg**	**MPS 31 barg**	**LPS 3.4 barg**	**HP T/h**	**MP T/h**	**LP T/h**	**T/h**	**T/h**	
Natural Gas Reforming & Shift												
	Reformer Burners	-0.09	-65.9									
	Reformer Exhaust Gas Fan	-2.11										
	Feeds to Reformer	0.0	-370.1		-82.7							
	Hydrogen Recycle Compressor	-0.01								0.0		
	Syngas Cooling & Heat Integration	0.0			191.0				32.4	-1244.2	-47.3	
	Steam Turbine	12.5			-111.6	103.0			111.6	-398.0		
	Natural Gas Reforming & Shift Net	10.3			-3.4					-1642.2	-47.3	
CO2 Capture												
	Pumps	-0.9										
	Thermal Reclaimer Reboiler	-0.3										
	CO2 Capture Unit Coolers	0.0								-7640.0		
	Solvent Reboiler	0.0				-103.0			103.0			
	CO2 Capture Net	-1.2								-7640.0		
CO2 Compression												
	CO2 Compressor	-6.1										
	CO2 Pumps	-0.2										
	CO2 Inter/After Coolers	0.0								-1016.8		
	CO2 Compression Net	-6.3								-1016.8	1.7	
Utilities & Offistes Systems												
	Cooling Water System	-1.8								10299.0	-149.4	
	Water Units	-0.1									195.0	
	Buildings and Misc extra loads	-0.9										
	Effluent Treatment	-0.1										
	U&O Net	-2.9										
Overall Performance												
	Gross Production	12.5	0	0.0	0.0	0.0	0.0	0.0	0.0	10299.0	195.0	
	Total Parasitic Loads	-12.5	-436	0.0	0.0	0.0	0.0	0.0	0.0	-10299.0	-195.0	
	Net Produced	0.0	0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	

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	Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333	Originated	SF			SHEET 1 of 7
	Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology	Checked	RR			
		Approved	TT			
	Case: Natural Gas Steam Methane Reformer with Post Combustion CO2 Capture	Date	06/10/2017			

EQUIPMENT NUMBER	DESCRIPTION	COMPRESSOR TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	ACTUAL CAPACITY m^3/hr	Cp/Cv INLET/ OUTLET	DIFF. PRESS. bar	PRESSURE INLET/OUTLET bara bara	TURB.DRIVE STEAM PRESS. barg	COMPRESSIBILITY INLET/OUTLET	POWER EST/RATED kW	MATERIAL CASING	MOLECULAR WEIGHT	REMARKS	REV
K-001	CO2 Compressor Package (4 stages)	Multi-Stage Integrally Geared	2 x 50%		14,591	1.284 / 1.598	66.0	2.0 / 68.0		0.991 / 0.825	3357	CrNi alloy	42.80	45 t/h CO2	
						/		/		/					
						/		/		/					
						/		/		/					
						/		/		/					
						/		/		/					

Notes:

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EQUIPMENT LIST FOR HEAT EXCHANGERS

Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333
Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology
Case: Natural Gas Steam Methane Reformer with Post Combustion CO2 Capture

Rev.	REV 01	REV 02	REV 03
Originated	SF		
Checked	RR		
Approved	TT		
Date	06/10/2017		

EQUIPMENT NUMBER	DESCRIPTION	EXCHANGER TYPE(1)/ SUB-TYPE	No.off x DUTY %	No.OF SHELLS/TUBES (ST)	TEMA TYPE(ST)/ HEADER CONST(AC) (2)	RATE(3)	DUTY MW	HEAT T'FER AREA(6) m^2	DESIGN CONDITIONS COLDSIDE(4) TEMP/PRESS °C / barg	DESIGN CONDITIONS HOTSIDE TEMP/PRESS °C / barg	MATERIAL PLATE/ SHELL	MATERIAL TUBE(ST/AC) HEAD(AC)	No.OF BAYS/FANS (AC)	FAN TYPE (5)	TOTAL FAN POWER kW	REMARKS	REV
E-001	DCC Cooler	Gasketed Plate and frame	1						51.0 / 10.0	80.0 / 7	SS304L	SS304L				CO2 Capture Unit Scope	
E-002	CO2 Wash Water cooler	Gasketed Plate and frame	1						80.0 / 3/FV	80.0 / 7.0	SS304L	SS304L				CO2 Capture Unit Scope	
E-003	Absorber Intercooler	Gasketed Plate and frame	1						51.0 / 10.0	140.0 / 9.5	SS304L	SS304L				CO2 Capture Unit Scope	
E-004	Lean amine cooler	Gasketed Plate and frame	1						51.0 / 10.0	140.0 / 9.5	SS304L	SS304L				CO2 Capture Unit Scope	
E-005	Lean/Rich exchangers	Gasketed Plate and frame	2						51.0 / 10.3	140.0 / 6	SS316L	SS316L				CO2 Capture Unit Scope	
E-006	Stripper condenser	Welded Plate and frame	1						51.0 / 10.0	150.0 2/FV	SS316L	SS316L				CO2 Capture Unit Scope	
E-007	Reboilers	Welded Plate and frame	2						150.0 / 2/FV	295.0 / 5/FV (tubeside)	SS316L	SS316L				CO2 Capture Unit Scope	
E-008	Treated Gas Reheater		1						105.0 / 2.000	125 / 2.0	CS	CS					
E-101	CO2 Compressor Cooler - Stage 1	Shell & Tube	2				1.76	320	tubeside 51.0 / 10.0	160.0 / 10	SS304	SS304					
E-102	CO2 Compressor Cooler - Stage 2	Shell & Tube	2				1.25	195	tubeside 51.0 / 10.0	160.0 / 12	SS304	SS304					
E-103	CO2 Compressor Cooler - Stage 3	Shell & Tube	2				1.32	181	tubeside 51.0 / 10.0	160.0 / 29	SS304	SS304					
E-104	CO2 Condenser	Shell & Tube	2				3.41	456	tubeside 51.0 / 10.0	160.0 / 70	SS304	SS304					
E-105	CO2 Product Cooler	Shell & Tube	1				0.57	179	tubeside 51.0 / 10.0	100.0 / 115	SS304	SS304					

Notes:
1. C - Condenser HE - Heat Exchanger RB - Reboiler STB - Steam Boiler 2. For Air Coolers CP - Cover Plate PT - Plug Type MT - Manifold Type BT - Billet Type
3. Rate = Total Fluid Entering Coldside And Applies To Condensers, Boilers And Heaters. 4. Coldside Design Temp Equals Design Air Temp. For Air Coolers 5. I - Induced F - Forced
6. For Air-Coolers, this is Bare Tube Area

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	Client:	Department for Business, Energy & Idustrial Strategy	Contract No: 13333	Originated	SF			SHEET 3 of 7
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology		Checked	RR			
				Approved	TT			
	Case:	Natural Gas Steam Methane Reformer with Post Combustion CO2 Capture		Date	06/10/2017			

EQUIPMENT NUMBER	DESCRIPTION	PUMP TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE (2) OP./SPARE	DESIGN CAPACITY t/h	DESIGN CAPACITY m3/hr	PUMP EFFIC'Y %	DIFF PRESSURE bar	TURB. DRIVE STEAM P barg	OPERATING CONDS TEMP / SG / VISC'Y °C cP	DESIGN TEMPERATURE /PRESSURE °C	POWER EST/RATED kW	MATERIAL CASING/ROTOR	REMARKS	REV
P-001	DCC Pump	Centrifugal	2 x 100%					6.0		30.0 1.004 0.797	Design Temp: 80/-10 Design Press.: 7.5		304L SS	**CO2 Capture Unit Scope**	
P-002	Absorber WW pump	Centrifugal	2 x 100%					3.6		30.0 1.004 0.797	Design Temp: 80/-10 Design Press.: 7		304L SS	**CO2 Capture Unit Scope**	
P-003	Absorber Intercooler Pump	Centrifugal	2 x 100%					3.5		30.0 1.004 0.797	Design Temp: 80/-10 Design Press.: 11		304L SS	**CO2 Capture Unit Scope**	
P-004	Rich amine pumps	Centrifugal	2 x 100%					12.9		36.0	Design Temp: 80/-10 Design Press.: 11		304L SS	**CO2 Capture Unit Scope**	
P-005	Stripper reflux pump	Centrifugal	2 x 100%					6.0		137.0 0.919 0.205	Design Temp: 295/-10 Design Press.: 12.4/FV		304L SS	**CO2 Capture Unit Scope**	
P-006	Lean amine pumps	Centrifugal	2 x 100%					3.9		119.0 0.95	Design Temp: 140/-10 Design Press.: 8/FV		316L SS	**CO2 Capture Unit Scope**	
P-007	Lean amine feed pumps	Centrifugal	2 x 100%					6.9		50.0 0.95	Design Temp: 140/-10 Design Press.: 8/FV		304L SS	**CO2 Capture Unit Scope**	
P-008	Amine Drain Pump	Centrifugal	1 x 100%					4		60.0 0.983	Design Temp: 100/-10 Design Press.:2.40		SS316L	**CO2 Capture Unit Scope**	
	Demin Water Pump	Pump	2					1		20.0	Design Temp- 80 / -10 Design Pressure: 5.20		Stainless Steel casing and impeller		
	Towns Water Transfer Pump	Centrifugal	2					1		20.0	Design Temp- 80 / -10 Design Pressure: 6.0		Cast iron casing with Stainless steel impeller		
	Firewater Pump Package	Pump	3							20.0	Design Temp- 80 / -10 Design Pressure: 19		CS		
	Supercritical CO2 Pump	Pump	2 x 100%		84	115.9		44		25.0 0.728	Design Temp- 80 / -10 Design Pressure: 5.0	189	CS		
	CO2 Compressor Condensate Return Pump	Pump	2 x 100%		1.7	1.7		3		25.0 1.007	Design Temp- 80 / -10 Design Pressure: 5.0	2	CS		

Notes:

EQUIPMENT LIST FOR VESSELS

Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Rev.**	**REV 01**	**REV 02**	**REV 03**
Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology		**Originated**	SF		
			Checked	RR		
			Approved	TT		
Case:	Natural Gas Steam Methane Reformer with Post Combustion CO2 Capture		**Date**	06/10/2017		

EQUIPMENT NUMBER	DESCRIPTION	VESSEL TYPE(1)/ SUB-TYPE	No.off x DUTY %	DIMENSIONS ID m	DIMENSIONS HEIGHT T/T m	TOTAL VOLUME m^3	V/H (2)	DESIGN CONDITIONS TEMP ^oC	DESIGN CONDITIONS PRESS barg	DESIGN CONDITIONS VACUUM FVPRESS bara	INTERNALS TYPE/No.OFF PACKED VOL. m^3 / PACKED HGT mm	MATERIALS OF CONST'N SHELL MAT./LINING/ CA	MATERIALS OF CONST'N INTERNALS MAT./LINING/ CA	REMARKS	REV
C-001	DCC (Direct Contact Cooler)	Rectangular column	1				V	150	0.30	1.013			SS304 packing and internals	**CO2 Capture Unit Scope**	
C-002	Absorber (absorber section)	Rectangular column	1				V	80/-10	0.10	1.013			SS304 packing and internals	**CO2 Capture Unit Scope**	
C-003	Absorber (water wash section)	Rectangular column	1				V	80/-10	0.10	1.013			SS304 packing and internals	**CO2 Capture Unit Scope**	
C-004	Stripper	Vertical cylinder	1				V	250/-10	3.50	FV				**CO2 Capture Unit Scope**	
V-002	CO2 Reflux Accumulator	Horizontal	1	3.0	4.0	23.0	H	295/-10	5.00	FV		304L SS		**CO2 Capture Unit Scope**	
V-101	CO2 Compressor Stage 1 KO drum	Vertical drum	2	1.80	3.60	9	V	80	3.50	1.013	demister	SS304	SS304	43 t/h CO2	
V-102	CO2 Compressor Stage 2 KO drum	Vertical drum	2	1.40	2.80	4	V	80	6.00	1.013	demister	SS304	SS304	43 t/h CO2	
V-103	CO2 Compressor Stage 3 KO drum	Vertical drum	2	1.10	2.20	2	V	80	10.00	1.013	demister	SS304	SS304	43 t/h CO2	
V-104	CO2 Compressor Stage 4 KO drum	Vertical drum	2	1.00	2.00	2	V	80	25.00	1.013	demister	SS304	SS304	43 t/h CO2	

Notes:

1. TW - Single Diameter Tower DDT - Double Diameter Tower HT - Horizontal Tank AT - Agitated Tank VT - Vertical Tank
2. V - Vertical H - Horizontal

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EQUIPMENT LIST FOR TANKS

Client:	Department for Business, Energy & Idustrial Strategy	**Contract No:** 13333	**Rev.**	**REV 01**	**REV 02**	**REV 03**
Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology		**Originated**	SF		
			Checked	RR		
			Approved	TT		
Case:	Natural Gas Steam Methane Reformer with Post Combustion CO2 Capture		**Date**	06/10/2017		

SHEET 5 **of** 7

EQUIPMENT NUMBER	DESCRIPTION	NO REQ'D	DIMENSIONS ID m	HEIGHT m	TOTAL VOLUME m^3	ROOF TYPE	BOTTOM TYPE	STORAGE PRESS	TEMP	SG	CORRO-SION	'MATERIALS	MANWAYS SHELL	ROOF	TYPE	HEATING COIL	INSUL-ATION	REMARKS	REV
T-001	Lean Amine Tank	1				Storage tank		0.02	30			316L SS lined CS			Vertical, sized for full inventory			**CO2 Capure Unit Scope**	
T-002	Amine Drain Tank	1				Storage tank		0.02	30			304L SS			Horizontal, underground.			**CO2 Capure Unit Scope**	
T-003	Absorbent Make-up tank	1				Storage tank		0.02	30			304L SS lined CS			Vertical			**CO2 Capure Unit Scope**	
T-006	Towns Water Storage Tank	1	5.00	5.00	100.00	Vertical cylindrical		0.0075 / -0.0025	20			Lined CS						Design Temp: 80/-10 deg C Design Pres. 0.0075 /	
T-007	Demin Water Tank	1	5.00	5.00	100.00	Tank		0.90	20			Lined CS						Design Temp: 80/-10 deg C	
T-008	Firewater Storage Tank	1	13.0	7.8		Tank			20			Lined CS						Design Pres.: 0.0075 / -0.0025 Design Temp: 80/-10	

Notes:

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EQUIPMENT LIST FOR PACKAGE EQUIPMENT

Client: Department for Business, Energy & Idustrial Strategy **Contract No:** 13333

Description: Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation) UK Carbon Capture Technology

Case: Natural Gas Steam Methane Reformer with Post Combustion CO2 Capture

Rev.	REV 01	REV 02	REV 03
Originated	SF		
Checked	RR		
Approved	TT		
Date	06/10/2017		

EQUIPMENT NUMBER	DESCRIPTION	EQUIPMENT TYPE(1)/ SUB-TYPE	No.off x DUTY %	DRIVE TYPE OP./SPARE	DIMENSIONS DIAM./HGT/ LENGTH mm	AREA mm^2	CAPACITY m^3/h	FLOW kg/hr	PRESS OPER./DIFF. barg / bar	DESIGN CONDS. TEMP/PRESS °C / barg	POWER EST/RATED kW	MATERIAL BODY/CA	COOL.TOWER WBT °C / APP °C / CWT °C (3)	REMARKS	REV
	SMR Hydrogen Production Unit - 100,000 Nm3/h Hydrogen @ 99.99% purity						**4840**	**9000**						**Steam Methane Reformer with standard Pressure Swing Adsoprtion H2 purification**	
	Steam Turbine generator package		**1**				**9580**	**111634**	**33.5 / 33.3**	**425 / 43**	**12568**			**2-stage steam turbine with condenser and generator**	
S-001	Ion Exchange Package		2				3.6	16 kg/h removal duty	5 kPag differential pressure			316L SS / Ion exchange resin 3600L of M600 resin per unit		**CO2 Capture Unit Scope**	
S-002	Thermal Reclaimer Package	Vacuum Distillation Column	1				0.4					316L SS		**CO2 Capture Unit Scope**	
S-003	CO2 Absorbent Filtration Unit	Cartridge Type Filter	1				80					304L SS		**CO2 Capture Unit Scope**	
S-004	Activated Carbon Filtration Unit	Fixed Bed Filter	1				80					304L SS		**CO2 Capture Unit Scope**	
S-003	Nitrogen Package	Tank/pump	1						7			CS			
S-004	Compressed Air Package	Compressor	1												
S-005	Cooling Tower Package		1				10300	10300						**O&U Scope**	
TEG-101	TEG Dehydration Package		2							Design Temp: 150/-10 deg C Design Press.: 76 barg				Dehydrates 180 t/h CO2, removes 31 kg/h water to get to spec of 50 ppm	
P-001	Flow metering and analyser package	Metering	1					85000		Design Temp: 150/-10 deg C Design Press.: 114 barg				Fiscal metering package	

Notes:

amec foster wheeler	EQUIPMENT LIST FOR MISCELLANEOUS EQUIPMENT				Rev.	REV 01	REV 02	REV 03	
	Client:	Department for Business, Energy & Idustrial Strategy	Contract No:	13333	Originated	SF			SHEET 7 of 7
	Description:	Assessing the Cost Reduction Potential and Competitiveness of Novel (Next Generation)			Checked	RR			
	Unit No:	UK Carbon Capture Technology			Approved	TT			
	Case:	Natural Gas Steam Methane Reformer with Post Combustion CO2 Capture			Date	06/10/2017			

EQUIPMENT NUMBER	DESCRIPTION	P&ID No.	SIZE	TYPE	ELECTRIC LOAD	MATERIAL OF CONSTR.	REMARKS	REV
	Flue gas ducting from reformer to direct contact cooler		1.7 m X 1.7 m Estimated Length 40 m	Square		CS	Flow Rate: 380 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	
	Flue gas ducting from single DCC to single absorber		1.3 m X 1.3 m Estimated Length 20 m	Square		CS	Flow Rate: 350 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 97 deg C Design/Operating Pressure: 0.30/0.10 barg	
	Flue gas ducting to stack		1.4 m X 1.4 m Estimated Length 40 m	Square		CS	Flow Rate: 264 t/h Design Temp:150/-10 deg C; Normal Op. Temp: 80 deg C Design/Operating Pressure: 0.30/0.10 barg	
	Stack							

Notes: