Programme Area: Carbon Capture and Storage

Project: Next Gen Capture Tech Benchmarking and Performance Analysis

Title: Benchmarking and Performance Analysis of Future CO2 Capture Technologies – Benchmarking Study

Abstract:

This report describes the work carried out for the Capture Benchmarking Study, and provides a techno-economic analysis of the performance of four state-of-theart power generation/CO2 capture designs (coal with post-combustion capture, coal with gasification/pre-combustion capture, coal with oxy-firing and combined cycle gas turbine with post-combustion capture). The original study provided results for 90% CO2 capture. This report now includes results for 85% and 95% capture rates. The report is partly superseded by work undertaken and reported in the Benchmark Refresh project.

Context:

This project provided ETI with an objective view of the techno-economic performance of a range of current and next generation CO2 capture technologies including pre and post combustion and oxyfuel CCS plant. The analysis that underpins these benchmarking studies was based on coal and gas fired power station designs typical of those found in the UK and considered parameters such as power station capital cost, efficiency and levelised cost of electricity (with and without CCS).

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ORIG. BY	SJ/GS/PB	SJ/GS/PB	SF/NG/GU	SF/NG/GU	SF/GS	SF
CHKD BY	TB	TB	TB	TB	TB	TB
APP. BY	TB	TB	TB	TB	TB	ME

Benchmarking and Performance Analysis of Future CO_2 Capture Technologies

Work Package 1 & 3 – Benchmarking Study

Final Report

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NOTE TO READER

This report has been prepared following two distinct yet similar work packages:

- Work Package 1 (WP1)
- Work Package 3 (WP3)

Work Package 1

Under WP1, the following cases were developed:

- Case 1A: IGCC Coal Power Plant with pre combustion CO_2 capture at a carbon capture level of 90%;
- Case 1B: IGCC Coal Power Plant without CO_2 capture;
- Case 2A: Pulverised coal power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 90%;
- Case 2B: Pulverised coal power plant without CO_2 capture;
- Case 3A: Natural gas CCGT power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 90%;
- Case 3B: Natural gas CCGT power plant without CO_2 capture;
- Case 4: Pulverised coal power plant with oxyfuel CO_2 capture at a carbon capture level of 90%;

The results and findings of this work package have been completed and have not been significantly amended during Work Package 3.

Work Package 3

Under WP3, the following cases were developed:

- Case $1A_{85\%}$: IGCC Coal Power Plant with pre combustion CO_2 capture at a carbon capture level of 85%;
- Case $1A_{95\%}$: IGCC Coal Power Plant with pre combustion CO_2 capture at a carbon capture level of 95%;
- Case $2A_{85\%}$: Pulverised coal power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 85%;
- Case $2A_{95\%}$: Pulverised coal power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 95%;
- Case $3A_{85\%}$: Natural gas CCGT power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 85%;
- Case $3A_{95\%}$: Natural gas CCGT power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 95%;

The reader is advised that the WP3 cases contained within this report are currently in the draft stage and subject to amendment. All changes relating to WP3 have been identified (for the draft report only) by blue text and a line in the left-hand margin as per this section.

The results and discussions surrounding WP3 have been integrated into the report to provide one consolidated report that considers both WP1 and WP3.

1. EXECUTIVE SUMMARY

1.1 Introduction

The Energy Technologies Institute (ETI) is a UK based company formed from global industry members - BP, Caterpillar, EDF Energy, E.ON, Rolls-Royce and Shell with the UK government to bring together projects that create affordable, reliable, clean energy for heat, power and transport. A target of the ETI is to accelerate the deployment of technologies that provide affordable, secure UK based low-carbon energy systems from 2020 to 2050 by supporting a range of energy production technology programmes. One of the ETI technology programmes is Carbon Capture and Storage (CCS).

At present fossil fuels play a key part in providing for the UK's energy demands. The development and implementation of Carbon Capture and Storage (CCS) technologies is an important option in reducing CO_2 emissions. The ETI is working along with others to ensure that investment and technology innovation is dedicated appropriately to CCS technologies, including the development and demonstration of lower cost capture technologies with a smaller environmental impact / performance.

The current range of commercially available technologies for CO_2 capture impose substantial capital cost and plant efficiency penalties on power production. The ETI has identified opportunities to bring forward capture technologies to demonstration that offer a step change in cost and / or efficiency penalty reductions over those currently available. A wide range of potential technologies exists, for example:

- Pre-combustion (e.g. novel solvents, sorbents, membranes, physical separation; H_2 firing (boiler and gas turbine); shift reactor design);

- Post-combustion (e.g. novel solvents, sorbents, membranes, physical separation; bio-fixation);

- Oxy-fuel (e.g. novel solvents, sorbents, membranes, physical separation);

- Novel combinations.

The ETI has undertaken an analysis of opportunities for application of CCS within the UK considering the current generating facilities and the likely roll out of new generating capacity over the next 20 to 30 years. The ETI is currently engaged in a consultation to identify promising technologies and hence potential ETI projects. As a key part of this prioritisation the ETI wishes to establish benchmark performances (operating performance, capital and operating costs) of currently considered "best available technologies" for the selected applications, and then assess the likely performance of the selected next generation technologies against these benchmarks.

1.2 Scope of Study

To support carbon capture technology programmes the ETI has commissioned a two phase study. The objective of the first phase – Work Package 1 (WP1) – is to establish a consistent assessment methodology and provide benchmark performances (operating performance, capital and operating costs) for agreed, full scale power plant designs in four application areas.

A future second work package (WP2) will involve performance and economic evaluations of up to six selected potential next generation technologies, using as far as possible the same basis and methodology as WP1, with resulting comparisons with the equivalent WP1 benchmarks.

This report covers the activities undertaken in support of WP1 and WP3 only.

The scope of WP1 included the evaluation of seven outline designs for coal and natural gas based power plants including:

- Case 1A - Integrated gasification combined cycle (IGCC) with pre-combustion CO_2 capture
- Case 1B - Integrated gasification combined cycle (IGCC) without CO_2 capture
- Case 2A - Ultra supercritical pulverised coal (USCPC) with post-combustion CO_2 capture
- Case 2B - Ultra supercritical pulverised coal (USCPC) without CO_2 capture
- Case 3A - Combined cycle gas turbine (CCGT) with post-combustion CO_2 capture
- Case 3B - Combined cycle gas turbine (CCGT) without CO_2 capture
- Case 4 - Oxy-fuel based ultra supercritical pulverised coal (USCPC) with purification, compression and dehydration.

In addition a number of sensitivity cases have been further evaluated:

- Case 3A(i) – modified Case 3A to consider a reduced capture rate of 75% (compared with 90% in Case 3A)
- Case 4A – modified Case 4 considering an alternate flowsheet incorporating an integrated CO_2 purifier arrangement.
- Case 4B – modified Case 4 considering an alternate flowsheet without CO_2 purifier and resulting in a higher concentration of O_2 in the product CO_2.

The scope of WP3 included the evaluation of six additional outline designs for coal and natural gas based power plants at varying levels of carbon capture, including:

- Case 1A$_{85\%}$: IGCC Coal Power Plant with pre combustion CO_2 capture at a carbon capture level of 85%;

- Case 1A$_{95\%}$: IGCC Coal Power Plant with pre combustion CO_2 capture at a carbon capture level of 95%;

- Case 2A$_{85\%}$: Pulverised coal power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 85%;

- Case 2A$_{95\%}$: Pulverised coal power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 95%;

- Case 3A$_{85\%}$: Natural gas CCGT power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 85%;

- Case 3A$_{95\%}$: Natural gas CCGT power plant with amine solvent post-combustion CO_2 capture at a carbon capture level of 95%;

This report details the evaluation, performance and results comprising WP1 and WP3 for each of the above benchmark cases.

1.3 Summary of Study Results

Figure 1-1 Summary Performance Figures

		IGCC		USCPC		Natural Gas CCGT			Oxyfuel		
Case		1A	1B	2A	2B	3A	3A(i)	3B	4	4A	4B
Capture Level		with 90% CC	no CC	with 90% CC	no CC	with 90% CC	with 75% CC	no CC	with 90% CC	with 90% CC	with 90% CC
Total gross installed capacity	MWe	923.8	878.1	745.7	836.2	955.3	970.7	1037.6	855.4	855.4	855.4
Total auxiliary loads	MWe	223.9	132.0	119.0	57.0	113.0	98.4	46.8	249.8	234.8	230.8
Net Power Export	MWe	699.9	746.1	626.7	779.2	842.3	872.3	990.8	605.6	620.6	624.6
Net Efficiency (LHV)	%	35.5	45.0	34.4	42.8	49.9	51.7	58.8	33.3	34.1	34.3
Carbon capture rate	%	90.0	0.0	90.0	0.0	90.1	75.2	0.0	90.4	90.3	90.4
Total CO_2 captured	tpd	13976	0	13124	0	7504	6265	0	13184	13177	13182
Total CO_2 emitted	tpd	1557	13150	1465	14590	825	2063	8329	1406	1412	1408
CO_2 emissions	g CO_2/ kWh_{Net}	92.7	734.4	97.4	780.2	40.8	98.5	350.2	96.7	94.8	93.9

Figure 1-2 Summary Economic Figures

		IGCC		USCPC		Natural gas CCGT			Oxyfuel
Case		1A - with 90% CC	1B - no CC	2A - with 90% CC	2B - no CC	3A – with 90% CC	3A(i) – with 75% CC	3B - no CC	4 – with 90% CC
Total CAPEX	GB£M	1556	1274	1607	1140	1024	976	544	1897
CAPEX efficiency	GB£ / kW_{Net}	2223	1708	2565	1463	1216	1118	549	3132
Total OPEX – incl. fuel	GB£M p.a.	208.2	175.8	211.5	187.3	297.7	296.0	279.6	205.9
Total OPEX – excl. fuel	GB£M p.a.	75.9	64	87.2	62.9	45.2	43.4	27	81.5
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	297.5	235.6	339.4	240.4	352.8	339.3	282.2	340.0
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	108.4	85.8	139.9	80.7	53.6	49.8	27.3	134.6
Levelised Cost of Electricity									
CO_2 emission cost = £ 0 / te CO_2	£ / MWh_{Net}	76.23	59.49	87.80	56.24	69.43	65.89	47.88	96.90
CO_2 emission cost = £ 20 / te CO_2	£ / MWh_{Net}	78.08	74.18	89.76	71.84	70.22	67.86	54.89	98.83
CO_2 emission cost = £ 40 / te CO_2	£ / MWh_{Net}	79.94	88.87	91.73	87.44	71.01	69.84	61.89	100.77
CO_2 emission cost = £ 60 / te CO_2	£ / MWh_{Net}	81.79	103.56	93.69	103.05	71.81	71.81	68.90	102.70
Cost of CO_2 Captured CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	20.12		35.58		58.00			44.82
Cost of CO_2 Avoided CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	26.08		46.28		69.38			59.48

Notes:

1. OPEX figures based on fuel costs: Coal £65/tonne, Natural Gas £0.02058/kWh.
2. OPEX figures calculated on the basis of [Load Factor x Availability] = 0.65 (Year 1), 0.75 (Year 2) and 0.85 (balance of operation)

3. 30 years operation assumed for coal fed plants, 20 years operation for natural gas fed plants.
4. Levelised cost of electricity figures are based upon a discount rate of 10%.
5. Economic Figures for 4A and 4B have not been generated.
6. Total CAPEX figures include Total Installed Cost, 5% land costs, 10% owner's costs and 25% contingency.
7. The CAPEX estimates detailed in Table 1.2 have been produced based on the specific technical definition and designs for the benchmark cases considered within this study. These may differ from those considered in similar studies prepared by others. For all of the benchmark cases Foster Wheeler has adopted a consistent estimating methodology using equipment costs and cost factors developed using cost estimating tools along with in-house data and experience. Where available estimates have been supported with data from previous work undertaken by Foster Wheeler for similar plant designs. For all of the cases any historic source estimate data has been adjusted to provide figures on a consistent and comparable 1st quarter 2009 (1Q2009) UK Basis. Estimates prepared at this level of technical definition and initial study phase of project development using the above methodology and associated qualifications / exclusions are considered to have an accuracy of +/-40%.

Figure 1-3 Graph Showing Total CAPEX of each Technology with differing % Carbon Capture

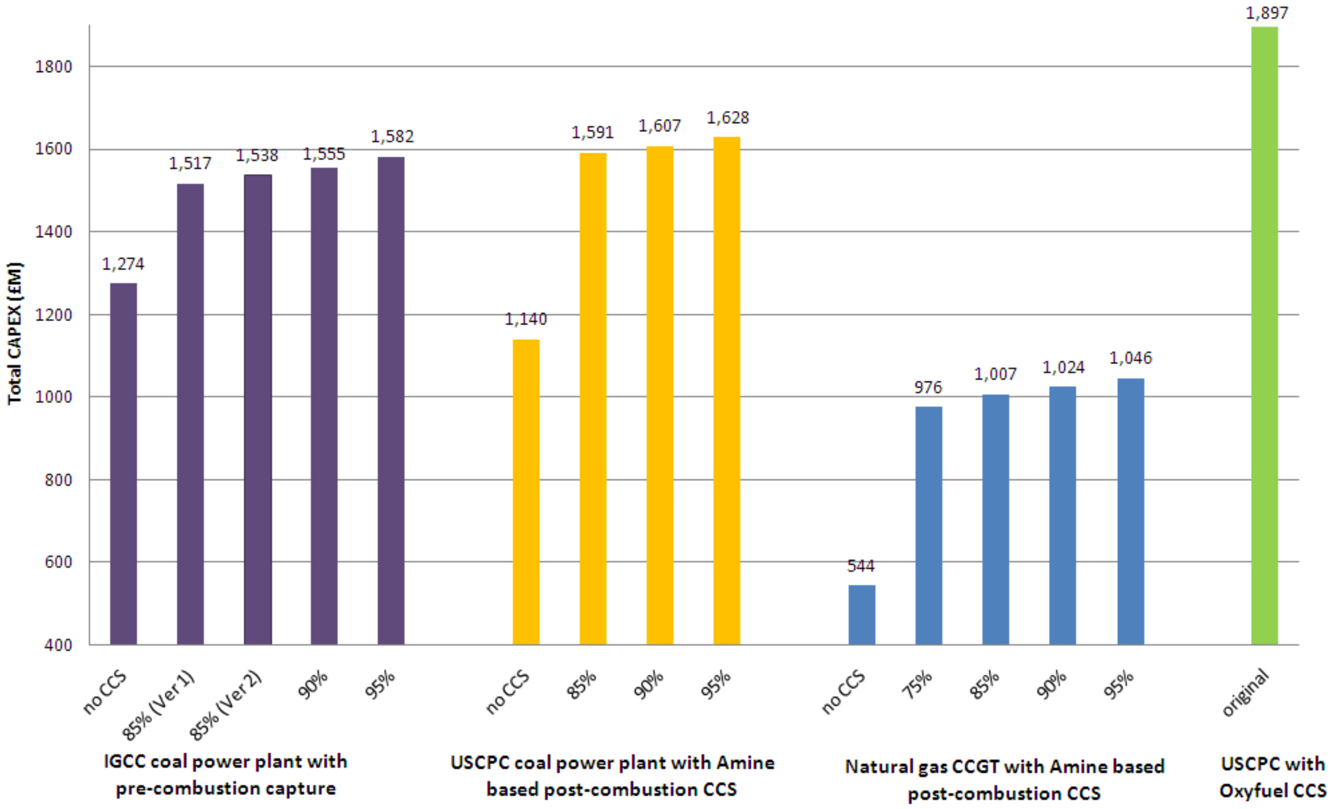


Figure 1-4 Graph Comparing Levelised Cost of Electricity for Different CO_2 Prices

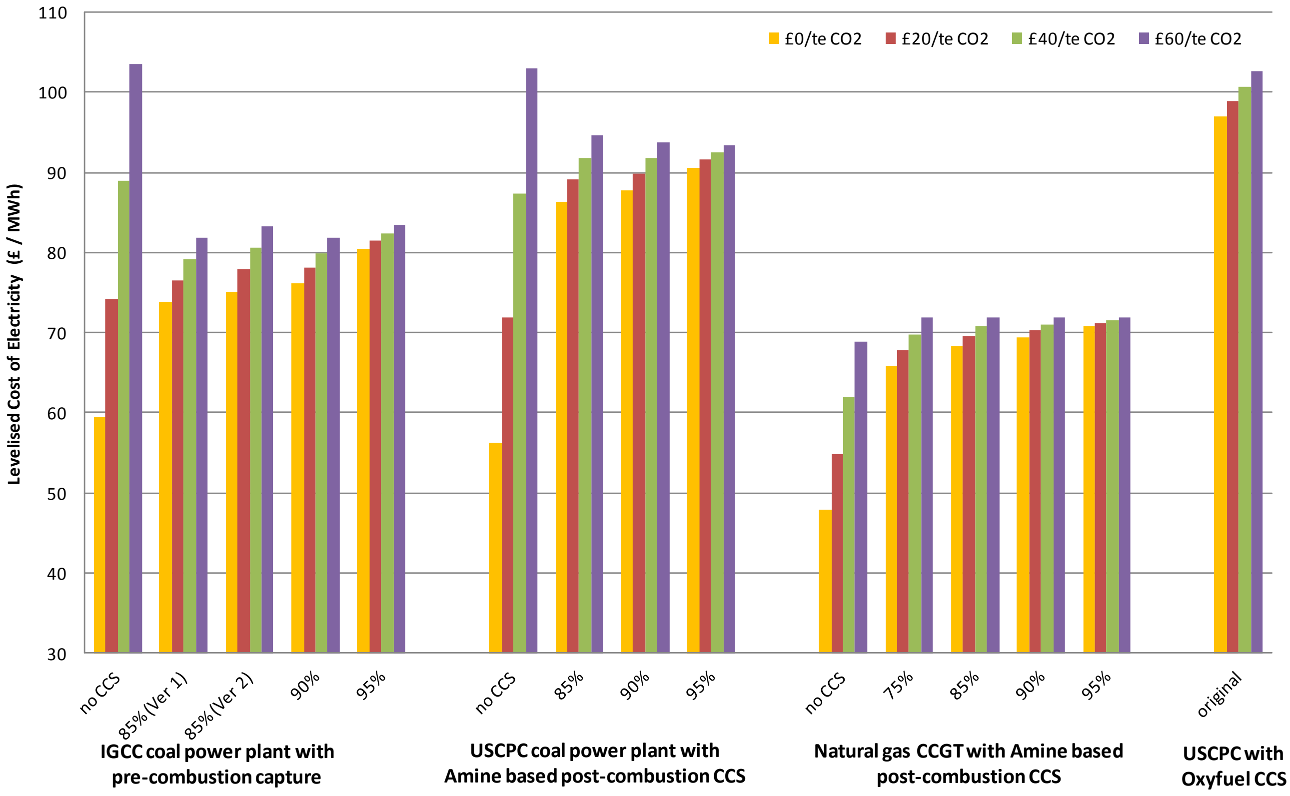


Figure 1-5 Graph Showing Net Efficiency (Based on LHV)

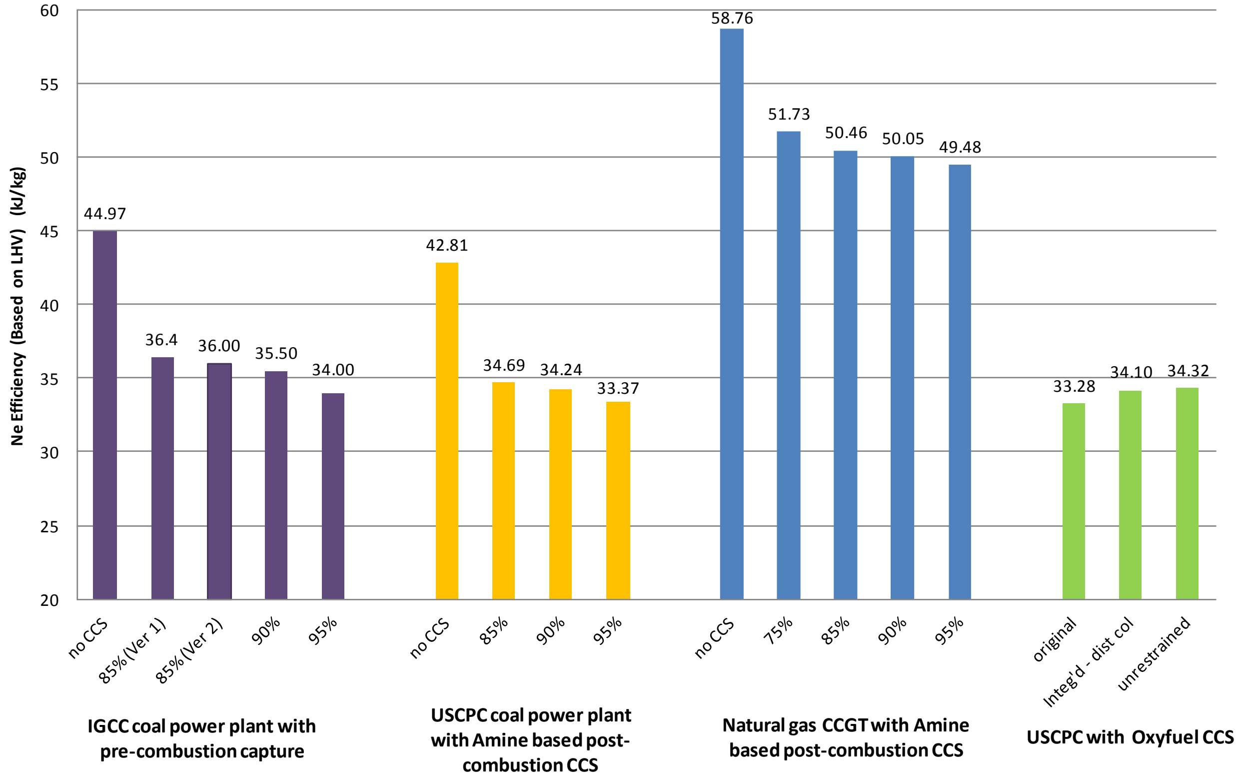
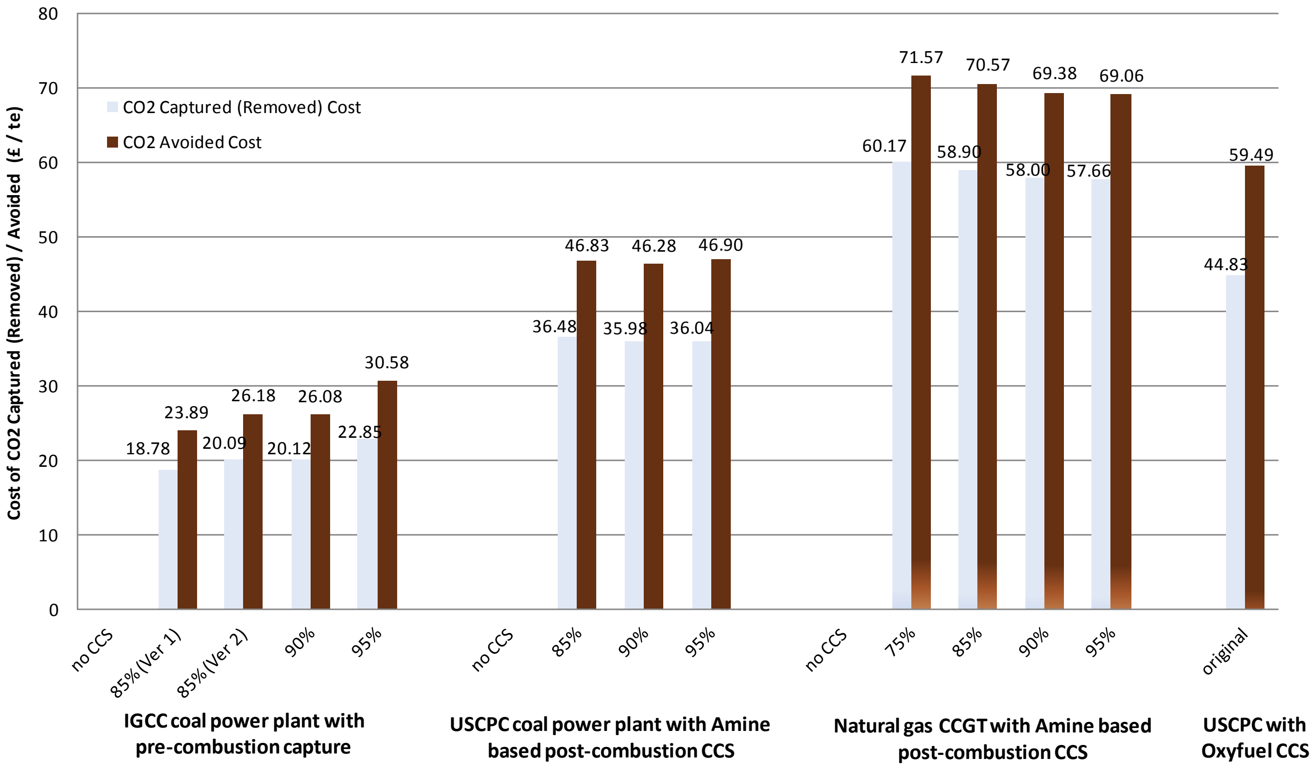


Figure 1-6 Graph Showing Cost of CO_2 Captured (Removed) and Avoided



1.4 Conclusions

The Report presents the results of Work Package 1 and Work Package 3, consisting of evaluation of selected conventional options for power generation from natural gas and bituminous coal, with a range of CO_2 capture levels (0%, 85%, 90% and 95%, plus 75% for case 3A(i)). All the cases have been developed on an impartial basis, using a common Basis of Design. The results are intended to be used as “benchmarks” of currently available technology, against which more advanced technology concepts were evaluated in Work Package 2.

The report provides a useful UK based update of power plant performance and economics. Key highlights of the performance results include:

- The calculated efficiency of the natural gas fired CCGT plant with post combustion carbon capture at 50% LHV basis is about 15 % points higher than the coal based options. The coal based carbon capture options all have efficiencies in the low to mid 30s%.
- The IGCC options have marginally the best efficiency of the coal cases at just over 35%. IGCC can be expected to gain significantly in CAPEX and OPEX when specific gas turbines become available for firing of H_2-rich fuel gases with conventional firing temperatures and with moderate exhaust NO_x. However, this important development may still be five or so years into the future.
- The oxyfuel option shows the lowest calculated efficiency. Two sensitivity cases have been considered (Case 4A and Case 4B) to focus on improvements that can be made particularly in the cryogenic purifier unit,

reducing its rather large power demand, and by relaxing the specified oxygen content of < 100 ppm in the captured CO_2.

- Varying the % CO_2 captured between 85% and 95% resulted in the expected trends of slightly increased costs and reduced power plant efficiency with increasing CO_2 capture level for all power generation schemes considered. However, there was some variation in the rate of change of these effects as a function of % CO_2 capture between the different schemes, resulting in different overall trends for the cost of CO_2 captured and avoided:
    - For the CCGT cases the cost per tonne of CO_2 captured (and avoided) decreases with increasing CO_2 capture. This is due to the relatively high additional capital cost of adding a post-combustion capture plant, operating at any capture level, to the comparatively cheap CCGT power plant, combined with the lower carbon intensity of the natural gas fuel and hence lower operating cost as a function of % CO_2 captured relative to the coal cases.
    - For the IGCC cases the cost per tonne of CO_2 captured (and avoided) increases with increasing CO_2 capture. This is due to the relatively low additional capital cost of modifying the acid gas removal process and adding CO_2 compression to the comparatively expensive IGCC power plant, combined with the higher carbon intensity of the coal feedstock and hence higher operating cost as a function of % CO_2 captured compared to the natural gas cases.
    - For the USCPC cases the cost per tonne of CO_2 captured (and avoided) remains almost constant with increasing CO_2 capture. This is due to the moderate additional capital cost of adding a post-combustion capture plant to the moderately expensive USCPC power plant, combined with the higher carbon intensity of the coal feedstock and hence higher operating cost as a function of % CO_2 captured compared to the natural gas cases.

The cost and economic figures generated are all presented on a UK, first quarter 2009 basis. They provide a useful update of earlier studies and provide a platform for WP2 evaluations. In general the results show good agreement with a number of previous studies available in the public domain in which equivalent cases have been evaluated. It is anticipated that further improvements could be made to the economics of individual cases with further refinement on a project specific basis.

2. INTRODUCTION

2.1 Objective and Approach

The study work scope is split into three Work Packages – WP1, WP2 and WP3. This report covers WP1 and WP3 activities only.

The objective of WP1 and WP3 is to establish a consistent assessment methodology and provide benchmark performances (operating performance, capital and operating costs) for agreed, full scale power plant designs in four application areas,

- Coal based pre-combustion CO_2 capture
- Coal based post-combustion CO_2 capture
- Natural gas based post-combustion CO_2 capture
- Oxy-fuel

Work Package 2 (WP2) considered the performance and economic evaluation of up to six selected potential next generation technologies, using as far as possible the same basis and methodology as WP1, with resulting comparisons with the equivalent WP1 benchmarks.

The scope of WP1 included the evaluation of seven outline designs for coal and natural gas based power plants including:

- IGCC with 90% CO_2 capture
- IGCC without CO_2 capture
- USCPC with 90% CO_2 capture
- USCPC without CO_2 capture
- CCGT with 90% CO_2 capture
- CCGT without CO_2 capture
- Oxy-fuel with CO_2 capture

In addition a number of sensitivity cases were further evaluated as part of WP1:

- Case 3A(i) – modified Case 3A to consider a reduced capture rate of 75% (compared with 90% in Case 3A)
- Case 4A – modified Case 4 considering an alternate flowsheet incorporating an integrated CO_2 purifier arrangement.

- Case 4B – modified Case 4 considering an alternate flowsheet without CO_2 purifier and resulting in a higher concentration of O_2 in the product CO_2.

After the completion of WP1 and WP2 a further benchmarking exercise; Work Package 3 (WP3), was performed in order to quantify the impact of varying levels of CO_2 capture on the performance, capital and operation costs of cases 1, 2 and 3.

The scope of WP3 included the evaluation of a further 6 outline designs based upon those developed for WP1, including:

- IGCC with 85% CO_2 capture
- IGCC with 95% CO_2 capture
- USCPC with 85% CO_2 capture
- USCPC with 95% CO_2 capture
- CCGT with 85% CO_2 capture
- CCGT with 95% CO_2 capture

No process configuration changes were made in order to move from the 90% CO_2 capture cases developed for WP1 to the 85% and 95% CO_2 capture cases developed for WP3. The change in CO_2 capture level was achieved by modifying parameters such as solvent flow rates, reboiler duties and CO_2 compression and dehydration capacity.

This report details the evaluation, performance and results for each of the above benchmark cases developed during WP1 and WP3.

2.2 Outline Case Descriptions

The following power plant configurations were evaluated in this study. Performance, capital and operating costs for each case are presented within this draft report. For Evaluation basis please refer to section 7. Key assumptions and uncertainties are highlighted as appropriate within each section.

2.2.1 Coal Based Pre-Combustion

- Case 1A – IGCC plant with CO_2 capture. Conventional pressure air separation unit (ASU), Shell gasifier, particulate removal filter, sour shift (will also accomplish COS hydrolysis), gas cooling with heat recovery, Selexol based acid gas unit for H_2S and then CO_2 removal. Multi stage integrally geared centrifugal CO_2 compression with CO_2 dehydration. Claus plant plus tail gas treating unit (TGTU). Fuel gas (H_2) dilution with nitrogen, before combustion in two GE 9F IGCC gas turbines each with associated heat recovery steam generator (HRSG) feeding a common steam turbine.

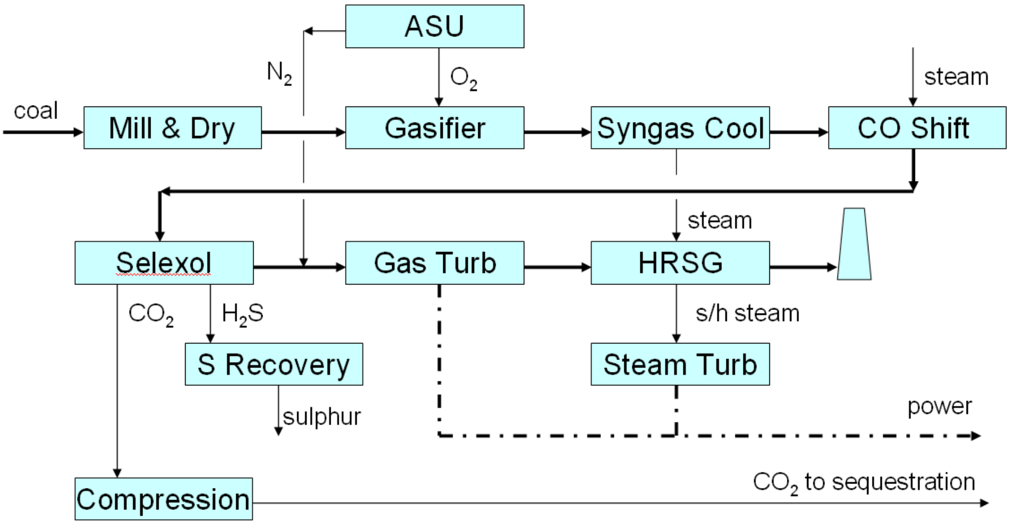
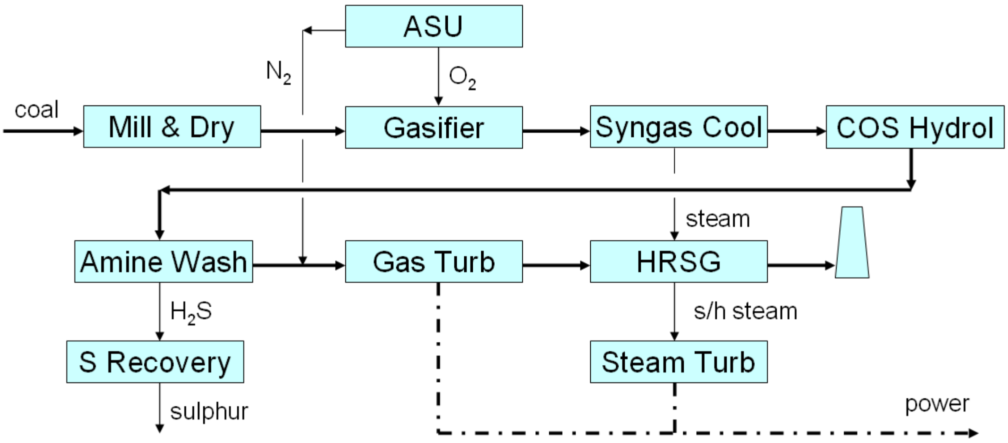


Figure 2-1 Case 1A - IGCC with Pre-Combustion CO_2 Capture

- Case 1B – IGCC plant without CO_2 capture. Conventional pressure air separation unit (ASU), Shell gasifier, particulate removal filter, COS hydrolysis, gas cooling with heat recovery, MDEA based acid gas unit for H2S removal. Claus plant plus tail gas treating unit (TGTU). Fuel gas (syngas) dilution with nitrogen, before combustion in two GE 9F IGCC gas turbines each with associated heat recovery steam generator (HRSG) feeding a common steam turbine.

Figure 2-2 Case 1B - IGCC without CO_2 Capture



2.2.2 Coal Based Post-Combustion

- Case 2A – Ultra supercritical pulverised coal (USCPC) with CO_2 capture. Single 800 MW (nominal) pulverised coal fired boiler raising steam at 275 bara, Electrostatic precipitator (ESP) for particulate removal, limestone scrubbing for SO_2 removal, selective catalytic reduction (SCR) for NOx removal, and MEA based CO_2 absorption system. Multi stage integrally geared centrifugal CO_2 compression with CO_2 dehydration.

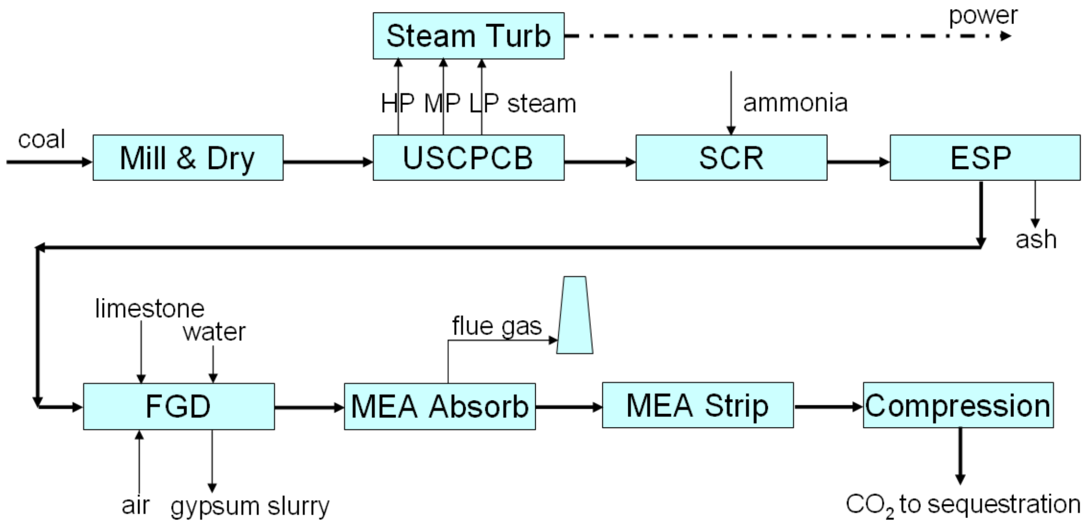
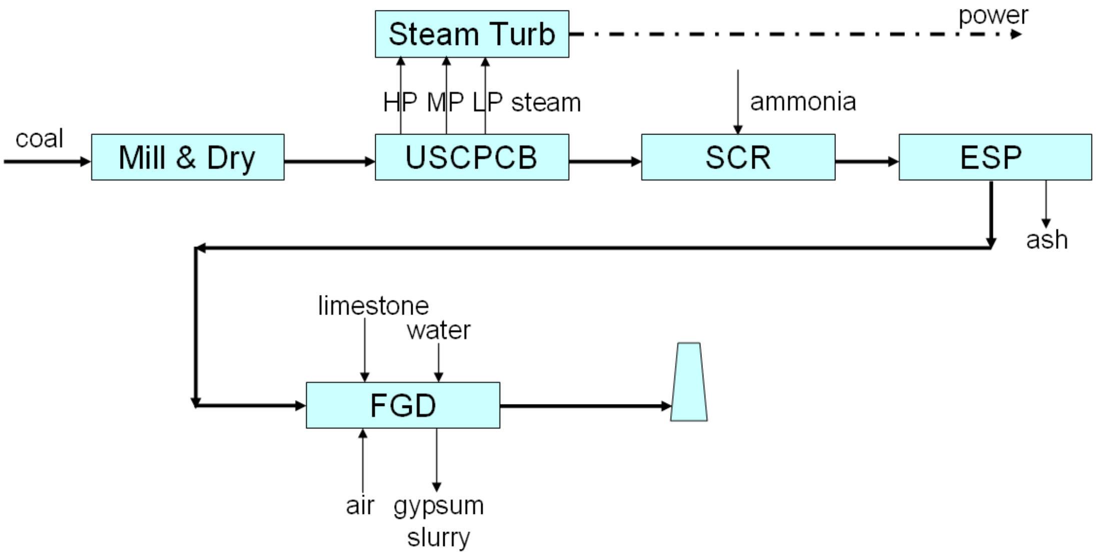


Figure 2-3 **Case 2A - USCPC with Post Combustion CO_2 Capture**

- Case 2B – Ultra supercritical pulverised coal (USCPC) without CO_2 capture. Single 800 MW (nominal) pulverised coal fired boiler raising steam at 275 bara, Electrostatic precipitator (ESP) for particulate removal, limestone scrubbing for SO_2 removal, selective catalytic reduction (SCR) for NOx removal.

Figure 2-4 **Case 2B - USCPC without CO_2 Capture**



2.2.3 Natural Gas Based Post-Combustion

- Case 3A – Combined Cycle Gas Turbine (CCGT) with CO_2 capture. Two trains of Mitsubishi Heavy Industries (MHI) M701G2 natural gas fed gas turbine, each with its own heat recovery steam generator (HRSG) feeding a single steam turbine. MEA based CO_2 absorption system. Multi stage integrally geared centrifugal CO_2 compression with CO_2 dehydration.

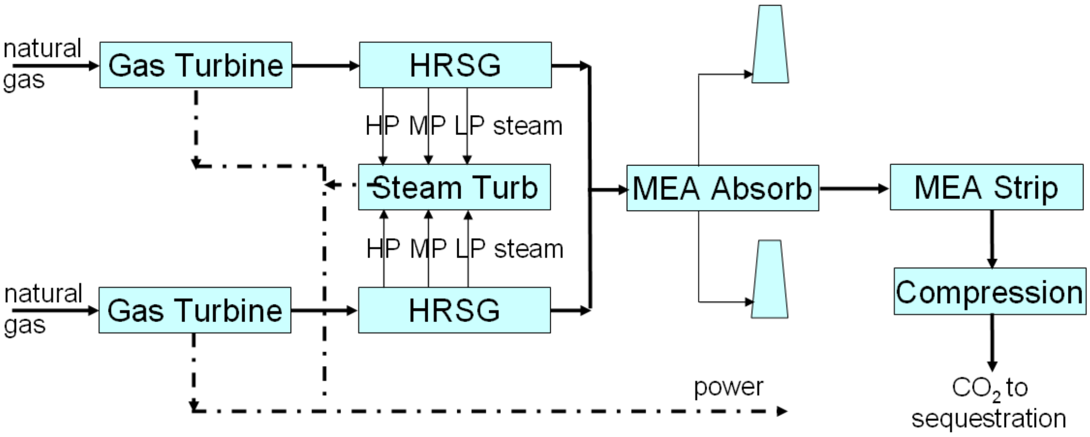


Figure 2-5 Case 3A - CCGT with Post Combustion CO_2 Capture

- Case 3B – Combined Cycle Gas Turbine (CCGT) without CO_2 capture. Two trains of Mitsubishi Heavy Industries (MHI) M701G2 natural gas fed gas turbine, each with its own heat recovery steam generator (HRSG) feeding a single steam turbine.

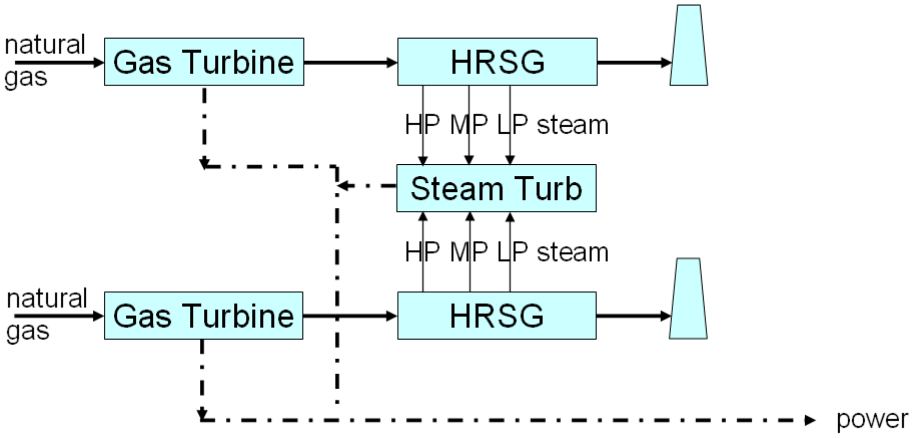


Figure 2-6 Case 3B - CCGT without CO_2 Capture

2.2.4 Oxy-fuel

- Case 4 – Coal based Oxy-Fuel. Single 800 MW (nominal) pulverised coal fired boiler with steam conditions as Cases 2A and 2B. SCR to remove residual NO_x to acceptable levels. "Hot" ESP located between the convection section of the boiler and a recycle gas heater, prior to the flue gas recycle to remove particulates. FGD located upstream of the flue gas recycle loop off-take removing SO_x from the boiler flue gas and thereby avoiding SO_x recycle to the boiler with resultant potential corrosion issues. Air Separation Unit. Cryogenic CO_2 purification including CO_2 dehydration, CO_2 liquefaction flash and low temperature separation of residual impurities, achieving the required design basis product CO_2 purity. Multi stage integrally geared centrifugal CO_2 compression.

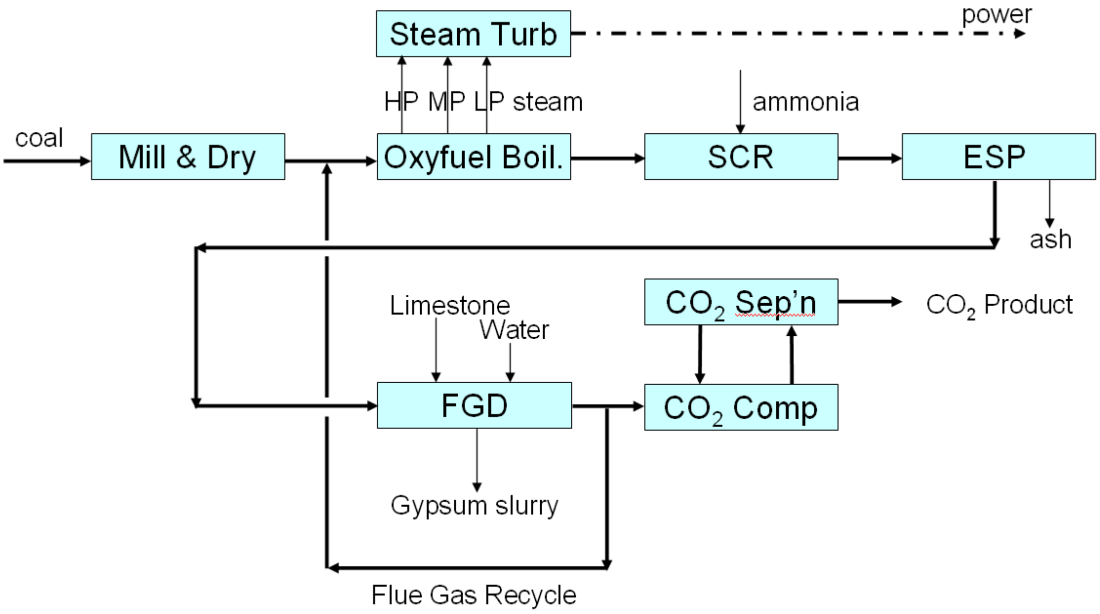


Figure 2-7 Case 4 - USCPC with Oxyfuel CO_2 Capture

2.3 Comparison with Previous Public Domain Studies

A number of studies have been conducted in the past to provide comparisons between the various different options for carbon capture from power generation. The findings of Work Package 1 were compared to results presented in the following public domain reports:

- "Evaluation of Innovative Fossil Fuel Power Plants with CO_2 Removal", by EPRI, December 2000.
- "Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Final Report", by the DOE and NETL, May 2007.
- "Fossil Fuel Fired Power Generation", IEA Clean Coal Centre, 2007.
- "CO_2 Capture in Low Rank Coal Power Plants", IEA Greenhouse Gas R&D Programme, November 2005.
- "Gas Turbine World" articles published in early 2009, in the IGCC Reference Guide.
- "Co-Production of Hydrogen and Electricity by Coal Gasification with CO_2 Capture – updated economic analysis", IEA Greenhouse Gas R&D Programme, August 2008.
- "Strategic Analysis of the Global Status of Carbon Capture and Storage, Report 5: Synthesis Report, Final Report", Global CCS Institute, September 2009.

Comparison Methodology

In order to quantitatively compare the specific capital cost (GB£/kW_{net}) trend across the four cases assessed in this study with the findings of the above open literature studies it was necessary to bring them as close as possible to the same basis.

It was not feasible in the scope of this study to attempt to adjust the figures for technology differences (such as supercritical versus ultra-supercritical coal plants), design basis differences (cooling towers versus sea water cooling), or factor capital cost assumptions differences (such as varying contingency levels), especially as these were not always specifically stated.

The figures must therefore be treated with caution when comparing differing study results. It was possible, however, to adjust the figures so that they reflect the correct currency and time basis. The figures presented below were calculated using the following method:

- Conversion to US$
- Escalation from historic study cost basis date to 1Q2009
- Conversion to GB£

The currency conversions were based on US Federal Reserve published historic currency figures while the escalation was based on the IHS-CERA indices for power plants from 2000 to present day.

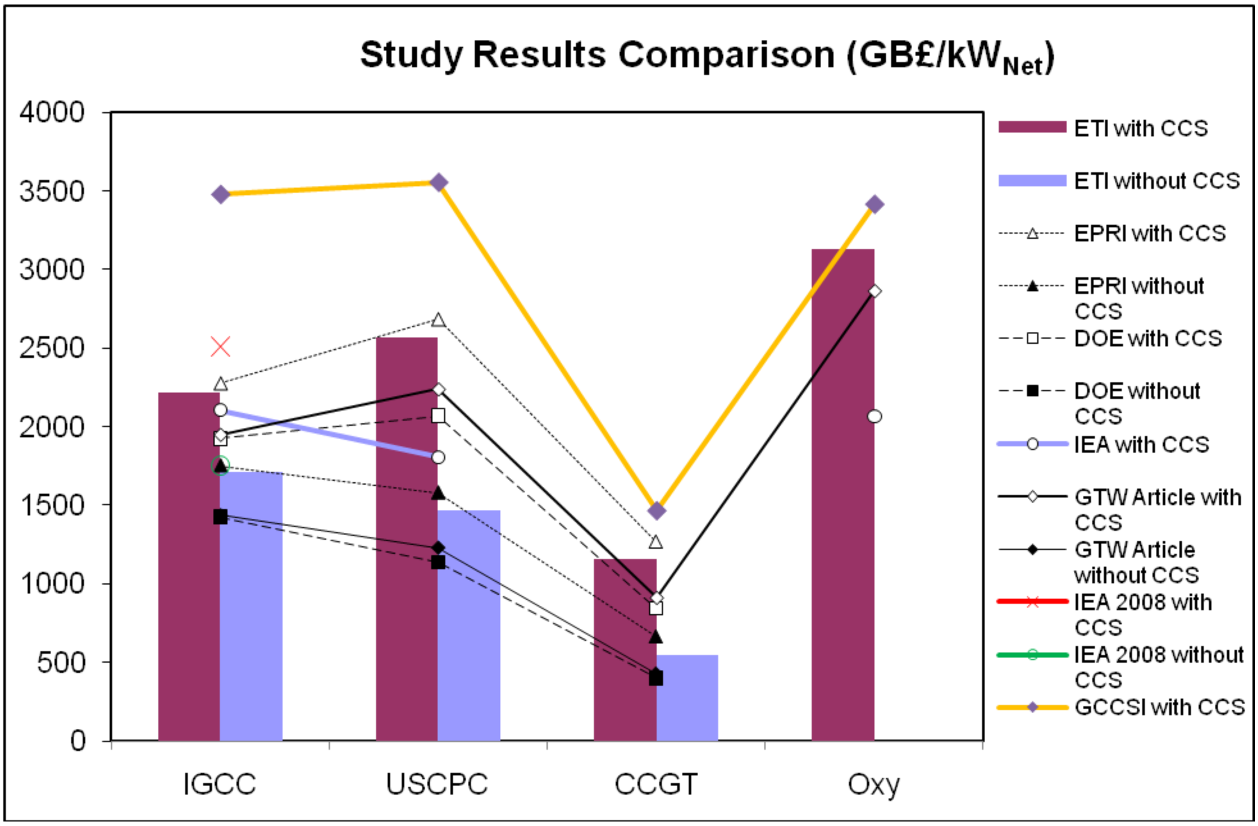


Figure 2-8 Results Comparison with Previous Studies

The bars represent the results for this study while the marked lines represent the open literature study data for the cases with and without carbon capture respectively.

EPRI Study

Despite the age of this study, both the final figures and the relationship between the costs of each case are very similar to the ETI study's results. The IGCC cases were particularly close with the EPRI study showing a very small increase from the ETI Study results for the USCPC and CCGT cases.

DOE Bituminous Coal Study

The figures presented in the DOE report are between 15% and 27% lower than the results of this study. The IGCC cases with and without capture are 15% and 17% cheaper than this study's results while there is a greater difference in the USCPC and CCGT cases. The USCPC case sees figures 19% and 22% lower in the DOE's results and the CCGT cases are both 27% lower than our results.

These variations may be due to variations in basis between the two pieces of work: the DOE coal plant is supercritical rather than ultra supercritical and the CCGT basis is F class compared to our G class machines. It should also be noted that lower figures for contingency are applied as the report moves from IGCC to USCPC then with CCGT cases having the lowest project contingency allowed.

IEA Reports

There is some limited cost data in the Fossil Fuel Fired Power Generation report which is roughly in-line with this study's figures, for example the incremental cost in $/kWh when changing an ultra supercritical boiler to oxy-firing.

The GHG Low Rank Coal study shows a rather smaller change when moving from USC coal to oxy-firing than the Fossil Fuel Fired Generation Report. It also does not give data for unabated cases for comparison. The "with capture" cases show a close comparison for the IGCC case but are significantly lower cost than this study for the USCPC and Oxy cases.

The IEA 2008 study only reports cost figures for IGCC cases with and without carbon capture and the results are similar to the ETI study figures. The "without capture" case shows particularly close agreement to equivalent IEA 2008 study case, whilst the "with capture" case figures show an approximately 10% difference.

GTW Articles

These articles are contemporary with the cost data used in this study and, like the DOE results, they show a significant and fairly consistent gap between the open literature study results and those presented in this study, with the GTW articles published results about 10% to 20% lower.

GCCSI Study

Like the GTW article, the GCCSI study is contemporary with the cost data used in this study. However, the results from the GCCSI study illustrate a significant gap with those reported in this study. This is particular evident in the GCCSI report figures for the IGCC and USCPC cases with CO_2 capture, which are 57% and 39% higher respectively than those reported in this study.

Comparison Conclusion

Overall it can be seen that there is generally good agreement between this study's results and the open literature in terms of both the relationship between the benchmark technologies and for the cost difference between the cases with and without carbon capture. The main notable exceptions to this are the results from the GCCSI study, particularly relating to IGCC and USCPC cases with CO_2 capture.

3. INTEGRATED GASIFICATION COMBINED CYCLE TECHNICAL DESCRIPTIONS

3.1 Case 1A – IGCC with CO_2 Capture

3.1.1 Introduction

The overall process scheme for this case is a coal fed integrated gasification combined cycle (IGCC) scheme, based upon dry feed, entrained flow gasification, a sour shift unit, acid gas removal unit, CO_2 compression and dehydration unit, and power island featuring 2 x frame F class gas turbines and single steam turbine generator (STG).

To ensure full utilisation of the gas turbines this case was sized to produce sufficient syngas to satisfy the full thermal appetite of the two gas turbines. The process conditions, including stream flows, pressures, temperatures and compositions, were produced to reflect this sizing basis.

Key features of the proposed configuration include:

- Gasification Unit – developed using in-house information, supported as required by Shell, based on Shell's dry feed gasifier with product gas cooling in a heat recovery boiler.
- Air Separation Unit – The cryogenic air separation unit (ASU) produces 95% purity gaseous oxygen at approximately 40 bar to the Gasification Unit plus a small flow of low pressure oxygen for the Sulphur Recovery Unit, gaseous nitrogen for injection into the gas turbine fuel for NO_X emission control, coal conveying and plant/instrument air. In addition, the ASU is able to produce quantities of liquid oxygen to maintain the desired backup inventory. The liquid oxygen storage system will be sized to provide 12 hours backup supply at full design rate of normal gaseous oxygen supply.
- Shift Unit – adiabatic three stage sour water gas shift unit to provide sufficient CO shift, in conjunction with AGR CO_2 slip, to achieve target CO_2 capture level.
- Acid Gas Removal Unit – both carbon dioxide and H_2S removal scheme developed using in-house information on the basis of a physical solvent type process.
- Sulphur Removal Unit / Tail Gas Treatment Unit – The sulphur recovery unit (SRU) converts the environmentally harmful sulphur emissions from the coal feed into a marketable sulphur product. The feed to the unit is composed of the acid gas from the AGR and the stripped gases from the Gasification unit's sour water treatment. Sulphur removal is achieved using a Claus process which can be split into two process steps, thermal and catalytic. The Claus unit removes approximately 96% of the sulphur from the acid gas stream, this recovery is increased to approximately 99.9% sulphur recovery by the tail gas treatment unit.
- Power Island Unit – comprising of two parallel trains, each with one F class 50 Hz gas turbine without air extraction and one heat recovery steam generator (HRSG), connected to a single condensing steam turbine, using seawater cooling. The configuration does not include air integration between the GT and the ASU.
- Carbon Dioxide Compression and Drying Units – dehydration and compression to 150 barg based on in-house Foster Wheeler knowledge of commercially available equipment.

In addition, a required feature of this case is the flexibility to produce raw hydrogen as a co-product. No adjustments to the configuration or sizing are proposed to accommodate this feature. In the event that raw hydrogen export is required, it is intended that the required quantity will taken as a side stream product at a point within the flow scheme downstream of the AGR at the naturally occurring process conditions, whilst the remaining raw hydrogen is fed to the downstream power island.

The following two bullet points describe the approach taken for the WP3 development of the 85% and 95% capture cases from the original 90% capture WP1 Case 1A bench mark case.

- In the 85% capture case two Versions were evaluated. In Version 1 the reduced CO_2 capture was achieved by reducing the flow of MP process steam to the CO shift, with the CO_2 removal unit remaining essentially unchanged. In Version 2 the CO shift steam flow was retained from the WP1 Case 1A (with 90% capture) and the reduced CO_2 capture was achieved by increasing the CO_2 content of the process gas at the outlet of the CO_2 removal unit.
- In the 95% capture case the flow of MP process steam to the CO shift was increased and the CO_2 removal unit was modified to reduce the residual CO_2 content of the process gas.

3.1.2 Plant Performance

The plant performance for the three levels of CO_2 capture are outlined in the table below:

Figure 3-1 Performance Figures for Case 1A – IGCC with CO_2 Capture

Case		**1A $_{85\%\ V1}$**	**1A$_{85\%\ V2}$**	**1A $_{90\%}$**	**1A $_{95\%}$**
Power					
Total gross installed capacity	MWe	923.8	922.9	923.8	906.2
Gas Turbine (s)	MWe	535.2	543.6	544.4	545.6
Steam Turbine	MWe	388.6	379.3	379.4	360.6
Others	MWe	0.0	0.0	0.0	0.0
Total auxiliary loads	MWe	216.5	218.4	223.9	236.2
ASU	MWe	98.1	100.2	101.6	103.2
Gasification / Boiler	MWe	12.8	13.0	13.0	13.0
Power Island	MWe	7.3	7.6	7.6	7.7
Acid Gas Removal	MWe	30.5	30.0	30.8	33.4
CO_2 compression	MWe	45.1	44.7	47.9	55.8
Others	MWe	22.6	23.0	23.0	23.0
Net Power Export	MWe	707.3	704.5	699.9	670.0
Net Efficiency (LHV)	%	36.4	36.0	35.5	34.0
Heat Rate	kJ/kWh	9860	10062	10129	10607
Flows					
Total fuel feed rate	tpd	6458.1	6559.2	6559.2	6565.3
Oxygen consumption	tpd	4640.3	4709.2	4709.2	4712.1
Water consumption	tpd	6329.9	6423.0	6429.8	6435.0
Cooling water (once through)	tpd	2190200	2140599	2160192	2133855
Carbon Balance					
Total carbon in feeds	tpd	4161.2	4251.6	4238.9	4242.9
Total carbon captured	tpd	3544.4	3589.9	3814.0	4029.8
Carbon capture rate	%	85.0	85.0	90.0	95.0
Total CO_2 captured	tpd	12988.2	13187.9	13976.0	14766.9
Total CO_2 emitted	tpd	2260	2298.5	1557	781

Case		**1A** $_{85\%\ V1}$	**1A**$_{85\%\ V2}$	**1A** $_{90\%}$	**1A** $_{95\%}$
CO_2 emissions	g CO_2/ kWh$_{Net}$	133.2	134.8	92.7	48.5

The variations in % carbon capture introduced in WP3 (85% CC and 95% CC) have been carried out with the same plant configuration as the original 90% CC Case 1A in WP1 (i.e. Shell gasification at 40 bar g followed by three-stage sour shift and DEPG (Selexol) wash for H_2S and CO_2 removal). It should be noted that the 95% case is close to the upper limit of feasibility with this configuration, and a revised configuration of CO_2 removal system may be required if a higher % CC were required. It was shown that the changes in power plant parasitic load and hence overall efficiency were not a linear function of the %CC, in particular more energy (as steam and electrical power per unit CO_2), was required to achieve 95%CC than to achieve 90%CC or 85%.

3.1.3 Process Description

3.1.3.1 Coal Receiving, Handling, Preparation and Feeding

Raw coal is crushed and fed to a pulveriser, where it is ground to a size suitable for the Shell gasifier (90 wt% < 100 microns). A hot nitrogen gas stream is introduced to the coal as it is being ground to reduce the moisture content to <5 wt%. The heat source for coal drying is a small quantity of natural gas. Controlling the nitrogen flow rate through the grinding chamber ensures only coal of a suitable size is conveyed, oversize coal particles remain in the grinding chamber. The coal particles are removed from the inert gas stream first in a centrifuge and then in a fabric dust collector. The moist inert gas stream is discharged.

The ground partially dried coal is then transferred to lock hoppers from which it is conveyed by high pressure nitrogen from the air separation unit before being introduced into the gasifier.

A small quantity of flux is mixed with the dried coal to facilitate withdrawal of ash from the gasifier.

3.1.3.2 Air Separation Unit

Cryogenic distillation of atmospheric air produces the oxygen and nitrogen required by the process. Oxygen at a purity of >95% and a pressure of approximately 48 barg is required for the Shell Gasifier. Lower pressure oxygen is also required to the Claus unit. A small quantity of nitrogen at high pressure, approximately 70 barg, is required as a transport medium for the pulverised coal into the gasifier. Nitrogen at lower pressure, approximately 32 barg, is required for blending with the hydrogen fuel to the gas turbines with ballast nitrogen at 25 barg required for injection into the gas turbines.

3.1.3.3 Gasification

The Shell gasifier is a dry-feed, pressurised, oxygen-blown, entrained flow slagging type. High pressure coal, oxygen and steam enter the gasifier through horizontally opposed burners at approximately 40 barg. The raw syngas produced in the reaction travels upwards through the gasifier at a temperature of approximately 1400 - 1500°C and exits together with entrained slag from the top of the gasifier. The gasification zone is lined with a membrane wall, in which medium pressure steam is generated. The high reaction temperature results in most of the mineral content of the feed being converted to molten slag, which forms a protective layer on the membrane wall. The slag flows down the wall and out of an opening at the bottom of the gasifier into a water quench chamber. The insulating properties of the slag help reduce heat loss to the membrane wall, ensuring a high cold gas efficiencies with minimum production of CO_2 in the syngas. On contact with the water quench the slag forms dense, glassy granules which are washed, depressured and then sent to storage for recycle and disposal.

Hot syngas leaving the gasifier is quenched with cooled, filtered syngas to a temperature of approximately 750 °C. The syngas is cooled further by generating first superheated HP steam, then saturated MP steam. Entrained solid residue is filtered from the syngas and recycled to the gasifier.

Any remaining particulate matter together with chlorides are washed from the syngas in the scrubber vessel before the syngas is routed to the scrubber/saturator, where it is contacted with hot circulating water. The water saturated syngas is heated against syngas leaving the third shift reactor before entering the first shift reactor where CO is converted to CO_2 by reaction with water, generating additional hydrogen.

$$CO + H_2O \leftrightarrows CO_2 + H_2$$

Carbonyl Sulphide (COS) is also hydrolysed to H_2S (and CO_2), eliminating the need for a separate COS hydrolysis reactor.

$$COS + H_2O \leftrightarrows CO_2 + H_2S$$

Heat from the reactions above is removed by generation of saturated HP and subsequently saturated MP steam before the syngas enters a second shift reactor where more CO is converted to CO_2. The heat of reaction from the second shift reactor is removed by generation of MP steam before the syngas enters the third shift reactor. The reaction heat is removed firstly by heating the syngas feed to the first shift reactor then by heating the recycled water from the syngas saturator.

The syngas stream then splits and is further cooled against the gas turbine fuel gas from AGRU and the ballast N_2 for injection into the GT. The gas turbine fuel gas comprises hydrogen-rich syngas from the AGR diluted with nitrogen from the ASU.

Condensed water is knocked out and routed to the sour water drum before the syngas is further cooled against vacuum condensate. A second knock out drum removes any water before the syngas stream is combined with recycled tailgas from the Claus unit and enters the Selexol acid gas removal unit.

3.1.3.4 Acid Gas Removal

The purpose of the Selexol unit is to preferentially remove H_2S as a product stream and then to remove CO_2 as a separate product stream. This is achieved in a dual-absorber Selexol process.

Raw syngas enters the unit and is cooled against product gas in the feed/product exchanger before entering the H_2S absorber. Gas flowing upwards through the packed beds is contacted with solvent CO_2 loaded solvent to absorb the H_2S and COS content of the syngas. Some CO_2 and H_2 are absorbed in the solvent before the remaining gas leaves the top of the column. The rich solvent loaded with CO_2 and H_2S leaves the bottom of the column and is heated against lean solvent leaving the H_2S stripper in the lean/rich solvent exchanger.

The treated gas leaves the top of the H_2S absorber and is sent to the CO_2 absorber. Gas flowing upwards through the packed beds is contacted with cool, lean solvent which enters the top of the absorber and cool, semi-lean solvent entering in the middle of the absorber. CO_2 is primarily absorbed by the solvent from the gas stream but some remaining H_2S as well as H_2 are also absorbed. The treated gas leaves the top of the absorber, is used to cool the raw syngas feed to the unit in the feed/product exchanger before being combined with nitrogen and routed to the GT Fuel Gas Heater where it is heated prior to firing in the gas turbine.
The CO_2 loaded solvent leaving the CO_2 Absorber is split with a portion of the flow being pumped via a chiller to the top of the H_2S Absorber. The remaining flow is sent to the first of a series of four flash drums to liberate CO_2 and partially regenerate the solvent.

The loaded solvent is reduced in pressure into the first flash drum evolving H_2, CO_2 and other dissolved gases back into the gas phase. This gas stream is compressed and recycled back into the CO_2 absorber. This recycle minimises the losses of hydrogen from the system and improves the purity of the CO_2 to meet the CO_2 product specification.

The solvent from the first flash drum is further reduced in pressure to approximately 8 bara before entering the CO_2 HP Flash Drum. CO_2 and traces of other dissolved gases are evolved and routed to the CO_2 compression and drying unit. The solvent is then further reduced in pressure to approximately 3.5 bara into the CO_2 MP Flash Drum, evolving more CO_2 which is routed to CO_2 compression and drying. The solvent leaving the drum is reduced in pressure to approximately 1.5 bara into the CO_2 LP Flash Drum. The CO_2 evolved is routed to the CO_2 compression and drying unit. The semi-lean solvent leaving the CO_2 LP Flash Drum is pumped and chilled before entering the middle of the CO_2 absorber.

The rich solvent after being heated in the lean/rich solvent exchanger is routed to the H_2S Concentrator where CO_2 and traces of other dissolved gases are stripped against a slipstream of sweet syngas which has been compressed from downstream the Feed/Product Exchanger. In the H_2S Concentrator, compounds with a lower relative solubility are preferentially stripped from the solvent, while those with a high relative solubility, primarily H_2S remain dissolved in the solvent. The gas stream leaving the top of the H_2S Concentrator, composed primarily of CO_2, N_2, CO and traces of H_2S, is combined with the compressed gas stream from the H_2S flash drum is cooled and routed to combine with the feed stream.

The solvent leaving the bottom of the concentrator is reduced in pressure into the H_2S Flash Drum where more low relative solubility compounds are evolved as gas. These gases are cooled, any solvent separated in the Flash Gas KO Drum before being compressed and combined with the H_2S Concentrator overheads back to blend with the feed stream. The solvent leaving the H_2S Flash Drum is combined with solvent separated in the Flash Gas KO Drum and sent to the H_2S Stripper. The solvent passes down the column where H_2S, COS and CO_2 are steam stripped. The steam and stripped gases pass up the column and leave the top, a condenser, condenses the stream and the acid gases (approx 40 mol% H_2S) leave the top of the reflux drum where they are routed to the Sulphur Recovery Unit.

Steam for stripping is generated in the Stripper Reboiler at the bottom of the H_2S Stripper. The lean solvent leaves the bottom of the column, its heat is recovered against rich solvent in the Lean/Rich Solvent Exchanger before it is pumped through a chiller and into the top of the CO_2 Absorber completing the cycle.

3.1.3.5 Sulphur Recovery Unit

Sour gas from the Acid Gas Removal Unit together with the sour gas stream from the Sour Water Stripper together with 95% purity oxygen from the Air Separation Unit and air is fed to the Claus combustor. The Claus combustor partially oxidises the H_2S to SO_2 and by ensuring the furnace temperature is maintained above 1350 °C, any NH_3 present is decomposed. The gases leaving the furnace section are cooled by generation of MP steam and any elemental sulphur formed is condensed and routed to the sulphur storage tank. The gas stream leaving the boiler section is cooled further by generating LP steam before entering the first Claus catalytic reactor. In the first Claus Catalytic Reactor H_2S is catalytically oxidised with SO_2 to elemental sulphur. The stream leaving the reactor is cooled by generation of LP steam. Any sulphur condensed is routed to sulphur storage. The gas stream is then heated against LP steam before entering the second Claus reactor. After reaction, the gas stream is cooled by generation of LP steam and condensed sulphur is routed to sulphur storage. The remaining tail gas is sent to the Tail Gas Treating Unit.

3.1.3.6 Tail Gas Treatment Unit

Tailgas from the Claus unit is combined with Syngas (H_2 rich gas) before being pre-heated against MP steam into the Hydrogenation Reactor. In the Hydrogenation Reactor, SO_2 and any elemental sulphur species are catalytically reduced in the presence of H_2 to H_2S. COS is also hydrolysed to H_2S. The hydrogenated stream is then cooled by generation of LP steam, quenched against water in a quench tower, before being compressed back to the Acid Gas Removal Unit.

A slipstream of sour water is removed and sent to the Sour Water Stripper to prevent accumulation in the circulating quench water.

3.1.3.7 Sour Water Stripper

Sour water received from various sources including syngas cooling, tailgas quench etc will be processed in the Sour Water Stripper unit to produce a sour gas stream and a sweet water stream to be reused in the gasification process.

Sour water received in the Sour Stripper Column is reboiled against LP steam. The sour gas leaving the top of the stripper column is routed to the Sulphur Recovery Unit. The sweet water leaving the bottom of the column is recycled back to the Syngas Saturator column.

3.1.3.8 CO_2 Compression and Drying

CO_2 compression is achieved in an 8 stage compressor with intermediate pressure adsorptive dryer. Stages 1 to 6 compress the CO_2 stream to approximately 35 bar. An intercooler is provided downstream each compression stage. Separators are provided after the intercoolers (except after the first intercooler) to remove condensed water. The cooled gas from Stage 6 is then routed to one of two dehydration vessels containing molecular sieve adsorbent. Two dehydration vessels are required as one molecular sieve bed will be in use whilst simultaneously the second bed will be in regeneration. The regeneration cycle uses a slipstream of dried gas exiting the operating molecular sieve bed. The gas is heated using the returning regeneration gas exiting the molecular sieve bed in regeneration. It is further heated under temperature control in an electric heater before entering the bed in a counter flow direction. The wet gas leaving the bed is cooled against incoming gas, any condensed water is separated in a knock out drum before it is passed through a fines filter and returned upstream of the 3rd stage compressor. The absorbent regeneration process takes several hours. When complete the heater is bypassed and the bed is cooled down over several hours before return to operation.

The molecular sieve adsorbs water to a level of < 50ppmv in the product gas. The dry CO_2 exiting the dehydration vessel is routed through a filter to remove any fines before entering the 7th stage compressor, then cooled before the final 8th stage of compression to reach the 150 bar product pressure required. The gas is cooled in a final cooler before pipeline export.

The water separated in the knock out drums is returned to the acid gas removal unit as make-up water.

3.1.3.9 Combined Cycle Power Generation

The power island is based on two General Electric Frame 9F IGCC gas turbines, two heat recovery steam generators (HRSGs) generating steam at two pressure levels and a single steam turbine.

In normal operation the gas turbine is fired with decarbonised fuel gas, comprising hydrogen from the AGRU diluted with nitrogen from the Air Separation unit. Additional nitrogen is injected directly into the GT as ballast flow for the expansion section. Dual fuel burners within the gas turbine allow start-up and shutdown to be carried out using natural gas at start-up or when syngas is not available.

The gas turbine exhaust gas flows to the Heat Recovery Steam Generator at approximately 550^0C. The thermal energy of the exhaust gases is used to superheat HP and MP steam, to generate HP steam in an HP boiler, heat the boiler feedwater, to generate LP steam in the LP boiler for use in the deaerator and to heat the vacuum condensate. The flue gases are emitted to atmosphere at approximately 85 °C.

The coil sequence in the HRSG is as follows:

- HP Superheater and MP Superheater are arranged in parallel across the exhaust duct.
- HP Boiler
- HP Economiser
- BFW Heater
- LP Boiler
- Vacuum Condensate Heater

Exhaust gas is discharged to the atmosphere through the HRSG stack, which is fitted with a continuous emission monitoring system and a silencer.

Vacuum condensate from the pumps is heated first against raw syngas before being heated further against flue gases and entering the deaerator. A small quantity of LP steam is generated in the LP boiler, from a slipstream of boiler feed water from the deaerator to strip the oxygen from the system. The remaining boiler feed water is routed to the boiler feedwater pumps, which delivers it via the boiler feed water heater and the HP economiser to the HP steam drum. A sidestream from downstream of the boiler feed water heater is routed to provide MP boiler feed water to the gasification unit. Water from the HP steam drum generates HP steam in the HP boiler. This HP steam combines with HP steam from the gasification unit before being superheated in the HP superheater. Superheated steam is combined with the superheated steam from the second HRSG and is sent to the HP stage of the steam turbine. The exhaust steam from the HP module of the steam turbine is split between the two HRSGs and is superheated in the two MP superheaters before being recombined and routed to the MP stage of the steam turbine. Exhaust steam from the MP stage of the steam turbine is routed to the LP stage of the steam turbine. Electricity is generated in the steam turbine generator on a common shaft with the 3 stages of steam turbine. Exhaust steam from the LP stage of the steam turbine is condensed against seawater in the vacuum condenser, before returning to the suction of the Vacuum Condensate Pumps completing the circuit.

3.1.3.10 Balance of Plant

The key balance of plant requirements for this scheme are the cooling water supply systems. A very large flow of cooling water is required to supply the steam turbine vacuum condenser. This duty is supplied using sea water in a once through flow scheme.

The ASU, AGRU and CO_2 compression and drying units also require a significant quantity of cooling medium. Where this cannot be supplied using heat integration within or between the process units, cooling water is required. This cooling water will be supplied as fresh cooling water in a closed circuit. The fresh water system is cooled against sea water.

Facilities are also required for storage and make up of the Selexol based solvent to the AGRU. Reuse and treatment of the numerous, mainly small, water streams produced from the cooling of water saturated gas streams will be integrated with the units where possible.

3.1.4 Plant Turndown

This case can be turned down to 50% output by idling one of the two gasification + CCGT lines. However, this will not be an attractive procedure, except when a long plant outage is foreseen, due to the time required and the operating expense of restarting a complete line. In normal operation it is expected that turn down would be achieved by reducing the output of one or both lines, down to around 50% of full capacity, with however, a significant loss of thermal efficiency.

3.1.5 Major Equipment List

A high level equipment list was developed for this case broken down according to unit. This equipment list is located in Appendix 4 of this report.

3.1.6 Capital Cost, Operating Cost and Economics

The economic results are outlined in the table below:

Figure 3-2 Economic Figures for Case 1A – IGCC with CO_2 Capture

Case		**$1A_{85\% V1}$**	**$1A_{85\% V2}$**	**$1A_{90\%}$**	**$1A_{95\%}$**
ASU	GB£M	144	146	145.8	146
Gasification / Boiler	GB£M	331	334	334.4	335
Power Island	GB£M	478	478	477.5	472
Acid Gas Removal	GB£M	161	163	169.4	175
CO_2 compression	GB£M	83	83	86.4	96
Others	GB£M	320	334	342.5	358
Total CAPEX	GB£M	1517	1538	1556	1582
CAPEX efficiency	GB£/kWh	2145	2166	2223	2361
Total OPEX – incl. fuel	GB£M p.a.	204.7	207.3	208.2	209.4
Total OPEX – excl. fuel	GB£M p.a.	74.5	75.0	75.9	77.0
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	289.4	294.3	297.5	312.5
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	105.3	106.5	108.4	114.9
Levelised Cost of Electricity					
CO_2 emission cost = £ 0 / te CO_2	£ / MWh_{Net}	73.86	75.16	76.23	80.47
CO_2 emission cost = £ 20 / te CO_2	£ / MWh_{Net}	76.53	77.88	78.08	81.44
CO_2 emission cost = £ 40 / te CO_2	£ / MWh_{Net}	79.19	80.60	79.94	82.41
CO_2 emission cost = £ 60 / te CO_2	£ / MWh_{Net}	81.85	83.32	81.79	83.38
Cost of CO_2 Captured					
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	18.78	20.09	20.12	22.85
Cost of CO_2 Avoided					
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	23.89	26.18	26.08	30.58

The CAPEX and OPEX figures for the varied % carbon capture (CC) cases reflect the trends shown in the power plant performance figures; costs are not a linear function of the %CC, in particular more incremental cost was incurred, to achieve 95%CC than 90%CC or 85%. The net effect of these cost and efficiency trends is that the costs of CO_2 captured and avoided appear to increase slightly with increasing %CC, highlighting the fact that the IGCC scheme is close to its maximum possible %CC level, without introducing process modifications.

3.2 Case 1B – IGCC without CO_2 Capture

3.2.1 Introduction

The overall process scheme for this case is a coal fed integrated gasification combined cycle (IGCC) scheme, upon dry feed, entrained flow gasification, acid gas removal unit, and power island featuring 2 x frame F class gas turbines and single steam turbine generator (STG).

As with the IGCC with CO_2 capture case this scheme was sized to ensure full utilisation of the gas turbines by producing sufficient syngas to provide the full thermal appetite of the two gas turbines. However, due to the higher LHV of the syngas relative to the IGCC with CO_2 capture case, a lower product syngas flow will be required resulting in smaller Gasification, ASU, AGR, SRU/TGT unit sizes and correspondingly lower coal feed rate, oxygen consumption and CO_2 product rate. The process conditions, including stream flows, pressures, temperatures and compositions, were produced to reflect this sizing basis.

Key features of the proposed configuration include:

- Gasification Unit – developed using in-house information, supported as required by Shell, based on Shell's dry feed gasifier with product gas cooling in a heat recovery boiler.
- Air Separation Unit – The cryogenic air separation unit (ASU) produces 95% purity gaseous oxygen at approximately 40 bar to the Gasification Unit plus a small flow of low pressure oxygen for the Sulphur Recovery Unit, gaseous nitrogen for injection into the gas turbine fuel for NOX emission control, coal conveying and plant/instrument air. In addition, the ASU is able to produce quantities of liquid oxygen to maintain the desired backup inventory. The liquid oxygen storage system will be sized to provide 12 hours backup supply at full design rate of normal gaseous oxygen supply.
- Acid Gas Removal Unit – for H_2S removal only a scheme developed using in-house information on the basis of a chemical solvent type process (e.g. MDEA).
- Sulphur Removal Unit / Tail Gas Treatment Unit – The sulphur recovery unit (SRU) converts the environmentally harmful sulphur emissions from the coal feed into a marketable product. The feed to the unit is composed of the acid gas from the AGR and the stripped gases from the Gasification unit's sour water treatment. Sulphur removal is achieved using a Claus process which can be split into two process steps, thermal and catalytic. The Claus unit removes approximately 96% of the sulphur from the acid gas stream, this recovery is increased to approximately 99.9% sulphur recovery by the tail gas treatment unit.
- Power Island Unit – comprising of two parallel trains, each with one F class 50 Hz gas turbine and one heat recovery steam generator (HRSG), connected to a single condensing steam turbine, using seawater cooling. The configuration will not include air integration between the GT and the ASU.

3.2.2 Plant Performance

The plant performance is outlined in the table below:

Figure 3-3 Performance Figures for Case 1B – IGCC without CO_2 Capture

		Case 1B
Power		
Total gross installed capacity	MWe	878.1
Gas Turbine (s)	MWe	514.5
Steam Turbine	MWe	363.6
Others	MWe	0.0
Total auxiliary loads	MWe	132.0
ASU	MWe	95.7
Gasification / Boiler	MWe	11.6
Power Island	MWe	10.5
Acid Gas Removal	MWe	0.3
CO_2 compression	MWe	0.0
Others	MWe	13.9
Net Power Export	MWe	746.1
Net Efficiency (LHV)	%	45.0
Heat Rate	kJ/kWh	8006
Flows		
Total fuel feed rate	tpd	5539.2
Oxygen consumption	tpd	4243.0
Water consumption	tpd	1624.5
Cooling water (once through)	tpd	1818240
Carbon Balance		
Total carbon in feeds	tpd	3588.5
Total carbon captured	tpd	0.0
Carbon capture rate	%	0.0
Total CO_2 captured	tpd	0
Total CO_2 emitted	tpd	13150
CO_2 emissions	g CO_2/ kWh_{Net}	734.4

3.2.3 Process Description

3.2.3.1 Coal Receiving, Handling, Preparation and Feeding

Raw coal is crushed and fed to a pulveriser, where it is ground to a size suitable for the Shell gasifier (90 wt% < 100 microns). A hot nitrogen gas stream is introduced to the coal as it is being ground to reduce the moisture content to <5 wt%. Controlling the nitrogen flow rate through the grinding chamber ensures only coal of a suitable size is conveyed, oversize coal particles remain in the grinding chamber. The coal particles are removed from the inert gas stream first in a centrifuge and then in a fabric dust collector. The moist inert gas stream is discharged.

The ground partially dried coal is then transferred to lock hoppers from which it is conveyed by high pressure nitrogen from the air separation unit before being introduced into the gasifier.

A small quantity of flux is mixed with the dried coal to facilitate withdrawal of ash from the gasifier.

3.2.3.2 Air Separation Unit

Cryogenic distillation of atmospheric air produces the oxygen and nitrogen required by the process. Oxygen at a purity of > 95% and a pressure of approximately 40 bara is required for the Shell Gasifier. Lower pressure oxygen is also required to the Claus unit. Nitrogen at high pressure, approximately 70 bar, is required as a transport medium for the pulverised coal into the gasifier. Nitrogen at lower pressure, approximately 33 bara, is required for blending with the hydrogen fuel to the gas turbines.

3.2.3.3 Gasification

The Shell gasifier is a dry-feed, pressurised, oxygen-blown, entrained flow slagging type. High pressure coal, oxygen and steam enter the gasifier through horizontally opposed burners at approximately 40 bar. The raw syngas produced in the reaction travels upwards through the gasifier at a temperature of approximately 1400 - 1500°C and exits together with entrained slag from the top of the gasifier. The gasification zone is lined with a membrane wall, in which medium pressure steam is generated. The high reaction temperature results in most of the mineral content of the feed being converted to molten slag, which forms a protective layer on the membrane wall. The slag flows down the wall and out of an opening at the bottom of the gasifier into a water quench chamber. The insulating properties of the slag help reduce heat loss to the membrane wall, ensuring a high cold gas efficiency with minimum production of CO_2 in the syngas. On contact with the water quench the slag forms dense, glassy granules which are washed, depressured and then sent to storage for recycle and disposal.

Hot syngas leaving the gasifier is quenched with cooled, filtered syngas to a temperature of approximately 750 °C. The syngas is cooled further by generating first superheated HP steam, then saturated MP steam. Entrained solid residue is filtered from the syngas before the stream splits with a partial recycle back to the gasifier. The remaining syngas is routed to the Syngas Scrubber vessel.

Any remaining particulate matter together with chlorides are washed from the syngas in the scrubber vessel. The water saturated syngas is heated first against syngas leaving the COS Hydrolysis Reactor, then against MP steam before entering the COS Hydrolysis Reactor where Carbonyl Sulphide (COS) is hydrolysed to H_2S (and CO_2).

$$COS + H_2O \leftrightarrows CO_2 + H_2S$$

The gas stream leaving the COS Hydrolysis Reactor is cooled first against the incoming gas and then against cooling water before entering the Acid Gas Removal Unit.

3.2.3.4 Sulphur Removal

The purpose of the acid gas removal unit is to remove CO_2 and H_2S from the syngas stream. In this scheme both CO_2 and H_2S are removed as a single product stream which is then routed to the Sulphur Recovery Unit.

The Syngas stream, combined with the tailgas from the Sulphur recovery Unit, is fed into the Acid Gas Removal Absorber where the gas is washed with lean Methyl Di-ethyl Amine (MDEA) solution. The solvent preferentially absorbs the H_2S and traces of CO_2 leaving the sweet syngas stream to exit the top of the absorber. The syngas stream is sent as fuel gas to the gas turbine generators. The Rich Amine leaving the bottom of the H_2S Absorber is then flashed in a drum to remove components such as hydrogen and nitrogen from the stream. The Rich MDEA is then passed through the Rich/Lean Amine Heat exchanger in order to bring it up to close to the Regeneration column feed temperature. Steam, produced in the reboilers, strips the H_2S and traces of CO_2 from the rich amine in the Regeneration Column. The water from the overhead acid gas stream is condensed against cooling water and returned to the column. The acid gas stream leaving the top of the Regeneration Column is routed to the Sulphur Recovery Unit. Lean amine leaving the bottom of the Regeneration Column is heat exchanged against incoming rich amine in the Lean/Rich Heat Exchanger before being cooled further against cooling water and returned to the H_2S Absorber, completing the circuit.

3.2.3.5 Sulphur Recovery Unit

Sour gas from the Acid Gas Removal Unit together with the sour gas stream from the Sour Water Stripper together with 95% purity oxygen from the Air Separation Unit and air is fed to the Claus combustor. The Claus combustor partially oxidises the H_2S to SO_2 and by ensuring the furnace temperature is maintained above 1350 °C, any NH_3 present is decomposed. The gases leaving the furnace section are cooled by generation of MP steam and any elemental sulphur formed is condensed and routed to the sulphur storage tank. The gas stream leaving the boiler section is cooled further by generating LP steam before entering the first Claus catalytic reactor. In the first Claus Catalytic Reactor H_2S is catalytically oxidised with SO_2 to elemental sulphur. The stream leaving the reactor is cooled by generation of LP steam. Any sulphur condensed is routed to sulphur storage. The gas stream is then heated against LP steam before entering the second Claus reactor. After reaction, the gas stream is cooled by generation of LP steam and condensed sulphur is routed to sulphur storage. The remaining tail gas is sent to the Tail Gas Treating Unit.

3.2.3.6 Tail Gas Treatment Unit

Tailgas from the Claus unit is combined with Syngas (H_2 rich gas) before being pre-heated against MP steam into the Hydrogenation Reactor. In the Hydrogenation Reactor, SO_2 and any elemental sulphur species are catalytically reduced in the presence of H_2 to H_2S. COS is also hydrolysed to H_2S. The hydrogenated stream is then cooled by generation of LP steam, quenched against water in a quench tower, before being compressed back to the Acid Gas Removal Unit.

A slipstream of sour water is removed and sent to the Sour Water Stripper to prevent accumulation in the circulating quench water.

3.2.3.7 Sour Water Stripper

Sour water received from various sources including syngas cooling and tailgas quench are processed in the Sour Water Stripper unit to produce a sour gas stream and a sweet water stream to be reused in the gasification process.

Sour water received in the Sour Stripper Column is reboiled against LP steam. The sour gas leaving the top of the stripper column is routed to the Sulphur Recovery Unit. The sweet water leaving the bottom of the column is recycled back to the Syngas Saturator column.

3.2.3.8 Combined Cycle Power Generation

The power island is based on two General Electric Frame 9F IGCC gas turbines, two heat recovery steam generators (HRSGs) generating steam at two pressure levels and a single steam turbine.

Syngas from the AGRU is heated against sour syngas in the GT Fuel Gas heater, E-2203 and subsequently against MP steam in the GT Fuel Steam Heater, E-2204, before being routed to the gas turbines. Nitrogen from the air separation unit is heated against sour syngas in Nitrogen Heater, E-2206 before being routed to the Gas Turbines. Dual fuel burners within the gas turbine allow start-up and shutdown to be carried out on natural gas when syngas supply is unavailable or unreliable.

The gas turbine exhaust gas flows to the Heat Recovery Steam Generator at approximately 530^{0}C. The thermal energy of the exhaust gases is used to superheat HP and MP steam, to generate HP and MP steam, heat the boiler feedwater, to generate LP steam in the LP boiler for use in the deaerator and to heat the vacuum condensate. The flue gases are emitted to atmosphere at approximately 85 °C.

The coil sequence in the HRSG is as follows:

- HP Superheater and MP Superheater Section 2 are arranged in parallel across the exhaust duct.
- HP Boiler
- MP Superheater Section 1
- HP Economiser
- MP Boiler
- BFW Heater
- LP Boiler
- Vacuum Condensate Heater

Exhaust gas is discharged to the atmosphere through the HRSG stack, which is fitted with a continuous emission monitoring system and a silencer.

Vacuum condensate from the pumps is heated first against raw syngas in E-2205 before being heated further against flue gases in E-3209 and entering the deaerator. A small quantity of LP steam is generated in the LP boiler, E-3208, from a slipstream of Boiler Feed Water from the deaerator to strip the oxygen from the system. The remaining boiler feed water is routed to the boiler feedwater pumps, P-3201, which delivers it via the boiler feed water heater, E-3207 to split, part of the stream routed to the MP Steam drum, D-3201. The remainder is heated further in the HP Economiser, E-3205 before entering the HP Steam Drum.

From the MP Steam Drum, MP Steam is generated in the MP Boiler, E-3206 and is superheated in the MP Superheater Section 1, E-3204 before combining with MP steam from the gasification unit and MP Steam from the exhaust of the HP Stage of the Steam Turbine.

From the HP Steam Drum, steam is raised in the HP Boiler, E-3203 before being combined with HP steam from the gasification unit and heated further in the HP Superheater, E-3201. Superheated steam is combined with the superheated steam from the second HRSG and is sent to the HP stage of the steam turbine, Z-3301. The exhaust steam from the HP module of the steam turbine is split between the two HRSGs, combined with steam from the MP Superheater section 1 together with steam from gasification unit and is superheated in the two MP Superheaters, E-3202 before being recombined and routed to the MP stage of the steam turbine. Exhaust steam from the MP stage of the steam turbine is routed to the LP stage of the steam turbine. Electricity is generated in the Steam Turbine Generator on a common shaft with the 3 stages of steam turbine. Exhaust steam from the LP stage of the steam turbine is condensed against seawater in the Vacuum Condenser, E-3210, before returning to the suction of the Vacuum Condensate Pumps, P-3202 completing the circuit.

3.2.3.9 Balance of Plant

The key balance of plant requirements for this scheme are the cooling water supply systems. A very large flow of cooling water is required to supply the steam turbine vacuum condenser. This duty is supplied using sea water in a once through flow scheme.

The ASU requires a significant quantity of cooling medium. Where this cannot be supplied using heat integration within or between the process units, cooling water is required. This cooling water is supplied as fresh cooling water in a closed circuit. The fresh water system is cooled against sea water.

3.2.4 Major Equipment List

A high level equipment list was developed for this case broken down according to unit. This equipment list is located in Appendix 4 of this report.

3.2.5 Capital Cost, Operating Cost and Economics

The economic results are outlined in the table below:

Figure 3-4 Economic Figures for Case 1B – IGCC without CO_2 Capture

		Case 1B
ASU	GB£M	129.9
Gasification / Boiler	GB£M	300.3
Power Island	GB£M	478.5
Acid Gas Removal	GB£M	87.5
CO_2 compression	GB£M	0.0
Others	GB£M	277.8
Total CAPEX	GB£M	1274
CAPEX efficiency	GB£/kWh	1708
Total OPEX – incl. fuel	GB£M p.a.	175.8
Total OPEX – excl. fuel	GB£M p.a.	64.0
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	235.6
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	85.8
Levelised Cost of Electricity		
CO_2 emission cost = £ 0 / te CO_2	£ / MWh_{Net}	59.49
CO_2 emission cost = £ 20 / te CO_2	£ / MWh_{Net}	74.18
CO_2 emission cost = £ 40 / te CO_2	£ / MWh_{Net}	88.87
CO_2 emission cost = £ 60 / te CO_2	£ / MWh_{Net}	103.56
Cost of CO_2 Captured		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	n/a
Cost of CO_2 Avoided		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	n/a

4. PULVERISED COAL POWER PLANT TECHNICAL DESCRIPTIONS

4.1 Case 2A – Ultra Supercritical plant with CO_2 Capture

4.1.1 Introduction

The overall process scheme for this case includes a pulverised coal fed ultra super critical boiler with downstream SCR for NOx removal, electrostatic precipitator (EP) for dust and particulate removal, FGD for SOx removal, CO_2 capture unit, and CO_2 compression and dehydration unit. Steam conditions are in the range 600-620°C. Consistent with the other cases the process conditions, including stream flows, pressures, temperatures and compositions, reflect that of a power plant nominally sized to produce the target net electrical power output. Key features of the configuration include:

- Power Island Unit – power island is a single train, composed of one ultra supercritical steam turbine and preheating line.

- Acid Gas Removal Unit – carbon dioxide removal scheme developed using in-house information on the basis of an MEA-based process such as Fluor Econamine FG+ carbon dioxide recovery technology. The unit includes a direct contact cooler incorporating chemical dosing to remove residual SOx to an appropriate level.

- Carbon Dioxide Compression and Drying Units – dehydration and compression to 150 barg based on in-house knowledge of commercially available equipment.

The carbon capture scheme is configured with two trains of MEA absorption, three trains of stripping and two trains of CO_2 compression and drying. The absorption trains are sized based upon the maximum size of the absorption column in the region of 15m diameter (larger column diameters up to 20m have been suggested where the vessel can be constructed on-site). The number of stripping trains was selected based upon the heat input required for the stripper reboilers with a maximum total reboiler duty of 200 MWth per train (this is based upon 4 x 50 MWth reboilers located around the column base). The number of CO_2 compression trains was selected based upon in house data and to keep a consistent order of compressor size across all of the cases considered in the scope of this study.

The lean/rich solvent exchanger, also known as the cross-over exchanger, is another very large and key equipment item in the post-combustion carbon capture scheme. This duty is most commonly met using a plate and frame type heat exchanger in the smaller scale plants currently in operation. A feature of this type of exchanger is its relative simplicity of scale up, achieved by adding frames and increasing the area of each frame. While it is unlikely that an exchanger of this type has yet been operated at the scale required for the benchmark cases, previous Foster Wheeler work with technology providers has shown that the sizes envisaged in this study are not infeasible (this case was calculated to require 3 x 11710m^2 heat transfer surface area exchangers with a duty of 120MW each).

4.1.2 Plant Performance

The plant performance is outlined in the table below:

Figure 4-1 Performance Figures for Case 2A – USCPC with CO_2 Capture

Case		**$2A_{85\%}$**	**$2A_{90\%}$**	**$2A_{95\%}$**
Power				
Total gross installed capacity	MWe	750.7	745.7	733.9
Gas Turbine (s)	MWe	0.0	0.0	0.0
Steam Turbine	MWe	750.7	745.7	733.9
Others	MWe	0.0	0.0	0.0
Total auxiliary loads	MWe	119.3	119.0	126.5
ASU	MWe	0.0	0.0	0.0
Gasification / Boiler	MWe	0.0	0.0	0.0
Power Island	MWe	29.0	29.0	29.0
Acid Gas Removal	MWe	7.8	7.4	8.1
CO_2 compression	MWe	49.1	48.6	54.5
Others	MWe	33.4	34.0	35.0
Net Power Export	MWe	631.3	626.7	607.3
Net Efficiency (LHV)	%	34.7	34.4	33.4
Heat Rate	kJ/kWh	10378	10455	10788
Flows				
Total fuel feed rate	tpd	6163.4	6163.4	6163.4
Oxygen consumption	tpd	0	0	0
Water consumption	tpd	3797	3828	3850
Cooling water (once through)	tpd	2918213	2,998,188	3113263
Carbon Balance				
Total carbon in feeds	tpd	3981.4	3981.4	3981.4
Total carbon captured	tpd	3415.0	3581.5	3794.5
Carbon capture rate	%	85.8	90.0	95.3
Total CO_2 captured	tpd	12514	13124	13905
Total CO_2 emitted	tpd	2074	1465	685
CO_2 emissions	g CO_2/ kWh_{Net}	137.0	97.4	47.0

The specific steam requirement for the 90% case was calculated as 1.31 kg steam / kg CO_2 captured.

No changes to the overall process scheme were made in order to vary the % CO_2 capture (CC). In order to capture more CO_2 the flow rate of solvent was increased, resulting in higher steam and power loads in the CO_2 capture and compression units. The figures above show that the changes in parasitic steam and power load, and hence overall power plant efficiency, were not a linear function of the %CC. In particular more energy, per unit CO_2, was required to achieve 95%CC than 90%CC or 85%.

4.1.3 Process Description

4.1.3.1 Coal Receiving, Handling, Preparation and Feeding

Coal is unloaded from the trains or barges and stored on-site. Vibratory feeders transfer the coal from the storage pile onto conveyors which in turn transport the coal to the preparation and feeding area.

Raw coal is crushed and fed to a pulveriser. Preheated air is introduced to the coal as it is being ground, then the air / coal mixture is fed to the boiler.

4.1.3.2 Coal Combustion System

The coal boiler system consists of a once through, wall-fired, balanced draft ultra-supercritical pulverised coal fired plant.

Forced draft fans supply air to the burner system via air preheaters in which the air is heated against the boiler flue gas on its way to the stack. This air is distributed to the burner windbox as secondary air. Air from the primary air fans is also preheated against flue gas and is used as combustion air, a portion of this air bypasses the preheaters and is used for tempering the pulverisers' fuel-air mixture outlet temperature.

The pulverised air-coal mixture is fed to the boiler via nozzles at various elevations. Each burner is designed as a low-NOx configuration and the boiler will be equipped with at least one level of overfire air nozzles to add additional air to cool the rising gases and inhibit further NOx formation.

The hot combustion products rise to the top of the boiler and pass horizontally through the secondary superheater and reheater. The gases then pass downwards over the primary superheater, economiser and air preheater.

4.1.3.3 Steam Turbine, Condensate and Feedwater Systems

Boiler feed water (BFW) passes through an economiser coil before being fed to the lower and then upper sections of the boiler water walls. The fluid then passes through the convection circuit followed by the primary and secondary superheaters. This superheated high pressure (HP) steam is then fed to the HP section of the steam turbine.

The exhaust of the HP steam turbine section is at medium pressure (MP) and is reheated in the boiler before being fed to the MP steam turbine section. Exhaust from the MP section is fed to the inlet of the low pressure (LP) section of the turbine. A portion of the steam from the MP to LP crossover is supplied to the MEA stripper columns, from which hot condensate is returned and added to the boiler feed water.

The LP steam turbine exhaust is condensed against sea water at sub-atmospheric pressure before being, treated and pumped to the BFW heaters. The first stage of BFW heating is supplied with hot water from the MEA unit, this water is cooled sufficiently to be used as cooling water for the flashed gas cooler and several of the compressor intercooler stages. The subsequent BFW heating consists of 8 heaters with deaeration and HP pumps between stages 4 and 5. The heat is supplied by steam extracted from the steam turbines at appropriate levels to supply each temperature required. It is possible to bypass the first two steam extraction BFW heaters while all three trains of carbon capture are in operation. The BFW is then ready to be fed back to the boiler.

4.1.3.4 Ash Handling System

The boiler is equipped with MP steam soot-blowing equipment in an array of retractable nozzles and lances. The furnace bottom comprises several refractory lined hoppers with water seal and cooling which feed the ash handling system.

4.1.3.5 NOx Prevention and Removal

The low-NOx burners and air overfiring in the boiler reduce the NOx concentration significantly, however, a selective catalytic reduction (SCR) unit is also required. The SCR uses ammonia and a catalyst to reduce the NOx to N_2 and water. The ammonia is injected into the gas immediately prior to entering the reactor vessel which holds the catalyst. The SCR system consists of a reactor vessel, ammonia storage and injection and gas flow control.

4.1.3.6 Electrostatic Precipitator

The NOx reduced flue gas then passes to an electrostatic precipitator (ESP) for particulates removal. The electrodes produce an ionising corona through which the flue gas passes. The ionised particles follow the negative electric field and move downwards where they are collected and removed to the ash handling system.

4.1.3.7 Flue Gas Desulphurisation

Downstream of the ESP induced draft fans (ID) ensure flue gas flow. The flue gas then enters a wet limestone forced oxidation flue gas desulphurisation (FGD) unit which removes 98% of the sulphur. The FGD system comprises a gas/gas heat exchanger followed by an absorber module. In the gas/gas exchanger the hot flue gas is cooled against the flue gas exiting the MEA carbon capture unit. The cooled gas entering the bottom of the absorber module is typically contacted with multiple sprays of limestone slurry with trays or packing providing enhanced contact. Finally the flue gas passes through moisture separators and on to the MEA carbon capture unit.

The slurry is collected in a large inventory of liquid held in the bottom of the absorber module. Air is added to this liquid inventory to promote oxidation to calcium sulphate (gypsum). The slurry is continuously agitated to ensure mixing and prevent solids settling out. Slurry is recirculated to the absorber spray section via recirculation pumps. Fresh make-up and spend reagent pumps are used to ensure the chemical balance is maintained in the absorber module. The gypsum slurry is dewatered in a slacking system.

4.1.3.8 CO_2 Removal

Flue gas is fed to a direct contact cooler (DCC) where much of the water present in the flue gas stream condenses as the gas is cooled to 30°C. The condensate is then recirculated through a cooler and returned to the contact tower. A small quantity of sodium hydroxide is added to the recirculating water in order to ensure that the remaining SO_2 in the flue gas is removed to meet the <10 ppm specification to prevent excessive solvent losses. Precipitates and excess water are removed from the system to waste water treatment.

A blower then boosts the pressure of the cooled flue gas sufficiently to overcome the pressure drop in the absorption column. In the lower portion of the column the flue gas is contacted with semi-lean and then lean amine which absorbs approximately 90% of the CO_2 content of the flue gas. This section also incorporates an extraction and cooling loop in order to ensure the cooler conditions more favourable to CO_2 absorption. In the top of the column the flue gas is washed with water to prevent solvent losses to the atmosphere. The flue gas is routed back to the gas / gas heat exchanger in the FGD unit, to ensure its temperature is sufficient for dispersion, then is released to atmosphere via the stack.

The CO_2 rich solvent stream exits the bottom of the absorber column and is pumped up to approximately 5 bara. The stream is then split, with approximately 25% of the flow passing through 2 stages of heating against warmer solvent streams before being flashed at a pressure of 1.3 bara. The semi lean solvent from the flash drum is then cooled against rich solvent and returned to the absorption column with the cooled extracted solvent. The remaining rich solvent is heated against lean solvent in the cross over exchanger and introduced to the stripper column.

In the stripper column the CO_2 desorbs from the rich solvent as it is heated producing a stream of hot lean solvent from the bottom of the stripper. This lean solvent is cooled against rich solvent and returned to the absorption column. The stripper overheads are cooled to 30°C, condensing a significant quantity of water, some of which is returned to the stripper as reflux with the rest being sent to treatment or recovery.

4.1.3.9 CO_2 Compression and Drying

The acid gas resulting from the semi lean amine flash is compressed in the first of 8 compression stages, after which it is cooled and passed through a knock out drum. After the first compression stage the main CO_2 stream, from the stripper column, is added to the flashed acid gas stream for all the subsequent compression steps. Between each of the next 4 steps is a cooler and knock out drum up to a pressure of 25 bara. The CO_2 is then dried by molecular sieve adsorption to reach the specification of < 50 ppmv moisture.

The final 3 compression stages include intercoolers and an after cooler and result in a final CO_2 product at specification of 150 barg and 30ºC.

Based on the total plant flue gas flow rate of this case three trains were selected for absorption, stripping and compression. This was determined by the maximum sizes of key equipment items; absorber diameter, stripper heat load and CO_2 compressor capacity.

4.1.3.10 Balance of Plant

The key balance of plant requirements for this scheme are the cooling water supply systems. A very large flow of cooling water is required to supply the steam turbine vacuum condenser. This duty is supplied using sea water in a once through flow scheme.

The AGRU and CO_2 compression and drying units also require a significant quantity of cooling medium. Where this cannot be supplied using heat integration within or between the process units, cooling water is required. This cooling water is supplied as fresh cooling water in a closed circuit. The fresh water system is cooled against sea water.

Facilities are also required for storage and make-up of the MEA based solvent to the AGRU. Reuse and treatment of the numerous, mainly small, water streams produced from the cooling of water saturated gas streams are integrated with the units where possible. Streams containing contaminants such as MEA and sodium sulphate are routed to an effluent treatment system.

4.1.3.11 Steam Systems and Heat Integration

The MEA process flow scheme was developed to maximise heat integration within the solvent loop system where much of this integration is aimed at minimising the heat input required to the stripper. The remaining heat required for the stripper is supplied as LP steam taken from the IP to LP steam turbine crossover line in the power island. This results in lower steam flow through the LP section of the turbine and hence lower power generation overall. However, some of the resultant loss in efficiency can be made up in other ways.

A significant quantity of cooling is required in both the MEA solvent scheme and in the CO_2 compressor intercoolers. Replacing the normal sea cooling water in a few key locations, such as the compressor intercoolers and the flashed acid gas cooler, with a fresh water closed loop system enables this heat to be transferred to the power island where it can be utilised for boiler feed water preheating. This reduces the quantity of LP steam extracted from the LP turbine which is required for boiler feed water preheating, hence increasing the overall power output and recovering some of the efficiency loss created by extracting LP steam for use in the stripper reboiler.

4.1.4 Plant Turndown

This case is potentially capable of turn down to outputs below 50%, sharing the characteristics of conventional steam plant. The turn-down capacity of the CO_2 compressors will typically be limited to 70% without recourse to suction recycle.

4.1.5 Major Equipment List

A high level equipment list was developed for this case broken down according to unit. This equipment list is located in Appendix 4 of this report.

4.1.6 Capital Cost, Operating Cost and Economics

The economic results are outlined in the table below:

Figure 4-2 Economic Figures for Case 2A – USCPC with CO_2 Capture

Case		**$2A_{85\%}$**	**$2A_{90\%}$**	**$2A_{95\%}$**
ASU	GB£M	0.0	0.0	0.0
Gasification / Boiler	GB£M	779.4	779.4	779.4
Power Island	GB£M	221.2	211.2	211.2
Acid Gas Removal	GB£M	356.0	366.9	381.0
CO_2 compression	GB£M	123.0	126.9	132.0
Others	GB£M	122.0	123.0	125.0
Total CAPEX	GB£M	1591.0	1607.4	1628.0
CAPEX efficiency	GB£/kWh	2520	2565	2681
Total OPEX – incl. fuel	GB£M p.a.	210.1	211.5	213.1
Total OPEX – excl. fuel	GB£M p.a.	85.8	87.2	88.8
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	332.8	339.4	350.9
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	135.9	139.9	146.2
Levelised Cost of Electricity				
CO_2 emission cost = £ 0 / te CO_2	£ / MWh_{Net}	86.36	87.80	90.63
CO_2 emission cost = £ 20 / te CO_2	£ / MWh_{Net}	89.10	89.76	91.57
CO_2 emission cost = £ 40 / te CO_2	£ / MWh_{Net}	91.84	91.73	92.50
CO_2 emission cost = £ 60 / te CO_2	£ / MWh_{Net}	94.58	93.69	93.44
Cost of CO_2 Captured				
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	36.48	35.98	36.04
Cost of CO_2 Avoided				
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	46.83	46.28	46.90

The CAPEX and OPEX figures for the varied % CO_2 capture (CC) cases reflect the trends shown in the power plant performance figures; costs were not a linear function of the %CC, in particular slightly more total cost was incurred, to achieve 95%CC than 90%CC or 85%.

Two variables appear to be in balance in this scenario; the effect of equipment economy of scale and the effect of beginning to approach the limit at which parasitic loads increase at greater than linear rate with increasing %CC. A slight minimum at 90%CC is seen in the cost of CO_2 captured and avoided in this case, showing that the dominance of the two variables changes either side of the 90%CC point.

4.2 Case 2B – Ultra Supercritical plant without CO_2 Capture

4.2.1 Introduction

An equivalent pulverized coal power plant without CO_2 capture benchmark will be developed. This will be based upon the same configuration as above, with the exclusion of the AGR and CO_2 compression and drying units. The case will be sized using the same coal feed rate as the Pulverised Coal Power Plant with CO_2 capture.

4.2.2 Plant Performance

The plant performance is outlined in the table below:

Figure 4-3 Performance Figures for Case 2B – USCPC without CO_2 Capture

Case		2B
Power		no CCS
Total gross installed capacity	MWe	836.2
Gas Turbine (s)	MWe	0.0
Steam Turbine	MWe	836.2
Others	MWe	0.0
Total auxiliary loads	MWe	57.0
ASU	MWe	0.0
Gasification / Boiler	MWe	0.0
Power Island	MWe	29.1
Acid Gas Removal	MWe	0.0
CO_2 compression	MWe	0.0
Others	MWe	27.9
Net Power Export	MWe	779.2
Net Efficiency (LHV)	%	42.8
Heat Rate	kJ/kWh	8409
Flows		
Total fuel feed rate	tpd	6163.4
Oxygen consumption	tpd	0
Water consumption	tpd	3379
Cooling water (once through)	tpd	2,283,811
Carbon Balance		
Total carbon in feeds	tpd	3981.4
Total carbon captured	tpd	0.0
Carbon capture rate	%	0.0
Total CO_2 captured	tpd	0
Total CO_2 emitted	tpd	14590
CO_2 emissions	g CO_2/ kWh_{Net}	780.2

4.2.3 Process Description

4.2.3.1 Coal Receiving, Handling, Preparation and Feeding

Coal is unloaded from the trains or barges and stored on-site. Vibratory feeders transfer the coal from the storage pile onto conveyors which in turn transport the coal to the preparation and feeding area.

Raw coal is crushed and fed to a pulveriser. Preheated air is introduced to the coal as it is being ground, then the air / coal mixture is fed to the boiler.

4.2.3.2 Coal Combustion System

The coal boiler system consists of a once through, wall-fired, balanced draft ultra-supercritical pulverised coal fired plant.

Forced draft fans supply air to the burner system via air preheaters in which the air is heated against the boiler flue gas on its way to the stack. This air is distributed to the burner windbox as secondary air. Air from the primary air fans is also preheated against flue gas and is used as combustion air, a portion of this air bypasses the preheaters and is used for tempering the pulverisers' fuel-air mixture outlet temperature.

The pulverised air-coal mixture is fed to the boiler via nozzles at various elevations. Each burner is designed as a low-NOx configuration and the boiler will be equipped with at least one level of overfire air nozzles to add additional air to cool the rising gases and inhibit further NOx formation.

The hot combustion products rise to the top of the boiler and pass horizontally through the secondary superheater and reheater. The gases then pass downwards over the primary superheater, economiser and air preheater

4.2.3.3 Steam Turbine, Condensate and Feedwater Systems

Boiler feed water (BFW) passes through an economiser coil before being fed to the lower and then upper sections of the boiler water walls. The fluid then passes through the convection circuit followed by the primary and secondary superheaters. This superheated high pressure (HP) steam is then fed to the HP section of the steam turbine.

The exhaust of the HP steam turbine section is at medium pressure (MP) and is reheated in the boiler before being fed to the MP steam turbine section. Exhaust from the MP section is fed to the inlet of the low pressure (LP) section of the turbine.

The LP steam turbine exhaust is condensed against sea water at sub-atmospheric pressure before being, treated and pumped to the BFW heaters. BFW heating consists of 8 heaters with deaeration and HP pumps between stages 4 and 5. The heat is supplied by steam extracted from the steam turbines at appropriate levels to supply each temperatures required. The BFW is then ready to be fed back to the boiler.

4.2.3.4 Ash Handling System

The boiler is equipped with MP steam soot-blowing equipment in an array of retractable nozzles and lances. The furnace bottom comprises several refractory lined hoppers with water seal and cooling which feed the ash handling system.

4.2.3.5 NOx Prevention and Removal

The low-NOx burners and air overfiring in the boiler reduce the NOx concentration significantly, however, a selective catalytic reduction (SCR) unit is also required. The SCR uses ammonia and a catalyst to reduce the NOx to N_2 and water. The ammonia is injected into the gas immediately prior to entering the reactor vessel which holds the catalyst. The SCR system consists of a reactor vessel, ammonia storage and injection and gas flow control.

4.2.3.6 Electrostatic Precipitator

The NOx reduced flue gas then passes to an electrostatic precipitator (ESP) for particulates removal. The electrodes produce an ionising corona through which the flue gas passes. The ionised particles follow the negative electric field and move downwards where they are collected and removed to the ash handling system.

4.2.3.7 Flue Gas Desulphurisation

Downstream of the ESP induced draft fans (ID) ensure flue gas flow. The flue gas then enters a wet limestone forced oxidation flue gas desulphurisation (FGD) unit which removes 98% of the sulphur. The FGD system comprises a gas/gas heat exchanger followed by an absorber module. In the gas/gas exchanger the hot flue gas is cooled against the flue gas exiting the FGD unit. The cooled gas entering the bottom of the absorber module is typically contacted with multiple sprays of limestone slurry with trays or packing providing enhanced contact. Finally the flue gas passes through moisture separators, the gas/gas exchanger and is discharged to atmosphere via the stack.

The slurry is collected in a large inventory of liquid held in the bottom of the absorber module. Air is added to this liquid inventory to promote oxidation to calcium sulphate (gypsum). The slurry is continuously agitated to ensure mixing and prevent solids settling out. Slurry is recirculated to the absorber spray section via recirculation pumps. Fresh make-up and spend reagent pumps are used to ensure the chemical balance is maintained in the absorber module. The gypsum slurry is dewatered in a slacking system.

4.2.3.8 Balance of Plant

The key balance of plant requirements for this scheme are the cooling water supply systems. A very large flow of cooling water is required to supply the steam turbine vacuum condenser. This duty is supplied using sea water in a once through flow scheme. Any other cooling duties are supplied using fresh cooling water, which is itself cooled against sea water.

4.2.4 Major Equipment List

A high level equipment list was developed for this case broken down according to unit. This equipment list is located in Appendix 4 of this report.

4.2.5 Capital Cost, Operating Cost and Economics

The economic results are outlined in the table below:

Figure 4-4 Economic Figures for Case 2B – USCPC without CO_2 Capture

		Case 2B
ASU	GB£M	0.0
Gasification / Boiler	GB£M	770.7
Power Island	GB£M	259.5
Acid Gas Removal	GB£M	0.0
CO_2 compression	GB£M	0.0
Others	GB£M	109.7
Total CAPEX	GB£M	1140
CAPEX efficiency	GB£/kWh	1463
Total OPEX – incl. fuel	GB£M p.a.	187.3
Total OPEX – excl. fuel	GB£M p.a.	62.9
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	240.4
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	80.7
Levelised Cost of Electricity		
CO_2 emission cost = £ 0 / te CO_2	£ / MWh_{Net}	56.24
CO_2 emission cost = £ 20 / te CO_2	£ / MWh_{Net}	71.84
CO_2 emission cost = £ 40 / te CO_2	£ / MWh_{Net}	87.44
CO_2 emission cost = £ 60 / te CO_2	£ / MWh_{Net}	103.05
Cost of CO_2 Captured		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	n/a
Cost of CO_2 Avoided		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	n/a

5. NATURAL GAS COMBINED CYCLES TECHNICAL DESCRIPTIONS

5.1 Case 3A – NGCC, G Class Turbine with CO_2 Capture

5.1.1 Introduction

The overall process scheme was based upon a natural gas fired combined cycle gas turbine (CCGT) using two G class gas turbines featuring dry low NOx (DLN) burners, each with downstream heat recovery steam generator (HRSG), and common single steam turbine generator (STG), CO_2 capture unit and CO_2 compression and dehydration unit.

In this case the natural gas feed rate was set to ensure full utilisation of the gas turbines with the supporting and downstream equipment items sized to process the generated gas turbine exhaust gas. The process conditions, including stream flows, pressures, temperatures and compositions, were produced to reflect this sizing basis. Key features of the configuration include:

- Power Island Unit – comprising of two parallel trains, each with one G class 50 Hz gas turbine and one heat recovery steam generator (HRSG), connected to a single condensing steam turbine, using seawater cooling.

- Acid Gas Removal Unit – CO_2 removal scheme developed using in-house information on the basis of an MEA-based process such as Fluor Econamine FG+ CO_2 recovery technology.

- Carbon Dioxide Compression and Drying Units – dehydration and compression to 150 barg based on in-house knowledge of commercially available equipment.

The carbon capture scheme is configured with three trains of MEA absorption, two trains of stripping and two trains of CO_2 compression and drying. The absorption trains are sized based upon the maximum size of the absorption column in the region of 15m diameter (larger column diameters up to 20m have been suggested where the vessel can be constructed on-site). The number of stripping trains was selected based upon the heat input required for the stripper reboilers with a maximum total reboiler duty of 200 MWth per train (this is based upon 4 x 50 MWth reboilers located around the column base). The number of CO_2 compression trains was selected based upon in house data and to keep a consistent order of compressor size across all 7 cases considered in the scope of this study.

The lean/rich solvent exchanger, also known as the cross-over exchanger, is another very large and key equipment item in the post-combustion carbon capture scheme. This duty is most commonly met using a plate and frame type heat exchanger in the smaller scale plants currently in operation. A feature of this type of exchanger is its relative simplicity of scale up, achieved by adding frames and increasing the area of each frame. While it is unlikely that an exchanger of this type has yet been operated at the scale required for the benchmark cases, previous Foster Wheeler work with technology providers has shown that the sizes envisaged in this study are not infeasible (this case was calculated to require 2 x 9460m^2 heat transfer surface area exchangers with a duty of 109MW each).

5.1.2 Plant Performance

The plant performance for this case was assessed at four different levels of CO_2 capture; firstly at 90% carbon capture, resulting in emissions to atmosphere of 41g CO_2/kWh_{Net} electricity exported to the grid. Secondly the CCGT case was assessed with 75% carbon capture in order to achieve emissions to atmosphere in the range of 90 to 100 g CO_2/kWh_{Net} electricity export, comparable with the emissions per kWh resulting from the coal cases. Finally cases at 85% and 95% carbon capture were developed resulting in emissions of 61g CO_2/kWh_{Net} and 20g CO_2/kWh_{Net} respectively. The performance figures for these four cases are outlined in the table below:

Figure 5-1 Performance Figures for Cases 3A & 3A(i) – NGCC with CO_2 Capture

		Case 3A (i) – with 75% CC	Case 3A – with 85% CC	Case 3A– with 90% CC	Case 3A – with 95% CC
Power					
Total gross installed capacity	MWe	970.7	959.4	954.6	947.3
Gas Turbine (s)	MWe	729.1	729.1	729.1	729.1
Steam Turbine	MWe	241.6	230.3	225.5	218.1
Others	MWe	0.0	0.0	0.0	0.0
Total auxiliary loads	MWe	98.4	108.5	110.6	112.9
ASU	MWe	0.0	0.0	0.0	0.0
Gasification / Boiler	MWe	0.0	0.0	0.0	0.0
Power Island	MWe	39.4	39.3	39.2	39.2
Acid Gas Removal	MWe	27.4	32.7	32.8	33.0
CO_2 compression	MWe	23.2	27.1	28.6	30.3
Others	MWe	8.4	9.5	10.0	10.4
Net Power Export	MWe	872.3	850.9	844.0	834.3
Net Efficiency (LHV)	%	51.7	50.5	50.05	49.5
Heat Rate	kJ/kWh	6959	7135	7129	7276
Flows					
Total fuel feed rate	tpd	3070.4	3070.4	3070.4	3070.4
Oxygen consumption	tpd	0	0	0	0.0
Water consumption	tpd	3084	2947	3110	3293
Cooling water (once through)	tpd	1,429,008	1,498,824	1,553,050	1,631,923
Carbon Balance					
Total carbon in feeds	tpd	2272.8	2272.8	2272.8	2272.8
Total carbon captured	tpd	1709.8	1941.8	2053.5	2174.2
Carbon capture rate	%	75.2	85.4	90.3	95.7
Total CO_2 captured	tpd	6265	7115.6	7524.8	7967.2
Total CO_2 emitted	tpd	2063	339.5	835	392.5
CO_2 emissions	g CO_2/kWh_{Net}	98.5	60.9	41.2	19.6

In order to arrive at the 75% carbon capture the 90% carbon capture model was modified to allow a portion of the flue gas to bypass the MEA unit and flow directly to the stack. This resulted in a reduction in the solvent circulation rate and blower power of the order of 17%.

The specific steam requirement for the 90% carbon capture case was calculated as 1.50 kg steam / kg CO_2 captured with a figure of 1.46 kg steam / kg CO_2 captured for the 75% carbon capture case.

No changes to the overall process scheme were made in order to vary the % CO_2 capture (CC) between 85% and 95%. In order to capture more CO_2 the flow rate of solvent was increased, resulting in higher steam and power loads in the CO_2 capture and compression units. The figures above show that the changes in parasitic steam and power load, and hence overall power plant efficiency, were almost exactly linear as a function of the %CC, suggesting that the process is some margin away from its maximum possible %CC.

5.1.3 Process Description

5.1.3.1 Gas Turbines, Heat Recovery Steam Generators and Steam Turbine

The power island is based on two Mitsubishi Heavy Industries (MHI) M701G2 natural gas fed gas turbines, each with its own heat recovery steam generator (HRSG). The two HRSGs are identical and are configured to generate steam at three pressure levels with full reheat of medium pressure steam. A single steam turbine receives the steam from both HRSGs and is equipped with a vacuum condenser and condensate treatment.

Natural gas is received from across the plant battery limits via a metering station before being heated against MP boiler feed water (BFW) and fed to the gas turbines (GTs).

The GT exhaust gases flow to the Heat Recovery Steam Generator, without additional duct firing. The thermal energy of the exhaust gases is used to raise and superheat steam at 3 pressure levels as well as preheating condensate and heating the BFW. The flue gases, at approximately 93 °C, leaving the HRSG are pressurised using a blower in order to overcome the pressure drop through the MEA based Acid Gas Removal unit. Once the CO_2 has been removed the flue gases are reheated against the hot flue gas from the HRSG to cool the gas entering the AGRU and ensure that the treated flue gases are warm enough for dispersion via the stack.

The coil sequence in the HRSG is summarised as follows:

- 2nd HP Superheater
- 2nd MP Reheater
- 1st HP Superheater
- 1st MP Reheater
- HP Evaporator
- MP Superheater
- 2nd HP Economiser
- MP Evaporator
- LP Superheater
- MP Economiser
- 1st HP Economiser
- LP Evaporator
- LP Economiser

- Condensate Preheater

Condensate from the steam turbine condenser is preheated and deaerated using LP steam in the deaerator. Boiler feed water from the deaerator is pumped up to the three pressure levels required by the boiler feed water pumps.

In the HP circuit the BFW is pumped to approximately 14,000 kPa, passing through the 1st and 2nd HP Economiser into the HP Steam Drum. Water from the HP Steam Drum passes through the HP Evaporator coil generating saturated HP steam which returns to the HP Steam Drum before passing through the 1st and 2nd HP Superheaters and then to the HP inlet of the Steam Turbine.

The MP BFW pumps pump BFW to approximately 3000 kpa, through the MP Economiser and into the MP Steam Drum. Water from the MP Steam Drum passes through the MP Evaporator generating MP steam which is returned to the MP Steam Drum before entering the MP Superheater. Exhaust steam from the HP stage Steam Turbine are combined with superheated MP steam which is subsequently further superheated in the 1st and 2nd MP Reheaters before being routed to the MP stage of the Steam Turbine.

Desuperheaters between the two HP superheaters and the two MP reheaters use boiler feed water to control the second superheater outlet temperatures to 565ºC for both pressure levels.

The LP BFW pumps pump the BFW to approximately 450 kpa, through the LP Economiser and into the LP Steam Drum. Water from the LP Steam Drum passes through the LP Evaporator generating LP steam which is returned to the LP Steam Drum before entering the LP Superheater. The superheated LP Steam is then split, with a portion of being used to supply the heat required for the Stripper reboiler in the AGRU with the remaining LP Steam being routed to the LP inlet of the Steam Turbine.

The exhaust gases from the LP stage of the steam turbine are combined with condensate from the Natural Gas Preheater and the condensate return from the AGRU Stripper Reboiler before being fully condensed against seawater in the Vacuum Condensate Condenser. The vacuum condensate is then returned to the Vacuum Condensate Pumps completing the circuit.

5.1.3.2 CO_2 Removal

Flue gas is fed to a direct contact cooler (DCC) where much of the water present in the flue gas stream condenses as the gas is cooled to 30°C. The condensate is then recirculated through a cooler and returned to the contact tower. A small quantity of sodium hydroxide is added to the recirculating water in order to ensure that the remaining SO_2 in the flue gas is removed to meet the <10 ppm specification to prevent excessive solvent losses. Precipitates and excess water are removed from the system to waste water treatment.

A blower then boosts the pressure of the cooled flue gas sufficiently to overcome the pressure drop in the absorption column. In the lower portion of the column the flue gas is contacted with semi-lean and then lean amine which absorbs approximately 90% of the CO_2 content of the flue gas. This section also incorporates an extraction and cooling loop in order to ensure the cooler conditions more favourable to CO_2 absorption. In the top of the column the flue gas is washed with water to prevent solvent losses to the atmosphere. The flue gas is routed back to the gas / gas heat exchanger in the FGD unit, to ensure its temperature is sufficient for dispersion, then is released to atmosphere via the stack.

The CO_2 rich solvent stream exits the bottom of the absorber column and is pumped to approximately 5 bara. The stream is then split, with approximately 25% of the flow passing through 2 stages of heating against warmer solvent streams before being flashed at a pressure of 1.3 bara. The semi lean solvent from the flash drum is then cooled against rich solvent and returned to the absorption column with the cooled extracted solvent. The remaining rich solvent is heated against lean solvent in the cross over exchanger and introduced to the stripper column.

In the stripper column the CO_2 desorbs from the rich solvent as it is heated producing a stream of hot lean solvent from the bottom of the stripper. This lean solvent is cooled against rich solvent and returned to the absorption column. The stripper overheads are cooled to 30°C, condensing a significant quantity of water, some of which is returned to the stripper as reflux with the rest being sent to treatment or recovery.

5.1.3.3 CO_2 Compression and Drying

The acid gas resulting from the semi lean amine flash is compressed in the first of 8 compression stages, after which it is cooled and passed through a knock out drum. After the first compression stage the main CO_2 stream, from the stripper column, is added to the flashed acid gas stream for all the subsequent compression steps. Between each of the next 4 steps is a cooler and knock out drum up to a pressure of 25 bara.

The CO_2 is then dried by molecular sieve adsorption to reach the specification of < 50 ppmv moisture. Two dehydration vessels are required since one bed will be in use whilst the second bed will be in regeneration. The regeneration cycle uses a slipstream of dried gas exiting the operating molecular sieve bed. The gas is heated using the returning regeneration gas exiting the molecular sieve bed in regeneration. It is further heated under temperature control in an electric heater before entering the bed in a counter flow direction. The wet gas leaving the bed is cooled against incoming gas, any condensed water is separated in a knock out drum before it is passed through a fines filter and returned upstream of the 3rd stage compressor. The absorbent regeneration process takes several hours. When complete the heater is bypassed and the bed is cooled down over several hours before return to operation.

The final 3 compression stages include intercoolers and an after cooler and result in a final CO_2 product at specification of 150 barg and 30ºC.

5.1.3.4 Balance of Plant

The key balance of plant requirements for this scheme are the cooling water supply systems. A very large flow of cooling water is required to supply the steam turbine vacuum condenser. This duty is supplied using sea water in a once through flow scheme.

The AGRU and CO_2 compression and drying units also require a significant quantity of cooling medium. Where this cannot be supplied using heat integration within or between the process units, cooling water is required. This cooling water is supplied as fresh cooling water in a closed circuit. The fresh water system is cooled against sea water.

Facilities are also required for storage and make up of the MEA based solvent to the AGRU. Reuse and treatment of the numerous, mainly small, water streams produced from the cooling of water saturated gas streams are integrated with the units where possible. Streams containing contaminants such as MEA are routed to an effluent treatment system.

5.1.4 Plant Turndown

This case should as easy to operate at reduced capacity as a normal CCGT plant, approximately 35% or even lower. There is now more emphasis on operating gas-fired units at part load, to avoid the high maintenance aspects of repeated stop-start operation.

5.1.5 Major Equipment List

A high level equipment list was developed for this case broken down according to unit. This equipment list is located in Appendix 4 of this report.

5.1.6 Capital Cost, Operating Cost and Economics

The economic results are outlined in the table below:

Figure 5-2 Economic Figures for Cases 3A & 3A(i) – NGCC with CO_2 Capture

		Case 3A(i) - with 75% CC	Case 3A - with 85% CC	Case 3A - with 90% CC	Case 3A - with 95% CC
ASU	GB£M	0.0	0.0	0.0	0.0
Gasification / Boiler	GB£M	0.0	0.0	0.0	0.0
Power Island	GB£M	447.5	448.0	447.6	448.0
Acid Gas Removal	GB£M	299.6	326.0	337.3	351.0
CO_2 compression	GB£M	86.7	94.0	97.3	101.0
Others	GB£M	141.8	139.0	141.8	146.0
Total CAPEX	GB£M	976	1007	1024	1046
CAPEX efficiency	GB£/kWh	1118	1184	1216	1254
Total OPEX – incl. fuel	GB£M p.a.	296.0	297.2	297.7	298.5
Total OPEX – excl. fuel	GB£M p.a.	43.4	44.6	45.2	45.9
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	339.3	349.3	352.8	357.8
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	49.8	52.4	53.6	55.0
Levelised Cost of Electricity					
CO_2 emission cost = £0 / te CO_2	£ / MWh_{Net}	65.89	68.41	69.43	70.83
CO_2 emission cost = £20 / te CO_2	£ / MWh_{Net}	67.86	69.60	70.22	71.19
CO_2 emission cost = £40 / te CO_2	£ / MWh_{Net}	69.84	70.78	71.01	71.55
CO_2 emission cost = £60 / te CO_2	£ / MWh_{Net}	71.81	71.97	71.81	71.91
Cost of CO_2 Captured					
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	60.17	58.90	58.00	57.66
Cost of CO_2 Avoided					
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	71.57	70.57	69.38	69.06

The CAPEX and OPEX figures for the varied % carbon capture (CC) cases reflect the trends shown in the power plant performance figures, however, the costs were further from being a linear function of the %CC, in particular more total cost was incurred, per unit CO_2 captured, to higher %CC, however, less cost per unit additional CO_2 captured was seen with increasing %CC. This shows a benefit of economy of scale with increasing quantity of CO_2 captured in these cases. Consequently the cost of CO_2 captured and avoided decreases with increasing %CC for the CCGT cases.

5.2 Case 3B – NGCC, G Class Turbine without CO_2 Capture

5.2.1 Introduction

A Natural Gas CCGT without CO_2 capture was developed. This was based upon the same configuration as Case 3A, with the exclusion of the AGR and CO_2 compression and drying units. The case uses the same natural gas feed rate as the Natural Gas CCGT Power Plant with CO_2 capture case.

5.2.2 Plant Performance

The plant performance is outlined in the table below:

Figure 5-3 Performance Figures for Case 3B – NGCC without CO_2 Capture

		Case 3B
Power		
Total gross installed capacity	MWe	1037.6
Gas Turbine (s)	MWe	729.1
Steam Turbine	MWe	308.5
Others	MWe	0.0
Total auxiliary loads	MWe	46.8
ASU	MWe	0.0
Gasification / Boiler	MWe	0.0
Power Island	MWe	40.1
Acid Gas Removal	MWe	0.0
CO_2 compression	MWe	0.0
Others	MWe	6.7
Net Power Export	MWe	990.8
Net Efficiency (LHV)	%	58.8
Heat Rate	kJ/kWh	6127
Flows		
Total fuel feed rate	tpd	3070.4
Oxygen consumption	tpd	0
Water consumption	tpd	509
Cooling water (once through)	tpd	1,222,104
Carbon Balance		
Total carbon in feeds	tpd	2272.8
Total carbon captured	tpd	0.0
Carbon capture rate	%	0.0
Total CO_2 captured	tpd	0
Total CO_2 emitted	tpd	8329
CO_2 emissions	g CO_2/ kWh_{Net}	350.2

5.2.3 Process Description

5.2.3.1 Gas Turbines, Heat Recovery Steam Generators and Steam Turbine

The power island is based on two Mitsubishi Heavy Industries (MHI) M701G2 natural gas fed gas turbines, each with its own heat recovery steam generator (HRSG). The two HRSGs are identical and are configured to generate steam at three pressure levels with full reheat of medium pressure steam. A single steam turbine receives the steam from both HRSGs and is equipped with a vacuum condenser and condensate treatment.

Natural gas is received from across the plant battery limits via a metering station before being heated against MP boiler feed water (BFW) and fed to the gas turbines (GTs).

The GT exhaust gases flow to the Heat Recovery Steam Generator, without additional duct firing. The thermal energy of the exhaust gases is used to raise and superheat steam at 3 pressure levels as well as preheating condensate and heating the BFW. The flue gases, at approximately 93 °C, leaving the HRSG are released to the atmosphere via a stack equipped with damper and continuous emissions monitoring.

The coil sequence in the HRSG is summarised as follows:

- 2nd HP Superheater
- 2nd MP Reheater
- 1st HP Superheater
- 1st MP Reheater
- HP Evaporator
- MP Superheater
- 2nd HP Economiser
- MP Evaporator
- LP Superheater
- MP Economiser
- 1st HP Economiser
- LP Evaporator
- LP Economiser
- Condensate Preheater

Condensate from the steam turbine condenser is preheated and deaerated using LP steam in the deaerator. Boiler feed water from the deaerator is pumped up to the three pressure levels required by the boiler feed water pumps.

In the HP circuit the BFW is pumped to approximately 14,000 kPa, passing through the 1st and 2nd HP Economiser into the HP Steam Drum. Water from the HP Steam Drum passes through the HP Evaporator coil generating saturated HP steam which returns to the HP Steam Drum before passing through the 1st and 2nd HP Superheaters and then to the HP inlet of the Steam Turbine.

The MP BFW pumps pump BFW to approximately 3000 kpa, through the MP Economiser and into the MP Steam Drum. Water from the MP Steam Drum passes through the MP Evaporator generating MP steam which is returned to the MP Steam Drum before entering the MP Superheater. Exhaust steam from the HP stage Steam Turbine are combined with superheated MP steam which is subsequently further superheated in the 1st and 2nd MP Reheaters before being routed to the MP stage of the Steam Turbine.

Desuperheaters between the two HP superheaters and the two MP reheaters use boiler feed water to control the second superheater outlet temperatures to 565ºC for both pressure levels.

The LP BFW pumps pump the BFW to approximately 450 kpa, through the LP Economiser and into the LP Steam Drum. Water from the LP Steam Drum passes through the LP Evaporator generating LP steam which is returned to the LP Steam Drum before entering the LP Superheater. The superheated LP Steam is then split, with a portion of being used to supply the heat required for the Stripper reboiler in the AGRU with the remaining LP Steam being routed to the LP inlet of the Steam Turbine.

The exhaust gases from the LP stage of the steam turbine are combined with condensate from the Natural Gas Preheater and the condensate return from the AGRU Stripper Reboiler before being fully condensed against seawater in the Vacuum Condensate Condenser. The vacuum condensate is then returned to the Vacuum Condensate Pumps completing the circuit.

5.2.3.2 Balance of Plant

The key balance of plant requirements for this scheme are the cooling water supply systems. A very large flow of cooling water is required to supply the steam turbine vacuum condenser. This duty is supplied using sea water in a once through flow scheme. Any other cooling duties are supplied using fresh cooling water, which is itself cooled against sea water.

5.2.4 Major Equipment List

A high level equipment list was developed for this case broken down according to unit. This equipment list is located in Appendix 4 of this report.

5.2.5 Capital Cost, Operating Cost and Economics

The economic results are outlined in the table below:

Figure 5-4 Economic Figures for Case 3B – NGCC without CO_2 Capture

		Case 3B
ASU	GB£M	0.0
Gasification / Boiler	GB£M	0.0
Power Island	GB£M	423.5
Acid Gas Removal	GB£M	0.0
CO_2 compression	GB£M	0.0
Others	GB£M	120.9
Total CAPEX	GB£M	544
CAPEX efficiency	GB£/kWh	549
Total OPEX – incl. fuel	GB£M p.a.	279.6
Total OPEX – excl. fuel	GB£M p.a.	27.0
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	282.2
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	27.3
Levelised Cost of Electricity		
CO_2 emission cost = £0 / te CO_2	£ / MWh_{Net}	47.88
CO_2 emission cost = £20 / te CO_2	£ / MWh_{Net}	54.89
CO_2 emission cost = £40 / te CO_2	£ / MWh_{Net}	61.89
CO_2 emission cost = £60 / te CO_2	£ / MWh_{Net}	68.90
Cost of CO_2 Captured		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	n/a
Cost of CO_2 Avoided		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	n/a

6. OXYFUEL BASED PULVERISED COAL POWER PLANT TECHNICAL DESCRIPTIONS

6.1 Case 4 – Oxyfuel with CO_2 Capture

6.1.1 Introduction

The overall process scheme for this case is based on a pulverised coal fired ultra super critical boiler with single stage of reheat. The boiler configuration is generally similar to that described under Cases 2A and 2B, but modified for oxy-combustion and CO_2 capture.

The case is based on the same coal feed rate as in the PF Pulverised Coal Power Plant cases described above (Cases 2A and 2B).

Key features of the configuration include:

- Pulverised coal steam generator including the following features:
    - 40 mol% oxygen content in combustion medium at the burners
    - Combustion medium contains 115% of the quantity of oxygen necessary fully to combust the coal feed.
    - Air in leakage at 2 mols air per 100 mols flue gas, entering downstream the convection section
    - Full air firing capability is not provided.

- Hot Electrostatic Precipitator (ESP) located between the outlet of the convection section and the inlet to the CO_2 recycle heater (which replaces the combustion air heater of a normal boiler).

- A catalytic ammonia-base Selective Catalytic Reduction (SCR) unit installed downstream the ESP to provide bulk NO_x removal, avoiding significant accumulation of nitrogen oxides in the flue gas recycle loop.

- FGD located upstream of the flue gas recycle loop off-take, thereby avoiding recycle of SO_x to the boiler with resultant potential corrosion issues.

- A cryogenic Air Separation Unit (ASU) produces 95% purity gaseous oxygen for supply to the boiler. In addition, a liquid oxygen storage system is sized to provide 12 hours backup supply at full design rate of normal gaseous oxygen supply. There is no integration of the ASU with the CO_2 purification unit described below. Three options for configuration of the CO_2 purification step (Cases 4, 4A & 4B) are described below, with the respective performance values shown in Table 6-1

- Two nominally 50% capacity cryogenic CO_2 purification units, each comprising a molecular sieve drier, a Joule Thomson CO_2 liquefier and a final purification column, integrated with an eight-stage CO_2 compressor and liquid CO_2 export pumps.

6.1.2 Plant Performance

The plant performance is outlined in the table below:

Figure 6-1 Performance Figures for Case 4 – Oxyfuel with CO_2 Capture

		Case 4	**Case 4A**	**Case 4B**
		Base Scheme: Separate Purifier	**Integrated CO_2 Purifier**	**No CO_2 Purifier**
O_2 content in CO_2 product		**< 100 vppm**	**< 100 vppm**	**1.3 mol%**
Total gross installed capacity	MWe	855.4	855.4	855.4
Gas Turbine (s)	MWe	0.0	0.0	0.0
Steam Turbine	MWe	855.4	855.4	855.4
Others	MWe	0.0	0.0	0.0
Total auxiliary loads	MWe	249.8	234.8	230.8
ASU	MWe	91.5	91.5	91.5
Gasification / Boiler	MWe	0.0	0.0	0.0
Power Island	MWe	28.2	28.2	28.2
Acid Gas Removal	MWe	0.0	0.0	0.0
CO_2 compression	MWe	91.4	76.7	72.9
Others	MWe	38.7	38.4	38.2
Net Power Export	MWe	605.6	620.6	624.6
Net Efficiency (LHV)	%	33.3	34.1	34.3
Heat Rate	kJ/kWh	10819	10557	10490
Flows				
Total fuel feed rate	tpd	6163.4	6163.4	6163.4
Oxygen consumption	tpd	13656	13656	13656
Water consumption	tpd	3200	3200	3200
Cooling water (once through)	tpd	3,076,512	2,990,611	2,977,171
Carbon Balance				
Total carbon in feeds	tpd	3981.4	3981.4	3981.4
Total carbon captured	tpd	3597.7	3596.0	3597.2
Carbon capture rate	%	90.4	90.3	90.4
Total CO_2 captured	tpd	13184	13177	13182
Total CO_2 emitted	tpd	1406	1412	1408
CO_2 emissions	g CO_2/ kWh_{Net}	96.7	94.8	93.9

6.1.3 Process Description

6.1.3.1 Coal Receiving, Handling, Preparation and Feeding

Coal is unloaded from trains or ships and is stored on site. Vibratory feeders transfer the coal from the storage pile onto conveyors which in turn transport the coal to the preparation and feeding area.

Raw coal is crushed and fed to pulverisers. Preheated CO_2-rich recycled flue gas sweeps the coal from the pulverisers as it is being ground and then the gas / coal stream is supplied to the boiler burners.

6.1.3.2 Coal Combustion System

The coal boiler system consists of a once through, wall-fired, balanced draft ultra-supercritical steam generator adapted for use with a combustion medium containing 40 mol% oxygen.

Forced draft fans supply the recycled CO_2 -rich flue gas to the burner system via a preheater, in which the recycled gas is heated against the boiler flue gas.

The larger part of the recycled flue gas is mixed with oxygen supplied from the ASU. The combined combustion medium is used as primary and secondary oxidant in the boiler burners.

The smaller part of the reheated recycle gas is fed to the pulverisers as described above, with a small part bypassing the reheater and used for adjustment of the temperature of the coal/recycle gas stream exiting the pulverisers.

The pulverised coal / recycle gas mixture is fed to the boiler via nozzles at various elevations. The boiler will be equipped with at least one level of overfire oxidant, which introduce additional oxidant to cool the rising gases and inhibit NO_x formation.

The hot combustion products rise to the top of the boiler and pass horizontally through the secondary superheater and reheater. The gases then pass downwards over the primary superheater, economiser and into the recycle gas heater.

6.1.3.3 Steam Turbine, Condensate and Feedwater Systems

Boiler feed water (BFW) passes through an economiser coil before being fed to the lower and then upper sections of the boiler water walls. The fluid then passes through the convection circuit followed by the primary and secondary steam generation surfaces. The superheated high pressure (HP) steam is then fed to the HP section of the steam turbine.

The exhaust of the HP steam turbine section is at medium pressure (MP) and is reheated in the boiler before being fed to the MP steam turbine section. Exhaust from the MP section is fed to the inlet of the low pressure (LP) section of the turbine. A portion of the steam from the MP to LP crossover is supplied to the MEA stripper columns, from which hot condensate is returned and added to the boiler feed water.

The LP steam turbine vacuum exhaust is condensed against sea water and is then sent to the condensate polishing unit. Treated BFW is pumped from the water treatment/polishing area to the BFW heaters. The first stage of BFW heating is supplied with hot water from the MEA unit, this water is cooled sufficiently to be used as cooling water for the flashed gas cooler and several of the compressor intercooler stages. The subsequent BFW heating consists of 8 heaters with deaeration and HP pumps between stages 4 and 5. The heat is supplied by steam extracted from the steam turbines at appropriate pressure levels to supply the temperatures required. It is possible to bypass the first two steam extraction BFW heaters when all three trains of carbon capture are in operation. The BFW is then ready to be fed back to the boiler.

A small stream of extracted LP steam is desuperheated to saturation and supplied to the ASU for regeneration of the air dryers.

6.1.3.4 Ash Handling System

The boiler is equipped with MP steam soot-blowing equipment in an array of retractable nozzles and lances. The furnace bottom comprises several refractory lined hoppers with water seal and cooling which feed the ash handling system.

6.1.3.5 NOx Prevention and Removal

The low-NO_x burners and oxygen overfiring in the boiler reduce the NO_x concentration significantly. A selective catalytic reduction (SCR) unit is also provided. The SCR uses ammonia and a catalyst to reduce the NOx to N_2 and water. The ammonia is injected into the gas immediately prior to entering the reactor vessel which holds the catalyst. The SCR system consists of a reactor vessel, ammonia storage and injection and gas flow control.

6.1.3.6 Electrostatic Precipitator

The NOx reduced flue gas then passes to an electrostatic precipitator (ESP) for particulates removal. The electrodes produce an ionising corona through which the flue gas passes. The ionised particles follow the negative electric field and move downwards where they are collected and removed to the ash handling system.

6.1.3.7 Flue Gas Desulphurisation

Downstream of the ESP the flue gas flows to induced draft (ID) fans. The flue gas first enters a wet limestone forced oxidation flue gas desulphurisation (FGD) unit which removes 98% of the sulphur. As a second stage the flue gas is contacted with dilute sodium hydroxide solution which reduces its SO_x concentration to < 30 ppmv. This additional purification is necessary in order to satisfy the < 50 ppmv SO_x content of the final CO_2 product leaving the cryogenic CO_2 purifier described below. The FGD system comprises a gas/gas heat exchanger followed by an absorber module. In the gas/gas exchanger the hot flue gas is cooled against the recycle flue gas stream exiting the FGD unit. The cooled gas entering the bottom of the absorber module is typically contacted with multiple sprays of limestone slurry with trays or packing providing enhanced contact. Finally the flue gas passes through moisture separators. The larger part flows to the cryogenic CO_2 purification unit described below and the smaller part is reheated in the above-mentioned gas/gas heat exchanger.

The slurry is collected in a large inventory of liquid held in the bottom of the absorber module. Oxygen is added to this liquid inventory to promote oxidation to calcium sulphate (gypsum). The slurry is continuously agitated to ensure mixing and prevent solids settling out. Slurry is recirculated to the absorber spray section via recirculation pumps. Fresh make-up and spend reagent pumps are used to ensure the chemical balance is maintained in the absorber module. The gypsum slurry is dewatered in a slacking system.

The CO_2 rich combustion product stream from the steam generator flows to a Flue Gas Desulphurisation (FGD) Unit. This employs conventional limestone based technology for nominally 98% removal of SOx, with a secondary step of washing with sodium hydroxide solution to bring the exit SO_2 concentration to < 30 ppmv.

This low SO_2 content is necessary to achieve the specified < 50 ppmv SO_2 content in the export CO_2 stream.

The combustion product exit the FGD consists of approximately 85 mol% CO_2 with smaller amounts of oxygen, nitrogen, and argon, plus saturation water vapour.

6.1.3.8 Cryogenic CO_2 Recovery and Purification

A cryogenic CO_2 Recovery and Purification Unit is provided to produce a specification export stream of CO_2 from the flue gas exiting the FGD Unit and a vent stream containing separated O_2, N_2 and Ar.

The variants have considered, Cases 4, 4A and 4B.

Case 4 – Separate CO_2 Purification Unit

This variants consists of a cryogenic Joule-Thomson CO_2 condenser and a separate downstream CO_2 which is provided to reduce the oxygen content of the exported CO_2 stream to the specified <100 vppm.

The process scheme comprises these main steps:

- compression of the FGD exit stream to 25 bar
- drying the gas with molecular sieve
- condensation of over 90% of the CO_2 content in a cryogenic separator
- reheating and venting of the uncondensed permanent gases
- re-evaporation of the condensed CO_2
- further compression to 70 bar followed by recondensation using cooling water
- removal of dissolved oxygen from the liquid CO_2 in an absorption column
- pumping of the CO_2 product to the specified 150 bar g delivery pressure.

The incoming saturated combustion product at essentially atmospheric pressure is compressed to 12 bar in the first four stages of a multistage centrifugal CO_2 compressor. A cooler and condensate separator is provided after each compression stage. Wet gas regeneration gas from the downstream gas dryer is recycled into the discharge stream from the 4th compression stage. The combined stream is then compressed to 25 bar in the 5th compression stage.

After cooling and separation of condensate, the gas enters an adsorption dryer, which consists of two adsorbers containing molecular sieve, one in operation and the other under regeneration. Regeneration is performed by recycle of 10% of the dried gas leaving the on-line adsorber. The recycled regeneration gas is heated in two stages, first in a feed/product heat exchanger and then by an electric heater. The wet regeneration gas stream is recycled to a point upstream the 5th compression stage as described above

The main stream of dried process gas is then combined with recycled overhead gas from the downstream absorption column (see below) and enters the cryogenic CO_2 condenser, in which approximately 90% of the CO_2 in the unit feed is condensed in two stages in the plate-fin heat exchangers.

The gas phase leaving from the cold end of the second plate fin exchanger is reheated to near-ambient temperature and sent to the plant battery limits at 23 bar as a reject vent gas stream. There is potential for future efficiency improvements, for example by heating this stream by a convenient means followed by a power recovery expander.

The condensed CO_2 streams leaving the plate-fin exchangers are re-evaporated and then recompressed in the final three stages of the CO_2 compressor. The compressed CO_2 is then condensed by heat exchanger with cooling water.

From the viewpoint of CO_2 purity it would only be necessary to flash this stream to a reduced pressure to meet the specified 97% mol CO_2 content. However its oxygen content (approx. 1.3 mol %) necessitates a further purification stage to meet the specified < 100 ppmv oxygen content.

This is achieved by means of a reboiled stripper column. The CO_2 stream is cooled to 13^0C in the column reboiler and is then flashed through a valve to 25 bar. The resulting two-phase mixture enters the top of the absorption column. The reboiler generates CO_2 vapour which strips oxygen and other gases from the liquid CO_2 descending the column. The liquid at the base of the column contains 99.9+ % CO_2 and < 100 ppmv O_2.

The liquid is pumped to 150 barg and is exported for sequestration.

The column overhead vapour is recycled to the inlet of the first plate-fin heat exchanger, as described above.

Further development may result in a reduction on the number of compression stages from 8 to 6 or 7.

Case 4A – Integrated CO_2 Purifier

This second variant incorporates the reboiled stripper column into the Joule-Thomson CO_2 condenser, instead of its locating as a separate downstream assembly in Case 4. This change has reduced the volume of the stripper overhead vapour reducing the total compression power required.

The stream exiting the fifth compressor stage enters the first plate-fin heat exchanger at approximately 25 bar. This stream leaves the exchanger at – 30^0C and flows to the reboiler of the stripper described below. It finally flows to the second plate-fin heat exchanger, exiting with approximately 90% of the inlet CO_2 condensed and enters a vapour/liquid separator. The vapour phase comprises the vent stream, which is reheated to ambient temperature as described above for Case 4.

The liquid phase from the separator is reheated and partially evaporated in the second plate-fin heat exchanger and flows then to the top of the stripper column, which operates at 13 bar. The overhead vapour from the stripper is reheated to ambient temperature in the first plate-fin exchanger and flows to the suction of the fifth compressor stage. A smaller part of this stream is separated from the total flow and is used for regeneration of the dryers.

The liquid from the bottom of the stripper, containing the specified < 100 vppm oxygen, is reheated and re-evaporated in the first plate-fin heat exchanger. It is compressed to 64 bar in the final three stages of the CO_2 compressor. Finally it is condensed with cooling water and pumped for export at 150 barg.

Case 4B – Without Purifier

This third case represents the widely published arrangement comprising a second Joule-Thomson CO_2 condenser without a final oxygen-removal stripper column. The configuration is identical to that of Case 4 except that the separate stripper sub-assembly is deleted.

The oxygen content of the exported CO_2 at 1.3 mol% is far above the 100 vppm value specified in the Basis of Design for the study.

This Case 4B is included only for comparison of the total power requirement and efficiency with Cases 4 and 4A.

In conclusion it appears that Option 4A is particularly attractive as it incurs only 0.2 % points penalty in overall plant efficiency while conforming to the specified 100 vppm oxygen content of the export CO_2, relative to Case 4B with unconstrained oxygen content.

Case 4 with a separate stripper unit incurs a larger efficiency penalty of 1% point in overall plant efficiency. It is recommended that Case 4A should be considered for a further stage of development.

6.1.4 Plant Turndown

The steam cycle based CCS cases (2A as well as 4) are potentially capable of turn down to outputs below 50%, sharing the characteristics of conventional steam plant. In this respect the oxy-fuel Case 4 is not thought to be significantly less flexible than the amine-wash Case 2A, although the turn-down capacity of the CO_2 compressors will typically be limited to 70% without recourse to suction recycle.

6.1.5 Major Equipment List

A high level equipment list was developed for this case broken down according to unit. This equipment list is located in Appendix 4 of this report.

6.1.6 Capital Cost, Operating Cost and Economics

The economic results are outlined in the table below:

Figure 6-2 Economic Figures for Case 4 – Oxyfuel with CO_2 Capture

		Case 4
ASU	GB£M	338.4
Gasification / Boiler	GB£M	779.4
Power Island	GB£M	234.4
Acid Gas Removal	GB£M	0.0
CO_2 compression	GB£M	217.8
Others	GB£M	326.5
Total CAPEX	GB£M	1897
CAPEX efficiency	GB£/kWh	3132
Total OPEX – incl. fuel	GB£M p.a.	205.9

Total OPEX – excl. fuel	GB£M p.a.	81.5
OPEX – incl. fuel	GB£ p.a. / kW_{Net}	340.0
OPEX – excl. fuel	GB£ p.a. / kW_{Net}	134.6
Levelised Cost of Electricity		
CO_2 emission cost = £0 / te CO_2	£ / MWh_{Net}	96.90
CO_2 emission cost = £20 / te CO_2	£ / MWh_{Net}	98.83
CO_2 emission cost = £40 / te CO_2	£ / MWh_{Net}	100.77
CO_2 emission cost = £60 / te CO_2	£ / MWh_{Net}	102.70
Cost of CO_2 Captured		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	44.82
Cost of CO_2 Avoided		
CO_2 emission cost = £ 0 / te CO_2	£ / te CO_2	59.48

7. EVALUATION BASES

7.1 Evaluation Basis

7.1.1 Technical Basis

To support the study objective of providing consistent benchmarks the following Design Basis features were agreed and have been applied across each of the benchmark cases.

- **Plant Location:** The site is assumed to be a green field coastal location on the NE coast of the UK, with adjacent deep sea access, thus limiting the length of the sea water lines (both the submarine line and the sea water pumps discharge line). The site is also assumed to be close to an existing harbour equipped with a suitable pier and coal bay to allow coal transport by large ships and associated ease of coal handling.

- **Site condition:** An assumed clear level obstruction free site (both under and above ground) free site, without the need for any special civil works

- **Plant capacity:** Each Benchmark case has been configured to produce electric energy (800 MWe nominal gross capacity without CO_2 capture) to be delivered to the UK National grid. For each of the Benchmarks considered, the design capacity will vary, determined by the full design capacity of key equipment items, for example, in the case of IGCC and CCGT scheme the full ”appetite” of the selected gas turbines.

- **Plant operating conditions:** The following climatic conditions marked (*) have been considered the reference conditions for plant performance evaluation.

    - Atmospheric pressure: 1013 mbar (*),

    - Relative humidity: average 60% (*), max: 95%, min: 40%

    - Ambient temperature: average 10°C (*), max 30°C, min -10°C

- **Natural Gas feedstock specification**: Natural gas National Transmission System (NTS) connection is available. National Grid electrical grid connection is available for “black start” scenarios. Natural gas feedstock specification (as NTS spec):

Figure 7-1 Natural Gas Specification

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

- **Coal feedstock specification:**
  - Proximate analysis (wt%): Inherent moisture: 9.50, Ash: 12.20, Coal (dry, ash free): 78.30
  - Ultimate analysis (wt%) (dry ash free): Carbon 82.50, Hydrogen 5.60, Nitrogen 1.77, Oxygen 9.00, Sulphur 1.10, Chlorine 0.03
  - Gross CV 27.06 MJ/kg, Net CV 25.87 MJ/kg
  - Hardgrove Index 45
  - Ash fusion point 1350 ^{o}C (reducing temperature)
  - The coal specification is based on an open-cut coal from Eastern Australia.

- **CO_2 product specification:** Carbon dioxide produced from the plant will be dried and compressed to 150 bar(g) for export from the facility. Product carbon dioxide conditions:
    - Pressure: 150 bar(g), Temperature: ≤ 30°C. The carbon dioxide is assumed to be used for Enhanced Oil Recovery (EOR), and therefore the target CO_2 export specification is: $H_2O < 50$ ppmv, $CO_2 > 97$ vol%, $SO_2 < 50$ ppm, $H_2S < 50$ ppm, $CO < 3$ vol%, $Ar < 3$ vol%, O_2 100 ppmv, $N_2 < 3$ vol%, $H_2 < 3$ vol%, $CH_4 < 2$ vol%, $COS < 50$ppm
- **Power:** Power will be generated from the complex at 275 kV and will be transmitted to an assumed existing HV substation for connection onto the UK National Grid. It is assumed that National Grid electrical grid connection is available. Electric Power.
    - Net Power Output 800 MWe nominal capacity. Voltage 275kV, frequency 50 Hz
- **CO_2 capture rate**: Each CO_2 abated case is designed to achieve a target carbon capture level of at least 90%, defined as:

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

7.1.2 Capital Cost Estimating Basis

7.1.2.1 Introduction

The Estimates contained within this study report have been based on the technical definition as for each of the benchmark cases considered. The estimate methodology is largely based on in-house data, available from previous work undertaken by Foster Wheeler for similar plants.

For all of the cases reported the source estimate data has been adjusted to provide figures on a consistent and comparable 1st quarter 2009 (1Q2009) UK Basis.

Estimates prepared using this methodology and associated qualifications/exclusions are normally considered to have an accuracy of +/-40%.

7.1.2.2 Currency

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

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

Base Currency	Exchange Rate
GB£ 1	US$ 1.53
GB£ 1	€ 1.12

7.1.2.3 Basis

These are Factored Estimates and are based on previous estimates prepared using Aspentech Kbase and Foster Wheeler in-house data for more complex specialist equipment, the description below is consistent with the production of these more detailed estimates.

All other costs associated with the project are factored from the equipment costs or other available information to give a Total Project Cost.

No Site specific costs have been included. Consistent with the study Design Basis the site has been assumed to be a generic site clear and level and free from underground obstructions. These Estimates reflect a 1Q2009 UK Site Basis with no allowance for future escalation.

7.1.2.4 Format

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

- Coal Handling, Storage, Milling and Gasification
- Air Separation / Oxidant Supply
- Acid Gas Removal
- Sulphur Plant
- Syngas Treatment Unit
- CO_2 compression and dehydration
- Power Block
- Common Facilities

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

- Interconnecting piping
- Electrical Switchgear/Transformers
- 275 kV cables to new switchyard
- DCS system
- Seawater Intake/Pumping/Outake System
- Demineralised Water system
- Natural gas system
- BFW Chemical Injection
- Condensate Polishing Package
- Chemicals
- Water treatment
- Flare Package
- N_2 Generation Package

- Instrument/Utility Air Package
- Fire fighting system

7.1.2.5 Direct Materials

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

7.1.2.6 Major Equipment

The majority of equipment item costs have been generated using the Aspentech Kbase estimating program indexed to reflect Foster Wheeler's experience of market conditions.

For some specialised major equipment items not covered by the Kbase database, costs have been based on in-house data and some budget prices from suppliers or licensors.

The main supplier/licensor budget prices, received for previous works made by Foster Wheeler, mainly include the following unit/equipment:

- Gasification unit of the different technologies;
- Air Separation unit;
- Acid Gas Removal unit;
- Dehydration Package;
- Plate heat exchangers;
- Power Island.

7.1.2.7 Bulk Materials

The bulk material costs have been factored from the major equipment costs using factors derived from a more detailed study for a very similar plant.

7.1.2.8 Catalysts & Chemicals

For the IGCC alternatives, the costs for the first fill Catalyst & Chemicals have been based on estimated volumes and in-house unit volume price data.

7.1.2.9 Spare Parts

Commissioning and 2 yrs Operational Spares costs have been estimated using historical percentage factors.

7.1.2.10 Shipping, Freight & Import Duties

Shipping & Freight costs have been estimated using historical percentage factors. Import duties have been excluded.

7.1.2.11 Other costs

Other costs, including material and Labour Sub Contractors, temporary facilities, construction management etc, have been derived using standard historical percentages.

7.1.2.12 Escalation

The estimates have been escalated depending on the date of the reference project, based on Foster Wheeler experience. No allowance has been made for future escalation.

7.1.2.13 Contingency

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

7.1.2.14 Owner's Costs

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

7.1.2.15 Land Costs

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

7.1.2.16 Exclusions

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

- Import Duties;
- Capital / Insurance Spares;
- Financing;
- Royalties & Process Guarantees;
- Piling;
- Removal of unseen/unidentified underground obstructions;
- Operating costs;
- Statutory Authority & Utility Company Costs & permits;
- Currency Fluctuations;
- PMC Costs;
- Contractors Fees;
- Contractors All Risk Insurance;
- Taxes;
- Metal pricing movements.

7.1.3 Operating and Maintenance Cost Estimating Basis

7.1.3.1 Introduction

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

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

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

7.1.3.2 Variable costs

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

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

7.1.3.3 Fixed costs

The fixed costs mainly include the following:

- Direct labour
- Administrative and general overheads
- Maintenance

7.1.3.4 Direct Labour

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

Table 1 – Personnel basis for USCPC plants without CO_2 capture

Operation	Boiler Island	CCU & Utilities	Total	Notes
Area Responsible	1	1	2	daily position
Assistant Area Responsible	1	1	2	daily position
Shift Superintendent	5		5	1 shift position
Electrical Assistant	5		5	1 shift position
Shift Supervisor	5	5	10	2 shift position
Control Room Operator	10	10	20	4 shift position
Field Operator	15	25	40	8 shift position
Subtotal			**84**	
Maintenance				
Mechanical group	6		6	daily position
Instrument group	6		6	daily position
Electrical group	5		5	daily position
Subtotal			**17**	
Laboratory				
Superintendent + Analysts	4		**4**	daily position
Total			**105**	

For USCPC plants with CO_2 capture, two additional individuals for each shift are included (one field operator plus one control room operator), leading to a total of 115 individuals.

Table 2 – Personnel basis for IGCC plants with CO_2 capture

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

The number of personnel required for the IGCC plant without pre-combustion CO_2 capture has been considered as 115. For the Oxyfuel plant same number as per IGCC with carbon capture has been considered.

Table 3 – Personnel of Combined Cycle Gas Turbine plants

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

The number of personnel required for the Combined Cycle Gas Turbine plants with post-combustion CO_2 capture has been considered as 60.

7.1.3.5 Administrative and General Overheads

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

- Management
- Personnel services
- Technical services
- Clerical staff

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

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

7.1.3.6 Maintenance

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

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

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

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

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

ATTACHMENT 1 BLOCK FLOW DIAGRAMS

1. **CASE 1A – IGCC WITH CO_2 CAPTURE**
2. **CASE 1B – IGCC WITHOUT CO_2 CAPTURE**
3. **CASE 2A – ULTRA SUPERCRITICAL PLANT WITH CO_2 CAPTURE**
4. **CASE 2B – ULTRA SUPERCRITICAL PLANT WITHOUT CO_2 CAPTURE**
5. **CASE 3A – NGCC, G CLASS TURBINE WITH CO_2 CAPTURE**
6. **CASE 3B – NGCC, G CLASS TURBINE WITHOUT CO_2 CAPTURE**
7. **CASE 4 – OXYFUEL WITH CO_2 CAPTURE**

ATTACHMENT 2 H&MB DIAGRAMS

1. CASE 1A – IGCC WITH CO_2 CAPTURE

1.1 85% CAPTURE (VERSION 1)

1.2 90% CAPTURE

1.3 95% CAPTURE

2. CASE 1B – IGCC WITHOUT CO_2 CAPTURE

3. CASE 2A – ULTRA SUPERCRITICAL PLANT WITH CO_2 CAPTURE

3.1 85% CAPTURE

3.2 90% CAPTURE

3.3 95% CAPTURE

4. CASE 2B – ULTRA SUPERCRITICAL PLANT WITHOUT CO_2 CAPTURE

5. CASE 3A – NGCC, G CLASS TURBINE WITH CO_2 CAPTURE

5.1 85% CAPTURE

5.2 90% CAPTURE

5.3 95% CAPTURE

6. CASE 3B – NGCC, G CLASS TURBINE WITHOUT CO_2 CAPTURE

7. CASE 4 – OXYFUEL WITH 90% CO_2 CAPTURE

ATTACHMENT 3 UTILITY SUMMARIES

1. CASE 1A – IGCC WITH CO$_2$ CAPTURE

1.1 85% CAPTURE (Version 1)

1.2 85% CAPTURE (Version 2)

1.3 90% CAPTURE

1.4 95% CAPTURE

2. CASE 1B – IGCC WITHOUT CO$_2$ CAPTURE

3. CASE 2A – ULTRA SUPERCRITICAL PLANT WITH CO$_2$ CAPTURE

3.1 85% CAPTURE

3.2 90% CAPTURE

3.3 95% CAPTURE

4. CASE 2B – ULTRA SUPERCRITICAL PLANT WITHOUT CO$_2$ CAPTURE

5. CASE 3A – NGCC, G CLASS TURBINE WITH CO$_2$ CAPTURE

5.1 85% CAPTURE

5.2 90% CAPTURE

5.3 95% CAPTURE

6. CASE 3B – NGCC, G CLASS TURBINE WITHOUT CO$_2$ CAPTURE

7. CASE 4 – OXYFUEL WITH 90% CO$_2$ CAPTURE

ATTACHMENT 4 EQUIPMENT LISTS

1. CASE 1A – IGCC WITH CO_2 CAPTURE

1.1 85% CAPTURE (Version 1)

1.2 85% CAPTURE (Version 2)

1.3 90% CAPTURE

1.4 95% CAPTURE

2. CASE 1B – IGCC WITHOUT CO_2 CAPTURE

3. CASE 2A – ULTRA SUPERCRITICAL PLANT WITH CO_2 CAPTURE

3.1 85% CAPTURE

3.2 90% CAPTURE

3.3 95% CAPTURE

4. CASE 2B – ULTRA SUPERCRITICAL PLANT WITHOUT CO_2 CAPTURE

5. CASE 3A – NGCC, G CLASS TURBINE WITH CO_2 CAPTURE

5.1 75% CAPTURE

5.2 85% CAPTURE

5.3 90% CAPTURE

5.4 95% CAPTURE

6. CASE 3B – NGCC, G CLASS TURBINE WITHOUT CO_2 CAPTURE

7. CASE 4 – OXYFUEL WITH 90% CO_2 CAPTURE

ATTACHMENT 5 CAPITAL COST ESTIMATES

1. CASE 1A – IGCC WITH CO_2 CAPTURE

2. CASE 1B – IGCC WITHOUT CO_2 CAPTURE

3. CASE 2A – ULTRA SUPERCRITICAL PLANT WITH CO_2 CAPTURE

4. CASE 2B – ULTRA SUPERCRITICAL PLANT WITHOUT CO_2 CAPTURE

5. CASE 3A – NGCC, G CLASS TURBINE WITH CO_2 CAPTURE

6. CASE 3B – NGCC, G CLASS TURBINE WITHOUT CO_2 CAPTURE

7. CASE 4 – OXYFUEL WITH CO_2 CAPTURE

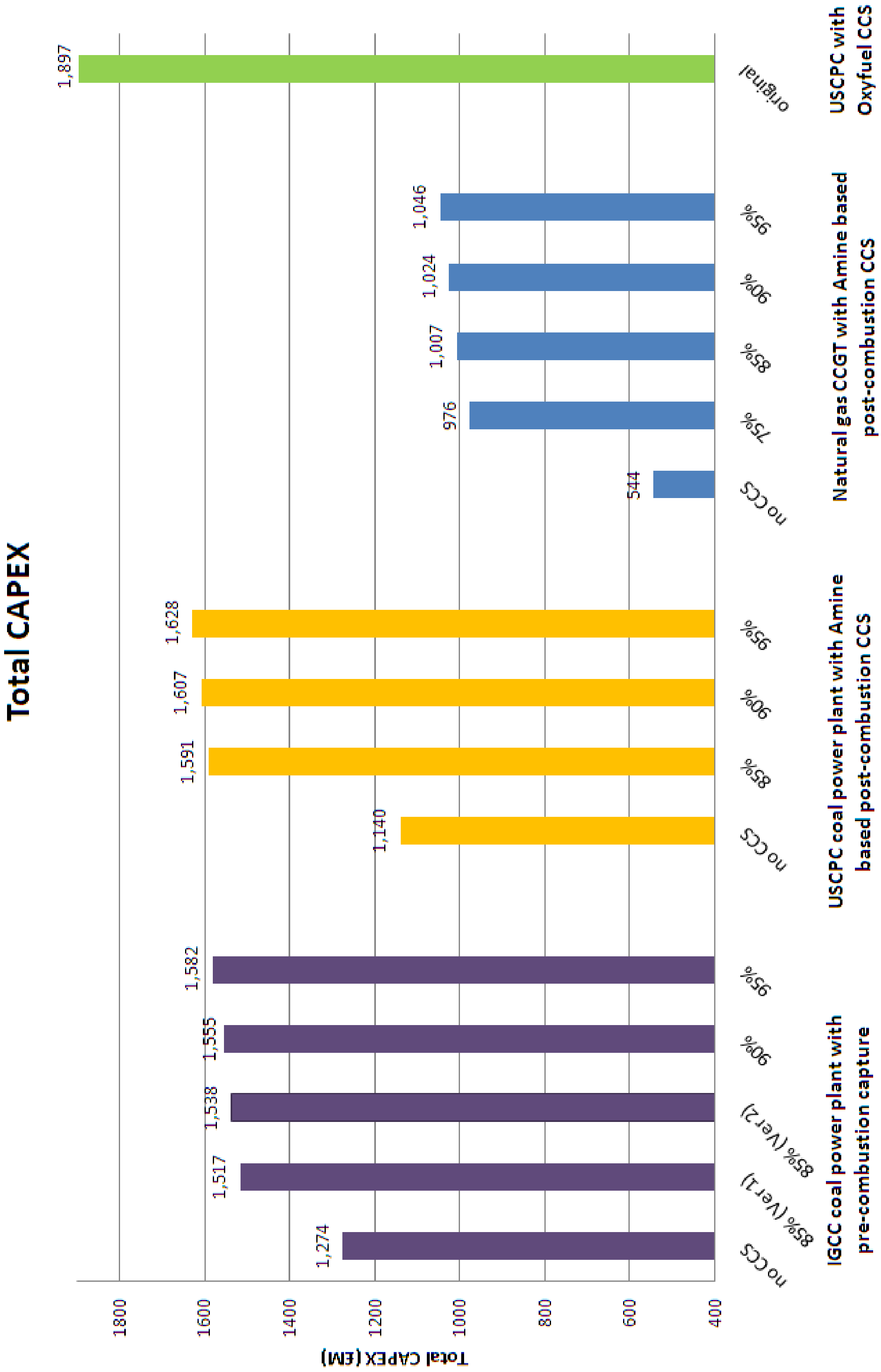
ATTACHMENT 6 OPERATING COST ESTIMATES

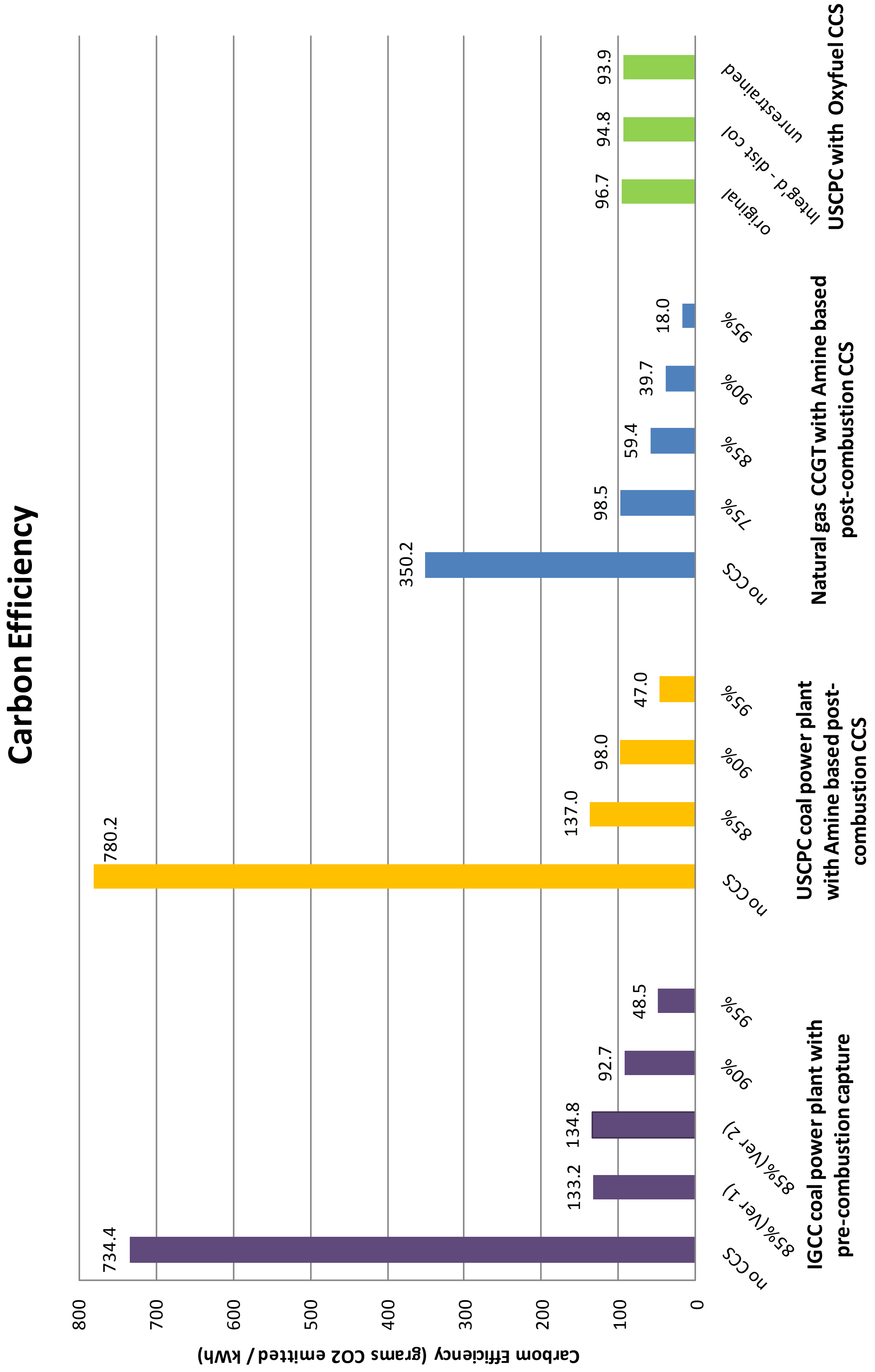
TABLE 6.1 OPERATION AND MAINTENANCE COSTS

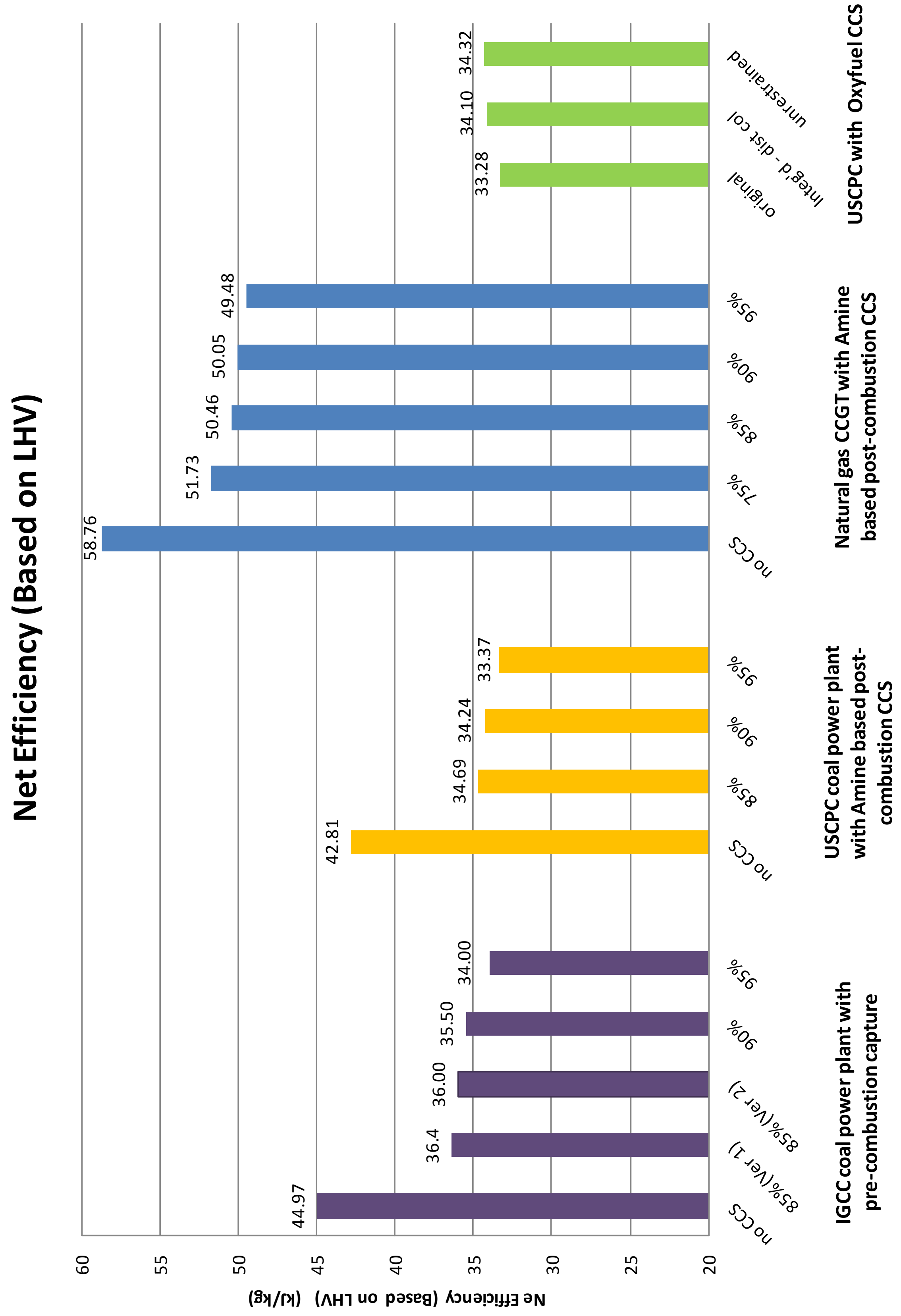
TABLE 6.2 TOTAL OPERATIONS AND MAINTENANCE COSTS

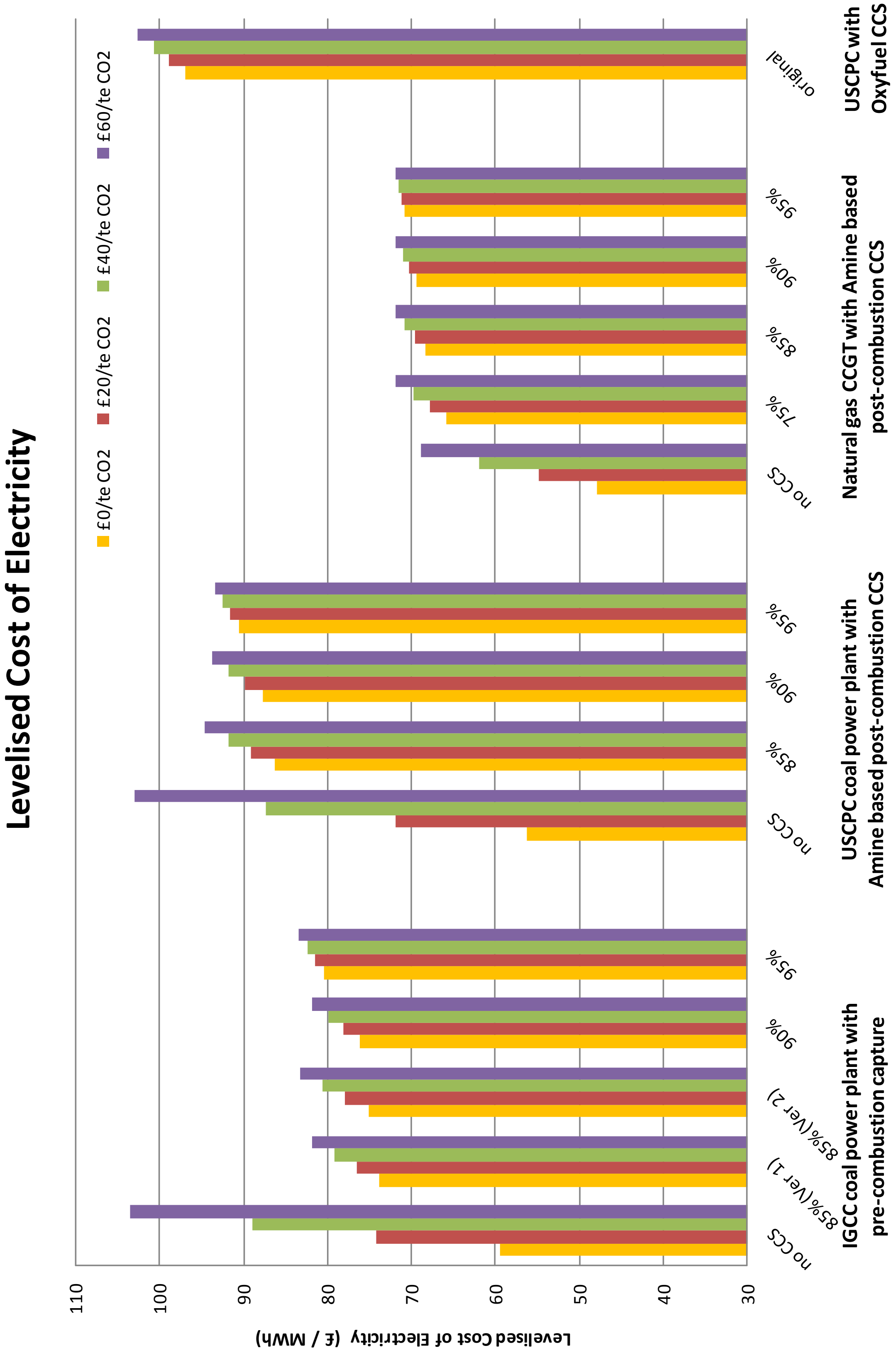
ATTACHMENT 7 BASIS OF DESIGN

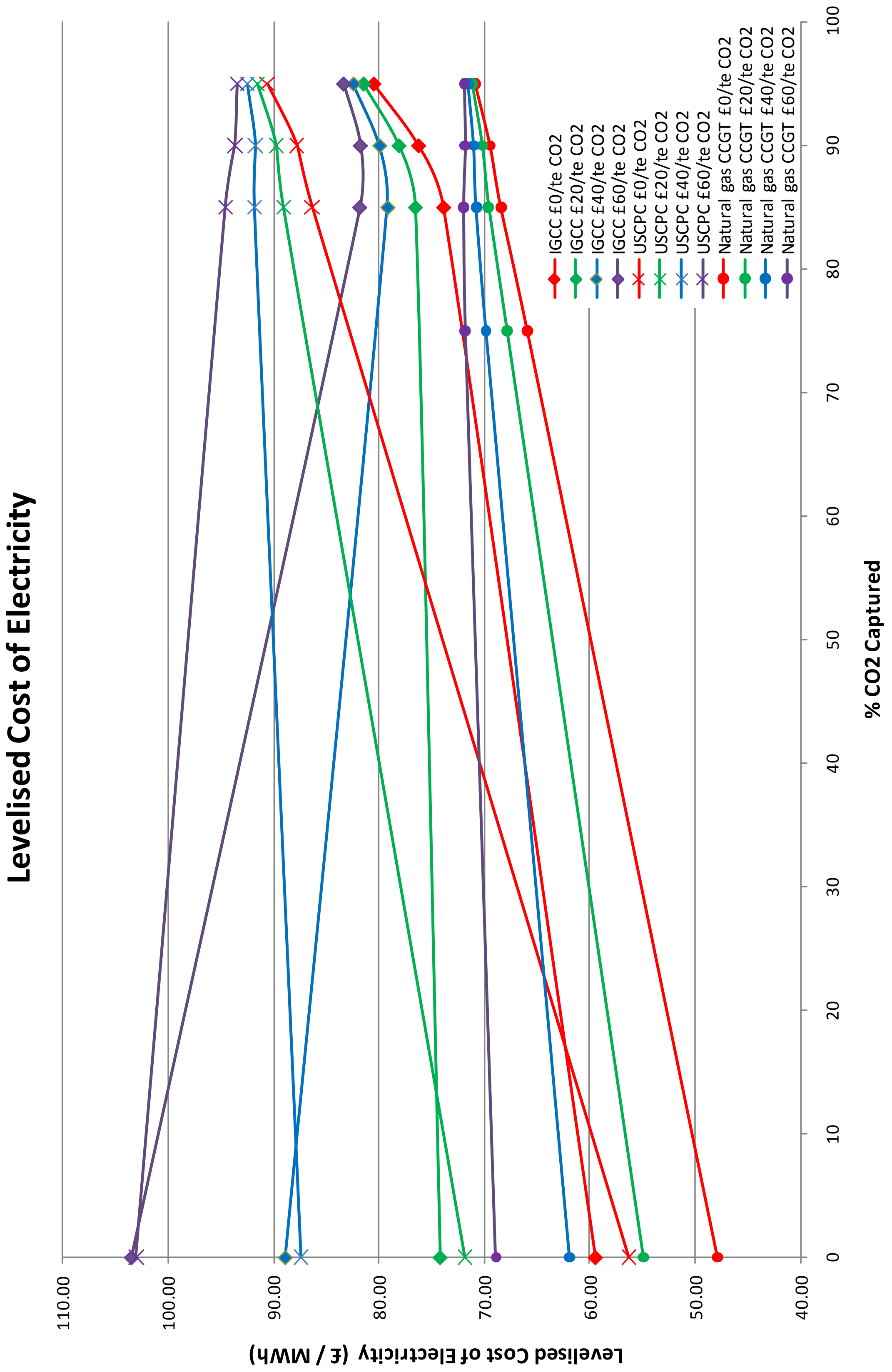
ATTACHMENT 8 GRAPHS SHOWING TRENDS

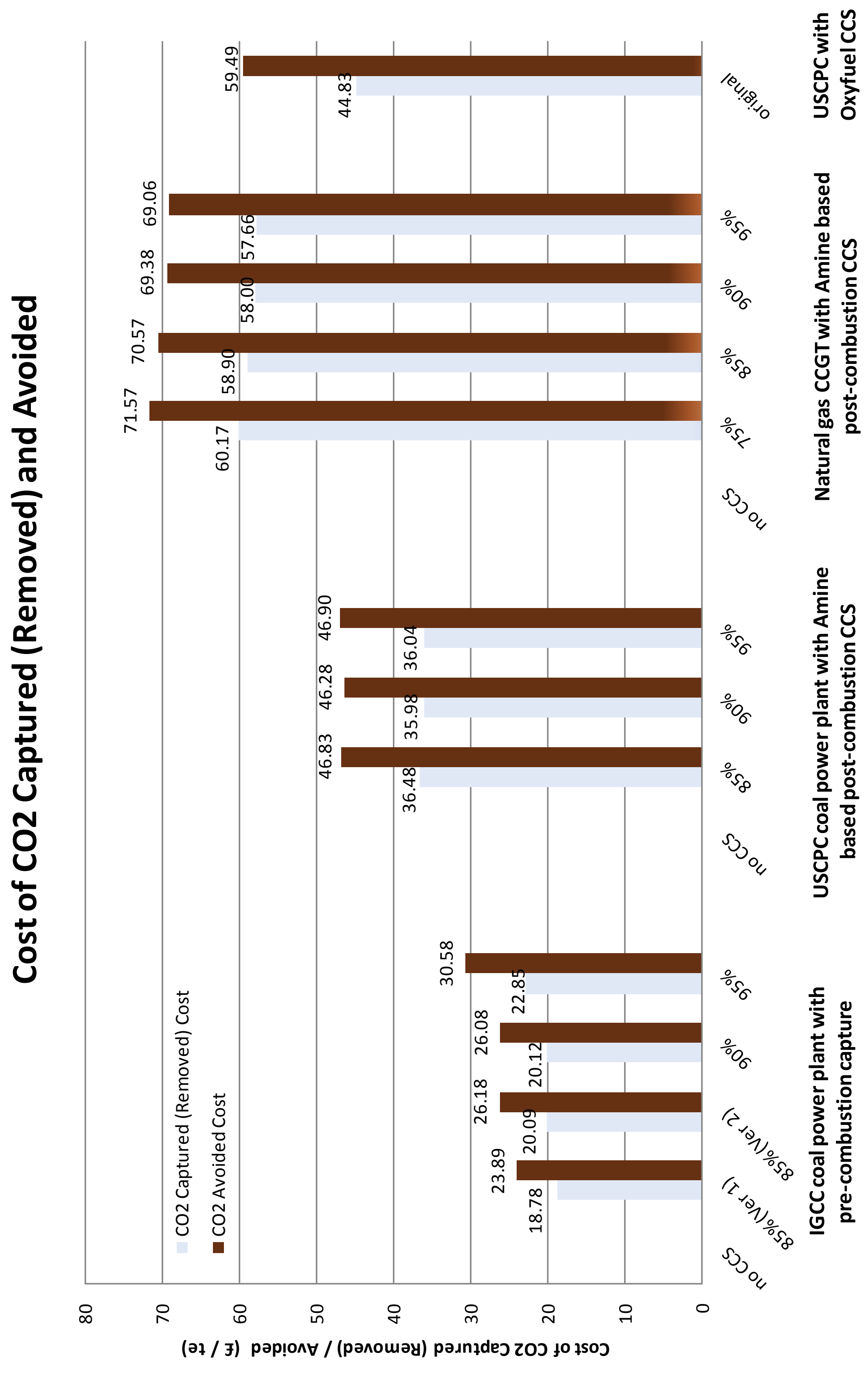



Carbon Efficiency
Carbom Efficiency (grams CO2 emitted / kWh)
0
100
200
300
400
500
600
700
800
734.4
133.2
134.8
92.7
48.5
no CCS
85%(Ver 1)
85%(Ver 2)
90%
95%
IGCC coal power plant with pre-combustion capture
780.2
137.0
98.0
47.0
no CCS
85%
90%
95%
USCPC coal power plant with Amine based post-combustion CCS
350.2
98.5
59.4
39.7
18.0
no CCS
75%
85%
90%
95%
Natural gas CCGT with Amine based post-combustion CCS
96.7
94.8
93.9
original
Integ'd - dist col
unrestrained
USCPC with Oxyfuel CCS


Net Efficiency (Based on LHV)
Ne Efficiency (Based on LHV) (kJ/kg)
60
55
50
45
40
35
30
25
20
44.97
36.4
36.00
35.50
34.00
no CCS
85%(Ver 1)
85%(Ver 2)
90%
95%
IGCC coal power plant with pre-combustion capture
42.81
34.69
34.24
33.37
no CCS
85%
90%
95%
USCPC coal power plant with Amine based post-combustion CCS
58.76
51.73
50.46
50.05
49.48
no CCS
75%
85%
90%
95%
Natural gas CCGT with Amine based post-combustion CCS
33.28
34.10
34.32
original
Integ'd - dist col
unrestrained
USCPC with Oxyfuel CCS


Levelised Cost of Electricity
£0/te CO2
£20/te CO2
£40/te CO2
£60/te CO2
Levelised Cost of Electricity (£ / MWh)
110
100
90
80
70
60
50
40
30
no CCS
85%(Ver 1)
85%(Ver 2)
90%
95%
IGCC coal power plant with pre-combustion capture
no CCS
85%
90%
95%
USCPC coal power plant with Amine based post-combustion CCS
no CCS
75%
85%
90%
95%
Natural gas CCGT with Amine based post-combustion CCS
original
USCPC with Oxyfuel CCS


Levelised Cost of Electricity
Levelised Cost of Electricity (£ / MWh)
110.00
100.00
90.00
80.00
70.00
60.00
50.00
40.00
% CO2 Captured
0
10
20
30
40
50
60
70
80
90
100
IGCC £0/te CO2
IGCC £20/te CO2
IGCC £40/te CO2
IGCC £60/te CO2
USCPC £0/te CO2
USCPC £20/te CO2
USCPC £40/te CO2
USCPC £60/te CO2
Natural gas CCGT £0/te CO2
Natural gas CCGT £20/te CO2
Natural gas CCGT £40/te CO2
Natural gas CCGT £60/te CO2


Cost of CO2 Captured (Removed) and Avoided
CO2 Captured (Removed) Cost
CO2 Avoided Cost
Cost of CO2 Captured (Removed) / Avoided (£ / te)
0
10
20
30
40
50
60
70
80
IGCC coal power plant with pre-combustion capture
no CCS
85%(Ver 1)
18.78
23.89
85%(Ver 2)
20.09
26.18
90%
20.12
26.08
95%
22.85
30.58
USCPC coal power plant with Amine based post-combustion CCS
no CCS
85%
36.48
46.83
90%
35.98
46.28
95%
36.04
46.90
Natural gas CCGT with Amine based post-combustion CCS
no CCS
75%
60.17
71.57
85%
58.90
70.57
90%
58.00
69.38
95%
57.66
69.06
USCPC with Oxyfuel CCS
original
44.83
59.49