Oxy Combustion Processes for CO_2 Capture from Power Plant

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Background to the study.

The IEAGHG R&D programme has completed several studies on the costs of CO_2 capture from power plants using post combustion and pre-combustion capture technology. The costs of oxy-combustion capture have not been studied to the same depth because of the immaturity of the technology. Although commercial examples of the technology are still not in existence it is now felt that there has been sufficient advance in knowledge to attempt a cost study of similar accuracy. The process is applicable to both natural gas and coal fired power plant although the equipment used for the different fuels is quite different. There are many oxy-combustion process variants some of which are still in early stages of development. For example processes are proposed which use dense oxygen conducting membranes or recycle of water rather than CO_2. It would be difficult to generate firm costs for process using such novel elements. The brief given to the contractor was to select a gas fired and a coal fired process which would make use of existing designs and not represent any significant stepout beyond accepted limits. For the coal fired case this essentially means maintaining sufficient CO_2 recirculation so as not to radically alter velocities and heat fluxes in the radiant and convective parts of a conventional pulverized coal boiler. For the gas fired case, which would be a combined cycle gas turbine with CO_2 recycle, the situation is more difficult since a new gas turbine model would have to be developed. The brief was to base this machine on existing design limitations and practices so that the changes would be minimized. The commercial costing of this machine remains difficult since the price will be highly dependent on the size of the market.

Approach adopted

The contract was awarded to a consortium consisting of Mitsui Babcock, Alstom and Air Products who collectively cover all the specialized equipment required for designing and costing state of the art oxy-combustion power plant. The team were asked to determine the optimum oxygen supply specifications and to design and cost cryogenic oxygen plant to match the requirement. The main considerations for the plant are minimizing parasitic power consumption, providing adequate but not excessive continuity of operation, and balancing oxygen and CO_2 purification costs. The contractors were asked to review the options and select the most proven technology for a coal fired and a gas fired plant.

The main brief for Mitsui Babcock was to select a process which would allow use of conventional coal milling and boiler components and not move into an operating regime outside currently accepted practices.

The main brief for Alstom was to evaluate a combined cycle gas turbine power plant which made maximum use of existing proven design principles and fully respected proven design parameters for gas turbine design.

The brief for Air Products was to specify and cost fit for purpose oxygen plants for the two power plants with appropriate consideration given to the trade offs between oxygen purity, cost, and parasitic power consumption and clean up processing for the captured CO_2.

Results and discussion

Process selection

Coal fired process

It was concluded that a pulverized fuel power station converted to oxy-fuel firing was the best proven technology basis. The amount of CO_2 recycled in such a process has a considerable effect on conditions in the plant. Most important are the effect on flame temperature and radiation intensity on the radiant section and the mass flow and hence convective heat transfer in the convection section. The further the recycle is reduced the more the design deviates from conventional practice. More heat has to be absorbed in the radiant section as flame temperature increases and the convection section would have to have a slimmer longer flow path to maintain suitable velocities. The contractor's judgment was that a recycle which had the net effect of feeding about 30% oxygen in CO_2, (compared to the 21% oxygen in air fed to a conventional boiler) would still allow essentially the same boiler designs to be used. There is considerable scope for reducing the recirculation further but this would require development and testing of radically altered boiler designs. The comparison was to be made on the basis of a net output of 500MW and this is achieved in a single train facility

Gas fired process

A process in which captured CO_2 is recycled was chosen. Other alternatives in which flue-gas, steam or water are recycled are also possible. However the recycle of CO_2 alone was considered to be the most conservative approach. The effects of higher steam partial pressures which other options involve were considered to be slightly less certain. The base case was taken to be a gas turbine combined cycle process since this has the best efficiency. When converted to oxy-combustion the gas turbine at the heart of the process would have to be a completely new machine because of the altered properties of the working fluid. As base case a single train gas turbine and steam turbine system fired on natural gas was chosen. The machine for the equivalent CO_2 capturing process was chosen to have a similar frame size but because of changes in the properties of the working fluid would result in a smaller output from a single train system. Thus a system with two gas turbines feeding one common steam turbine was chosen for analysis. This gave the closest approach in terms of net power output to that of the base case.

The key, fundamental design parameters for a gas turbine were considered by Alstom to be the stress in the metal of the turbine blades in relation to their creep resistance and the stage loading of each rotor in the compressor. A further design constraint was that the last stage of the expansion turbine would not require cooling and turbine inlet temperatures would have to be limited to respect this limitation. Recent advances such as steam cooling and reheat were not incorporated in the design which took as its starting point the General Electric 9FA machine.

Air separation plant

The key characteristics of the air separation units when compared with typical plants in operation are the very large capacity requirement, the need to minimize power consumption, the relaxed optimum O_2 purity, no requirement for co-production of nitrogen, and the requirement only for a gaseous oxygen stream which for the PF plant is also at low pressure. The gas fired plant requires gaseous oxygen at high pressure because it is mixed with recirculating CO_2 just before the gas burners after the CO_2 has been compressed. The cycle chosen for this duty is one in which gaseous oxygen (GOX) is produced by pumping liquid oxygen (LOX) to the required 40 barg pressure and allowing this stream to be boiled against condensing high pressure air. This avoids a separate oxygen compressor. An alternative would be to mix low pressure gaseous oxygen into the inlet of the gas turbine compressors which would have to be larger to handle the extra flow. The main reason for choosing to mix at high rather than low pressure was that it was considered easier and more efficient to integrate this into the oxygen plant rather than increase the power of the gas turbine compressors.

Accordingly Air Products proposed a three column design which enabled the capacity to be satisfied with two trains and also resulted in slightly reduced power consumption. Air is compressed to two pressures and the addition of an intermediate pressure third column enables a greater proportion of the compression to be to the lower pressure giving a small reduction is specific power.

The oxygen plant is stand alone and there is no integration of the air compressor with other turbo machinery. This choice was made because it avoids costly modifications to any standard machinery. The only integration is heat recovery from the compressed air to the power plant feedwater heating system.

CO_2 clean up.

A simple cryogenic purification of the captured CO_2 was chosen. Depending on the purity a two or three stage flash process or one with a distillation stage were specified. The effects on cost are minimal but designs for higher purity do carry a small parasitic power penalty. The main effect of producing purer CO_2 is a reduction in the percentage of the CO_2 which is captured as more has to be vented with uncondensables. If purity is increased from 95% to 98% there is a reduction of 2.3% in the amount of CO_2 captured but little effect on the cost of electricity. The clean up process is relatively simple and cheap. Producing high purity oxygen is significantly more expensive and air ingress and stoichiometric requirements for complete combustion contaminate the CO_2 with nitrogen and oxygen anyway. Hence optimum oxygen purity was considered to be just 95%. An optional addition is extraction of SO_2 and NO_x as separate streams which can in principle be used as feedstock to other industrial processes. If these components are not separated the issue of storage of these components along with CO_2 would have to be addressed. Their presence could make it more difficult to acquire the necessary permits. SO_2 has some significant specialized industrial uses or it can be converted to sulphuric acid for which global production is well over 100 million tpa. Thus there should be little difficulty in absorbing the production into the market place. NO_x has limited industrial application and would probably have to go into nitric acid manufacture and hence into fertilizer production. World annual nitrogenous fertilizer production is in excess of 100 million tpa. and this should be sufficient to absorb the much smaller amounts of NO_x which would be co-produced.

Design considerations

Having chosen basic concepts the main components of the power processes both with and without capture were developed to enable costs and performance to be determined. When designing the oxy-combustion plants a number design considerations came up and choices had to be made. The important ones are outlined below.

Coal fired plant

Some of the re-circulating CO_2 has to replace air used in milling coal and transporting it to the burners. This stream, the primary recirculation, has to be dry enough to take up moisture from the coal feed and has to be preheated to 250-300C. Oxygen should not be added to this stream to minimize risk of fires and explosions in the mills.

The rest of the CO_2 recycle, the secondary recycle, can be re-circulated at higher temperature without drying. Thus the primary stream has to be cooled and dried before it is directed to the coal mills. The oxygen content of the secondary recycle should not exceed 40% to avoid the need to specify pure oxygen construction materials standards for the ducting. Thus a portion of the oxygen is added directly at the burner.

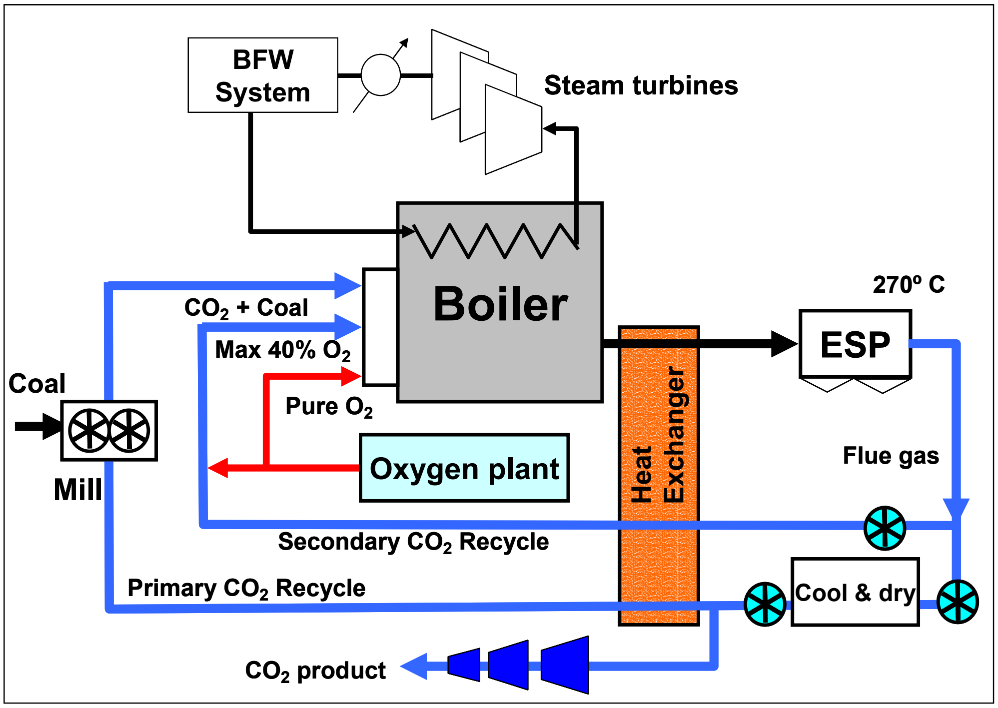

Fig 1 Flow diagram Coal plant with CO_2 capture

Flue gas from conventional coal fired power plants passes through an electrostatic precipitator (ESP) in order to reduce particulate emissions to the atmosphere. In the oxy-combustion process there are no stack emissions, except during startup, so the ESP's role is changed. Fine particle removal is still required in order to protect downstream heat exchangers, recycle blowers and in particular the CO_2 compression equipment. Thus an ESP remains a requirement.

A main considerations in placing the ESP is the operating temperature since higher temperatures increase the cost. However there is an efficiency advantage in taking the secondary recycle at elevated temperature directly from the outlet of the convection section at about 300C. This eliminates thermodynamic losses caused by cooling and then re-heating this stream. In this study a compromise in which the flue gas is cooled slightly to 230-270C has been adopted.

A significant simplification of the coal fired oxy-combustion capture process is the elimination of the Flue Gas Desulphurisation (FGD) unit and NO_x reduction equipment. Much of the SO_2 and NO_x are captured with the CO_2 although some will be removed with water condensing out during CO_2 compression. It is simple and cheap to extract these components during the CO_2 purification which already has to use cryogenic flashes to separate incondensible components such as O_2 , Ar and N_2. An extra distillation step is all that has to be added. Alternatively they can be left to be stored along with the CO_2. For more information on the co-capture and storage of these and other contaminants of captured CO_2 refer to report PH4/32.

Gas fired plant

The key design parameter for a new turbine required for an oxy-combustion cycle is limiting the stress in the turbine blades such that creep life is not reduced. A machine operating on CO_2 needs a lower rotational speed which reduces blade stress allowing a slightly higher maximum metal temperature for equal creep life. A further effect is the lower temperature drop with expansion over each stage of the turbine as compared to an air fired system, which is again due to the different properties of CO_2. This has the effect of significantly raising metal temperatures and it turns out that these two effects more or less counterbalance one another. The blade metal temperature is also affected by the cooling properties of CO_2 which are somewhat different to those of air. The chosen design has taken into account all of these factors so that the resulting machine does not overstep any of the principle boundaries of state of the art turbine design.

When CO_2 rather than air is used as the main component of the working fluid gas turbine efficiency with the same pressure ratio is significantly reduced. A final choice in the selected design was to increase turbine inlet pressure to 30 bar compared to the typical 15-20 bar of large conventional air machines. This pressure is considered to be the current proven limit for large gas turbines.

The re-circulated CO_2 and the fresh oxygen supplied to the gas turbine are clean so the need for air filtration is eliminated in the oxy-combustion process.

Performance and costs

The cost of electricity and cost of CO_2 avoidance for the cases evaluated are shown in the table below. The increase in electricity price is slightly lower for the coal fired processes. The capture cost per ton of CO_2 emission avoided for the gas fired process is considerably higher than for the coal plant. One reason for this is that large amounts of oxygen have to be produced simply to burn the hydrogen content of the natural gas which is much higher than that of coal. Oxy-combustion thus appears to be much more favourable for high carbon content fuels such as coal.

Fuel →	Coal Fired PF		Gas fired NGCC	
Case→	Base Case	Oxy-fired	Base Case	Oxy-fired
Cost of electricity ¢t/kwh	4.9	7.28	3.35	6.13
Cost of CO_2 avoided $/ton	-	37	-	77
Net power MW	677	532	388	440
Capital cost $million	1024	1246	217	658
Specific capital $/kw net	1513	2342	559	1495
Efficiency % LHV (net)	44.2	35.4	56	44.7

US$ costs are estimated on Q1 2005 basis €1 = $1.2

Note that an exchange rate of $1.2 per € has been used for capital cost estimating. Earlier studies have been made with an exchange rate of 1.0 and this should be taken into account when making comparisons.

Expert Reviewers' Comments

Expert reviewers were in general agreement with the results of the design and costing exercise for coal fired oxy-combustion plant. There were some reservations about the results obtained for the gas fired process which centered mainly on the costing for the new gas turbine which would have to be developed. This remains an area of uncertainty because the true extent of development costs and the size of the market from which these can be recovered is not known. The assumptions made in this report are that the market will be sufficiently large for development costs to be similar to those of machines in current large scale production. Because there would be no real out-steps in design with the approach adopted the development and testing of components and complete assemblies would be similar to that required for new models of conventional machines. The success or otherwise in the market place is not possible to predict. A successful machine with no competition could achieve very satisfactory results. However if competitors decided to make design step-outs which improved performance the conservative machine depicted in this study might soon be seen as obsolescent and might not have many sales. It does seem possible that the mature machine costs would only be reached after a period of technological innovation for the new type of machinery.

Some reviewers were concerned about the boundary parameters chosen for the new type of gas turbine. Turbine inlet temperature is often seen as a key design parameter when comparing gas turbines and one which should perhaps have been kept constant. However while this is fundamental to the efficiency of the thermodynamic cycle it is not the underlying constraint. This report attempts to explain that the fundamental constraint is the creep resistance of the metal of the turbine blades which operate at far lower temperatures than the gas entering the turbine. Fair comparison thus requires similar metals, coatings and blade cooling strategies to be used.

Reviewers also felt that a more extensive acknowledgement of previous work should be made. The authors have attempted to improve referencing. The main aim of this study has been to document the expected costs of the oxy-combustion option using IEAGHG standard evaluation methods as far as possible on the basis of today’s knowledge rather than to further research oxy-combustion technology.

Major Conclusions

The main conclusion is that the extra costs of electricity in CO_2 capturing oxy-combustion processes are similar to those for coal fired pre and post combustion options but somewhat higher for gas fired plant. Efficiency losses and costs of CO_2 avoidance are also of similar magnitude for coal fired plant but significantly higher for the gas fired process. As yet there is no obvious CO_2 capture technology preference, thus choices seem likely to determined by factors other than cost and efficiency at this stage. IEAGHG is preparing a comparison of the performance of all the processes which have been studied by the programme and this will be issued in due course.

The need for a completely new gas turbine remains a major hurdle for application of this technology to a gas fired CCGT process.

Recommendations

There are some early initiatives in the oxy-combustion field such as the CES water recycling process which uses rocket motor technology and the AZEP process which uses a dense oxygen membrane in conjunction with a gas turbine. A watching brief should be kept on these and other similar developments in oxy-combustion technology. The programme can provide limited guidance to these developments on performance benchmarks.

It is also proposed that a network be set up through which experts in the field of oxy-combustion can share information and better co-ordinate their research.

Oxy-combustion CO_2 capture process enable SO_2 and NOx to be separated rather easily as pure streams which in principle can be used to make sulphuric and nitric acids. It is important to establish whether this is a viable and practical destination for these byproducts. A small study to establish exactly how these co-captured materials could be recycled is thus recommended.

Technology & Engineering

IEA Greenhouse Gas R&D Programme: Oxy-Combustion Processes for CO_2 Capture from Power Plant

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P J Melling

SUMMARY

The IEA Greenhouse Gas R&D Programme has reported the cost of pre-combustion and post-combustion CO_2 capture technologies for power generation, except for oxy-combustion based technologies.

The use of oxy-combustion (fossil fuel combusted with near pure oxygen and recycled flue gas or CO_2 or water/steam) to produce a flue gas consisting essentially of CO_2 and water is seen as having potential as a means of disposing of combustion related CO_2. The advantage is that the flue gas is not diluted with nitrogen as when air is used for firing, and therefore can be disposed of with minimal further downstream processing.

A study was therefore commissioned by the IEA Greenhouse Gas R&D Programme and led by Mitsui Babcock with project partners Alstom Power, Air Products plc, and Imperial College London to confirm the feasibility and projected costs for CO_2 capture using oxy-combustion technology as applied to new-build power generation plant.

The study covered the following greenfield power generation plant with CO_2 capture:-

- Advanced Supercritical bituminous Pulverised Fuel (ASC PF)-Fired Power Plant
- Natural Gas Combined Cycle (NGCC) Power Plant

In compliance with the IEA GHG study specification, the target net power output of the power plant was aimed at 500 MW_e. To represent a reasonable baseline of oxy-combustion, the process evaluation was based on proven technology, where possible, with oxygen production based on cryogenic separation technology and the recycle system utilising either flue gas or CO_2.

The minimum CO_2 capture level specified for the study was 85% of all the CO_2 generated. The CO_2 was compressed to 110 bara before injection into the transfer pipeline. The economic analysis featured within this study is based on capture and CO_2 compression, costs associated with transport and storage are excluded.

This report presents the results of the study:

- An overview of oxy-combustion processes.
- Process flow diagrams and heat and mass balance calculations for the ASC PF Power Plant with CO_2 capture.
- Process flow diagrams and heat and mass balance calculations for the NGCC Power Plant with CO_2 capture.
- Discussion of safety and operability issues.
- Costs and economics associated with both types of CO_2 capture plants, including sensitivity to fuel price, discount rate and investment costs.
- Development requirements of oxy-combustion application for ASC PF and NGCC Power generation plants.

Process flow diagrams (PFDs) were developed using proprietary software and heat and mass balances (HMBs) compiled to model the overall oxy-combustion process. For the PF-fired power plant case, the required recycle flue gas system took into consideration dust loading and milling plant requirements. Consideration was also given to possible integration opportunities and heat and power utilisation to maximise plant performance.

Optimised plant performance parameters were used for cryogenic air separation with 95% v/v oxygen purity to suit the required CO_2 purity of 95% v/v with recognition of excess combustion oxygen and furnace/boiler tramp air ingress for the PF-fired power plant case.

The study compared both PF and NGCC plant with CO_2 capture, therefore the size of the associated plants were required to be similar. The target net power output for both plants was specified as 500 $MW_{e.}$ After detailed modelling of the ancillary plant, particularly the ASUs and CO_2 treatment plants, the net power output with CO_2 capture achieved were:-

- 532 MW_e for the ASC PF plant with CO_2 capture
- 440 MW_e for the NGCC plant power plant with CO_2 capture.

The PFDs/HMBs analysis produced data which allowed specifications for the major plant items to be developed; this in turn allowed budget capital costs for the major components to be assessed and operating philosophies to be investigated.

The table below summarises the main findings of the above activities:

		ASC PF Air Fired Power Plant Without CO_2 Capture	ASC PF Oxy-Combustion Power Plant With CO_2 Capture	Typical NGCC Air Fired Power Plant Without CO_2 Capture	NGCC Oxy-Combustion Power Plant With CO_2 Capture
Fuel Input	kg/s	59.19	58.09	14.77	20.32
Fuel Heating Value	MJ/kg (LHV)	25.86	25.86	46.90	48.45
Fuel Heat Input	MW_{th} (LHV)	1530.8	1502.2	692.9	984.5
O_2 Input (contained)	tonne/day	-	10,373	-	6,840
Gross Power Output	MW_e	740	737	400	575
ASU Power	MW_e	-	87	-	90
CO_2 Compression & Purification	MW_e	-	65	-	28
Power Plant Auxiliaries	MW_e	63	54	12	18
Net Power Output	MW_e	677	532	388	440
Gross Efficiency	% LHV	48.3	49.1	57.7	58.4
Net Efficiency	% LHV	44.3	35.4	56.0	44.7
CO_2 Capture Loss	**% points**	**-**	**8.9**	**-**	**11.3**
Unit Investment Costs:					
(Including Contingency, Fees & Owners Cost)*					
Coal & Ash Handling	US$ x 10^6	79	79	-	-
Boiler Island	US$ x 10^6	341	341	-	-
FGD Plant	US$ x 10^6	119	-	-	-
DeNOx Plant	US$ x 10^6	26	-	-	-
Steam Turbine Island	US$ x 10^6	195	195	-	-
ASU	US$ x 10^6	-	258	-	172
CO_2 Comp. & Purif.	US$ x 10^6	-	102	-	39
Balance of Plant	US$ x 10^6	264	272	-	-
NGCC Power Plant	US$ x 10^6	-	-	217	447
Total Investment Costs	US$ x 10^6	1024	1246	217	658
Specific Investment Cost	US$/$kW_e$ gross	1384	1691	543	1144
Specific Investment Cost	US$/$kW_e$ net	1513	2342	559	1495
Fuel Cost	US$/GJ	1.5	1.5	3.0	3.0
Cost of Electricity	**US cents/kWh**	**4.98**	**7.28**	**3.35**	**6.13**
CO_2 Emissions	t/h	489	45	144	5
CO_2 Captured	g/kWh (net)	-	831	-	403

*For PF plant: contingency 10% of Installed Costs; Fees 2% of Installed Costs; Owners Costs 16% of Installed Costs.
*For NGCC plant: contingency 10% of Installed Costs; Fees 2% of Installed Costs; Owners Costs 5% of Installed Costs.
Load Factor defined as achieved output as a percentage of rated / nameplate capacity (taken as 45% for 1st year. 85% every year after).
Exchange rate 1.2US$ = 1€
Capital costs +/- 25%

The costs of electricity production for each of the above cases were based on the IEA Greenhouse Gas Standard Economic Spreadsheet and assessment criteria at a 10% discount rate and an operating life of 25 years at an assumed 85% availability. The gas base fuel price applied was 3.0 US$/GJ for natural gas and 1.5 US$/GJ for coal.

A rate of exchange of 1.2 US dollar to 1 Euro was applied to capital cost estimates in line with current exchange rates.

The two main factors found to influence the cost of electricity were unsurprisingly, the investment cost and the fuel price.

Based on the technical and economic parameters presented in the report, the results show that for the ASC PF plant, the costs associated with the CO_2 capture plant compared to the cost of the reference ASC PF plant without capture is equivalent to a cost delta of 2.3 US Cents/kWh. For the NGCC plant the cost associated with the CO_2 capture plant compared to the cost of the plant without capture is equivalent to a cost delta of 2.8 Cents/kWh.

Whilst illustrating the envisaged potential, it is recognise that oxy-combustion technology needs further development in certain key areas in order to introduce the technology successfully to the marketplace.

Such areas requiring refinement are recognised as plant start up and control systems, burner and flame characterisation and materials issues.

A “stepped” approach in development is recommended for the case of the ASC PF plant by developing a “capture ready” plant that addresses many of the technical issues without committing utilities entirely to a CO_2 capture route.

Similarly for both types of oxy-combustion plant, the opportunity for further improvements in efficiency and reductions in costs are envisaged as oxy-combustion boilers, ASU, CO_2 compressors and gas turbine technologies advance.

GLOSSARY OF TERMS

ASC	Advanced Supercritical
ASU	Air Supply Unit
ADP	Acid Dew Point
BFW	Boiler Feed Water
BMS	Burner Management System
BOP	Balance of Plant
CoE	Cost of Electricity
DCAC	Direct Contact After Cooler
DCC	Direct Contact Coolers
DCS	Distribution Control System
DTI	Department of Trade and Industry
ESP	Electro Static Precipitators
FD	Forced Draught
FGD	Flue Gas Desulphurisation
FGR	Flue Gas Recycle
FWH	Feed Water Heating
GAS-ZEP	Gas-Zero Emission Power Plant
GHG	Green House Gases
GOX	Gaseous Oxygen
HMBs	Heat and Mass Balances
HP	High Pressure
HRSG	Heat Recovery Steam Generator
ID	Induced Draught
IEA	International Energy Agency
ITM	Ion Transport Membranes
IGCC	Integrated Gasification Combined Cycle
LOX	Liquid Oxygen
LP	Low Pressure
MAC	Main Air Compressor
MCR	Maximum Conditions Rating
MHE	Main Heat Exchanger
MP	Medium Pressure
NGCC	Natural Gas Combined Cycle
OTM	Oxygen Transport Membranes
OTSC	Once Through SuperCritical
PA	Primary Air
PFDs	Process Flow Diagrams
PF	Pulverised Fuel
PIC	Pressure Integral Controller
R&D	Research and Development
SCR	Selective Catalytic Reduction
TSA	Thermal Swing Adsorption
TIC	Total Investment Cost
UK	United Kingdom
ZEP	Zero Emission Power plant

CONTENTS

1. INTRODUCTION

1.1 Background

The IEA Greenhouse Gas R&D Programme has assessed and reported the relative cost of pre-combustion[1] and post-combustion[1A] CO_2 capture technologies for power generation, except for oxy-combustion based technologies. The use of oxy-combustion (fossil fuel combusted with near pure oxygen and recycled flue gas or CO_2 or water/steam) to produce a flue gas consisting essentially of CO_2 and water is seen as having potential as a means of disposing of combustion related CO_2. The advantage is that the flue gas is not diluted with nitrogen as when air is used for firing and therefore can be disposed of with minimal further downstream processing.

Previous work from a number of authors [2,3,4,5] as well as pilot and laboratory scale studies[2] have indicated that retrofit of oxy-combustion based technology to existing fossil fuel fired plant is feasible with little or no changes to the existing boiler pressure parts. Projected costs have generally been comparable with those for current pre- and post combustion capture technologies, the major costs compared with conventional air firing being associated with the production of the oxygen and the compression and treatment of the CO_2 product.

Given the perceived benefits, the IEA Greenhouse Gas R&D Programme commissioned Mitsui Babcock to undertake a study to confirm the feasibility and projected costs for CO_2 capture using oxy-combustion technology as applied to new-build power generation plant. The study was to be based on a bituminous pulverised fuel (PF) power plant and a natural gas combined cycle power plant.

A small experienced project team was assembled by Mitsui Babcock to execute the oxy-combustion study. The project team comprised of the following partners along with their anticipated split in the scope of work.

Mitsui Babcock acted as main contractor. As a manufacturer of large combustion plant, Mitsui Babcock assessed the aspects relating to the application of oxy-combustion technology to a new-build state-of-the-art PF-fired power generation plant.

Alstom Power developed the oxy-combustion process as applied to the natural gas power generation plant as well as providing details on the steam turbine island for both power generation schemes. Application of oxy-combustion technology to a natural gas combined cycle power plant will necessitate the design of a novel gas turbine due to the very large changes in gas properties from air to carbon dioxide and water.

Air Products plc applied their considerable expertise in air separation technology, downstream CO_2 processing and systems integration to the two power generation plants considered in this study. Oxygen separation was based on cryogenic air separation.

Imperial College London investigated possible integration opportunities and heat and power utilisation to maximise efficiency and minimise the cost of CO_2 capture for the two power generation plants considered in the study.

1.2 Scope Of Work

This report presents the results of the following activities: -

1.2.1 Overview of Oxy-Combustion Processes

A general overview of the applicable oxy-combustion process with simplified block schemes of the main component variants. The reasoning behind the selection of the two representative processes for large-scale power generation based on a PF fired plant and a natural gas fired plant is clarified.

1.2.2 Process Flow Diagram and Heat & Mass Balance Calculations for the PF Plant

A high level process flow diagram (PFD) is developed and heat and mass balances (HMBs) compiled to model the overall oxy-combustion process for PF plant. The overall PFD is generated from iterations between independent flowsheets. Mitsui Babcock was responsible for the Once Through Super Critical Boiler Island (OTSC) and flue gas recycle components; Air Products for the Air Separation Unit (ASU) and CO_2 purification and compression components and integration with the power plant, and Alstom Power for providing the steam turbine island input. Mitsui Babcock collated the overall PFD based on partners inputs. Consideration was given to possible integration opportunities and heat and power utilisation to maximise plant efficiency as investigated by Imperial College London. A criteria of 500MW$_e$ net power output, as set by the IEA in line with previous studies on post- and pre-combustion technologies was targeted.

1.2.3 Process Flow Diagram and Heat & Mass Balance Calculations for the NGCC Plant

A high level PFD is developed and HMBs compiled to model the overall oxy-combustion process for NGCC plant. This work being led by Alstom Power with input from Air Products regarding the ASU and CO_2 compression components. A criteria of 500MW$_e$ net power output, as set by the IEA in line with previous studies on post and pre combustion technologies was targeted, however, in practise due to Gas Turbine sizing, a lower figure was actually achieved.

1.2.4 Safety and Operability

Any aspects of the oxy-combustion process which impacts on the safety and operability of the power plant are highlighted in the study. The study outlines the safety and operability requirements, however, it is not intended to resolve these issues within this report.

1.2.5 Development Requirements

The principle equipment development requirements leading to demonstration of the oxy-combustion technology as perceived by the Project Team are outlined together with an order of magnitude of development costs.

1.2.6 Costs and Economics

In order to arrive at the power cost per kWh, an economic analysis has been executed by Mitsui Babcock using the IEA Greenhouse Gas R&D Programme ‘Technical and Financial Assessment Criteria : Revision B : May 2003 and IEA Greenhouse Gas standard economic assessment spreadsheet. The basic cost elements being capital cost (± 25%), operating cost and agreed economic parameters.

In order to compare the influence of discount rate and fuel costs on power costs, a sensitivity analysis is undertaken based on a 5% discount rate (compared to 10% for the base case) and fuel price variation from the base case of ±100%.

1.3 Deliverables

The deliverables of the study for each of the two greenfield power generation plants are as follows:

- General overview of oxy-combustion processes and simplified block diagram.
- Process description.
- High level process flow diagram and heat & mass balance.
- Major equipment list.
- Budget capital cost estimate (to within ± 25%).
- Operating cost estimate.
- Economic analysis: power cost in US c/kWh and CO_2 emissions in g/kWh.
- Overview of integration opportunities and heat and power utilisation to minimise the cost of CO_2 capture.
- Sensitivity of performance and costs to CO_2 purity: 95% mol vs. 98%mol, at 110 bara.
- Sensitivity of abatement costs to discount rate and fuel cost.
- Statement regarding additional processing requirements for reduced SO_2 and O_2 contaminants in the final CO_2 product.

- Statement on issues associated with materials, corrosion and build-up of impurities.
- Statement on the requirements for plant start-up and shutdown.
- Development requirements and order of magnitude of development costs.
- Statement on safety requirements for oxygen handling.

2. OVERVIEW OF OXY-COMBUSTION PROCESSES

2.1 Oxy-Combustion Application to Utility Power Plant

2.1.1 Basic Oxy-Combustion Pulverised Coal Concept

An outline block flow diagram for conventional air-fired pulverised coal combustion is shown in Figure 1. Combustion air is divided into two flows. Approximately 25%, the primary air, is mixed with coal in the mills where (after heating by the flue gases) it serves to dry the coal, to circulate it within the mill, to carry it to the burners to provide pre-mixed oxygen for the early combustion stages. The secondary air is also heated by the flue gas and sent to the burners to provide the balance of the combustion air.

The 21/79 volumetric oxygen/nitrogen (atmospheric nitrogen) ratio in air is the main factor determining peak flame temperatures in the furnace (~ 2000 ºC adiabatic flame temperature), with the nitrogen acting as an inert dilutent. The furnace is designed to cool the flue gases by radiation to a temperature of about 1000 ºC at the exit so that fly ash particles have solidified sufficiently not to form intractable deposits on the convective passes.

In oxy-combustion the desired overall combustion process is coal with essentially near pure oxygen but, to use existing, proven boiler technology, flue gas must be recycled to take the place of primary air for pulverised fuel transport from the mills and to be an inert dilutent to moderate peak and exit combustion temperatures in the furnace, as shown in Figure 2. Because of the higher heat capacity of carbon dioxide a reduced inert gas/O2 ratio is required (approaching 2.5:1 instead of 4:1 for air).

The primary and secondary recycle flows are each approximately equal to the net products of combustion, but exact values vary with configuration and operating parameters such as required temperatures, fuel/oxidant ratios and assumed air in-leakage rates. The flow configuration options for cleaning and heating the recycle flows can also vary, as discussed below.

2.1.2 Black Coal Characteristics and Consequences for Primary Recycle Requirements

Black (Bituminous) coals are characterised by:

- low reactivity - must be finely ground (e.g. 80% <75µm) for satisfactory ignition and burnout

- low grindability – rolling element pulverised coal mills with classifiers and oversize recycling are required

- hence large coal inventory in the mill - danger of fires and deflagrations

- Limited mill exit temperatures (typically 60-80ºC) - due to fire risk and also tendency for some coals to melt and adhere to the burner (can then cause severe overheating).

These characteristics have important consequences for black coal oxy-combustion plants.

- Primary recycle must be dry - needs to be able to carry coal moisture as vapour at relatively low temperatures.

- No oxygen to the mill – due to the risk of explosion if oxygen gets too high or recycle gets too low (this makes coal ignition more difficult and requires a higher concentration of O_2 in the secondary recycle and/or direct oxygen addition at the burner). No safety case based on O_2 detection and interlocks is likely to be acceptable given the number of mills, the frequency of mill start/stops and the very serious consequences of a mill explosion at elevated oxygen levels.

- Primary recycle temperatures limited by mill constraints such as mill bearings to 250-300ºC.

- Limited range of primary recycle flow rates that will give acceptable solids entrainment within the mill and through the classifier and out of the mill to the burners.

- Cannot have risk of acid corrosion in the complex mill machinery, so must have low SO_3 content in the primary recycle.

(Note that the situation is entirely different for low rank brown coal plants. These are already designed to recycle flue gas from the furnace exit for the primary recycle stream and can use once-through beater mills/blowers because brown coals are friable and reactive.)

Although other configurations are possible in theory, the constraints above leave little latitude in practice in the way the primary recycle is handled with black coals. The primary stream must be cooled and scrubbed (i.e. with the product gas) to remove moisture and soluble acid gas components such as SO_3 and HCl and then be reheated to 250-300ºC before feeding the mills.

The ratio of coal to primary recycle is also fixed for current mill technologies. A similar volumetric flow rate would be required to give the same solids-carrying ability and hence a proportionally higher mass flow rate.

It would probably be feasible as a future development to design pulverised coal mills to handle acid recycle gases, although the widespread use of corrosion-resistant alloys that this would require would add considerably to the capital cost. It is also possible, however, that an entirely satisfactory compromise between corrosion and abrasion resistance would not be achievable, leading to high maintenance costs. It would also be feasible to operate with lower primary recycle flows, although the energy required to dry the coal will limit the amount this can be reduced, depending on coal moisture content and allowable/achievable primary recycle temperatures.

An alternative option would be essentially an indirect milling system, with hot air or dry nitrogen from the ASU used to dry the coal and transport it through the mill. After the mill the pulverised coal and air or nitrogen is separated; the air or nitrogen is vented and the coal is re-entrained in recycled flue gas. There is no loss in discarding the latent heat in the coal moisture, since the mill exit temperature will almost certainly be lower than the effective final boiler exit temperature. Such an indirect milling system would remove the need for the dry primary recycle, thus allowing all recycle gas to be recycled hot, benefiting overall plant efficiency. However, the milling plant becomes more complex and difficult to operate, and significantly less similar to the proven and robust direct fired pulverised coal plant.

2.1.3 Secondary Recycle Configuration Options

Options do exist for the way the secondary flue gas recycle is handled (see Table 1). In all cases solids removal is required in an electrostatic precipitator (ESP) to avoid excessive ash concentrations in the boiler flue gas passes and erosion of the hot recirculating fan. The secondary recycle must then be recycled to the boiler at ~ 300ºC, to promote ignition, and possibly have oxygen added. Three options are possible, characterised by increasing operating temperatures for the ESP and recirculating fan and greater gas/gas heat exchanger duties.

A) Figure 3 - Secondary recycle cooled close to ambient (i.e. below acid dewpoint), moisture condensed and SO_3 removed before recycling through a gas/gas heater. ESP temperature limited by the acid dewpoint to about ~160ºC. Note: Acid dewpoint based on method illustrated by Pierce[5A] with assumption made on $SO_3 / (SO_2 + SO_3)$ conversion rate of 1.0%.

B) Figure 4 - Secondary recycle cooled but not below acid dewpoint, no moisture removal, recycled through a gas/gas heater. ESP temperature determined by heat balance; 230-270ºC depending on recycle flows and economiser gas exit temperatures.

C1) Figure 5 - Secondary recycle not cooled, recycled at ESP temperature ~300ºC, variation C1 (parallel gas heater and FWH) for subsequent heat transfer from combined product and primary recycle stream.

C2) Figure 6 - Secondary recycle not cooled, recycled at ESP temperature ~300ºC, variation C2 (series gas heater and FWH) for subsequent heat transfer from combined product and primary recycle stream.

C3) Figure 7 - Secondary recycle not cooled, recycled at ESP temperature ~300ºC, variation C3 (FWH only) for subsequent heat transfer from combined product and primary recycle stream.

Efficiencies are likely to increase slightly from A to C because the average temperature for potential low temperature heat recovery in the flue gas feed water heater will increase, but there is a finite scope to use heat at these temperatures in the steam cycle (for feed water heating). Equipment design moves further from conventional configurations from options A to C, possibly leading to higher costs and certainly affecting the ability to switch between oxy-combustion and air-fired operation.

In this study option B has been used as representing a reasonable compromise between optimising efficiency and matching current practice.

2.1.4 Excess Oxidant, Oxygen Addition and Air Leakage

Typically 15-20% excess air is required in pulverised coal combustion to ensure adequate burnout. The amount of excess air is thought to be largely determined by inevitable variations in air/fuel distribution between burners in multi-burner installations rather than the need to achieve a particular oxygen concentration. It is essential that burners with high fuel loadings and low air flow rates still have a slight excess of air, otherwise very high unburnt carbon levels will inevitable occur (typical burnout levels are of the order of 99% conversion). Non-ideal fuel distributions are still likely to occur with oxy-combustion. Oxygen distribution may also be at least as variable if it is added to the secondary recycle, since any mixing variations must be compounded with total flow variations. Improved (and more controllable) oxygen distribution could probably be achieved with direct oxygen supply to the burners. An excess oxygen level at the burner from all sources (recycle and added oxygen) equivalent to a 10-20% excess air would therefore appear reasonable.

As already discussed due to safety considerations, it appears inadvisable to add oxygen to the primary recycle. If all the combustion oxygen is added to the secondary recycle oxygen concentrations will be in the region of 50% by volume. Systems with oxygen purities greater than 40% by volume must comply with strict rules for metals selection and vapour velocity. Systems with oxygen purities from 23% to 40% by volume, in regards to metals selection, may be designed as an air system and the rules for oxygen velocity do not apply. However, the system shall still be designed as a clean gas system with full compliance for oxygen cleanliness and compatibility of non-metallic materials. It therefore appears that at least some of the oxygen must be added at the burner and that 23 mol% oxygen should be the limit set for the secondary recycle.

Virtually all coal-fired boilers are designed to be operated under slight suction to avoid danger to operators and equipment from leakage of hot gases, dust and carbon monoxide and current construction methods appear unlikely to be suitable for any other approach. Even if an unusually gas-tight containment could be achieved with a newly commissioned plant thermal distortion and cracking is likely to lead to a decrease in containment integrity between major overhauls at 3-5 year intervals. Tramp air Leakage levels in line with expectations for current new plant (see Section 3.1.1) therefore appear likely. (Note: oil and gas boilers can be designed for pressurised operation).

With any appreciable level of air in-leakage it becomes impossible to avoid the need for inert gas removal from the CO_2 product to achieve design purity. The extra capital and power associated with high purity oxygen is not justified. Lowering the oxygen purity will significantly reduce the power consumption of the ASU. An oxygen purity of 95% v/v used in this study is considered to represent a reasonable compromise between inert removal and oxygen production costs.

2.1.5 Integration and Performance Optimisation

The main scope for integration to optimise overall system performance is the appropriate use of the additional low-temperature heat available in the oxy-combustion configuration from:

- Flue gas cooling to ambient (in modern air-fired combustion plant with FGD, flue gas feed water heaters are in use for a similar duty).
- Heat rejected from gas compressors (the main air compressor in the ASU and the CO_2 compressor are operated adiabatically with heat recovery to the condensate and boiler feedwater.)

Heat can (if available) be used at all temperatures up to the boiler feed water temperature at the economiser inlet for feed water heating (FWH). This displaces steam bled from the steam turbine for feed water heating and allows more power to be produced in the turbine.

Only limited amounts of heat can be used over any given temperature range, but sufficient scope exists for both of the above heat recovery options. It is also necessary to coordinate auxiliary feed water heating with the constraints on steam tapping pressures in the main steam cycle for the deaerator and higher pressure FWH.

In this study, two stages (before and after the deaerator) of flue gas FWH are employed using heat from the combined primary recycle and product gas stream after the ESP. Other feed water heating from the air separation unit compressor aftercoolers and CO_2 compressor aftercoolers is used to replace steam abstraction from the low and intermediate pressure steam turbines for most of the conventional steam feed water heating requirements.

2.2 Oxy-Combustion Application to NGCC Power Plant

2.2.1 Gas Turbine Oxy-Combustion Cycle Concept Selection

To date, numerous studies have been undertaken with regards gas turbine oxy-combustion concepts, notably by Mathieu and Bolland [6-25].

There is a choice[26] when considering a gas turbine combined cycle power generation system designed for CO_2 capture.

- Pre-combustion CO_2 removal from a fossil fuel[1] which would be converted into hydrogen which would fuel a conventional air cycle gas turbine having a combustion system designed for use with a hydrogen rich fuel gas.

- Oxy-combustion of a fossil fuel in a gas turbine involves combustion of the fossil fuel in near pure oxygen diluted with compressed recycled flue gas which would be predominantly CO_2. This means that the gas turbine working fluid is CO_2 rich gas and that all of the oxygen required for combustion must be provided from an air separation system. This gas turbine configuration will require a completely new gas turbine design since the physical properties of CO_2 are different from air.

- Post-Combustion CO_2 removal involving the separation of CO_2 from flue gas by means of scrubbing the flue gas with a chemical solvent (usually amine) and then heating the solvent to release high purity CO_2[1A].

The oxy-combustion gas turbine with CO_2-rich gas as the working fluid is the process system considered in this study.

3. ADVANCED SUPERCRITICAL (ASC) PF POWER PLANT WITH CO_2 CAPTURE

3.1 Design Basis, Process Description and Performance

Section 3.1.1 below presents the summary of the ground rules used as the design basis for the study, as agreed with project partners. Descriptions of the base case ASC PF power plant without CO_2 capture and the ASCI PF power plant with CO_2 capture are presented alongside heat and mass balances and performance data for the respective processes are presented in Sections 3.1.2 and 3.1.3 respectively.

3.1.1 Technical & Financial Ground Rules

The main technical / financial ground rules for this study have been based on IEA GHG R&D programme Technical & Financial Assessment Criteria [27]. Key points are summarised below:

Technical Design Basis:

Site Conditions

Ambient Air temperature: 9 °C

Ambient air relative humidity: 60%

Ambient air pressure 1.013 bar

CO_2 Processing

Minimum CO_2 capture level is to be 80%, with the preferred level 85%

CO_2 is to be compressed to 110 bara before injection into the transfer pipeline.

Further detailed operating parameters for the ASC PF plant with CO_2 capture were established based on discussions and agreement with all partners and the appropriate IEA GHG R&D programme representative. These parameters being specifically:

Size of Power Plant

The study would compare both PF and NGCC plant with CO_2 capture, therefore the size of the associated plants were required to be similar. The net power output for both power plants was specified[27] as 500 MW_e. Therefore, based on plant ancillary power approximations, a gross power output of 740 MW_e was targeted for the PF power plant with CO_2 capture and gross output of 575 MW_e targeted for NGCC plant with CO_2 capture based on GT size.

After detailed modelling of the ancillary plant, particularly the ASUs and CO_2 treatment plants, the net power outputs with CO_2 capture were 532 MW_e for the PF plant and 440 MW_e for the NGCC plant. The PF oxy-combustion process could be remodelled to converge on 500 MW_e net output but time did not allow this level of interaction between all of the integrated plant elements.

O_2 Purity

The O_2 purity supplied by the Air Separation Unit (ASU) was specified at 95 %v/v, this figure being derived from previous studies[28] as the optimum purity associated with oxy-combustion PF plant given tramp air ingress as discussed below.

Recycle rate

A recycle rate of approximately two thirds of the flue gas was used in this study for the ASC PF Oxy-Combustion Power Plant, and is consistent with previous studies in the oxy-combustion field. This rate was set to maintain acceptable combustion temperatures based on current air–fired PF boiler technology.

Coal Specification

The bituminous coal specification[27] is presented in Table 2.

Tramp Air Ingress

Tramp air ingress is a fact of life on large boiler plant with balanced draught. For safety reasons these boilers operate under suction from the burner exit (i.e. in the furnace chamber) through to the induced draft fan. Hot flue gases are therefore prevented from leaking out of the boiler casing; any leakage is of cold air into the flue gas stream. In conventional plant such in-leakage represents a loss in efficiency and active measures are taken to keep any such infiltration to a minimum. The major sources of air in-leakage are through openings in the furnace membrane wall, penetrations in the boiler casing for the hot pipework to the boiler heating surfaces, at the induced draft fan, in the ash removal unit and through any gaps/penetrations/expansion joints in the ducts and casings.

Obviously any air in-leakage has a detrimental effect on the efficiency of the proposed process as it can significantly increase the duty of the CO_2 purification plant.

The ASC PF plant with CO_2 capture performance model was based on sensible levels of tramp air ingress associated with boiler and balance of plant:

- Furnace + boiler air infiltration rate = 0.01kg air / kg flue gas
- Electrostatic precipitator (ESP) air infiltration rate = 0.02kg air / kg flue gas

Further cost may be incurred throughout the lifetime of the boiler, to ensure reasonable levels of sealing are maintained.

Water

The use of sea water is assumed for cooling; Average inlet temperature 12°C; maximum temperature rise 7°C; salinity 22 grams/litre.

Fuels

Bituminous coal and natural gas are used as the design fuels. The cost of the coal delivered to site is assumed to be 1.5 US$/GJ (LHV Basis). The cost of the natural gas delivered by pipeline to site is assumed to be 3.0 US$/GJ (LHV Basis). Sensitivity to the cost of power generated for a range of fuel prices (0-3 US$/GJ for coal and 0-6 US$/GJ) for gas is considered.

Financial Assessment Base Data

Location

A green field site with no special civil works will be assumed. Adequate plant and facilities to make the plant self sufficient in site services will be included in the investment costs.

Currency

The results of the studies will be expressed in US $ applicable to a specific year. (Note: An exchange rate of 1 Euro = 1.2 US $ has been applied).

Design and Construction

As projected finances can be sensitive to the time required to erect the plant the following factors have been applied:

For PF:	Year 1	20% investment
	Year 2	45% investment
	Year 3	35% investment
For NGCC	Year 1	40% investment
	Year 2	40% investment
	Year 3	20% investment

Load Factor

Defined as the achieved output as a percentage of rated nameplate capacity. A load factor of 85% has been applied to both PF and NGCC cases.

Project Life

A design life of twenty five years is assumed as the basis for the economic appraisal.

Discount Rates

Discounted cash flow calculations will be expressed at a discount rate of 10% and to illustrate sensitivity, at 5%.

Contingencies

A contingency is added to the capital cost to allow for unforeseen set backs, cost under estimates, programme overruns. The value of the contingency is taken as 10% of the installed plant costs.

Fees

Fees are taken as 2% of the installed plant costs. These include process/patent fees and costs, fees for agents or consultants, legal and planning costs land purchase surveys and general site preparation.

Owners Costs

Owners costs are 20% of the installed plant costs for the PF air firing plant. For the PF capture plant the owners costs were assumed as 16% due to the influence of the ASU and CO_2 compression train which have a lower owners cost attributed of 5% of the installed plant costs.

Owners costs are 5% of the installed plant costs for the NGCC plant for both air-fired and CO_2 capture plants.

Operating Costs

Operating costs are made up from variable costs and fixed costs. The fixed costs are generated from the sum of direct labour, administration and maintenance, where administration is considered at 30% of the labour costs and the maintenance 4% of the installed costs (excluding contingency). The variable cost is essentially the fuel costs.

3.1.2 ASC PF Air Fired Power Plant Without CO_2 Capture: Base Case Process Description

3.1.2.1 Description

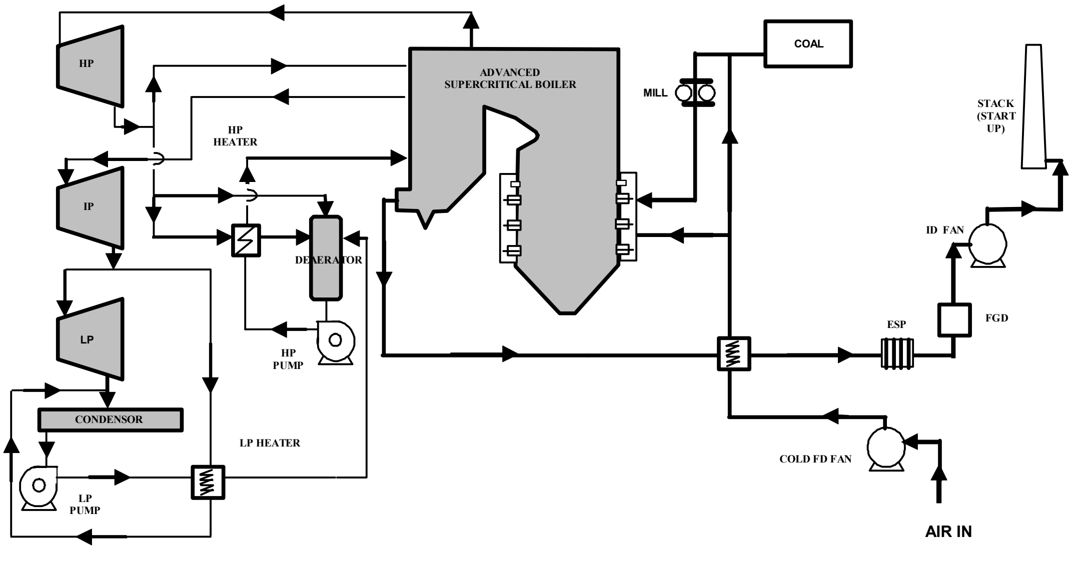


Figure 8: Conventional Air Fired PF Bituminous Plant Arrangement

Figure 8 outlines the conventional plant arrangement for the combustion of pulverised coal for large scale steam raising electricity generation without CO_2 capture.

Wet coal, containing around 10% moisture, is supplied to the milling plant where it is pulverised to typically 70% passing 75 micron. Whilst in the mill the coal is also dried, the air temperature at the mill inlet being controlled to give a mill outlet temperature of typically 90°C. The combustion air itself is preheated to typically 270°C in a regenerative airheater by the flue gases leaving the boiler. Around 20% of the combustion air is supplied to the mills and transports the pulverised coal pneumatically to the burners. The remaining 80% of the combustion air is supplied directly to the furnace chamber, either at the burners or as overfire air in the case of furnace air staging for NO_x control.

The coal is burnt in a highly radiant flame where it releases its heat. About 35% of the heat is absorbed in the radiant furnace chamber, 15% to the radiant platen superheaters suspended in the upper region of the furnace chamber, and 40% by convective heat transfer to superheater, reheater and economiser surfaces in the boiler and to the airheater. The remaining 10% is lost, predominantly in the sensible heat of the flue gases exhausted to the stack, giving a typical overall thermal efficiency of the plant (i.e. chemical heat to steam) of around 90%.

In modern pulverised coal fired utility furnaces the coal is burnt with typically 17½% excess air, giving rise to a flue gas at the boiler exit of the following composition:-

	% w/w	**% v/v**
Oxygen	3.45	3.21
Carbon dioxide	20.92	14.14
Sulphur dioxide	0.39	0.18
Water	5.02	8.30
Nitrogen	70.21	74.18
	100.00	100.00

The flue gases also contain NO_x ash and Hg. The concentrations of these species can vary considerably from plant to plant and are dependent upon the equipment installed and how well it is maintained. For a modern well maintained plant within low NO_x burners installed NO_x emissions are typically around 350 ppm at 6% O_2 v/v dry.

The scale of the required CO_2 separation duty becomes apparent when one considers that a single 660 MW_e unit produces 680 kg/s of this flue gas. Whilst there are a number of processes available to remove the CO_2 from the flue gas the problem is one of size - the cost of treating such a large quantity of flue gas economically. It is by seeking to minimise the problem rather than treating the products arising from combustion which gives rise to the CO_2 removal system upon which the project is based.

3.1.2.2 Case 1: ASC PF Air fired Power Plant without CO_2 Capture: Process Flow Diagram

For the costing purposes of this study and to highlight differences between conventional plant and plant designed for CO_2 capture, the key components of the ASC PF air-fired power plant without CO_2 capture have been identified as units. PFD 1 illustrates the major components and process streams for a conventional ASC PF power plant. Namely:

Unit 100: Coal and ash handling

Unit 200: Boiler island

Unit 300: FGD and handling plant

Unit 400: $DeNO_x$ plant

Unit 500: Steam turbine island

Unit 800: BOP, electrical C&I

These can therefore be directly compared with the key equipment requirements of the PF power plant with CO_2 capture.

3.1.2.3 Case1: ASC PF Air Fired Power Plant without CO_2 Capture: Performance

A 740 MW_e gross (677MW_e net), balanced draught, advanced supercritical PF power plant with 290 bara/600°C main steam and single reheat to 620°C was employed as the base case (Case 1) scenario. The overall net cycle efficiency for the plant is 44.2% (LHV) based on a condenser pressure of 40mbar. The PF plant included Selective Catalytic Reduction (SCR) and Flue Gas Desulphurisation (FGD) equipment to meet current NO_x and SO_x emissions legislation.

The breakdown of the total auxiliary power consumption associated with this plant is presented in Table 3.

3.1.3 Case 2: ASC PF Oxy-Combustion Power Plant with CO_2 Capture: Process Description

3.1.3.1 General Description

The flue gas produced by the combustion of coal in air is mostly nitrogen. If the air is separated into its constituent components prior to combustion and only oxygen is supplied to the furnace then the resulting flue gas will contain only the products of combustion - the inert nitrogen “ballast” will have been eliminated and the quantity of flue gas to be treated will be significantly reduced.

This removal of the nitrogen ballast is at the heart of the proposed process. However the reduced mass and volume flow through the plant is perceived to lead to a number of difficulties if applied directly to conventional combustion plant. In the furnace chamber the introduction of the same quantity of heat to a reduced mass of combustion products is believed to result in greatly increased temperatures. As a result increased radiant heat pick-up, greater slagging and higher NO_x emissions are all anticipated. Furthermore, the reduced volumetric flow (and hence gas velocity) in the convective passes of the boiler leads to lower heat transfer coefficients and reduced heat absorption. Therefore the overall balance of the heat absorbed throughout the unit is likely to be so disturbed as to make the plant inoperable without substantial modification to the heating surfaces.

The problem is simply resolved by recycling a proportion of the flue gas back to the furnace so as to maintain the mass/volume flow at an acceptable level. It is therefore possible to devise a conceptual process diagram whereby an existing pulverised coal fired utility boiler can be operated without nitrogen being present in the flue gas, resulting in a substantial reduction in the quantity of flue gas that must be treated in downstream processing equipment to capture the CO_2.

Figure 9 shows the arrangement selected for the basis of this particular study. There are three key features that distinguish the process from conventional coal combustion plant

- The combustion air is separated prior to the furnace in an air separation plant - only the oxygen (95 mol% oxygen with 2 mol% nitrogen and 3 mol% argon) is supplied to the combustion process, the nitrogen/argon is vented to atmosphere.
- Part of the flue gas leaving the boiler, around one third of the total flow, is recirculated back to the burners to provide the part of the ballasting effect formerly carried out by the atmospheric nitrogen.
- The net flue gas is cooled and passed to the CO_2 compression and inerts separation plant. The gas is further cooled by direct water contact which removes residual ash, the primary recycle is separated and recycled to the coal mill. The net product is compressed, dried, the inerts separated and the pure CO_2 is further compressed to pipeline pressure of 110 bara.

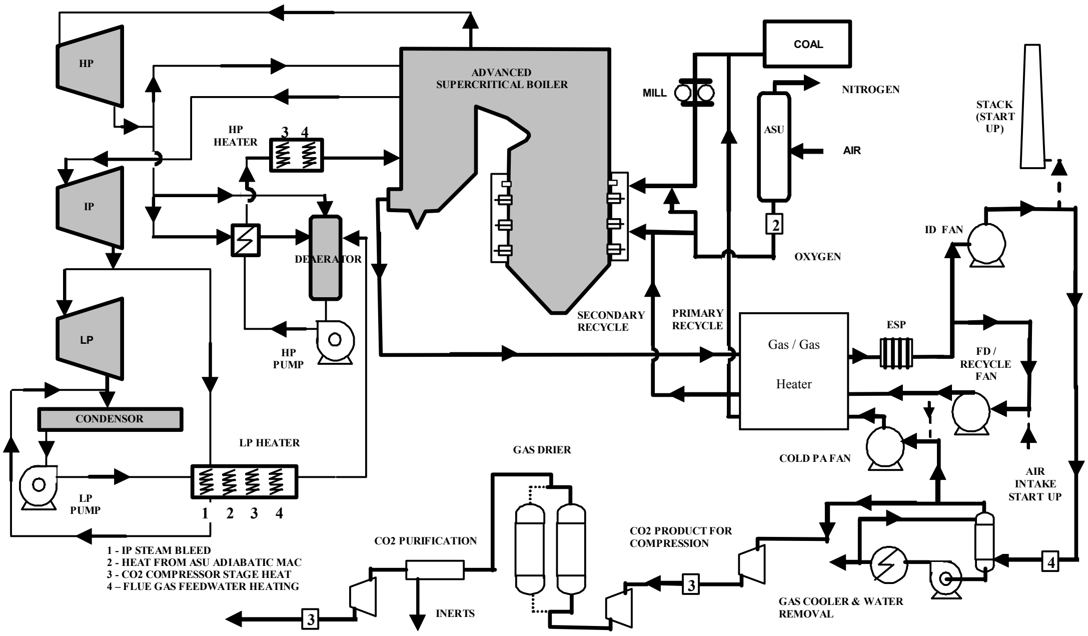


Figure 9: Oxy-Combustion ASC PF Power Plant with CO_2 Capture: Plant Arrangement

Utilising oxy-combustion and flue gas recycling, the supercritical boiler takes 524 kg/s of feed water (325 bara @ 270°C) and generates 516 kg/s of HP steam (290 bara @ 600°C) and 411 kg/s of reheat steam (60 bara @ 620°C) at the steam turbines.

To ensure complete combustion excess O_2 at the burners is maintained at 19% over the amount required for stoichiometric combustion (i.e. similar levels of excess air with air firing). The oxygen flow (127.1 kg/s) is supplied at 95%v/v purity to the burners via the ASU. This cryogenic system developed by Air Products is described in detail in (Section 3.1.3.2.3).

Two streams of recycle flue gas are required for the oxy-combustion system:

- Primary recycle (152 kg/s) which passes through the coal mills and transports the PF to the burners. The volumetric flow rate of primary recycle gas is maintained at value required for air firing.

- Secondary recycle (258 kg/s) which provides the additional gas ballast to the burners to maintain temperatures within the furnace at similar levels to air firing.

The combined primary and secondary gas recycle is approximately 67% of the original flue gas leaving the economiser.

The flue gas exiting the boiler at 340°C is used to heat the primary and secondary recycle flue streams via a regenerative gas / gas heater. The flue gas is de-dusted via the ESP. The clean flue gas is then split into two, with one stream forming the secondary recycle and returning back through the gas / gas heater (exit temp 330°C) to the burners. The remaining stream is cooled, dried and split again to form primary recycle and CO_2 product streams respectively. The primary recycle passes through the gas / gas heater (exit temperature 250°C) and is delivered to the coal mills. The PF is dried in the mill using this flow (mill exit temperature 105°C) and transported to the burners.

The net flue gas is then passed through a compression and inerts removal stage that delivers a final CO_2 product (126.96 kg/s) of 95% v/v purity, at 110 bara. The details of the compression and inerts removal are described in (Section 3.1.3.2.4).

Although the process concept is very simple in principle there are a number of possible options for its implementation on existing and new plant designs. These options have been discussed previously in Section 2.0 of the report. However, the main advantages / disadvantages associated with the process scheme (option B) adopted are as follows:

Advantages

- Low moisture content in primary recycle flue gas to milling plant.
- Minimum quantity of recycle gas being cooled and reheated, thus efficiency benefits.
- Concept applicable to normal start up with air firing.
- Concept utilises conventional plant equipment and layout (e.g. Gas Airheater and ESP positioning).
- Minimum number of fans required for air start up and oxy-combustion operation (3 fans x 2 flue gas streams or 3 fans x 1 flue gas stream depending on boiler flue gas configuration).

Disadvantages

- The design and performance of the ESP will require to take into consideration the impact of elevated flue gas inlet temperatures on ash resistivity based on the sulphur content of the range of coal qualities fired.
- Higher levels of flue gas recycle system impurities such as SO_2 and HCl and their impact on the boiler flue gas passes.

3.1.3.2 Case 2: ASC PF Oxy-Combustion Power Plant with CO_2 Capture: Process Flow Diagram

PFD 2 illustrates the major components and process streams the Case 2 ASC PF Power plant with CO_2 capture:

Unit 100: Coal and ash handling

Unit 200: Boiler island

Unit 500: Steam turbine island

Unit 600: Air Separation Unit

Unit 700: CO_2 compression & inerts removal

Unit 800: BOP, electrical C&I

In comparison with the base case plant without CO_2 capture (Case 1), Unit 300: FGD and Unit 400: $DeNO_x$ are no longer required with the addition of Unit: 600 Air Separation Unit and Unit: 700 CO_2 Compression & Inerts Removal, a detailed description of these units and their stream data is provided in Sections 3.1.3.2.3 and 3.1.3.2.4.

Furthermore, for the CO_2 capture Case 2, a greater degree of heat integration among separate units is possible. Both condensate heating and boiler feedwater heating is utilised between the steam turbine island, the ASU plant, the boiler island, and the CO_2 compression and inerts removal plant. This is further discussed in Section 3.1.3.2.4.

3.1.3.2.1 Boiler Island

Mitsui Babcock, utilising the Eclipse simulation package generated a heat and mass balance for the overall power plant oxy-combustion process. The model is based around the boiler island with inputs from Alstom Power for Unit 500: steam turbine island and Air Products for Units 600: ASU and 700: CO_2 compression / inerts removal. The detailed results of the Eclipse model are included in Appendix 1.

3.1.3.2.2 Steam Turbine Island

Figure 10 presents the proposed arrangement for the turbine island for the oxy-combustion PF power plant with CO_2 Capture. The diagram shows the condensate preheat and feed water heating train utilising the available heat from the ASU, CO_2 compression / inerts removal and flue gas sources in order to maximise the overall efficiency of the plant.

For air firing plant the condensate leaving the condenser would conventionally be heated utilising several feed water heaters fed with turbine bled steam, however, for the CO_2 capture plant, only a single feed heater is required for condensate preheating prior to the deaerator, as some $124.3MW_t$ of heat is sourced from the other plant units ($18.7MW_t$ from the flue gas, $55.3MW_t$ from the ASU and $50.3MW_t$ from the CO_2 plant). This integration is later discussed in detail in Section 3.1.3.7.

Following the condensate preheating the water is passed through the deaerator (operating at 6 bara) and then pumped to the required operating pressure (339 bara). The high pressure stream is then split to make use of heat from two different sources. The first stream is heated by the flue gas ($28MW_t$) and then further heated by a feed water heater using turbine bleed. The second stream bypasses the feed heater and is heated exclusively by the CO_2 compression unit (16MW_t) before being re-combined with the original stream. Two further feed heaters using turbine extracted stream, raise the temperature to the required economiser inlet temperature.

The ASC boiler elevates the temperature of the feedwater to 290 bara @ 600°C steam condition which is then delivered to the HP steam turbine. Steam is extracted from the later stages of the HP turbine to feed the last feed water heater (HP FWH 5). Upon exiting the HP turbine, a portion of steam is bled and utilised in the second to last feed water heater (HP FWH 4) with the remaining steam returned to the boiler to be reheated. Following reheat, the steam enters the IP turbine at 60 bara @ 620°C where a bleed is taken in the later stages of the turbine to feed the first stage feed water heater (HP FWH 3).

Some of the steam exiting the IP turbine en route to the LP turbine is sent to the deaerator. Within the LP turbine, steam is bled to the remaining single condensate feed heater (LP FWH 1). Finally, the vapour exiting the LP turbine is sent to the condenser (40 mbara) where seawater at 12°C provides the source of cooling that returns the stream to a condensate ready to be recirculated.

3.1.3.2.3 Air Separation Unit

The amount of oxygen required for the ASC PF boiler is 10,400 tonne/day. Currently, the largest plants in construction are 3,750 tonnes/day. The proposal for the production of oxygen in this case is to use two cryogenic ASUs of 5,200 tonnes/day. This is within the range of plant output currently being offered for sale. The single train axial flow air compressors required for this duty are available commercially.

The cycle chosen is one in which gaseous oxygen (GOX) is produced by boiling liquid oxygen (LOX) which is ideally suited to this application as the delivery pressure required is low. There is no requirement for either pumping the liquid O_2 or compressing the gaseous product.

A low purity cycle was chosen, which produces 95% oxygen purity. Other studies have been carried out to show that for oxy-combustion plants this is the optimum purity. Even new balanced-draught boiler plant are expected to have air in-leakage, and therefore there will always be some inerts that must be removed in the CO_2 purification plant. Also, the increase in power required for the ASU to produce 99.5% purity oxygen is greater than the increase in CO_2 compression power required to remove the inerts introduced due to the lower purity oxygen.

Cycle Description

To minimise the ASU power consumption because of its importance in this application, an innovative cycle has been chosen that uses two high pressure columns. A process flow diagram of the process and the mass balance is given in PFD 3. A summary of the utility requirements is given below:

Summary of Utility Requirements		
	Per Plant	Total
Power, (MWe)	43.37	86.74
Cooling Water, (tonne/hr)	417	834
Steam for ASU TSA regeneration (at 170°C), (MW_t)	3.5	7.0
Condensate Preheating, (MW_t)	27.6	55.3

The standard double column cycle has a low pressure column (C105) with its reboiler (E103) integrated with the condenser of a high pressure column (C104). The column pressures are set to give a temperature driving force in the reboiler/condenser E103. In this cycle an extra column is added operating at an intermediate pressure (C103). The condenser (E104) for this column also integrates with a reboiler in the low pressure column but at a lower temperature, boiling a liquid stream higher up within the low pressure column.

This arrangement minimises the amount of feed air that must be compressed to the higher pressure of C104, leading to the low power requirement of this process cycle.

Due to the size of this plant and the two pressure levels of compression, another feature of this cycle is that there are dryers in two locations; after compression for feed to the intermediate pressure column and after compression to the high pressure column pressure, Thus requiring smaller vessels than if only one system were used for the total flow of air.

The plant consists of:

1) A compression system

2) An adsorption front end air purification system

3) A cold box containing the separation and the heat exchanger equipment

This process offers the benefits of high reliability, low maintenance cost and is simple to install and operate.

Air Compression and Cooling

Air is taken in through an inlet filter to remove dust and particulate matter prior to entering the main air compressor (MAC) where it is compressed to 3.5 bara using an adiabatic compression arrangement, shown in Figure 11. An axial compressor is used to compress the feed air without intercooling so as to provide a higher temperature air stream to use as a source of heat for preheating condensate for the ASC PF power plant.

The air discharge is further cooled in the Direct Contact Aftercooler (DCAC) with chilled water from the Chiller Tower which uses evaporation of water into the dry waste nitrogen stream leaving the ASU cold box to further cool part of the plant cooling water. The air is cooled to a temperature of around 12°C. The main air compressor will be an in-line axial compressor driven by an electric motor. Around half of this compressed air stream is then further compressed in a single radial wheel to 5.0 bara, cooled to ambient and compressed in the compressor wheel of the coupled compressor/expander K103/K104 to 5.4 bara. The air is then cooled to 12°C in a second direct contact column.

Air Cleanup

Before the air is cooled to cryogenic temperatures, water vapour and carbon dioxide and other trace impurities such as hydrocarbons and nitrous oxide are removed in a pair of dual bed adsorbers. One pair is used to purify the 3.5 bara air stream and the other purifies the 5.4 bara stream. Removal of carbon dioxide and water avoids blockage of cryogenic equipment. The removal of impurities results in a clean, dry air stream free from contaminants which might cause blockages or safety problems in ASU operation. The adsorber operates on a staggered cycle, i.e. one vessel is adsorbing the contained impurities while the other is being reactivated by low pressure gaseous waste nitrogen using a temperature swing adsorber cycle. The nitrogen is heated to around 160°C against condensing steam in a reactivation gas heater followed by a period in which the bed is cooled down with ambient temperature nitrogen which bypasses the heater. The adsorbents used are generally selected for optimum operation at the particular site. They consist of layers of alumina or silica gel plus layers of zeolite. The adsorber vessels are vertical cylindrical units having annular adsorbent beds.

Principle of Cryogenic Air Separation

The industry standard method of cryogenic air separation consists of a double column distillation cycle comprising a high pressure (HP) column (C104) and a low pressure (LP) column (C105) as shown in PFD 3. The high pressure, higher temperature cryogenic distillation produces an overhead nitrogen product that is condensed against the low pressure, low temperature liquid oxygen in the LP column sump. The plate-fin condenser-reboiler (E103) sits in the LP column sump and thermally links the HP and LP column. The HP column nitrogen provides the boil up for the LP distillation column and the LP column oxygen provides the condensing duty for the HP column. Some of the condensed nitrogen returns to the high pressure column as reflux. The balance of the pure nitrogen reflux is cooled in the subcooler (E102) and flashed into the top of the low pressure column as reflux. The columns have aluminium structured packing optimised for cryogenic separation. In this cycle an extra column is added operating at an intermediate pressure (C103). The condenser E104 for this column also integrates with a reboiler in the low pressure column but at a lower temperature, boiling a liquid stream higher up within the low pressure column.

Cooling and Refrigeration

Following the two front end adsorber systems (C101 and C102), both the intermediate and high pressure air streams are split in two. These four streams (4, 6, 14 and 18 as shown in PFD 3) are fed directly to the main heat exchanger (E101). This consists of a number of parallel aluminium plate-fin heat exchanger blocks manifolded together.

The intermediate pressure stream 4 is cooled close to its dew point (-178°C) and fed to the bottom of the intermediate pressure column (C103). The second intermediate pressure stream 6 is removed from the main heat exchanger at -171°C then expanded in a centrifugal single wheel expansion turbine K104 running on the same shaft as a single wheel centrifugal compressor K103 which adsorbs the expander power. The expanded air is fed to the middle of the low pressure column (C105) at a pressure of about 1.4 bara and –188°C to provide refrigeration for the operation of the ASU. The high pressure stream 18 is cooled close to its dew point (-173°C) and fed to the bottom of the high pressure column (C104). The second high pressure air stream is cooled and condensed in the main heat exchanger against boiling oxygen. The resulting liquid air from the main exchanger is fed to the middle of both the high pressure and intermediate pressure columns.

Distillation System

In the high (C104) and intermediate pressure (C103) columns, the gaseous air feed is separated in the distillation packing into an overhead nitrogen vapour and an oxygen-enriched bottom liquid. The nitrogen vapour from the high pressure column is condensed against boiling oxygen in the low pressure column sump and split into two parts. The first part is returned to the high pressure column as reflux, whilst the second part is subcooled, reduced in pressure and fed to the low pressure column (C105) as reflux. The nitrogen from the intermediate pressure column (C103) is condensed against a boiling liquid stream in the low pressure column. Part of this nitrogen is used as column reflux in the intermediate pressure column and part is subcooled and added to the reflux to the low pressure column.

Crude liquid oxygen is withdrawn from the sumps of the high and intermediate pressure columns, cooled in the subcooler (E102) against warming waste nitrogen and is flashed to the low pressure column as intermediate feeds. A portion of liquid air is also withdrawn from the middle of the high pressure column. This liquid is subcooled in the subcooler and fed to the middle of the low pressure column.

Low Pressure Column

The feeds to the low pressure column are separated into a waste nitrogen overhead vapour and a liquid oxygen bottom product, which reaches the required purity of 95% by volume. At present the nitrogen is vented to atmosphere, however, there is potential to utilise this warm dry nitrogen stream within the coal drying process.

The waste nitrogen is withdrawn from the top of the low pressure column and warmed in the subcooler and the main heat exchanger. A portion of the nitrogen stream from the main exchanger is used for adsorber reactivation. The remaining dry nitrogen is vented through a Chilled Water Tower to produce chilled water by evaporative cooling. The chilled water is used to provide additional feed air cooling in the top section of the DCACs.

Pure liquid oxygen is withdrawn from the reboiler sump of the low pressure column and is returned to the main heat exchanger where it is vaporised and warmed up to ambient conditions against boosted air feed to the columns. The gaseous O_2 is then regulated and supplied to the power plant. The pressure in the low pressure column is typically 1.35 bara. The hydrostatic head between the sump of the LP Column and the LOX boil heat exchanger results in the O_2 product being available at approximately 0.6 barg.

Oxygen Backup

The PF boilers will be designed in such a way as to allow air-firing as a fall-back position should there be an interruption in supply from the ASUs. Therefore, adequate backup for the ASUs should be provided in order to allow a controlled change-over to air-firing.

Backup will be in the form of liquid oxygen (LOX) enough of which will be stored on site to allow controlled changeover to air-firing. A PFD for this backup system is shown in PFD 4. An isometric of the oxygen backup system is shown in Figure 12. The LOX will be held at a pressure of 2.5 bara in a 200 tonne capacity vacuum insulated storage tank which can be filled by gravity from the ASU. If backup oxygen is required from storage, detected by a pressure controller on the GOX header, the control valves will open to allow LOX to enter the vaporiser. Because of the short time lag in the system to initiate the GOX backup flow through the vaporiser, a temporary means of providing GOX is required. The GOX pressure is maintained in the system using a GOX buffer vessel kept at 30 bara pressure, which discharges into the GOX header under pressure control.

Air Separation Equipment

Multiple structural steel cold boxes and one column can are supplied as part of the equipment. The column can is a cylindrical enclosure of pre-formed/pre-rolled flanged sections which bolt together at site to complete the structure. Steel jacket panels can be welded or bolted for equipment access to the framework. The cold boxes and column vessel are inclusive of process equipment. The process equipment is supplied and constructed of material suitable for use at low temperature.

The column can encloses the high, intermediate and low pressure columns. The reboiler and condensers are contained in the low pressure column. All heat exchangers in the cold air separation equipment are multi-passage, extended surface aluminium / aluminium alloy, plate-fin heat exchangers.

The main heat exchanger and subcooler are prefabricated. The main heat exchanger (MHE) box houses the main heat exchanger and the expander units. The subcooler box contains a multi-passage, extended surface plate fin heat exchanger.

The primary insulation material is expanded pearlite. Certain areas are packed with rockwool to allow access for maintenance of valves without pearlite removal.

A dry nitrogen purge system is included on all cold boxes and cans to prevent moist atmospheric air from leaking into the cold box/can during normal operation.

3.1.3.2.4 CO_2 Treatment

The net flue gas from the 740 MWe gross ASC PF oxy-combustion coal-fired boiler must be cooled, dried, compressed, and purified to the required level. A summary of the performance of this system is shown below for one boiler unit:

Performance Summary for CO_2 Treatment System : 95mol% CO_2 Purity		
Flue Gas Heater	(from 170 to 300°C) (MWt)	5.0
Net Compressor/Expander/Pump Power (MWe)		64.7
Cooling Water (tonne/hr)		1,635
Seawater (tonne/hr)		10,711
Condensate Preheating (MWt)		50.3
Boiler Feedwater Preheating (MWt)		16.0
CO_2 Captured:	Purity (%v/v)	95.8
	Contained CO_2 (tonne/hr)	441.9
	Recovery (%)	90.4

The CO_2 treatment plant consists of:

- A venturi scrubber;V201
- An indirect contact cooler; C204
- A direct contact seawater cooler; C205
- A compression system; K205, K204, K202,K201
- BFW and Condensate preheating exchangers; E206 and E207
- A drier system
- A cold box containing CO_2 purification equipment

The process is shown in the PFD 5 and PFD 6.

The CO_2-rich flue gas leaves the heat recovery system of the ASC PF oxy-combustion power plant at approximately 110°C. The first part of the CO_2 treatment system, PFD 5, cools the flue gas, thus removing the moisture by condensation, and compresses it to 30 bara.

A venturi mixer V201 is used to quench the gas with water to a temperature where a conventional indirect seawater contact cooler can be used with standard plastic packing. A two column system is used. The first column C204 cools all of the flue gas to 35°C by direct contact with condensate that has been cooled against seawater in titanium plate-frame heat exchangers E205. Around half of this flue gas is then recycled to the boiler system as primary recycle gas, stream 4. The rest, stream 5, is further cooled against seawater to 13°C (stream 14) in a second direct contact cooler C205.

The net flue gas is now around 75% by volume CO_2 and at atmospheric pressure and should be compressed to 30 bara for further drying before purification. Compression to 30 bara is carried out in two stages. First K205 compresses the CO_2 adiabatically to 15bara. This compressor has 8 axial stages followed by 3 radial stages (see Figure 13 for compressor train layout details from MAN Turbocompressors). The heat of compression is then used to preheat 151 tonne/hr of boiler feedwater in E206 and 330 tonne/hr of condensate in E207. These two heat exchangers are stainless steel diffusion bonded compact heat exchangers. Cooling water is used in E208 to further cool the raw CO_2 to 20°C before compression in a single radial stage K204 of the integrally geared CO_2 compressor shown in Figure 13 to 30 bara with aftercooling in E209 using cooling water.

The process then moves on to PFD 6 where the raw CO_2 is dried and the inerts (N_2 and Ar) and oxygen are separated to give >95 mol% CO_2. The CO_2 is then compressed to 110 bara for pipeline transmission. Any excess O_2, SO_2 or NO_X present in the CO_2 need not be removed as the final CO_2 product will be used either for enhanced oil recovery (EOR) or stored in aquifers. For specific EOR applications requiring higher purity CO_2 >98% the process flowsheet can be modified to achieve this requirement as discussed in Section 3.1.3.5.

The raw CO_2 gas passes through a temperature swing dual bed desiccant dryer (C201) to reach a dew point of below -55°C before entering the “cold box”. This desiccant dryer system prevents ice formation which could cause a blockage in the cold box as well as causing corrosion in the pipeline. The cold equipment is contained in a steel jacketed container or “cold box” with pearlite granular insulation.

The inerts removal process uses the principle of phase separation between condensed liquid CO_2 and insoluble inerts gas at a temperature of –55°C which is very close to the triple point, or freezing temperature, of CO_2. The CO_2 partial pressure can be reduced to about 7 bara. The actual CO_2 pressure levels used for the separation are fixed by the specification of >95 mol% CO_2 product purity and the need to reduce the CO_2 vented with the inerts to an economic minimum.

The system proposed uses two flash separators C202 and C203 at temperatures of 25°C and -55°C. The CO_2 feed gas pressure is at 30 bara. The necessary refrigeration for plant operation is obtained by evaporating liquid CO_2 at pressure levels of 18.6 bara (stream 20) an 9.3 bara (stream 16) and compressing these two low pressure gas streams in the main CO_2 product compressor to the final pipeline delivery pressure of 110 bara. The separated inert gas leaving the cold box at 29 bara (stream 7) can be heated and passed through a power recovery turbine. It is possible to reach a CO_2 purity in excess of 96% using this method at inlet CO_2 concentrations as low as 77% by volume with a CO_2 recovery of better than 90%.

The dry gas is fed to the cold box and is cooled by heat exchange to –25°C with the returning evaporating and superheating CO_2 streams and the waste streams in the main exchanger. The main heat exchangers, E201 and E202, are multi-stream plate-fin aluminium blocks. The cooled feed stream 3 is sent to a separator pot C202 at a temperature of –25°C where it is split into liquid and vapour; the liquid product, stream 18, contains part of the required CO_2 product at 29.7 bara.

The vapour from the separator, stream 4, still contains a large proportion of CO_2. In order to recover this CO_2 the vapour is cooled further to –54°C where it partially condenses and is passed to another separator pot C203. The pressure at this point is critical in controlling the process since cooling the vapour below –56.2°C would lead to the formation of solid carbon dioxide. The vapour, stream 6, from the second separator, containing the separated inerts together with some CO_2 at a partial pressure of about 7 bara, is sent back through the heat exchangers E202 and E201 where it is heated to 8°C. This stream of inerts, which is at a pressure of 29 bara, is then heated against hot compressed CO_2 product (E210) and hot flue gas in the boiler area (E203) and expanded in a power producing turbo-expander (K203) before being vented.

Liquid, stream 18, from the first separator C202 containing part of the CO_2 is expanded through a J-T valve to 18.8 bara (stream 19) and heated to 8°C (stream 20). The liquid, stream 12, from the second separator C203 , is heated, expanded through a valve to 9.7 bara and a temperature of about –55°C (stream 13) to provide refrigeration in E202 by evaporation, while the vapour formed is heated to 8°C. The CO_2 vapour stream leaving E202 at 9.5 bara is then compressed in a single radial wheel (K202) to 18.7 bara, the same pressure as the CO_2 stream from the first separator C202. The two streams are combined and compressed to the required pressure of 110 bara. This machine (K201) is a four stage integrally geared unit (Figure 13) which could be operated from the 18.7 bara to 110 bara level as either an intercooled compressor or as an adiabatic compressor with an aftercooler used to heat flue gas before expansion and condensate for the boiler system. In the latter case no cooling water would be required for this section of the compressor.

This project has selected K201 to be run adiabatically, with condensate being preheated in the aftercooler along with some of the flue gas heating duty. This has the benefit of simplifying the final stages of K201 since it avoids supercritical dense fluid CO_2 forming in K201. The likelihood of dense fluid CO_2 forming in K201 has meant that the four stage isothermal option only had one intercooler, to prevent the dense phase forming within the machine. Therefore the power penalty in removing this intercooler to give an adiabatic compressor is small but gives the benefit of a simpler machine, reduced cooling water requirement and saves low pressure steam that would have otherwise been used to preheat the condensate.

The plant is furnished complete with all structural, mechanical equipment, piping, supports, anchor bolts, electrical equipment, instrumentation, controls and accessories as required for continuous automatic operation. The controls are designed to interface with the existing systems. The intent would be to operate the plant from a remote central control room with periodic inspection.

The material of construction of the low pressure inlet gas piping, the venturi cooler, direct contact coolers and the parts of the compressor upstream of the drier must be resistant to wet gas corrosion taking account of the possible gas composition. Suitable materials are selected for piping and compressor parts in contact with the flue gas.

3.1.3.3 ASC PF Oxy-Combustion Power Plant With CO_2 Capture: Performance

ASC PF oxy-combustion power plant with CO_2 capture generates 737 MW_e gross (532 MW_e net) power output at a net cycle efficiency of 35.4% (LHV).

The penalty therefore in terms of net cycle efficiency associated with the oxy-combustion with CO_2 capture process is calculated at 8.9 % points when compared to the efficiency of reference power plant without CO_2 capture of 44.2% LHV.

The auxiliary power consumption for the ASC PF oxy-combustion power plant with CO_2 capture is shown in Table 3. The draught plant power consumption has reduced as a direct consequence of the lower mass of flue gas being circulated compared to the air firing case. The power consumption of the ASU and CO_2 compression & inerts removal units are as anticipated by Air Products.

The effect of removing the nitrogen when operating an oxy-combustion boiler is to increase the partial pressure of the CO_2 and H_2O which in turn raises the emissivity of the flue gas and improves the radiative heat transfer within the furnace. For the same fuel firing rate, an oxy-combustion boiler with a smaller sized furnace can therefore achieve similar steam conditions as its air-firing counterpart. The reduction in cost associated with the smaller furnace, however, has not been considered within this study as detail design of the furnace (i.e. furnace heat surface sizing) has not been undertaken. This cost saving is somewhat minor compared to cost of the overall oxy-combustion power plant and considering the target budget capital cost estimate of ± 30% stated as the basis of this study.

3.1.3.4 Emissions

In comparison with the air-fired PF plant base case, NO_x and SO_x emissions for the CO_2 capture plant are proposed to be uncontrolled (i.e. no SCR and no FGD plant).

Experimental testwork[29] has shown oxy-combustion will inherently result in a reduction in NO_x emissions to levels comparable with SCR technology (i.e. 200 mg/Nm^3 @ 6% O_2).

All soluble acid gas components in the primary recycle and net CO_2 product gas streams are removed by direct contact water scrubbing in C204 and C205. There is no tendency for any condensation of acid components or water in the coal mills. No FGD is required for an oxy-combustion PF coal fired boiler. SO_2 and NO_x are presently included in the CO_2 product stream and have no diverse effect on hardware as long as sufficient dehydration is present to avoid acid corrosion. If the CO_2 specification calls for very low levels of SO_2 and NO_x then a distillation unit can be included in the CO_2 purification system which will remove all SO_2 plus NO_2 as a concentrated liquid at 30 bara pressure. This will result in a small cost increase.

NO_x Emissions

The level of NO_x present in the flue gas exiting the boiler (115 mg/MJ) in the oxy combustion PF case is approximately 50% of the NO_x emissions expected for similar sized air fired PF Plant. (650 mg/Nm^3 @ 6% O_2 or 227mg/MJ). This assumption is based on published literature[29] from previous combustion tests of oxy-combustion with FGR.

Figure 14 shows the NO_x levels in each of the major streams (note: these emissions presented are at actual oxygen levels). The primary and the secondary recycles return 32 mg/MJ and 49 mg/MJ respectively back to the furnace. Approximately 26 mg/MJ is removed via the CO_2 compression and inerts removal process leaving 6 mg/MJ contained in the high pressure CO_2 product stream for storage. Note the only emission of NO_x to atmosphere is therefore 26 mg/MJ resulting from the vented inert gases. There is an option to remove all NO_x in the low temperature CO_2 purification plant by distillation. This would result in a NO_2 component stream which would require further disposal. The associated disposal costs have not been considered under this study.

SO_X Emissions

Figure 14 shows the SO_x levels in each of the major streams (note: these emissions presented are at actual oxygen levels). The level of SO_x present in the flue gas exiting the boiler in the oxy-combustion case is calculated at 2046 mg/MJ. The primary and the secondary recycles, return 559 mg/MJ and 872 mg/MJ respectively back to the furnace. Some 80 mg/MJ is believed to be removed in the condensate as a result of the contact cooling process prior to CO_2 compression, leaving 534 mg/MJ remaining in the high pressure CO_2 product stream for storage. There is an option to remove all SO_2 in the low temperature CO_2 purification plant by distillation. This would result in a SO_2 component stream which would require further disposal. The associated disposal costs have not been considered under this study.

Particulates

Figure 14 shows the particulate levels in each of the major streams (note: these emissions presented are at actual oxygen levels). The level of particulates present in the flue gas exiting the boiler in the oxy combustion case is calculated at 11 mg/MJ. The secondary recycle returns 5 mg/MJ to the furnace with the remaining particulates (5mg/MJ) removed upstream of the recirculation pump from the direct contact cooler C204.

Mercury Removal

If mercury is present in the coal it will appear as a contaminant in the feed to the CO_2 compression and purification system. Mercury is a potential source of corrosion in the aluminium heat exchangers used in the CO_2 purification plants. It can be removed from the 30 bar compressed cycle CO_2 stream by adsorption and charcoal impregnated with sulphur. This technology was developed to remove mercury from natural gas streams prior to liquefaction in aluminium heat exchangers.

The Effect on Impurity Concentrations in the Flue Gas from Varying Recycle Flow

The concentration of SO_2, SO_3 and HCl in the net flue gas product and recycle flue gas streams depends only on the sulphur content and the chlorine content of the coal feed respectively and is independent of the recycle rate back to the boiler which is adjusted to control the combustion temperature.

3.1.3.5 Clean Up Technologies for CO_2 Purity

The base case of this study produces CO_2 at > 95 mol% using a simple process that gives the required purity at high (> 90%) recovery. This process is easily modified to give higher purities.

As it stands the basic process cycle produces a CO_2 purity of 96.5 mol%. To produce 98 mol% purity CO_2 a modification is required to the process. This is achieved by the addition of a third separation vessel to remove the non-condensed portion of stream 14 in PFD 6, shown in bold in PFD 7. To prevent a loss of CO_2 from the process, since there will be non-condensed CO_2 in this stream, this stream is compressed from around 10 bara to 30 bara, in a two-stage intercooled compressor and recycled by adding to the cold box feed after the dryers.

Pure (>99.99mol%) CO_2 is also possible using distillation in place of separation in this flowsheet to remove a stream of higher boiling components comprising all of the nitrogen oxides as NO_2, and SO_2 plus some CO_2 as a liquid stream from the bottom of a distillation column which replaces C202. A second column in place of C203 at the cold end removes traces of inert gases including O_2.

3.1.3.6 Performance and Cost Sensitivity to CO_2 Purity

The performance of the CO_2 treatment plant with 98 mol% purity CO_2 is shown below. Compared to the 95 mol% purity case 0.5 MW_e less power is required but the flue gas heater is larger by 1.9MW to accommodate the extra flow of vented inerts. A small amount of extra cooling water is also required. There is a reduction in CO_2 capture efficiency of the process, down from 90.4% to 88.4%, resulting in 2.3% less CO_2 being captured.

Capital costs will be slightly increased due to the recycle compressor, extra separation vessel and the extra passages through the exchangers, but this will be minimal. The main penalties for the increased CO_2 purity are a reduction in CO_2 captured from the process and an increase in the amounts of NO_2 and SO_2 vented.

Performance Summary for the CO_2 Treatment System for 98 mol% CO_2 Purity			
Flue Gas Heater	(from 170 to 300°C) (MWt)		6.9
Net Compressor/Expander/Pump Power (MWe)			64.2
Cooling Water (tonne/hr)			1,770
Seawater (tonne/hr)			10,771
Condensate Preheating (MWt)			50.2
Boiler Feedwater Preheating (MWt)			16.0
CO_2 Captured	Purity	(% v/v)	98.3
	Contained CO_2	(tonne/hr)	431.9
	Recovery	(%)	88.4

Although not presented the cost and performance penalties of pure CO_2 will be higher than those for 98 mol% purity CO_2.

3.1.3.7 Overview of Integration Activities

Oxy-combustion PF power plants offer potential opportunities for efficiency improvements through integration of the additional oxy-combustion process units with the steam cycle:

Option A: Heat recovery from additional flue gas cooling

Option B: Heat recovery from adiabatic compression in the ASU and CO_2 treatment plants

Option C: Direct mechanical drives (i.e. small steam turbines) for compressors

In this study options A and B have been implemented.

Option C was considered but rejected. The isentropic efficiency of the large main sream turbine for power generation was much higher than could be achieved in smaller turbines (5-10 percentage points) and the latter would also have greatly complicated the steam cycle and start-up procedures. Consequently all compressor drives are electric.

The relevant areas of the power plant where heat integration takes place are listed in the table below:

Summary of Heat Integration Sources		
Heat Source	**Amount (MWt)**	**Heat Transferred To**
ASU Adiabatic Compressors	55	Condensate
CO_2 Adiabatic Compression (Early & later stages)	50	Condensate
CO_2 Adiabatic Compression (Early stages)	16	BFW
Flue Gas	28	BFW
Flue Gas	18	Condensate
Flue Gas	5	Vented Inert Gases

Heat Recovery From Additional Flue Gas Cooling

Heat recovered from cooling flue gases after the ESP can be used for condensate/feed water heating and avoids the need to extract steam from the turbine for this purpose. With the acid dew point approximately 160 °C and the flue gas temperature exiting heat exchangers approximately 110°C, acid condensation will occur in these coolers and appropriate construction materials must be used. Proven designs are available for similar duties in modern conventional PF plants to cool flue gases before the FGD unit.

It is obviously advantageous to use the heat recovered at as high a temperature as possible so multiple (two in this study) stages of water heating are likely to be used. In practice, economics will determine the degree of heat recovery from the net product flue gas.

Heat Recovery From Compressors

Adiabatic compression has been used in the oxy-combustion process flowsheet place of isothermal compression in several places. The heat of compression is then used to preheat either condensate or boiler feedwater (see Section 3.1.3.2.2) depending upon the temperature of the compressed gas. This eliminates the requirement for intercoolers and aftercoolers and reduces the overall cooling water requirement. The extra power that is required for the compressors is more than compensated by the saving in bleed steam that would have been used to preheat the same condensate and boiler feedwater for the boiler system, thus allowing this steam to produce more power in the steam turbines. The design of the compressors is also simplified by using adiabatic compression since no intercoolers are required, removing equipment, pressure drop and complex piping and resulting in less plot area and lower construction costs.

In this work, the main compressors for the two ASUs are adiabatic, as is the first (K205) section of CO_2 compression, to 15 bara, and the final stages of CO_2 compression (K201) to 110 bara.

In order to produce power from the high pressure inert stream and to maintain an outlet temperature in which the vent gases are buoyant, the inerts, at around 30 bara, from the CO_2 purification system are heated to around 300°C before being expanded to atmospheric pressure. Part of this heat is provided by cooling part of the hot CO_2 at 110 bara, leaving the adiabatic compressor K201. The remaining heat, from 170°C to 300°C, could be provided by heat interchange of a suitable high temperature stream within the boiler system.

3.1.3.8 Materials and Corrosion Issues

In this section a short description of the principal technical concerns associated with the selection of the materials of construction of key plant components is presented. This description is restricted to the identification of the key technical issues, and no attempt is made to provide appropriate detailed design and materials solutions.

The Coal Milling and Firing System

It is proposed that the medium for milling the coal and conveying the pulverised coal to the burners will be recycled flue gases, from the primary recycle system. This stream has a significant SO_2 concentration, and is saturated with water at 35ºC, but very low SO_3 and HCl concentrations, since the HCl and SO_3 have been removed to high efficiency in the gas cooler and water removal system in the primary recycle.

The Furnace and Boiler Convective Section

The principal concern here is the increased acid gas (SO_2, SO_3 and HCl) concentrations in the combustion gases, and the potential impacts on the metal wastage rates due to gas-side corrosion of the boiler tubes, particularly in the furnace and the superheater/reheater sections of the boiler. The calculated gas composition data indicate that could be as much as a five-fold increase in these concentrations over the normal levels associated with conventional pulverised coal combustion systems.

In the furnace, the principal concern is associated with the increased risks of accelerated metal wastage of sidewall tubes, associated with advanced primary NO_x control techniques, and exacerbated by the increased concentrations of HCl and the sulphur oxides in the furnace gases.

The main concerns in the superheater/reheater section are associated with the risks of increased gas-side corrosion rates due to the increased acid gas concentrations.

These issues will have to be examined carefully in that long-term plant operating experience and measured corrosion rate data for very high acid gas concentrations in boiler flue gases are relatively rare. Careful consideration of the metal surface temperatures and the materials will be required.

The Gas-Gas Heater

The gas-gas heater is located upstream of the electrostatic precipitators and operates under the following conditions:

Flue gas inlet and outlet temperatures	340-264ºC
Primary recycle inlet temperature	35ºC
Secondary recycle inlet temperature	264ºC

It is inevitable, therefore, that the gas-gas heater at the primary recycle inlet will be operating at temperatures below the acid dew point, and perhaps the water dew point, on the flue gas side, and that the condensation of acidic aqueous species on the heat exchanger surfaces will occur. For this reason, the gas-gas heat exchanger will need very careful design, particularly with regard to the materials of construction, and to avoid excessive coal ash fouling at the flue gas entry.

The ID fan, Gas Recycle Fans and Ductwork

The ID fan, the secondary recycle fan and the associated ductwork and windboxes are located downstream of the gas-gas heater and the electrostatic precipitators. These gases have relatively high SO_2/SO_3 and HCl concentrations, but are operating at temperatures around 264ºC and, as such, there are no dew point issues or high temperature corrosion issues.

The primary recycle fan and associated ductwork will operate with flue gas from the gas cooler and water removal system. This gas stream is at 35ºC, has a significant SO_2 concentration, and is saturated with water. The HCl and SO_3 have been removed to high efficiency. The gases are at or close to the water dew point, however, and the design and materials selection in this section will need careful consideration.

Feed Water Pre-Heaters and the Gas Cooler and Water Removal System

The primary gas recycle system has a series of feed water heaters designed to reduce the gas temperatures from 264 to 110ºC prior to entry to the gas cooler and water removal system. The feedwater heaters will be operating at temperatures below the sulphuric acid dew point temperature for the recycled gas, and will require very careful design and materials selection.

The gas cooler and water removal vessel and associated systems will reduce the gas temperatures from 100 to 35ºC, with the condensation of water and removal of both SO_3 and HCl from the gases. The liquors contained in the vessel and the water recycle system will be highly acidic and extremely corrosive, and the system will require very careful design and materials selection.

Under normal operating conditions, the mill outlet temperature will be significantly in excess of the water dew point, and no water condensation on the mill internal surfaces or the pulverised coal pipework is anticipated. It may be necessary to purge the mills and pulverised coal pipework with fresh air if the mill is to be shut down for a prolonged period of time, otherwise the primary recycle stream appears to be a suitable coal milling and firing medium. There are no specific materials issues.

Low-Temperature Corrosion

Low temperature corrosion occurs when temperature conditions favour the condensation of sulphuric acid. The sulphur present in pulverised coal will oxidise to SO_2 and SO_3, and combining with the water vapour, will form sulphurous and sulphuric acids. The temperature of the condensation (the acid dew point) is important as surfaces below this temperature will condense acid: the lower the temperature of the surface the greater the corrosion.

If such a condition arises it is necessary to protect the exposed surfaces using materials that offer exceptional resistance to sulphuric acid corrosion, for example high silicon cast-iron inserts, polypropylene and glass reinforced plastics.

Under oxygen-combustion conditions, the dew point increases to approximately 160°C and thus corrosion in both the later stage of heat exchangers and the cooler parts that follow, such as ducting and draught plant, become a distinct possibility. In addition, with oxy-combustion the problem is aggravated by the fact that recycle streams themselves require to be heated whilst containing some oxidised sulphur compounds.

For the ASC PF oxy-combustion boiler case with CO_2 capture, the temperature in the early stages of the economiser is not a concern as the high inlet temperature (~300°C) of the feedwater at this stage is well above the acid dew point of the flue gas. Likewise, secondary recycle, whilst containing SO_x, only reduces to temperatures of approximately 260°C on its course back to the furnace, therefore the acid dew point is not an issue. The ESP is also operated at temperatures (260°C) above the acid dew point.

However, the heat exchangers positioned in the flue gas stream, which are utilised for boiler feed water heating and condensate preheating, feature inlet water-side temperatures below the acid dew point and would therefore require to be protected against corrosion. The associated ductwork in this area would also be required to be lined to prevent acid corrosion damage.

Following these exchangers the flue gas is passed through the venturi water contactor and the two direct contact coolers which condense the water content of the flue gas. The water saturated stream leaving the first direct contact cooler at 35°C is split with one stream forming the primary recycle and the other the feed to the second direct contact cooler which further cools the net flue gas to 13°C before entering the CO_2 compression and inerts removal stage. The primary recycle stream contains no soluble acid components as a result of the direct water cooling removing any SO_3 and HCl present.

CO_2 Compressor Materials

Compressor material selection for the wet CO_2 compression needs careful consideration due to the possibility of wet SO_2 being present. Previous studies have suggested doubling the nickel content in 316 stainless steel to 904 austenitic stainless steel to combat this problem. An appropriate material specification would be:

Alloy 20Cb-3 (UNS No: NO8020) – 20Cr2.2Mo34Ni3.5Cu austenitic stainless steel for impact areas or cold areas such as volutes, impeller, intercoolers and internals
Alloy 2205 – 22Cr5Ni3Mo duplex stainless steel for shafting.

Fouling and Deposition

The recycled flue gas from oxy-combustion of pulverised coal will contain particulate material (predominantly ash with a small amount of unburnt carbon). Removal of this dust burden upstream of the FGR fans will be necessary to avoid premature wear/erosion of the impellers.

However, complete removal of the particulate burden is unlikely, therefore there is risk of increased dust/ash deposition within the boiler flue gas passes, particularly within the convective pass.

Rig scale tests[2] indicated that sootblowing frequency may have to be increased in order to prevent detrimental deposition.

In addition these tests have indicated that deposition rates during flue gas recycle are greater than conventional air firing, however, quantitative comparison was not possible. High flue gas recycle rates were not observed to exacerbate deposition when compared to low recycle rates. Further investigation is recognised as being required in this area.

3.2 Safety and Operability

It is of paramount importance that oxy-combustion technology can be implemented safely and without significant impact upon operational flexibility.

The following sections highlight the key safety and operability concerns relating to oxy-combustion, with specific regard to pulverised coal utility plant.

3.2.1 Control Strategy

For the process flow sheets developed under this work the boiler will be started on air with the FD fan drawing air into the process and a PA fan stripping a portion of this air for delivery to the mills. An ID fan positioned after the ESP will draw the combustion products through the boiler to the stack as shown in Figure 9.

Power produced by the steam turbine-generator whilst operating on air can be used to start the ASU. The boiler will then be converted to oxy-combustion firing by progressively increasing the oxygen flow to the burners and reducing the air flow while simultaneously introducing the hot flue gas recycle and increasing the recycle rate to maintain acceptable combustion temperatures. During this period, the net flue gas production will be vented through the main stack. It, however, will be necessary to cool a part of the flue gas and separate the water vapour and this can be accomplished by commissioning the water venturi quench and the direct contact cooling tower. Once the boiler is operating with no air feed, and the monitored flue gas CO_2/N_2 levels at the ID fan exit are within acceptable limits, then the full net flue gas flow can be directed to the cooling system and the CO_2 compressors and low temperature CO_2 purifier plant can be started.

Returning the plant to air-firing requires a reversal of the above procedure, i.e. divert the majority of the flue gas to the flue stack whilst maintaining a primary recycle to the coal mills via the direct contact cooler. The compression plant will then be taken off line. The oxygen supply to the furnace will then be reduced whilst reintroducing air rather than the recycled flue gas, through the FD fan. Finally, the switch to an air supply for the mills will be made with the primary recycle reduced accordingly.

It is initially envisaged that the boiler will operate on low loads only up to the point of minimum stable combustion and no higher whilst firing on air. Therefore in this study, for a new build oxy-combustion plant, air firing mode is perceived as a means to achieving oxy-combustion. The plant is not intended to operate for any persistent length of time on air firing other than that necessary to permit changeover to and from oxy-combustion mode of operation.

Note: for the short period of time envisaged in which the air-firing is utilised, no means of NO_x and SO_x emission abatement are present, i.e. SCR and FGD plants exclusively for the start-up and shut down scenarios are not costed within this study. At the point of power generation, the plant will operate in oxy-combustion mode in order to provide compliance with emission limits.

3.2.1.1 Economiser Exit Flue Gas Oxygen

For air-firing plant, the control of combustion air is based on a function of the measured fuel flow, the measured air flow at the FD fan and the measured economiser exit flue gas oxygen content.

For oxy-combustion firing, a new control loop would be required to control the oxygen flow, the fuel flow and the recycle flue gas flow to the burners. This control loop would measure the oxygen concentration downstream of the oxygen injection point and drive the oxygen injection flow demand for that stream, as a function of this, the FGR flow and the economiser exit flue gas oxygen. This loop may also require additional feed forward terms relating to load depending on the response time of the ASU plant.

Given the oxygen concentration is maintained in the combined O_2 and CO_2 recycle stream at the windbox by the above control loop during oxygen/flue gas recycle firing, then the economiser exit flue gas oxygen control loop can now be a similar function to that used for air-firing, but driving the FGR flow demands instead of the air flow demands.

3.2.1.2 Furnace Draught Pressure

The furnace draught pressure can remain as a control function of the ID fan speed.

3.2.1.3 ID Fan Exit Process Control

The ID fan exit ducts will require control dampers to divert the flue gas flow from the flue stack inlet to the flue gas cooler and CO_2 compression plant. In addition flue gas monitoring equipment will indicate to the boiler operator when it is safe to start supplying the CO_2 plant with boiler exit flue gases.

If the CO_2 compressor flow demand creates a negative pressure local to the control dampers at the flue stack, then this will tend to induce a back flow of leakage ambient air from the flue stack to the compressor inlet which will contaminate the flue gas flow with N_2 and O_2. The ID fan exit pressure should therefore be maintained at slight positive gauge pressure during oxy-combustion firing. Depending on the rundown time of the compressor when tripped, the stack inlet / compressor inlet diverting dampers may need to act quickly.

Whilst guillotine dampers are the power industry norm for gas tight applications, the use of a water sealing mechanism for the dampers may be required to provide the level of sealing necessary to minimise any losses to stack or further in-leakage to the compression system.

3.2.1.4 Alarms

The following additional alarms are envisaged for the ASC PF oxy-combustion power plant when operating in oxygen-combustion mode:

- ASU O_2 Outlet Temperature Low - To detect a failure of the O_2 heating supply from the ASU.

- Combined O_2 and CO_2 recycle stream Temperature High - To detect the possibility of combustion taking place before the burner, between some of the added O_2 and any unburned combustibles in the FGR flow.

- Superheater Metal Temperature High - To detect high metal temperatures due to burner/flame parameters changing due to oxygen/flue gas recycle combustion.

- ID Fan Exit Pressure Low - To detect if the duct pressure becomes negative enough for there to be a danger of inducing a reverse flow in the flue stack and drawing air into the CO_2 compression stream through the damper.

3.2.1.5 Trips

Trips envisaged for oxy-combustion firing would include but not be limited to the following:

Fan Trips	For oxy-combustion firing a 'Windbox Gas Flow Low' trip is required to replace the 'FD Fan Tripped' on the air-firing case. This allows the draft plant BMS trips to remain active during the changeover from air firing to oxy-combustion firing.
ASU Trip	Assuming there is sufficient O_2 storage to maintain the plant while air-firing is brought on line, then changeover to the air-firing would keep the boiler on line. This would decrease the sensitivity of the plant in the unlikely event of an ASU trip (ASU availability will be 98-99% and this includes planned shutdowns which could be scheduled with planned boiler shutdowns and to the effect of the ASU on the overall availability will be very low). If the ASU tripped a trip signal would be sent to the boiler controller to initiate changeover to air-firing.
Boiler Trip	Any 'Boiler Trip' would require the ASU feed to be tripped and the ID exit flue gases to be directed to the flue stack.
Compressor Trip	When oxy-combustion firing, a 'Compressor Trip' will require that the ID fan exit flue gases be diverted to the flue stack. If this can be done within the rundown time of the compressor, the boiler need not be tripped. If this is not the case, then the back pressure through the flue gas pass will lift the furnace pressure and trip the boiler. If it is deemed unlikely that the diversion to the stack can take place quickly enough to avoid a boiler trip, then the plant should be tripped directly from a 'Compressor Trip'.

3.2.1.6 Purging

The presence of FGR flow loops necessitates that these flue gas passes are adequately purged prior to the introduction of fuel to the furnace.

Flow will be required to be generated through the FGR ducts as part of the flue gas pass purge. Some extension of furnace purge time is required to take account of the re-introduction of furnace exit purge flue gases being re-introduced to the furnace by the FGR fans. In addition, since it is possible for combustibles to enter the compressor flue gas pass (under ‘Boiler Trip’ conditions, or economiser exit flue gas oxygen control loop failure), then this flue gas pass will require to be purged in addition to the flue stack duct, prior to trial for ignition.

In this context the oxygen system will require to be proven to be isolated from the boiler for purging purposes.

3.2.2 ASU Safety Issues

Safety is a major factor in the design and operating strategy of ASUs[30A]:

- Rapid oxidation (which falls into two categories: Accumulating fuel in O_2 enriched streams, and O_2 enriched streams reacting with normally non-combustible materials).
- Interfaces between the ASU and downstream equipment, with the risk for sending high pressures and cold temperatures that are incompatible with the downstream equipment.
- Building high pressures due to vaporising cryogenic liquids.
- Oxygen enriched and deficient atmospheres.

There is a strong commitment to safety from the ASU equipment manufacturers based on nearly 100 years of operating experience as the air separation industry has developed. Safety standards are the responsibility of the industry as a whole[30B] and are a result of the cooperation between companies on a continuing basis. Notable areas of activity in recent years have included:

- Standards for materials compatibility with oxygen, covering flammability and material properties
- Design standards for oxygen compressor systems, both centrifugal and reciprocating
- Considerations for the design of reboiler/condenser systems

These specific safety considerations are backed by procedures used in the design, construction, operation and maintenance of air separation equipment.

The following specific items form an integral part of the design of the plant to ensure safe operation:

- The design allows for the elimination of potential hydrocarbon build-up due to the location of the fuel burning power generation facility.
- The oxygen plant is designed for fail-safe emergency shutdown as a result of internal or external upset. All process control power supplies are connected to an uninterruptible power supply, that will provide a power back-up for 30 minutes.
- The design of oxygen injection system into the fuel burners which would include direct oxygen injection into the burner and mixing of oxygen with recirculating hot flue gas.
- The location of vents and drains to avoid discharge of oxygen deficient or oxygen rich gas or liquid streams that might be hazardous to the surroundings.

The plant and equipment is designed in accordance with recognised national/international codes and standards appropriate to its location and its point of manufacture.

3.2.3 ASU Plant Process Control

3.2.3.1 Control System Design Philosophy

The control system is designed to meet the following overall objectives:

- To provide a safe system.
- To meet the plant reliability and availability targets.
- To enable the plant to run routinely within the particular operation constraints.

3.2.3.2 Control Strategy

The following control strategy outlines the control loops in place on a typical companded LOX boil ASU such as the one considered for this study. There will be an overall supervisory control program which will allow the ASU and CO_2 compression and purification systems to be adjusted automatically in response to planned load changes on the power boiler system in response to changing electrical demand from the grid. This control program will also allow for a controlled rate change of oxygen from the ASU by ramping the plant up or down at a rate between 2-3% of full flow per minute.

3.2.3.3 ASU Air Supply

Main Air Compressor

Air to the cold box supplied by the Main Air Compressor (MAC) is flow controlled by varying the guide vanes on the compressor. As the guide vanes are opened up, the flow rate through the compressor will increase which will ultimately increase the product flow of O_2 with a delay subject to the time constant of the system. There will be an associated increase in the discharge pressure of the MAC and this is monitored by a separate pressure control loop which will vent the MAC product air should the calculated approach to the compressor surge line become unacceptably close.

Air Purification System

The air to the main exchanger is passed through the adsorber beds where water, carbon dioxide, acetylene and heavy hydrocarbons and some N_2O are adsorbed.

The beds operate on a Thermal Swing Adsorption process (TSA), with bed regeneration obtained by heating the adsorbent at low pressure. Regeneration heat is provided by heating part of the low pressure waste N_2 stream leaving the Main Heat Exchanger (E101) as shown in PFD 3 in a steam reactivation heater and passing it through the bed in the reverse direction. This is followed by a cooling period when the heater is by-passed. The on-stream time for each bed is typically 3 to 6 hours, and while one vessel is on-line the other undergoes regeneration. The changeover of the on-line bed and subsequent regeneration is controlled entirely by a pre-programmed sequence in the DCS.

During the sequence when no re-generation gas is required, all waste gas is vented under pressure control. When the sequence moves from the depressurisation step onto the heating step the required valves are ramped slowly under automatic control minimising disturbance to the plant. Feed forward control is used when switching from the cooling step to the re-pressurisation step to further minimise disturbance to the ASU when the TSA inlet valve closes.

The temperature of the regeneration gas is controlled by regulating the regen gas flow through the reactivation heater. The steam supply through the heater has no temperature control and can vary, therefore there is a bypass of the reactivation gas around the heater to help to control the heating of the reactivation gas. The temperature controller acts to limit the reactivation feed flow through the heater by increasing the bypass flow to control the outlet temperature.

Automated front-end regeneration is provided for plant start-up, returning air from the on-line bed as regeneration gas through the reactivation heater. A coldbox trip also sets the TSA to front-end regeneration, allowing the compressor and TSA to continue running. There is a CO_2 analyser to sample the CO_2 content of the air to the coldbox leaving the TSA. The analyser is switched to sample the reboiler sump at regular intervals, generally coinciding with molecular sieve bed changeover.

Main Heat Exchanger

The air leaving the TSA adsorbers is passed through the main heat exchanger (E101) as shown in PFD 3, where it is cooled by counter current heat transfer with the returning waste N_2 and product GOX flows. The total air flow is controlled by the MAC guide vanes, with the split between expander and HP column flow controlled by the expander inlet guide vanes. The boosted air flow is determined by a temperature controller at the cold end of the main heat exchanger which regulates the flow of liquid air to the HP column via the JT valve.

ASU Compander

The energy produced by the expander part of the compander is used to drive the air booster compressor part. The Medium Pressure (MP) air flow through the expander is controlled by the expander inlet guide vanes and the booster discharge pressure varies as the booster and expander flows are adjusted.

3.2.3.4 ASU Column System

The general principle for maintaining the correct mass balance for the column section is detailed in the following sections. Air flow into the plant and product flows out are all flow controlled hence any gas not taken as product leaves the plant as waste.

Distillation Columns

The amount of nitrogen reflux flow from the condensers to the high and intermediate pressure columns is modulated using remotely operated control valves to maintain the correct column operating composition profile in both columns. Liquid air is withdrawn from the middle of the HP column and fed to the LP column via the subcooler under flow control.

Level controllers maintain the sump levels in the high and intermediate pressure columns by controlling the transfer of liquid to the LP column. These streams enter the LP column as crude LOX.

The LP column is split into three sections. In the top section, a waste nitrogen stream is taken off under pressure control at the warm end of the MHE. Liquid air from the high pressure column enters at the top of the second section and the Crude LOX (from the high and intermediate pressure column sumps) and LP air (from the expander) are fed to the column at the top of the third section. The LOX product collects in the column sump and is fed to the main heat exchanger where it is boiled and warmed to form the GOX product against cooling and condensing liquid air.

GOX Product

LOX from the LP column sump is vaporised and warmed in the main heat exchanger as described above. The GOX product is delivered to the plant on both pressure control and flow control. The pressure control acts by controlling the amount of LOX taken off the LP column sump and the flow rate is controlled by adjustment of the air flow from the MAC. A vent valve is provided to discharge excess oxygen if downstream problems arise.

The oxygen flowrate required by the oxy-combustion system determines the total demand on the two ASUs. At any total flow below the maximum, the two ASUs will automatically adjust their flows to produce the required total oxygen flow. As the total oxygen flow approaches maximum, the computer system will be programmed to indicate to the operators the oxygen flow and enable adjustments in demand to be made.

LOX Storage Control

Should an ASU trip a 30 minute supply of back-up oxygen at full flowrate can be provided to allow the ASC boiler to be changed from oxy-combustion to air-firing with no disturbance to steam generation or electrical power output or to allow the ASC boiler to be reduced in load to match the available oxygen form the remaining ASU. Backup LOX flows from the storage tank through a steam vaporiser into the back-up oxygen pipeline. To minimise any pressure disturbance an ASU trip signal initiates the back-up system response activating the LOX flow to the vaporiser.

Note: despite the high reliability of the ASU, in the event of one unit being out of service it is envisaged that the ASC boiler would be able to operate at 50% MCR load on a single ASU.

A pressure controller on backup GOX header opens the control valve on the discharge of the vaporiser. Because of the time lag in the system to initiate the GOX back-up flow through the vaporiser, a temporary means of providing GOX is required. The GOX pressure is maintained by a GOX buffer vessel at 30 bara pressure, which discharges into the GOX header under pressure control (PIC). Once the backup supply from storage is at full capacity the buffer vessel supply route is backed out, and the vessels are recharged by a small compressor, set to start on a low pressure switch and stop on a high pressure switch.

Liquid oxygen is stored in a vacuum insulated cryogenic vessel under a pressure of 2.5 bara. Two pressure controllers control the storage tank pressure. One acts on the tank vent valve to reduce the storage pressure. The other allows liquid to pass through a vaporiser and return to the tank as pressurising vapour. The set point of the vent valve controller is set to a higher value to avoid controller run-away.

3.2.4 CO_2 Recovery Plant Process Control

The CO_2 recovery plants would typically be designed to interface with the boiler control system, allowing ease of use and high reliability.

The design of the instrumentation and Distributed Control System (DCS) enables safe operation of the CO_2 recovery plant and provides the necessary control while handling disturbances and certain levels of process or instrumentation degradation.

The instrumentation and distributed control system is made to be sufficiently reliable and robust so as to require no operator involvement during normal plant operation.

The cooling water flow demand to the venturi mixers and Direct Contact Coolers (DCC) will be controlled under flow control. The level in the DCC sumps is controlled by the liquid flow leaving the vessel.

The drier operation is controlled by the DCS. Valve switching and sequencing will be automated for the entire adsorber cycle, including the on-stream and regeneration steps. The amount of moisture in the gas stream leaving the drier is monitored to ensure the performance of the system.

Dry crude CO_2 at 30 bara enters the plate fin heat exchanger where it is cooled to –55°C in two stages. The cooling is accomplished by evaporating two CO_2 streams taken from the intermediate separator at 18.4 bara and from the cold separator at 10.5 bara. These pressures are controlled using inlet guide vanes for flow control of the CO_2 product compressor at the appropriate points.

Level control is used to remove condensed liquid CO_2 in the two separators. The critical cold end temperature which ensures maximum inerts removal without CO_2 freezeout is obtained by partial bypass of the low pressure CO_2 refrigeration stream around the cold end section of the heat exchanger. The inerts stream is removed from the final separator warmed to ambient and passed through an expansion turbine whose guide vanes are on pressure control.

3.2.5 Oxygen Injection at Burners / Oxygen Handling

The oxygen generated in the ASUs must be distributed to the ASC boiler. In order to be able to use carbon steel piping, the pipeline network must be designed to a velocity limit to avoid the risk of fire caused by impingement of foreign objects within the piping against the pipe walls. In addition to the velocity restriction, there is also a restriction on the configuration of the piping so as to avoid situations in which impingement would be worse. Therefore, only long radius bends are used and T-junctions can only be used when flow goes from the main into the branch.

The key factor in supplying the oxygen for PF oxy-combustion combustion is that oxygen cannot safely be injected into the recycle flue gas flow through the coal mills i.e. via the primary recycle route. Whilst safe concentrations for oxygen in the primary recycle flow are likely to be of an order of similar to air, any significant increase above this is likely to result in the PF dust inside the mill igniting, with potentially disastrous results. In practical power plant operation, with mills being brought in and out of service to meet changes in electricity demand, the possibility of transient mismatches between recycle and oxygen flows appears significant, in addition to risks from equipment failures (e.g. oxygen flow control valve, recycle fan failure, blockages due to excessive coal in the mill).

Therefore the oxygen for the burners will be injected into the hot secondary gas recycle up to a maximum oxygen composition of 23% by volume. Any additional oxygen requirements will be injected directly into the burners and/or through overfire ‘air’ ports.

Injection of O_2 directly at the burner may lead to poor and unstable combustion as the full mixing requirements may not be achieveable in the furnace volume.

The alternative is to inject O_2 downstream of the mills but upstream of the burner ensuring that the O_2 levels added to the PA / PF stream are no greater than those which would be present in the PA / PF stream on an air-fired plant. The advantage of this scenario is that better mixing is envisaged which would directly contribute to burner combustion stability. A conventional burner design is also retained.

Note that this pulverised coal preparation/combustion arrangement is similar to that found in current pulverised coal plants burning high moisture (e.g. ~ 60% by weight), friable low rank brown coals (principally in Germany and Australia). Flue gas is recycled at approximately 1000ºC from the furnace exit to contact raw coal as it falls down a 'drying shaft' into a beater mill that also acts as circulating fan. Gas temperatures fall rapidly in the shaft, and drying is completed in the mill to give an exit temperature of about 120ºC and a coal moisture content of approximately 15% by weight. A cyclone may be inserted in the pulverised coal/flue gas riser to inject a fuel-rich mixture at the main burner level, promoting stable ignition. The moisture-laden remainder flue gases are injected at a higher level, after combustion is established. The oxygen for combustion, apart from a small amount in the recycle flow, is all supplied from the secondary air, which must be mixed with the primary coal/flue gas stream before combustion can take place.

Oxy-combustion arrangements do not match so directly those in pulverised coal plants for less reactive, higher rank coals that require much finer grinding in vertical spindle or ball/rod mills with internal oversize particle recycling and, conventionally, a flow of primary air through the mill to dry and transport the coal. In this case, volatiles combustion can be initiated using the oxygen in the primary air stream, with no need for mixing with the secondary air.

No insuperable difficulties are anticipated for oxy-combustion in this latter, more widely used, pulverised coal preparation/combustion configuration, but modified burners will need to be developed to ensure efficient and stable combustion.

The relative advantages for combustion, operability and safety of the various possible permutations of:

- Injecting some or all oxygen directly at the burner
- Injecting a fraction of the oxygen in the secondary recycle flow
- Injecting a fraction of the oxygen in the primary recycle flow downstream of the milling plant to promote stable combustion and to avoid burner modifications for direct oxygen injection

- Injecting oxygen at an overfire 'air' port

are beyond the scope of this study to determine, but overall plant cost and performance can be estimated irrespective of such configuration details.

4. NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE

4.1 Process Description and Performance

Section 4.1 initially presents the ground rule basis for the study, as agreed by the project partners based on the requirements of the IEA GHG R&D programme. Descriptions of a base case typical NGCC air-fired power plant without CO_2 capture and the NGCC oxy-combustion power plant with CO_2 capture are presented alongside heat and mass balances and performance data for the respective processes.

The NGCC Power Plant, designed for CO_2 capture, utilises turbomachinery designed to operate with thermodynamic and transport properties of CO_2/steam mixtures, using conventional design criteria. This plant is based essentially on the Gas-ZEP concept [31]. Other cycle options with water recycle or steam/CO_2 recycle are possible, however, the more conventional CO_2 recycle has been selected as the basis for this study. Characteristically, turbo-machinery speeds are reduced in CO_2 relative to air due to differences in gas properties which translates into smaller power-plant output at synchronous speeds for heavy duty machines.

4.1.1 Technical and Financial Ground Rules

The main technical and financial ground rules for this study are included in Section 3.0 based on the study requirements as stated in the IEA GHG R&D programme document[27].

Size of plant

The net power output[27] of the NGCC plant was targeted as $500MW_e$ in order to provide a reasonable comparison with the ASC PF power plant. In practice it is not possible to match the $500MW_e$ criterion due to gas turbine size. For this study there was a choice between the NGCC plant being based on one GT resulting in a power output of less than $500MW_e$ or two GTs with greater power output. Therefore the reference Case 3 is 1 x GT and the CO_2 capture case is 2 x GT.

Oxygen Purity

The O_2 purity supplied by the ASU was specified at 95% v/v. The justification for this is described in Section 4.1.3.1.2

4.1.2 Case 3: Typical NGCC Air Fired Power Plant Without CO_2 Capture Base Case Process Description

4.1.2.1 Description and Performance

A schematic of a simple combined cycle system consisting of a single gas turbine generator, a HRSG a single steam turbine generator, a condenser and associated auxiliary systems is shown in Figure 15.

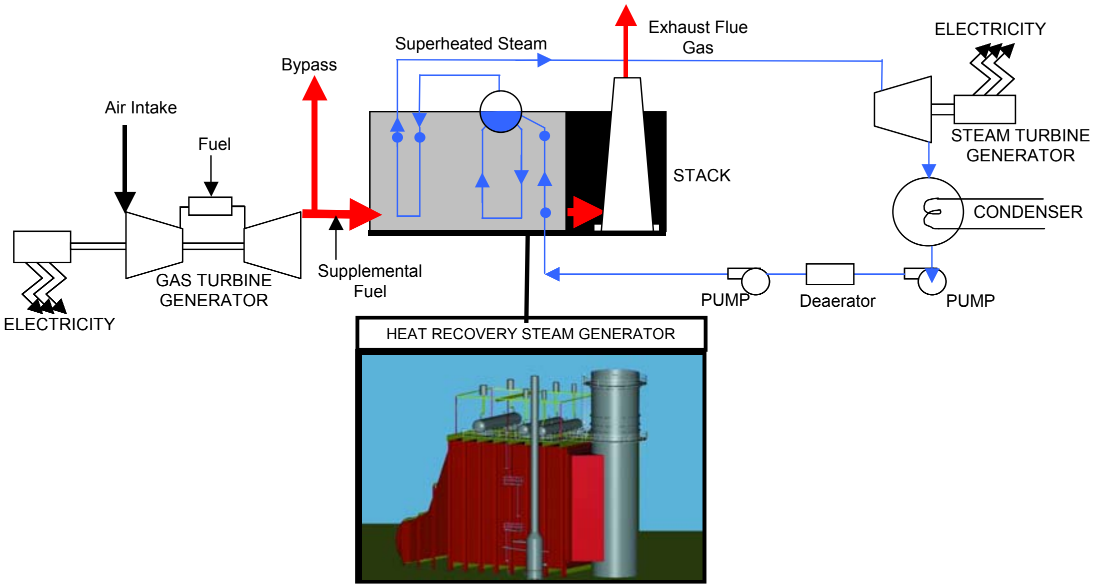


Figure 15: Conventional Air Fired NGCC Power Plant Arrangement

The 400MW$_e$ gross (388MW$_e$ net) base case originates from a current state-of-the-art design where a single 9FA type gas turbine provides direct electrical power and a Heat Recovery Steam Generator (HRSG) generates steam at three pressure levels from the flue gases which in turn provides additional power via the use of a single steam turbine set. The overall net cycle efficiency associated with this NGCC plant was taken as 56.0% (LHV).

4.1.2.2 Typical NGCC Air Fired Power Plant Without CO_2 Capture: Process Flow Diagram

For the costing purposes of this study and to highlight differences between conventional NGCC plant which uses once through air as the working fluid and plant designed for CO_2 capture the key components of the NGCC plant are listed below:

Unit 1000: NGCC Power plant

(1 Gas Turbine Island; 1HRSG Island; 1 Steam Turbine Island)

PFD 8 illustrates the major components and process streams for a conventional NGCC plant:

4.1.3 Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture Process Description

The combined cycle power plant concept employs a semi closed loop gas turbine with near stoichiometric combustion of natural gas in a mixture of CO_2 and oxygen to produce a mixture of CO_2 and steam (PFD 8). CO_2 is captured from the products of combustion after flue gas condensation and cooling to standard atmospheric conditions to facilitate conventional turbomachinery mechanical design.

Near stoichiometric combustion with 2% O_2 margin is required to ensure complete combustion of fuel for the expected degree of gas mixing in the combustion chamber (i.e. 2% over and above the stoichiometric requirement). Products of combustion are a mixture of CO_2 and H_2O which is expanded through the turbine for power generation in the gas turbine and to provide heat for the steam cycle.

Flue gas emerges from the HRSG at its dewpoint, 84°C, and is subsequently cooled to ~15°C, and water removed for recycle to GT, in order to preserve conventional turbomachinery design inlet conditions, minimise compressor flow size and maximise cycle performance potential.

Figure 16 shows the impact of recirculation temperature on net cycle performance and volumetric flow rate into the compressor.

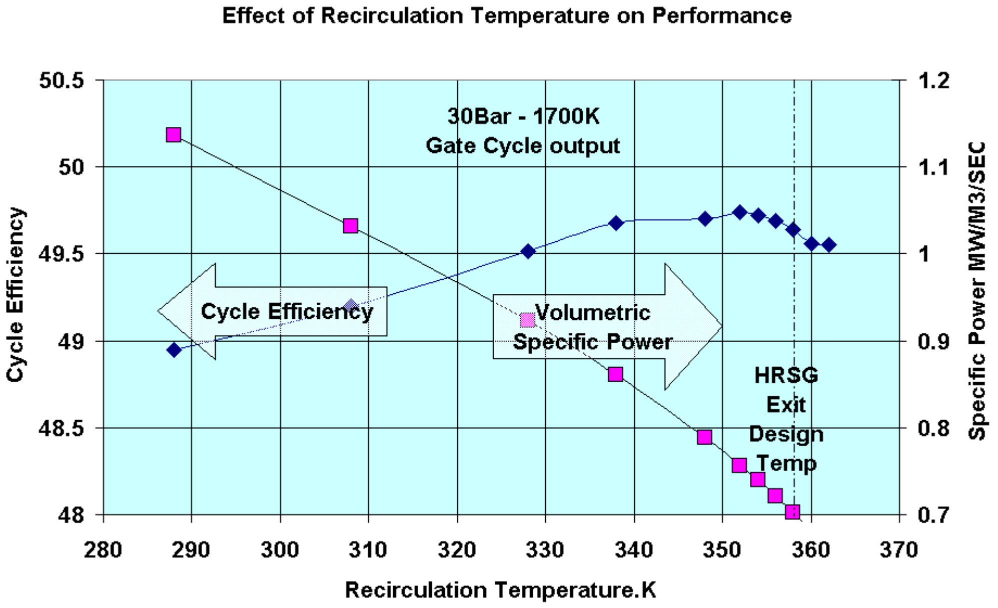


Figure 16: NGCC Power Plant with CO_2 Capture, The Effect of Flue gas Recirculation Temperature on Gas Turbine Performance

Cycle performance improves with less flue-gas cooling but volumetric specific power reduces by ~40%. Implication is that maximum volumetric specific power equals minimum capital cost in spite of the marginal loss of cycle efficiency. Also dry CO_2 at 288K is more attractive from compressor materials corrosion point of view.

Compression work is increased by higher compressor inlet temperature and flow with water content without complete flue gas condensation.

Water is removed from the flue gas, by direct contact with cold water, in readiness for capture and purification of surplus CO_2 produced in combustion. The flue gas cooler may form part of a district heating system and benefit the power plant thermo economics.

Captured CO_2 (~8% flue gas) is compressed to 30 bara, 31°C, in readiness for purification and separation of residual oxygen, argon and nitrogen, prior to sequestration at 110 bara. Inter-cooling requirement prevents more efficient compression in the Gas Turbine compressor.

Remaining CO_2 is recirculated and compressed to 30 bara from standard atmospheric conditions, in the gas turbine, prior to the addition of sufficient oxygen, provided by means of a cryogenic air separation unit, for near stoichiometric combustion.

Gas turbine compressor inlet flow is dictated by the requirement to deliver power at synchronous speed, 3000 RPM. Speed is determined by compressor stage loading required for adequate performance and stall margin. Stage loading is defined in terms of stage vector triangles which define blade angles and velocities in terms of stage work and mean blade speed (Enthalpy change / Mean blade speed squared).

Velocity triangles then provide an empirical description of aerodynamic diffusion loading for a given aerofoil section. Change in gas properties from air to CO_2 produces a reduction in speed for CO_2 because the enthalpy reduces for a given pressure ratio in CO_2 relative to air. Figure 17 shows typical compressor stage vector triangles illustrating the relationship between stage work and blade relative velocities.

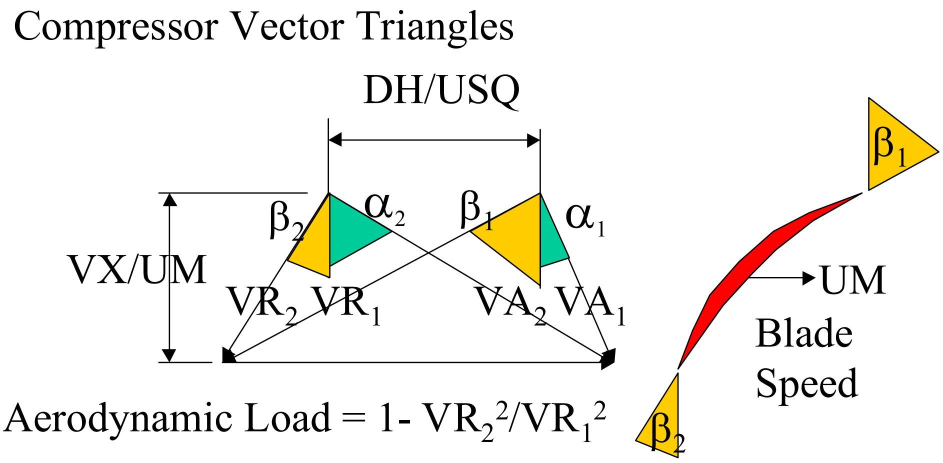


Figure 17: GT Compressor Vector Triangles

Figure 18 shows the difference between combined cycle power output for CO_2 and air for cycle pressure ratio of 30. Speed reduction for similar flow and power translates into significant reduction in system power at the same synchronous power output.

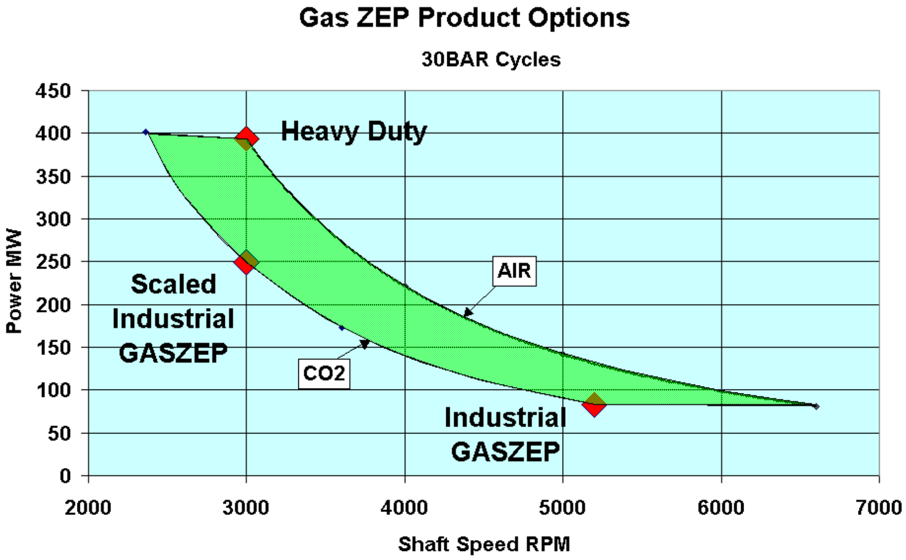


Figure 18: NGCC Power Plant, Gas Turbine Effect of Synchronous Speed Requirements on Combined Cycle Power Requirements

Firing temperature (1389°C) is adjusted to facilitate single shaft rotor design with zero turbine blade cooling requirement for the last turbine stage. Turbine exit temperature is higher than conventional air design due to the effects of CO_2 gas properties. However, the reduction in rotational speed, due to the compressor loading requirements for CO_2, reduces turbine blade stress levels and compensates for the increased temperature to maintain adequate blade mechanical creep rupture life. Figure 19 shows the relationship between metal temperature requirements for adequate creep rupture life for a range of unspecified materials.

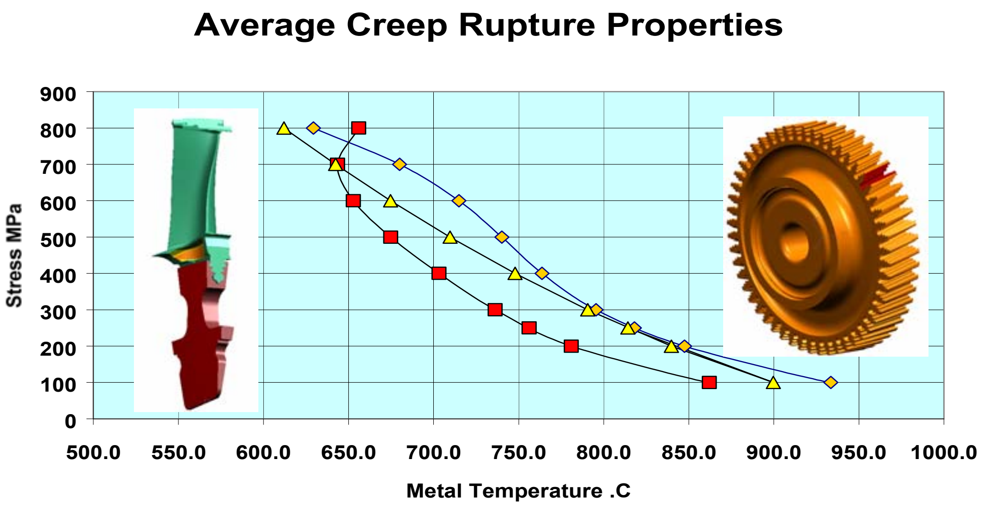


Figure 19: NGCC Power Plant, Average Creep Rupture Properties for Constant Life

Twin shaft design could provide relief for the last turbine stage rotor mechanical design by facilitating a reduction of power turbine rotational speed. This could facilitate a firing temperature increase, for performance improvement, but may exceed the limitations set by steam generator/steam turbine materials as well as high pressure turbine cooling design technology requirements.

Twin shaft justification with further increase in cycle pressure ratio would tend to reduce the gas turbine exit temperature, due to increased compression work, in a way that is sympathetic to increasing turbine rotor blade stress due to resulting flow increase.

Alternatively the use of reheat may facilitate increased power density without exceeding HP turbine cooling design technology while achieving maximum exhaust heat for the steam cycle.

Single shaft design is more desirable, for system operability, since shaft speed excursions due to rapid load changes are resisted by turbomachinery inertia. Twin shaft system would require blow off valves to release the stored energy in the gas path during rapid load reduction. This prevents power turbine over-speed and avoids exceeding disc-bursting speeds.

Turbine cooling is achieved with CO_2 bled from compressor inter-stage ports with appropriate source pressure levels to ensure adequate pressure margins for nozzle guide vane leading edge film cooling in each turbine stage. Normal cooling flow requirements for adequate blade life are facilitated by combination of colder coolant temperature and lower shaft speed due to CO_2 gas properties.

The steam cycle is conventional and has three pressure levels (HP, IP and LP) with reheat with maximum steam temperature of 600°C entering the high pressure and intermediate pressure steam turbines.

4.1.3.1 Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture Process Flow Diagram

For the purposes of this study, two gas turbines each with their associated HRSGs were combined to supply steam to a single steam turbine set on the basis of targeting 500MWe net power output. PFD 9 illustrates the major components for the NGCC power plant with CO_2 capture:

Unit 1000: NGCC Power Plant

Unit 2000: ASU

Unit 3000: CO_2 Purification and Compression

Properties of the fluid streams connecting power plant components are shown in PFD 9. Note: all the flows are the combined values for the duplicate streams.

4.1.3.1.1 Thermodynamic Process

The power plant system comprises a Brayton cycle gas turbine and a Rankine steam cycle linked by the Heat Recovery Steam Generator (HRSG) designed to utilise gas turbine waste heat for steam production.

Recirculated CO_2 is mixed with sufficient oxygen, from the Air Separation Unit (ASU) after compression, to provide for near stoichiometric combustion of natural gas with 2% excess O_2. The GT compressor outlet flow is depleted by the need for turbine cooling.

PFD 9 shows the relationship between components in the Gas ZEP (Zero Emission Power plant) process. The outputs from the system include combined cycle power and a flow of 88% purity net CO_2 product flow at 30 bara which is then ready for purification to 95% by volume and further compression to 110 bara.

4.1.3.1.2 Air Separation Unit

The NGCC plant requires two ASUs each providing 3420 tonne/day of contained oxygen at 40 barg. The cycle chosen for this duty is one in which gaseous oxygen (GOX) is produced by pumping liquid oxygen (LOX) to the required 40 barg pressure and allowing this stream to be boiled against condensing high pressure air. There is no requirement for a separate oxygen compressor.

A low purity pumped LOX cycle was chosen, which produces 95% by volume oxygen purity. Other studies[4] have been carried out to show that for oxy-combustion plants this is the optimum purity. Even with no air inleakage the increase in power required for the ASU to produce 99.5% purity oxygen is greater than the increase in CO_2 compression power required to remove the inerts introduced due to the lower purity oxygen. This is illustrated in Figure 20 which gives a plot of the net power required for oxygen production and CO_2 treatment for given purities of oxygen.

Cycle Description

To minimise the ASU power, a cycle has been chosen that uses two high pressure columns. A PFD and the mass balance of the process is given in PFD 10. A summary of the utility requirements are given in the table below. The standard double column cycle has a low pressure column (C105) with its reboiler (E103) integrated with the condenser of a high pressure column (C104). The column pressures are set to give a temperature driving force in the reboiler/condenser E103. In this cycle an extra column is added operating at an intermediate pressure (C103). The condenser (E104) for this column also integrates with a reboiler in the low pressure column but at a lower temperature, boiling a liquid stream higher up within the low pressure column which has a lower boiling point.

	Per Plant	Total
Power, (MWe)	45.0	90.0
Cooling Water, (tonne/hr)	3767	7534
Steam for ASU TSA regeneration (at 170°C), (MW)	3.0	6.0

This arrangement minimises the amount of feed air that must be compressed to the higher pressure of C104, leading to the low power requirement of this process cycle.

The plant consists of: -

- a compression system;
- an adsorption front-end air purification system;
- a cold box containing the separation and the heat exchanger equipment.

This process offers the benefits of high reliability, low maintenance cost and it is simple to install and operate.

Air Compression and Cooling

Air is taken in through an inlet filter to remove dust and particulate matter prior to entering the main air compressor (K101A) where it is compressed to 3.4 bara using an intercooled axial compressor. The air discharge is cooled in the Direct Contact Aftercooler (DCAC), in the lower section with cooling water and in the upper section with chilled water from the chiller tower. This uses evaporation of water into the dry waste nitrogen stream leaving the ASU cold box to further cool part of the plant cooling water. The air is cooled to a temperature of around 12°C. The main air compressor will be an in-line axial compressor driven by an electric motor. Around half of this compressed air stream is then further compressed in a single radial wheel (K101B) to 5.2 bara, cooled to ambient and compressed in the compressor/expander

K103/K104 to 5.4 bara. Part of this high pressure air stream is further compressed to 90 bara in a multi-stage integrally geared centrifugal compressor with intercooling.

Air Cleanup

Before the air is cooled to cryogenic temperatures, water vapour and carbon dioxide and other trace impurities such as hydrocarbons and nitrous oxide are removed in a dual bed adsorber (C101). Removal of carbon dioxide and water avoids blockage of cryogenic equipment. The removal of impurities results in a clean, dry air stream free from contaminants which might cause blockages or safety problems in ASU operation. The adsorber operates on a staggered cycle, i.e. one vessel is adsorbing the contained impurities while the other is being reactivated by low pressure gaseous waste nitrogen using a temperature swing adsorber cycle. The nitrogen is heated to around 160°C against condensing steam in a reactivation gas heater followed by a period in which the bed is cooled down with ambient temperature nitrogen which bypasses the heater. The adsorbents used are generally selected for optimum operation at the particular site. They consist of layers of alumina or silica gel plus layers of zeolite. The adsorber vessels are vertical cylindrical units having annular adsorbent beds.

Principle of Cryogenic Air Separation

The industry standard method of cryogenic air separation consists of a double column distillation cycle comprising a high pressure and a low pressure column. The high pressure, higher temperature cryogenic distillation produces an overhead nitrogen product that is condensed against the low pressure, low temperature liquid O_2 in the LP column sump. The plate-fin condenser-reboiler sits in the LP column sump and thermally links the HP and LP column. The HP column provides the boil up for the LP distillation column and the LP column O_2 provides the condensing duty for the HP column. Some of the condensed nitrogen returns to the high pressure column as reflux. The balance of the pure nitrogen reflux is cooled in the subcooler and flashed into the top of the low pressure column as reflux. The columns have aluminium structured packing optimised for cryogenic separation. In this cycle an extra column is added operating at an intermediate pressure (C103 as shown in PFD 10). The condenser (E104) for this column also integrates with a reboiler in the low pressure column but at a lower temperature, boiling a liquid stream higher up within the low pressure column.

Cooling and Refrigeration

Following the front end adsorber and compression of around half of the air stream to the pressure of the high pressure column, both the intermediate and high pressure air streams are split in two. Three of these four streams (7, 9 and 22 as shown in PFD 10) are fed directly to the main heat exchanger, E101, which consists of a number of parallel aluminium plate-fin heat exchanger blocks manifolded together. The fourth, stream 17, is compressed in a multi-stage intercooled compressor to 90 bara (K105) before entering E101 as stream 18.

The intermediate pressure stream 7 in PFD 10 is cooled close to its dew point (-171°C) and fed to the bottom of the intermediate pressure column. The second intermediate pressure stream 9 is removed from the main heat exchanger at -175°C then expanded in a centrifugal single wheel expansion turbine K104 running on the same shaft as a single wheel centrifugal compressor K103 which adsorbs the expander power. The expanded air is fed to the middle of the low pressure column at a pressure of about 1.4 bara and –184°C to provide refrigeration for the operation of the ASU. The high pressure stream 22 is cooled close to its dew point (-173°C) and fed to the bottom of the high pressure column. The high pressure air stream that has been boosted to 90 bara is cooled and condensed in the main heat exchanger against boiling oxygen at 40 barg. The resulting liquid air from the main exchanger is fed to the middle of both the high pressure and intermediate pressure columns.

Distillation System

In the high and intermediate pressure columns, the gaseous air feed is separated in the distillation packing into an overhead nitrogen vapour and an oxygen-enriched bottom liquid. The nitrogen vapour from the high pressure column (C104) is condensed against boiling oxygen in the low pressure column sump, and split into two parts. The first part is returned to the high pressure column as reflux, whilst the second part is subcooled in the subcooler (E102), reduced in pressure and fed to the low pressure column as reflux. The nitrogen from the intermediate pressure column is condensed against a boiling liquid stream in the low pressure column (C105). Part of this nitrogen is used as column reflux in the intermediate pressure column and part is subcooled and added to the reflux to the low pressure column.

Crude liquid oxygen is withdrawn from the sumps of the high and intermediate pressure columns, cooled in the subcooler (E102) against warming waste nitrogen and is flashed to the low pressure column as intermediate feeds. A portion of liquid air is also withdrawn from the middle of the high pressure column. This liquid is subcooled in the subcooler (E102) and fed to the middle of the low pressure column.

Low Pressure Column

The feeds to the low pressure column are separated into a waste nitrogen overhead vapour and a liquid oxygen bottom product, which reaches the required purity of 95% by volume.

The waste nitrogen is withdrawn from the top of the low pressure column and warmed in the subcooler (E102) and the main heat exchanger (E101). A portion of the nitrogen stream from the main exchanger is used for adsorber reactivation. The remaining dry nitrogen is vented through a chilled water tower to produce chilled water by evaporative cooling. The chilled water is used to provide additional feed air cooling in the top section of the DCAC.

Unlike the ASC PF Oxy-combustion plant pure liquid oxygen is withdrawn from the reboiler sump of the low pressure column, pumped to 40 barg and is returned to the main heat exchanger where it is vaporised and warmed up to ambient conditions against boosted air feed to the columns. The gaseous oxygen is then regulated and supplied to the NGCC power plant.

Oxygen Backup

Since the NGCC will not be designed to be able to operate on air, there is no benefit in the extra expense of backup for the ASU. Therefore, should the ASU fail to produce oxygen, the NGCC plant will be tripped.

Air Separation Equipment

Multiple structural steel cold boxes and one column can are supplied as part of the equipment. The column can is a cylindrical enclosure of pre-formed / pre-rolled flanged sections which bolt together at site to complete the structure. Steel jacket panels can be welded or bolted for equipment access to the framework. The cold boxes and column vessel are inclusive of process equipment. The process equipment is supplied and constructed of material suitable for use at low temperature.

The column can encloses the high, intermediate and low pressure columns. The reboiler and condensers are contained in the low pressure column. All heat exchangers in the cold air separation equipment are multi-passage, extended surface aluminium / aluminium alloy, plate-fin heat exchangers.

The main heat exchanger and subcooler are prefabricated. The main heat exchanger (MHE) box houses the main heat exchanger and the expander units. The subcooler box contains a multi-passage, extended surface plate fin heat exchanger.

The primary insulation material is expanded pearlite. Certain areas are packed with rockwool to allow access for maintenance of valves without pearlite removal.

A dry nitrogen purge system is included on all cold boxes and cans to prevent moist atmospheric air from leaking into the cold box/can during normal operation.

4.1.3.1.3 CO_2 Treatment

The net flue gas, essentially raw CO_2, leaves the NGCC plant at 1.03 bara and is compressed to 30 bara and is the feed to the process shown in PFD 11. A summary of the performance of this system is given below.

Performance Summary for CO_2 Treatment System : 95% CO2 purity			
Flue Gas Heater (to 325°C)	(MWe)	1.3	
Feed CO_2 Compressor power to 30 bar	(MWe)	23.6	
Net Compressor/Expander Power (MWe)		5.4	
Cooling Water (tonne/hr)		1554.92	
CO_2 leaving cold box	@ 9.7 bar	11.5	(%)
	@ 21.5 bar	88.5	(%)
CO_2 Captured	Purity	95.89	(% v/v)
	Contained CO_2	177.18	(tonne/hr)
	Recovery	97.16	(%)

The CO_2 treatment plant consists of:

- A drier system
- A cold box containing CO_2 purification equipment
- A compression system

The raw CO_2 is dried and the inerts (N_2 and Ar) and oxygen are separated to give >95 mol% CO_2. The CO_2 is then compressed to 110 bara for pipeline transmission. Any excess oxygen (there is no SO_2 or NO_X in the flue gas) present in the CO_2 need not be removed as the final CO_2 product will be used either for enhanced oil recovery (EOR) or stored in aquifers. The heat and mass balance for this process is given in PFD 11.

The raw CO_2 gas passes through a temperature swing dual bed desiccant dryer C201 to reach a dew point of below -55°C before entering the "cold box". The cold equipment is contained in a steel jacketed container or "cold box" with pearlite granular insulation. This desiccant dryer system prevents ice formation which could cause a blockage in the cold box as well as causing corrosion in the pipeline.

The inerts removal process uses the principle of phase separation between condensed liquid CO_2 and insoluble inerts gas at a temperature of –55°C which is very close to the triple point or freezing temperature of CO_2. The CO_2 partial pressure can be reduced to about 7 bara. The actual CO_2 pressure levels used for the separation are fixed by the specification of >95 mol% CO_2 product purity and the need to reduce the CO_2 vented with the inert to an economic minimum.

As shown in PDF 11, the system proposed uses two flash separators, C202 and C203, at temperatures of -27°C and -55°C. The CO_2 feed gas pressure is at 30 bara.

The necessary refrigeration for plant operation is obtained by evaporating liquid CO_2 at pressure levels of 21.5 bara (stream 19) and 10.1 bara (stream 13) and compressing these two low pressure gas streams in the main CO_2 product compressor to the final pipeline delivery pressure of 110 bara. The separated inert gas at 29 bara leaving the cold box can be heated and passed through a power recovery turbine. It is possible to reach a CO_2 purity in excess of 96% by volume using this method at inlet CO_2 concentrations as low as 77% by volume with a CO_2 recovery of better than 90%.

The dry gas is fed to the cold box and is cooled by heat exchange to –27°C with the returning evaporating and superheating CO_2 streams and the waste streams in the main exchanger. The main heat exchangers, E201 and E202, are multi-stream plate-fin aluminium blocks. The cooled feed stream 3 is sent to a separator pot C202, the stream is split into liquid and vapour; the liquid produced, stream 17, contains part of the required CO_2 product.

The vapour from the separator, stream 4, still contains a large proportion of CO_2. In order to recover this CO_2 the vapour is cooled further to –55°C where it partially condenses and is passed to another separator pot C203. The pressure at this point is critical in controlling the process, cooling the vapour below –56.2°C would lead to the formation of solid carbon dioxide. The vapour, stream 6, from the second separator (C203), containing the separated inerts together with some CO_2 at a partial pressure of about 7 bara, is sent back through the heat exchangers E201 and E202 where it is heated to 21.6°C. This stream of inerts which is at a pressure of 29 bar is heated against hot flue gas in the boiler island and expanded in a power producing turbo-expander (K203) before being vented.

Liquid, stream 17, from the first separator C202 containing part of the CO_2 is expanded through a J-T valve to 21.7 bara and heated to 21.6°C. The liquid, stream 11, from the second separator, is heated, expanded through a valve to 10.1 bara and a temperature of about –56°C to provide refrigeration in E202 by evaporation, while the vapour formed is heated to 21.6°C in E202 and E201. The CO_2 vapour is then compressed in a single radial wheel (K202) to 21.7 bara (C202 stream 19), the same pressure as the CO_2 stream from the first separator. The two streams are combined (stream 20) and compressed to the required pressure of 110 bara (stream 21) This machine (K201) is a four stage unit (see Figure 21) which could be operated from the 21.6 bara to 110 bara level as either an intercooled compressor or as an adiabatic compressor with an aftercooler used to heat condensate. Since the adiabatic compression heat cannot be used in this case to preheat condensate, only the intercooled option is considered here. The likelihood of dense fluid CO_2 forming in K201 has meant that the machine has only one intercooler, to prevent the dense phase forming within the machine

The plant is furnished complete with all structural, mechanical equipment, piping, supports, anchor bolts, electrical equipment, instrumentation, controls and accessories as required for continuous automatic operation. The controls are to be designed to interface with the power plant systems. The intent would be to operate the plant from a remote central control room with periodic inspection.

4.1.3.2 Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture: Performance

The actual combined cycle power plant system, being studied, consists of two gas turbines with corresponding ASU oxygen supply accompanied by a single steam cycle to provide 575MW_e gross (440MW_e net) at 44.7% net efficiency (LHV) with allowance for ASU power, feedwater pumping power, CO_2 compression and inerts removal, and generator losses. This allows comparison of this type of configuration with the 532MW_e net ASC PF oxy-combustion power plant equivalent.

The penalty therefore in terms of net cycle efficiency associated with the NGCC CO_2 capture plant is calculated at 11.3% points when compared to the reference NGCC power plant without CO_2 capture

The auxiliary power consumption for the NGCC CO_2 capture plant is shown in Table 3.

4.1.3.2.1 CO_2 vs Air-Firing Cycle Performance

Comparison of CO_2 with air at 20 bar cycle pressure ratio shows the impact of ASU power requirement on gas turbine efficiency which is reflected in the overall combined cycle efficiency. The table below shows the effect of CO_2 and cycle pressure ratio on performance. Note CO_2 capture power loss is excluded from these figures.

CO_2 vs. Air Cycle Performance

	AIR	CO_2	CO_2
GT Pressure Ratio	20	20	30
ST Inlet Pressure (BAR)	145	190	190
Steam Process	3 Press RH	3 Press RH	3 Press RH
GT Efficiency	0.38	0.202	0.24
ST Efficiency	0.308	0.299	0.297
CC Efficiency	0.57	0.441	0.466
GT Specific Power kJ/Kg	393	232	297
ST Specific Power kJ/Kg	198	273	278
CC Specific Power kJ/Kg	591	524	596

Note: CO_2 capture power is excluded from these figures

Comparison between 30 bar and 20 bar cycles, with CO_2 working fluid, shows a 4% points improvement in the gas turbine efficiency. The GT pressure ratio of 30 bar represents the current limit for conventional gas turbines and was therefore selected for in depth design comparison with conventional gas turbine design technology.

4.1.3.2.2 Performance Potential

Figure 22 below indicates the remaining performance potential as a function of cycle pressure ratio, for single shaft design, where the design is constrained by gas turbine cooling requirements for zero cooling in the final stage turbine. Full potential of such a system is likely to be ~50% cycle efficiency using existing cryogenic air separation unit and CO_2 capture systems and with the additional benefit of steam cooling which can be worth ~2% points cycle efficiency.

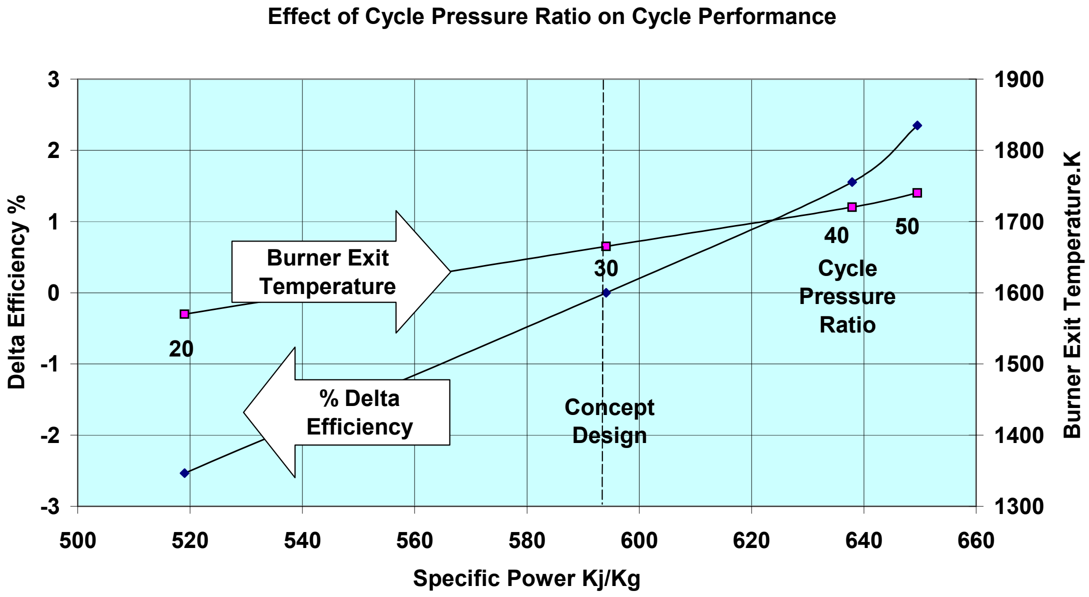


Figure 22: NGCC Power Plant with CO2 Capture, Effect of GT Pressure Ratio on Cycle Efficiency

4.1.3.2.3 NGCC GT Compressor Design

Table 4 shows how much the change in gas properties from air to CO_2 changes the speed requirement for equal stage loading with corresponding reduction in GT compressor exit temperature. Changing the working fluid from air to CO_2 effectively reduces the speed to 77% of the value for air with the same number of stages. Increasing the pressure ratio at the same speed increases the stage number to 18 for the same stage loading. Loading is defined as:-

(Compressor Enthalpy Rise per Stage) / (Mean blade speed) ^{2}

4.1.3.2.4 Turbine Blade Cooling

The corresponding effect of changed gas properties during expansion is to increase the turbine exit temperature which conspires to make turbine cooling and mechanics more severe by virtue of the higher temperature gradient through the turbine blade surface with cooler CO_2 coolant. Table 5 illustrates how the turbine boundary conditions alter when changing the working fluid from air to CO_2 indicating a significant increase in turbine exit temperature due to gas properties. Increasing cycle pressure ratio for CO_2 allows an increase of firing temperature for the same turbine exit temperature and rotational speed.

However, the associated reduction in speed compensates for the increased severity in temperature gradient to provide a satisfactory blade cooling solution. Figure 23 illustrates the difference in cooling characteristics for a 30 bar Gas ZEP compared to a 20 bar conventional industrial machine. Although the heat transfer is higher for Gas ZEP the coolant volumetric flow rate is lower for blades and vanes due to the density increase. This implies a relatively lower blade cross sectional area requirement for CO_2 comparing V1 and R1 for each design.

(Nomenclature of chart is V1 = Stage 1 Vane, R1 = Stage 1 Rotor Blade etc).

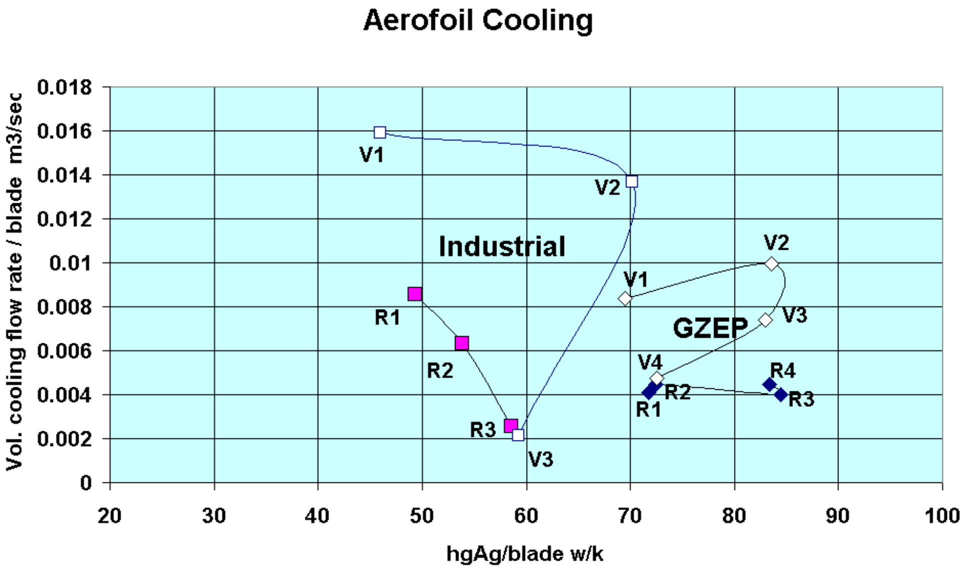


Figure 23: NGCC Power Plant with CO_2 Capture, Effect of CO_2 and Cycle Pressure Ratio on GT Turbine Rotor Blade Cooling

4.1.3.2.5 Effect of Oxygen and CO_2 Purity on Cycle Performance

Interaction of supplied oxygen purity with captured CO_2 purity, with cycle performance, is shown in Figure 24, indicating increasing power losses with increasing purity of both CO_2 and oxygen.

4.1.3.3 Clean Up Technologies for CO_2 Purity

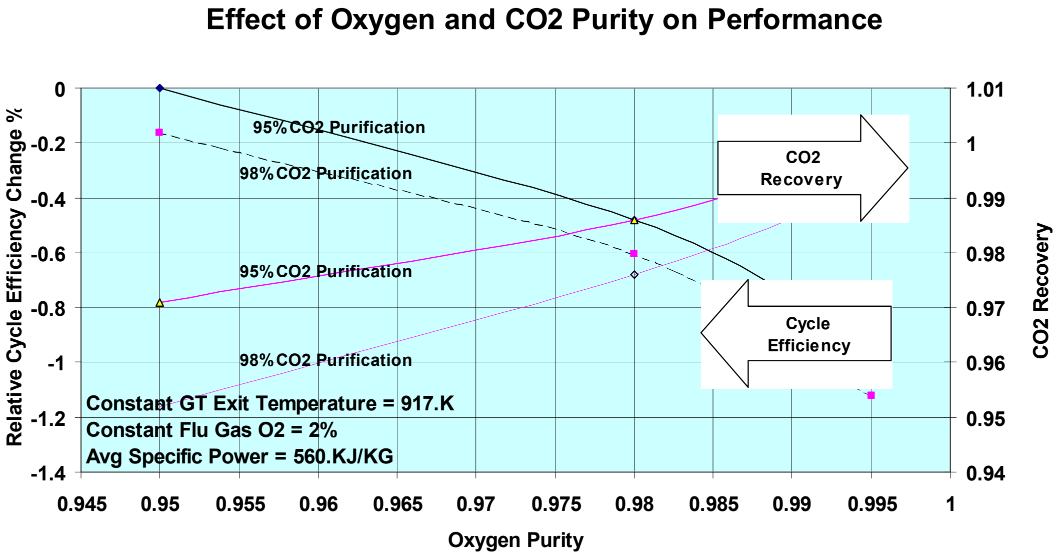


Figure 24: NGCC Power Plant with CO_2 Capture, Effect on Oxygen and CO_2 purity on Cycle Performance

The base case of this study produces CO_2 at > 95 mol% using a simple process that gives the required purity at high (> 90%) recovery. This process is easily modified to give higher CO_2 purities.

As it stands the basic process cycle produces a CO_2 purity of almost 96% by volume. To produce higher, 98% purity CO_2, this process requires a modification. This is achieved by the addition of a third separation vessel to remove the non-condensed portion of stream 13 in PFD 11, shown in bold in PFD 12. To prevent a loss of CO_2 from the process, since there will be non-condensed CO_2 in this stream, this stream is compressed from around 10 bara to 30 bara, in a two stage intercooled compressor, and recycled by adding to the cold box feed after the dryers.

Pure (>99.99mol%) CO_2 is also possible using distillation in place of separation in this flowsheet.

4.1.3.4 CO_2 Treatment Plant Performance & Cost Sensitivity to CO_2 Purity

The performance of the CO_2 treatment plant with 98 mol% purity CO_2 is shown in the table below. A small amount of extra power and cooling water are required over the 95 mol% purity case.

Capital costs will be slightly increased due to the recycle compressor, extra separation vessel and the extra passages through the exchangers, but this will be minimal. There is also a small reduction in CO_2 capture efficiency of the process, down from 97.2% to 96.3%, resulting in a small reduction in CO_2 captured.

Performance Summary for the CO_2 Treatment System for 98 mol% CO_2 Purity			
Flue Gas Heater (to 325°C)		1.64	MWt
Net Compressor/Expander Power		5.6	MWe
Cooling Water		1605.5	tonne/hr
CO_2 leaving cold box	@ 10.7 bar	44.1	%
	@ 22.6 bar	55.9	%
CO_2 Captured	Purity	98.02	%v/v
	Contained CO_2	175.64	tonne/hr
	Recovery	96.32	%

The cost and performance penalties of pure CO_2 will be higher than those for 98% purity CO_2.

4.1.3.5 Materials, Corrosion and Impurities Build Up

Gas Turbine Issues

The response of high temperature materials to flow environment which contains steam and CO_2 is currently unknown and is the subject of a materials test programme[31].

It is likely that some impurities may build up in concentration in the system since it is semi-closed loop. However, some of this contamination will be removed by the flue gas condenser which effectively washes the flue gas with water to provide direct contact gas cooling.

CO_2 Compressor Issues

The degree of contamination also depends on composition of the fuel which may be subject to periodic variation in quality.

Compressor material selection for the wet CO_2 compression needs careful consideration. In this case SO_2 concentration will be low. Previous studies have suggested doubling the nickel content in 316 stainless steel to 904 austenitic stainless steel to combat this potential corrosion issue. An appropriate material specification would be:

Alloy 20Cb-3 (UNS No: NO8020) – 20Cr2.2Mo34Ni3.5Cu austenitic stainless steel for impact areas or cold areas such as volutes, impeller, intercoolers and internals;

Alloy 2205 – 22Cr5Ni3Mo duplex stainless steel for shafting.

4.2 Safety and Operability

4.2.1 NGCC Power Plant Operability and Control

Semi-closed loop power plant operability will require increased controls and instrumentation than conventional air-fired NGCC plant to account for the variation in gas turbine working fluid composition due to variation in fuel composition, ASU supply oxygen purity, CO_2 capture purity, flue gas cooling variation and power demand during normal operation.

Compressor inlet conditions must be maintained near standard atmospheric values of 15°C and 1 bar at whatever flow is demanded to satisfy power requirements for any gas composition.

Purity of the CO_2 for capture will depend on composition of the fuel as well as oxygen purity delivered from the ASU and therefore gas turbine control system needs to be integrated with the ASU and CO_2 capture sub-systems.

Transient loads will require some form of compensation for the amount of stored energy contained within the flue gas return ducting. Recirculation conduit is required to transfer flue gas from HRSG exit to Compressor Inlet. Typically blow-off valves are used to release stored energy to accommodate such load transients. It is anticipated that normal load transients will cycle between 50% and 100% power at constant speed. This variation is facilitated by compressor variable guide vanes in steady state operation.

Load rejection may require more drastic action since complete shutdown is implied with the possibility of a hot restart. Special procedures will be required to ensure that all systems can function immediately following a shutdown.

Thermal soakage distorts casing and rotors, creating a tip clearance problem which is distinct from a cold restart. This would degrade design operation if it were uncontrolled.

4.2.2 ASU Safety Issues

See section 3.2.2.

4.2.3 ASU Plant Process Control

4.2.3.1 Control System Design Philosophy

The control system is designed to meet the following overall objectives:

- To provide a safe system.
- To meet the plant reliability and availability targets.
- To enable the plant to run routinely within the particular operation constraints.

4.2.3.2 Control Strategy

The following control strategy outlines the control loops in place on a typical pumped LOX ASU such as the one considered for this study. There will be an overall supervisory control program which will allow the ASU and CO_2 compression and purification systems to be adjusted automatically in response to planned load changes on the NGCC plant in response to changing electrical demand from the grid. This control program will also allow for a controlled rate change of oxygen from the ASU by ramping the plant up or down at a rate between 2-3% of full flow per minute.

4.2.3.3 ASU Air Supply

Main Air Compressor

Air to the cold box supplied by the Main Air Compressor (MAC) is flow controlled by varying the guide vanes on the compressor. As the guide vanes are opened up, the flow rate through the compressor will increase which will ultimately increase the product flow of O_2 with a delay subject to the time constant of the system. There will be an associated increase in the discharge pressure of the MAC and this is monitored by a separate pressure control loop which will vent the MAC product air should the calculated approach to the compressor surge line become unacceptably close.

Air Purification System

All the air to the ASU at a pressure of 3.36 bara is passed through a set of adsorber beds where water, carbon dioxide, acetylene and heavy hydrocarbons and some N_2O are adsorbed.

The beds operate on a Thermal Swing Adsorption process (TSA), with bed regeneration obtained by heating the adsorbent at low pressure. Regeneration heat is provided by heating part of the low pressure waste N_2 stream leaving the Main Heat Exchanger (E101) as shown in PFD 10 in a steam reactivation heater and passing it through the bed in the reverse direction to the air. This is followed by a cooling period when the heater is by-passed. The on-stream time for each bed is typically 3 to 6 hours, and while one vessel is on-line the other undergoes regeneration. The changeover of the on-line bed and subsequent regeneration is controlled entirely by a pre-programmed sequence in the control system.

During the sequence when no re-generation gas is required, all waste gas is vented under pressure control. When the sequence moves from the depressurisation step onto the heating step the required valves are ramped slowly under automatic control minimising disturbance to the plant. Feed forward control is used when switching from the cooling step to the re-pressurisation step to further minimise disturbance to the ASU when the TSA inlet valve closes.

The temperature of the regeneration gas is controlled by regulating the regen gas flow through the reactivation heater. The steam supply through the heater has no temperature control and can vary, therefore there is a bypass of the reactivation gas around the heater to help to control the temperature of the reactivation gas. The temperature controller acts to limit the reactivation feed flow through the heater by increasing the bypass flow to control the outlet temperature.

Automated front-end regeneration is provided for plant start-up, returning air from the on-line bed as regeneration gas through the reactivation heater. A coldbox trip also sets the TSA to front-end regeneration, allowing the compressor and TSA to continue running. There is a CO_2 analyser to sample the CO_2 content of the air to the coldbox leaving the TSA. The analyser is switched to sample the reboiler sump at regular intervals, generally coinciding with molecular sieve bed changeover.

Main Heat Exchanger

The air leaving the TSA adsorbers is passed through the main heat exchanger (E101 as shown in PFD 10), where it is cooled by counter current heat transfer with the returning waste N_2 and product GOX flows. The total air flow is controlled by the MAC guide vanes, with the split between LP and MP plus HP column flow controlled by the guide vanes on the K102 air compressor. The expander flow is controlled by the setting up of the expander inlet guide vanes. The boosted air flow is determined by a temperature controller at the cold end of the main heat exchanger which regulates the flow of liquid air to the HP column via the JT valve. The booster compressor guide vanes control the air flow.

ASU Compander

The energy produced by the expander part of the compander is used to drive the air booster compressor part. The medium pressure (MP) air flow through the expander is controlled by the expander inlet guide vanes and the booster discharge pressure varies as the booster and expander flows are adjusted.

4.2.3.4 ASU Column System

The general principle for maintaining the correct mass balance for the column section is detailed in the following sections. Air flow into the plant and product flows out are all flow controlled hence any gas not taken as product leaves the plant as waste.

Distillation Columns

The amount of nitrogen reflux flow from the condensers to the high and intermediate pressure columns is modulated using remotely operated control valves to maintain the correct column operating composition profile in both columns. Liquid air is withdrawn from the middle of the HP column and fed to the LP column via the subcooler under flow control.

Level controllers maintain the sump levels in the high and intermediate pressure columns by controlling the transfer of liquid to the LP column. These streams enter the LP column as crude LOX.

The LP column is split into three sections. In the top section, a waste nitrogen stream is taken off under pressure control at the warm end of the MHE. Liquid air from the high pressure column enters at the top of the second section and the Crude LOX (from the high and intermediate pressure column sumps) and LP air (from the expander) are fed to the column at the top of the third section. The LOX product collects in the column.

GOX Product

LOX from the LP column sump is pressurised in a liquid pump, vaporised and warmed in the main heat exchanger against cooling and condensing air. The GOX product is delivered to the plant on both pressure control and flow control. The pressure control acts by controlling the amount of LOX taken off the LP column sump and the flow rate is controlled by adjustment of the air flow from the MAC. A vent valve is provided to discharge excess oxygen if downstream problems arise.

The oxygen flowrate required by the oxy-combustion system determines the total demand on the two ASUs. At any total flow below the maximum, the two ASUs will automatically adjust their flows to produce the required total oxygen flow. As the total oxygen flow approaches maximum, the computer system will be programmed to indicate to the operators the oxygen flow and enable adjustments in demand to be made.

4.2.4 CO_2 Recovery Plant Process Control

The CO_2 recovery plants would typically be designed to interface with the Gas turbine control system, allowing ease of use and high reliability.

The design of the instrumentation and Distributed Control System (DCS) enables safe operation of the CO_2 recovery plant and provides the necessary control while handling disturbances and certain levels of process or instrumentation degradation. The instrumentation and distributed control system is made to be sufficiently reliable and robust so as to require no operator involvement during normal plant operation.

The drier operation is controlled by the DCS. Valve switching and sequencing will be automated for the entire adsorber cycle, including the on-stream and regeneration steps. The amount of moisture in the gas stream leaving the drier is monitored to ensure the good performance of the system.

Dry crude CO_2 at 30 bara enters the plate fin heat exchanger where it is cooled to –55°C in two stages. The cooling is accomplished by evaporating two CO_2 streams taken from the intermediate separator at 21.5 bara and from the cold separator at 10.1 bara. These pressures are controlled using inlet guide vanes for flow control of the CO_2 product compressor at the appropriate points.

Level control is used to remove condensed liquid CO_2 in the two separators. The critical cold end temperature which ensure maximum inerts removal without CO_2 freezeout is obtained by partial bypass of the low pressure CO_2 refrigeration stream around the cold end section of the heat exchanger. The inerts stream is removed from the final separator warmed to ambient and passed through an expansion turbine whose guide vanes are on pressure control.

4.2.5 NGCC Oxy-Combustion Power Plant Start Up Issues

Start-up is anticipated to require special procedures since the Gas Turbine flowpath will need to be primed with CO_2 in order to satisfy the turbomachinery flow requirements until self sustaining speed is reached by the system, typically 50%. This will necessitate a CO_2 liquid storage and vaporisation system requirement.

5. ECONOMICS

The technical and financial ground rules used for this study are included in Section 3.0

5.1 ASC PF Oxy-Combustion Power Plant with CO_2 Capture

In order to evaluate the economics of the ASC PF oxy-combustion power plant with CO_2 capture, investment costs and O&M costs were generated for:

Case 1: 740MWe gross ASC PF air-fired plant without CO_2 capture; and

Case 2: 740MWe gross ASC PF oxy-combustion power plant with CO_2 capture.

Note the economic analysis featured within this study is based on CO_2 capture and CO_2 compression to 110 bara, costs associated with transport and storage are excluded as they are generally site specific.

Investment Costs

Tables 6 and 7 illustrates the installed capital costs in Euros associated with Case 1 and Case 2 respectively. The installed costs are given and the owners costs (including contingency and fees) are added as a percentage of the installed costs to arrive at the Total Investment Costs (TIC).

The estimated accuracy of figures quoted is ± 25%.

The costs have been normalised to US$ in line with the IEA guidelines[27]. Note: In compiling costs some of the units were estimated in US$ whilst others were in Euros. In the case of dollar pricing a conversion has been made at the rate of 1 Euro = 1.2 US$.

For each of the cases the following major power plant units have been identified and the costs associated with the respective unit expressed.

- Case 1: 740 MW_e Gross ASC PF Air-Fired Power Plant Without CO_2 Capture

Unit 100	Coal and Ash Handling
Unit 200	Boiler Island
Unit 300	FGD Plant and Handling Plant
Unit 400	DeNOx Plant
Unit 500	Steam Turbine Island
Unit 800	BoP, Electrical, I&C

- Case 2: 740 MW_e Gross ASC PF Oxy-Combustion Power Plant with CO_2 Capture

Unit 100	Coal and Ash Handling
Unit 200	Boiler Island
Unit 500	Steam Turbine Island
Unit 600	ASU
Unit 700	Inert Removal and CO_2 Compression
Unit 800	BoP, Electrical, I&C

The outline equipment list for each of the above cases is given in Tables 8 and 9.

The direct field costs DFC are defined as

- Direct Materials: Including equipment and bulk materials.
- Construction: Including mechanical erection, instrument and electrical installation, civil works and where applicable buildings and site preparation.
- Other costs Including temporary facilities, solvents, catalysts, chemicals, training, commissioning, start up costs and spare parts.
- EPC services Including contractors home services and construction supervision and freight.

The indirect field costs (IFC) are:

- Contingency: 10% of the total installed costs
- Fees: 2% of the total installed costs
- Owner Costs: 16% of the total installed costs

Note: these costs are inherently site specific. The IEA GHG basic criteria for owners costs was suggested as 5% and this was used for ASU and CO_2 compression plant equipment. However, the resulting 16% is the overall weighted average figure for the power plant with CO_2 capture [32].

Land, taxes, project management, transmission lines to an existing grid system, other infrastructure changes and financing can add >40%[32] onto a basic EPC price. For the purposes of this study, the indirect field costs have been taken as 32% of the EPC of the basic PF Power Plant based on Mitsui Babcock's in-house cost data base.

The summing of the above costs generates the Total Investment Cost (TIC) for each of the major power plant units. (i.e. TIC = DFC + IFC).

Table 10 shows the percentage contribution of each unit to the overall Total Investment Cost for Case 1 and Case 2.

The Total Investment Costs for Case 1 and Case 2 are expressed as a function of their gross and net power output to arrive at the Specific Investment costs associated with each power plant, namely:

- Case 1: 1152 Euro/kW$_e$ Gross (1382 US$/kW$_e$ Gross)*
  1260 Euro/kW$_e$ Net (1512 US$/kW$_e$ Net)*
- Case 2: 1408 Euro/kW$_e$ Gross (1690 US$/kW$_e$ Gross)*
  1951 Euro/kW$_e$ Net (2342 US$/kW$_e$ Net)*

*Note: the above figures are based on exchange rate of 1 Euro = 1.2 US$.

Where possible the resulting specific investment cost have been checked against market prices to ensure the estimates prepared in this study are representative of the current market prices.

For comparisons capital costs of supercritical PF plant quoted in literature range widely. Kjaer[32] quotes 1400 US$/kW$_e$ (assumed gross) for a 400 MW$_e$ plant in Europe, while the world bank quotes 1000-1200 US$/kW$_e$ for advanced plant. However, site specific factors, exchange rate, price of steel and owners costs have a significant influence on the TIC.

Operation and Maintenance Costs

The operating and maintenance cost figure is formed from a fixed cost and variable cost.

The variable cost element in the total operating and maintenance cost figure is derived from the consumables associated with each plant. The consumables associated with Case 1and Case 2 are listed in Table 11 and are based on 7446 operating hours in a year (85% load factor following commissioning and start up).

The fixed cost contribution to the total operating and maintenance cost figure is shown in Table 12. These costs are calculated using:

- Operator cost and supervision (50k US$/yr per operator)

- Administration and general overheads (30% of direct labour)
- Maintenance 4% of installed costs per year as excluding contingency fees and owner costs as per IEA technical specification[27].

Therefore the total operation and maintenance costs calculated for both plants are:

Case 1: 99.571 M Euro per year (119.485 M US$ per year)

Case 2: 102.140 M Euro per year (122.568 M US$ per year)

CO_2 Abatement Costs for PF Plant

In order to generate the Cost of Electricity (CoE) in USc/kWh and the Cost of CO_2 avoidance (US$/ton CO_2) associated with the PF base case and the CO_2 capture case, the figures generated above were entered into IEA's GHG standard economic spreadsheet version 01, February 2003.

The sensitivity of the model results to varying fuel cost (± 100%), discount rate (10% and 5%) and Total Investment Cost (-20%) were then evaluated.

The results of the spreadsheet models are shown in:

Case 1: PF Power Plant without CO_2 Capture

Table 13: Case 1: ASC PF Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate

Table 14: Case 1: ASC PF Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate & +100% Fuel Price

Table 15: Case 1: ASC PF Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate & -100% Fuel Price

Table 16: Case 1: ASC PF Air Fired Power Plant Without CO_2 Capture: 5% Discount Rate

Table 17: Case 1: ASC PF Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate & -20% TIC

Case 2: PF Oxy-combustion Power Plant with CO_2 Capture

Table 18: Case 2: ASC PF Oxy-Combustion Power Plant with CO_2 capture: 10% Discount Rate

Table 19: Case 2: ASC PF Oxy-Combustion Power Plant with CO_2 capture: 10% Discount Rate & +100% Fuel Price

Table 20: Case 2: ASC PF Oxy-Combustion Power Plant with CO_2 capture: 10% Discount Rate & -100% Fuel Price

Table 21: Case 2: ASC PF Oxy-Combustion Power Plant with CO_2 capture: 5% Discount Rate

Table 22: Case 2: ASC PF Oxy-Combustion Power Plant with CO_2 capture: 10% Discount Rate & -20% TIC

The Effect on Cost of Electricity Resulting from CO_2 Capture on ASC PF Plant

The results of the above simulations and sensitivities are summarised in Table 23.

As expected the investment cost and the fuel price dominate the CoE. The reference air-fired case for the ASC PF plant resulted in a calculated CoE of 4.98 US Cents/kWh.

This study is 44.2% net LHV, 1152 Euros/kW_e gross (1384 US\$/kW_e gross) and IEA GHG economic parameters give 4.98 US Cents/kWh. Previous studies[33] quote for 41.2% LHV (assumed net) and 1150 US\$/kW_e (assumed gross) a CoE of 4.4 US Cents/kWh. This difference in CoE is due to TIC, thermal efficiency and economic assumptions in fuel price, load factor and capital charges. Reference[34] for 46% LHV (assumed net) and 1020 US\$/kW_e (assumed gross) gives 3.7 US Cents/kWh based on the same economic parameters based as this study.

The CoE associated with the ASC PF with CO_2 capture plant is calculated at 7.28 USc/kWh assuming no CO_2 credits. Therefore the penalty in CoE associated with CO_2 capture is 2.3 US Cents/kWh.

The cost of avoidance of CO_2 emission is calculated at 36 US\$/t CO_2 (30 Euros/t CO_2). Under the recently introduced (2005) EU Emissions Trading Scheme Directive (EU ETS), utilities would need to sell their CO_2 emission quota at a price higher than that associated with capture. However, until such time as emission trading is fully established and penalties enforced, there can be no true market price placed on avoided CO_2 emissions.

Therefore, the only accurate comparison that can be currently undertaken is the difference in cost per kWh sold which results between equivalent technologies with and without CO_2 capture.

As shown in Table 23 fuel cost is seen to influence the penalty in CoE associated with CO_2 capture from 2.3 US Cents/kWh to 2.0 US Cents/kWh. However, the discount rate from 10% to 5% resulted in the lowest penalty in CoE associated with CO_2 capture at 1.7 US Cents/kWh. This discount rate 5% reduction was also more significant in influencing the CoE associated with CO_2 capture than the effect of a 20 % reduction in Total Investment Cost.

5.2 NGCC Oxy-Combustion Power Plant with CO_2 Capture

In order to evaluate the economics of the NGCC oxy-combustion power plant with CO_2 capture, Investment costs and O&M costs were generated for:

- Case 3: 400MW$_e$ gross NGCC conventional air-fired power plant without CO_2 capture; and
- Case 4: 575MW$_e$ gross NGCC oxy-combustion power plant with CO_2 capture.

The costs for the latter were based on the costs of a mature product with the development costs included in the final figure.

Investment Costs

Tables 24 and 25 illustrate the installed capital costs associated with Case 3 and Case 4 respectively (Note: The estimated accuracy of figures quoted is ±25%).

For each of the cases the following major power plant units have been identified and the costs associated with the respective unit expressed.

- Case 3: 400MWe MW$_e$ Gross Typical NGCC Air Fired Power Plant Without CO_2 Capture

Unit 1000	NGCC Power Plant

- Case 4: 575 MW$_e$ Gross NGCC Oxy-Combustion Power Plant With CO_2 Capture

Unit 1000	NGCC Power Plant
Unit 2000	ASU
Unit 5000	CO_2 Purification & Compression

The outline equipment list for each of the above cases is given in Tables 26 and 27.

The direct field costs (DFC)[1] are defined as:

Direct Materials:	Including equipment and bulk materials.
Construction:	Including mechanical erection, instrument and electrical installation, civil works and where applicable buildings and site preparation.
Other costs	Including temporary facilities, solvents, catalysts, chemicals, training, commissioning, start up costs and spare parts.
EPC services	including contractors home services and construction supervision and freight.

The indirect field costs (IFC) are:

Contingency:	10% of the total installed costs
Fees:	2% of the total installed costs

- Owner Costs: 5% of the total installed costs

Note as discussed earlier, owner costs are inherently site specific. The IEA GHG economic criteria[27] for owners costs propose 5% for NGCC plant and this has been used in the economic analysis presented within this study. The 5% owners costs figure is viewed by Alstom Power as conservative and may in fact rise to some 13% depending on the site.

The summing of the above generates the Total Investment Cost (TIC) for each of the major power plant units. (i.e. TIC = DFC + IFC).

Table 28 shows the percentage contribution of each unit to the overall Total Investment Cost for Case 3 and Case 4.

The Total Investment Cost for Case 3 and Case 4 are expressed as a function of their gross and net power output to arrive at a the Specific Investment Cost associated with each plant; namely:

- Case 3:	452 Euro/kW$_e$ Gross	(542 US$/kW$_e$ Gross)*
	466 Euro/kW$_e$ Net	(559 US$/kW$_e$ Net)*
- Case 4:	954 Euro/kW$_e$ Gross	(1144 US$/kW$_e$ Gross)*
	1246 Euro/kW$_e$ Net	(1495 US$/kW$_e$ Net)*

*Note: the above figures are based on exchange rate of 1 Euro = 1.2 US$.

Operation and Maintenance Costs

The operating and maintenance cost figure is formed from a fixed cost and variable cost.

The variable cost element in the total operating and maintenance cost figure is derived from the consumables associated with each plant. The consumables associated with Case 3 and Case 4 are listed in Table 29 and are based on 7446 operating hours in a year (85% load factor following commissioning and start up compared to 90% specified by IEA GHG[27]).

The fixed cost contribution to the total operating and maintenance cost figure for both plants are shown in Table 30. These costs are calculated using:

- Operator cost and supervision (50k US$/yr per operator)
- Administration and general overheads (30% of direct labour)
- Maintenance 4% of installed cost per year compared to 2% for NGCC[27]. (However, it is believed this percentage should be equal for both the NGCC plant and the ASC PF Plant case as the actual costs are reflected in the difference in installed costs.)

Therefore the total operation and maintenance costs calculated for the capture plant are:

- Case 3: 65.2 M Euro per year (78.2 M US$ per year)
- Case 4: 98.7 M Euro per year (118.4 M US$ per year)

CO_2 Abatement Costs for NGCC Plant

To generate the CoE and the Cost of CO_2 avoidance associated with the NGCC base case and the NGCC CO_2 capture case, the figures generated above were entered into IEA's GHG standard economic spreadsheet version 01, February 2005.

The sensitivity of the model results to varying fuel cost (±100%), discount rate (10% and 5%) and Total Investment Cost (-20%) were then evaluated.

The results of the spreadsheet models are shown in:

Case 3: NGCC Power Plant without CO_2 Capture

Table 31:	Case 3: Typical NGCC Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate
Table 32:	Case 3: Typical NGCC Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate & +100% Fuel Price
Table 33:	Case 3: Typical NGCC Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate & -100% Fuel Price
Table 34:	Case 3: Typical NGCC Air Fired Power Plant Without CO_2 Capture: 5% Discount Rate
Table 35:	Case 3: Typical NGCC Air Fired Power Plant Without CO_2 Capture: 10% Discount Rate & -20% TIC

Case 4: NGCC Oxy-Combustion Power Plant with CO_2 Capture

Table 36:	Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture: 10% Discount Rate
Table 37:	Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture: 10% Discount Rate & +100% Fuel Price
Table 38:	Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture: 10% Discount Rate & -100% Fuel Price
Table 39:	Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture: 5% Discount Rate

Table 40: Case 4: NGCC Oxy-Combustion Power Plant With CO_2 Capture: 10% Discount Rate & -20% TIC

The Effect on Cost of Electricity Resulting from CO_2 Capture on NGCC Plant

The results of the above simulations and sensitivities are summarised in Table 41.

The reference case for the typical NGCC Air Fired Power Plant Without CO_2 Capture resulted in a calculated CoE of 3.35 US Cents/kWh.

The CoE associated with the NGCC Oxy-Combustion Power Plant With CO_2 Capture is calculated at 6.13 US Cents/kWh assuming no CO_2 credits. Therefore the penalty in CoE associated with CO_2 capture is 2.8 US Cents/kWh.

The cost of avoidance of CO_2 emission is calculated at 78 US$/t CO_2 (65 Euros/t CO_2).

A fuel cost variation of -100% is seen to influence the penalty in CoE associated with CO_2 capture reducing from 2.8 US Cents/kWh to 2.3 US Cents/kWh. However, the discount rate from 10% to 5% resulted in the lowest CoE associated with CO_2 capture at 2.2 US Cents/kWh. This discount rate 5% reduction was also more significant in influencing the CoE associated with CO_2 capture than the effect of a 20 % reduction in Total Investment Cost.

5.3 Summary of Economic Results for CO_2 Capture

The penalty in the cost of electricity as a result of CO_2 capture is slightly lower for an ASC PF plant (2.3 US Cents/kWh) than the NGCC plant (2.8 US Cents/kWh) assuming no CO_2 tax credits and baseline economic parameters.

Figure 25 shows the cost of electricity as a function of the fuel price for both plants with and without CO_2 capture.

The fuel price applied for the purposes of the study was 3.0 US $/GJ for natural gas and 1.5 US $/GJ for coal. Whilst the coal price has remained relatively stable, the present gas price is somewhat higher (current figures suggest 5.1 US $/GJ based on UK NBP gas price (Jan 2005) of 30 pence/therm [35]: exchange rate £ to US$ 1 to 1.8). The graph shows that a gas cost in excess of 5.5 US $/GJ would result in the cost of electricity generated from NGCC plant being greater than that of electricity generated via its ASC PF counterpart.

6. FUTURE DEVELOPMENTS

6.1 ASC PF Plant with CO_2 Capture

For a realistic chance of success, the development of a purpose designed oxy-combustion PF boiler must follow a stepwise procedure from research through prototype development to large-scale demonstration. To offset the considerable decrease in efficiency, the plant design must take advantage of the continuous development of higher efficiency units[36] such as advanced supercritical pulverised coal technology that produces less CO_2 per megawatt generated. (Figure 26)

Research programmes aimed at finding cost effective ways to recover CO_2 from an oxygen fired fossil fuel combustion process are underway in Canada, USA, Australia and more recently in Europe (as part of the EC 6th Framework Programme ENCAP) as well as the UK DTI 4th Call Cleaner Coal Technology R&D Programme for retrofit options for Advanced Supercritical PF CO_2 Capture Plant.

The oxy-combustion process concept lends itself for accommodating a staged approach to the implementation of CO_2 capture into a new build power plant as 'capture ready' plant specifically designed for subsequent easy retrofit of suitable CO_2 capture plant.

This novel integrated multi-pollutant control scheme could potentially have some economic advantages[37] when compared to a conventional air-fired PF plant and importantly provides a stepping stone to the full CO_2 capture plant.

The financial advantages perceived from the oxy-combustion boiler are the elimination of the FGD, DeNOx and Hg plant due to the possibility of total removal of virtually all normal emission of SO_x, NO_x, HCl, Hg and dust from the compressed net product CO_2. Furthermore the enhanced heat transfer from the CO_2 rich combustion products of oxy-combustion allow the potential for a reduction in boiler's heating surface area and thus reduced capital outlay of the boiler. However, these benefits have to be offset against the substantial outlay for an ASU and a CO_2 treatment system.

As part of the development of the ASC oxy-combustion 'capture ready' plant, the key areas required to be addressed are: -

- The successful demonstration of full-scale burners under conditions of combustion in oxygen and recycle flue gas building on the results of previous pilot-scale test work[2, 29].
- Full appraisal of the slagging and fouling nature of the ash arising from the oxy-combustion process and their impact on boiler heating surface arrangement and choice of boiler materials.
- The impact of radiant and convective heat transfer [3] from the novel flue gas composition on boiler plant design.
- General issues associated with materials, corrosion, build-up of impurities, and requirements for plant start-up, shutdown, oxy-combustion boiler control systems, recycle flue gas purging and the effect of plant trips on the boiler system.

- Further process optimisation, integration and automation to reduce cost, improve performance and increase energy efficiency.

An important milestone in realisation of the oxy-combustion technology for carbon Capture and Storage (CCS) is the European 30MW_{th} Oxyfuel PF Pilot Plant[38] planned for operation in 2008. The pilot plant test programme would aim to address the key areas outlined above and provide the design basis for scale-up of the technology to European power generation scale of 600$MW_{e.}$

In order to attribute sensible costs to an oxy-combustion process for comparison with other CO_2 capture technologies the oxy-combustion boiler plant considered within this study is recognised as somewhat conservative in its design (the cost of a basic boiler typically accounts for around 25% of the cost of a new power plant).

Based essentially on current air fired plant technology, albeit applied to oxy-combustion, it is far from a radical, innovative purpose built oxy combustion boiler. It is however recognised that once 'capture ready' technology is established and accepted in the marketplace, more radical steps to refine the plant can be attempted and that such steps may considerably reduce cost and complexity. More radical areas of oxy-combustion boiler development for CO_2 capture currently recognised are:

- Combustion process intensification: Oxy-combustion burners/slagging combustor designs aimed at combustion of PF in essentially pure O_2 or in a mixture of O_2 and alternative recycle flue gas rates. Potentially this could lead to significant reductions in the overall size of the boiler plant [39] resulting in significant savings in both capital and operating costs when compared to conventional power generation plant.

- Plant start up based purely on oxygen combustion, which would remove the need for initial air firing and shut down and the associated equipment and additional control systems this currently generates.

A further long term oxy-combustion possibility is high pressure oxy-combustion burners with direct water quench for high carbon content fuels, such as coals, bitumen or petroleum coke. Potentially, it may be possible to produce fully oxidised combustion products, with a steam content of typically 80-90% at pressures of, say, 250-500 bar, and temperatures of 600-1200°C. The combustion products would contain predominantly CO_2 with fully oxidised impurities from the coal such as SO_2, NO_x, HCl. Advanced steam turbines based on gas turbine technology would be required for power generation. CO_2 rich gas would be produced from the steam condenser at the back end of the plant. Theoretical efficiencies with CO_2 capture of over 50% might be possible with such a system. Effective ash removal systems would need to be derived and significant efforts required with regards to fouling and corrosion prevention. High pressure direct oxidation systems with water quench have been demonstrated[40] for megawatt production of steam using both hydrogen and natural gas as fuel burning almost stoichiometrically with oxygen.

6.2 NGCC Oxy-Combustion Plant with CO_2 Capture

Principal development requirements include combustion and materials technology together with demonstration of system components into a validated power plant to demonstrate operability of system and components.

Near stoichiometric combustion with CO_2 will require non-standard design to ensure that mixing of oxygen takes place properly and appropriate cooling and flow dilution takes place with CO_2 distribution in the burner. Development testing will be required to validate design assumptions.

Materials test programs are currently in place to screen typical materials found in high temperature regions of the gas turbine and HRSG. Some experience has been gained in the nuclear power industry with different materials.

Turbomachinery is feasible using conventional design techniques but requires validation of aeromechanics, aero-cooling, lubrication and control systems. Blade aero-mechanics subject to forced vibration in higher density CO_2 may require aspect ratio tuning to avoid flutter response in both compressor and turbine high aspect ratio blade rows. Turbine blade fracture mechanics may require special attention given the higher temperature gradient between gas and coolant described earlier. Special lubrication fluid may be required to avoid contamination and consequent degradation.

Compressor operability would need to be understood empirically since the stage matching requirements would change as a result of different gas properties.

This programme would be a major undertaking and constitute a novel engine development programme for which there is no precedent. Cost is therefore difficult to predict.

Further long-term advancements might include increased cycle pressure ratio with reheat, component efficiency and advanced blade cooling (utilising steam). These additions will add growth potential and increase both cycle efficiency and specific power. This should form the basis for further study in order to realise the full potential of the Gas-ZEP concept which may approach 50% combined cycle efficiency without change to ASU technology. Clearly the incentive exists for development of a non-cryogenic ASU in order to reduce the power associated with oxygen production.

6.3 Air Separation

Current knowledge of cryogenic oxygen plant design and components such as air compressors, drive motors and cold equipment fabrication leads to the conclusion that the large size oxygen plants required for the PF and NGCC oxy-combustion systems can be constructed now at capacities of currently up to about 7000 tonne/day. Cryogenic air separation is a mature technology and major improvements in either efficiency or capital cost cannot be expected in the future.

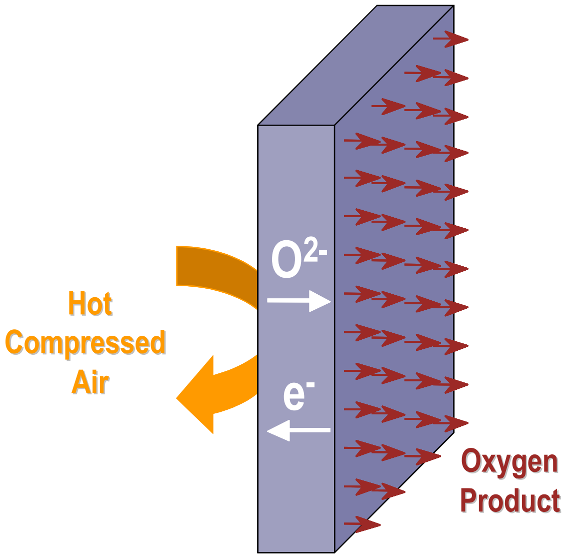


Figure 27: Mixed conducting Ion Transport Membrane

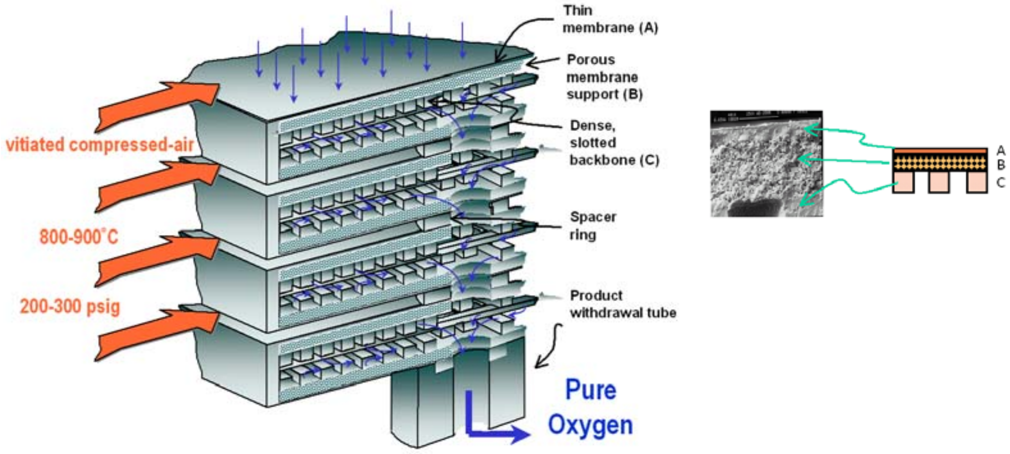


Figure 28 : Schematic of ITM Oxygen Planar Supported Membrane Device

Ion transport membranes (ITM or OTM) are being developed for production of oxygen using high temperature mixed oxide materials such as perovskites which will simultaneously conduct electrons and oxygen ions (see Figure 27) when there is a difference in the activity such as a difference in oxygen partial pressure across the material. The ITM system, as configured by Air Products, consists of flat square section hollow plates arranged sequentially on a hollow collector tube Figure 28 and mounted in a pressure vessel through which heated air from the gas turbine flows over the “fins”. Oxygen diffuses through the membrane surfaces due to a pressure gradient and pure oxygen is collected, cooled and compressed to delivery pressure.

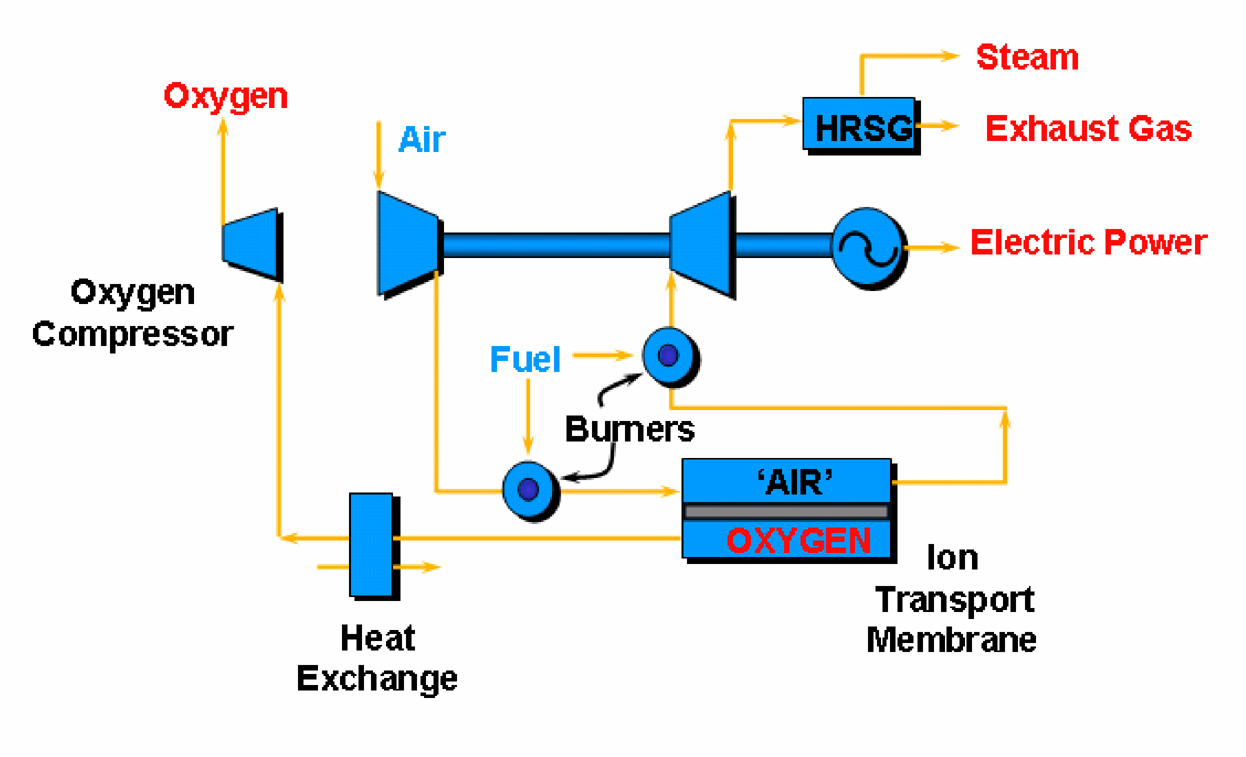


Figure 29: Basic Scheme for Integration of an ITM Oxygen Air Separation Unit into a Gas Turbine Power Cycle

The integration of the ITM with a gas turbine is shown with integration into a coal fired integrated gasification combined cycle (IGCC) system in Figure 29. The air from the compressor section of the gas turbine is heated to the required operating temperatures of the ITM (800°C to 900°C) and passed through the ITM tubes. The air flows between the fins, oxygen diffuses into the fins and then to the central collector tube. The ITM tubes are sealed into metallic collector headers which are manifolded together within the pressure vessel.

Allam et al [41] studied the oxyfuel conversion of boilers and heaters on a refinery site using an ITM Oxygen system to produce the oxygen. This showed that when integrated into the current steam system the ITM Oxygen system resulted in a cost of CO_2 capture around half that of the traditional cryogenic ASU. These advantages would be most apparent in this study with the PF oxyfuel system since the gas turbine HRSG could be used for condensate and boiler feed water heating duties within the boiler steam cycle. With the NGCC oxyfuel system in this report the integration with the ITM Oxygen system would most likely be achieved by combining the HRSG duties of the NGCC gas turbine and the ITM Oxygen gas turbine.

The integration of the ITM with a gas turbine is shown with integration into a coal fired integrated gasification combined cycle (IGCC) system in Figure 30. Here the fuel for the gas turbine is provided using gasified coal.

One interesting feature of this type of integration arises when producing CO_2-free power by burning hydrogen, by shifting the syngas produced in the gasifier, or a natural gas reformer, and then capturing the CO_2 from this stream. In this case there is no separate N_2 stream available for hydrogen dilution, but the hydrogen would be burnt in a nitrogen/oxygen stream from which part of the oxygen had been removed. The maximum oxygen product produced from a gas turbine will depend on the design of the combustor considering flammability limits and flame stability when burning pure hydrogen in O_2 depleted air.

In order to take advantage of the potential cost reduction, both capital and operating, inherent in the use of ITM oxygen production systems, current systems analysis would suggest the integration of ITM with a gas turbine as shown Figure 30. This would imply conversion of part of the coal in a gasification system with shift conversion and CO_2 removal to produce pure hydrogen as fuel for the gas turbine. Hydrogen can undergo stable combustion with a much lower oxygen concentration than natural gas which increases the oxygen recovery from a given size of gas turbine with a given maximum air flow.

It has been proposed to use the ITM principle in a fuel rich burner rather than an oxygen rich burner configuration. The equilibrium partial pressure on the combustion side in a reducing atmosphere would be so low that diffusion of oxygen could take place into a high pressure stream from air at atmospheric pressure. The power producing cycle would then involve expansion of the high pressure CO_2-rich combustion gas with separation of excess fuel gas for recycle to the burner. Burners of this type may become available in the 2020 time frame[42].

6.4 CO_2 Compression and Purification

CO_2 compressors of the necessary single train size (50 to 100 MW) can be supplied now and constructed using proven components and design procedures. Drive systems using steam turbines, or two pole motors are commercially available.

The low temperature CO_2 purification system, although never before built, is made up of proven equipment and is designed using known technology procedures and data.

The challenge in the design of the CO_2 compression system is producing a mechanical design that can accommodate the supercritical properties of CO_2 at the pressure required for CO_2 capture processes.

The critical pressure of pure CO_2 is around 75 bar. With impurities this pressure increases. Supercritical fluid in the dense phase will not be suitable for compression within a compressor and should be pumped. However, compression is required to reach this point. The compression system design needs to accommodate the change from gas, or gas-like properties, to dense fluid, incompressible liquid-like properties, with the change happening due to cooling in a condenser, rather than within the compression machinery. Also, the compression train should be designed for the highest possible efficiencies. The current design fulfils these objectives by using an adiabatic high pressure system which avoids gross changes in CO_2 physical properties while recovering heat of compression for condensate and boiler feed water heating.

CO_2 compression trains for the pressures required by CO_2 capture projects are relatively rare at the moment. Therefore, the compression equipment, wheel sizes, etc, are not best suited to this duty. Compressor manufacturers will, once the market for such compressors opens, optimise their offerings for CO_2 compression and will have a better understanding of the transition from compression to pumping. Therefore, costs should reduce and efficiencies improve.

Where it is possible to use adiabatic compression, and use the low grade heat for condensate or boiler feedwater preheating, further developments in the compression equipment will be required to accommodate the higher compressor discharge temperatures.

All of these scenarios require large amounts of motive force to drive the compressors. If the process into which these compressors are integrated allows, steam turbines could be used to drive the compressor trains. At the moment the large steam turbines used in the power generation process are 10% or so higher in efficiency than the small turbines that would be needed to drive the CO_2 compression train, negating the savings in motor and gear box losses by using steam turbine drives. However, should the efficiency of the smaller steam turbine drives increase then the CO_2 compression train would be made simpler by the use of steam turbine drives.

7. CONCLUSIONS AND RECOMMENDATIONS

Plant Performance:

The study has shown that for an advanced supercritical PF plant designed to capture CO_2 the penalty in terms of cycle efficiency loss is approximately 9 percentage points compared to a similar sized plant with no CO_2 capture. (I.e. a plant net efficiency reduction of 44.3% to 35.4% LHV).

Careful consideration of integration of feed water heating with available heat from the flue gas and oxy-combustion plant components such as the air separation unit and CO_2 compression unit has contributed significantly to reducing the cycle efficiency penalty of the advanced supercritical PF oxy-combustion plant with CO_2 capture.

For the case of a NGCC plant with CO_2 capture the penalty in terms of cycle efficiency loss is calculated at approximately 11 percentage points compared to a similar sized NGCC plant with no CO_2 capture (i.e. a plant net efficiency reduction of 56.0% to 44.7%, LHV).

Economics

The cost associated with the advanced supercritical PF power plant with CO_2 capture compared to the study reference case advanced supercritical plant without CO_2 capture is equivalent to a cost delta of 2.3 US Cents/kWh.

For the NGCC plant the cost associated with the capture plant compared to the cost of the plant without capture is equivalent to a cost delta of 2.8 US Cents/kWh.

These costs were based on coal price of 1.5 US$/GJ and a gas price of 3.0 US$/GJ with no allowance for CO_2 credits.

Recommendations

Whilst illustrating the envisaged potential, it is recognised that oxy-combustion technology applied to both plant types requires further development in key areas in order introduce the technology successfully to the marketplace. The specific areas requiring refinement are identified as

Plant start up and control systems;

Burner and flame characterisation;

Materials issues

None of these development areas are likely to represent technical show-stoppers.

To offset the considerable efficiency penalties, it is recommended that ASC PF plant must take advantage of the continuous development of higher efficiency units that produce less CO_2 per Megawatt generated.

Further radical improvements in oxygen production technology using high temperature ceramic membranes currently at an advanced state of development will lead to significant reduction in oxygen costs compared to the current cryogenic air separation systems.

The application and integration of ITM technology to NGCC and PF plant is recommended as a subject for further study.

Similarly, developments in air-fired gas turbine technology materials and blade cooling techniques need to be continually incorporated into the oxy-combustion gas turbine design.

A "stepwise" approach for the case of the ASC PF plant from research through prototype development, to large scale demonstration is suggested. A 30 MW_{th} PF oxy-combustion pilot plant[38] is already planned for operation in 2008 and will represent an important milestone in technology development.

A recognised significant "stepping stone" to a fully operational CO_2 capture plant is the "capture ready" plant. This novel integrated pollutant control scheme could potentially have economic advantages compared with conventional air fired PF plant and should be investigated on this basis. The attractiveness of a "capture ready" plant is that it may provide a pathway which addresses the technical issues associated with oxy-combustion and competes with existing PF clean coal technology systems in its own right, without committing vendors entirely to CO_2 capture.

Radical areas of oxy-combustion boiler development recognised within this study are:

- Combustion process intensification (i.e. very little, if any, flue gas recycle).
- Start up based on oxygen combustion exclusively.

It is recommended that the feasibility of these options be explored accordingly.

8. ACKNOWLEDGEMENTS

Mitsui Babcock on behalf of all the project partners wishes to thank the International Energy Agency Greenhouse Gas R&D Programme for the funding of this study.

9. REFERENCES

[1] IEA Greenhouse Gas R&D Programme. *Potential for Improvements in Gasification Combined Cycle Power Generation with CO_2 Capture,* IEA Greenhouse Gas R&D Programme Report PH4/19, May 2003.

[1A] IEA Greenhouse Gas R&D Programme. *Potential for improvement in Power Generation with Post-Combustion capture of CO_2,* IEA Greenhouse Gas Report PH/33, November 2004.

[2] Mitsui Babcock, Air Products plc, University of Ulster and University of Naples. *Pulverised coal combustion system for CO_2 capture*, CEC Joule II Programme Contract No JOU2-CT92-0062.

[3] Wilkinson, M B, Boden, J C, Panesar, R S and Allam, R J. *A study on the capture of carbon dioxide from a large refinery power station boiler by conversion to oxyfuel operation*, International Conference on Greenhouse Gas Control Technologies, Cairns, Australia, 13-16 August 2000.

[4] Wilkinson, M B, Boden, J C, Panesar, R S and Allam, R J. *CO_2 capture via Oxyfuel Firing: optimisation of a retrofit design concept for a refinery power station*, First National Conference on Carbon Sequestration, 15-17 May 2001, Washington DC.

[5] Wilkinson, M B, Simmonds M, Allam, R J and White, V. *Oxyfuel conversion of heaters and boilers for CO_2 capture,* Second National Conference on Carbon Sequestration, 5-8 May 2003, Washington DC.

[5A] RR Pierce. *Estimating Acid dewpoints in stack gases.* Chemical Engineering 1977 April.

[6] Mathieu, P et al. *Behaviour of Gas Turbine Cycles using CO2/H2O/Ar as the working fluid*, International Symposium on CO_2 Fixation and Efficient Utilization of Energy, Kyoto, Japan, 1994.

[7] Mathieu, P et al. *The Use of CO_2 Gas Turbines in Combined Cycle Plants*, ECOS 95 Conference, Istanbul, Turkey, 1995.

[8] Mathieu, P. *The use of CO_2 Gas Turbines*, Power-Gen Europe 94 Conference, Cologne, Germany, 1994.

[9] Mathieu, P. *Zero Emission Power Generation : a dream or reality?*, Newsletter published by the IEA Greenhouse Gas R&D Programme, 1st annual workshop on Zero Emission Power Plants, organized in Liège (Begium) Greenhouse Issue No. 30, 1997.

[10] Mathieu, P and Bolland, O. *Comparison of two CO_2 removal options in Combined Cycle Power Plants*, FLOWERS '97, Florence (Italy), July 1997.

[11] Mathieu, P and Iantovski, E. *Highly efficient zero emission CO_2 based power plants*, Energy Conversion and Management, 1997,Vol 38 ,suppl 1, pp S141-146.

[12] Mathieu, P. *Report on the 2nd annual workshop on Zero Emission Power Plants*, organized in Liège (Begium)", Greenhouse Issue No. 34, 1998.

[13] Mathieu, P. *Presentation of an innovative zero emission cycle for mitigation of the global climate change*, Int Jl of Applied Thermodynamics, First issue, vol 1, pp 21-31, 1998.

[14] Mathieu, P and Nihart, R. *Sensitivity analysis of the Matiant cycle,* Energy Conversion and Management, Vol 40 (15-16 issues), pp 1687-1700, 1999.

[15] Mathieu, P. *Zero Emission MATIANT cycle*, Transactions of the ASME, Journal of Engineering for Gas Turbines and Power Plants, pp. 116-121, 1999.

[16] Mathieu, P. *AMPERE* Report, Belgian Ministry of Economic Affairs, 2000 (contribution on advanced coal and gas technologies and on nuclear power).

[17] Mathieu, P, Houyou, S, and Nihart, R. *Techno-economic comparison of different options of very low CO_2 emission technologies*, 5th International Conference on GreenHouse Gas Control Technologies, Cairns (Australia) 2000.

[18] Mathieu, P and Mathieu, P P. *Zero Emission Technologies- a key strategic option for climate change mitigation*, Mathieu Greenhouse Issue No. 63, 2002.

[19] Mathieu, P. *Towards a world based on near zero CO_2 emission energy systems*, International Conference on Clean Coal Technologies for our Future, Sardinia, 2002.

[20] Mathieu, P, *Cycles leading to a full CO_2 sequestration*, Experts conference on "Cycle for low CO_2 production", Cranfield University (UK), 2003.

[21] Mathieu, P. *Zero Emission from coal*, Energy choices for Europe, 2003, European Parliament, Brussels.

[22] Mathieu, P. *Overview of CO_2 capture technologies*, CAN Europe (Climate Action Network, NGO) workshop on CO_2 capture and storage and Hydrogen, Brussels, 23-24 Sept 2003.

[23] Mathieu, P. *Mitigation of CO_2 emission using low and near zero emission power plants*; Int Jl on Energy for a Clean Environment, 2003, vol 4 also presented at the 6th International Conference on Technologies and Combustion for a Clean Environment, Porto (Portugal), 2001.

[24] Mathieu, P. *Near zero emission power cycles : performance and costs*, First Int Conf on Industrial Gas Turbines Technologies, Thematic Network CAME-GT, Brussels, 2003.

[25] Mathieu, P and Mathieu, P P. *Economics of climate change mitigation using zero emission technologies*, International Conference on Technologies and Combustion for a Clean Environment, Lisbon (Portugal), 2003.

[26] Bolland, O. *Options for oxy-fuels & pre-combustion decarbonisation cycles*, Presentation for Alstom Cross Segment CO_2 Mitigation Group, October 11th

2002 (http://www.tev.ntnu.no/Olav.Bolland/pdf/Alstom_Power_Oct_2002_Bolland.pdf).

[27] IEA Greenhouse Gas R&D Programme, *Technical and Financial Assessment Criteria,* Revision B May 2000.

[28] Panesar, R S (Mitsui Babcock), Wheeler, R J (Mitsui Babcock) and Allam, R J (Air Products plc). *Oxyfuel Combustion Concept for BP Amoco*, Mitsui Babcock Report No. E/99/059.

[29] Châtel-Pélage, F et al. *A pilot-scale demonstration of oxy-combustion with flue gas recirculation in a pulverized coal-fired boiler*, The 28th International Technical Conference on Coal Utilization & Fuel Systems, 10-13 March 2003, Clearwater, FL.

[30A] Schmidt, W P et al. *Safe Design and Operation of a Cryogenic Air Separation Unit*, AIChE 35th Annual Loss Prevention Symposium, April 2001, TX, USA.

[30B] CGA Publication P-9, *Safe Practice Guide for Cryogenic Air Separation Plants.*

[31] Foresight Link Award (FLA3) Project, *Natural Gas Fired Zero Emission Plant (GAS-ZEP)- Revised Specification for GAS-ZEP Optimum Concept OXY1,* Alstom Power Technology Centre Report No. U21625/009, July 2003.

[32] *Supercritical Steam Cycles for Power Generation Applications*, DTI Technology Status Report, TSR009, January 1999.

[33] IEA Working Party on Fossil Fuels. *Solutions for the 21st Century: Zero Emission Technologies for Fossil Fuels,* Technology Status Report, May 2002.

[34] IEA Working Party on Fossil Fuels. *CO_2 Capture and Storage,* Technology Status Report, June 2000.

[35] UK NBP Gas Price, Power In Europe, Issue 442, 17 January 2005.

[36] Kjaer, S et al. *The advanced supercritical 700°C pulverised coal-fired power plant*, VGB Power Tech 7/2002.

[37] Marin, O et al. *Economic analysis of oxygen-fired coal boilers*, The 28th International Technical Conference on Coal Utilization & Fuel Systems, 10-13 March 2003, Clearwater, FL.

[38] *Vattenfall commits to CCS Plant*, Power In Europe, 23 May 2005.

[39] Thambimuthu, K. *CANMET Energy Technology Centre - CETC's Oxyfuel Combustion Project Brochure*, January 2004.

[40] D'Aquino, R. *Zero Emission Power Generation – Pure Rocket Science*, Chemical Engineering Progress Magazine, August 2003.

[41] R J Allam, C J McDonald, V White, V E Stein, M Simmonds. *Oxyfuel Conversion Of Refinery Process Equipment Utilising Flue Gas Recycle For CO_2 Capture*, Proceedings of the 7th International Conference on Greenhouse

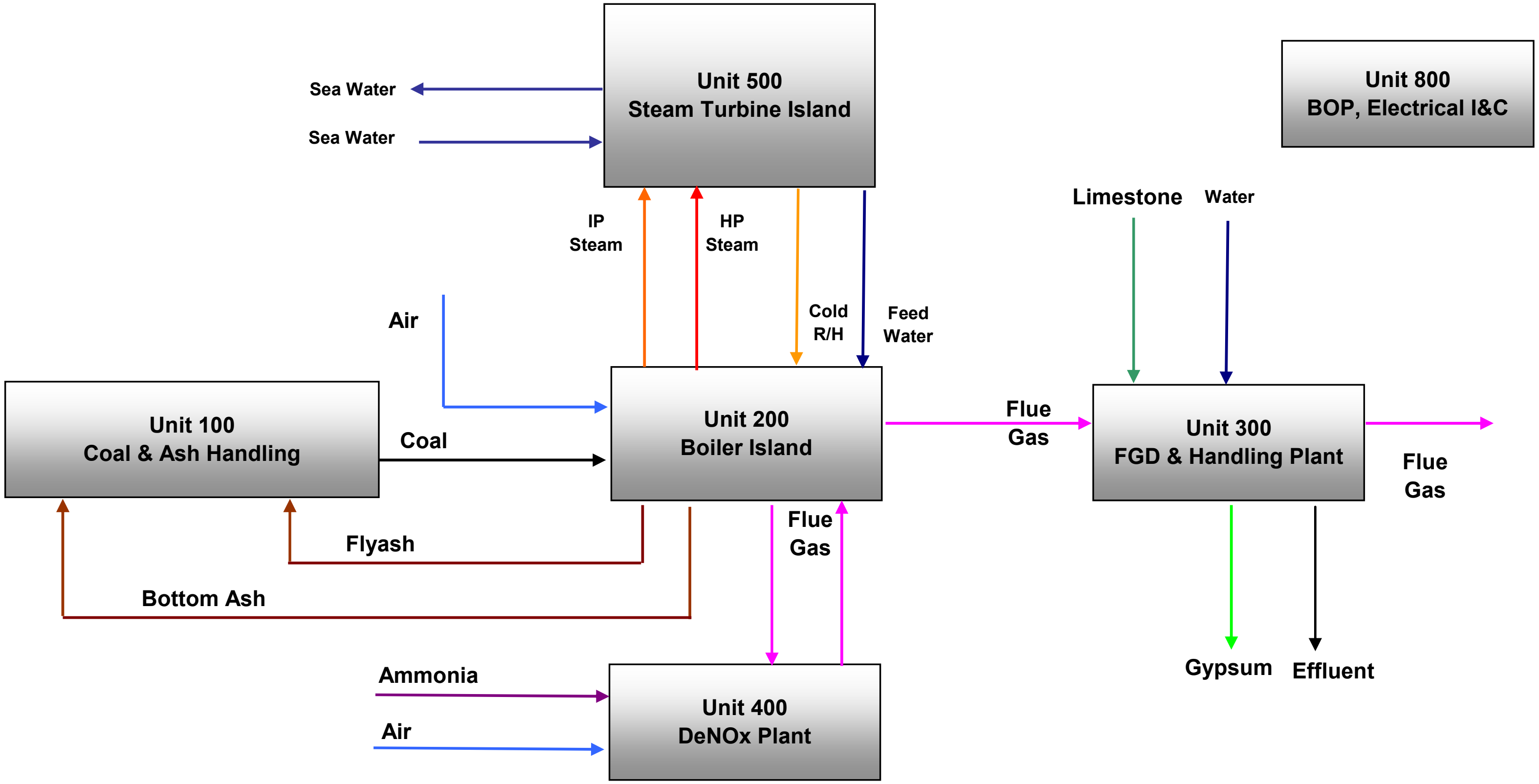
Gas Control Technologies. *Volume 1: Peer-Reviewed Papers and Plenary Presentations*, Vancouver, BC, September 5-9, 2004.

[42] Asen, I K et al. *Process For Generating Power and/or Heat Comprising a Mixed Conducting Membrane Reactor*, United States Patent, Patent No. US 6,237,339 B1, 29th May 2001.

PROCESS FLOW DIAGRAMS

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO_2 Capture from Power Plant
>> PF Case 1 Block Diag DJD Iss 1.xls : DJD Issue 1 09.05.04 Updated Issue 1 09.05.04 <<

ASC PF Case 1 : 740 MWe Gross PF Power Plant : Block Flow Diagram

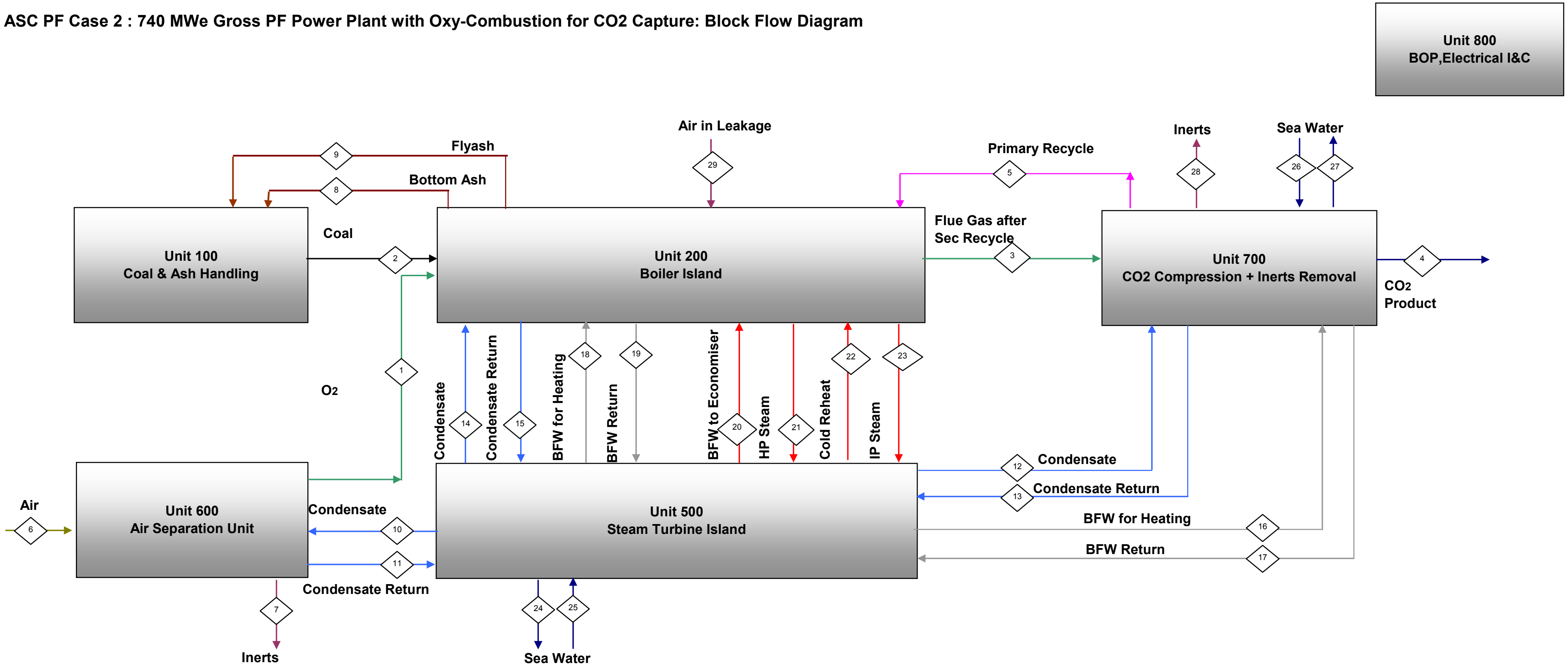


CASE 1 : AIR-FIRED ASC PF POWER PLANT PROCESS BLOCK FLOW DIAGRAM

PFD 1

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO_2 Capture from Power Plant
>> PF Case 2 Block Diag DJD Iss 1.xls : DJD Issue 1 29.04.04 Updated Issue 1 29.04.04 <<

ASC PF Case 2 : 740 MWe Gross PF Power Plant with Oxy-Combustion for CO2 Capture: Block Flow Diagram



CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : PROCESS FLOW BLOCK DIAGRAM

PFD 2

Stream ID		1	2	3	4	5	6	7	8	9	10
Material		Oxygen	Coal	Flue Gas -sec	prod	p rycl	air in	Asu inerts	btm ash	Fly ash	Condensate
model stream No.		Ap A 34	Ec N1b	Ap C 1	Ap C 24	Ap C 4	Ap A 1	Ap A 32	Ec S17	Ec N11	Ec AS1
Total mass flow	kg/s	127.1	58.09	347.65	126.96	154.87	534.68	403.91	1.446	5.767	243.7
- Coal	kg/s	0	45.484	0	0	0	0	0	0	0	0
- Air	kg/s	0	0.0	0	0	0	0	0	0	0	0
- Flue Gas	kg/s	0	0	0	0	0	0	0	0	0	0
- Ash	kg/s	0	7.087	0	0	0	0	0	1.446	5.767	0
- Water	kg/s	0	5.5186	30.41	0	3.964	3.334	0	0	0	243.7
- Steam	kg/s	0	0	0	0	0	0	0	0	0	0
- Argon	kg/s	4.7	0	7.59	0.70911	3.615	6.807	2.022	0	0	0
- Nitrogen	kg/s	2.2	0	33.09	1.655	15.741	401.23	399.064	0	0	0
- Oxygen	kg/s	120.1	0	15.51	0.9779	7.373	122.9	2.798	0	0.00	0
- Carbon Dioxide	kg/s	0	0	259.09	122.764	123.24	0.326	0	0	0	0
- Sulphur Dioxide	kg/s	0	0	1.82	0.839	0.8786	0	0	0	0	0
- Hydrogen Chloride	kg/s	0	0	0.027	0	0.0125	0	0	0	0	0
- Nitric Oxide	kg/s	0	0	0.099	0.01135	0.047	0	0	0	0	0
- Nitrogen dioxide	kg/s	0	0	0.0044	0	0.0018	0	0	0	0	0
- NOx	kg/s	0	0	0.10	0.01	0.05	0	0	0	0	0
Props											
- Phase		Gas	Solid	Gas	liquid	Gas	Gas	Gas	Solid	Solid	Liquid
- Temperature	°C	16	15	110	43	35	9	16	1102	264	29
- Pressure	bara	1.600	-	1.020	110.000	1.020	1.010	1.2	-	-	16.0
- Density	kg/m^3	-	-	-	-	-	-	-	-	-	-
Composition											
O_2	%v/v,wet	94.94	-	5.12	1.05	5.88	20.73	0.608	-	-	-
CO_2	%v/v,wet	0	-	62.20	95.84	71.46	0.04	0	-	-	-
SO_2	%v/v,wet	0	-	0.30	0.45	0.35	0.00	0	-	-	-
H_2O	%v/v,wet	0	-	17.85	0.00	5.62	1.00	0	-	-	-
N_2	%v/v,wet	1.98	-	12.48	2.03	14.34	77.30	99.04	-	-	-
Ar	%v/v,wet	3.03	-	2.01	0.61	2.31	0.92	0.352	-	-	-
NO	%v/v,wet	0	-	0.035	0.01	0.04	0.00	0	-	-	-
NO_2	%v/v,wet	0	-	0.001	0	0.001	0.00	0	-	-	-
molecular weight	kg/kmol	32.1	-	36.70	43.62	39.52	28.86	28.08	-	-	-
Emissions											
NOx	mg/MJ	-	-	66	6	32	-	-	-	-	-
SOx	mg/MJ	-	-	1174	534	559	-	-	-	-	-
Particulates	mg/MJ	-	-	6	0	0	-	-	-	-	-

CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : PROCESS FLOW BLOCK DIAGRAM STREAMS 1 - 10

PFD 2

Stream ID		11	12	13	14	15	16	17	18	19	20
Material		Cond Return	Condensate	Cond Return	Condensate	Cond Return	BFW	BFW return	BFW	BFW return	BFW to Econ
model stream No.		Ec AS2	Ec NN6A	Ec S47B	Ec FG1	Ec FG2	Ec NN9A	Ec S54A	Ec NN9B	Ec S52	Ec NN10
Total mass flow	kg/s	243.7	95	95	69.95	69.95	91.65	91.65	432.71	432.71	524.372
- Coal	kg/s	0	0	0	0	0	0	0	0	0	0
- Air	kg/s	0	0	0	0	0	0	0	0	0	0
- Flue Gas	kg/s	0	0	0	0	0	0	0	0	0	0
- Ash	kg/s	0	0	0	0	0	0	0	0	0	0
- Water	kg/s	243.7	95	95	69.95	69.95	91.65	91.65	432.71	432.71	524.372
- Steam	kg/s	0	0	0	0	0	0	0	0	0	0
- Argon	kg/s	0	0	0	0	0	0	0	0	0	0
- Nitrogen	kg/s	0	0	0	0	0	0	0	0	0	0
- Oxygen	kg/s	0	0	0	0	0	0	0	0		0
- Carbon Dioxide	kg/s	0	0	0	0	0	0	0	0	0	0
- Sulphur Dioxide	kg/s	0	0	0	0	0	0	0	0	0	0
- Hydrogen Chloride	kg/s	0	0	0	0	0	0	0	0	0	0
- Nitric Oxide	kg/s	0	0	0	0	0	0	0	0	0	0
- Nitrogen dioxide	kg/s	0	0	0	0	0	0	0	0	0	0
- NOx	kg/s	0	0	0	0	0.00	0	0	0	0	0
Props											
- Phase		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
- Temperature	°C	83	29	155	29	93	165	206	165	180	270
- Pressure	bara	16	16	16	16	13	339	339	339	335	329
- Density	kg/m^3	-	-	-	-	-	-	-	-	-	801.45
Composition											
O_2	%v/v,wet	-	-	-	-	-	-	-	-	-	-
CO_2	%v/v,wet	-	-	-	-	-	-	-	-	-	-
SO_2	%v/v,wet	-	-	-	-	-	-	-	-	-	-
H_2O	%v/v,wet	-	-	-	-	-	-	-	-	-	-
N_2	%v/v,wet	-	-	-	-	-	-	-	-	-	-
Ar	%v/v,wet	-	-	-	-	-	-	-	-	-	-
NO	%v/v,wet	-	-	-	-	-	-	-	-	-	-
NO_2	%v/v,wet	-	-	-	-	-	-	-	-	-	-
molecular weight	kg/kmol	-	-	-	-	-	-	-	-	-	-
Emissions @ 6%O_2 Dry											
NOx	mg/MJ	-	-	-	-	-	-	-	-	-	-
SOx	mg/MJ	-	-	-	-	-	-	-	-	-	-
Particulates	mg/MJ	-	-	-	-	-	-	-	-	-	-

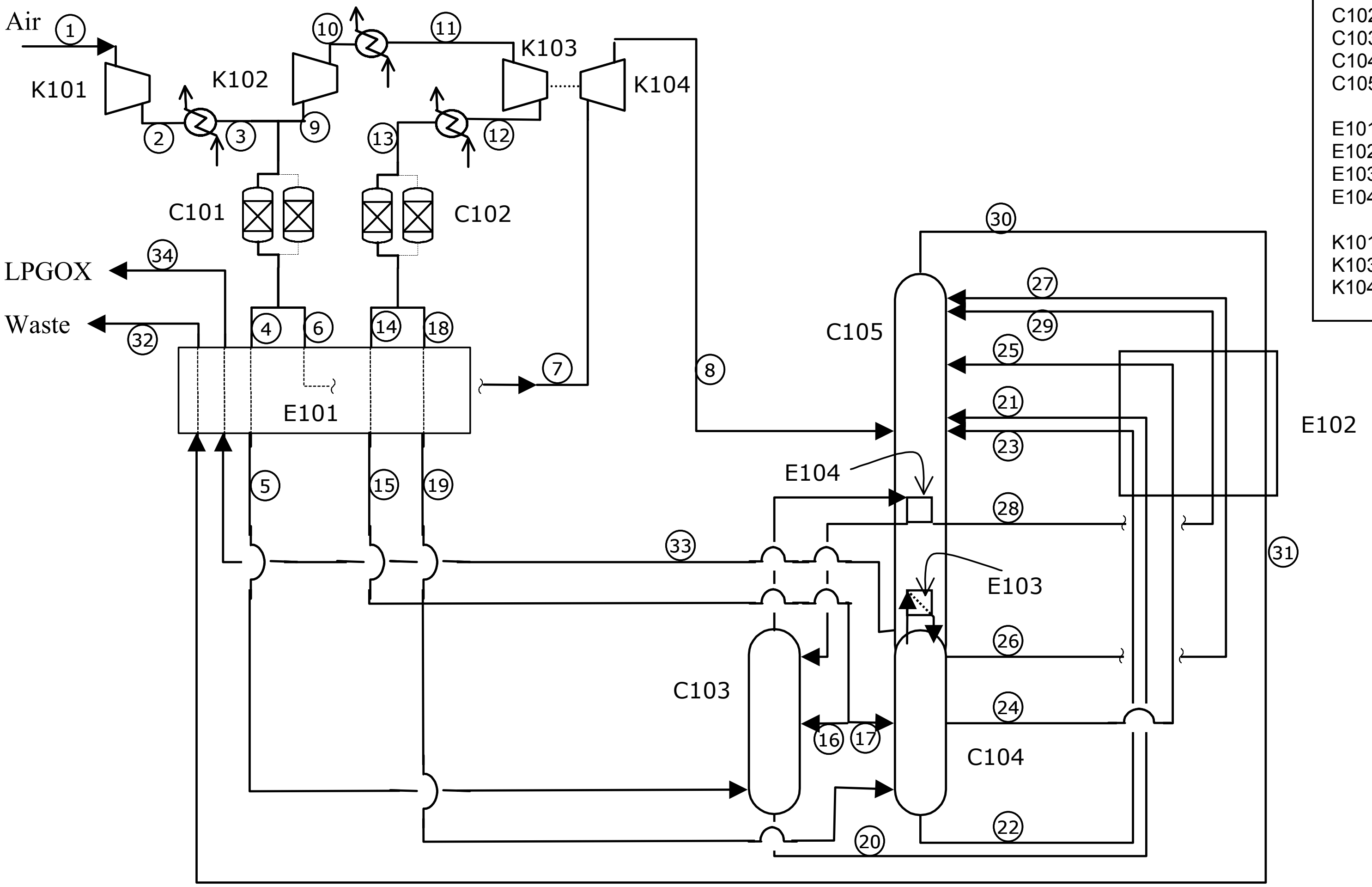
CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : PROCESS FLOW BLOCK DIAGRAM STREAMS 11 - 20

PFD 2

Stream ID		**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**
Material		HP Steam	Cold RH	IP Steam	Cond Sea water in	Cond Seawater out	Comp Sea water in	omp Sea water ou	CO2 Inerts	Air in leakage
model stream No.		Ec S24	Ec S26	Ec NN3	Ec Utility	Ec Utility	Ap CO 12&6	Ap CO 13&7	Ap CO 11	Ec S16C/S13C
Total mass flow	kg/s	516.536	410.807	410.807	20891	20891	2975.2	2978.9	38.6	18.8
- Coal	kg/s	0	0	0	0	0	0	0	0	0
- Air	kg/s	0	0	0	0	0	0	0	0	0
- Flue Gas	kg/s	0	0	0	0	0	0	0	0	0
- Ash	kg/s	0	0	0	0	0	0	0	0	0
- Water	kg/s	0	0	0	20891	20891	2975.2	2975.2	0	0.1167
- Steam	kg/s	516.536	410.807	410.807	0	0	0	0	0	0
- Argon	kg/s	0	0	0	0	0	0	0	3.2688	0.2388
- Nitrogen	kg/s	0	0	0	0	0	0	0	15.688	14.112
- Oxygen	kg/s	0	0	0	0	0	0	0	7.143	4.323
- Carbon Dioxide	kg/s	0	0	0	0	0	0	0.535	12.455	0.0079
- Sulphur Dioxide	kg/s	0	0	0	0	0	0	0.129	0	0
- Hydrogen Chloride	kg/s	0	0	0	0	0	0	0	-	-
- Nitric Oxide	kg/s	0	0	0	0	0	0	0	0.04069	0
- Nitrogen dioxide	kg/s	0	0	0	0	0	0	0	0	0
- NOx	kg/s	0	0	0	0	0	0	0	0.04	0
Props										
- Phase		Gas	Gas	Gas	Liquid	Liquid	Liquid	Liquid	Gas	Gas
- Temperature	°C	597	360	620	12	-	12	19	20.17	15
- Pressure	bara	290.0	64.50	61.14	-	-	4.0	3.0	1.01	1.013
- Density	kg/m^3	84.61	25.10	15.23	-	-	-	-	-	-
Composition										
O_2	%v/v,wet	-	-	-	-	-	-	-	19.42	20.73
CO_2	%v/v,wet	-	-	-	-	-	-	-	24.62	0.028
SO_2	%v/v,wet	-	-	-	-	-	-	-	0	0
H_2O	%v/v,wet	-	-	-	-	-	-	-	0	0.995
N_2	%v/v,wet	-	-	-	-	-	-	-	48.72	77.328
Ar	%v/v,wet	-	-	-	-	-	-	-	7.12	0.92
NO	%v/v,wet	-	-	-	-	-	-	-	0.118	0
NO_2	%v/v,wet	-	-	-	-	-	-	-	0	0
molecular weight	kg/kmol	-	-	-	18.02	18.02	18.02	18.02	33.58	28.96
Emissions										
NOx	mg/MJ	-	-	-	-	-	-	-	26	-
SOx	mg/MJ	-	-	-	-	-	-	82	0	-
Particulates	mg/MJ	-	-	-	-	-	-	-	0	-

CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : PROCESS FLOW BLOCK DIAGRAM STREAMS 21 - 29

PFD 2



C101	Dual Bed Absorber
C102	Dual Bed Absorber
C103	IP Column
C104	HP Column
C105	LP Column
E101	Main Heat Exchanger
E102	Subcooler
E103	Reboiler
E104	Condensor
K101	Main Air Compressor
K103	Centrifugal Compressor
K104	Expansion Turbine

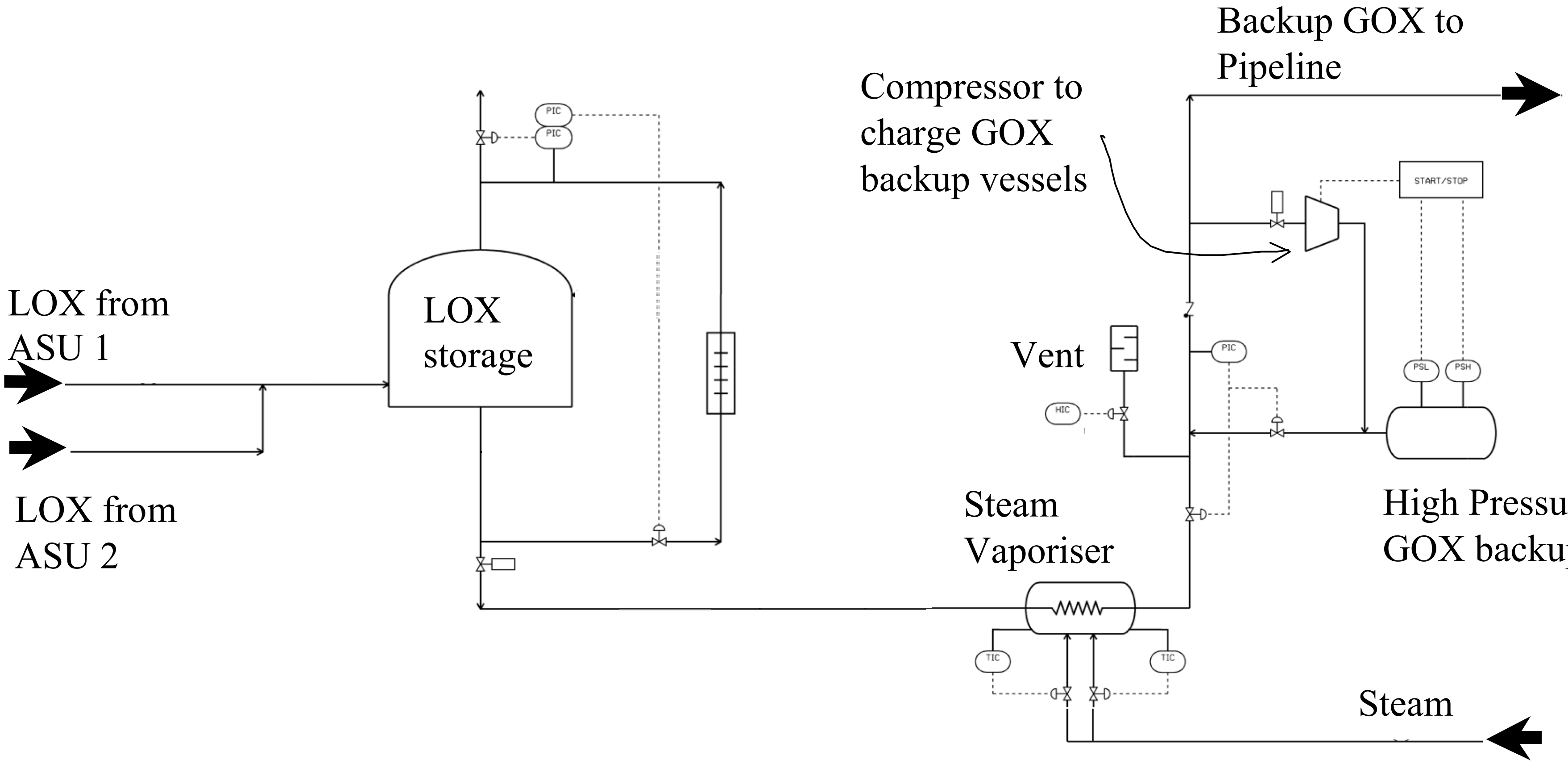
CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : ASU PROCESS FLOW DIAGRAM

PFD 3

STREAM No.		1	2	3	4	5	6	7	8	9	10	11	12	13
Composition - (mol%)														
Nitrogen		77.308	77.308	77.763	78.120	78.120	78.120	78.120	78.120	77.763	77.763	77.763	77.763	77.763
Argon		0.920	0.920	0.926	0.930	0.930	0.930	0.930	0.930	0.926	0.926	0.926	0.926	0.926
Oxygen		20.732	20.732	20.854	20.950	20.950	20.950	20.950	20.950	20.854	20.854	20.854	20.854	20.854
Water		1.000	1.000	0.417	0.000	0.000	0.000	0.000	0.000	0.417	0.417	0.417	0.417	0.417
Carbon Dioxide		0.040	0.040	0.040	0.000	0.000	0.000	0.000	0.000	0.040	0.040	0.040	0.040	0.040
Molecular Weight	kg/kmol	28.86	28.86	28.92	28.96	28.96	28.96	28.96	28.96	28.92	28.92	28.92	28.92	28.92
Flowrate	kg/hr	962,422	962,422	958,904	188,577	188,577	290,223	290,223	290,223	478,563	478,563	478,563	478,563	478,563
	Nm3/hr	747,095	747,095	742,721	145,862	145,862	224,485	224,485	224,485	370,672	370,672	370,672	370,672	370,672
Phase		Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour
Pressure	bar(a)	1.01	3.50	3.50	3.10	3.02	3.10	3.01	1.46	3.50	4.96	4.96	5.41	5.41
Temperature	°C	9.00	144.39	12.00	20.00	-178.54	20.00	-171.44	-188.16	12.00	46.19	20.00	28.92	20.00
STREAM No.		**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Composition - (mol%)														
Nitrogen		78.120	78.120	78.120	78.120	78.120	78.120	54.410	54.410	58.892	58.892	78.120	78.120	98.822
Argon		0.930	0.930	0.930	0.930	0.930	0.930	1.554	1.554	1.527	1.527	0.930	0.930	0.287
Oxygen		20.950	20.950	20.950	20.950	20.950	20.950	44.036	44.036	39.581	39.581	20.950	20.950	0.891
Molecular Weight	kg/kmol	28.96	28.96	28.96	28.96	28.960	28.96	29.954	29.954	29.773	29.773	28.960	28.960	28.084
Flowrate	kg/hr	240,378	240,378	44,788	195,590	236,650	236,650	110,843	110,843	152,635	152,635	145,882	145,882	133,723
	Nm3/hr	185,930	185,930	34,643	151,287	183,046	183,046	82,890	82,890	114,836	114,836	112,839	112,839	106,659
Phase		Vapour	Liquid	Liquid	Liquid	Vapour	Vapour	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Pressure	bar(a)	5.30	5.10	5.10	5.10	5.30	5.09	3.02	2.92	5.09	4.99	5.10	5.00	4.99
Temperature	°C	20.00	-176.75	-176.75	-176.75	20.00	-173.52	-180.78	-187.04	-174.64	-183.74	-176.75	-188.68	-179.06
STREAM No.		**27**	**28**	**29**	**30**	**31**	**32**	**33**	**34**					
Composition - (mol%)														
Nitrogen		98.822	98.254	98.254	99.040	99.040	99.040	1.981	1.981					
Argon		0.287	0.400	0.400	0.352	0.352	0.352	3.033	3.033					
Oxygen		0.891	1.347	1.347	0.608	0.608	0.608	94.985	94.985					
Molecular Weight	kg/kmol	28.08	28.12	28.12	28.08	28.08	28.08	32.16	32.16					
Flowrate	kg/hr	133,723	122,522	122,522	727,040	727,040	727,040	228,788	228,788					
	Nm3/hr	106,659	97,615	97,615	579,970	579,970	579,970	159,354	159,354					
Phase		Liquid	Liquid	Liquid	Vapour	Vapour	Vapour	Liquid	Vapour					
Pressure	bar(a)	4.89	2.92	2.82	1.36	1.31	1.20	1.72	1.60					
Temperature	°C	-190.52	-185.39	-190.43	-193.00	-178.53	15.54	-180.05	15.54					

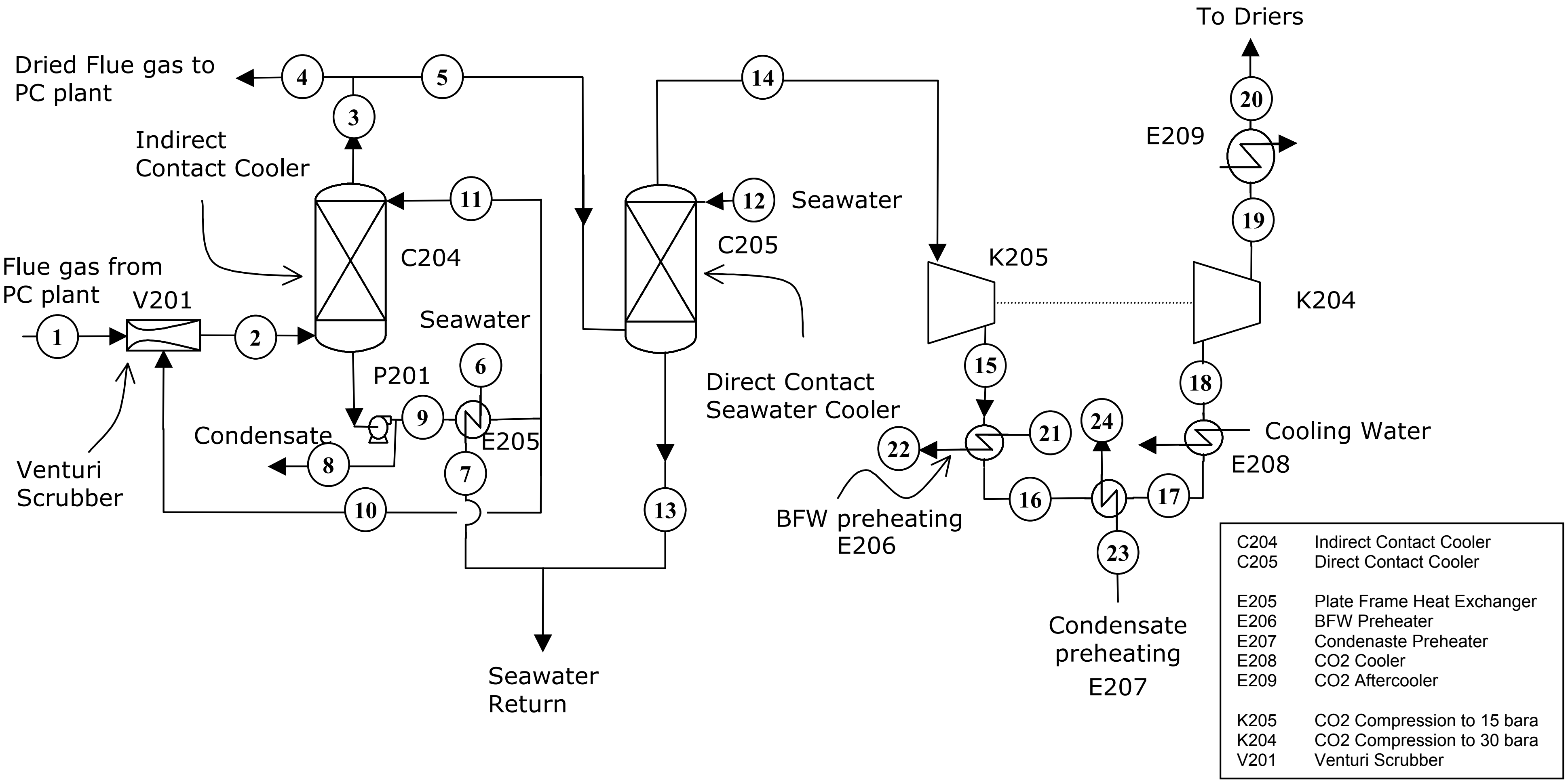
CASE 2 : ASC PF POWER PLANT WITH CO$_2$ CAPTURE PLANT : ASU PROCESS FLOW DIAGRAM STREAMS 1 - 34

PFD 3



CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : OXYGEN BACK UP SYSTEM

PFD 4



C204	Indirect Contact Cooler
C205	Direct Contact Cooler
E205	Plate Frame Heat Exchanger
E206	BFW Preheater
E207	Condenaste Preheater
E208	CO2 Cooler
E209	CO2 Aftercooler
K205	CO2 Compression to 15 bara
K204	CO2 Compression to 30 bara
V201	Venturi Scrubber

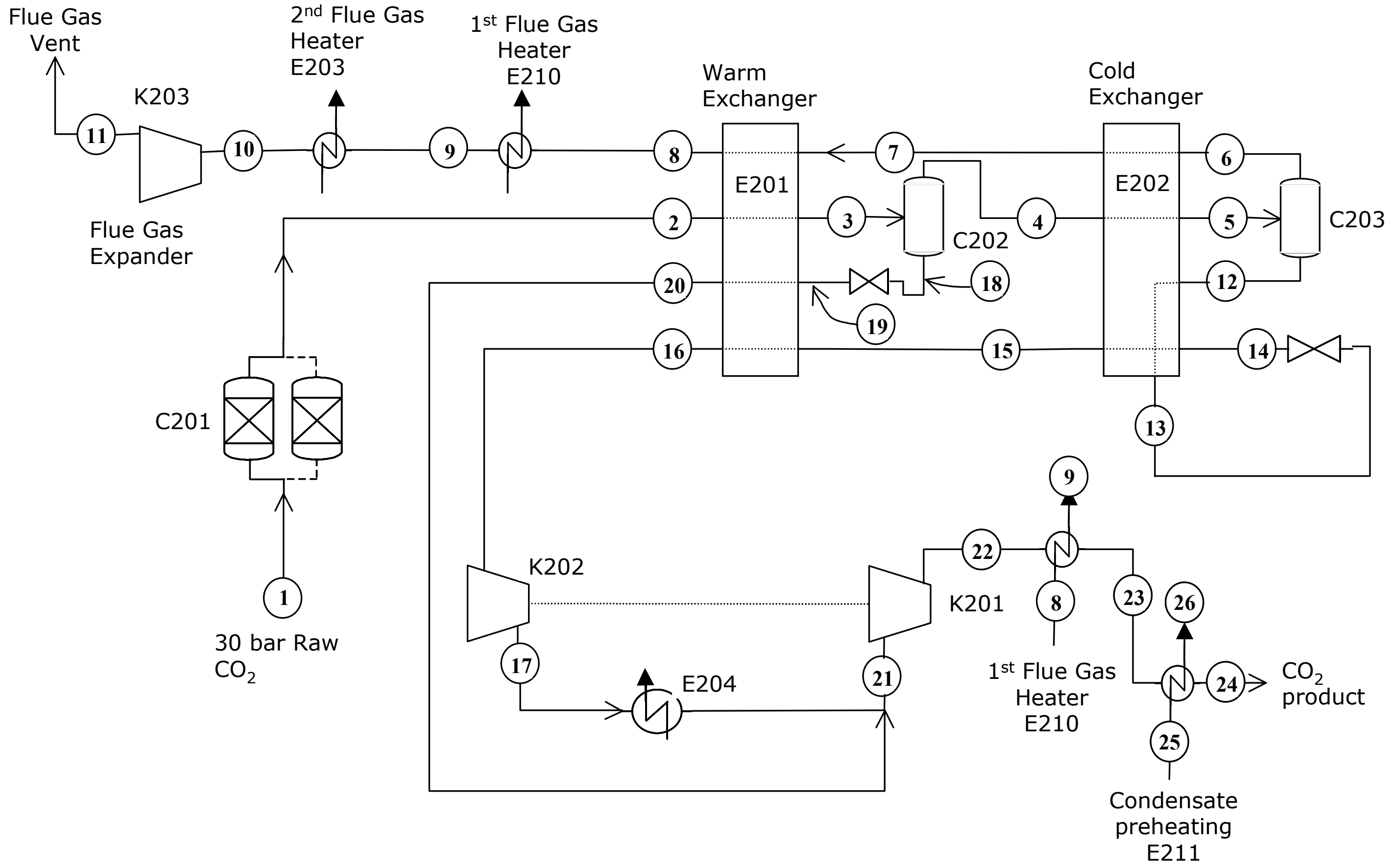
CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : CO_2 COOLING AND COMPRESSION TO 30 BAR (a)

PFD 5

STREAM No.		1	2	3	4	5	6	7	8	9	10	11	12
Composition - (mol%)													
Carbon Dioxide		62.20	59.27	71.46	71.46	71.46	0.00	0.00	0.04	0.04	0.04	0.04	0.00
Oxygen		5.12	4.88	5.88	5.88	5.88	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Argon		2.01	1.91	2.31	2.31	2.31	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen		12.48	11.89	14.34	14.34	14.34	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Water		17.85	21.71	5.62	5.62	5.62	100.00	100.00	99.95	99.95	99.95	99.95	100.00
Sulphur Dioxide		0.30	0.29	0.35	0.35	0.35	0.00	0.00	0.01	0.01	0.01	0.01	0.00
NO		0.035	0.033	0.040	0.040	0.040	0.000	0.000	0.000	0.000	0.000	0.000	0.000
NO2		0.001	0.001	0.001	0.001	0.001	0.000	0.000	0.002	0.002	0.002	0.002	0.000
Molecular Weight	kg/kmol	36.73	35.85	39.52	39.52	39.52	18.02	18.02	18.03	18.03	18.03	18.03	18.02
Flow	kg/hr	1,251,499	1,281,820	1,171,841	557,562	614,279	9,407,225	9,407,225	79,657	3,670,468	30,321	3,640,147	1,303,688
	Nm3/hr	763,192	800,867	664,217	316,035	348,183	11,697,021	11,697,021	98,975	4,560,598	37,675	4,522,923	1,621,016
Phase		Vapour	2 Phase	Vapour	Vapour	Vapour	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Pressure	bar(a)	1.02	1.02	1.02	1.02	1.02	4.00	3.00	4.00	4.00	3.00	3.00	1.00
Temperature	°C	110.95	61.09	35.00	35.00	35.00	12.00	19.00	35.02	35.02	17.00	17.00	12.00
STREAM No.		13	14	15	16	17	18	19	20	21	22	23	24
Composition - (mol%)													
Carbon Dioxide		0.06	74.34	74.34	74.34	74.34	75.54	75.54	75.54	0.00	0.00	0.00	0.00
Oxygen		0.00	6.14	6.14	6.14	6.14	6.24	6.24	6.24	0.00	0.00	0.00	0.00
Argon		0.00	2.41	2.41	2.41	2.41	2.45	2.45	2.45	0.00	0.00	0.00	0.00
Nitrogen		0.00	14.98	14.98	14.98	14.98	15.22	15.22	15.22	0.00	0.00	0.00	0.00
Water		99.93	1.77	1.77	1.77	1.77	0.18	0.18	0.18	100.00	100.00	100.00	100.00
Sulphur Dioxide		0.01	0.32	0.32	0.32	0.32	0.32	0.32	0.32	0.00	0.00	0.00	0.00
NO		0.000	0.042	0.042	0.042	0.042	0.042	0.042	0.042	0.000	0.000	0.000	0.000
NO2		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Molecular Weight	kg/kmol	18.04	40.38	40.38	40.38	40.38	40.74	40.74	40.74	18.02	18.02	18.02	18.02
Flow	kg/hr	1,317,061	600,906	600,906	600,906	600,906	596,546	596,546	596,546	150,956	150,956	330,635	330,635
	Nm3/hr	1,635,819	333,380	333,380	333,380	333,380	328,031	328,031	328,031	187,700	187,700	411,114	411,114
Phase		Liquid	Vapour	Vapour	Vapour	2 Phase	Vapour	Vapour	Vapour	Liquid	Liquid	Liquid	Liquid
Pressure	bar(a)	1.01	1.01	15.00	15.00	15.00	15.00	30.00	30.00	338.53	338.53	6.00	6.00
Temperature	°C	19.00	13.01	278.27	185.00	50.00	20.00	82.18	20.00	165.00	250.00	33.37	93.20

CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : CO_2 COOLING AND COMPRESSION TO 30 BAR (a) STREAMS 1 - 24

PFD 5



C201	Dual Bed Dryer
C202	Separator Pot
C203	Separator Pot
E201	Warm Heat Exhanger
E202	Cold Heat Exchanger
E203	2nd Flue Gas Heater
E210	1st Flue Gas Heater
E211	Condensate Preheater
K201	Adiabatic Compressor
K202	Compressor
K203	Flue Gas Expander

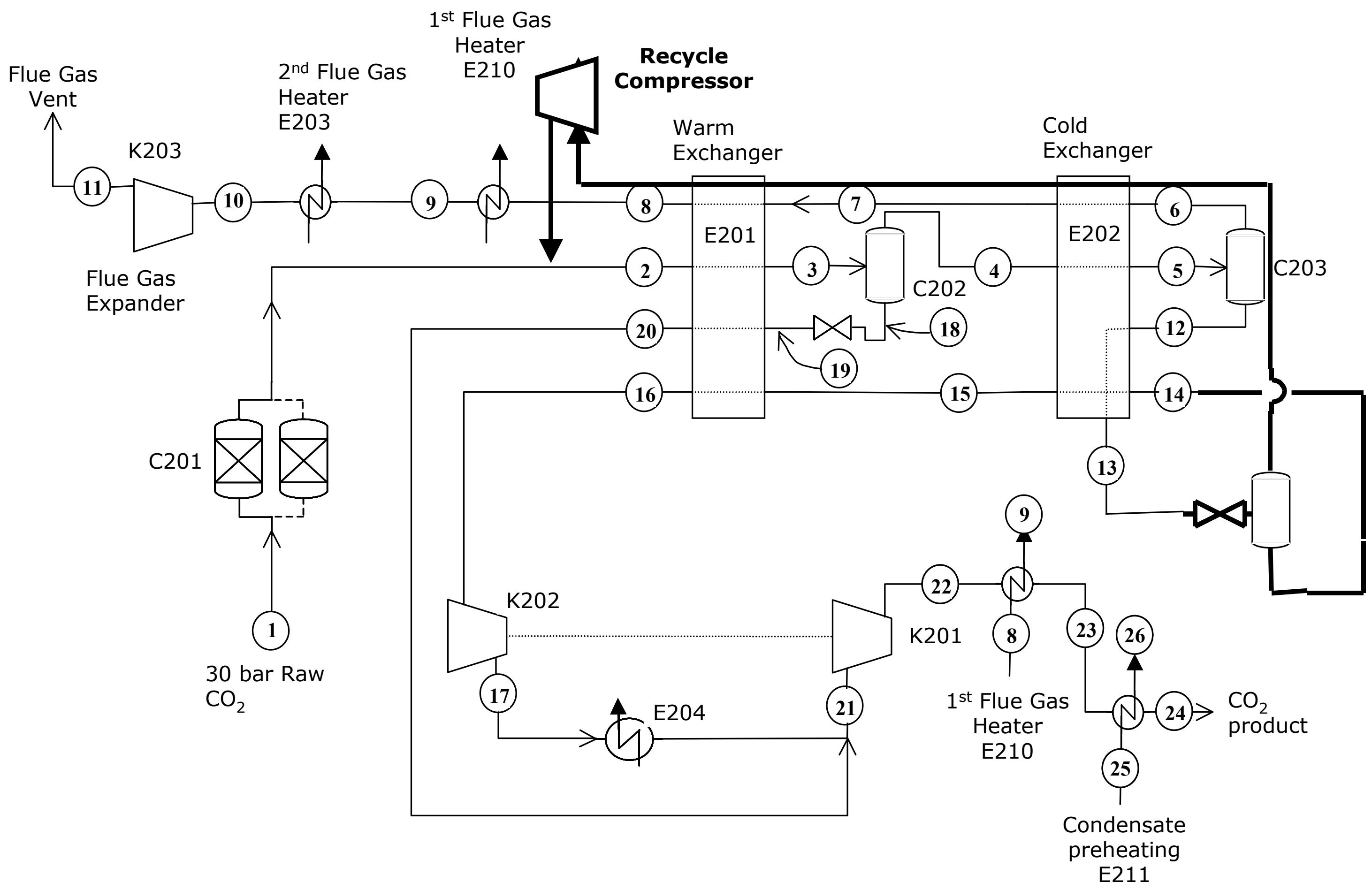
CASE 2 : ASC PF POWER PLANT WITH CO$_2$ CAPTURE : CO$_2$ INERTS REMOVAL AND COMPRESSION TO 110 BAR (a)

PFD 6

STREAM No.		1	2	3	4	5	6	7	8	9	10	11	12	13
Composition - (mol%)														
Carbon Dioxide		75.54	75.67	75.67	63.69	63.69	24.62	24.62	24.62	24.62	24.62	24.62	95.06	95.06
Oxygen		6.24	6.25	6.25	9.41	9.41	19.42	19.42	19.42	19.42	19.42	19.42	1.37	1.37
Argon		2.45	2.45	2.45	3.62	3.62	7.12	7.12	7.12	7.12	7.12	7.12	0.82	0.82
Nitrogen		15.22	15.25	15.25	23.14	23.14	48.72	48.72	48.72	48.72	48.72	48.72	2.59	2.59
Water		0.18	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide		0.32	0.32	0.32	0.07	0.07	0.00	0.00	0.00	0.00	0.00	0.00	0.13	0.13
NO		0.042	0.043	0.043	0.063	0.063	0.118	0.118	0.118	0.118	0.118	0.118	0.018	0.018
NO2		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Molecular Weight	kg/kmol	40.74	40.78	40.78	39.04	39.04	33.58	33.58	33.58	33.58	33.58	33.58	43.42	43.42
Flow	kg/hr	596,546	596,069	596,069	362,868	362,868	138,989	138,989	138,989	138,989	138,989	138,989	223,879	223,879
	Nm3/hr	328,031	327,438	327,438	208,218	208,218	92,720	92,720	92,720	92,720	92,720	92,720	115,498	115,498
Phase		Vapour	Vapour	2 Phase	Vapour	2 Phase	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Liquid	2 Phase
Pressure	bar(a)	30.00	30.00	29.72	29.72	29.45	29.45	29.17	28.90	28.90	28.90	1.10	29.45	29.24
Temperature	°C	20.00	20.00	-24.51	-24.51	-54.69	-54.69	-42.17	7.47	170.00	300.00	20.17	-54.69	-46.44
STREAM No.		**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Composition - (mol%)														
Carbon Dioxide		95.06	95.06	95.06	95.06	96.60	96.60	96.60	95.84	95.84	95.84	95.84	0.00	0.00
Oxygen		1.37	1.37	1.37	1.37	0.74	0.74	0.74	1.05	1.05	1.05	1.05	0.00	0.00
Argon		0.82	0.82	0.82	0.82	0.41	0.41	0.41	0.61	0.61	0.61	0.61	0.00	0.00
Nitrogen		2.59	2.59	2.59	2.59	1.48	1.48	1.48	2.03	2.03	2.03	2.03	0.00	0.00
Water		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	100.00	100.00
Sulphur Dioxide		0.13	0.13	0.13	0.13	0.75	0.75	0.75	0.45	0.45	0.45	0.45	0.00	0.00
NO		0.018	0.018	0.018	0.018	0.007	0.007	0.007	0.013	0.013	0.013	0.013	0.000	0.000
NO2		0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Molecular Weight	kg/kmol	43.42	43.42	43.42	43.42	43.82	43.82	43.82	43.62	43.62	43.62	43.62	18.02	18.02
Flow	kg/hr	223,879	223,879	223,879	223,879	233,201	233,201	233,201	457,080	457,080	457,080	457,080	378,478	378,478
	Nm3/hr	115,498	115,498	115,498	115,498	119,220	119,220	119,220	234,718	234,718	234,718	234,718	470,602	470,602
Phase		2 Phase	2 Phase	Vapour	Vapour	Liquid	2 Phase	Vapour	Vapour	Vapour	Vapour	Liquid	Liquid	Liquid
Pressure	bar(a)	9.74	9.54	9.33	18.69	29.72	18.80	18.59	18.59	110.00	110.00	110.00	6.00	6.00
Temperature	°C	-55.69	-42.17	7.47	65.63	-24.51	-31.27	7.47	19.39	187.79	148.55	43.00	33.37	93.20

CASE 2 : ASC PF POWER PLANT WITH CO$_2$ CAPTURE : CO$_2$ INERTS REMOVAL AND COMPRESSION TO 110 BAR(a)
STREAMS 1 – 26

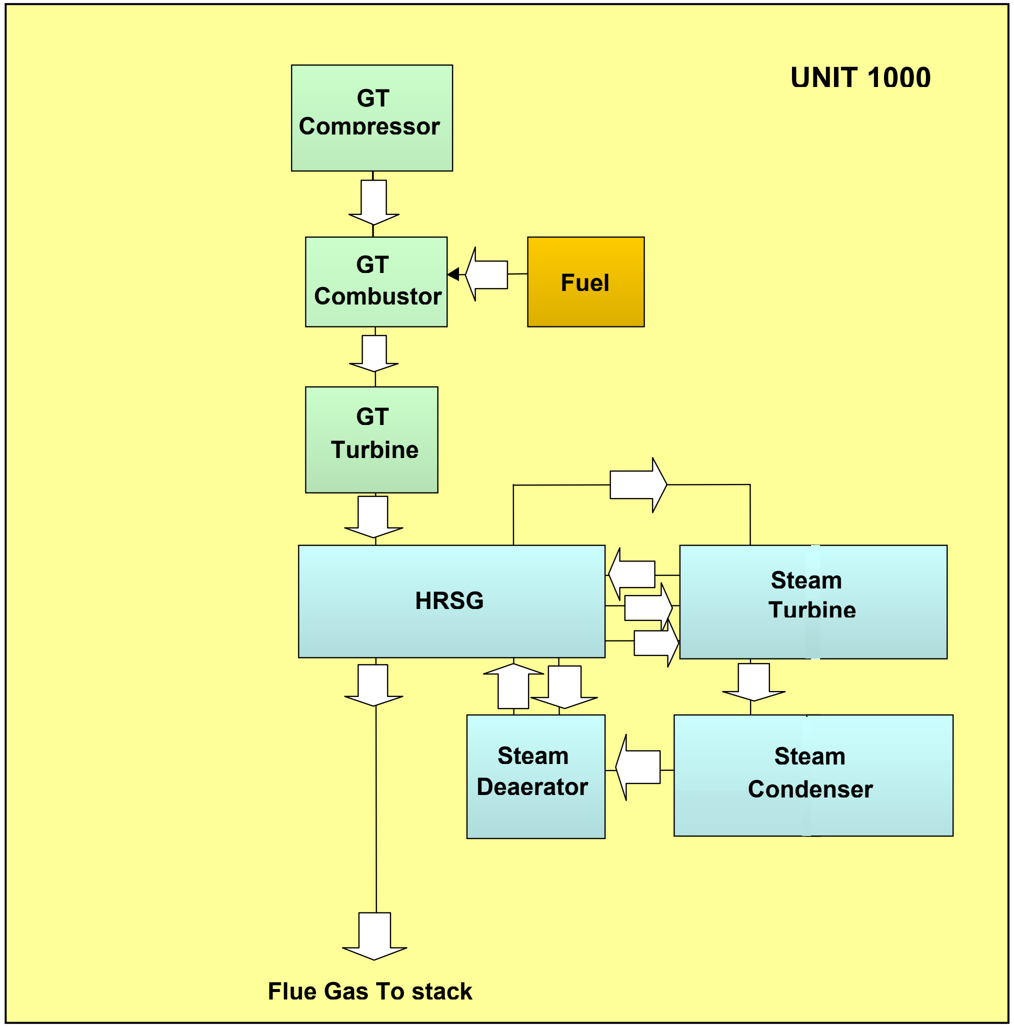
PFD 6



C201	Dual Bed Dryer
C202	Separator Pot
C203	Separator Pot
E201	Warm Heat Exhanger
E202	Cold Heat Exchanger
E203	2nd Flue Gas Heater
E210	1st Flue Gas Heater
E211	Condensate Preheater
K201	Adiabatic Compressor
K202	Compressor
K203	Flue Gas Expander

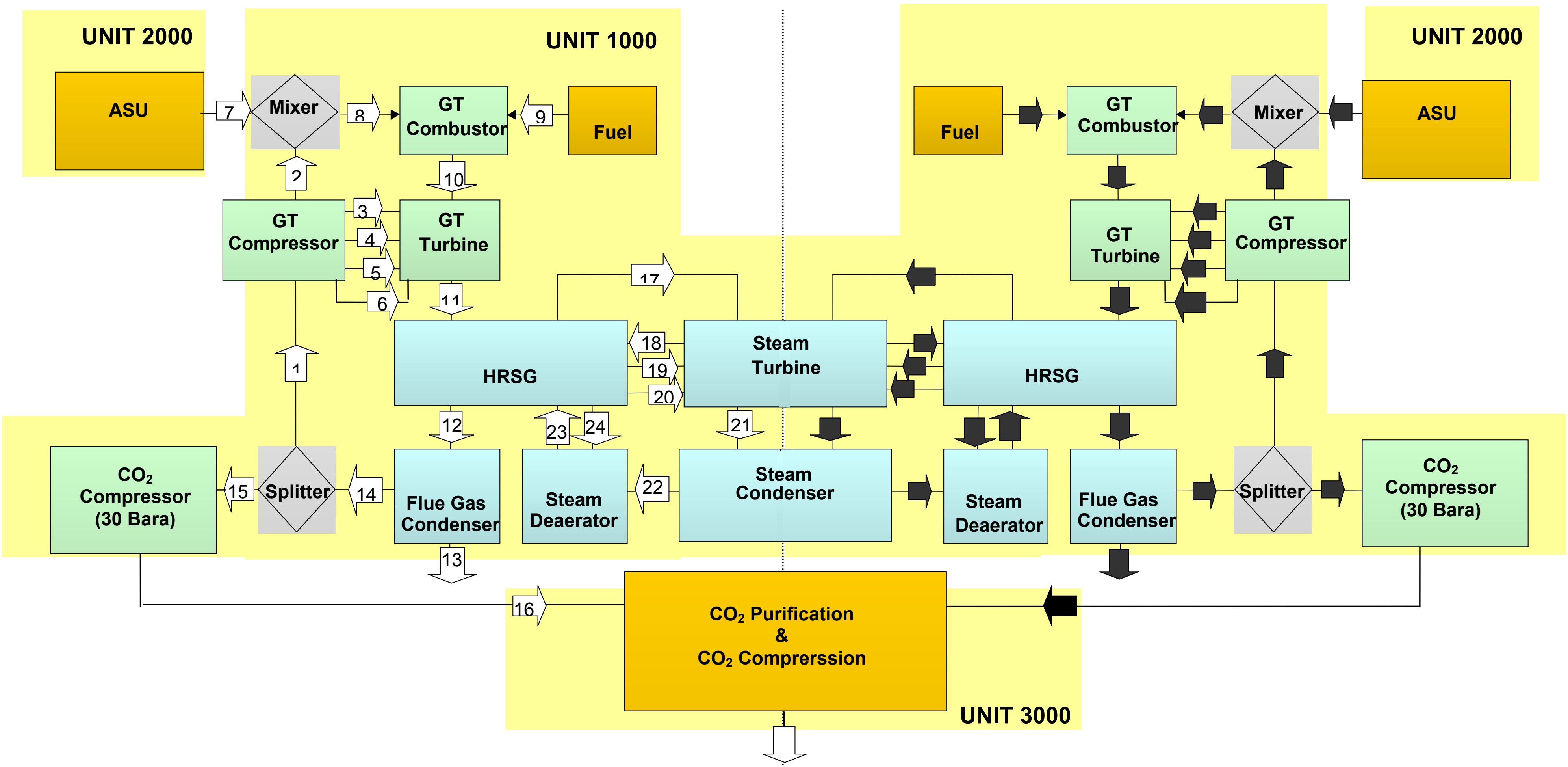
CASE 2 : ASC PF POWER PLANT WITH CO_2 CAPTURE : CO_2 COMPRESSION AND INERTS REMOVAL (98% v/v CO_2 PURITY)

PFD 7



CASE 3 : CONVENTIONAL NGCC POWER PLANT : PROCESS FLOW BLOCK DIAGRAM

PFD 8



CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : PROCESS FLOW BLOCK DIAGRAM

PFD 9

Gas Turbine Cold Flow

	Stream	1	2	3	4	5	6
		Comp	Comp	Turbine 1	Turbine 2	Turbine 3	Turbine 4
		Inlet	Exit	Cool/Leak	Cool/Leak	Cool/Leak	Cool/Leak
Argon	molar fraction	0.05563	0				
Carbon Dioxide	molar fraction	0.868674	0.05563				
H2O	molar fraction	0.016409	0.016409				
Hydro Carbons	molar fraction	0	0				
Molecular Weight	molar fraction	42.49803	42.49803				
Nitrogen	molar fraction	0.036862	0.036862				
Oxygen	molar fraction	0.022425	0.022425				
Enthalpy	kj/kg	-0.58672	316.002	316.002014	232.7036	183.3461	98.05312
Flow	kg/s	833.0602	600	83.6784363	70.28989	51.90784	27.18406
Pressure	kpa	101.3003	3039.01	3039.01025	1539.254	978.0485	394.8759
Quality		0	0	0	0	0	0
Temperature	k	288	618.1804	618.18042	538.1429	488.9767	399.9981
LHV	KJ/KG						
H/C							

CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : PROCESS FLOW BLOCK DIAGRAM
STREAMS 1 - 6

PFD 9

Hot Gas Flow

	Stream	7	8	9	10	11	12	13	14	15	16
		ASU	Combustor	Fuel	Combustor	Turbine	HRSG	Flugas Condenser	Flugas Condenser	Splitter	Purification
		Exit	Inlet	Exit	Exit	Exit	Exit	Water Exit	Gas Exit	CO2 Capture	Entry
Argon	molar fraction	0.03172	0.0518725	0	0.048485	0.050173	0.050173		0.0501735	0.050173	0.0563828
Carbon Dioxide	molar fraction	0	0.7321672	0	0.757101	0.783471	0.783471		0.7834714	0.783471	0.880431
H2O	molar fraction	0	0.0138303	0	0.142741	0.112883	0.112883	1	0.1128827	0.112883	0.0030962
Hydro Carbons	molar fraction	0	0	0.994501	0	0	0		0	0	0
Molecular Weight	molar fraction	32.1781	40.87632	15.67648	39.35347	40.09668	40.09668		40.096676	40.09668	42.829395
Nitrogen	molar fraction	0.01828	0.0339423	0.005499	0.032128	0.033247	0.033247		0.0332468	0.033247	0.0373613
Oxygen	molar fraction	0.95	0.1681876	0	0.019546	0.020226	0.020226		0.0202256	0.020226	0.0227287
Enthalpy	kj/kg	0.23	276.39627	-1.58	1677.077	667.3077	62.8054	62.410084	-0.5867206	-0.58672	11.076401
Flow	kg/s	83.48	677	20.32	697.3203	929.4368	930.3804	41.000305	889.38013	56	56.001049
Pressure	kpa	4040	2990.3862	1500	2885.723	118.6	106.0237	102.84299	102.84299	102.843	3038.9763
Quality		0	0	0	3.56872	3.56872	3.56872	0	0	0	0
Temperature	k	288	580.83984	288	1662.389	907.0369	357.8909	288	288	288	304
LHV	KJ/KG			48452							
H/C				3.5687							

CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO$_2$ CAPTURE : PROCESS FLOW BLOCK DIAGRAM
STREAMS 7 - 16

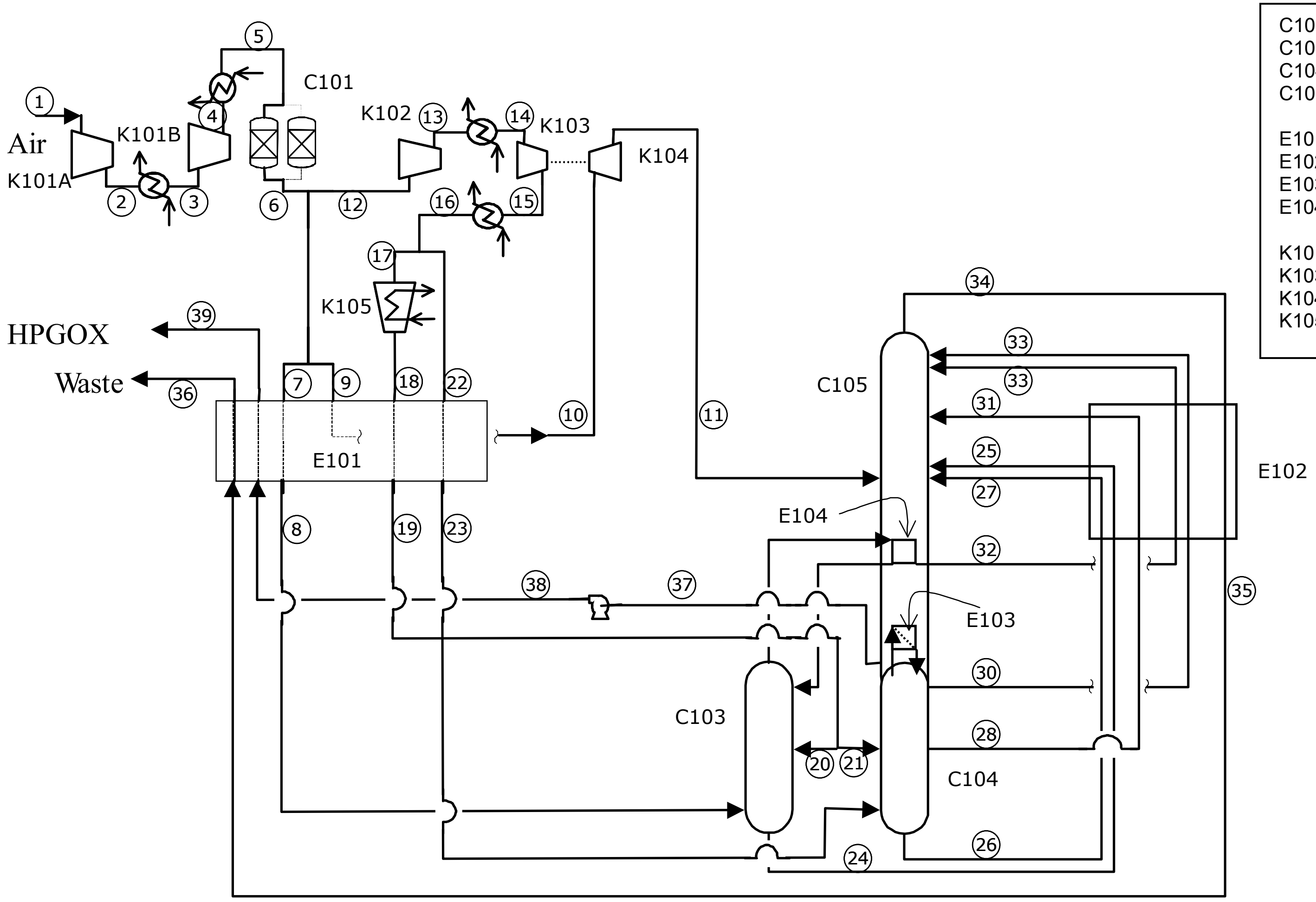
PFD 9

Steam Cycle

	Stream	17	18	19	20	21	22	23	24
		HP Steam Turbine	HP Steam Turbine	IP Steam Turbine	LP Steam Turbine	Steam Condenser	DeAerator	DeAerator	DeAerator Steam
		Inlet	Exit	Inlet	Inlet	Inlet	Inlet	Exit	Inlet
Argon	molar fraction								
Carbon Dioxide	molar fraction								
H2O	molar fraction								
Hydro Carbons	molar fraction								
Molecular Weight	molar fraction								
Nitrogen	molar fraction								
Oxygen	molar fraction								
Enthalpy	kj/kg	3550.374	3043.432	3692.98	2877.575	2395.3486	394.8398	439.3643	2877.575
Flow	kg/s	102.59	102.6121	134.9259	145.6711	145.67018	355.9949	362.4986	6.828516
Pressure	kpa	19212.21	2660.204	2325.018	120	4.5450001	125.44	120	120
Quality		1	1	1	1	0.9327778	0	0	1
Temperature	k	876.0002	588.5786	876.0002	474.7449	304.35962	367.3943	377.9582	474.7449
LHV	KJ/KG								
H/C									

CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : PROCESS FLOW BLOCK DIAGRAM
STREAMS 17-24

C101	Dual Bed Absorber
C103	IP Column
C104	HP Column
C105	LP Column
E101	Main Heat Exchanger
E102	Subcooler
E103	Reboiler
E104	Condensor
K101	Main Air Compressor
K103	Centrifugal Compressor
K104	Expansion Turbine
K105	Multi-Stage Intercooled Compressor



CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : ASU (PUMPED LOX) PROCESS FLOW DIAGRAM

PFD 10

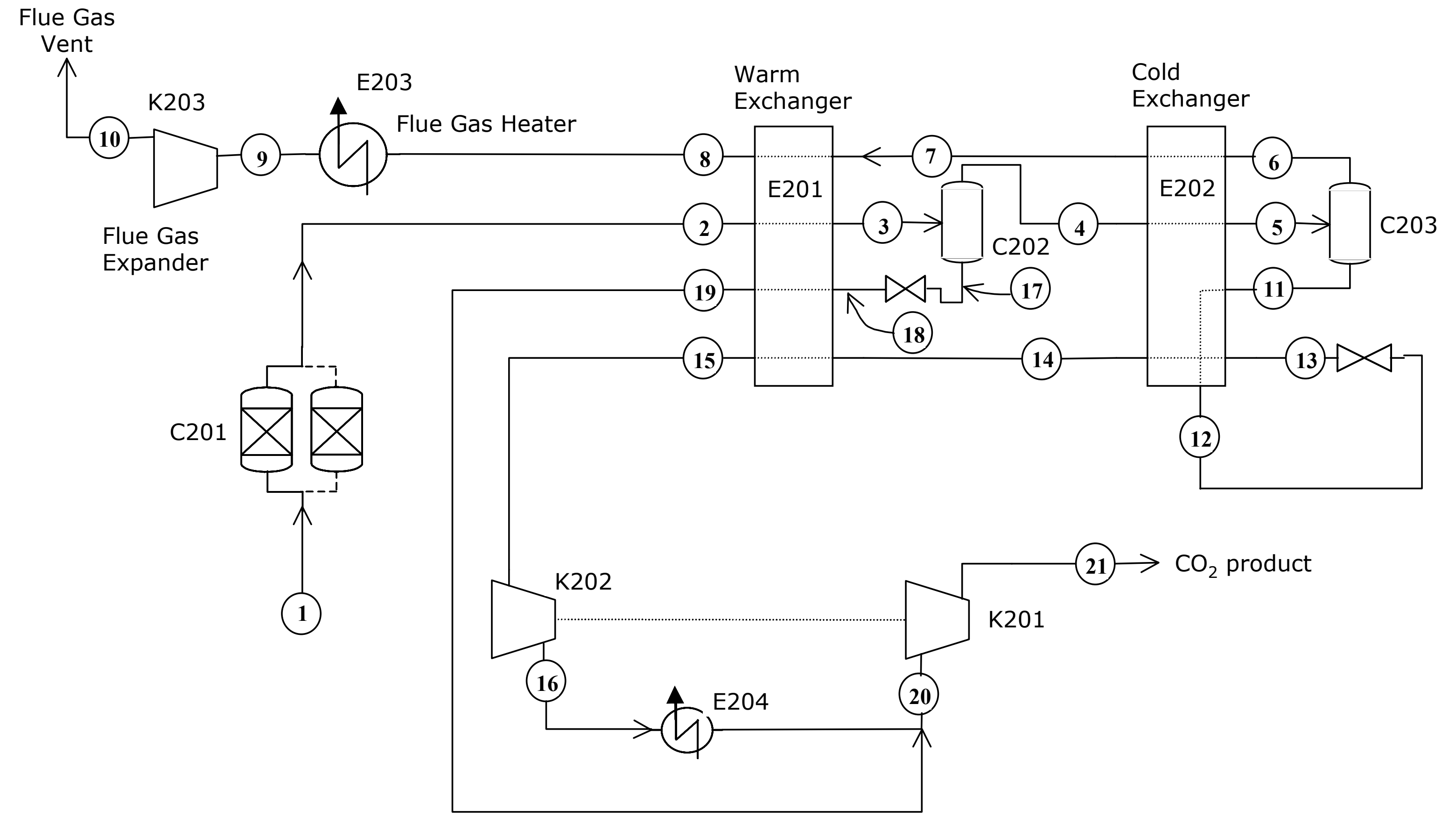
STREAM No.		1	2	3	4	5	6	7	8	9	10	11	12	13
Composition - (mol%)														
Nitrogen		77.308	77.308	77.308	77.308	77.525	78.120	78.120	78.120	78.120	78.120	78.120	78.120	78.120
Argon		0.920	0.920	0.920	0.920	0.923	0.930	0.930	0.930	0.930	0.930	0.930	0.930	0.930
Oxygen		20.732	20.732	20.732	20.732	20.790	20.950	20.950	20.950	20.950	20.950	20.950	20.950	20.950
Water		1.000	1.000	1.000	1.000	0.722	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Carbon Dioxide		0.040	0.040	0.040	0.040	0.040	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Molecular Weight	kg/kmol	28.86	28.86	28.86	28.86	28.89	28.96	28.96	28.96	28.96	28.96	28.96	28.96	28.96
Flowrate	kg/hr	629,481	629,481	629,481	629,481	628,382	625,173	149,219	149,219	152,557	152,557	152,557	323,397	323,397
	Nm3/hr	488,643	488,643	488,643	488,643	487,277	483,565	115,419	115,419	118,002	118,002	118,002	250,144	250,144
Phase		Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	Vapour	2 Phase	Vapour	Vapour
Pressure	bar(a)	1.01	1.84	1.84	3.36	3.36	3.36	3.36	3.24	3.36	3.27	2.31	3.36	5.19
Temperature	°C	9.00	67.98	20.00	81.34	20.00	20.00	20.00	-170.83	20.00	-175.41	-183.64	20.00	64.57

STREAM No.		14	15	16	17	18	19	20	21	22	23	24	25	26
Composition - (mol%)														
Nitrogen		78.120	78.120	78.120	78.120	78.120	78.120	78.120	78.120	78.120	78.120	53.871	53.871	64.134
Argon		0.930	0.930	0.930	0.930	0.930	0.930	0.930	0.930	0.930	0.930	1.653	1.653	1.391
Oxygen		20.950	20.950	20.950	20.950	20.950	20.950	20.950	20.950	20.950	20.950	44.476	44.476	34.475
Molecular Weight	kg/kmol	28.96	28.96	28.96	28.96	28.960	28.960	28.960	28.960	28.960	28.960	29.983	29.983	29.554
Flowrate	kg/hr	323,397	323,397	323,397	198,867	198,867	198,867	34,860	164,007	124,530	124,530	88,961	88,961	122,262
	Nm3/hr	250,144	250,144	250,144	153,821	153,821	153,821	26,964	126,858	96,323	96,323	66,462	66,462	92,668
Phase		Vapour	Vapour	Vapour	Vapour	Vapour	Liquid	Liquid	Liquid	Vapour	Vapour	Liquid	Liquid	Liquid
Pressure	bar(a)	5.19	5.36	5.26	5.26	90.00	89.80	89.80	89.80	5.26	5.14	3.24	3.14	5.14
Temperature	°C	20.00	23.26	20.00	20.00	20.00	-160.32	-160.32	-160.32	20.00	-173.38	-179.86	-183.78	-175.12

STREAM No.		27	28	29	30	31	32	33	34	35	36	37	38	39
Composition - (mol%)														
Nitrogen		64.134	78.120	78.120	98.420	98.420	99.347	99.347	99.266	99.266	99.266	1.922	1.922	1.922
Argon		1.391	0.930	0.930	0.260	0.260	0.297	0.297	0.323	0.323	0.323	3.118	3.118	3.118
Oxygen		34.475	20.950	20.950	1.320	1.320	0.355	0.355	0.411	0.411	0.411	94.960	94.960	94.960
Molecular Weight	kg/kmol	29.55	28.96	28.96	28.10	28.10	28.06	28.06	28.07	28.07	28.07	32.17	32.17	32.17
Flowrate	kg/hr	122,262	86,189	86,189	80,087	80,087	95,117	95,117	474,316	474,316	474,316	150,857	150,857	150,857
	Nm3/hr	92,668	66,666	66,666	63,847	63,847	75,921	75,921	378,521	378,521	378,521	105,044	105,044	105,044
Phase		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Vapour	Vapour	Vapour	Liquid	Liquid	Vapour
Pressure	bar(a)	5.04	89.80	89.70	5.04	4.94	3.14	3.04	1.38	1.33	1.20	1.48	41.24	41.00
Temperature	°C	-183.95	-160.32	-185.32	-178.88	-190.04	-184.66	-188.84	-192.91	-174.82	16.18	-179.90	-176.92	16.18

CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : ASU (PUMPED LOX) PROCESS FLOW DIAGRAM STREAMS 1 - 39

PFD 10



C201	Dual Bed Dryer
C202	Flash Separator
C203	Flash Separator
E201	Warm Heat Exhanger
E202	Cold Heat Exchanger
E203	Flue Gas Heater
K201	Adiabatic Compressor
K202	Compressor
K203	Flue Gas Expander

CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE CO_2 COMPRESSION AND INERTS REMOVAL PROCESS FLOW DIAGRAM

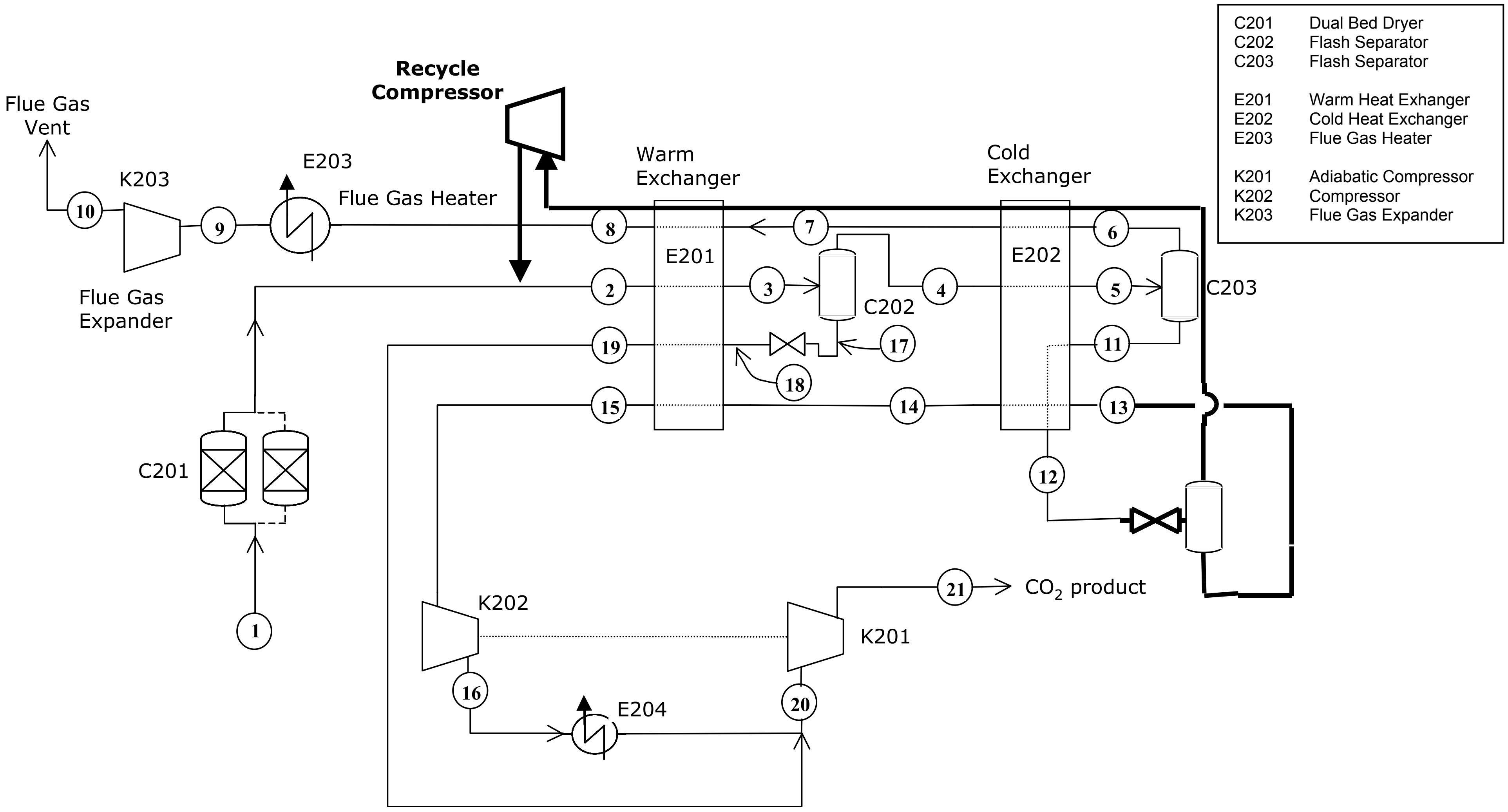
PFD 11

STREAM No.		1	2	3	4	5	6	7	8	9	10	11
Composition - (mol%)												
Carbon Dioxide		88.14	88.32	88.32	58.34	58.34	23.85	23.85	23.85	23.85	23.85	93.39
Oxygen		2.28	2.28	2.28	8.40	8.40	15.55	15.55	15.55	15.55	15.55	1.14
Argon		5.64	5.66	5.66	18.82	18.82	33.44	33.44	33.44	33.44	33.44	3.97
Nitrogen		3.74	3.75	3.75	14.44	14.44	27.16	27.16	27.16	27.16	27.16	1.51
Water		0.20	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Molecular Weight	kg/kmol	42.86	42.91	42.91	39.93	39.93	36.44	36.44	36.44	36.44	36.44	43.47
Flow	kg/hr	201,505	201,338	201,338	39,082	39,082	17,976	17,976	17,976	17,976	17,976	21,106
	Nm3/hr	105,320	105,113	105,113	21,926	21,926	11,051	11,051	11,051	11,051	11,051	10,875
Phase		Vapour	Vapour	2 Phase	Vapour	2 Phase	Vapour	Vapour	Vapour	Vapour	Vapour	Liquid
Pressure	bar(a)	30.39	30.39	30.11	30.11	29.84	29.84	29.56	29.29	29.29	1.05	29.84
Temperature	°C	31.00	31.00	-27.45	-27.45	-55.00	-55.00	-42.25	21.63	325.37	35.00	-55.00

STREAM No.		12	13	14	15	16	17	18	19	20	21	
Composition - (mol%)												
Carbon Dioxide		93.39	93.39	93.39	93.39	93.39	96.22	96.22	96.22	95.89	95.89	
Oxygen		1.14	1.14	1.14	1.14	1.14	0.67	0.67	0.67	0.72	0.72	
Argon		3.97	3.97	3.97	3.97	3.97	2.19	2.19	2.19	2.39	2.39	
Nitrogen		1.51	1.51	1.51	1.51	1.51	0.93	0.93	0.93	1.00	1.00	
Water		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Molecular Weight	kg/kmol	43.47	43.47	43.47	43.47	43.47	43.69	43.69	43.69	43.67	43.67	
Flow	kg/hr	21,106	21,106	21,106	21,106	21,106	162,256	162,256	162,256	183,362	183,362	
	Nm3/hr	10,875	10,875	10,875	10,875	10,875	83,187	83,187	83,187	94,062	94,062	
Phase		2 Phase	2 Phase	2 Phase	Vapour	Vapour	Liquid	2 Phase	Vapour	Vapour	Liquid	
Pressure	bar(a)	29.63	10.08	9.87	9.66	21.64	30.11	21.74	21.54	21.54	110.00	
Temperature	°C	-45.03	-56.35	-42.25	21.63	95.25	-27.45	-31.87	21.63	21.43	20.00	

CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE
CO_2 COMPRESSION AND INERTS REMOVAL PROCESS FLOW DIAGRAM STREAMS 1 – 21

PFD 11



C201	Dual Bed Dryer
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E201	Warm Heat Exhanger
E202	Cold Heat Exchanger
E203	Flue Gas Heater
K201	Adiabatic Compressor
K202	Compressor
K203	Flue Gas Expander

CASE 4 : NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE
CO_2 COMPRESSION AND INERTS REMOVAL PROCESS FLOW DIAGRAM - HIGH CO_2 PURITY CASE (98% v/v)

PFD 12

TABLES

Option		**Secondary Flue Gas Recycle**				**Primary Flue Gas Recycle**				**Figure No.**
	ESP Temp	**Cooled**	**Cleaned**	**Dried**	**Reheated**	**Cooled**	**Cleaned**	**Dried**	**Reheated**	
A	Cold	✓	✓	✓	✓	Gas/gas heater prior to ESP FWH after ESP	✓	✓	✓	**3**
B	Warm	✓	✓	No	✓	Gas/gas heater prior to ESP FWH after ESP	✓	✓	✓	**4**
C_1	Hot	No	✓	No	No	Parallel gas / gas heater and FWH after ESP	✓	✓	✓	**5**
C_2	Hot	No	✓	No	No	Series gas / gas heater and FWH after ESP	✓	✓	✓	**6**
C_3	Hot	No	✓	No	No	FWH after ESP Primary recycle water heated	✓	✓	✓	**7**

FLUE GAS RECYCLE ARRANGEMENT OPTIONS FOR ASC PF OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE

TABLE 1

Proximate analysis	weight %
Coal (dry, ash-free)	78.3
Ash	12.2
Moisture	9.5
Ultimate Analysis:	DAF
Carbon	82.5
Hydrogen	5.6
Oxygen	9.0
Nitrogen	1.8
Sulphur	1.1
Chlorine	0.03
Ash Analysis:	
SiO_2	50.0
Al_2O_3	30.0
TiO_2	2.0
Fe_2O_3	9.7
CaO	3.9
MgO	0.4
NA_2O	0.1
K_2O	0.1
P_2O_5	1.7
SO_3	1.7
Gross CV	27.06 MJ/kg
Net CV	25.87 MJ/kg

Hardgrove index	45
Ash Fusion Point (reducing atmosphere)	1350 °C

This coal specification is based on an open cut coal from Eastern Australia

COAL SPECIFICATION (BITUMINOUS)

TABLE 2

Case		**CASE 1**	**CASE 2**	**CASE 3**	**CASE 4**
Name		**ASC PF**	**ASC PF CAP**	**NGCC**	**NGCC CAP**
ALSTOM Turbine Heat Balance Ref					
Fuel Input	kg/s	59.19	58.09	14.77	20.32
Fuel NCV	MJ/kg	25.86	25.86	46.90	48.452
Fuel Heat Input (Wf x NCV)	MWt	1530.8	1502.2	692.9	984.5
GROSS OUTPUT	**MWe**	**740**	**737**	**400**	**575**
Power plant auxiliaries	MWe			8	12
FW Pumps, Condensate, CW Pumps	MWe	35	35		
Draught Plant	MWe	7	5		
Coal & Ash Handling, Mills & Feeders	MWe	4	4		
ESP	MWe	2	2		
Miscellaneous	MWe	7	8		
FGD	MWe	7	0		
DENOX	MWe	0.3	0.0		
Sub-total Power Plant	MWe	62.5	53.7	8	12
ASU	MWe	0	87		90
CO2 Compression & Inerts Removal	MWe	0	65		28
Utility Sytems	MWe	0	0	4	6
Sub-Total Capture Costs	MWe	0	152	4	124
Total Losses	**MWe**	**63**	**205**	**12**	**135**
Contingency (0% of subtotal)	%	0	0	0	0
TOTAL LOSS	**MWe**	63	205	12	135
NET OUTPUT	**MWe**	**677**	**532**	**388**	**440**
Gross Efficiency	**% LHV**	**48.3**	**49.1**	**57.7**	**58.4**
Net Efficiency	**% LHV**	**44.3**	**35.4**	**56.0**	**44.7**
Delta Net Efficiency (Base - CO2 Capture)			**8.9**		**11.3**

ASC PF & NGCC POWER PLANTS
AUXILIARY POWER CONSUMPTION WITH & WITHOUT CO_2 CAPTURE

TABLE 3

Gas **GT Pressure Ratio**		**Air** **20**	**CO_2** **20**	**CO_2** **30**
Gamma		1.3965	1.299	1.299
Rgas	kJ/kg/K	0.287	0.194	0.194
WRT/AP(0.5Mn)		30.19	35.50	35.50
Relative Inlet Flow		1.0	1.175	1.175
Enthalpy Rise	kJ/kg	438.3	261.6	311.7
Comp. Exit Temp	K	711	561	608
Stage No. (Nstg)		15	15	18
Rel. Stage Load		1.0	1.0	1
Rel. Shaft Speed		1.0	0.772	0.772

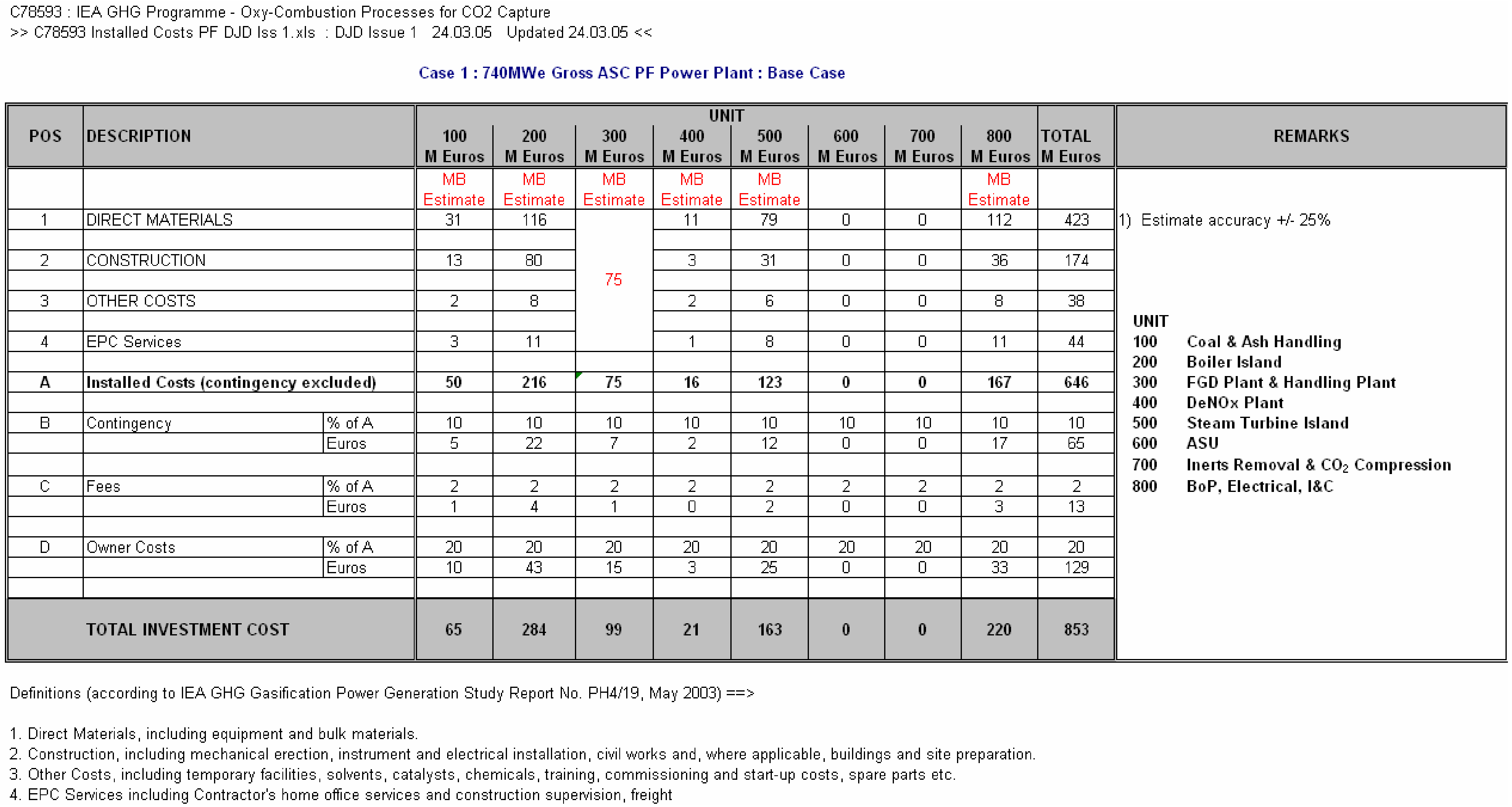
NGCC POWER PLANT
EFFECT OF CO_2 PRESSURE RATIO ON GT COMPRESSOR DESIGN

TABLE 4

GT Working Fluid **GT Pressure Ratio**		**Air** **20**	**CO_2** **20**	**CO_2** **30**
Gas Turbine Entry Temp.	K	1700	1570	1665
Gas Turbine Exit Temp.	K	853	910	912
Steam Turbine Inlet Steam Temp.	K	813	873	873
HRSG Flue Gas Exit Temp.	K	370	355	355

NGCC POWER PLANT
EFFECT OF CO_2 AND CYCLE PRESSURE RATIO ON GT TURBINE BOUNDARY CONDITIONS

TABLE 5



C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 Installed Costs PF DJD Iss 1.xls : DJD Issue 1 24.03.05 Updated 24.03.05 <<

Case 1 : 740MWe Gross ASC PF Power Plant : Base Case

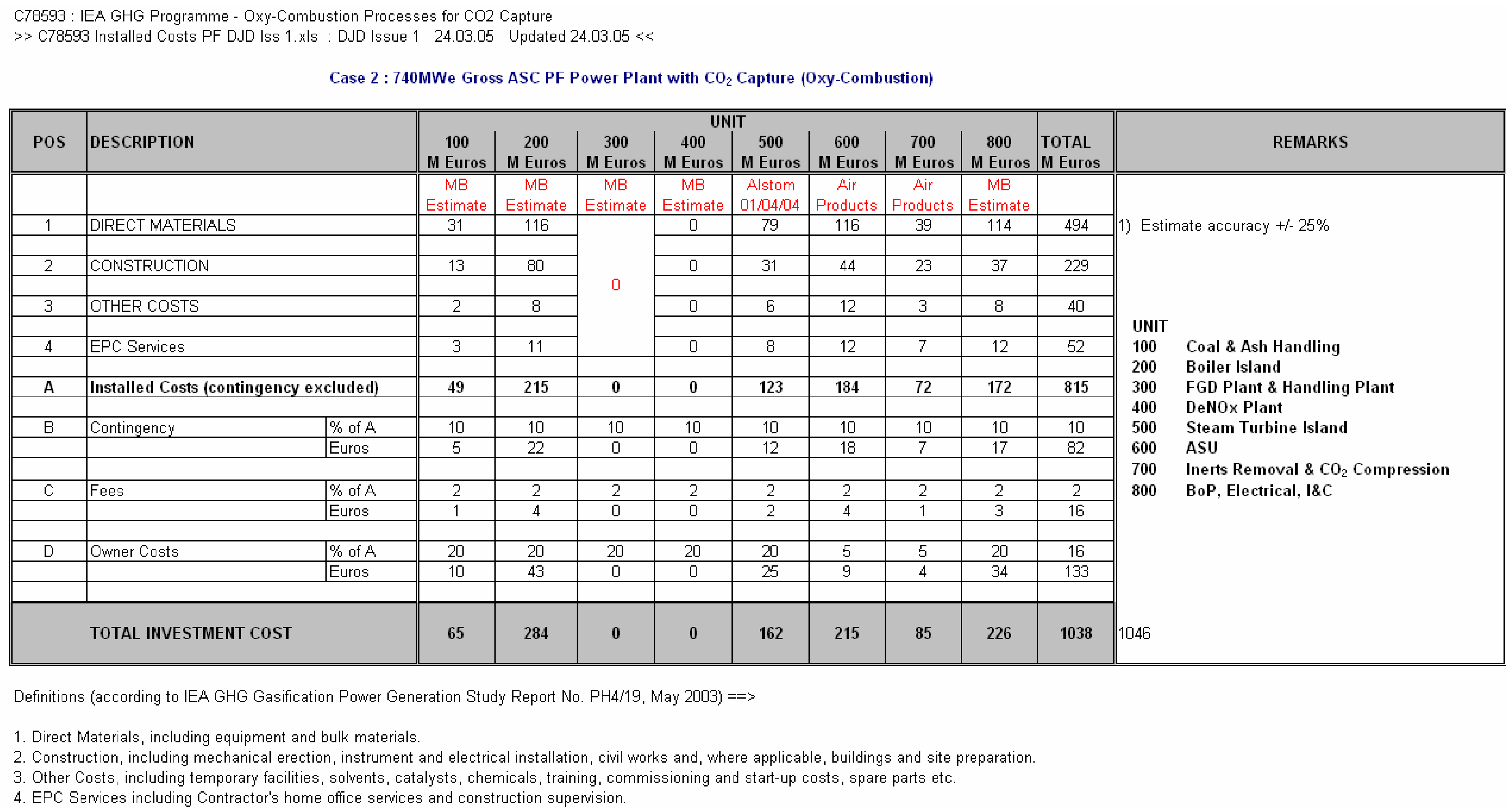
POS	DESCRIPTION		UNIT									REMARKS
			100 M Euros	200 M Euros	300 M Euros	400 M Euros	500 M Euros	600 M Euros	700 M Euros	800 M Euros	TOTAL M Euros	
			MB Estimate	MB Estimate	MB Estimate	MB Estimate	MB Estimate			MB Estimate		
1	DIRECT MATERIALS		31	116	75	11	79	0	0	112	423	1) Estimate accuracy +/- 25%
2	CONSTRUCTION		13	80		3	31	0	0	36	174	
3	OTHER COSTS		2	8		2	6	0	0	8	38	
												UNIT
4	EPC Services		3	11		1	8	0	0	11	44	**100 Coal & Ash Handling**
												200 Boiler Island
A	**Installed Costs (contingency excluded)**		**50**	**216**	**75**	**16**	**123**	**0**	**0**	**167**	**646**	**300 FGD Plant & Handling Plant**
												400 DeNOx Plant
B	Contingency	% of A	10	10	10	10	10	10	10	10	10	**500 Steam Turbine Island**
		Euros	5	22	7	2	12	0	0	17	65	**600 ASU**
												700 Inerts Removal & CO_2 Compression
C	Fees	% of A	2	2	2	2	2	2	2	2	2	**800 BoP, Electrical, I&C**
		Euros	1	4	1	0	2	0	0	3	13	
D	Owner Costs	% of A	20	20	20	20	20	20	20	20	20	
		Euros	10	43	15	3	25	0	0	33	129	
	TOTAL INVESTMENT COST		**65**	**284**	**99**	**21**	**163**	**0**	**0**	**220**	**853**	

Definitions (according to IEA GHG Gasification Power Generation Study Report No. PH4/19, May 2003) ==>

1. Direct Materials, including equipment and bulk materials.
2. Construction, including mechanical erection, instrument and electrical installation, civil works and, where applicable, buildings and site preparation.
3. Other Costs, including temporary facilities, solvents, catalysts, chemicals, training, commissioning and start-up costs, spare parts etc.
4. EPC Services including Contractor's home office services and construction supervision, freight

INSTALLED CAPITAL COSTS CASE 1: 740 MWe (Gross) ASC PF AIR-FIRED POWER PLANT WITHOUT CO_2 CAPTURE

TABLE 6



C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 Installed Costs PF DJD Iss 1.xls : DJD Issue 1 24.03.05 Updated 24.03.05 <<

Case 2 : 740MWe Gross ASC PF Power Plant with CO_2 Capture (Oxy-Combustion)

POS	DESCRIPTION		UNIT									REMARKS
			100 M Euros	200 M Euros	300 M Euros	400 M Euros	500 M Euros	600 M Euros	700 M Euros	800 M Euros	TOTAL M Euros	
			MB Estimate	MB Estimate	MB Estimate	MB Estimate	Alstom 01/04/04	Air Products	Air Products	MB Estimate		
1	DIRECT MATERIALS		31	116	0	0	79	116	39	114	494	1) Estimate accuracy +/- 25%
2	CONSTRUCTION		13	80		0	31	44	23	37	229	
3	OTHER COSTS		2	8		0	6	12	3	8	40	
												UNIT
4	EPC Services		3	11		0	8	12	7	12	52	**100 Coal & Ash Handling**
												200 Boiler Island
A	**Installed Costs (contingency excluded)**		**49**	**215**	**0**	**0**	**123**	**184**	**72**	**172**	**815**	**300 FGD Plant & Handling Plant**
												400 DeNOx Plant
B	Contingency	% of A	10	10	10	10	10	10	10	10	10	**500 Steam Turbine Island**
		Euros	5	22	0	0	12	18	7	17	82	**600 ASU**
												700 Inerts Removal & CO_2 Compression
C	Fees	% of A	2	2	2	2	2	2	2	2	2	**800 BoP, Electrical, I&C**
		Euros	1	4	0	0	2	4	1	3	16	
D	Owner Costs	% of A	20	20	20	20	20	5	5	20	16	
		Euros	10	43	0	0	25	9	4	34	133	
	TOTAL INVESTMENT COST		**65**	**284**	**0**	**0**	**162**	**215**	**85**	**226**	**1038**	1046

Definitions (according to IEA GHG Gasification Power Generation Study Report No. PH4/19, May 2003) ==>

1. Direct Materials, including equipment and bulk materials.
2. Construction, including mechanical erection, instrument and electrical installation, civil works and, where applicable, buildings and site preparation.
3. Other Costs, including temporary facilities, solvents, catalysts, chemicals, training, commissioning and start-up costs, spare parts etc.
4. EPC Services including Contractor's home office services and construction supervision.

INSTALLED CAPITAL COSTS CASE 2: 740 MWe (Gross) ASC PF OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE

TABLE 7

CASE 1: EQUIPMENT LIST

100 UNIT : COAL & ASH HANDLING

Coal delivery equipment
Bunkers
Yard equipment
Transfer towers
Dust suppression equipment
Ventilation equipment
Belt feeders
Metal detection
Belt weighing equipment
Miscellaneous equipment
Bottom ash systems
Fly ash systems

200 UNIT : BOILER ISLAND

Furnace
Reheater
Superheater
Economiser
Regenerative Airheaters
Piping
Air / Flue Gas Draught Plant
Structures
Fans: ID, FD and PA
Pumps
Coal feeders
Soot blowers
Drains systems
Dosing equipment
Mills
Auxiliary boiler
Miscellaneous equipment
Burners
ESP

300 UNIT : FGD PLANT & HANDLING

Priced as a turnkey package

400 UNIT : DENOx PLANT

Flues
Reactor casing
Catalyst
Ammonia injection equipment
Handling equipment
Control system

500 UNIT : STEAM TURBINE

HP, IP & LP Turbines
Associated Pipework
Feedwater heaters
Deaerator
Condenser
LP Pump
HP Pump
Sea water Circulation Pumps
Water treatment plant
Sea water inlet /outlet works

800 UNIT : BoP, ELECTRICAL, I&C

Balance of Power Plant
Controls
Instruments
Electrics

Buildings

CASE 1: ASC PF POWER PLANT WITHOUT CO$_2$ CAPTURE
OUTLINE EQUIPMENT LIST

TABLE 8

CASE 2: EQUIPMENT LIST

100 UNIT : COAL & ASH HANDLING

- Coal delivery equipment
- Bunkers
- Yard equipment
- Transfer towers
- Dust suppression equipment
- Ventilation equipment
- Belt feeders
- Metal detection
- Belt weighing equipment
- Miscellaneous equipment
- Bottom ash systems
- Fly ash systems

200 UNIT : BOILER ISLAND

- Furnace
- Reheater
- Superheater
- Economiser
- Regenerative Gas / Gas heaters
- Piping
- Flue gas recycle system
- Structures
- Fans: ID, FD and PA
- Pumps
- Coal feeders
- Soot blowers
- Drains systems
- Dosing equipment
- Mills
- Auxiliary boiler
- Miscellaneous equipment
- Burners
- ESP

500 UNIT : STEAM TURBINE

- HP, IP & LP Turbines
- Associated Pipework
- Feedwater heaters
- Deaerator
- Condenser
- LP Pump
- HP Pump
- Sea water Circulation Pumps
- Water treatment plant
- Sea water inlet /outlet works

600 UNIT : ASU

- Main air compressors
- Air purification system
- main heat exchanger
- ASU compander
- ASU Column System
- Pumps
- Backup storage vessel

700 UNIT : INERTS REMOVAL & CO_2

- Venturi scrubber
- Indirect contact cooler
- Direct contact seawater cooler
- Compressors
- Heat exchangers
- Flue gas expander

800 UNIT : BoP, ELECTRICAL, I&C

- Balance of Power Plant
- Controls
- Instruments
- Electrics

Buildings

CASE 2: ASC PF POWER PLANT WITH CO_2 CAPTURE
OUTLINE EQUIPMENT LIST

TABLE 9

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 Installed Costs PF DJD Iss 1.xls : DJD Issue 1 24.03.05 Updated 24.03.05 <<

Estimate Summary

UNIT	DESCRIPTION	Case 1 PF Power Plant Base Case		Case 2 PF Power Plant with CO_2 Capture	
		M Euros	%	M Euros	%
100	Coal and Ash Handling	65	8	65	6
200	Boiler Island	284	33	284	27
300	FGD Plant	99	12	0	0
400	DeNOx Plant	21	2	0	0
500	Steam Turbine Island	163	19	162	16
600	ASU	0	0	215	21
700	CO2 Compression & Drying	0	0	85	8
800	BoP, Electrical, I&C	220	26	226	22
	TOTAL INVESTMENT COST	**853**	**100**	**1038**	**100**

GROSS POWER OUTPUT	**MWe**	**740**	**737**
NET POWER OUTPUT	**MWe**	**677**	**532**
SPECIFIC INVESTMENT COST	**Euro/kWe Gross**	**1152**	**1408**
SPECIFIC INVESTMENT COST	**Euro/kWe Net**	**1260**	**1951**

CASE 1 & 2 : ASC PF POWER PLANT
PERCENTAGE CONTRIBUTION OF EACH UNIT TO TOTAL INVESTMENT COST

TABLE 10

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 O&M Costs PF DJD Iss 1.xls : DJD Issue 1 07.04.05 Updated Issue 1 07.04.05 <<

Yearly Variable Costs : Case 1 and Case 2

Yearly Operating hours =	**7446**	**Case 1**			**Case 2**		
		740 MWe PF Power Plant Base Case			**740 MWe PF Power Plant Oxy-Combustion & CO_2 Capture**		
Consumables	**Unit Cost €/t**	**Consumption Hourly kg/h**	**Consumption Yearly t/y**	**Operating Costs (yearly basis)**	**Consumption Hourly kg/h**	**Consumption Yearly t/y**	**Operating Costs (yearly basis)**
Fuel Fired (kg/s)		59.164			58.090		
Feedstock							
Coal (see notes 1, 2)	38.8	212,990	1,585,927	61,541,879	209,124	1,557,137	60,424,713
Auxiliary Feedstock							
FGD Make-up water	0.1	41,012	305,376	30,538	0	0	0
Solvents							
DeNOx Catalyst							
Plate-type Catalyst (3 years life) (see note 3)				2,960,000			0
Chemicals							
Limestone (see note 4)	17.0	5,694	42,401	720,813	0	0	0
Ammonia (see note 5)	330.8	374	2,788	922,201	0	0	0
Waste Disposal							
Bottom Ash	0.0	6,496	48,371	0	6,378	47,493	0
Fly Ash	0.0	19,489	145,112	0	19,135	142,478	0
Gypsum	0.0	10,324	76,875	0	0	0	0
FGD Chloride Purge	0.0	530	3,950	0	0	0	0
Miscellaneous (see note 6)				275,502			275,502
TOTAL YEARLY OPERATING COSTS				**66,450,932**			**60,700,215**

Notes
1 - Exchange Rate assumed as $1= €1.0
2 - Based on $1.5/GJ and Fuel NCV = 25870 kJ/kg from IEA GHG Tech Spec IEA/CON/03/97
3 - Based on $300/ft^3 from World Bank
4 - Based on $15/ton from National Lime Association January 2003
5 - Based on $2.55/lb-mole nitrogen from EPRI
6 - Based on €0.05/MWh

CASE 1 & 2 : ASC PF POWER PLANT - CONSUMABLES

TABLE 11

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 O&M Costs PF DJD Iss 1.xls : DJD Issue 1 07.04.05 Updated Issue 1 07.04.05 <<

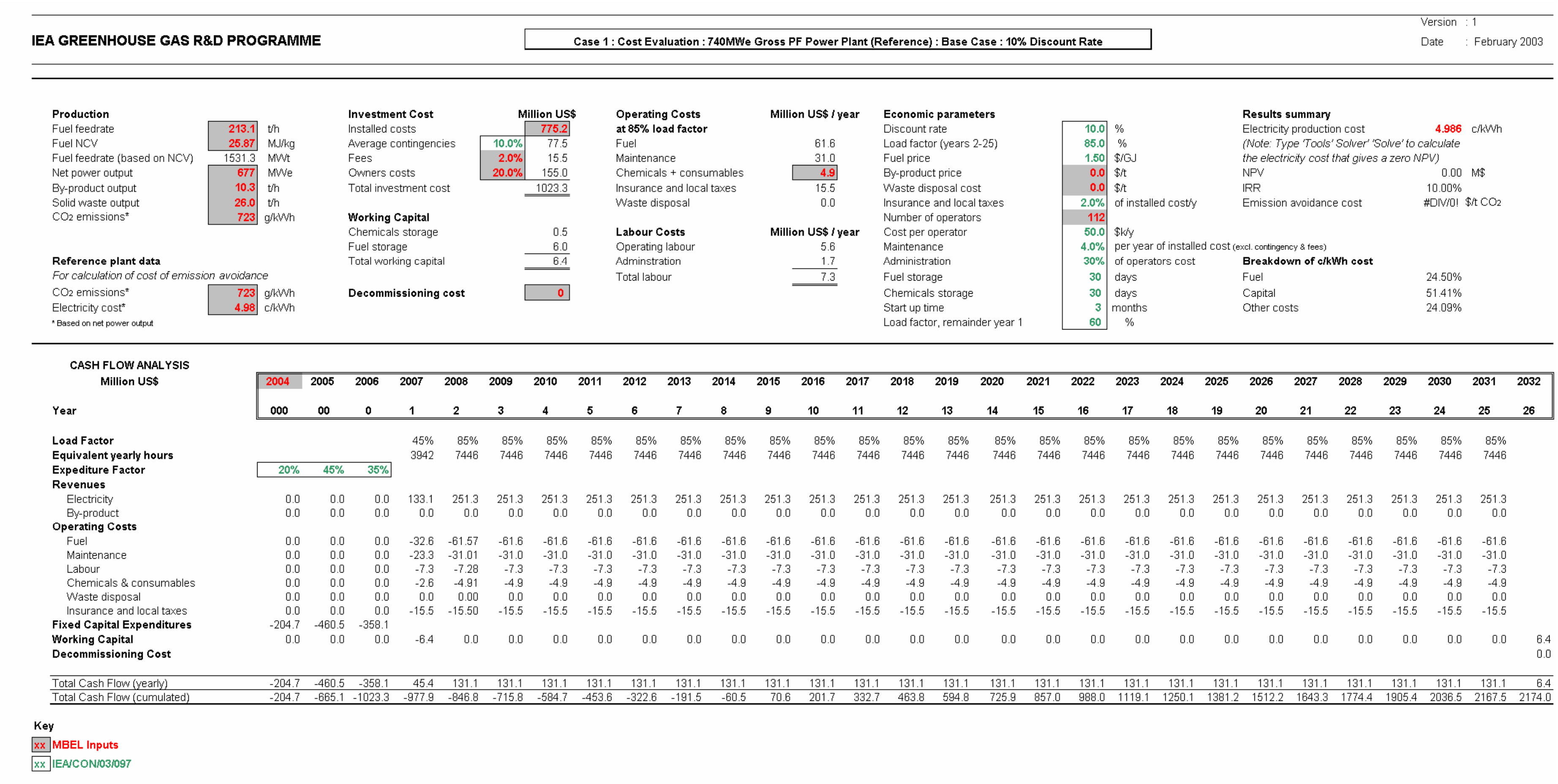
Total O&M Costs : Case 3 and Case 4c

POS	DESCRIPTION	Case 1 740 MWe Base Case €/year	Case 2 740 MWe CO_2 Capture €/year
	No. of Operators	112	136
A1	Direct Labour (see note 1, 2)	5,600,000	6,800,000
A2	Administration / General Overheads (30% of A1) (see Note 3)	1,680,000	2,040,000
A3	Maintenance (4% of **Installed costs** (contingency excluded))	25,840,000	32,600,000
A	**FIXED COSTS**	**33,120,000**	**41,440,000**
B	**VARIABLE COSTS**	**66,451,000**	**60,700,000**
	TOTAL OPERATION AND MAINTENANCE COSTS	**99,571,000**	**102,140,000**

Notes
1 - Base Case : €7/kW per year and 8 FGD Operators from National Lime Association January 2003 assumed at €50,000 per year
2 - CO2 Capture Case : €7/kW per year and 32 ASU+CO2 Plant Operators from Air Products assumed at €50,000 per year
3 - Based on EPRI, Technical Assessment Guide for the Power Industry

CASE 1 & 2: ASC PF POWER PLANT
TOTAL OPERATING AND MAINTENANCE COSTS

TABLE 12



IEA GREENHOUSE GAS R&D PROGRAMME

Case 1 : Cost Evaluation : 740MWe Gross PF Power Plant (Reference) : Base Case : 10% Discount Rate

Version : 1
Date : February 2003

Production		
Fuel feedrate	213.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1531.3	MWt
Net power output	677	MWe
By-product output	10.3	t/h
Solid waste output	26.0	t/h
CO_2 emissions*	723	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	4.98	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		775.2
Average contingencies	10.0%	77.5
Fees	2.0%	15.5
Owners costs	20.0%	155.0
Total investment cost		1023.3
Working Capital		
Chemicals storage		0.5
Fuel storage		6.0
Total working capital		6.4
Decommissioning cost		0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	61.6
Maintenance	31.0
Chemicals + consumables	4.9
Insurance and local taxes	15.5
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	5.6
Adminstration	1.7
Total labour	7.3

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	1.50	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	112	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	4.986	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2
Breakdown of c/kWh cost		
Fuel	24.50%	
Capital	51.41%	
Other costs	24.09%	

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electricity	0.0	0.0	0.0	133.1	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-32.6	-61.57	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6
Maintenance	0.0	0.0	0.0	-23.3	-31.01	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0
Labour	0.0	0.0	0.0	-7.3	-7.28	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3
Chemicals & consumables	0.0	0.0	0.0	-2.6	-4.91	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-15.5	-15.50	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5
Fixed Capital Expenditures	-204.7	-460.5	-358.1												
Working Capital	0.0	0.0	0.0	-6.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-204.7	-460.5	-358.1	45.4	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1
Total Cash Flow (cumulated)	-204.7	-665.1	-1023.3	-977.9	-846.8	-715.8	-584.7	-453.6	-322.6	-191.5	-60.5	70.6	201.7	332.7	463.8

	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electricity	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	251.3	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	
Maintenance	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	
Labour	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	
Chemicals & consumables	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	
Fixed Capital Expenditures														
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	6.4
Decommissioning Cost														0.0
Total Cash Flow (yearly)	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	131.1	6.4
Total Cash Flow (cumulated)	594.8	725.9	857.0	988.0	1119.1	1250.1	1381.2	1512.2	1643.3	1774.4	1905.4	2036.5	2167.5	2174.0

Key

xx MBEL Inputs
xx IEA/CON/03/097

CASE 1: ASC PF AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE

TABLE 13

Version : 1
Date : February 2003

IEA GREENHOUSE GAS R&D PROGRAMME

Case 1 : Cost Evaluation : 740MWe Gross PF Power Plant (Reference) : 10% Discount Rate & +100% Fuel Price

Production		
Fuel feedrate	213.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1531.3	MWt
Net power output	677	MWe
By-product output	10.3	t/h
Solid waste output	26.0	t/h
CO_2 emissions*	723	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	4.98	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		775.2
Average contingencies	10.0%	77.5
Fees	2.0%	15.5
Owners costs	20.0%	155.0
Total investment cost		1023.3
Working Capital		
Chemicals storage		0.5
Fuel storage		11.9
Total working capital		12.4
Decommissioning cost		0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	123.1
Maintenance	31.0
Chemicals + consumables	4.9
Insurance and local taxes	15.5
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	5.6
Adminstration	1.7
Total labour	7.3

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	112	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	6.219	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2
Breakdown of c/kWh cost		
Fuel	39.28%	
Capital	41.40%	
Other costs	19.32%	

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electricity	0.0	0.0	0.0	166.0	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-65.2	-123.15	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1
Maintenance	0.0	0.0	0.0	-23.3	-31.01	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0
Labour	0.0	0.0	0.0	-7.3	-7.28	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3
Chemicals & consumables	0.0	0.0	0.0	-2.6	-4.91	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-15.5	-15.50	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5
Fixed Capital Expenditures	-204.7	-460.5	-358.1												
Working Capital	0.0	0.0	0.0	-12.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-204.7	-460.5	-358.1	39.7	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6
Total Cash Flow (cumulated)	-204.7	-665.1	-1023.3	-983.5	-851.9	-720.3	-588.6	-457.0	-325.4	-193.8	-62.1	69.5	201.1	332.7	464.4

	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electricity	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	313.5	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	-123.1	
Maintenance	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	
Labour	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	
Chemicals & consumables	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	
Fixed Capital Expenditures														
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	12.4
Decommissioning Cost														0.0
Total Cash Flow (yearly)	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	131.6	12.4
Total Cash Flow (cumulated)	596.0	727.6	859.3	990.9	1122.5	1254.1	1385.8	1517.4	1649.0	1780.6	1912.3	2043.9	2175.5	2187.9

Key

xx	MBEL Inputs
xx	IEA/CON/03/097

CASE 1: ASC PF AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE & +100% FUEL PRICE

TABLE 14

Version : 1
Date : February 2003

IEA GREENHOUSE GAS R&D PROGRAMME

Case 1 : Cost Evaluation : 740MWe Gross PF Power Plant (Reference) : 10% Discount Rate & -100% Fuel Price

Production		
Fuel feedrate	213.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1531.3	MWt
Net power output	677	MWe
By-product output	10.3	t/h
Solid waste output	26.0	t/h
CO_2 emissions*	723	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	4.98	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		775.2
Average contingencies	10.0%	77.5
Fees	2.0%	15.5
Owners costs	20.0%	155.0
Total investment cost		1023.3

Working Capital	
Chemicals storage	0.5
Fuel storage	0.0
Total working capital	0.5
Decommissioning cost	0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	0.0
Maintenance	31.0
Chemicals + consumables	4.9
Insurance and local taxes	15.5
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	5.6
Adminstration	1.7
Total labour	7.3

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	0.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	112	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	3.753	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2

Breakdown of c/kWh cost	
Fuel	0.00%
Capital	68.00%
Other costs	32.00%

CASH FLOW ANALYSIS
Million US$

	2004	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	100.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Maintenance	0.0	0.0	0.0	-23.3	-31.01	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0
Labour	0.0	0.0	0.0	-7.3	-7.28	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3
Chemicals & consumables	0.0	0.0	0.0	-2.6	-4.91	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-15.5	-15.50	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5
Fixed Capital Expenditures	-204.7	-460.5	-358.1											
Working Capital	0.0	0.0	0.0	-0.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-204.7	-460.5	-358.1	51.0	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5
Total Cash Flow (cumulated)	-204.7	-665.1	-1023.3	-972.2	-841.7	-711.2	-580.7	-450.3	-319.8	-189.3	-58.8	71.7	202.2	332.7

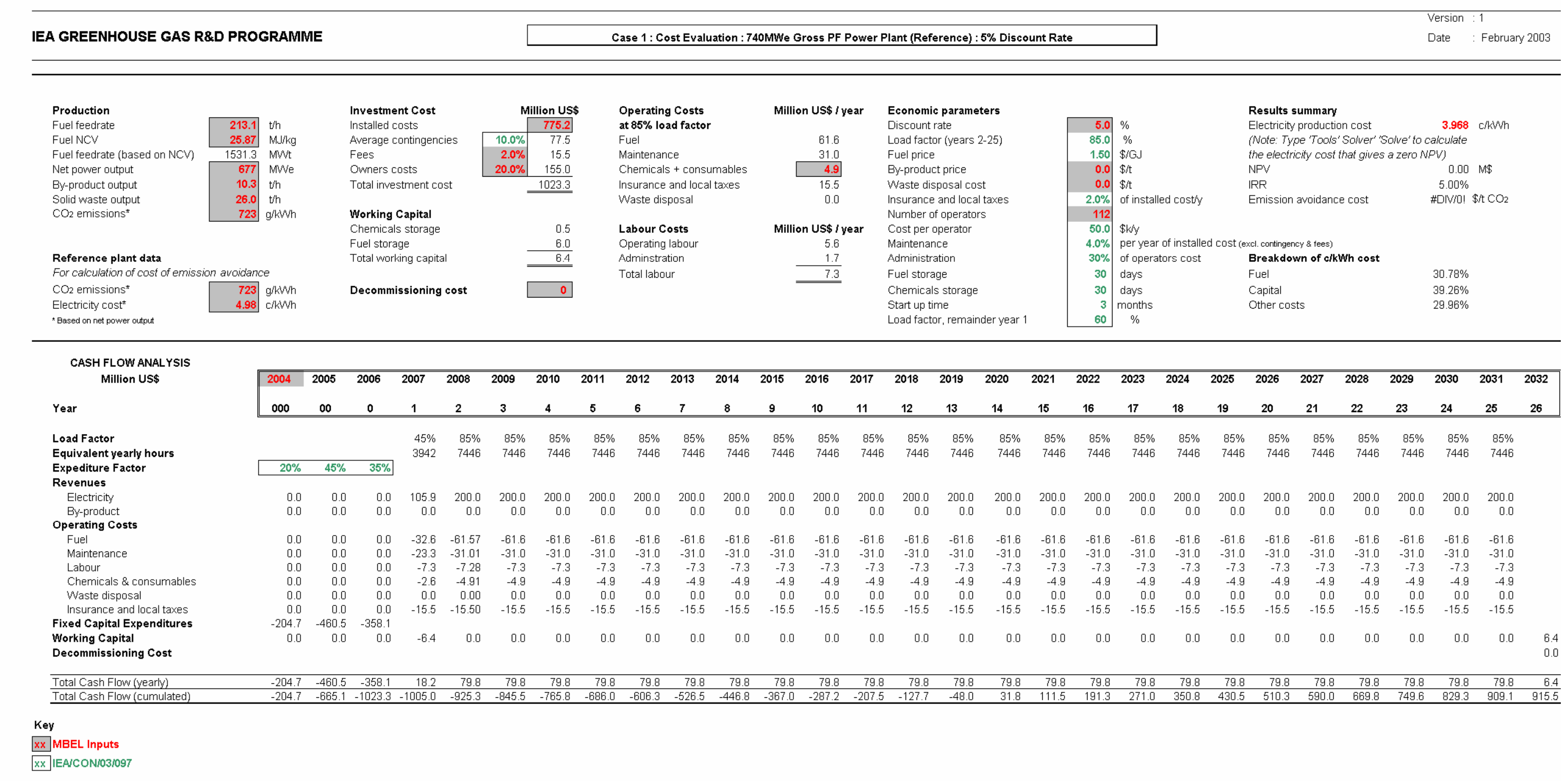
	2018	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	189.2	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Maintenance	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	
Labour	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	
Chemicals & consumables	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.5
Decommissioning Cost															0.0
Total Cash Flow (yearly)	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	130.5	0.5
Total Cash Flow (cumulated)	463.2	593.7	724.2	854.7	985.1	1115.6	1246.1	1376.6	1507.1	1637.6	1768.1	1898.6	2029.1	2159.6	2160.0

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 1: ASC PF AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE & -100% FUEL PRICE

TABLE 15



IEA GREENHOUSE GAS R&D PROGRAMME

Case 1 : Cost Evaluation : 740MWe Gross PF Power Plant (Reference) : 5% Discount Rate

Version : 1
Date : February 2003

Production		
Fuel feedrate	213.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1531.3	MWt
Net power output	677	MWe
By-product output	10.3	t/h
Solid waste output	26.0	t/h
CO_2 emissions*	723	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	4.98	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		775.2
Average contingencies	10.0%	77.5
Fees	2.0%	15.5
Owners costs	20.0%	155.0
Total investment cost		1023.3

Working Capital	
Chemicals storage	0.5
Fuel storage	6.0
Total working capital	6.4

Decommissioning cost	0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	61.6
Maintenance	31.0
Chemicals + consumables	4.9
Insurance and local taxes	15.5
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	5.6
Adminstration	1.7
Total labour	7.3

Economic parameters		
Discount rate	5.0	%
Load factor (years 2-25)	85.0	%
Fuel price	1.50	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	112	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	3.968	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	5.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2

Breakdown of c/kWh cost	
Fuel	30.78%
Capital	39.26%
Other costs	29.96%

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	105.9	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-32.6	-61.57	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6
Maintenance	0.0	0.0	0.0	-23.3	-31.01	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0
Labour	0.0	0.0	0.0	-7.3	-7.28	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3
Chemicals & consumables	0.0	0.0	0.0	-2.6	-4.91	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-15.5	-15.50	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5
Fixed Capital Expenditures	-204.7	-460.5	-358.1											
Working Capital	0.0	0.0	0.0	-6.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-204.7	-460.5	-358.1	18.2	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8
Total Cash Flow (cumulated)	-204.7	-665.1	-1023.3	-1005.0	-925.3	-845.5	-765.8	-686.0	-606.3	-526.5	-446.8	-367.0	-287.2	-207.5

	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	200.0	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	
Maintenance	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	-31.0	
Labour	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	
Chemicals & consumables	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	-15.5	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	6.4
Decommissioning Cost															0.0
Total Cash Flow (yearly)	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	79.8	6.4
Total Cash Flow (cumulated)	-127.7	-48.0	31.8	111.5	191.3	271.0	350.8	430.5	510.3	590.0	669.8	749.6	829.3	909.1	915.5

Key

▨ xx	MBEL Inputs
☐ xx	IEA/CON/03/097

CASE 1: ASC PF AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 5% DISCOUNT RATE

TABLE 16

Version : 1

IEA GREENHOUSE GAS R&D PROGRAMME

Case 1 : Cost Evaluation : 740MWe Gross PF Power Plant (Reference) : 10% Discount Rate & -20% CAPEX

Date : February 2003

Production		
Fuel feedrate	213.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1531.3	MWt
Net power output	677	MWe
By-product output	10.3	t/h
Solid waste output	26.0	t/h
CO_2 emissions*	723	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	4.98	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		620.2
Average contingencies	10.0%	62.0
Fees	2.0%	12.4
Owners costs	20.0%	124.0
Total investment cost		818.6
Working Capital		
Chemicals storage		0.5
Fuel storage		6.0
Total working capital		6.4
Decommissioning cost		0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	61.6
Maintenance	24.8
Chemicals + consumables	4.9
Insurance and local taxes	12.4
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	5.6
Adminstration	1.7
Total labour	7.3

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	1.50	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	112	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	4.285	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2
Breakdown of c/kWh cost		
Fuel	28.50%	
Capital	47.91%	
Other costs	23.59%	

CASH FLOW ANALYSIS
Million US$

	2004	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	114.4	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-32.6	-61.57	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6
Maintenance	0.0	0.0	0.0	-18.6	-24.81	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8
Labour	0.0	0.0	0.0	-7.3	-7.28	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3
Chemicals & consumables	0.0	0.0	0.0	-2.6	-4.91	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-12.4	-12.40	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4
Fixed Capital Expenditures	-163.7	-368.4	-286.5											
Working Capital	0.0	0.0	0.0	-6.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-163.7	-368.4	-286.5	34.4	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0
Total Cash Flow (cumulated)	-163.7	-532.1	-818.6	-784.2	-679.1	-574.1	-469.0	-364.0	-259.0	-153.9	-48.9	56.2	161.2	266.3

	2018	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	216.0	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	-61.6	
Maintenance	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	-24.8	
Labour	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	-7.3	
Chemicals & consumables	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	-4.9	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	-12.4	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	6.4
Decommissioning Cost															0.0
Total Cash Flow (yearly)	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	105.0	6.4
Total Cash Flow (cumulated)	371.3	476.3	581.4	686.4	791.5	896.5	1001.5	1106.6	1211.6	1316.7	1421.7	1526.8	1631.8	1736.8	1743.3

Key

- xx **MBEL Inputs**
- xx **IEA/CON/03/097**

CASE 1: ASC PF AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE & -20% TIC

TABLE 17

IEA GREENHOUSE GAS R&D PROGRAMME

Case 2 : Cost Evaluation : 740MWe Gross PF Power Plant with CO_2 Capture : Base Case : 10% Discount Rate

Version : 1
Date : February 2003

Production		
Fuel feedrate	209.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1502.8	MWt
Net power output	532	MWe
By-product output	0.0	t/h
Solid waste output	25.5	t/h
CO_2 emissions*	84	g/kWh

Reference plant data
For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	4.98	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		978.0
Average contingencies	10.0%	97.8
Fees	2.0%	19.6
Owners costs	16.0%	156.5
Total investment cost		1251.8
Working Capital		
Chemicals storage		0.0
Fuel storage		5.8
Total working capital		5.9
Decommissioning cost		0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	60.4
Maintenance	39.1
Chemicals + consumables	0.3
Insurance and local taxes	19.6
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	6.8
Adminstration	2.0
Total labour	8.8

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	1.50	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	136	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	7.281	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	36.0	$/t CO_2
Breakdown of c/kWh cost		
Fuel	20.95%	
Capital	54.74%	
Other costs	24.31%	

CASH FLOW ANALYSIS

Million US$	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	152.7	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-32.0	-60.42	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4
Maintenance	0.0	0.0	0.0	-29.3	-39.12	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1
Labour	0.0	0.0	0.0	-8.8	-8.84	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.28	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-19.6	-19.56	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6
Fixed Capital Expenditures	-250.4	-563.3	-438.1											
Working Capital	0.0	0.0	0.0	-5.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-250.4	-563.3	-438.1	56.9	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2
Total Cash Flow (cumulated)	-250.4	-813.7	-1251.8	-1194.9	-1034.7	-874.5	-714.3	-554.1	-393.9	-233.7	-73.5	86.7	246.9	407.1

Million US$	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	288.4	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	
Maintenance	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	
Labour	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	
Chemicals & consumables	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	5.9
Decommissioning Cost															0.0
Total Cash Flow (yearly)	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	160.2	5.9
Total Cash Flow (cumulated)	567.3	727.5	887.7	1047.9	1208.1	1368.3	1528.5	1688.7	1848.9	2009.1	2169.3	2329.5	2489.7	2649.9	2655.7

Key

xx	MBEL Inputs
xx	IEA/CON/03/097

CASE 2: ASC PF OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : 10% DISCOUNT RATE

TABLE 18

IEA GREENHOUSE GAS R&D PROGRAMME

Case 2 : Cost Evaluation : 740MWe Gross PF Power Plant with CO_2 Capture : 10% Discount Rate & +100% Fuel Price

Version : 1
Date : February 2003

Production		
Fuel feedrate	209.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1502.8	MWt
Net power output	532	MWe
By-product output	0.0	t/h
Solid waste output	25.5	t/h
CO_2 emissions*	84	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	6.22	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		978.0
Average contingencies	10.0%	97.8
Fees	2.0%	19.6
Owners costs	16.0%	156.5
Total investment cost		1251.8

Working Capital	
Chemicals storage	0.0
Fuel storage	11.7
Total working capital	11.7

Decommissioning cost 0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	120.8
Maintenance	39.1
Chemicals + consumables	0.3
Insurance and local taxes	19.6
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	6.8
Adminstration	2.0
Total labour	8.8

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	136	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	8.820	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	40.7	$/t CO_2

Breakdown of c/kWh cost	
Fuel	34.59%
Capital	45.35%
Other costs	20.07%

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	185.0	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-64.0	-120.85	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8
Maintenance	0.0	0.0	0.0	-29.3	-39.12	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1
Labour	0.0	0.0	0.0	-8.8	-8.84	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.28	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-19.6	-19.56	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6
Fixed Capital Expenditures	-250.4	-563.3	-438.1											
Working Capital	0.0	0.0	0.0	-11.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-250.4	-563.3	-438.1	51.4	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8
Total Cash Flow (cumulated)	-250.4	-813.7	-1251.8	-1200.4	-1039.7	-878.9	-718.2	-557.4	-396.7	-235.9	-75.2	85.6	246.4	407.1

	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	349.4	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	-120.8	
Maintenance	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	
Labour	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	
Chemicals & consumables	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	11.7
Decommissioning Cost															0.0
Total Cash Flow (yearly)	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	160.8	11.7
Total Cash Flow (cumulated)	567.9	728.6	889.4	1050.1	1210.9	1371.6	1532.4	1693.2	1853.9	2014.7	2175.4	2336.2	2496.9	2657.7	2669.4

Key

xx	MBEL Inputs
xx	IEA/CON/03/097

CASE 2: ASC PF OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : 10% DISCOUNT RATE & +100% FUEL PRICE

TABLE 19

IEA GREENHOUSE GAS R&D PROGRAMME

Case 2 : Cost Evaluation : 740MWe Gross PF Power Plant with CO_2 Capture : 10% Discount Rate & -100% Fuel Price

Version : 1
Date : February 2003

Production		
Fuel feedrate	209.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1502.8	MWt
Net power output	532	MWe
By-product output	0.0	t/h
Solid waste output	25.5	t/h
CO_2 emissions*	84	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	3.75	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		978.0
Average contingencies	10.0%	97.8
Fees	2.0%	19.6
Owners costs	16.0%	156.5
Total investment cost		1251.8

Working Capital	
Chemicals storage	0.0
Fuel storage	0.0
Total working capital	0.0

Decommissioning cost	0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	0.0
Maintenance	39.1
Chemicals + consumables	0.3
Insurance and local taxes	19.6
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	6.8
Adminstration	2.0
Total labour	8.8

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	0.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	136	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	5.741	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	31.2	$/t CO_2

Breakdown of c/kWh cost	
Fuel	0.00%
Capital	69.17%
Other costs	30.83%

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	120.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Maintenance	0.0	0.0	0.0	-29.3	-39.12	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1
Labour	0.0	0.0	0.0	-8.8	-8.84	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.28	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-19.6	-19.56	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6
Fixed Capital Expenditures	-250.4	-563.3	-438.1											
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-250.4	-563.3	-438.1	62.5	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6
Total Cash Flow (cumulated)	-250.4	-813.7	-1251.8	-1189.3	-1029.7	-870.1	-710.4	-550.8	-391.1	-231.5	-71.9	87.8	247.4	407.1

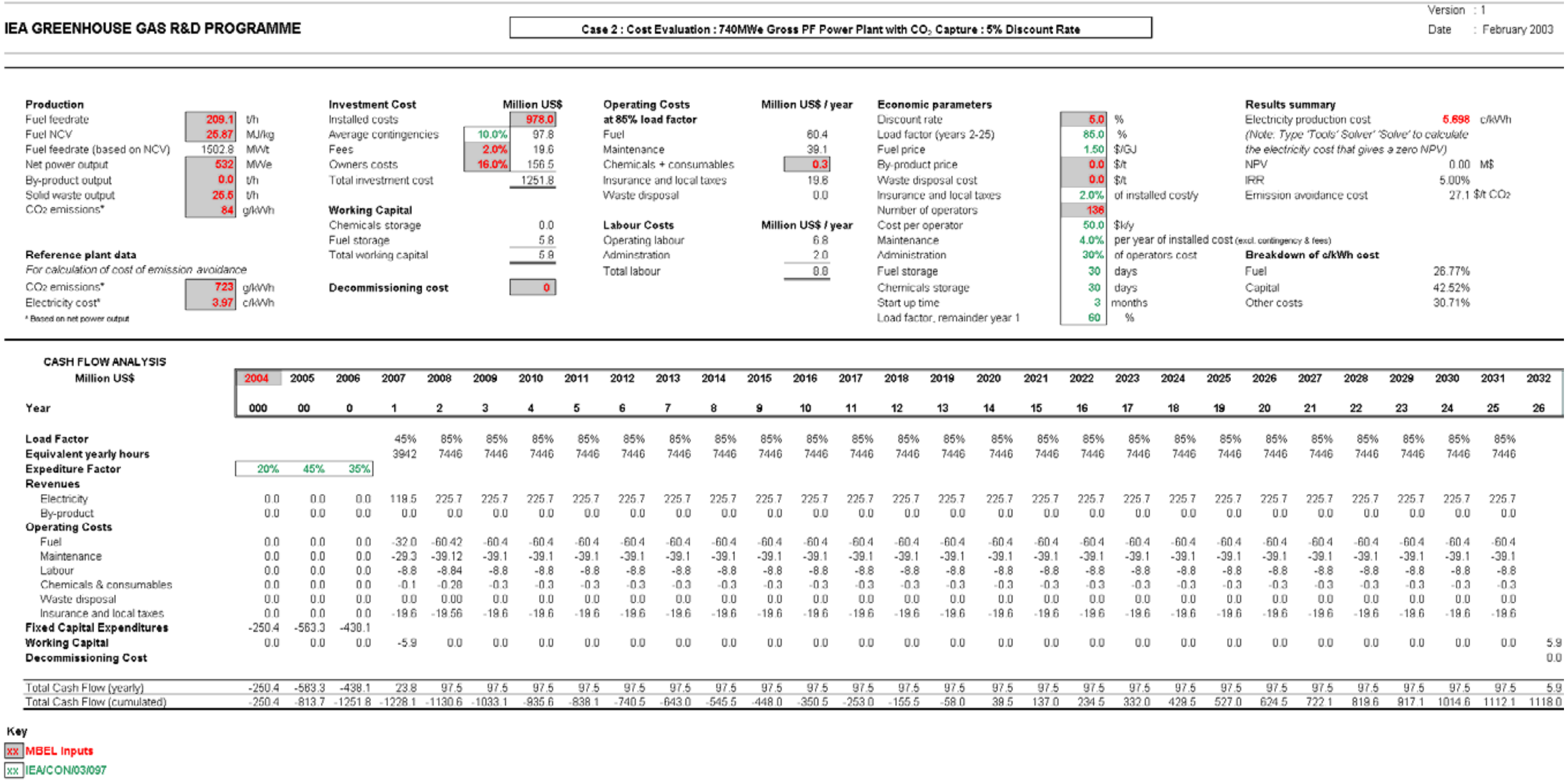
	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	227.4	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Maintenance	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	
Labour	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	
Chemicals & consumables	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															0.0
Total Cash Flow (yearly)	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	159.6	0.0
Total Cash Flow (cumulated)	566.7	726.3	886.0	1045.6	1205.3	1364.9	1524.5	1684.2	1843.8	2003.5	2163.1	2322.7	2482.4	2642.0	2642.1

Key

xx **MBEL Inputs**
xx **IEA/CON/03/097**

CASE 2: ASC PF OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : 10% DISCOUNT RATE & -100% FUEL PRICE

TABLE 20



IEA GREENHOUSE GAS R&D PROGRAMME

Case 2 : Cost Evaluation : 740MWe Gross PF Power Plant with CO_2 Capture : 5% Discount Rate

Version : 1
Date : February 2003

Production		
Fuel feedrate	209.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1502.8	MWt
Net power output	532	MWe
By-product output	0.0	t/h
Solid waste output	25.5	t/h
CO_2 emissions*	84	g/kWh

Reference plant data
For calculation of cost of emission avoidance

CO_2 emissions*	723	g/kWh
Electricity cost*	3.97	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		978.0
Average contingencies	10.0%	97.8
Fees	2.0%	19.6
Owners costs	16.0%	156.5
Total investment cost		1251.8
Working Capital		
Chemicals storage		0.0
Fuel storage		5.8
Total working capital		5.8
Decommissioning cost		0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	60.4
Maintenance	39.1
Chemicals + consumables	0.3
Insurance and local taxes	19.6
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	6.8
Administration	2.0
Total labour	8.8

Economic parameters		
Discount rate	5.0	%
Load factor (years 2-25)	85.0	%
Fuel price	1.50	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	136	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	5.698	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	5.00%	
Emission avoidance cost	27.1	$/t CO_2

Breakdown of c/kWh cost	
Fuel	26.77%
Capital	42.52%
Other costs	30.71%

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electricity	0.0	0.0	0.0	119.5	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-32.0	-60.42	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4
Maintenance	0.0	0.0	0.0	-29.3	-39.12	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1
Labour	0.0	0.0	0.0	-8.8	-8.84	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.28	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-19.6	-19.56	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6
Fixed Capital Expenditures	-250.4	-563.3	-438.1												
Working Capital	0.0	0.0	0.0	-5.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-250.4	-563.3	-438.1	23.8	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5
Total Cash Flow (cumulated)	-250.4	-813.7	-1251.8	-1228.1	-1130.6	-1033.1	-935.6	-838.1	-740.5	-643.0	-545.5	-448.0	-350.5	-253.0	-155.5

	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electricity	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	225.7	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	
Maintenance	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	-39.1	
Labour	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	
Chemicals & consumables	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	-19.6	
Fixed Capital Expenditures														
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	5.9
Decommissioning Cost														0.0
Total Cash Flow (yearly)	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	97.5	5.9
Total Cash Flow (cumulated)	-58.0	39.5	137.0	234.5	332.0	429.5	527.0	624.5	722.1	819.6	917.1	1014.6	1112.1	1118.0

Key

xx	MBEL Inputs
xx	IEA/CON/03/097

CASE 2: ASC PF OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE:5% DISCOUNT RATE TABLE 21

IEA GREENHOUSE GAS R&D PROGRAMME

Case 2 : Cost Evaluation : 740MWe Gross PF Power Plant with CO_2 Capture : 10% Discount Rate & -20%CAPEX

Version : 1
Date : February 2003

Production		
Fuel feedrate	209.1	t/h
Fuel NCV	25.87	MJ/kg
Fuel feedrate (based on NCV)	1502.8	MWt
Net power output	532	MWe
By-product output	0.0	t/h
Solid waste output	25.5	t/h
CO2 emissions*	84	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO2 emissions*	723	g/kWh
Electricity cost*	4.29	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		782.4
Average contingencies	10.0%	78.2
Fees	2.0%	15.6
Owners costs	16.0%	125.2
Total investment cost		1001.5

Working Capital	
Chemicals storage	0.0
Fuel storage	5.8
Total working capital	5.9

Decommissioning cost	0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	60.4
Maintenance	31.3
Chemicals + consumables	0.3
Insurance and local taxes	15.6
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	6.8
Adminstration	2.0
Total labour	8.8

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	1.50	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	136	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	30	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	6.181	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	29.6	$/t CO2

Breakdown of c/kWh cost	
Fuel	24.68%
Capital	51.63%
Other costs	23.69%

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	129.6	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-32.0	-60.42	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4
Maintenance	0.0	0.0	0.0	-23.5	-31.30	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3
Labour	0.0	0.0	0.0	-8.8	-8.84	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.28	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-15.6	-15.65	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6
Fixed Capital Expenditures	-200.3	-450.7	-350.5											
Working Capital	0.0	0.0	0.0	-5.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-200.3	-450.7	-350.5	43.7	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4
Total Cash Flow (cumulated)	-200.3	-651.0	-1001.5	-957.8	-829.5	-701.1	-572.7	-444.4	-316.0	-187.7	-59.3	69.0	197.4	325.8

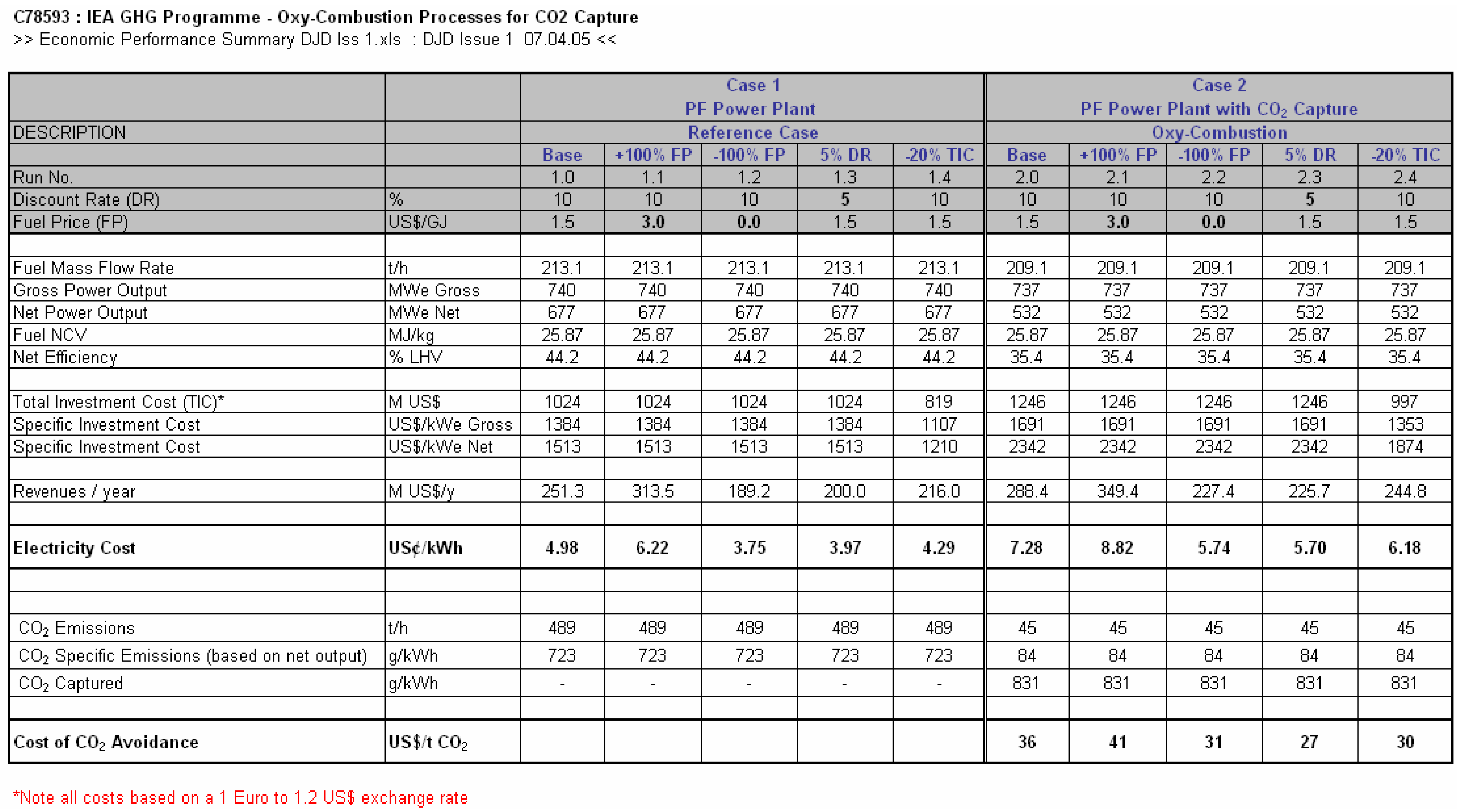
	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	244.8	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	-60.4	
Maintenance	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	-31.3	
Labour	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	
Chemicals & consumables	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	-15.6	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	5.9
Decommissioning Cost															0.0
Total Cash Flow (yearly)	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	128.4	5.9
Total Cash Flow (cumulated)	454.1	582.5	710.8	839.2	967.6	1095.9	1224.3	1352.6	1481.0	1609.3	1737.7	1866.1	1994.4	2122.8	2128.6

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 2: ASC PF POWER PLANT WITH CO_2 CAPTURE : 10% DISCOUNT RATE & -20% TIC

TABLE 22



C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> Economic Performance Summary DJD Iss 1.xls : DJD Issue 1 07.04.05 <<

		Case 1					Case 2				
		PF Power Plant					PF Power Plant with CO_2 Capture				
DESCRIPTION		Reference Case					Oxy-Combustion				
		Base	+100% FP	-100% FP	5% DR	-20% TIC	Base	+100% FP	-100% FP	5% DR	-20% TIC
Run No.		1.0	1.1	1.2	1.3	1.4	2.0	2.1	2.2	2.3	2.4
Discount Rate (DR)	%	10	10	10	**5**	10	10	10	10	**5**	10
Fuel Price (FP)	US$/GJ	1.5	**3.0**	**0.0**	1.5	1.5	1.5	**3.0**	**0.0**	1.5	1.5
Fuel Mass Flow Rate	t/h	213.1	213.1	213.1	213.1	213.1	209.1	209.1	209.1	209.1	209.1
Gross Power Output	MWe Gross	740	740	740	740	740	737	737	737	737	737
Net Power Output	MWe Net	677	677	677	677	677	532	532	532	532	532
Fuel NCV	MJ/kg	25.87	25.87	25.87	25.87	25.87	25.87	25.87	25.87	25.87	25.87
Net Efficiency	% LHV	44.2	44.2	44.2	44.2	44.2	35.4	35.4	35.4	35.4	35.4
Total Investment Cost (TIC)*	M US$	1024	1024	1024	1024	819	1246	1246	1246	1246	997
Specific Investment Cost	US$/kWe Gross	1384	1384	1384	1384	1107	1691	1691	1691	1691	1353
Specific Investment Cost	US$/kWe Net	1513	1513	1513	1513	1210	2342	2342	2342	2342	1874
Revenues / year	M US$/y	251.3	313.5	189.2	200.0	216.0	288.4	349.4	227.4	225.7	244.8
Electricity Cost	**US¢/kWh**	**4.98**	**6.22**	**3.75**	**3.97**	**4.29**	**7.28**	**8.82**	**5.74**	**5.70**	**6.18**
CO_2 Emissions	t/h	489	489	489	489	489	45	45	45	45	45
CO_2 Specific Emissions (based on net output)	g/kWh	723	723	723	723	723	84	84	84	84	84
CO_2 Captured	g/kWh	-	-	-	-	-	831	831	831	831	831
Cost of CO_2 Avoidance	**US$/t CO_2**						**36**	**41**	**31**	**27**	**30**

*Note all costs based on a 1 Euro to 1.2 US$ exchange rate

ASC PF PLANT: ECONOMIC PERFORMANCE SUMMARY WITH & WITHOUT CO_2 CAPTURE

TABLE 23

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 Installed Costs NGCC DJD Iss 1.xls : DJD Issue 1 24.03.05 Updated 24.03.05 <<

Case 3 : 400 MWe Gross NGCC Power Plant : Base Case

POS	DESCRIPTION		UNIT 1000 M Euros			TOTAL M Euros	REMARKS
1	DIRECT MATERIALS		-			0	1) Estimate accuracy +/- 25%
2	CONSTRUCTION		-			0	
3	OTHER COSTS		-			0	
							UNIT
4	EPC Services		-			0	**1000 NGCC Power Plant [1]**
A	**Installed Costs (contingency excluded)**		**154**			**154**	
B	Contingency	% of A	10			10	
		Euros	15			15	
C	Fees	% of A	2			2	
		Euros	3			3	
D	Owner Costs	% of A	5			5	
		Euros	8			8	
	TOTAL INVESTMENT COST		**181**	**0**	**0**	**181**	

Definitions (according to IEA GHG Gasification Power Generation Study Report No. PH4/19, May 2003) ==>

1. Direct Materials, including equipment and bulk materials.
2. Construction, including mechanical erection, instrument and electrical installation, civil works and, where applicable, buildings and site preparation.
3. Other Costs, including temporary facilities, solvents, catalysts, chemicals, training, commissioning and start-up costs, spare parts etc.
4. EPC Services including Contractor's home office services and construction supervision, freight

Notes
1. TIC Based on 542 US$/kWe Gross (ref: IEA Technology Status Report : Solutions for 21st Century, May 2002, p25)
2. NGCC Plant based on : 1 x 9FA GT (250MWe) + 1 ST (150MWe) = 400MWe (Combined Cycle)

INSTALLED CAPITAL COSTS
CASE 3 : 400 MWe (Gross) NGCC AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE

TABLE 24

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 Installed Costs NGCC DJD Iss 1.xls : DJD Issue 1 24.03.05 Updated 24.03.05 <<

Case 4 : 575 MWe Gross NGCC Power Plant with CO2 Capture

POS	DESCRIPTION		UNIT			TOTAL M Euros	REMARKS
			1000 M Euros	2000 M Euros	3000 M Euros		
			Alstom	Air Products	Air Products		
1	DIRECT MATERIALS		172	73	13	258	1) Estimate accuracy +/- 25%
2	CONSTRUCTION		78	32	7	117	
3	OTHER COSTS		17	4	2	23	
							UNIT
4	EPC Services		52	13	5	70	**1000 NGCC Power Plant**
							2000 ASU
A	**Installed Costs (contingency excluded)**		**318**	**123**	**28**	**469**	**3000 CO_2 Purification & Compression**
B	Contingency	% of A	10	10	10	10	
		Euros	32	12	3	47	
C	Fees	% of A	2	2	2	2	
		Euros	6	2	1	9	
D	Owner Costs	% of A	5	5	5	5	
		Euros	16	6	1	23	
	TOTAL INVESTMENT COST		**373**	**144**	**32**	**548**	548

Definitions (according to IEA GHG Gasification Power Generation Study Report No. PH4/19, May 2003) ==>

1. Direct Materials, including equipment and bulk materials.
2. Construction, including mechanical erection, instrument and electrical installation, civil works and, where applicable, buildings and site preparation.
3. Other Costs, including temporary facilities, solvents, catalysts, chemicals, training, commissioning and start-up costs, spare parts etc.
4. EPC Services including Contractor's home office services and construction supervision, freight

Notes
1. Unit 1000 derived from Alstom Power email dated 25.05.04 for total plant costs, contingency of 10% separated.
2. Units 2000 & Unit 3000 costs as per Air Products email dated 21.05.04, contingency of 10% separated.

INSTALLED CAPITAL COSTS
CASE 4: 575 MWe (Gross) NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE

TABLE 25

CASE 3: EQUIPMENT LIST

1000 UNIT : NGCC Power Plant

Gas Turbine power generators
Steam Turbine power generators
Triple pressure HRSG with reheat
Condensor
Feed water pumps
Condensate pumps
Deaerator
Blowdown & Drains vessel
Controls
Instruments
Electrics
Water treatment plant
Flue Stack
Flue gas ductwork

CASE 3: CONVENTIONAL NGCC POWER PLANT WITHOUT CO_2 CAPTURE OUTLINE EQUIPMENT LIST

TABLE 26

CASE 4: EQUIPMENT LIST

1000 UNIT : NGCC Oxy-Combustion

Gas Zep Gas Turbine power generator
Direct contact cooler
Dampers
Steam turbine power generators
Triple pressure HRSG with reheat
Steam Condensor
feed water pumps
Condensate pumps
Deaerator
Blowdown & Drains vessel
Controls
Instruments
Electrics
Water treatment plant
Flue gas ductwork

5000 UNIT : CO_2 Compression

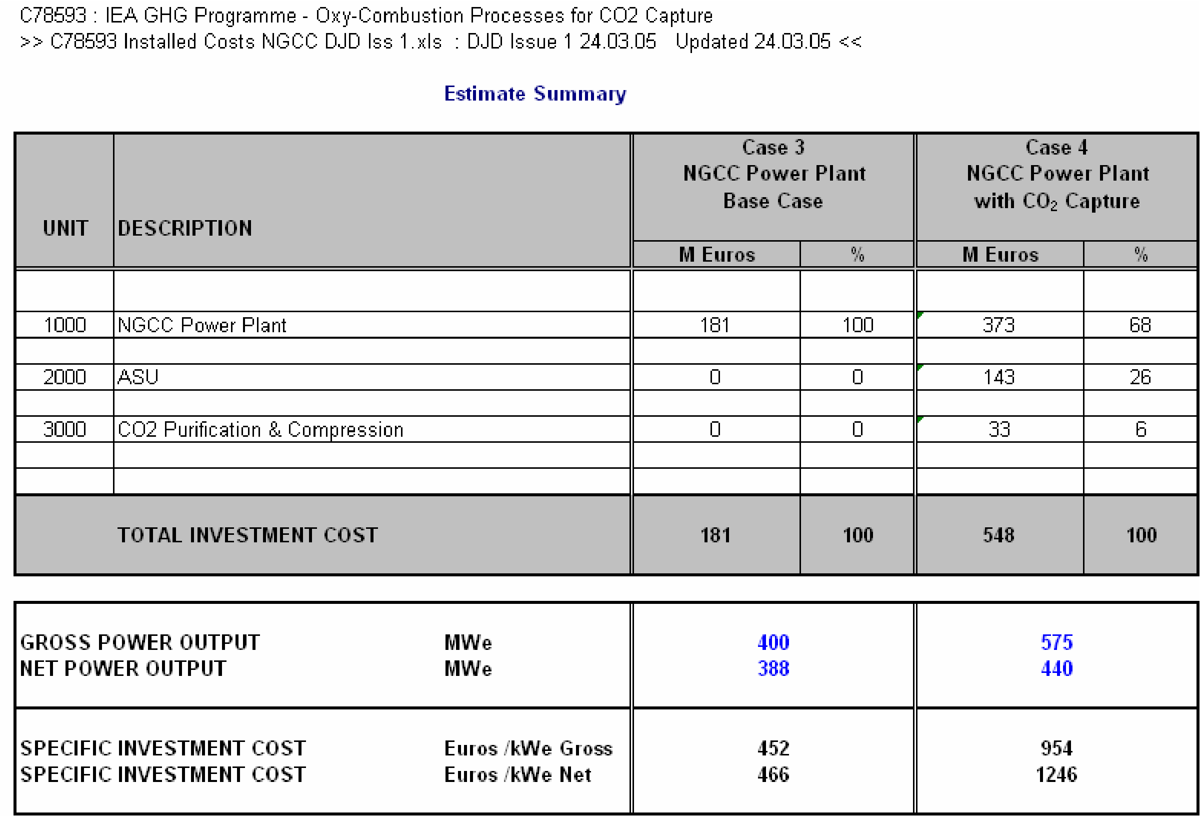
CO_2 adiabatic compressors
Heat exhangers
Inerts removal package
Pumps

2000 UNIT : ASU

Main air compressors
Air purification system
Main heat exchanger
ASU compander
ASU Column System
Pumps
Backup storage vessel

CASE 4: NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE OUTLINE EQUIPMENT LIST

TABLE 27



C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 Installed Costs NGCC DJD Iss 1.xls : DJD Issue 1 24.03.05 Updated 24.03.05 <<

Estimate Summary

UNIT	DESCRIPTION	Case 3 NGCC Power Plant Base Case		Case 4 NGCC Power Plant with CO_2 Capture	
		M Euros	%	M Euros	%
1000	NGCC Power Plant	181	100	373	68
2000	ASU	0	0	143	26
3000	CO2 Purification & Compression	0	0	33	6
	TOTAL INVESTMENT COST	**181**	**100**	**548**	**100**

		Case 3	Case 4
GROSS POWER OUTPUT	MWe	400	575
NET POWER OUTPUT	MWe	388	440
SPECIFIC INVESTMENT COST	Euros /kWe Gross	452	954
SPECIFIC INVESTMENT COST	Euros /kWe Net	466	1246

CASE 3 & 4 : NGCC POWER PLANT
PERCENTAGE CONTRIBUTION OF EACH UNIT TO TOTAL INVESTMENT COST

TABLE 28

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 O&M Costs NGCC DJD Iss 1.xls : DJD Issue 1 07.04.05 Updated Issue 1 07.04.05 <<

Yearly Variable Costs : Case 3 and Case 4

Yearly Operating hours =	7446	NGCC Plant Case			NGCC Plant Case		
		400MWe Gross NGCC Power Plant Case 3			575MWe Gross NGCC Power Plant Case 4		
Consumables	Unit Cost €/t	Consumption Hourly t/h	Consumption Yearly t/y	Operating Costs (yearly basis) €	Consumption Hourly t/h	Consumption Yearly t/y	Operating Costs (yearly basis) €
Feedstock							
Natural Gas (€/MWh, MWh/h, GWh/a)	145.3	51.5	383	55,718,046	73.2	545	79,195,358
Auxiliary Feedstock							
-	0.0	0	0	0	0	0	0
Solvents							
-	0.0	0	0	0	0	0	0
Catalyst							
no SCR	0.0	0	0	0	0	0	0
Chemicals							
neglectible	0.0	0	0	0	0	0	0
Waste Disposal							
neglectible (e.g. waste water)	0.0	0	0	0	0	0	0
Miscellaneous							
neglectible (e.g. propane, service water)	0.0	0	0	0	0	0	0
TOTAL YEARLY OPERATING COSTS			**€**	**55,718,046**		**€**	**79,195,358**

CASE 3 & 4 : NGCC POWER PLANT – CONSUMABLES

TABLE 29

C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> C78593 O&M Costs NGCC DJD Iss 1.xls : DJD Issue 1 07.04.05 Updated Issue 1 07.04.05 <<

Total O&M Costs : Case 3 and Case 4

POS	DESCRIPTION	Case 3 400MWe NGCC Plant €/year	Case 4 575MWe NGCC Plant €/year
A1	Direct Labour	1,600,000	3,000,000
A2	Administration/General Overheads (30% of A1)	480,000	900,000
A3	Maintenance (4% of **Installed costs** (contingency excluded))	6,160,000	18,760,000
A	**FIXED COSTS**	**8,240,000**	**22,660,000**
B	**VARIABLE COSTS (Fuel)**	**55,718,000**	**79,195,000**
	TOTAL OPERATION AND MAINTENANCE COSTS	**63,958,000**	**101,855,000**

CASE 3 & 4: NGCC POWER PLANT
TOTAL OPERATING AND MAINTENANCE COSTS

TABLE 30

IEA GREENHOUSE GAS R&D PROGRAMME

Case 3 : Cost Evaluation : 400MWe Gross NGCC Power Plant (Reference) : Base Case : 10% Discount Rate

Version : 1
Date : February 2003

Production		
Fuel feedrate	53.2	t/h
Fuel NCV	46.90	MJ/kg
Fuel feedrate (based on NCV)	692.9	MWt
Net power output	388	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	370	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	370	g/kWh
Electricity cost*	3.35	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		184.8
Average contingencies	10.0%	18.5
Fees	2.0%	3.7
Owners costs	5.0%	9.2
Total investment cost		216.2
Working Capital		
Chemicals storage		0.0
Fuel storage		0.0
Total working capital		0.0
Decommissioning cost		0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	55.7
Maintenance	7.4
Chemicals + consumables	0.1
Insurance and local taxes	3.7
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	1.6
Adminstration	0.5
Total labour	2.0

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	31	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	3.343	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2
Breakdown of c/kWh cost		
Fuel	57.68%	
Capital	28.13%	
Other costs	14.18%	

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	51.1	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-29.5	-55.72	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7
Maintenance	0.0	0.0	0.0	-5.5	-7.39	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4
Labour	0.0	0.0	0.0	-2.0	-2.02	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-3.7	-3.70	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7
Fixed Capital Expenditures	-43.2	-97.3	-75.7											
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-43.2	-97.3	-75.7	10.3	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6
Total Cash Flow (cumulated)	-43.2	-140.5	-216.2	-205.9	-178.3	-150.7	-123.0	-95.4	-67.8	-40.2	-12.5	15.1	42.7	70.4

	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	96.6	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	
Maintenance	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	
Labour	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															0.0
Total Cash Flow (yearly)	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	0.0
Total Cash Flow (cumulated)	98.0	125.6	153.2	180.9	208.5	236.1	263.8	291.4	319.0	346.6	374.3	401.9	429.5	457.2	457.2

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 3: NGCC AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE

TABLE 31

IEA GREENHOUSE GAS R&D PROGRAMME

Case 3 : Cost Evaluation : 400MWe Gross NGCC Power Plant (Reference) : 10% Discount Rate & +100% Fuel Price

Version : 1
Date : February 2003

Production		
Fuel feedrate	53.2	t/h
Fuel NCV	46.90	MJ/kg
Fuel feedrate (based on NCV)	692.9	MWt
Net power output	388	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	370	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	370	g/kWh
Electricity cost*	3.35	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		184.8
Average contingencies	10.0%	18.5
Fees	2.0%	3.7
Owners costs	5.0%	9.2
Total investment cost		216.2
Working Capital		
Chemicals storage		0.0
Fuel storage		0.0
Total working capital		0.0
Decommissioning cost		0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	111.4
Maintenance	7.4
Chemicals + consumables	0.1
Insurance and local taxes	3.7
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	1.6
Adminstration	0.5
Total labour	2.0

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	6.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	31	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	5.272	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2
Breakdown of c/kWh cost		
Fuel	73.16%	
Capital	17.84%	
Other costs	9.00%	

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	80.6	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-59.0	-111.43	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4
Maintenance	0.0	0.0	0.0	-5.5	-7.39	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4
Labour	0.0	0.0	0.0	-2.0	-2.02	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-3.7	-3.70	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7
Fixed Capital Expenditures	-43.2	-97.3	-75.7											
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-43.2	-97.3	-75.7	10.3	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6
Total Cash Flow (cumulated)	-43.2	-140.5	-216.2	-205.9	-178.3	-150.7	-123.0	-95.4	-67.8	-40.2	-12.5	15.1	42.7	70.4

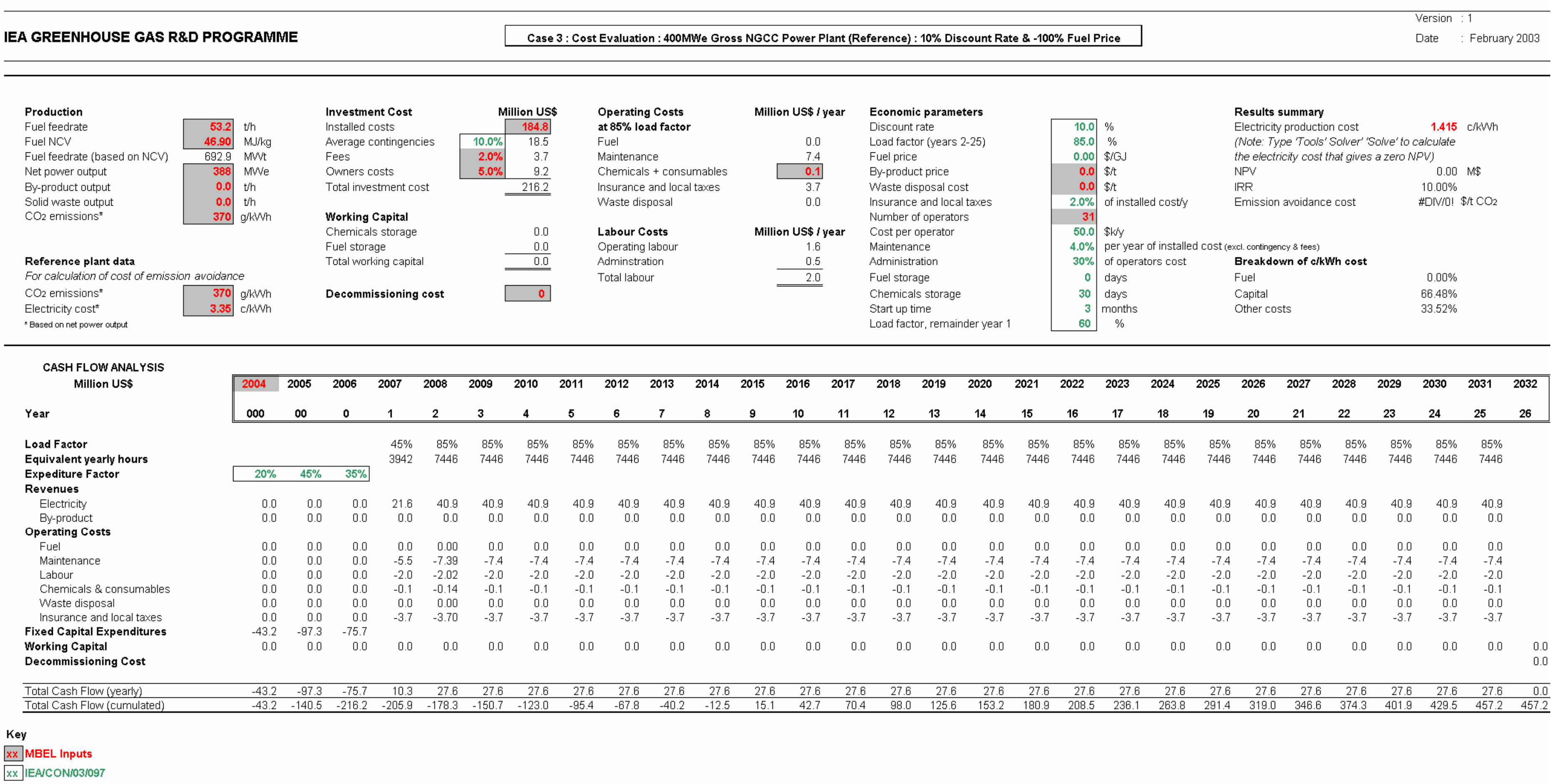
	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	152.3	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	-111.4	
Maintenance	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	
Labour	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															0.0
Total Cash Flow (yearly)	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	0.0
Total Cash Flow (cumulated)	98.0	125.6	153.2	180.9	208.5	236.1	263.8	291.4	319.0	346.6	374.3	401.9	429.5	457.2	457.2

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 3: NGCC AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE & +100% FUEL PRICE

TABLE 32



Version : 1
Date : February 2003

IEA GREENHOUSE GAS R&D PROGRAMME

Case 3 : Cost Evaluation : 400MWe Gross NGCC Power Plant (Reference) : 10% Discount Rate & -100% Fuel Price

Production		
Fuel feedrate	53.2	t/h
Fuel NCV	46.90	MJ/kg
Fuel feedrate (based on NCV)	692.9	MWt
Net power output	388	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	370	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	370	g/kWh
Electricity cost*	3.35	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		184.8
Average contingencies	10.0%	18.5
Fees	2.0%	3.7
Owners costs	5.0%	9.2
Total investment cost		216.2
Working Capital		
Chemicals storage		0.0
Fuel storage		0.0
Total working capital		0.0
Decommissioning cost		0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	0.0
Maintenance	7.4
Chemicals + consumables	0.1
Insurance and local taxes	3.7
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	1.6
Adminstration	0.5
Total labour	2.0

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	0.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	31	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	1.415	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2
Breakdown of c/kWh cost		
Fuel	0.00%	
Capital	66.48%	
Other costs	33.52%	

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	21.6	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Maintenance	0.0	0.0	0.0	-5.5	-7.39	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4
Labour	0.0	0.0	0.0	-2.0	-2.02	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-3.7	-3.70	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7
Fixed Capital Expenditures	-43.2	-97.3	-75.7											
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-43.2	-97.3	-75.7	10.3	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6
Total Cash Flow (cumulated)	-43.2	-140.5	-216.2	-205.9	-178.3	-150.7	-123.0	-95.4	-67.8	-40.2	-12.5	15.1	42.7	70.4

	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	40.9	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Maintenance	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	
Labour	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															0.0
Total Cash Flow (yearly)	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	27.6	0.0
Total Cash Flow (cumulated)	98.0	125.6	153.2	180.9	208.5	236.1	263.8	291.4	319.0	346.6	374.3	401.9	429.5	457.2	457.2

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 3: NGCC AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE & -100% FUEL PRICE

TABLE 33

IEA GREENHOUSE GAS R&D PROGRAMME

Case 3 : Cost Evaluation : 400MWe Gross NGCC Power Plant (Reference) : 5% Discount Rate

Version : 1
Date : February 2003

Production		
Fuel feedrate	53.2	t/h
Fuel NCV	46.90	MJ/kg
Fuel feedrate (based on NCV)	692.9	MWt
Net power output	388	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	370	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	370	g/kWh
Electricity cost*	3.35	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		184.8
Average contingencies	10.0%	18.5
Fees	2.0%	3.7
Owners costs	5.0%	9.2
Total investment cost		216.2

Working Capital	
Chemicals storage	0.0
Fuel storage	0.0
Total working capital	0.0
Decommissioning cost	0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	55.7
Maintenance	7.4
Chemicals + consumables	0.1
Insurance and local taxes	3.7
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	1.6
Adminstration	0.5
Total labour	2.0

Economic parameters		
Discount rate	5.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	31	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	2.970	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	5.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2

Breakdown of c/kWh cost	
Fuel	64.94%
Capital	19.26%
Other costs	15.79%

CASH FLOW ANALYSIS

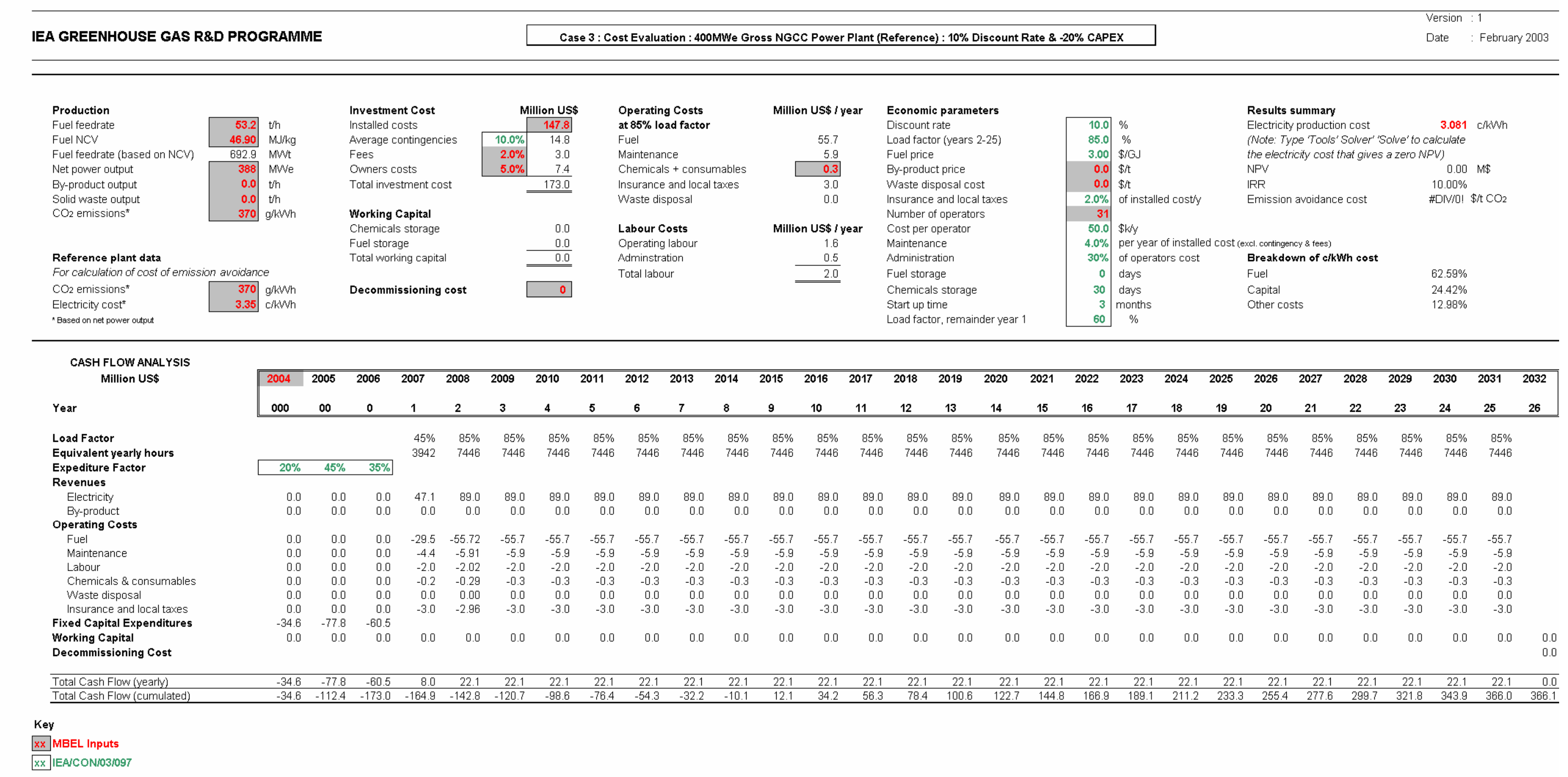
Million US$	**2004**	**2005**	**2006**	**2007**	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor	20%	45%	35%																										
Revenues																													
Electricity	0.0	0.0	0.0	45.4	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	85.8	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs																													
Fuel	0.0	0.0	0.0	-29.5	-55.72	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	
Maintenance	0.0	0.0	0.0	-5.5	-7.39	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	-7.4	
Labour	0.0	0.0	0.0	-2.0	-2.02	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	0.0	0.0	0.0	-3.7	-3.70	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	-3.7	
Fixed Capital Expenditures	-43.2	-97.3	-75.7																										
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost																													0.0
Total Cash Flow (yearly)	-43.2	-97.3	-75.7	4.6	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	0.0
Total Cash Flow (cumulated)	-43.2	-140.5	-216.2	-211.6	-194.8	-178.0	-161.2	-144.3	-127.5	-110.7	-93.8	-77.0	-60.2	-43.4	-26.5	-9.7	7.1	24.0	40.8	57.6	74.4	91.3	108.1	124.9	141.8	158.6	175.4	192.2	192.3

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 3: NGCC AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 5% DISCOUNT RATE

TABLE 34



Version : 1

IEA GREENHOUSE GAS R&D PROGRAMME

Case 3 : Cost Evaluation : 400MWe Gross NGCC Power Plant (Reference) : 10% Discount Rate & -20% CAPEX

Date : February 2003

Production		
Fuel feedrate	53.2	t/h
Fuel NCV	46.90	MJ/kg
Fuel feedrate (based on NCV)	692.9	MWt
Net power output	388	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	370	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	370	g/kWh
Electricity cost*	3.35	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		147.8
Average contingencies	10.0%	14.8
Fees	2.0%	3.0
Owners costs	5.0%	7.4
Total investment cost		173.0
Working Capital		
Chemicals storage		0.0
Fuel storage		0.0
Total working capital		0.0
Decommissioning cost		0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	55.7
Maintenance	5.9
Chemicals + consumables	0.3
Insurance and local taxes	3.0
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	1.6
Adminstration	0.5
Total labour	2.0

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	31	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	3.081	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	#DIV/0!	$/t CO_2
Breakdown of c/kWh cost		
Fuel	62.59%	
Capital	24.42%	
Other costs	12.98%	

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016
Year	000	00	0	1	2	3	4	5	6	7	8	9	10
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%										
Revenues													
Electricity	0.0	0.0	0.0	47.1	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs													
Fuel	0.0	0.0	0.0	-29.5	-55.72	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7
Maintenance	0.0	0.0	0.0	-4.4	-5.91	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9
Labour	0.0	0.0	0.0	-2.0	-2.02	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0
Chemicals & consumables	0.0	0.0	0.0	-0.2	-0.29	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-3.0	-2.96	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0
Fixed Capital Expenditures	-34.6	-77.8	-60.5										
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost													
Total Cash Flow (yearly)	-34.6	-77.8	-60.5	8.0	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1
Total Cash Flow (cumulated)	-34.6	-112.4	-173.0	-164.9	-142.8	-120.7	-98.6	-76.4	-54.3	-32.2	-10.1	12.1	34.2

	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor																
Revenues																
Electricity	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	89.0	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs																
Fuel	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	-55.7	
Maintenance	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	-5.9	
Labour	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	-2.0	
Chemicals & consumables	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	-0.3	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	-3.0	
Fixed Capital Expenditures																
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost																0.0
Total Cash Flow (yearly)	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	0.0
Total Cash Flow (cumulated)	56.3	78.4	100.6	122.7	144.8	166.9	189.1	211.2	233.3	255.4	277.6	299.7	321.8	343.9	366.0	366.1

Key

xx	MBEL Inputs
xx	IEA/CON/03/097

CASE 3: NGCC AIR FIRED POWER PLANT WITHOUT CO_2 CAPTURE : 10% DISCOUNT RATE & -20% TIC

TABLE 35

Version : 1
Date : February 2003

IEA GREENHOUSE GAS R&D PROGRAMME

Case 4 : Cost Evaluation : 575MWe Gross NGCC Power Plant with CO_2 Capture : Base Case : 10% Discount Rate

Production		
Fuel feedrate	73.2	t/h
Fuel NCV	48.45	MJ/kg
Fuel feedrate (based on NCV)	984.5	MWt
Net power output	440	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	12	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	370	g/kWh
Electricity cost*	3.35	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		562.8
Average contingencies	10.0%	56.3
Fees	2.0%	11.3
Owners costs	5.0%	28.1
Total investment cost		658.5
Working Capital		
Chemicals storage		0.0
Fuel storage		0.0
Total working capital		0.0
Decommissioning cost		0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	79.2
Maintenance	22.5
Chemicals + consumables	0.1
Insurance and local taxes	11.3
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	3.0
Adminstration	0.9
Total labour	3.9

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	60	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	6.136	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	77.8	$/t CO_2
Breakdown of c/kWh cost		
Fuel	39.39%	
Capital	41.17%	
Other costs	19.44%	

CASH FLOW ANALYSIS
Million US$

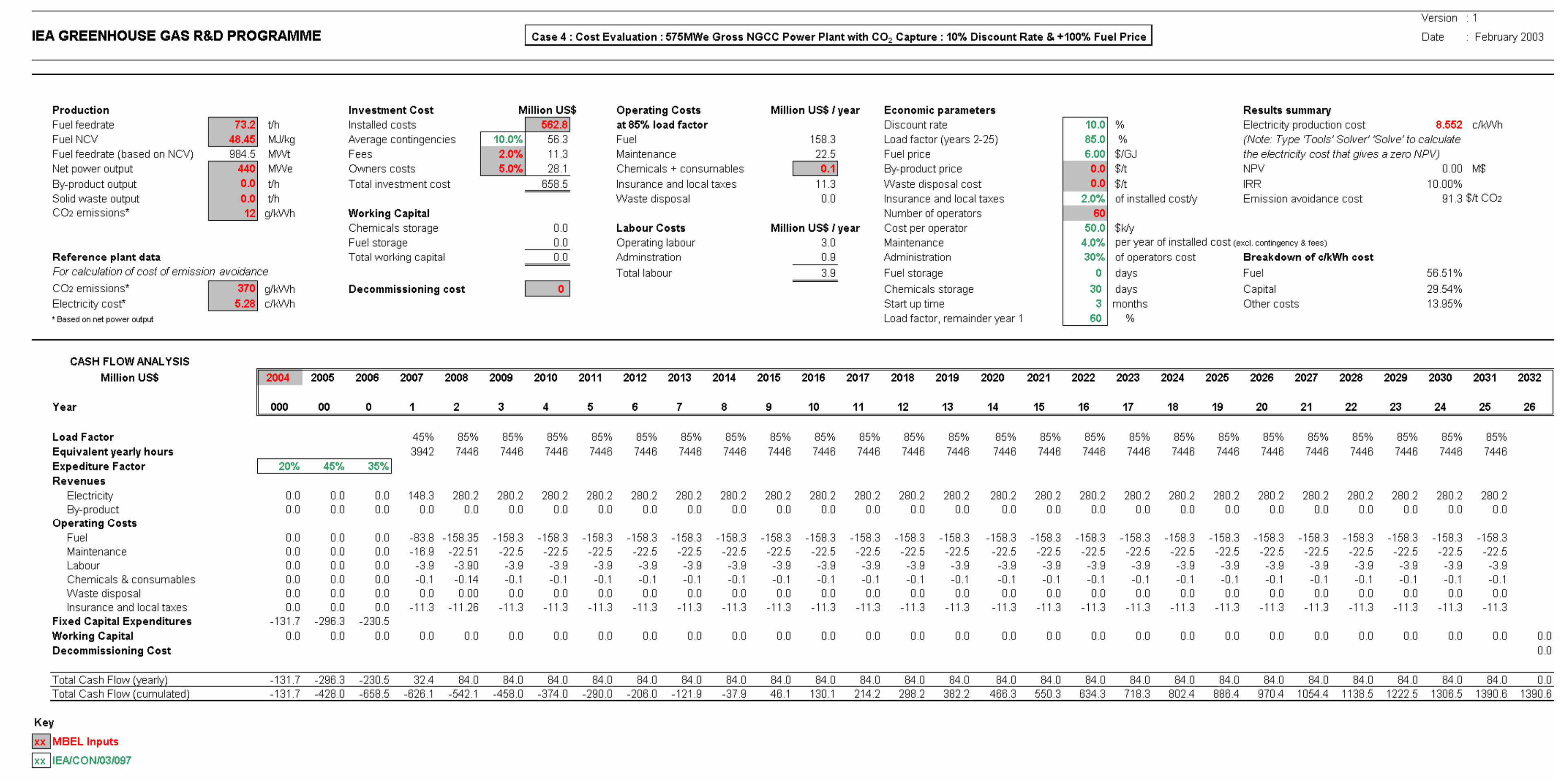
	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	106.4	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-41.9	-79.17	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2
Maintenance	0.0	0.0	0.0	-16.9	-22.51	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5
Labour	0.0	0.0	0.0	-3.9	-3.90	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-11.3	-11.26	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3
Fixed Capital Expenditures	-131.7	-296.3	-230.5											
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-131.7	-296.3	-230.5	32.4	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0
Total Cash Flow (cumulated)	-131.7	-428.0	-658.5	-626.1	-542.1	-458.0	-374.0	-290.0	-206.0	-121.9	-37.9	46.1	130.1	214.2

	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	201.0	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	
Maintenance	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	
Labour	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															0.0
Total Cash Flow (yearly)	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	0.0
Total Cash Flow (cumulated)	298.2	382.2	466.3	550.3	634.3	718.3	802.4	886.4	970.4	1054.4	1138.5	1222.5	1306.5	1390.6	1390.6

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 4: NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE :10% DISCOUNT RATE **TABLE 36**



IEA GREENHOUSE GAS R&D PROGRAMME

Case 4 : Cost Evaluation : 575MWe Gross NGCC Power Plant with CO_2 Capture : 10% Discount Rate & +100% Fuel Price

Version : 1
Date : February 2003

Production		
Fuel feedrate	73.2	t/h
Fuel NCV	48.45	MJ/kg
Fuel feedrate (based on NCV)	984.5	MWt
Net power output	440	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO2 emissions*	12	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO2 emissions*	370	g/kWh
Electricity cost*	5.28	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		562.8
Average contingencies	10.0%	56.3
Fees	2.0%	11.3
Owners costs	5.0%	28.1
Total investment cost		658.5
Working Capital		
Chemicals storage		0.0
Fuel storage		0.0
Total working capital		0.0
Decommissioning cost		0

Operating Costs at 85% load factor	**Million US$ / year**
Fuel	158.3
Maintenance	22.5
Chemicals + consumables	0.1
Insurance and local taxes	11.3
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	3.0
Adminstration	0.9
Total labour	3.9

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	6.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	60	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	8.552	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	91.3	$/t CO2

Breakdown of c/kWh cost	
Fuel	56.51%
Capital	29.54%
Other costs	13.95%

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electricity	0.0	0.0	0.0	148.3	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-83.8	-158.35	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3
Maintenance	0.0	0.0	0.0	-16.9	-22.51	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5
Labour	0.0	0.0	0.0	-3.9	-3.90	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-11.3	-11.26	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3
Fixed Capital Expenditures	-131.7	-296.3	-230.5												
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-131.7	-296.3	-230.5	32.4	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0
Total Cash Flow (cumulated)	-131.7	-428.0	-658.5	-626.1	-542.1	-458.0	-374.0	-290.0	-206.0	-121.9	-37.9	46.1	130.1	214.2	298.2

	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electricity	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	280.2	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	-158.3	
Maintenance	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	
Labour	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	
Fixed Capital Expenditures														
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														0.0
Total Cash Flow (yearly)	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	0.0
Total Cash Flow (cumulated)	382.2	466.3	550.3	634.3	718.3	802.4	886.4	970.4	1054.4	1138.5	1222.5	1306.5	1390.6	1390.6

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 4: NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : 10% DISCOUNT RATE & +100% FUEL PRICE

TABLE 37

Version : 1
Date : February 2003

IEA GREENHOUSE GAS R&D PROGRAMME

Case 4 : Cost Evaluation : 575MWe Gross NGCC Power Plant with CO_2 Capture : 10% Discount Rate & -100% Fuel Price

Production		
Fuel feedrate	73.2	t/h
Fuel NCV	48.45	MJ/kg
Fuel feedrate (based on NCV)	984.5	MWt
Net power output	440	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO2 emissions*	12	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO2 emissions*	370	g/kWh
Electricity cost*	1.42	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		562.8
Average contingencies	10.0%	56.3
Fees	2.0%	11.3
Owners costs	5.0%	28.1
Total investment cost		658.5

Working Capital	
Chemicals storage	0.0
Fuel storage	0.0
Total working capital	0.0

Decommissioning cost 0

Operating Costs	**Million US$ / year**
at 85% load factor	
Fuel	0.0
Maintenance	22.5
Chemicals + consumables	0.1
Insurance and local taxes	11.3
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	3.0
Adminstration	0.9
Total labour	3.9

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	0.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	60	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	3.719	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	64.2	$/t CO2

Breakdown of c/kWh cost	
Fuel	0.00%
Capital	67.92%
Other costs	32.08%

CASH FLOW ANALYSIS
Million US$

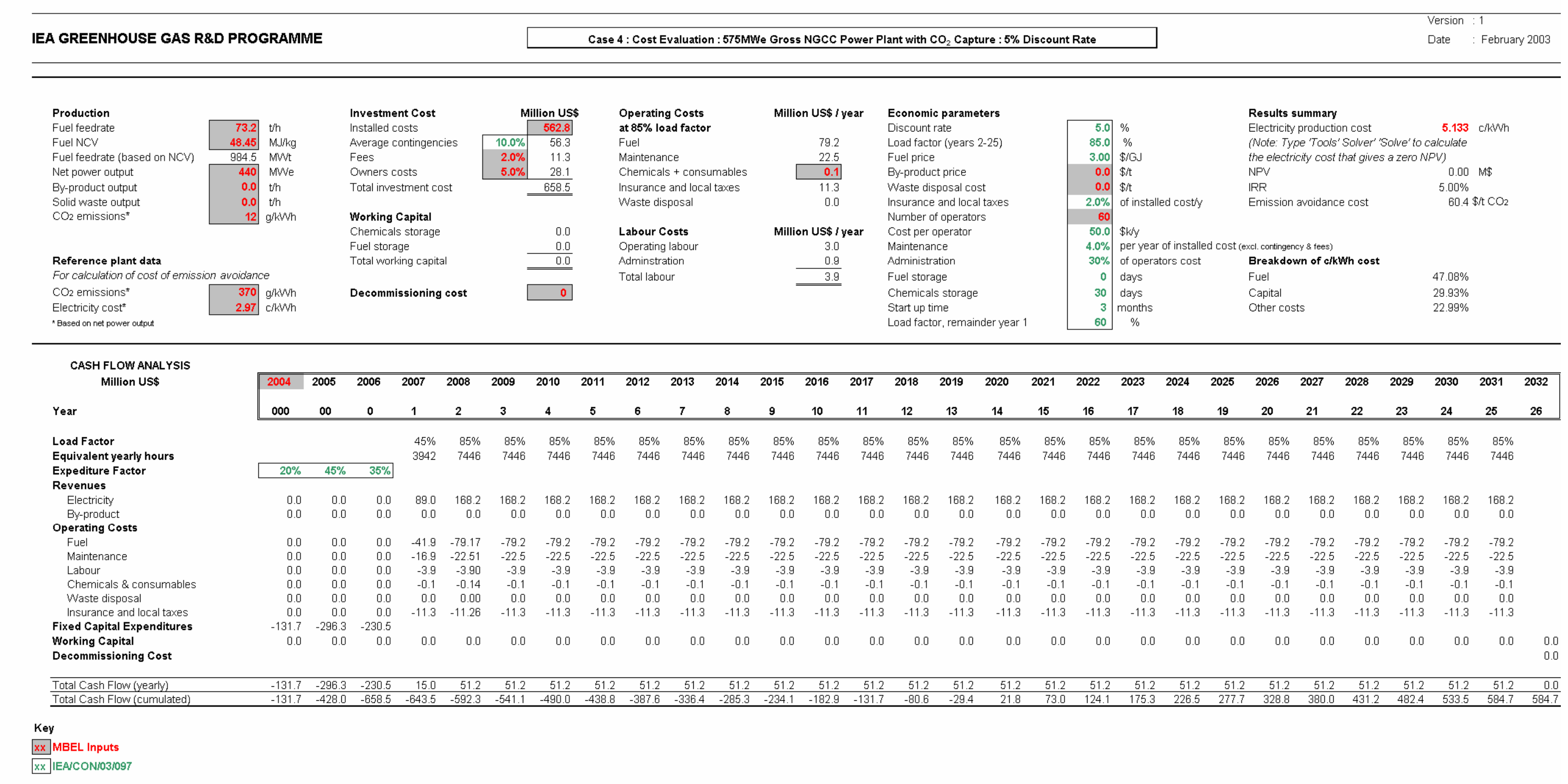
	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	64.5	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Maintenance	0.0	0.0	0.0	-16.9	-22.51	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5
Labour	0.0	0.0	0.0	-3.9	-3.90	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-11.3	-11.26	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3
Fixed Capital Expenditures	-131.7	-296.3	-230.5											
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-131.7	-296.3	-230.5	32.4	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0
Total Cash Flow (cumulated)	-131.7	-428.0	-658.5	-626.1	-542.1	-458.0	-374.0	-290.0	-206.0	-121.9	-37.9	46.1	130.1	214.2

	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	121.8	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Maintenance	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	
Labour	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															0.0
Total Cash Flow (yearly)	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	84.0	0.0
Total Cash Flow (cumulated)	298.2	382.2	466.3	550.3	634.3	718.3	802.4	886.4	970.4	1054.4	1138.5	1222.5	1306.5	1390.6	1390.6

Key

xx	**MBEL Inputs**
xx	**IEA/CON/03/097**

CASE 4: NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE:10% DISCOUNT RATE & -100% FUEL PRICE TABLE 38



IEA GREENHOUSE GAS R&D PROGRAMME

Case 4 : Cost Evaluation : 575MWe Gross NGCC Power Plant with CO_2 Capture : 5% Discount Rate

Version : 1
Date : February 2003

Production		
Fuel feedrate	73.2	t/h
Fuel NCV	48.45	MJ/kg
Fuel feedrate (based on NCV)	984.5	MWt
Net power output	440	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO2 emissions*	12	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO2 emissions*	370	g/kWh
Electricity cost*	2.97	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		562.8
Average contingencies	10.0%	56.3
Fees	2.0%	11.3
Owners costs	5.0%	28.1
Total investment cost		658.5

Working Capital	
Chemicals storage	0.0
Fuel storage	0.0
Total working capital	0.0
Decommissioning cost	0

Operating Costs **at 85% load factor**	**Million US$ / year**
Fuel	79.2
Maintenance	22.5
Chemicals + consumables	0.1
Insurance and local taxes	11.3
Waste disposal	0.0

Labour Costs	**Million US$ / year**
Operating labour	3.0
Adminstration	0.9
Total labour	3.9

Economic parameters		
Discount rate	5.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	60	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	5.133	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	5.00%	
Emission avoidance cost	60.4	$/t CO2

Breakdown of c/kWh cost	
Fuel	47.08%
Capital	29.93%
Other costs	22.99%

CASH FLOW ANALYSIS
Million US$

	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%											
Revenues														
Electricity	0.0	0.0	0.0	89.0	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-41.9	-79.17	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2
Maintenance	0.0	0.0	0.0	-16.9	-22.51	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5
Labour	0.0	0.0	0.0	-3.9	-3.90	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-11.3	-11.26	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3
Fixed Capital Expenditures	-131.7	-296.3	-230.5											
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-131.7	-296.3	-230.5	15.0	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2
Total Cash Flow (cumulated)	-131.7	-428.0	-658.5	-643.5	-592.3	-541.1	-490.0	-438.8	-387.6	-336.4	-285.3	-234.1	-182.9	-131.7

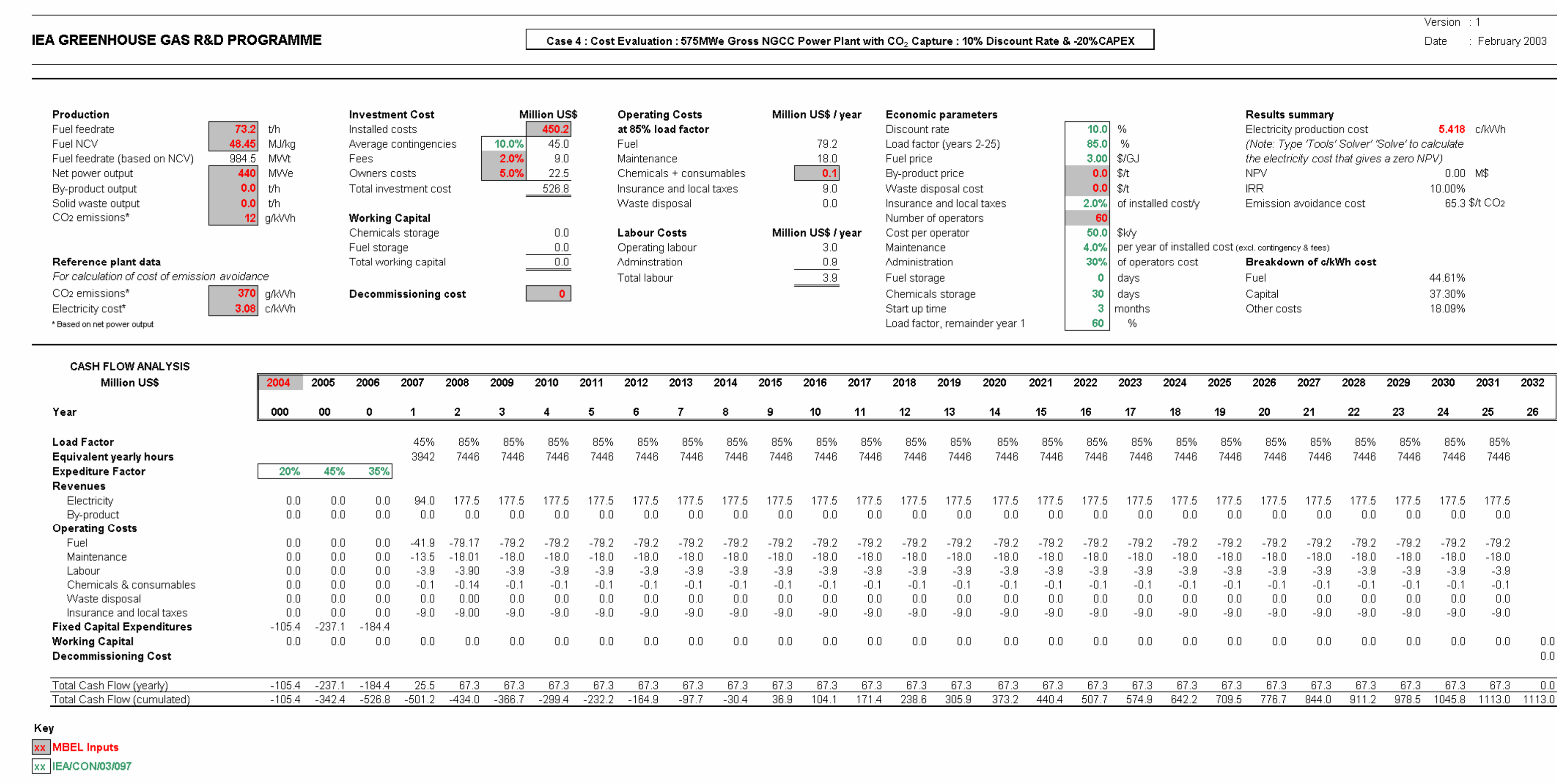
	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor															
Revenues															
Electricity	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	168.2	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	
Maintenance	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	-22.5	
Labour	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	-11.3	
Fixed Capital Expenditures															
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															0.0
Total Cash Flow (yearly)	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	51.2	0.0
Total Cash Flow (cumulated)	-80.6	-29.4	21.8	73.0	124.1	175.3	226.5	277.7	328.8	380.0	431.2	482.4	533.5	584.7	584.7

Key

xx **MBEL Inputs**
xx **IEA/CON/03/097**

CASE 4: NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : 5% DISCOUNT RATE

TABLE 39



IEA GREENHOUSE GAS R&D PROGRAMME

Case 4 : Cost Evaluation : 575MWe Gross NGCC Power Plant with CO_2 Capture : 10% Discount Rate & -20%CAPEX

Version : 1
Date : February 2003

Production		
Fuel feedrate	73.2	t/h
Fuel NCV	48.45	MJ/kg
Fuel feedrate (based on NCV)	984.5	MWt
Net power output	440	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	12	g/kWh

Reference plant data		
For calculation of cost of emission avoidance		
CO_2 emissions*	370	g/kWh
Electricity cost*	3.08	c/kWh

* Based on net power output

Investment Cost		**Million US$**
Installed costs		450.2
Average contingencies	10.0%	45.0
Fees	2.0%	9.0
Owners costs	5.0%	22.5
Total investment cost		526.8
Working Capital		
Chemicals storage		0.0
Fuel storage		0.0
Total working capital		0.0
Decommissioning cost		0

Operating Costs	**Million US$ / year**
at 85% load factor	
Fuel	79.2
Maintenance	18.0
Chemicals + consumables	0.1
Insurance and local taxes	9.0
Waste disposal	0.0
Labour Costs	**Million US$ / year**
Operating labour	3.0
Adminstration	0.9
Total labour	3.9

Economic parameters		
Discount rate	10.0	%
Load factor (years 2-25)	85.0	%
Fuel price	3.00	$/GJ
By-product price	0.0	$/t
Waste disposal cost	0.0	$/t
Insurance and local taxes	2.0%	of installed cost/y
Number of operators	60	
Cost per operator	50.0	$k/y
Maintenance	4.0%	per year of installed cost (excl. contingency & fees)
Administration	30%	of operators cost
Fuel storage	0	days
Chemicals storage	30	days
Start up time	3	months
Load factor, remainder year 1	60	%

Results summary		
Electricity production cost	5.418	c/kWh
(Note: Type 'Tools' Solver' 'Solve' to calculate the electricity cost that gives a zero NPV)		
NPV	0.00	M$
IRR	10.00%	
Emission avoidance cost	65.3	$/t CO_2
Breakdown of c/kWh cost		
Fuel	44.61%	
Capital	37.30%	
Other costs	18.09%	

CASH FLOW ANALYSIS
Million US$

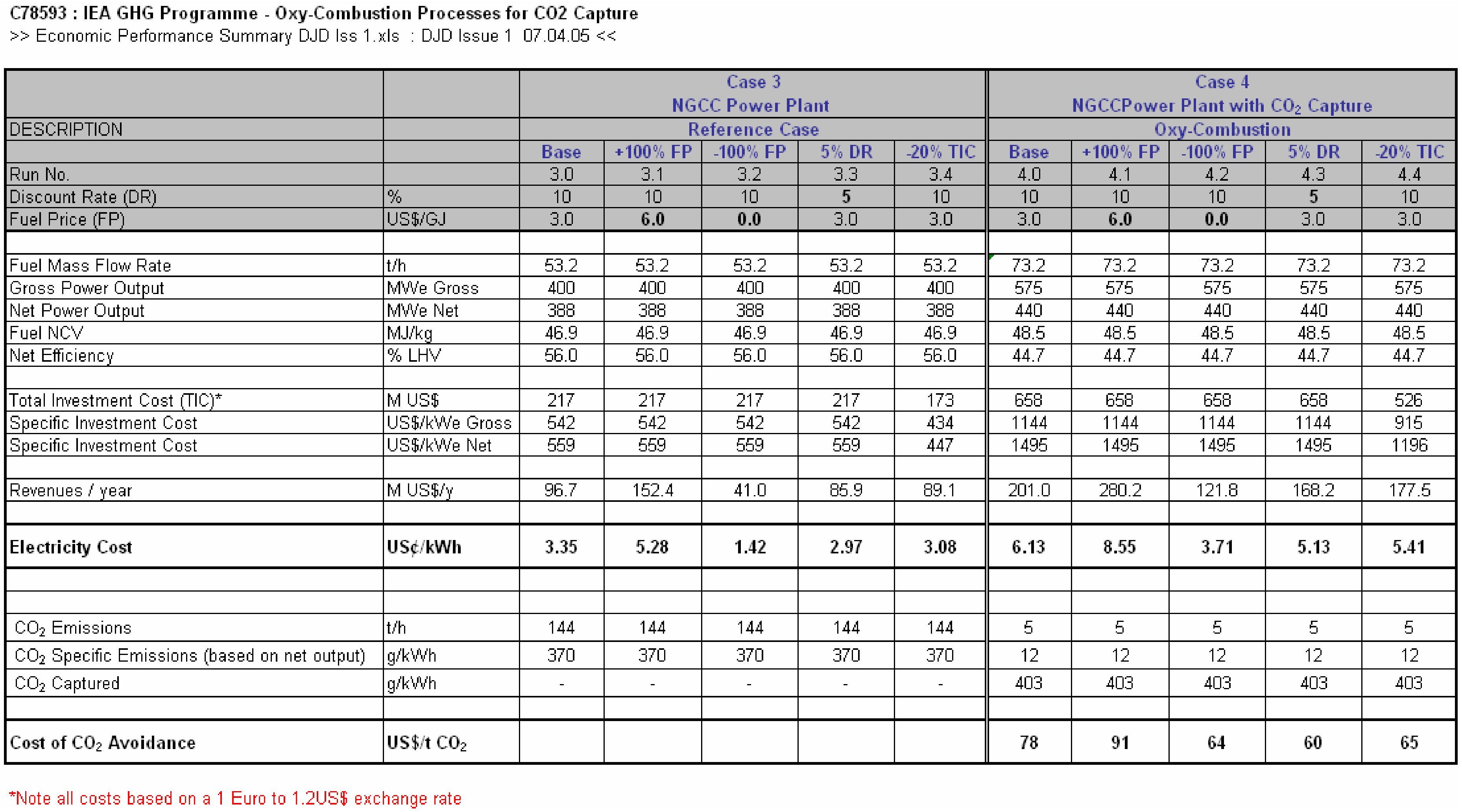
	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				45%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%
Equivalent yearly hours				3942	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446
Expediture Factor	20%	45%	35%												
Revenues															
Electricity	0.0	0.0	0.0	94.0	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-41.9	-79.17	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2
Maintenance	0.0	0.0	0.0	-13.5	-18.01	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0
Labour	0.0	0.0	0.0	-3.9	-3.90	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9
Chemicals & consumables	0.0	0.0	0.0	-0.1	-0.14	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1
Waste disposal	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-9.0	-9.00	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0
Fixed Capital Expenditures	-105.4	-237.1	-184.4												
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-105.4	-237.1	-184.4	25.5	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3
Total Cash Flow (cumulated)	-105.4	-342.4	-526.8	-501.2	-434.0	-366.7	-299.4	-232.2	-164.9	-97.7	-30.4	36.9	104.1	171.4	238.6

	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	85%	
Equivalent yearly hours	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	7446	
Expediture Factor														
Revenues														
Electricity	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	177.5	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	
Maintenance	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	-18.0	
Labour	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	-3.9	
Chemicals & consumables	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	-0.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	-9.0	
Fixed Capital Expenditures														
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														0.0
Total Cash Flow (yearly)	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	67.3	0.0
Total Cash Flow (cumulated)	305.9	373.2	440.4	507.7	574.9	642.2	709.5	776.7	844.0	911.2	978.5	1045.8	1113.0	1113.0

Key
xx **MBEL Inputs**
xx **IEA/CON/03/097**

CASE 4: NGCC OXY-COMBUSTION POWER PLANT WITH CO_2 CAPTURE : 10% DISCOUNT RATE & -20% TIC

TABLE 40



C78593 : IEA GHG Programme - Oxy-Combustion Processes for CO2 Capture
>> Economic Performance Summary DJD Iss 1.xls : DJD Issue 1 07.04.05 <<

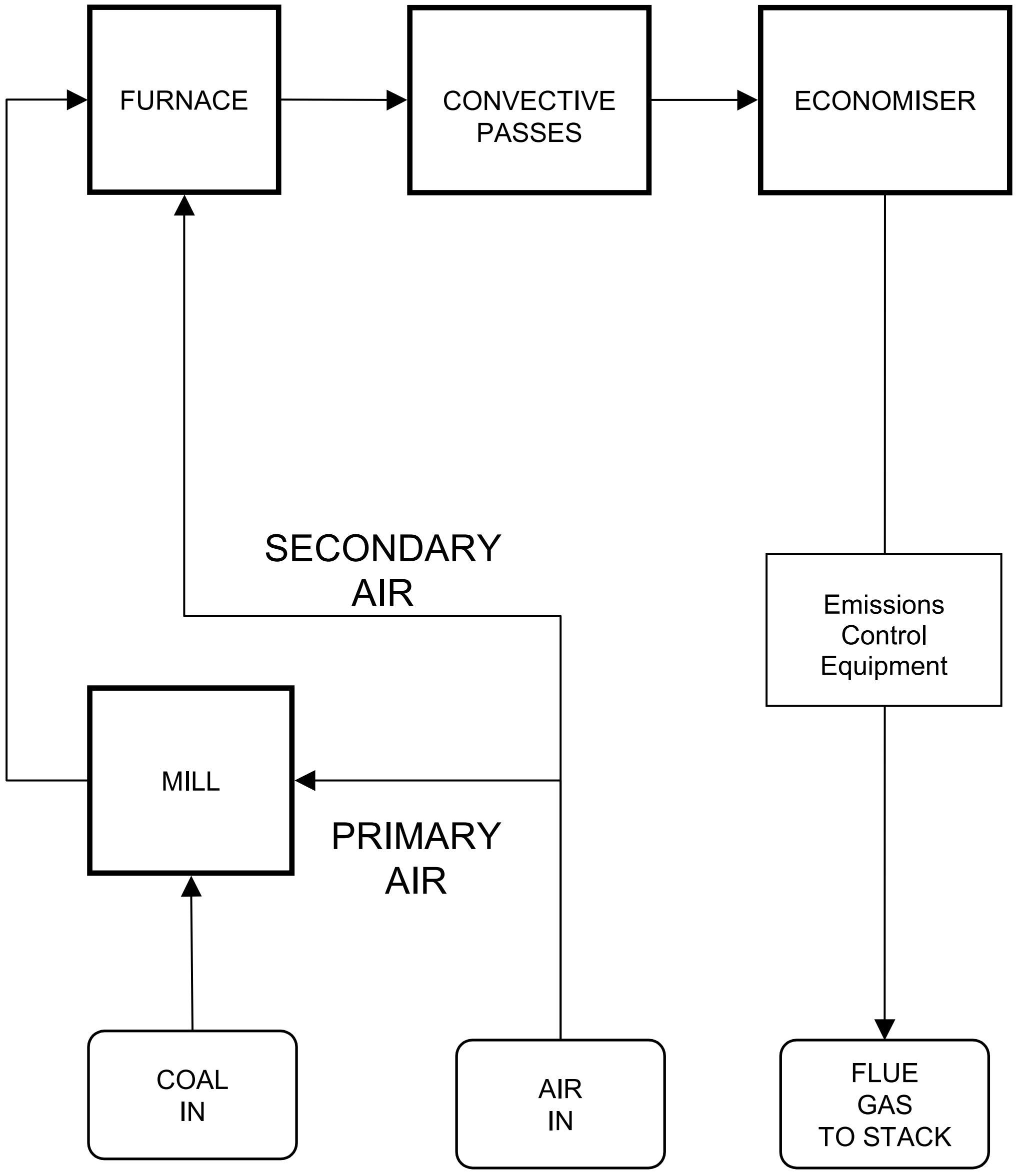
DESCRIPTION		Case 3 NGCC Power Plant Reference Case					Case 4 NGCCPower Plant with CO_2 Capture Oxy-Combustion				
		Base	+100% FP	-100% FP	5% DR	-20% TIC	Base	+100% FP	-100% FP	5% DR	-20% TIC
Run No.		3.0	3.1	3.2	3.3	3.4	4.0	4.1	4.2	4.3	4.4
Discount Rate (DR)	%	10	10	10	**5**	10	10	10	10	**5**	10
Fuel Price (FP)	US$/GJ	3.0	**6.0**	**0.0**	3.0	3.0	3.0	**6.0**	**0.0**	3.0	3.0
Fuel Mass Flow Rate	t/h	53.2	53.2	53.2	53.2	53.2	73.2	73.2	73.2	73.2	73.2
Gross Power Output	MWe Gross	400	400	400	400	400	575	575	575	575	575
Net Power Output	MWe Net	388	388	388	388	388	440	440	440	440	440
Fuel NCV	MJ/kg	46.9	46.9	46.9	46.9	46.9	48.5	48.5	48.5	48.5	48.5
Net Efficiency	% LHV	56.0	56.0	56.0	56.0	56.0	44.7	44.7	44.7	44.7	44.7
Total Investment Cost (TIC)*	M US$	217	217	217	217	173	658	658	658	658	526
Specific Investment Cost	US$/kWe Gross	542	542	542	542	434	1144	1144	1144	1144	915
Specific Investment Cost	US$/kWe Net	559	559	559	559	447	1495	1495	1495	1495	1196
Revenues / year	M US$/y	96.7	152.4	41.0	85.9	89.1	201.0	280.2	121.8	168.2	177.5
Electricity Cost	**US¢/kWh**	**3.35**	**5.28**	**1.42**	**2.97**	**3.08**	**6.13**	**8.55**	**3.71**	**5.13**	**5.41**
CO_2 Emissions	t/h	144	144	144	144	144	5	5	5	5	5
CO_2 Specific Emissions (based on net output)	g/kWh	370	370	370	370	370	12	12	12	12	12
CO_2 Captured	g/kWh	-	-	-	-	-	403	403	403	403	403
Cost of CO_2 Avoidance	**US$/t CO_2**						**78**	**91**	**64**	**60**	**65**

*Note all costs based on a 1 Euro to 1.2US$ exchange rate

NGCC POWER PLANT : ECONOMIC PERFORMANCE SUMMARY WITH & WITHOUT CO_2 CAPTURE

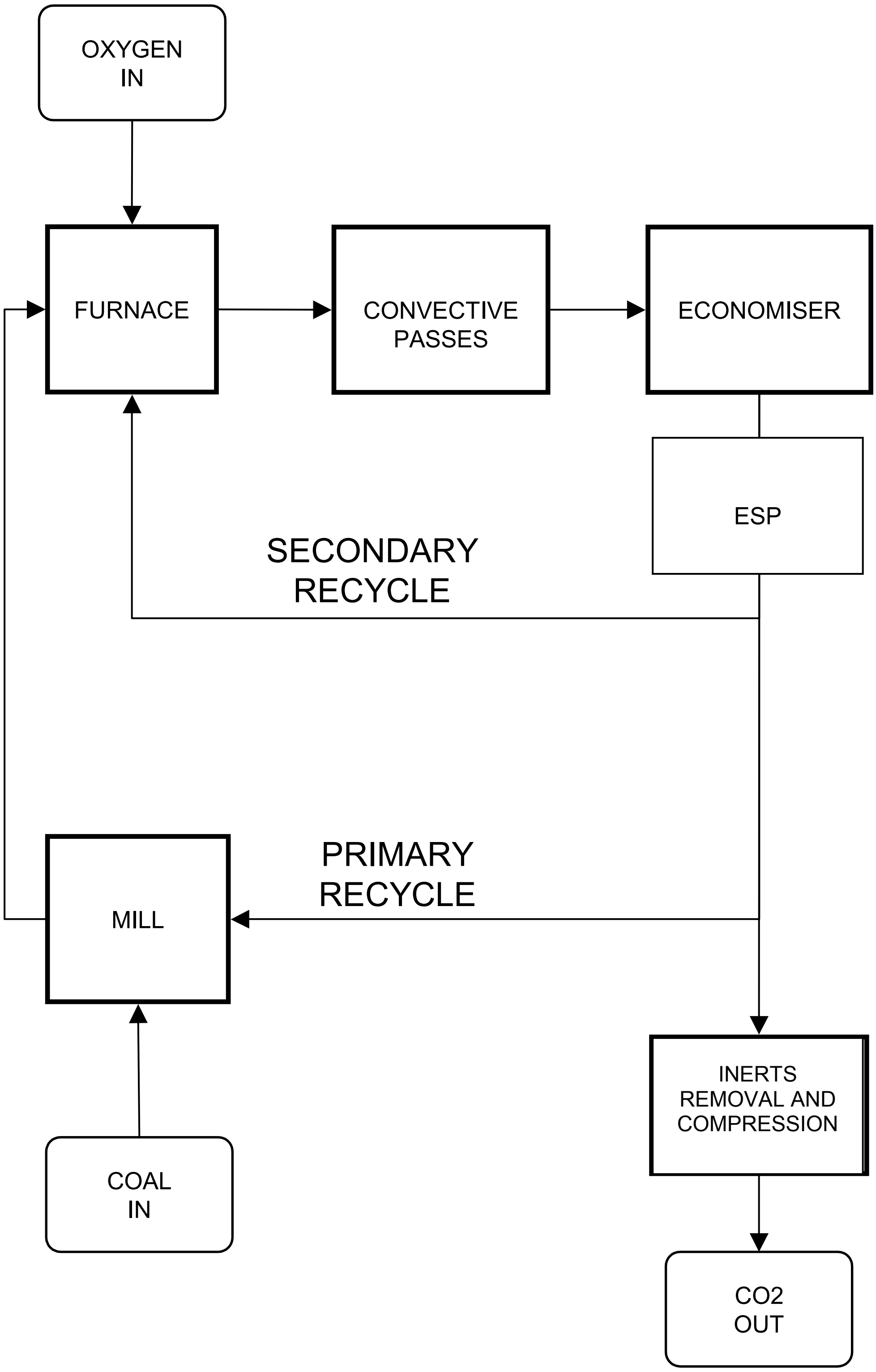
TABLE 41

FIGURES



CONVENTIONAL AIR-FIRED PULVERISED COAL (BITUMINOUS) COMBUSTION POWER PLANT
BASIC AIR COMBUSTION BLOCK DIAGRAM

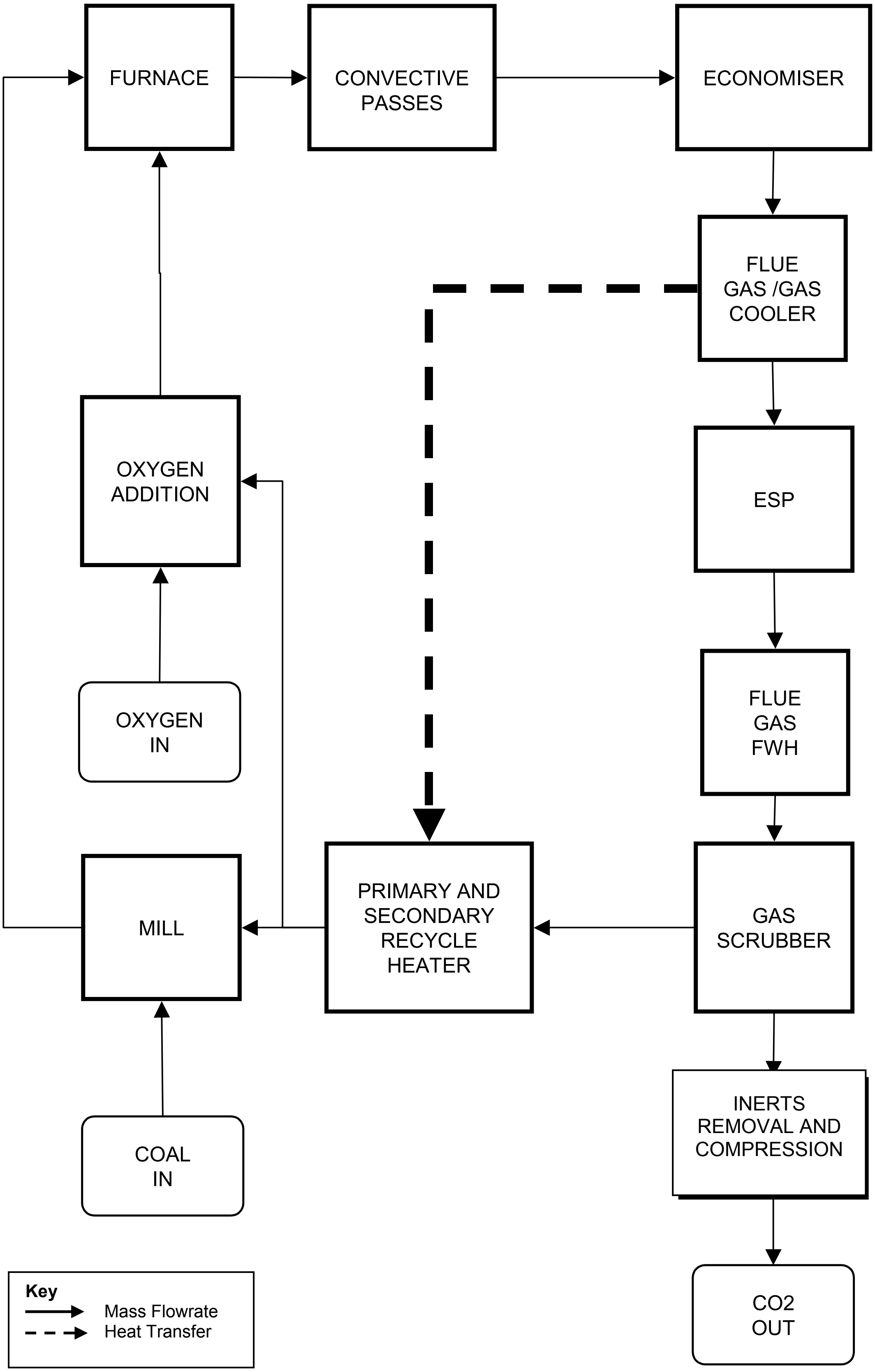
FIGURE 1



OXY-COMBUSTION PULVERISED COAL (BITUMINOUS) COMBUSTION POWER PLANT

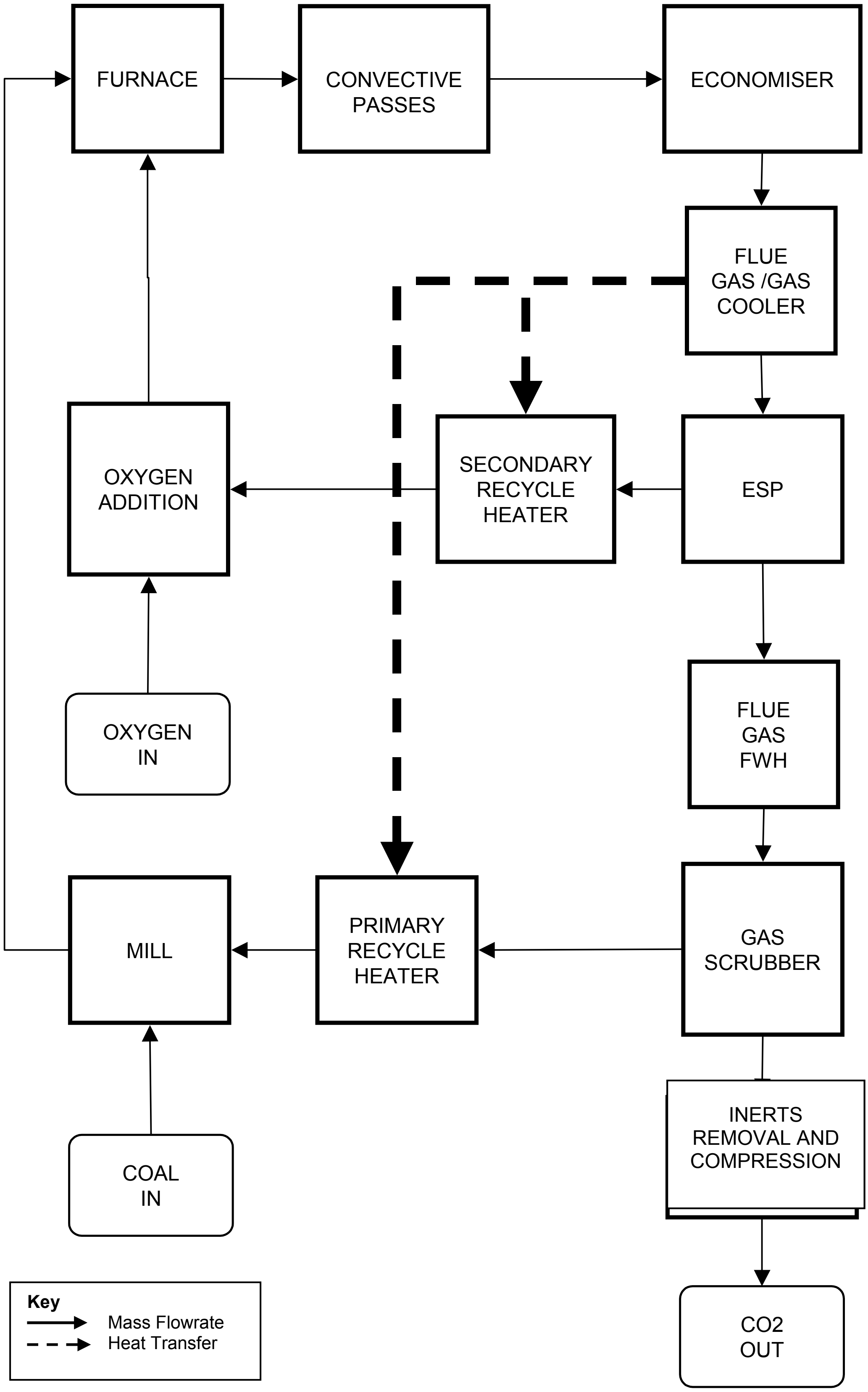
BASIC OXY-COMBUSTION BLOCK DIAGRAM

FIGURE 2



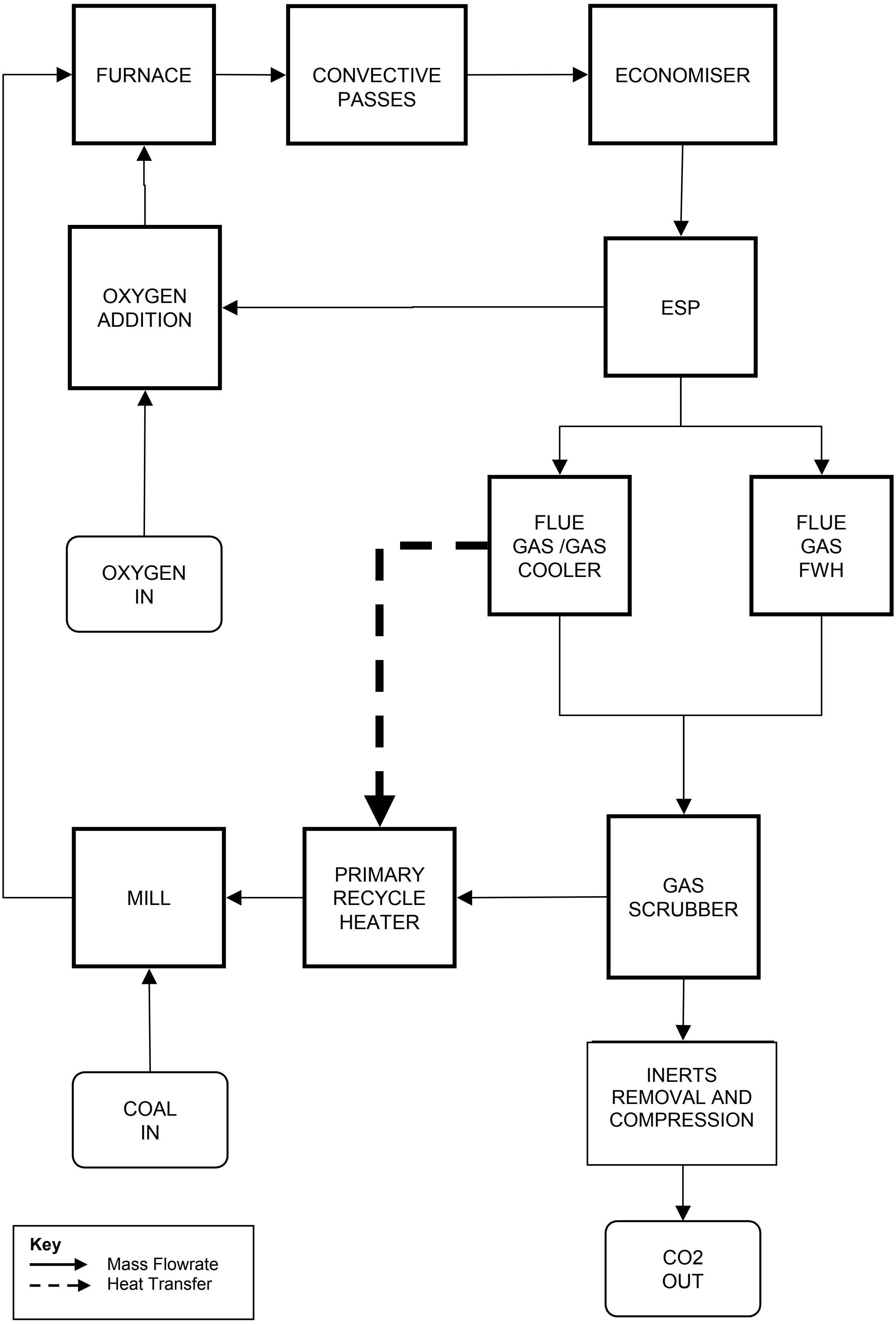
OXY-COMBUSTION PULVERISED COAL (BITUMINOUS) COMBUSTION OPTION A : COLD SECONDARY RECYCLE

FIGURE 3



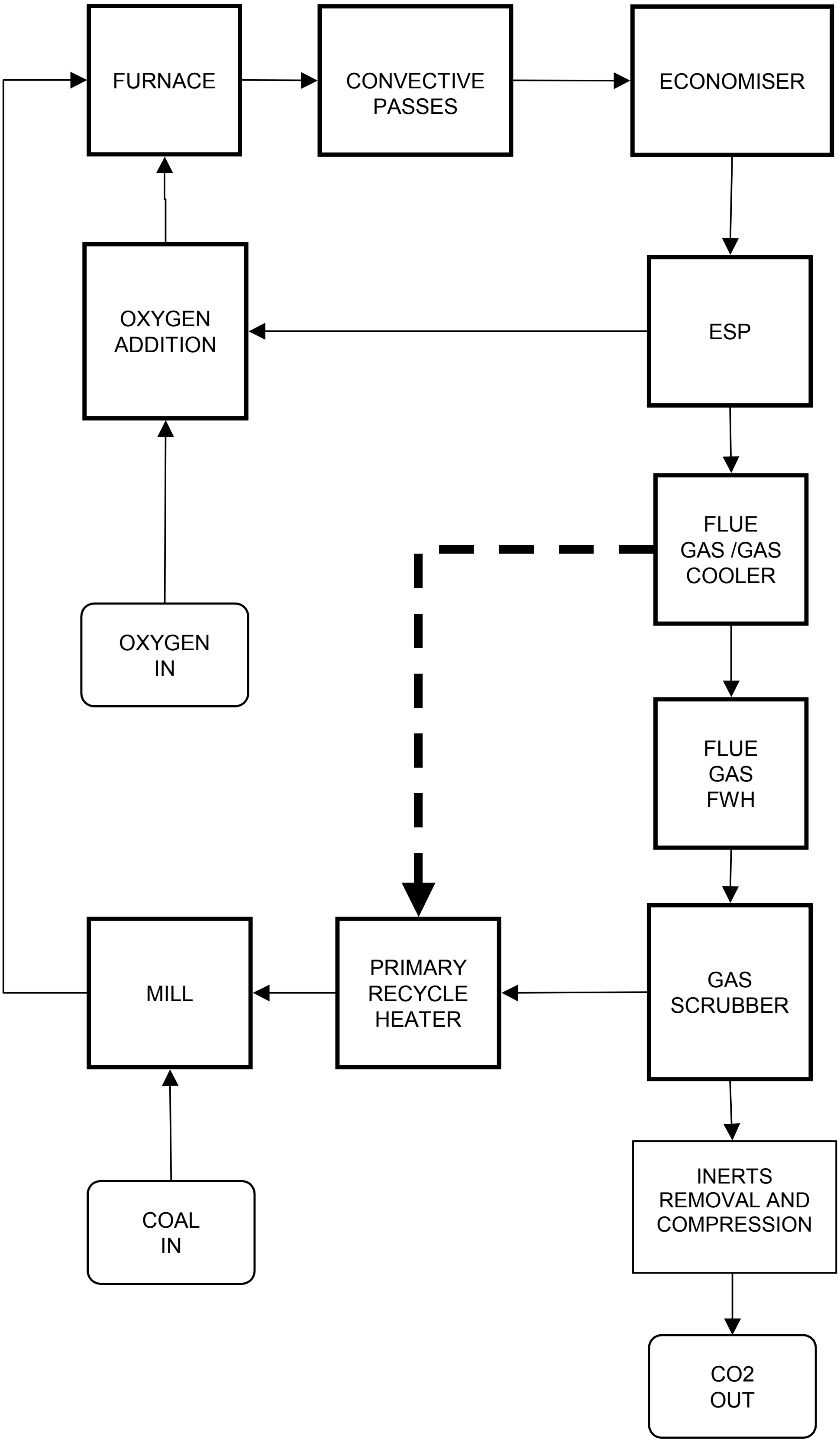
OXY-COMBUSTION PULVERISED COAL (BITUMINOUS) COMBUSTION
OPTION B : WARM SECONDARY RECYCLE

FIGURE 4



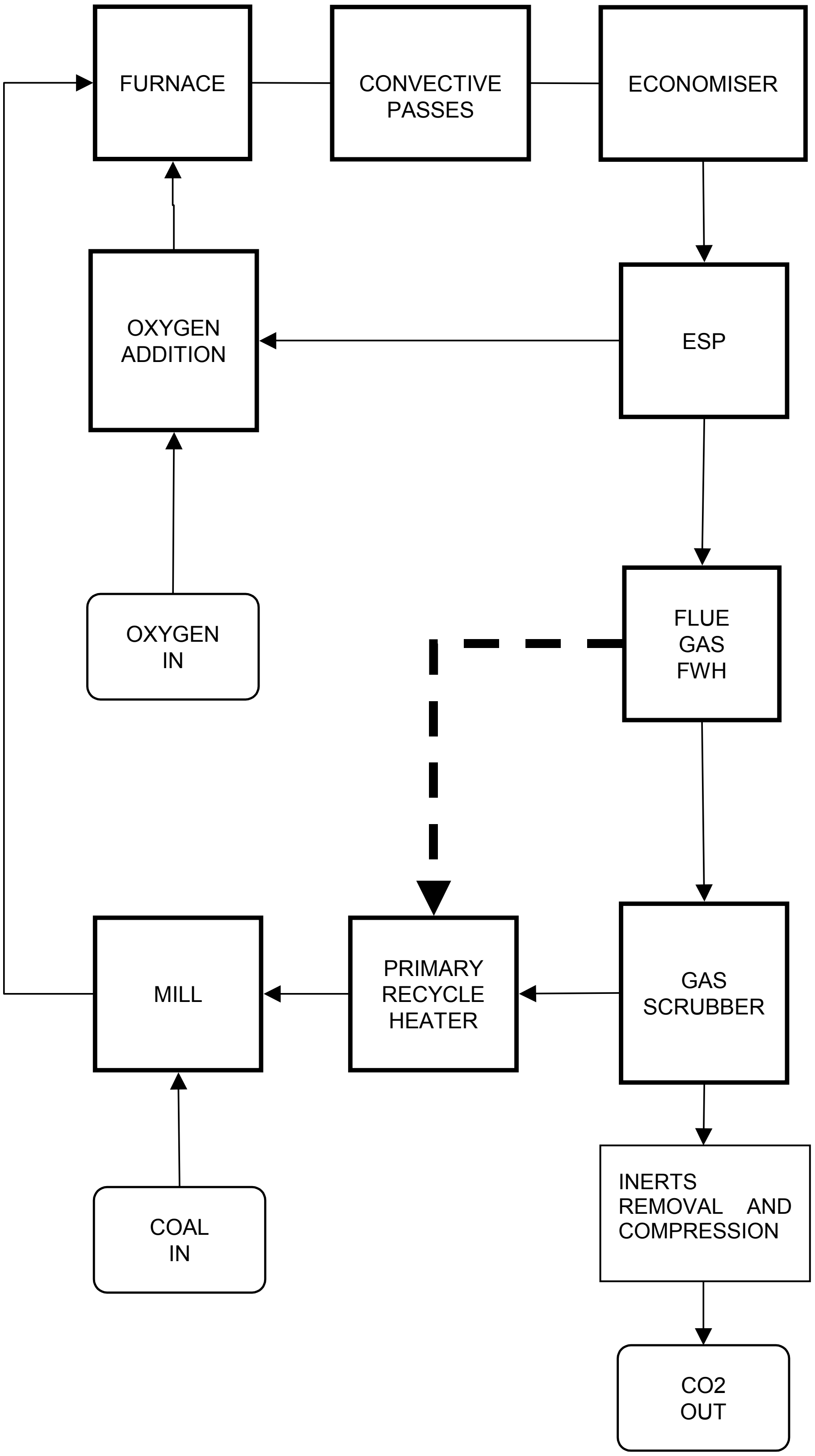
**OXY-COMBUSTION PULVERISED COAL (BITUMINOUS) COMBUSTION
OPTION C1 : HOT SECONDARY RECYCLE / PARALLEL GAS HEATER AND FWH**

FIGURE 5



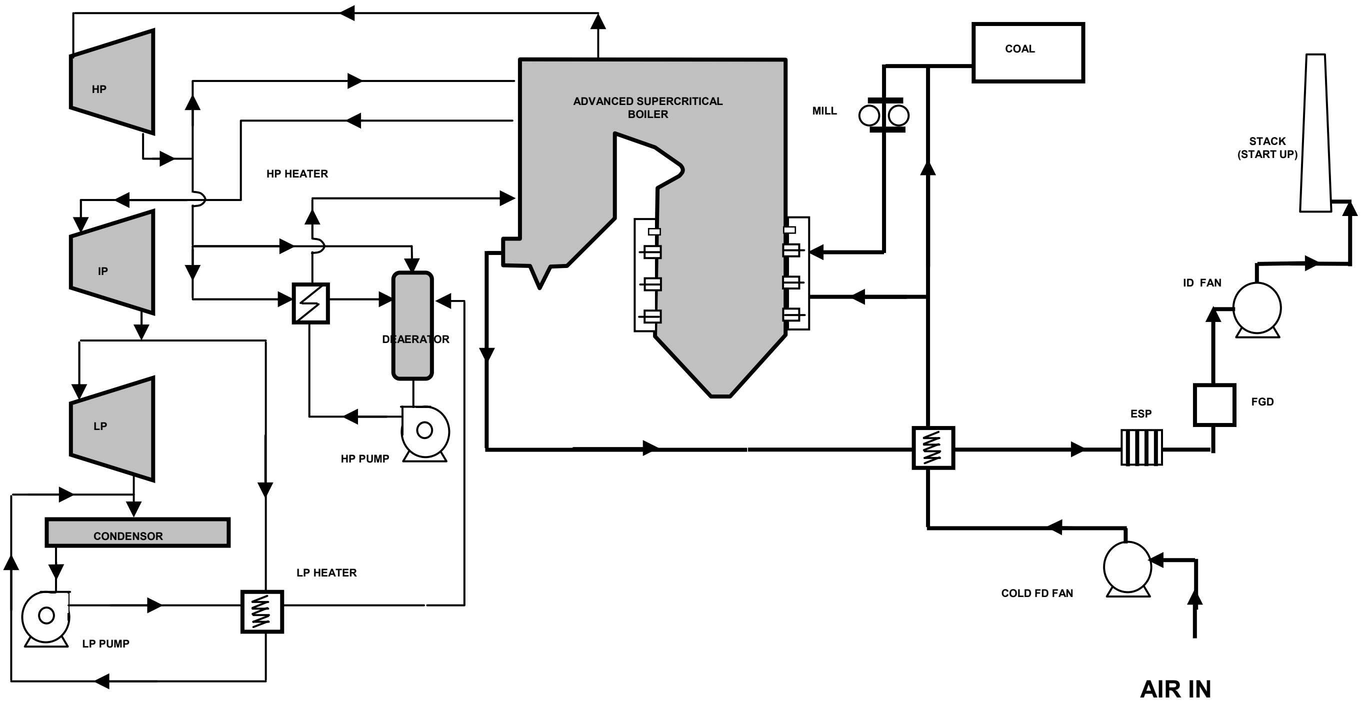
OXY-COMBUSTION PULVERISED COAL (BITUMINOUS) COMBUSTION
OPTION C2 : HOT SECONDARY RECYCLE / SERIES GAS HEATER AND FWH

FIGURE 6



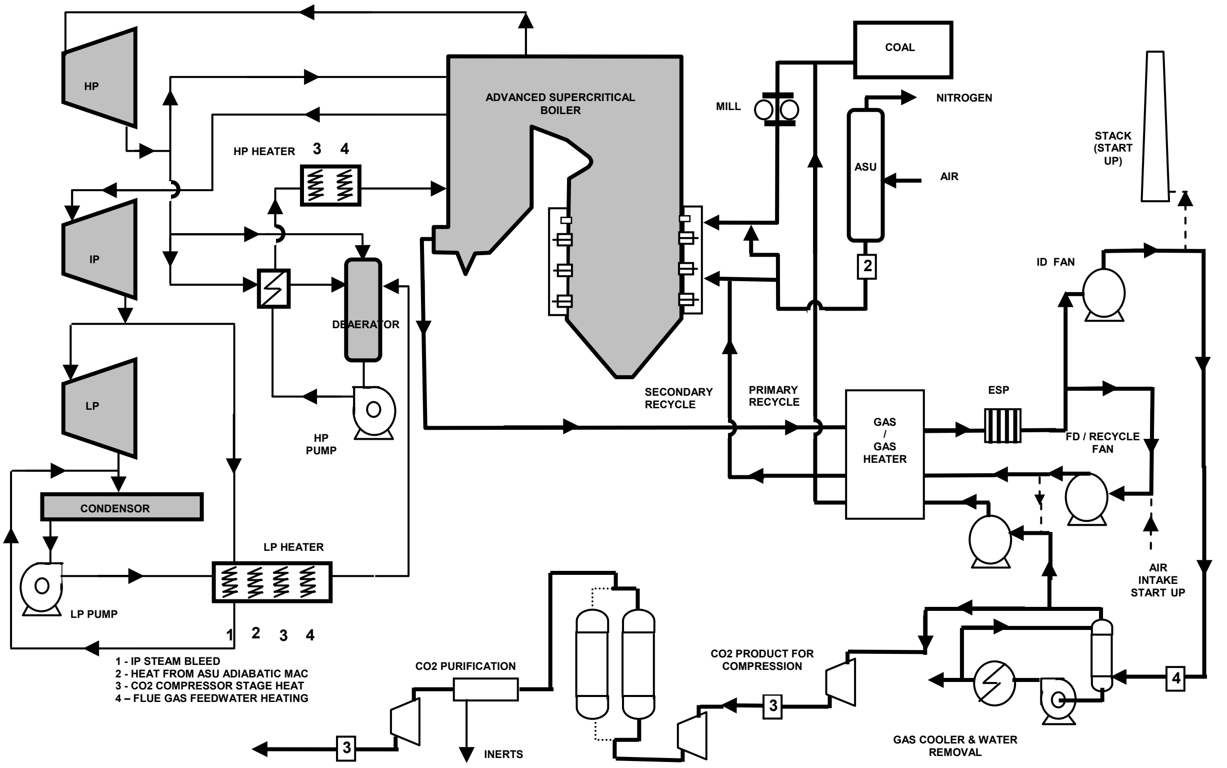
OXY-COMBUSTION PULVERISED COAL (BITUMINOUS) COMBUSTION
OPTION C3 : HOT SECONDARY RECYCLE / WATER HEATING FOR PRIMARY RECYCLE

FIGURE 7



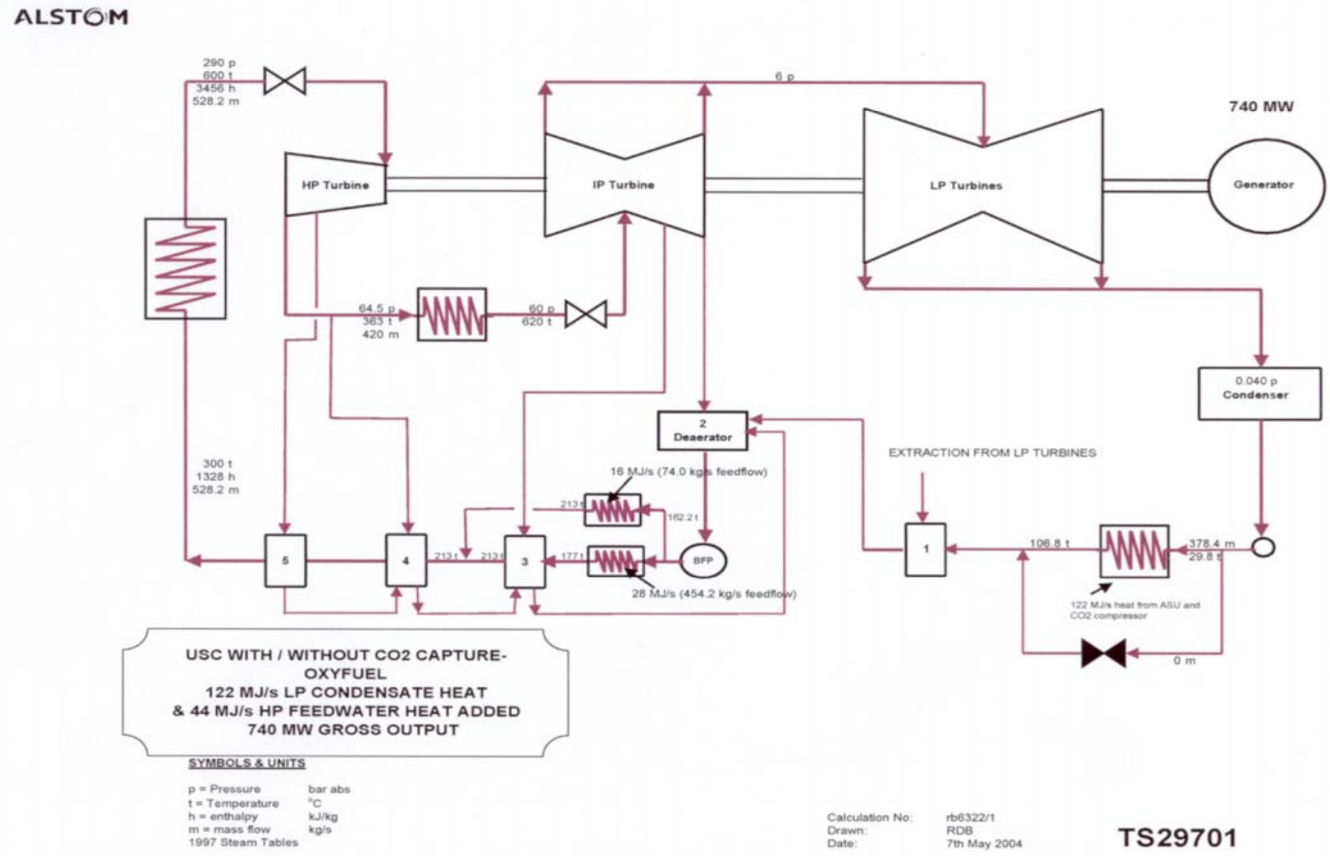
CONVENTIONAL AIR FIRED PF (BITUMINOUS) PLANT ARRANGEMENT

FIGURE 8



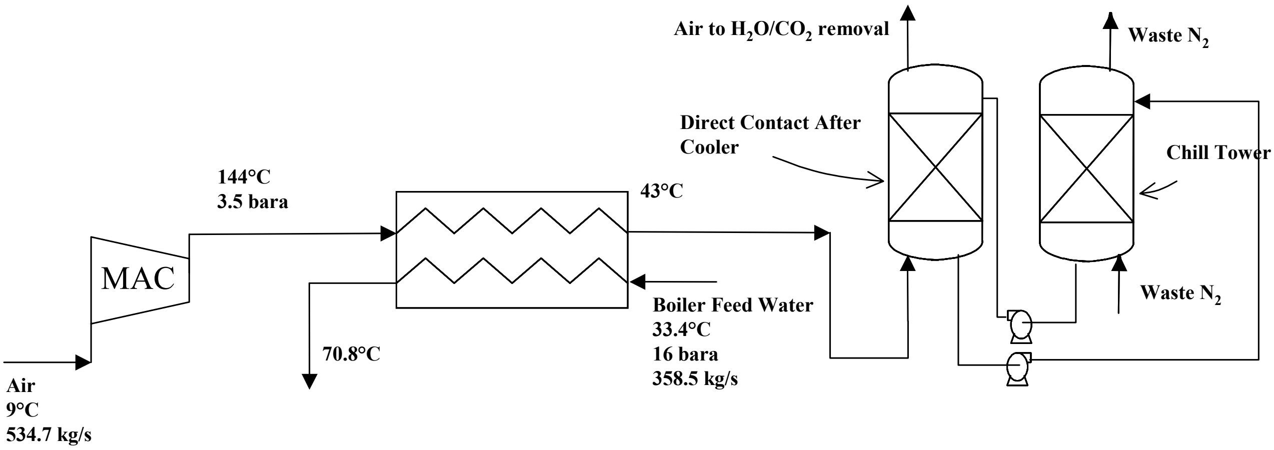
OXY-COMBUSTION ASC PF POWER PLANT WITH CO_2 CAPTURE ARRANGEMENT

FIGURE 9



OXY-COMBUSTION ASC PF POWER PLANT WITH CO_2 CAPTURE PLANT
FEED HEATING ARRANGEMENT

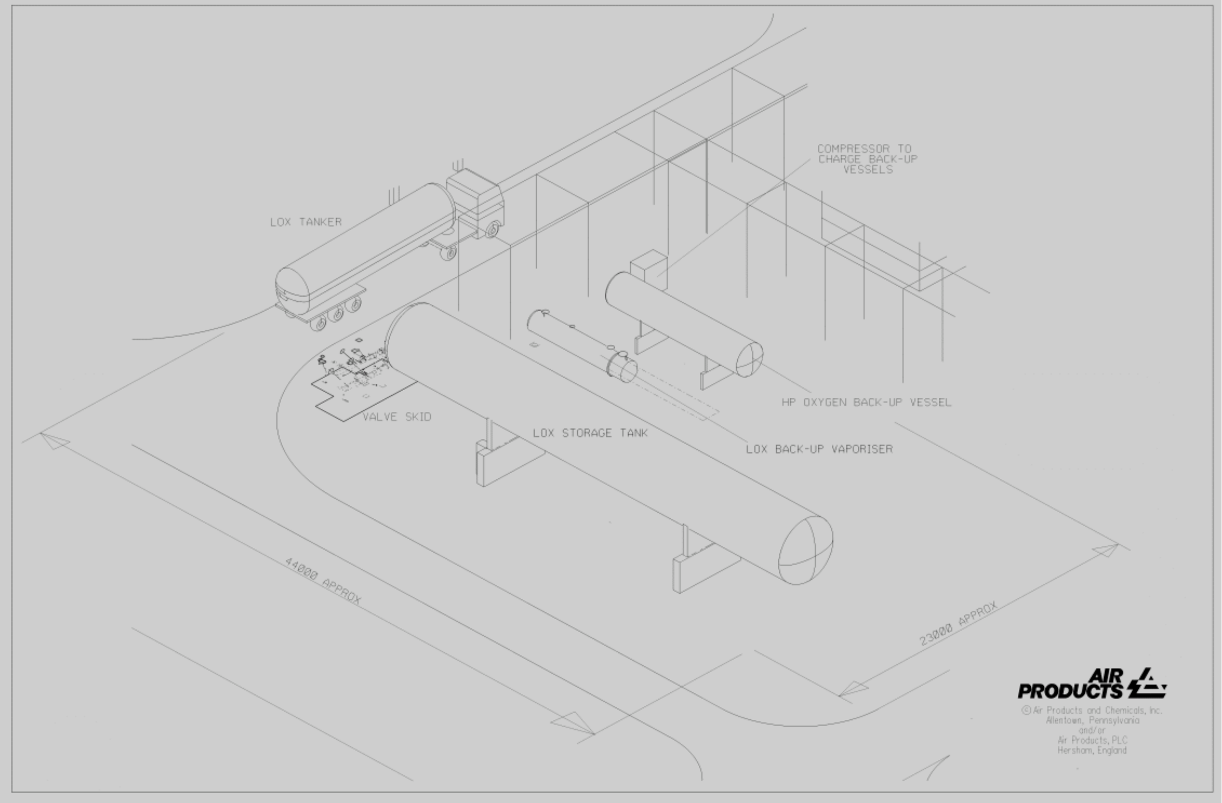
FIGURE 10



Note: Flowrates are totals over both ASUs

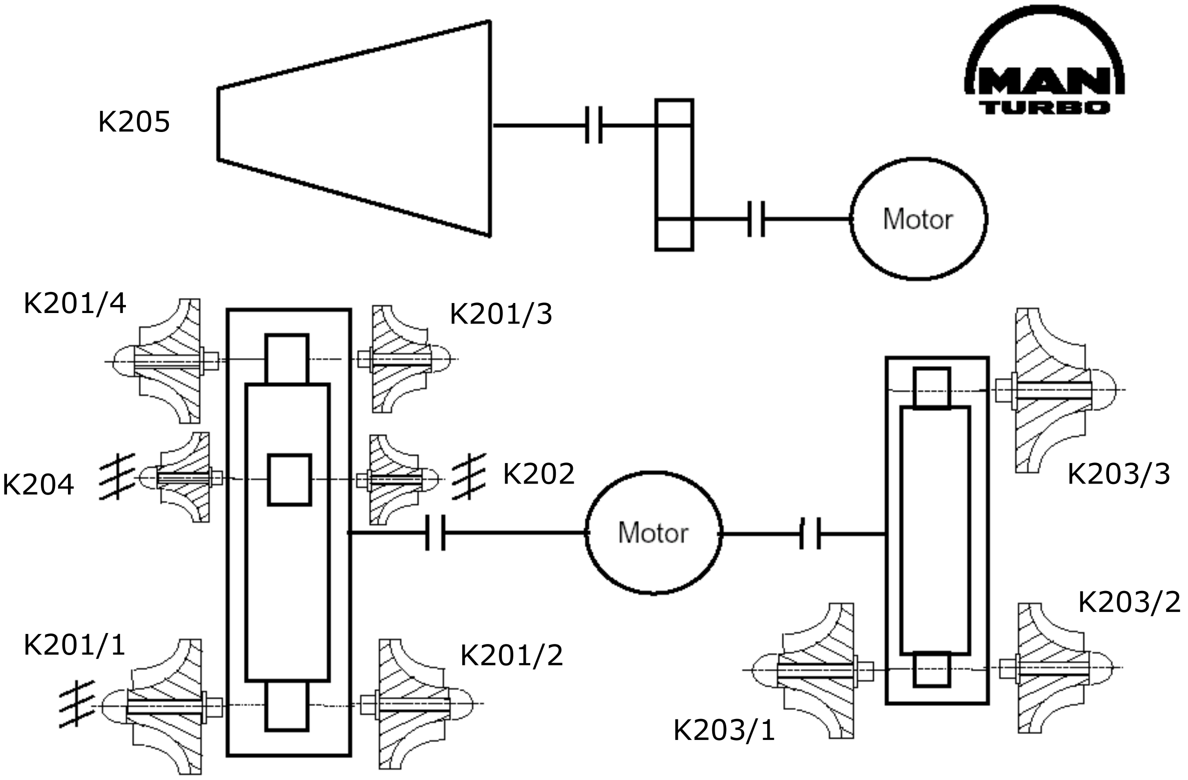
ADIABATIC MAIN AIR COMPRESSOR (MAC) ARRANGEMENT

FIGURE 11



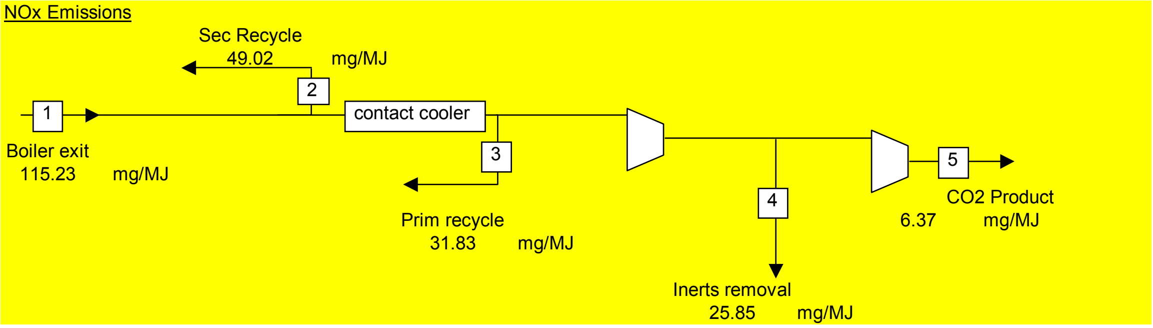
OXYGEN BACK UP SYSTEM ISOMETRIC

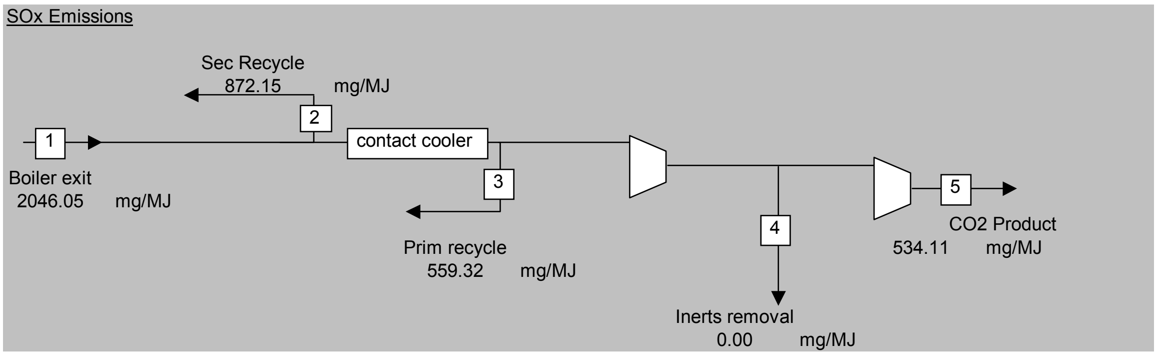
FIGURE 12

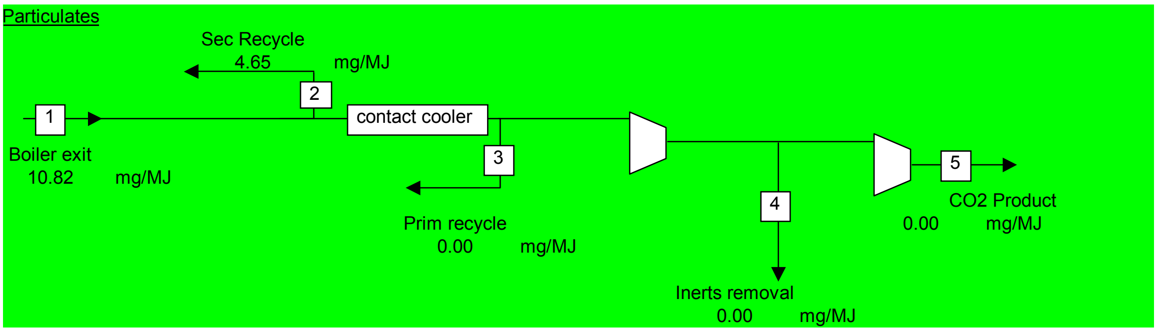


ASC PF POWER PLANT WITH CO_2 CAPTURE
CO_2 COMPRESSION TRAIN LAYOUT

FIGURE 13



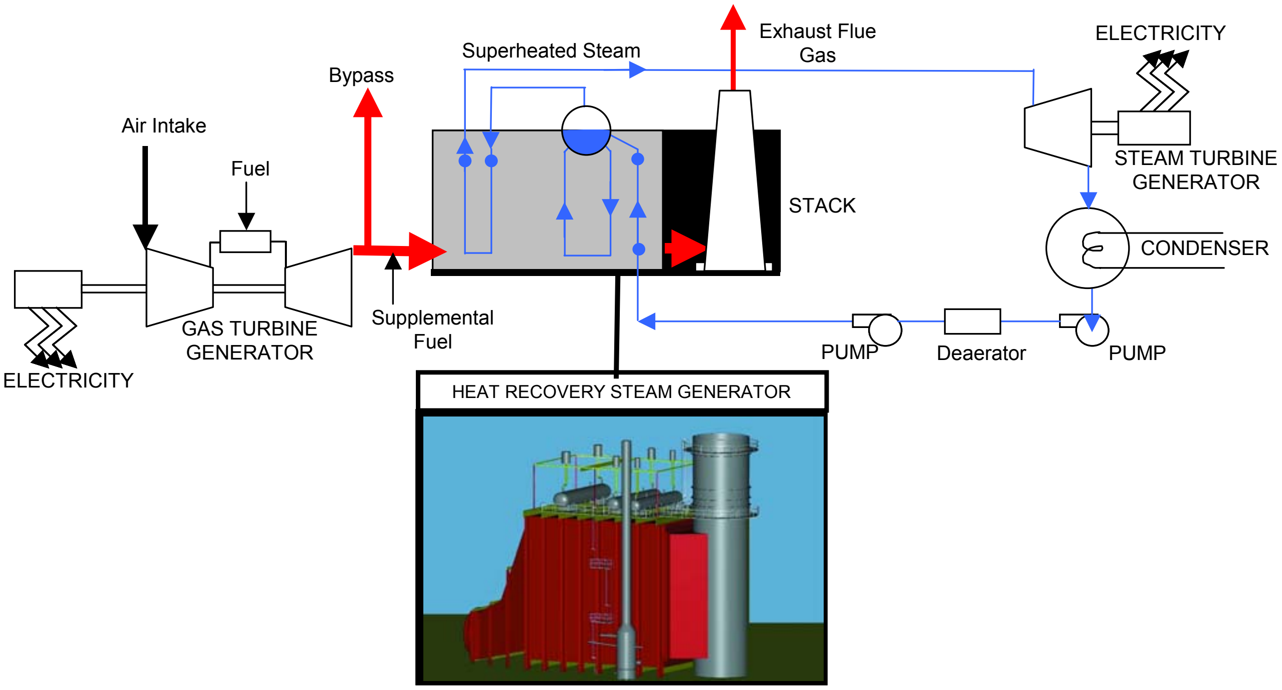




Note : Emissions presented are at actual oxygen levels and have not been normalised to 6% O_2, v/v, dry

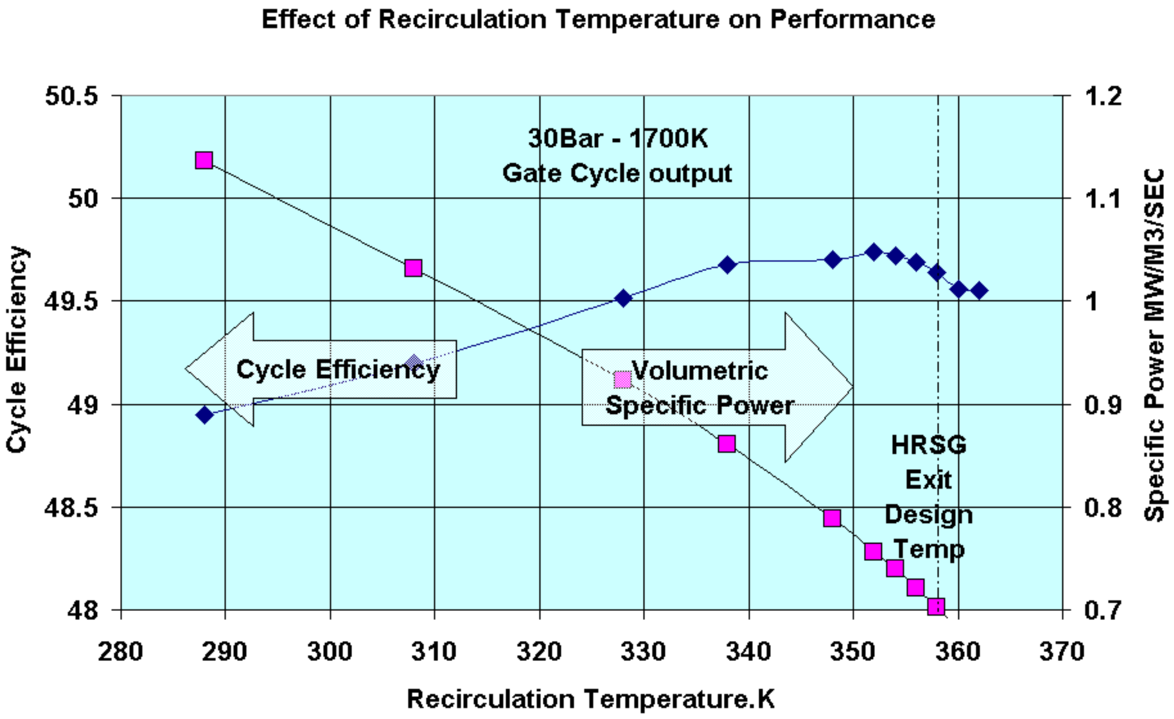
OXY-COMBUSTION ASC PF POWER PLANT WITH CO_2 CAPTURE
EMISSONS SUMMARY

FIGURE 14



CONVENTIONAL AIR FIRED NGCC POWER PLANT ARRANGEMENT

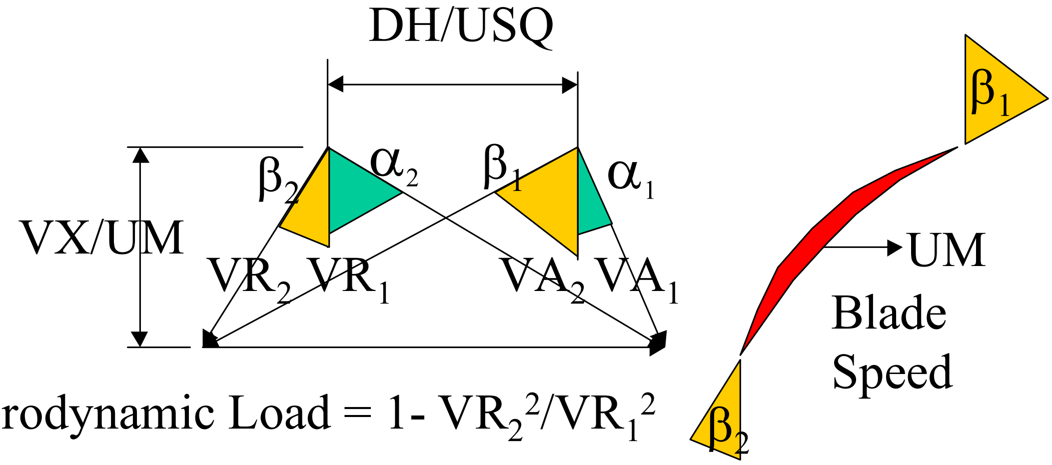
FIGURE 15



NGCC POWER PLANT WITH CO_2 CAPTURE
THE EFFECT OF FLUE GAS RECRCULATION TEMPERATURE ON GAS TURBINE PERFORMANCE

FIGURE 16

Compressor Vector Triangles



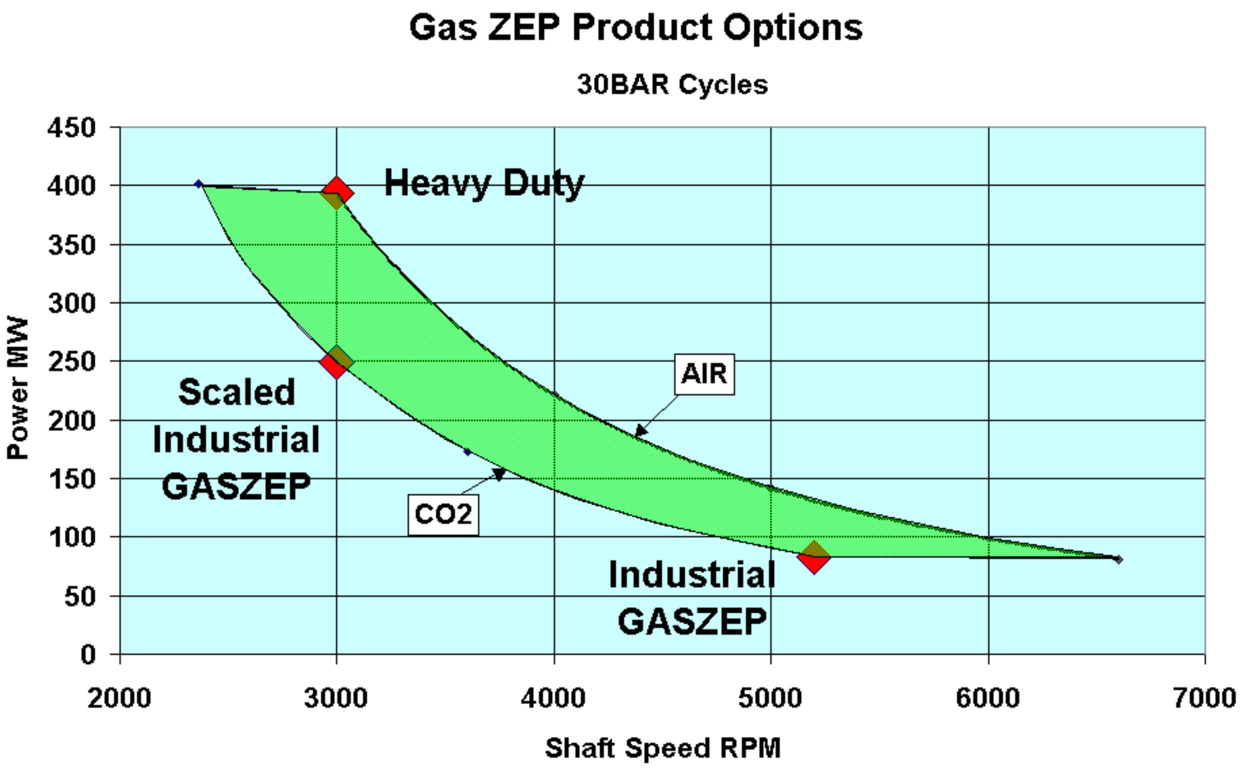
Aerodynamic Load = 1- VR_2^2/VR_1^2

Nomenclature for Vector Triangles :

DH	Stage Enthalpy rise
VX	Axial Velocity
UM	Mean Blade Speed
VA_1	Rotor Inlet Absolute Velocity
VA_2	Rotor Exit Absolute Velocity
VR_1	Rotor Inlet Relative Velocity
VR_2	Rotor Exit Relative Velocity
α_1	Rotor Inlet Absolute Flow Angle
α_2	Rotor Exit Absolute Flow Angle
β_1	Rotor Inlet Relative Flow Angle
β_2	Rotor Exit Relative Flow Angle

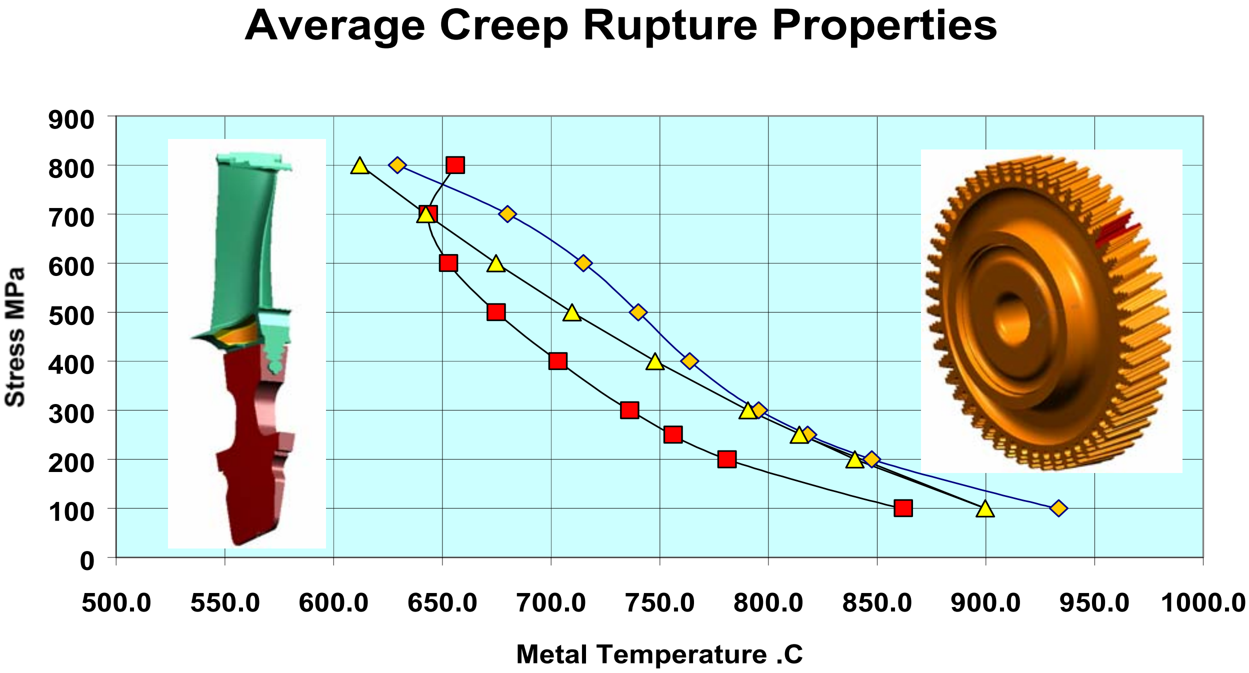
GT COMPRESSOR VECTOR TRIANGLES

FIGURE 17



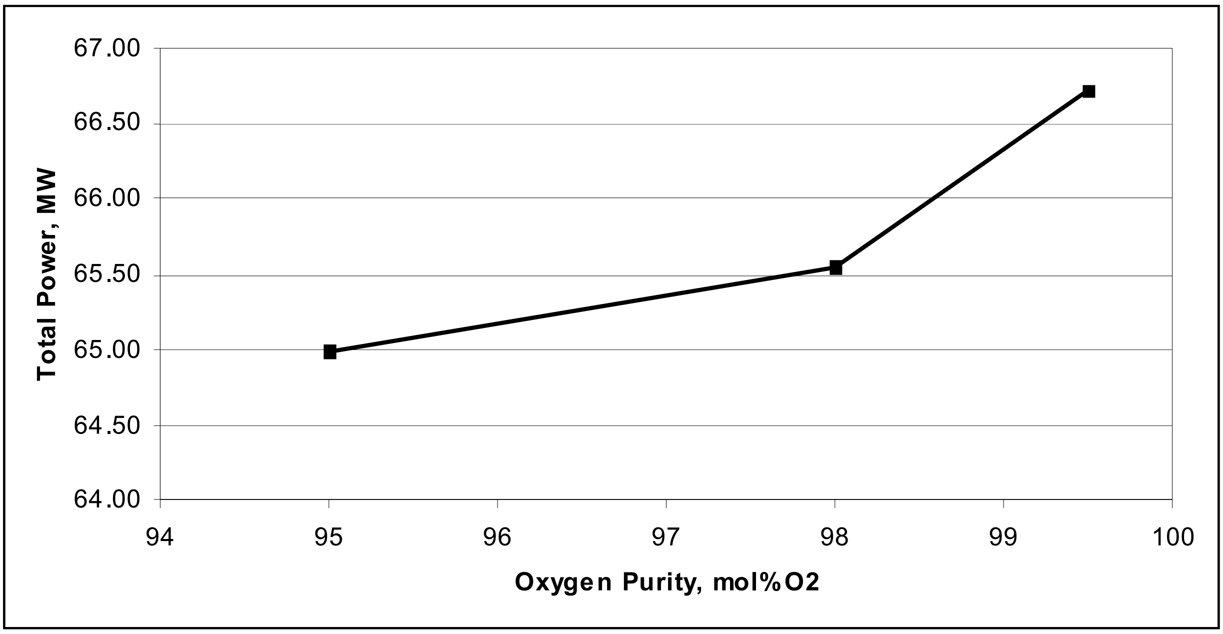
NGCC POWER PLANT : GAS TURBINE
EFFECT OF SYNCHRONOUS SPEED REQUIREMENTS ON COMBINED CYCLE POWER OUTPUT

FIGURE 18



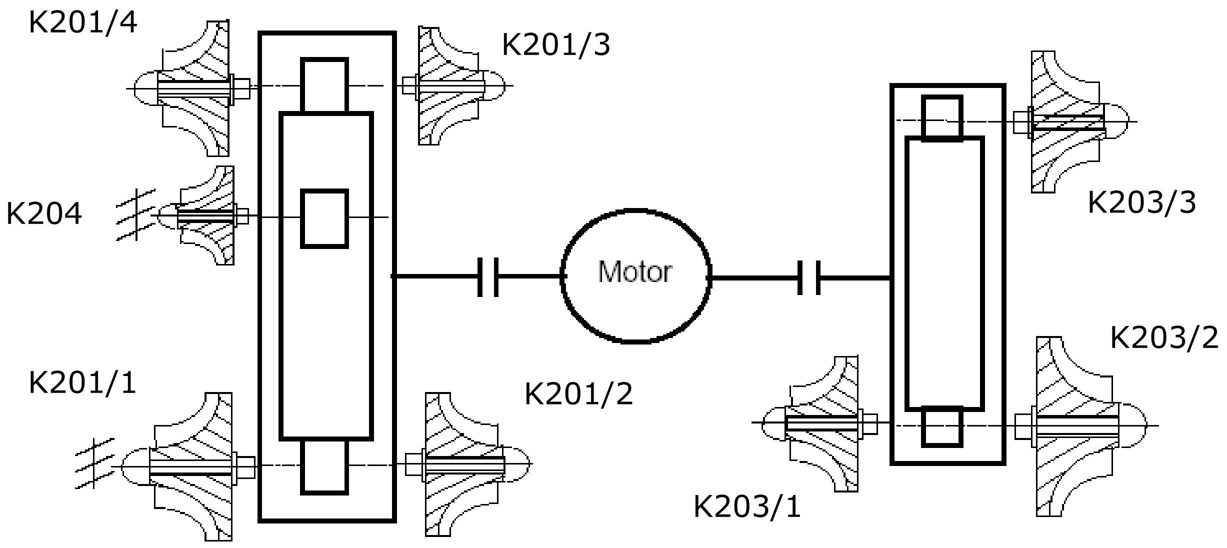
NGCC POWER PLANT : GAS TURBINE
AVERAGE CREEP RUPTURE PROPERTIES FOR CONSTANT LIFE

FIGURE 19



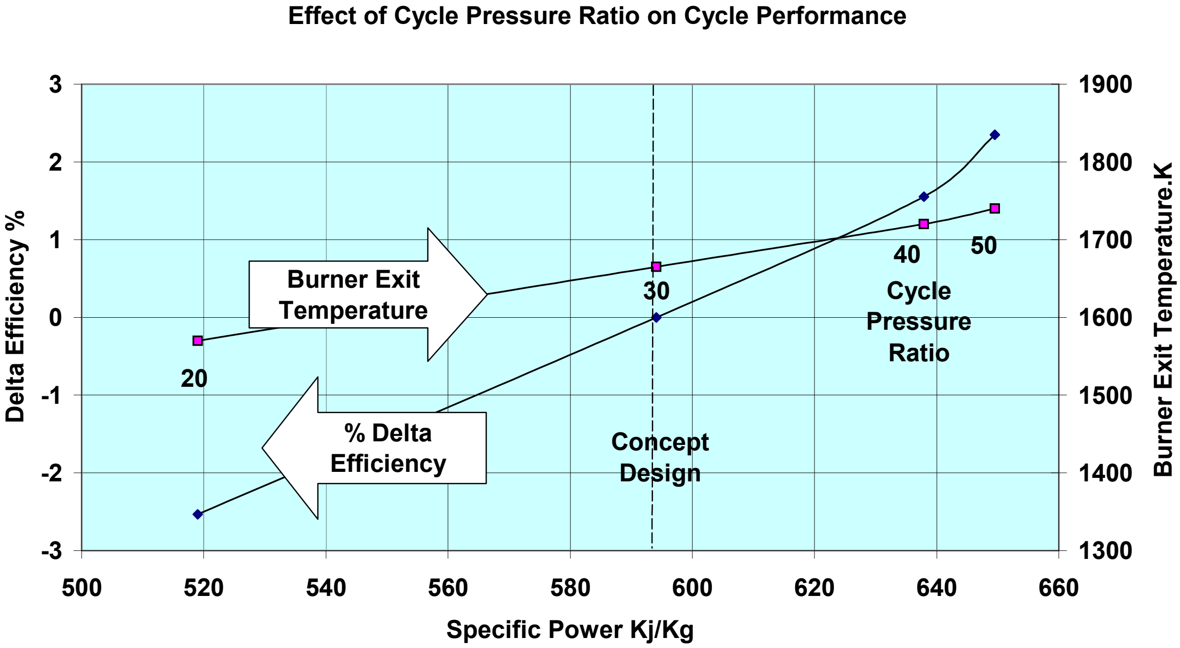
NGCC POWER PLANT WITH CO_2 CAPTURE
EFFECT OF OXYGEN PURITY ON TOTAL POWER REQUIREMENTS (ASU & CO_2 TREATMENT)

FIGURE 20



NGCC POWER PLANT WITH CO_2 CAPTURE
CO_2 COMPRESSION TRAIN LAYOUT

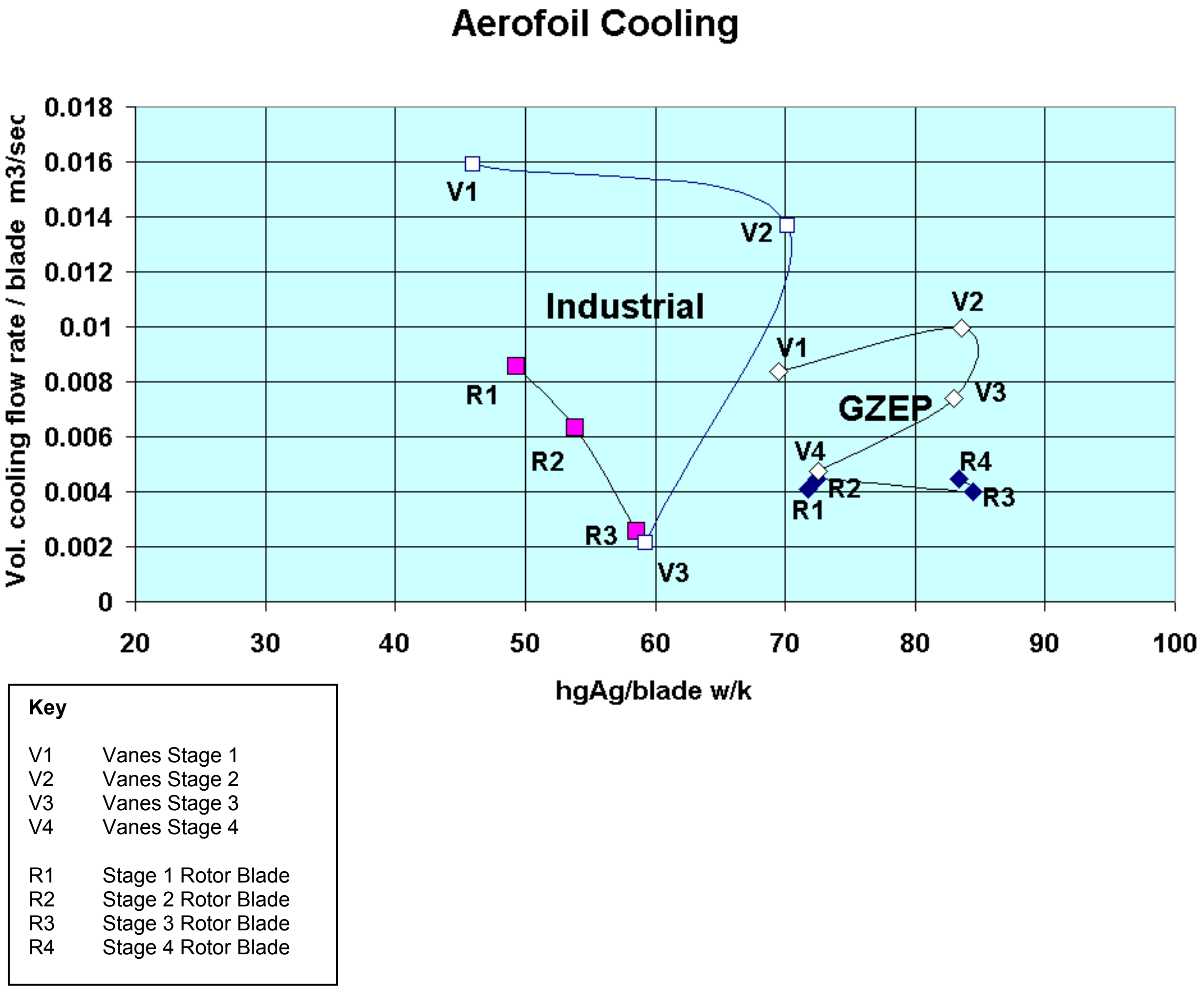
FIGURE 21



Note : Delta cycle efficiency excludes CO_2 compression power requirements

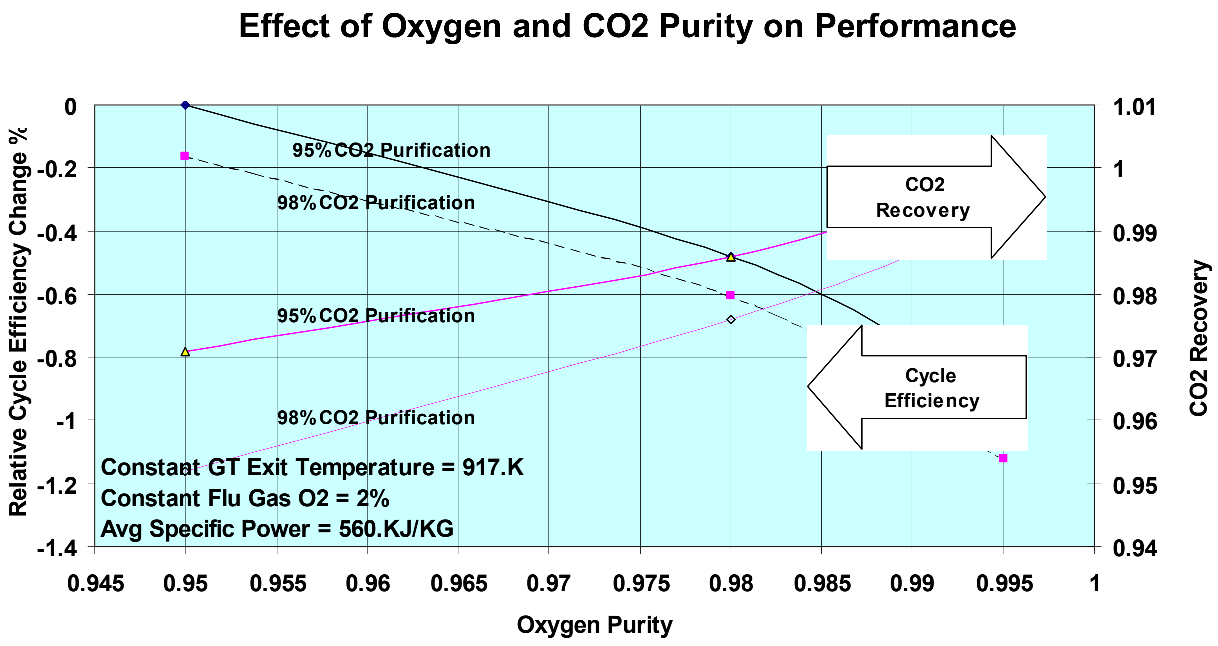
NGCC POWER PLANT WITH CO_2 CAPTURE EFFECT OF GT PRESSURE RATIO ON CYCLE EFFICIENCY

FIGURE 22



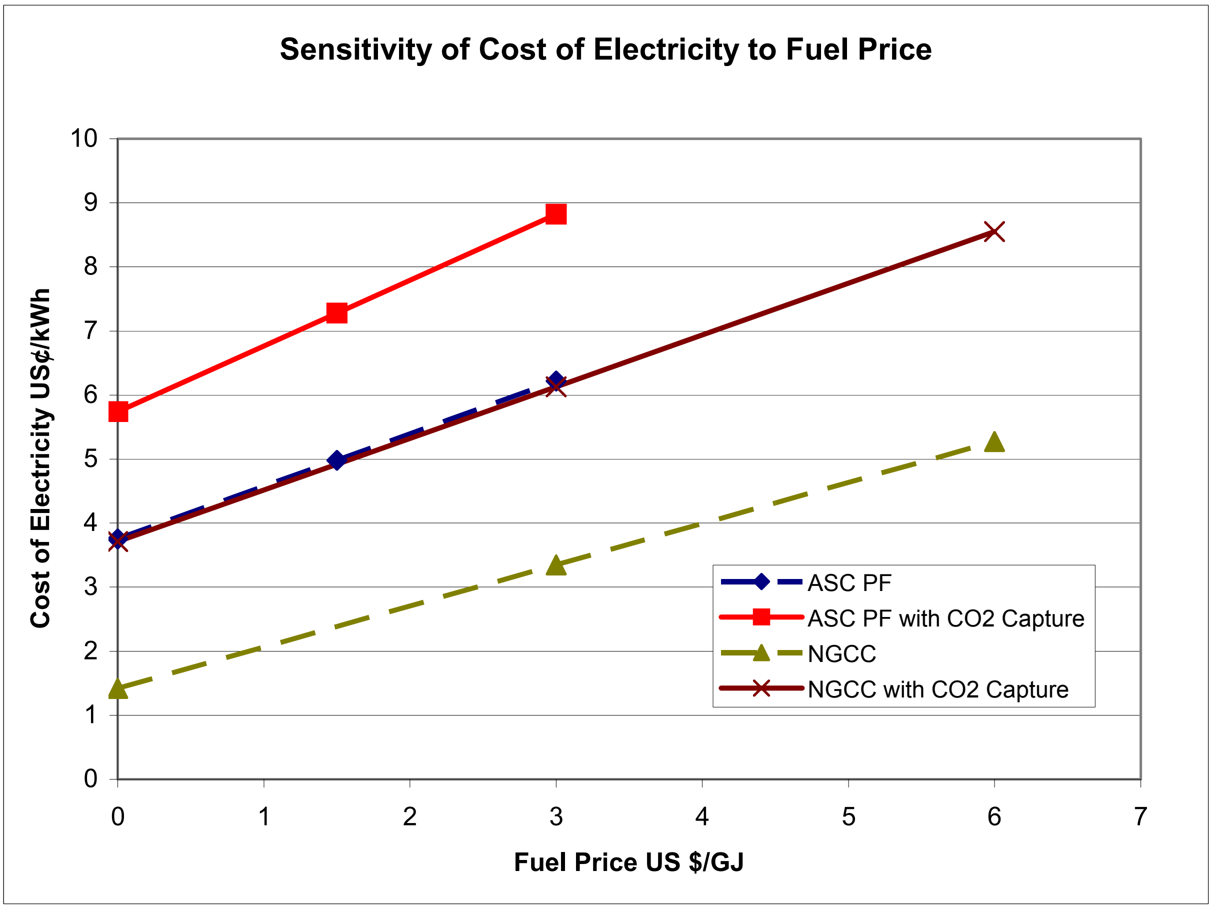
NGCC POWER PLANT WITH CO_2 CAPTURE
EFFECT OF CO_2 & CYCLE PRESSURE RATIO ON GT TURBINE ROTOR BLADE COOLING

FIGURE 23



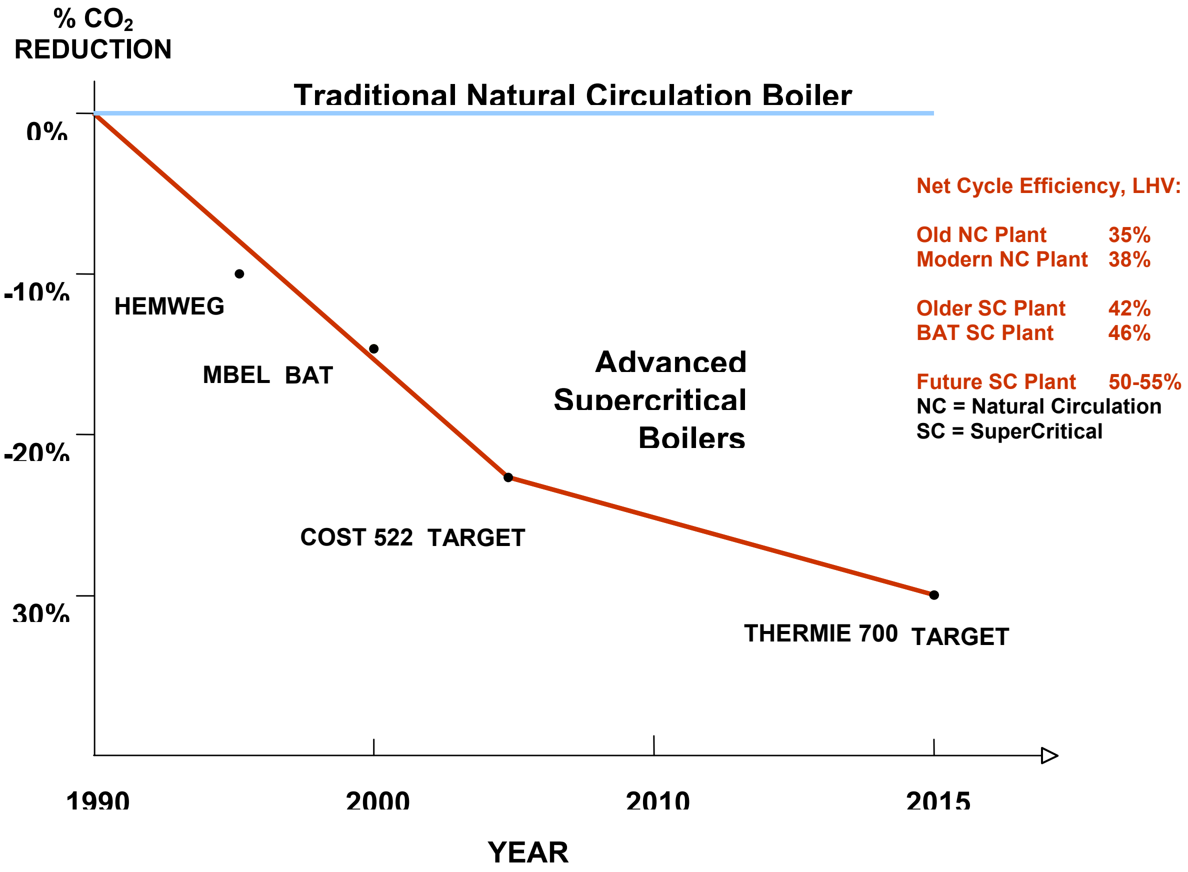
NGCC POWER PLANT WITH CO_2 CAPTURE
EFFECT OF OXYGEN AND CO_2 PURITY ON CYCLE PERFORMANCE

FIGURE 24



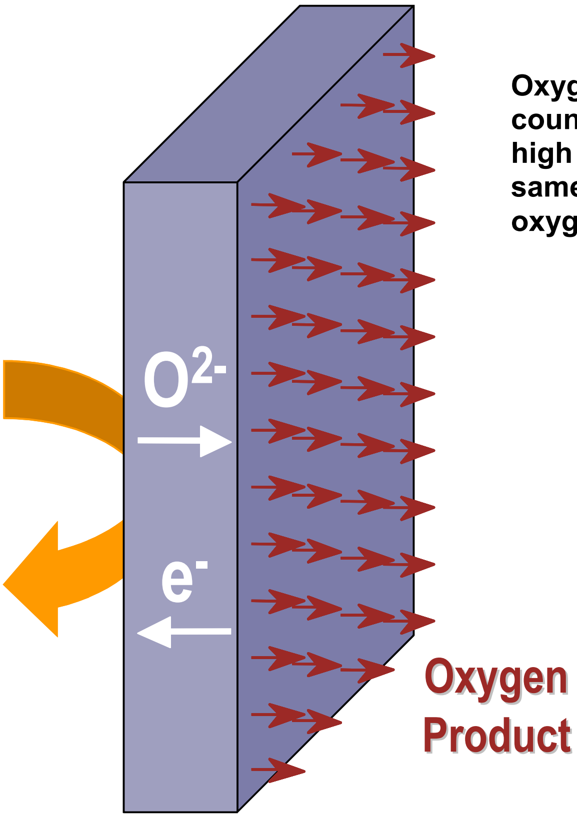
ASC PF POWER PLANT & NGCC POWER PLANT
SENSITIVITY OF COST OF ELECTRICITY TO FUEL PRICE : WITH & WITHOUT CO2 CAPTURE

FIGURE 25



CO_2 ABATEMENT BY EFFICIENCY IMPROVEMENT

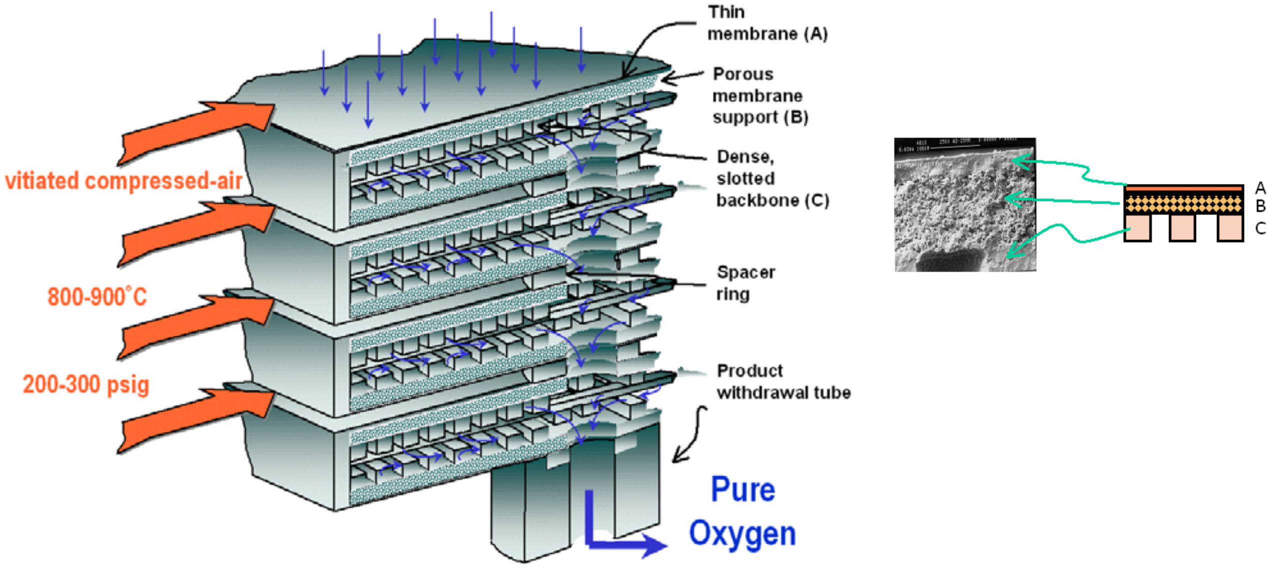
FIGURE 26



Oxygen anions move counter to electrons at high temperature in the same material under an oxygen activity gradient

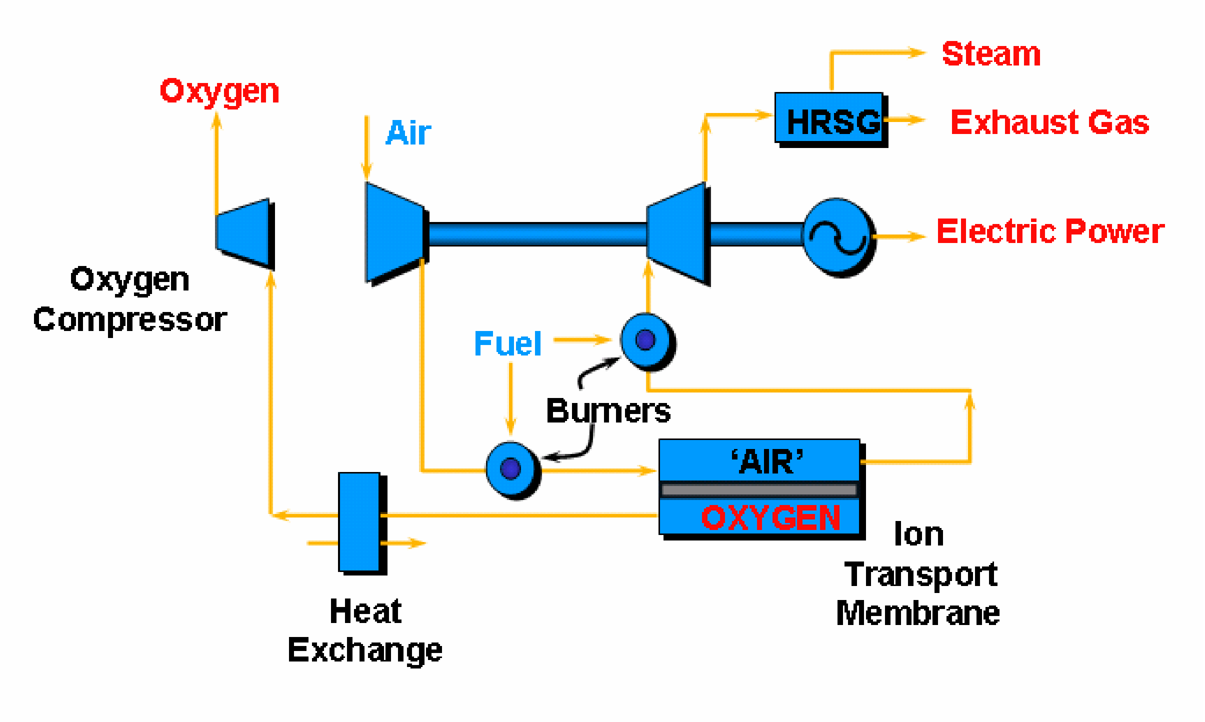
MIXED CONDUCTING ION TRANSPORT MEMBRANE

FIGURE 27



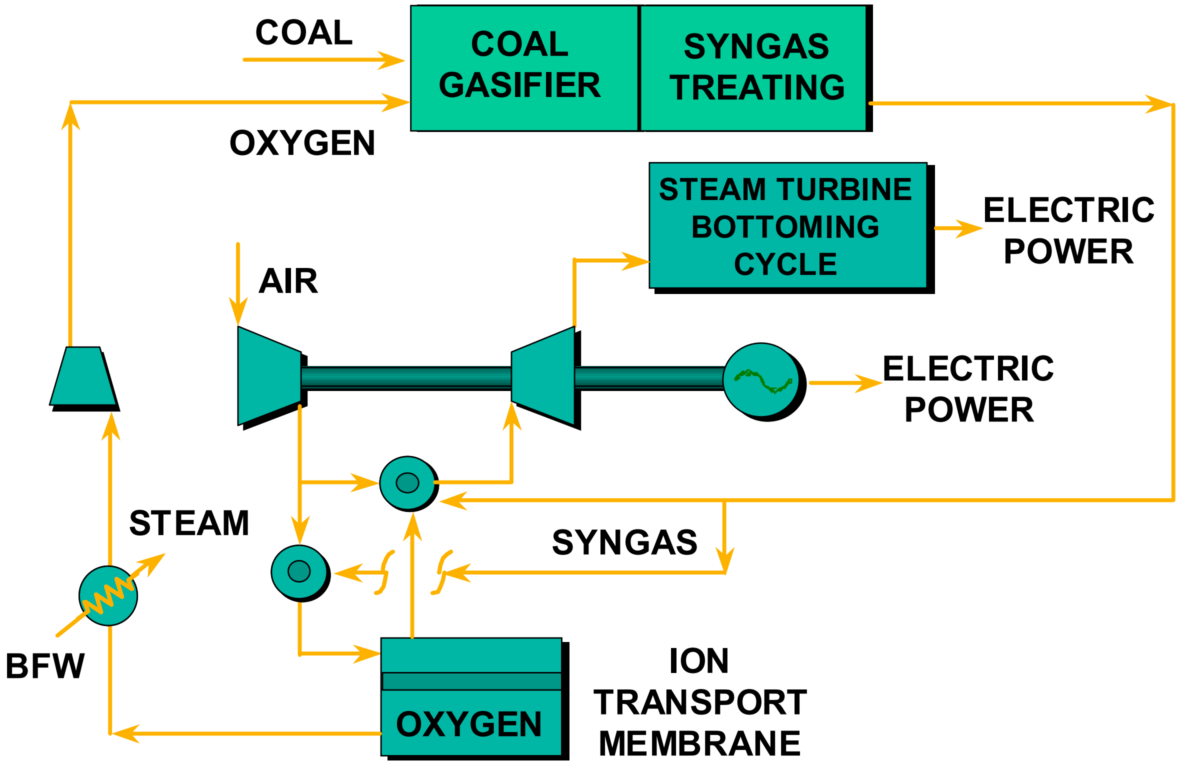
SCHEMATIC OF ITM OXYGEN PLANAR SUPPORTED MEMBRANE DEVICE

FIGURE 28



NGCC POWER PLANT
BASIC SCHEME FOR INTEGRATION OF AN ITM OXYGEN AIR SEPARATION UNIT INTO A GAS TURBINE POWER CYCLE (Process outputs are oxygen, power and steam)

FIGURE 29



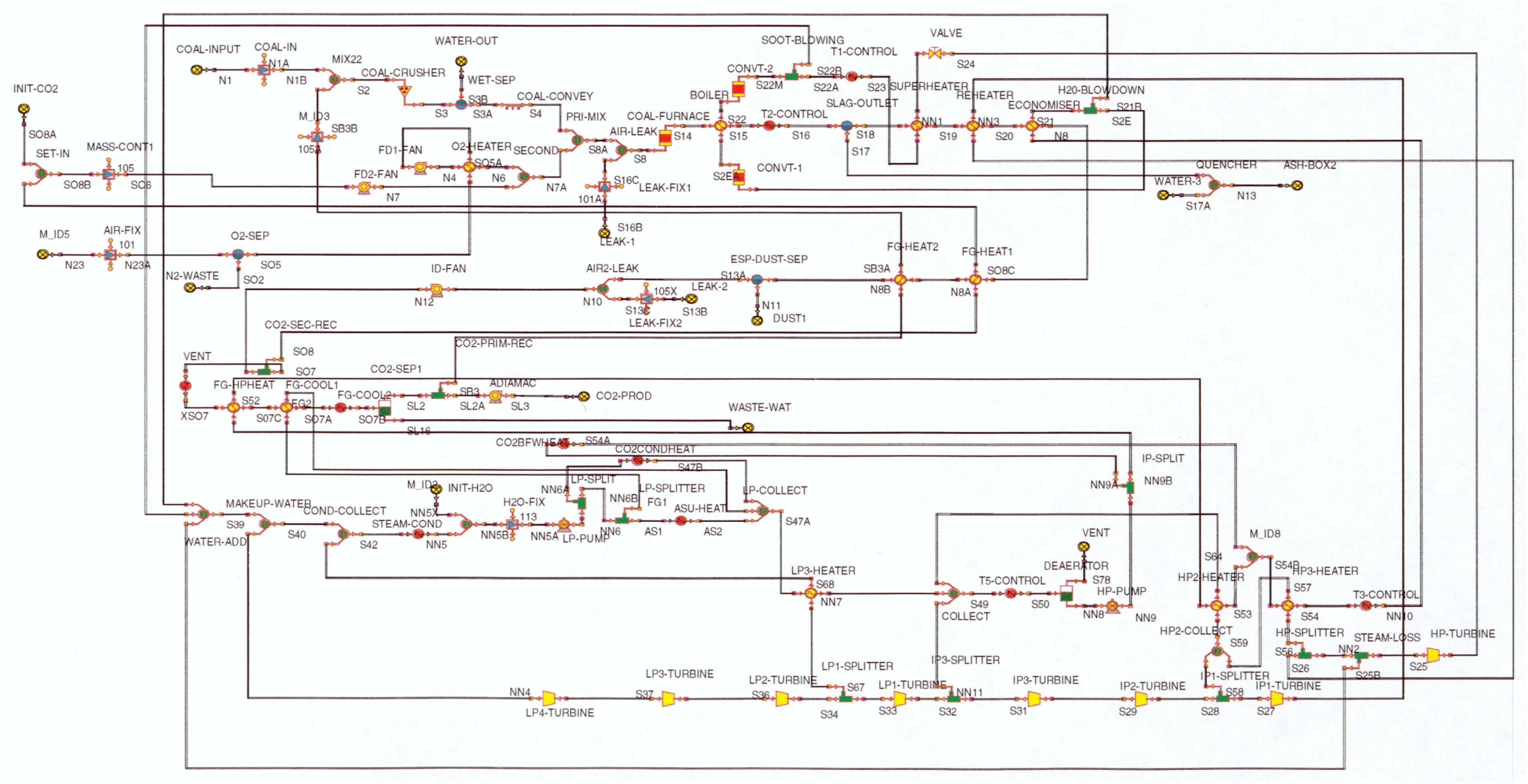
BASIC SCHEME FOR INTEGRATION OF AN ITM OXYGEN AIR SEPARATION UNIT INTO A IGCC POWER CYCLE

FIGURE 30

APPENDIX 1

ECLIPSE Modelling of ASC PF Power Plant with CO_2 Capture

ECLIPSE Model as described in MBEL Document No: 78593/B251/DS/31000/X/0001/A6 dated 27/04/04



Tank/Vessel ID	Pressure (Bar ABS)	Temperature (C)	Flash (%)		
DEAERATOR	6.019	158.988	0		
CO2-SEP1	1.016	35.101	92.704		
Heater or Cooler ID	**Inlet Temperature(C)**	**Outlet Temperature(C)**	**Heat Transfer (MJ/s)**		
T5-CONTROL	158.99	158.99	-0.388		
STEAM-COND	28.97	28.97	-873.261		
T3-CONTROL	270.38	270	-0.942		
T2-CONTROL	1101.69	1102	0.26		
T1-CONTROL	520	520	0		
FG-COOL2	122.97	35	-93.221		
ASU-HEAT	29.1	83.36	55.3		
VENT	263.44	250.3	-5.057		
CO2CONDHEAT	29.1	154.6	50.3		
CO2BFWHEAT	164.99	205.58	16		
Heat Exchanger ID	**Hot Side Inlet TEMP (Deg C)**	**Hot Side Outlet TEMP (Deg C)**	**Cold Side Inlet TEMP (Deg C)**	**Cold Side Outlet TEMP (Deg C)**	**Heat Transfer (MJ/s)**
LP3-HEATER	179.37	111.82	101.82	114.58	21.932
HP2-HEATER	223.98	223.98	180.3	213.98	63.15
HP3-HEATER	359.88	280.38	212.53	270.38	137.512
BOILER	1935.54	1101.69	333.69	520	814.106
SUPERHEATER	1102	919.7	520	600	150.13
REHEATER	919.7	578.79	359.88	620	264.942
ECONOMISER	578.79	340	270	333.69	171.346
O2-HEATER	22.32	22.32	15	15	0
FG-HEAT1	340	311.54	263.44	330	19.413
FG-HEAT2	311.54	264.17	35.1	250	31.747
FG-COOL1	174.99	122.97	29.1	93.2	18.744
FG-HPHEAT	250.3	174.99	164.99	180.3	28.171
Reactor (ID)	**Enthalpy Change**	**Elemental Balance Error**			
COAL-FURNACE	0	0			
CONVT-1	-8181.19	0			
CONVT-2	8222.896	0			

Name	**101**	**101A**	**103**	**103A**	**105**	**105A**	**105X**	**107**	**107**	**107X**
% Solid	0	0	0	0	0	0	0	0	0	0
% Liquid	0	0	0	0	0	0	0	0	0	0
% Gas	0	100	100	0	100	100	0	0	0	100
Pressure (Bar Abs)	0.1	1.013	1.013	0.1	1.016	1.016	0.1	0.1	0.1	1.013
Temperature (Deg C)	0	15	15	0	330	250	0	0	0	15
Enthalpy (MJ/s)	0	-0.242	-6.4	0	-0.251	-0.286	0	0	0	-0.79
Density (kg/m3)	0	1.22	1.22	0	0.74	0.93	0	0	0	1.22
Flow Rate (kg/s)	0	2.504	66.925	0	0.033	0.04	0	0	0	8.272
Compound Flow (kg/s)										
ARGON		0.0319	0.8499		0.0007	0.0009				0.105
ASH										
C										
CO2		0.0011	0.0268		0.0249	0.0318				0.0033
COAL										
H2O										
HCL										
N2		1.8794	50.2403		0.0032	0.0041				6.2101
NO2										
O2		0.5759	15.3927		0.0015	0.0019				1.9025
SO2					0.0002	0.0002				
WATER		0.0156	0.4149		0.0029	0.0008				0.0512
Name	**113**	**115**	**AS1**	**AS2**	**FG1**	**FG2**	**N1**	**N10**	**N11**	**N12**
% Solid	0	0	0	0	0	0	90.5	0.003	100	0.003
% Liquid	0	0	100	100	100	100	9.5	0	0	0
% Gas	100	0	0	0	0	0	0	99.997	0	99.997
Pressure (Bar Abs)	0.04	0.1	16.019	16.019	16.019	12.686	1.013	0.992	0.992	1.016
Temperature (Deg C)	28.97	0	29.1	83.36	29.1	93.2	15	259.37	264.17	263.44
Enthalpy (MJ/s)	0	0	30.061	85.361	8.628	27.372	-349.383	-4607.61	-71.79	-4604.876
Density (kg/m3)	996.02	0	996.69	970.21	996.69	963.49	1365.15	Two Phase	2740.27	Two Phase
Flow Rate (kg/s)	0	0	243.727	243.727	69.95	69.95	100	606.082	5.767	606.082
Compound Flow (kg/s)										
ARGON								13.2377		13.2377
ASH							12.2	0.017	5.6584	0.017
C								0.0003	0.1089	0.0003
CO2								451.6211		451.6211
COAL							78.3			
H2O	0.0002		243.7272	243.7272	69.95	69.95				
HCL								0.0462		0.0462
N2								57.6939		57.6939
NO2								0.1814		0.1814
O2								27.0121		27.0121
SO2								3.214		3.214
WATER							9.5	53.0581		53.0581

Name	**N13**	**N1A**	**N1B**	**N1C**	**N23**	**N23A**	**N4**	**N6**	**N7**	**N7A**
% Solid	75.425	90.5	90.5	0	0	0	0	0	0.003	0.002
% Liquid	0	9.5	9.5	0	0	0	0	0	0	0
% Gas	24.575	0	0	0	100	100	100	100	99.997	99.998
Pressure (Bar Abs)	0.992	1.013	1.013	0.1	1.013	1.013	1.086	1.086	1.041	1.041
Temperature (Deg C)	235.79	15	15	0	15	15	22.32	22.32	333.68	243.39
Enthalpy (MJ/s)	-24.082	-146.426	-202.957	0	-43.549	-49.949	-0.311	-0.311	-1942.768	-1943.079
Density (kg/m3)	Two Phase	1365.15	1365.15	0	1.22	1.22	1.42	1.42	Two Phase	Two Phase
Flow Rate (kg/s)	1.917	41.91	58.09	0	455.375	522.3	126.446	126.446	258.4	384.846
Compound Flow (kg/s)										
ARGON					5.7833	6.6332	3.7677	3.7677	5.6438	9.4115
ASH	1.4188	5.113	7.087						0.0073	0.0073
C	0.0273								0.0001	0.0001
CO2					0.1822	0.209			192.5465	192.5465
COAL		32.8155	45.4845							
H2O										
HCL									0.0197	0.0197
N2					341.85	392.0903	2.5486	2.5486	24.5975	27.1461
NO2									0.0773	0.0773
O2					104.7366	120.1293	120.1293	120.1293	11.5165	131.6458
SO2									1.3703	1.3703
WATER	0.4712	3.9814	5.5186		2.8233	3.2382			22.621	22.621

Name	**N8**	**N8A**	**N8B**	**NN1**	**NN10**	**NN11**	**NN2**	**NN3**	**NN4**	**NN5**
% Solid	0.965	0.965	0.965	0	0	0	0	0	0	0
% Liquid	0	0	0	0	100	0	0	0	10.744	100
% Gas	99.035	99.035	99.035	100	0	100	100	100	89.256	0
Pressure (Bar Abs)	0.992	0.992	0.992	299.99	328.58	6.019	64.5	61.14	0.04	0.04
Temperature (Deg C)	340	311.54	264.17	600	270	282.97	359.88	620	28.97	28.97
Enthalpy (MJ/s)	-4626.786	-4646.297	-4678.204	1778.349	619.781	8.381	1570.747	1521.539	890.486	49.589
Density (kg/m3)	Two Phase	Two Phase	Two Phase	87.47	801.45	2.38	25.1	15.23	Two Phase	996.02
Flow Rate (kg/s)	599.329	599.329	599.329	516.536	524.372	2.77	513.509	410.807	388.488	408.7
Compound Flow (kg/s)										
ARGON	13.0788	13.0788	13.0788							
ASH	5.6754	5.6754	5.6754							
C	0.1093	0.1093	0.1093							
CO2	451.6161	451.6161	451.6161							
COAL										
H2O				516.5356	524.3718	2.77	513.5087	410.8069	388.4878	408.7001
HCL	0.0462	0.0462	0.0462							
N2	48.2948	48.2948	48.2948							
NO2	0.1814	0.1814	0.1814							
O2	24.1326	24.1326	24.1326							
SO2	3.214	3.214	3.214							
WATER	52.9805	52.9805	52.9805							

Name	**NN5A**	**NN5B**	**NN5X**	**NN6**	**NN6A**	**NN6B**	**NN7**	**NN8**	**NN9**	**NN9A**
% Solid	0	0	0	0	0	0	0	0	0	0
% Liquid	100	100	100	100	100	100	100	100	100	100
% Gas	0	0	0	0	0	0	0	0	0	0
Pressure (Bar Abs)	0.04	0.04	0.06	16.019	16.019	16.019	9.356	6.019	338.53	338.53
Temperature (Deg C)	28.97	28.97	35	29.1	29.1	29.1	114.58	158.99	164.99	164.99
Enthalpy (MJ/s)	49.589	49.589	0	50.409	11.72	38.689	196.685	351.886	375.889	65.705
Density (kg/m3)	996.02	996.02	994.08	996.69	996.69	996.69	947.54	908.23	921.67	921.67
Flow Rate (kg/s)	408.7	408.7	0	408.7	95.023	313.677	408.7	524.372	524.372	91.66
Compound Flow (kg/s)										
ARGON										
ASH										
C										
CO2										
COAL										
H2O	408.7	408.7002	0.0001	408.7	95.0227	313.6773	408.7	524.3718	524.3718	91.6602
HCL										
N2										
NO2										
O2										
SO2										
WATER										

Name	**NN9B**	**S07C**	**S13A**	**S13B**	**S13C**	**S14**	**S15**	**S16**	**S16B**	**S16C**
% Solid	0	0.003	0.003	0	0	1.204	1.204	1.204	0	0
% Liquid	100	0	0	0	0	0	0	0	0	0
% Gas	0	99.997	99.997	100	100	98.796	98.796	98.796	100	100
Pressure (Bar Abs)	338.53	1.016	0.992	1.013	1.013	1.013	1.013	1.013	1.013	1.013
Temperature (Deg C)	164.99	174.99	264.17	15	15	1935.54	1101.69	1102	15	15
Enthalpy (MJ/s)	310.184	-2674.727	-4606.414	-0.406	-1.196	-3236.073	-4054.27	-4054.011	-0.849	-0.607
Density (kg/m3)	921.67	Two Phase	Two Phase	1.22	1.22	Two Phase	Two Phase	Two Phase	1.22	1.22
Flow Rate (kg/s)	432.712	347.649	593.562	4.248	12.52	600.775	600.775	600.775	8.784	6.28
Compound Flow (kg/s)										
ARGON		7.5931	13.0788	0.0539	0.1589	13.0788	13.0788	13.0788	0.1118	0.0799
ASH		0.0098	0.017			7.0942	7.0942	7.0942		
C		0.0002	0.0003			0.1366	0.1366	0.1366		
CO2		259.0499	451.6161	0.0017	0.005	451.6161	451.6161	451.6161	0.004	0.0029
COAL										
H2O	432.7116									
HCL		0.0265	0.0462			0.0462	0.0462	0.0462		
N2		33.0932	48.2948	3.189	9.3991	48.2948	48.2948	48.2948	6.593	4.7136
NO2		0.1041	0.1814			0.1814	0.1814	0.1814		
O2		15.4942	24.1326	0.977	2.8795	24.1326	24.1326	24.1326	2.0203	1.4444
SO2		1.8435	3.214			3.214	3.214	3.214		
WATER		30.4341	52.9805	0.0263	0.0775	52.9805	52.9805	52.9805	0.0548	0.0392

Name	**S17**	**S17A**	**S18**	**S19**	**S2**	**S20**	**S21**	**S21B**	**S22**	**S22A**
% Solid	100	0	0.965	0.965	24.976	0.965	0	0	0	0
% Liquid	0	100	0	0	0	0	100	100	0	0
% Gas	0	0	99.035	99.035	75.024	99.035	0	0	100	100
Pressure (Bar Abs)	0.992	1.013	0.992	0.992	1.013	0.992	319.05	319.05	309.52	309.52
Temperature (Deg C)	1102	15	1102	919.7	108.25	578.79	333.69	333.69	520	520
Enthalpy (MJ/s)	-16.589	-7.493	-4037.422	-4188.306	-1303.364	-4454.579	791.127	5.49	-6581.447	1628.22
Density (kg/m3)	2740.27	871.96	Two Phase	Two Phase	Two Phase	Two Phase	688.85	688.85	111.53	111.99
Flow Rate (kg/s)	1.446	0.471	599.329	599.329	210.49	599.329	524.372	3.639	520.733	516.536
Compound Flow (kg/s)										
ARGON			13.0788	13.0788	3.5874	13.0788				
ASH	1.4188		5.6754	5.6754	7.087	5.6754				
C	0.0273		0.1093	0.1093		0.1093				
CO2			451.6161	451.6161	122.0724	451.6161				
COAL					45.4845					
H2O							524.3718	3.6391		516.5356
HCL			0.0462	0.0462	0.0125	0.0462				
N2			48.2948	48.2948	15.6343	48.2948				
NO2			0.1814	0.1814	0.0451	0.1814				
O2			24.1326	24.1326	7.3205	24.1326				
SO2			3.214	3.214	0.8441	3.214				
WATER		0.4712	52.9805	52.9805	8.4023	52.9805			520.7327	
Name	**S22B**	**S22M**	**S23**	**S24**	**S25**	**S25B**	**S26**	**S27**	**S28**	**S29**
% Solid	0	0	0	0	0	0	0	0	0	0
% Liquid	0	0	0	0	0	0	0	0	0	0
% Gas	100	100	100	100	100	100	100	100	100	100
Pressure (Bar Abs)	309.52	309.52	309.52	290	64.5	64.5	64.5	25	25	13.7
Temperature (Deg C)	520	520	520	596.92	359.88	359.88	359.88	474.22	474.22	387.52
Enthalpy (MJ/s)	13.23	1641.45	1628.22	1778.349	1580.005	9.259	1256.597	1398.74	1364.01	1294.353
Density (kg/m3)	111.99	111.99	111.99	84.61	25.1	25.1	25.1	7.41	7.41	4.58
Flow Rate (kg/s)	4.197	520.733	516.536	516.536	516.536	3.027	410.807	410.807	400.607	400.607
Compound Flow (kg/s)										
ARGON										
ASH										
C										
CO2										
COAL										
H2O	4.1971	520.7327	516.5356	516.5356	516.5356	3.0269	410.8069	410.8069	400.6069	400.6069
HCL										
N2										
NO2										
O2										
SO2										
WATER										

Name	**S2E**	**S2EA**	**S3**	**S31**	**S32**	**S33**	**S34**	**S36**	**S37**	**S39**
% Solid	0	0	24.976	0	0	0	0	0	0	0
% Liquid	100	100	0	0	0	0	0	0.084	5.117	13.186
% Gas	0	0	75.024	100	100	100	100	99.916	94.883	86.814
Pressure (Bar Abs)	319.05	319.05	1.013	6.019	6.019	2.3	2.3	0.88	0.247	64.5
Temperature (Deg C)	333.69	333.69	108.25	282.97	282.97	179.37	179.37	96.1	64.71	280.38
Enthalpy (MJ/s)	785.637	-7395.553	-1303.364	1212.105	1203.724	1124.562	1098.135	1036.293	970.023	27.979
Density (kg/m3)	688.85	550.79	Two Phase	2.38	2.38	1.12	1.12	Two Phase	Two Phase	Two Phase
Flow Rate (kg/s)	520.733	520.733	210.49	400.607	397.837	397.837	388.488	388.488	388.488	10.863
Compound Flow (kg/s)										
ARGON			3.5874							
ASH			7.087							
C										
CO2			122.0724							
COAL			45.4845							
H2O	520.7327			400.6069	397.8369	397.8369	388.4878	388.4878	388.4878	10.8632
HCL			0.0125							
N2			15.6343							
NO2			0.0451							
O2			7.3205							
SO2			0.8441							
WATER		520.7327	8.4023							

Name	**S3A**	**S3B**	**S4**	**S40**	**S42**	**S47A**	**S47B**	**S49**	**S50**	**S52**
% Solid	25.076	0	25.076	0	0	0	0	0	0	0
% Liquid	0	0	0	10.427	12.149	100	100	99.965	100	100
% Gas	74.924	100	74.924	89.573	87.851	0	0	0.035	0	0
Pressure (Bar Abs)	1.013	1.013	1.033	0.04	0.04	12.686	16.019	6.019	6.019	335.2
Temperature (Deg C)	108.25	108.25	108.26	28.97	28.97	101.82	154.6	158.99	158.99	180.3
Enthalpy (MJ/s)	-1292.387	-10.98	-1292.387	918.465	922.851	174.753	62.02	352.274	351.886	338.355
Density (kg/m3)	Two Phase	0.58	Two Phase	Two Phase	Two Phase	957.37	913.08	Two Phase	908.23	907.07
Flow Rate (kg/s)	209.65	0.84	209.65	399.351	408.7	408.7	95.023	524.372	524.372	432.712
Compound Flow (kg/s)										
ARGON	3.5874		3.5874							
ASH	7.087		7.087							
C										
CO2	122.0724		122.0724							
COAL	45.4845		45.4845							
H2O				399.3509	408.7001	408.7	95.0227	524.3718	524.3718	432.7116
HCL	0.0125		0.0125							
N2	15.6343		15.6343							
NO2	0.0451		0.0451							
O2	7.3205		7.3205							
SO2	0.8441		0.8441							
WATER	7.5621	0.8402	7.5621							

Name	**S53**	**S54**	**S54A**	**S54B**	**S56**	**S57**	**S58**	**S59**	**S64**	**S67**
% Solid	0	0	0	0	0	0	0	0	0	0
% Liquid	100	100	100	100	0	69.202	0	50.895	81.436	0
% Gas	0	0	0	0	100	30.798	100	49.105	18.564	100
Pressure (Bar Abs)	331.89	328.58	338.53	331.89	64.5	64.5	25	25	25	2.3
Temperature (Deg C)	213.98	270.38	205.58	212.53	359.88	280.38	474.22	223.98	223.98	179.37
Enthalpy (MJ/s)	401.505	620.723	81.705	483.211	314.149	175.946	34.73	210.676	147.208	26.427
Density (kg/m3)	871.89	800.91	881.49	873.5	25.1	Two Phase	7.41	Two Phase	Two Phase	1.12
Flow Rate (kg/s)	432.712	524.372	91.66	524.372	102.702	102.702	10.2	112.902	112.902	9.349
Compound Flow (kg/s)										
ARGON										
ASH										
C										
CO2										
COAL										
H2O	432.7116	524.3718	91.6602	524.3718	102.7017	102.7018	10.2	112.9018	112.9018	9.3492
HCL										
N2										
NO2										
O2										
SO2										
WATER										

Name	**S68**	**S78**	**S8**	**S8A**	**SB3**	**SB3A**	**SB3B**	**SL16**	**SL2**	**SL2A**
% Solid	0	0	8.752	8.844	0	0	0	0.039	0	0
% Liquid	100	0	0	0	0	0	0	99.961	0	0
% Gas	0	0	91.248	91.156	100	100	100	0	100	100
Pressure (Bar Abs)	2.3	0.1	1.013	1.033	1.016	1.016	1.016	1.016	1.016	1.016
Temperature (Deg C)	111.82	0	191.55	193.33	35.1	250	250	34.9	35.1	35.1
Enthalpy (MJ/s)	4.385	0	-3236.073	-3235.466	-1132.441	-1100.694	-1100.407	-392.603	-2394.167	-1261.726
Density (kg/m3)	949.33	0	Two Phase	Two Phase	1.58	0.93	0.93	865.17	1.58	1.58
Flow Rate (kg/s)	9.349	0	600.775	594.495	152.44	152.44	152.4	25.366	322.283	169.843
Compound Flow (kg/s)										
ARGON			13.0788	12.9989	3.5883	3.5883	3.5874	0.0068	7.5863	3.998
ASH			7.0942	7.0942				0.0098		
C			0.0001	0.0001				0.0002		
CO2			314.6217	314.6189	122.1042	122.1042	122.0724	0.9015	258.1484	136.0442
COAL			45.4845	45.4845						
H2O	9.3492									
HCL			0.0322	0.0322	0.0125	0.0125	0.0125	0.0001	0.0264	0.0139
N2			47.4941	42.7804	15.6384	15.6384	15.6343	0.031	33.0622	17.4238
NO2			0.1225	0.1225	0.0451	0.0451	0.0451	0.0086	0.0954	0.0503
O2			140.4106	138.9662	7.3224	7.3224	7.3205	0.0135	15.4807	8.1583
SO2			2.2143	2.2143	0.8443	0.8443	0.8441	0.0586	1.785	0.9407
WATER			30.2223	30.1831	2.8845	2.8845	2.8837	24.3358	6.0983	3.2138

Name	**SL3**	**SO2**	**SO5**	**SO5A**	**SO6**	**SO7**	**SO7A**	**SO7B**	**SO8**	**SO8A**
% Solid	0	0	0	0	0.003	0.003	0.003	0.003	0.003	0
% Liquid	0	0	0	0	0	0	0	7.294	0	0
% Gas	100	100	100	100	99.997	99.997	99.997	92.704	99.997	100
Pressure (Bar Abs)	15	1.013	1.013	1.013	1.016	1.016	1.016	1.016	1.016	2
Temperature (Deg C)	334.18	15	15	15	330	263.44	122.97	35	263.44	20
Enthalpy (MJ/s)	-1211.578	-48.799	-1.15	-1.15	-1943.855	-2641.357	-2693.565	-2786.786	-1963.519	-0.001
Density (kg/m3)	11.84	1.18	1.36	1.36	Two Phase	Two Phase	Two Phase	Two Phase	Two Phase	3.65
Flow Rate (kg/s)	169.843	395.854	126.446	126.446	258.4	347.649	347.649	347.649	258.433	0
Compound Flow (kg/s)										
ARGON	3.998	2.8656	3.7677	3.7677	5.6438	7.5931	7.5931	7.5931	5.6445	
ASH					0.0073	0.0098	0.0098	0.0098	0.0073	
C					0.0001	0.0002	0.0002	0.0002	0.0001	
CO2	136.0442	0.209			192.5465	259.0499	259.0499	259.0499	192.5713	0.0001
COAL										
H2O										
HCL	0.0139				0.0197	0.0265	0.0265	0.0265	0.0197	
N2	17.4238	389.5417	2.5486	2.5486	24.5975	33.0932	33.0932	33.0932	24.6007	
NO2	0.0503				0.0773	0.1041	0.1041	0.1041	0.0774	
O2	8.1583		120.1293	120.1293	11.5165	15.4942	15.4942	15.4942	11.518	
SO2	0.9407				1.3703	1.8435	1.8435	1.8435	1.3704	
WATER	3.2138	3.2382			22.621	30.4341	30.4341	30.4341	22.624	

Name	**SO8B**	**SO8C**	**XSO7**
% Solid	0.003	0.003	0.003
% Liquid	0	0	0
% Gas	99.997	99.997	99.997
Pressure (Bar Abs)	1.016	1.016	1.016
Temperature (Deg C)	330	330	250.3
Enthalpy (MJ/s)	-1944.107	-1944.106	-2646.414
Density (kg/m3)	Two Phase	Two Phase	Two Phase
Flow Rate (kg/s)	258.433	258.433	347.649
Compound Flow (kg/s)			
ARGON	5.6445	5.6445	7.5931
ASH	0.0073	0.0073	0.0098
C	0.0001	0.0001	0.0002
CO2	192.5714	192.5713	259.0499
COAL			
H2O			
HCL	0.0197	0.0197	0.0265
N2	24.6007	24.6007	33.0932
NO2	0.0774	0.0774	0.1041
O2	11.518	11.518	15.4942
SO2	1.3704	1.3704	1.8435
WATER	22.624	22.624	30.4341