



Programme Area: Bioenergy

Project: Biomass to Power with CCS

Title: WP2 High-level Engineering Study: Report on Selected Technology Combinations

Abstract:

This report, the 2nd deliverable from the project, is a high level study of the engineering associated with implementing each of the 8 technologies identified in the Technology Landscape report (1st deliverable: WP1). It includes, for each technology, an overview evaluation of system performance. It provides a critical assessment of knowledge gaps, technical risks and subject areas that will require further development work. It also includes an overview of the total process for the combined technologies, preliminary process flow diagrams, high level process control philosophy, environmental performance summary, estimate of the project development and capital costs and an estimate of operating and maintenance costs

Context:

The Biomass to Power with CCS Phase 1 project consisted of four work packages: WP1: Landscape review of current developments; WP2: High Level Engineering Study (down-selecting from 24 to 8 Biomass to Power with CCS technologies); WP3: Parameterised Sub-System Models development; and WP4: Technology benchmarking and recommendation report. Reports generally follow this coding. We would suggest that you do not read any of the earlier deliverables in isolation as some assumptions in the reports were shown to be invalid. We would recommend that you read the project executive summaries as they provide a good summary of the overall conclusions. This work demonstrated the potential value of Biomass to Power with CCS technologies as a family, but it was clear at the time of the project, that the individual technologies were insufficiently mature to be able to 'pick a winner', due to the uncertainties around cost and performance associated with lower Technology Readiness Levels (TRLs).

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Biomass to Power with CCS Project

TESBiC: Techno-Economic Study of Biomass to power with CCS

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High-level Engineering Study:

Report on Selected Technology Combinations

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Authors	The TESBiC consortium:



cmcl innovations



Doosan Babcock



Drax



EDF



E4tech



Imperial College London



University of Cambridge



University of Leeds

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EXECUTIVE SUMMARY

The **Techno-Economic Study of Biomass to Power with CCS (TESBiC)** project is aimed at performing a techno-economic assessment of the current and potential future approaches to the combination of technologies which involve the generation of electricity from biomass materials, and those which involve carbon capture and storage (CCS).

This current report is concerned with the results of the work done under Work Package, WP2, on the key technology combinations identified under Work Package, WP1 for further study. For each of the selected technologies, the following items of engineering and other work have been prepared, viz:

- An overview of the total process for the combined technologies and the relevant engineering standards,
- A preliminary process flow diagram with energy, material and carbon balances,
- A list of the major items of equipment with short performance specifications covering the key process factors,
- A high level process control philosophy,
- An environmental performance summary, including the process inputs and estimates of the emissions to air, land and water, including the major prescribed pollutant species and carbon in all forms,
- An estimate of the project development and capital costs for new build and retrofit projects, as appropriate,
- An estimate of the fixed and variable operating and maintenance costs,
- An overview evaluation of the system performance and a critical assessment of the knowledge gaps, technical risks and subject areas that will require further development work.

The selected technologies and their component parts spanned a very wide range of technology readiness levels (TRLs), from as low as TRL 3, i.e. characteristic proof of concept up to 8, i.e. fully operational system with integrated hardware and software systems. For this reason, it proved to be challenging, and in some cases just not possible to obtain sufficiently detailed information on the key criteria across all the technologies considered. In WP1, a total of twenty eight technology combinations were assessed. These technology combinations are displayed in Table 0.1 below. One of the outcomes of WP1 was the identification of a shortlist of eight technology combinations which were selected by the project team, following consultation with ETI for further study under WP2. The following technology combinations were selected for further study:

- The co-firing of biomass materials in a pulverised coal-fired power plant, with post-combustion carbon dioxide capture by amine scrubbing,
- The combustion of biomass materials in a dedicated power plant with post-combustion carbon dioxide capture by amine scrubbing,
- The co-firing of biomass materials in a pulverised coal power plant, with carbon dioxide capture by using oxyfuel firing technology, with cryogenic O₂ separation,

- The combustion of biomass materials in a dedicated power plant with carbon dioxide capture by oxyfuel firing technology, with cryogenic O₂ separation,
- Biomass co-firing combustion, with CO₂ capture by post-combustion carbonate looping,
- Dedicated biomass chemical-looping-combustion using solid oxygen carriers,
- Co-firing Integrated gasification combined cycle (IGCC), with CO₂ capture by physical absorption, and
- Dedicated biomass, i.e. BIGCC (for small scale, i.e. <50 MWe), with CO₂ capture by physical absorption.

Thus, the key carbon capture technologies from all three categories (pre-combustion, post-combustion and oxy-fuel) are covered in WP2. An overview of the key features of the selected technology combinations are presented in Table 0.1 below. In each case, the basic technology and the current status of development have been described in the report. Outline process flow diagrams with listings of the key flows and process conditions have been presented, and calculations of the overall process efficiencies (LHV basis) have been prepared. High level assessments of the process control and environmental issues have also been performed. Preliminary study estimates of the capital and operations and maintenance (O&M) costs have also been prepared and presented in the report. The level of detail varies, depending largely on the status of development of the technologies involved. In summary, a comparison of the key technical and economic features of the technology combinations, highlighting the current status of development and the barriers to further implementation, and the potential technical advantages and disadvantages of the technologies are also presented. A direct comparison of the estimated capital and operating costs of the technology combinations, as far as is possible with the economic data available for the less well developed technologies, is also presented.

Overall, the great majority of the industrial expenditure to date on the development of carbon dioxide capture technology has been, and continues to be, on more conventional approaches involving large coal-fired power plants, i.e.

- The application of CO₂ capture by solvent scrubbing technologies to the flue gas streams from large pulverised coal-fired boilers and fluidised bed boilers firing coal, and the compression and drying of the CO₂ prior to transportation offsite,
- The application of oxyfuel firing technologies to large coal-fired boilers, and the purification, compression and drying of the product CO₂ prior to transportation off site,
- The application of physical CO₂ absorption techniques to the syngas produced by large coal-fired IGCC power plants.

To date, the development of all three of these technologies has been pursued in parallel, with each currently at the pilot- and small demonstration scale. It has, however, proven to be difficult for the electricity supply industry and their equipment suppliers to come to a clear decision as to which technology is to be preferred for future application. These developments are now moving to the larger scale demonstration projects but, because these projects are relatively expensive and require significant government subsidies, the progression to the demonstration phase has been relatively slow.

It is also clear that there has been relatively little activity on the application of the carbon capture technologies to dedicated biomass power plants or to coal power plants which are co-firing biomass. For most applications, the biomass fuel is significantly more expensive than coal, generally by a factor of more than two on a £/GJ basis, and the application of CO₂ capture and storage technology is also relatively expensive.

At the present time there is no commercial incentive to apply CO₂ capture technologies to dedicated biomass power plants, and specific incentives, associated with the removal of CO₂ from the atmosphere will need to be introduced to support these activities. This is not as yet being discussed seriously at government level.

It should also be noted that the power industry is historically very conservative in character, and the current electricity generation, distribution and trading systems have become accustomed over many decades to the degree of flexibility of operation provided by the pulverised coal boiler. It may also be the case that in the future, when a higher portion of the UK's power generation will be from intermittent sources, the flexibility of operation provided by thermal power plants will be of even more importance.

The technical options for the capture of CO₂ from thermal power plants can be described in terms of the following three categories, viz:

Post-combustion technologies involve the addition of a chemical process plant to the flue gas outlets, downstream of the current environmental control equipment. These technologies are the most familiar to the electricity supply industry, since they have been applying the liquid or slurry scrubbing of flue gases for the control of gaseous emissions for several decades. In general terms, these will not have a significant impact on the flexibility of operation of the power plant, except in cases where there is a requirement for an oxygen supply, e.g. for the carbonate looping system, where there is oxygen blown combustion of a secondary fuel in the calciner. The addition of an air separation unit (ASU) may have a significant impact on the flexibility of operation of the overall plant. This can be reduced by the provision of oxygen storage facilities on site, although is clearly a process and site specific issue.

Oxyfuel technologies conventionally involve a fairly major reconfiguration of the fuel firing systems, the draft plant and the boiler, and the installation of an air separation unit for oxygen production. In the case of chemical looping combustion technologies, the reconfiguration of the firing and heat transfer systems is fairly radical. In this case again, the ASU will have a significant impact on the flexibility of operation of the plant, although the impact of this can be reduced by the provision of oxygen storage facilities on site.

Pre-combustion technologies involve the gasification of the solid fuel followed by CO₂ capture by physical absorption, with combustion of the hydrogen-rich product gas in a gas turbine in a combined cycle configuration. The co-firing of biomass in an IGCC system is relatively well understood technology and the capture of CO₂ from the syngas is fairly conventional. The dedicated biomass IGCC system on the other hand is more complex and less well understood, and the uncertainties become particularly larger due to lack of experience at smaller scales (~30-50 MW_e).

Table 0.1: An overview of the eight selected technology combinations, studied in WP2.

Criteria	Co-firing amine scrubbing	Dedicated biomass with amine scrubbing	Co-firing oxy-fuel	Dedicated biomass oxy-fuel	Co-firing carbonate looping	Dedicated biomass chemical looping	Co-firing IGCC	Dedicated biomass BIGCC
Current TRL	6 to 7	4	6	5	4 to 5	4	5 to 6	4
Key technical issues	Scale-up, amine degradation,	Scale-up, amine degradation,	Corrosion, O ₂ energy costs, slow response	Corrosion, O ₂ energy costs, slow response	Calcliner firing, solid degradation, large purge of CaO	Loss in activity, reaction rates, dual bed operation	Complex operation, slow response, tar cleaning, retrofit impractical	Complex operation, slow response, tar cleaning, retrofit impractical
Suitability for small scale	Low	High	Low	High	Low	High	Low	High
Plant efficiency with capture	OK	Low	OK	Low	Good	Good	High,	Good
Capital costs with capture	OK	Expensive	OK	High ASU costs	OK	Low cost	OK	Expensive,
UK deployment potential	Immediate capture retrofit opportunities,	retrofit opportunities high long-term potential	retrofit opportunities, long-term doubtful	retrofit opportunities, high long-term potential	capture retrofit opportunities, cement integration	Likely first demos in Europe, UK in ~2020. High long term potential	No current UK plants, several demos by 2020 Long-term doubt	No current UK plants, demo unlikely by 2020. High long-term potential

During the performance of the high-level engineering evaluation of the eight technology combinations, one of the key challenges encountered was associated with the wide variation in the scales of operation and technology readiness levels of the technology combinations studied. These varied from TRL 3-4, i.e. only proof of concept, to values as high as TRL 7-8, i.e. commercial plant in operation. One of the consequences of the variability in the TRL values was that the quality of the technical and economic data available for the technology combinations varied accordingly.

In general terms, the quality of the technical and economic data for the technology combinations 1-4 and 7, i.e. the technology combinations with relatively high TRL values was in general very good. These data were largely from published literature, supplemented with in-house information and using validated open source simulation tools. Overall, the gaps in knowledge of the more advanced technologies are modest and these will be filled by proceeding to the large scale demonstration stage, with the associated engineering and R&D activities required to address any emergent technical difficulties. This will be done in the first instance for fossil fuel-fired plants, possibly with biomass co-firing. It is likely that the application of the carbon dioxide capture technologies to dedicated biomass plants can then be achieved without a further specific demonstration phase on a biomass plant. It is anticipated that these technologies will be ready for market in the second half of the current decade.

The equivalent technical and economic data for the technology combinations 5, 6 and 8 are not available in anything like a comparable level of detail, and the process flow sheets and other data have had to be built up using commercial or bespoke academic software. Similarly, the economic data are based on very limited engineering knowledge or experience of the proposed plant configuration, and hence are subject to relatively large bounds of uncertainty.

The gaps in the technical and economic data on the less well developed technology combinations are substantial, and to fill these gaps will require significant further work at laboratory and pilot scale, extensive programmes of technical work at large pilot scale, significant engineering work and finally component testing and demonstration projects to provide information similar to that available for technology combinations 1-4. In some cases, there are ongoing collaborative projects in these subject areas with financial support from the British government and the European Commission.

The following approach was adopted to help take account of this issue:

- As far as possible common approaches were adopted for both the technical and economic assessment work,
- Where possible, the technical performance and economic parameters for items of equipment, e.g. steam raising, ASU, fuel reception storage and handling etc. which are common to a number of the technology combinations have been harmonised, and
- The basic assumptions about some items of the capital costs (operation and utilities, civils and land costs, project development costs, etc.), of the fixed plant operating costs (maintenance and labour, insurance, etc.), of the feedstock prices, and of the plant capacity factors, etc. were harmonised as far as was possible. These were employed in calculating the

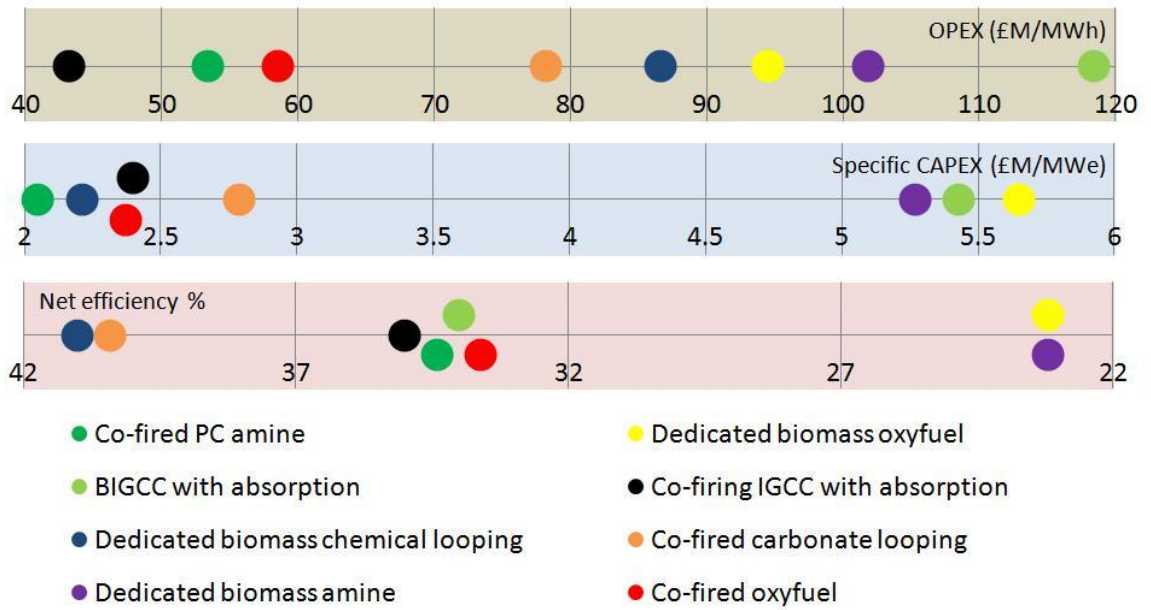
overall capital expenditure (CAPEX) and operating and maintenance costs (OPEX) figures for the eight technology combinations.

Table 0.2a: Summary of economic parameters for technology combinations 1 to 4.

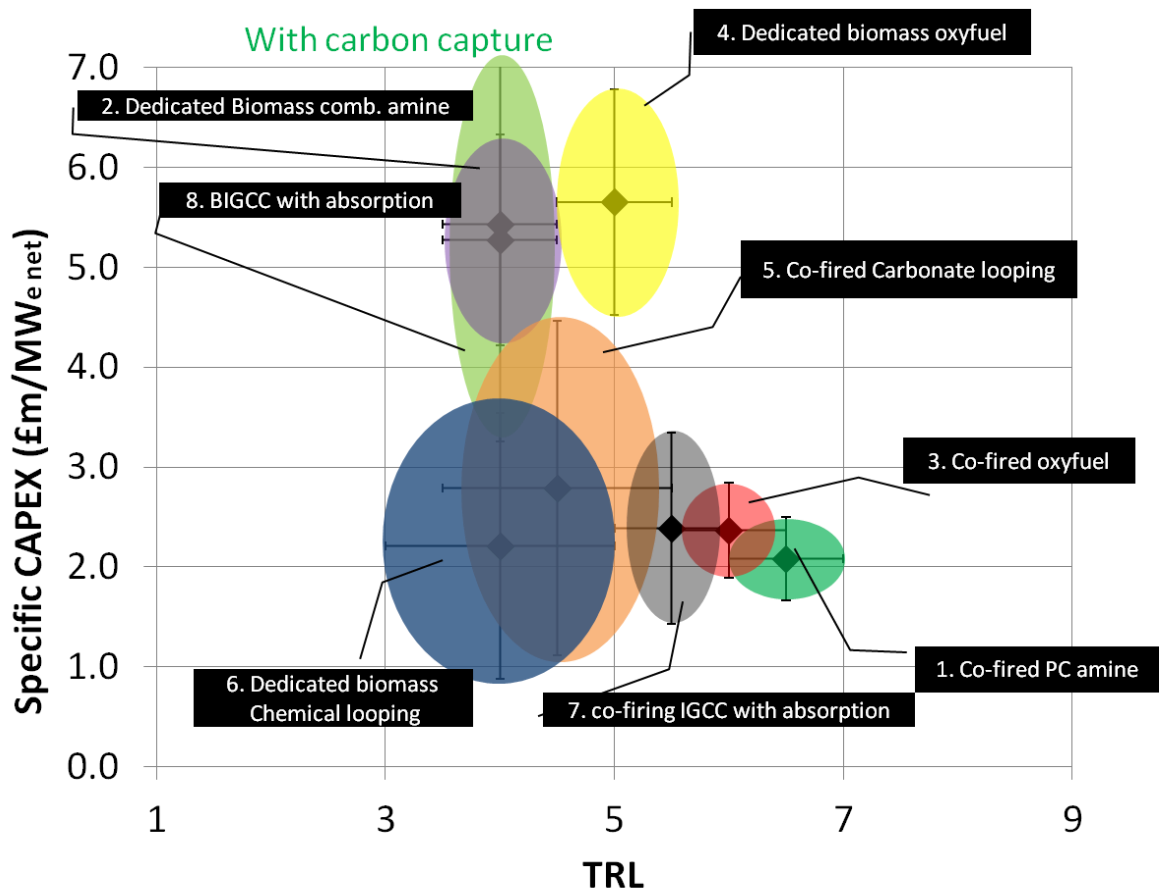
	1 Biomass co-firing in pulverised coal boiler		2 Dedicated biomass combustion		3. Biomass co-firing in pulverised coal boiler		4 Dedicated biomass combustion	
	Without CCS	with solvent scrubbing	Without CCS	With solvent scrubbing	Without CCS	With oxyfuel firing	Without CCS	With oxyfuel firing
Gross power (MW _e)	545.2	474.1	83.5	68.5	545.2	545.2	83.4	84.6
Net power (MW _e)	518.9	398.9	75.8	48.9	518.9	388.7	75.79	48.85
Gross efficiency %	47.1	40.9	39.6	32.5	47.1	47.1	39.5	40.1
Net efficiency %	44.8	34.4	35.9	23.2	44.8	33.6	35.9	23.2
Total installed cost (£M)	509.3	638.2	143.5	198.1	509.3	709.1	143.5	213.2
Total investment cost (£M)	662	830	187	258	662	922	187	277
Specific CAPEX (£M/MW _e)	1.28	2.08	2.46	5.27	1.28	2.37	2.46	5.65
Total O&M costs (£M/yr)	146	158.5	33.3	37.1	175	173.4	33.3	34.3
Energy generated (MWh/yr) x 10 ⁶	3.86	2.97	0.56	0.36	3.86	2.89	0.56	0.36
OPEX & fuel (£/MWh)	37.78	53.36	59	101.9	45.64	58.52	59	94.5

Table 0.2b: Summary of economic parameters for technologies 5 to 8.

	5 Carbonate looping		6 Chemical looping		7. IGCC co-firing with CO ₂ absorption		8 BIGCC	
	Without CCS	with carbonate looping	Without CCS	With chemical looping	Without CCS	With CO ₂ capture	Without CCS	with CCS
Gross power (MW _e)	N.A.	326	N.A.	325	549	549	64.7	73
Net power (MW _e)		300		300	484	461	47.5	40
Gross efficiency %		43.9		44.5	45.8	41.7	56.0	56.0
Net efficiency %		40.4		41.0	40.4	35.0	41.0	34.0
Total installed cost (£M)		644		509	650	848	109	167
Total investment cost (£M)		837		662	845	1102	142	217
Specific CAPEX (£M/MW _e)		2.79		2.21	1.74	2.40	2.98	5.43
Total O&M costs (£M/yr)		174.70		192.36	128.47	148.45	22.55	35.25
Energy generated (MWh/yr) x 10 ⁶		2.23		2.23	3.61	3.43	0.35	0.29
OPEX & fuel (£/MWh)		78.20		86.60	35.60	43.20	63.80	118.40



(a): Main performance and economic parameters



(b): Capital costs vs. technical risk

Figure 0.1: Accounting for uncertainties in specific capital costs (£M/MWe net) and technical risk represented in terms of technology readiness level (TRL) for all 8 technology combinations (with CCS).

Furthermore, it should be also noted that an additional challenge encountered was the lack of a single common engineering software platform that can evaluate all eight biomass CCS technology combinations. Consequently, four software platforms ranging from commercial flow-sheeting software, validated open-source power plant and carbon capture simulation tools to in-house bespoke academic computer programmes were used to assess the eight technologies in detail. The harmonization explained above was beneficial in systematically specifying the boundary conditions for the techno-economic simulations.

The key performance and economic parameters for the eight technology combinations are listed in Table 0.2a and Table 0.2b. All the eight combinations are also represented in terms of their three main measures of performance and costing, i.e. overall net efficiencies, CAPEX and OPEX in Figure 0.1a. Looking in the first instance, at the gross and net electricity generation efficiency values obtained for all eight technology combinations, the negative impacts of the carbon dioxide capture technologies was evident. In the cases studied, the operation of the CO₂ capture and compression/cleaning systems wherever directly comparable, decreased the net electrical efficiency of the power plant in the range of 6 to 13% points. This also had the effect of increasing the specific investment costs significantly. In cases where direct comparison was possible, the increase in the specific CAPEX was in the range of 45 to 130%, as would be expected. The impact of the operation of the CO₂ capture plant on the total annual O&M costs of the plants was more modest, i.e. in the range, 4% to 58%. The specific O&M cost per MWh of power was clearly much higher for the cases with CO₂ capture, because of the significant reduction in the net power output of the plant.

The data presented here also illustrates the difficulties the industry faces in identification of the preferred technology between solvent scrubbing and oxyfuel firing. The estimated capital costs of these technologies and the impacts on the net plant efficiencies are generally of the same order. The oxyfuel technology involves the installation and operation of an ASU, with the associated costs penalties and impacts on power plant flexibility of operation. The solvent scrubbing process does not require an ASU but has a similar impact on the net power plant efficiency, and there are additional environmental and other issues associated with the on-site storage handling and use of large quantities of organic chemicals.

For the carbonate looping, chemical looping combustion and the BIGCC systems, the quoted net electrical efficiency of the systems were of the order of 40-41%. These values are significantly higher than those for the more conventional combustion-based solvent scrubbing and oxyfuel firing systems studied here.

The specific OPEX + fuel cost estimates for the technology combinations indicated that the specific operating costs for the large coal power plants co-firing biomass, and IGCC plants were significantly lower than those for the smaller dedicated biomass combustion and biomass gasification power plants. The differences between the specific fuel + OPEX costs for the solvent scrubbing and the oxyfuel firing processes were fairly modest.

As stated above, the study involved technology combinations (and the associated individual components) spanning a wide range of technology readiness levels, from as low as TRL 3 to 4, up to 7 to 8, it was deemed necessary to account for the uncertainties associated with most of the process and economic factors. For instance, the estimation of the overall electrical generation efficiency

values for the technologies is also clearly affected by the TRL. For Case studies 1-4 and 7, where good quality data for current technology at state of the art steam conditions, are available, these have been employed in the assessment. For Case Studies 5,6 and 8 where good quality data are not available and the level of engineering detail of the configuration of the process plant and heat exchangers are not available, significant speculation about future operating conditions was required.

The estimated capital costs of the eight technology combinations are plotted against the TRLs in Figure 0.1b. It is evident that the uncertainties associated with economic parameters for the better-known and understood technology combinations, for example, co-firing and dedicated biomass combustion with amine scrubbing and oxyfuel are relatively low, i.e. of the order of 20%, whereas for relatively new technologies such as dedicated biomass combustion and gasification as well as chemical looping (TRL around 4), the uncertainties were considered to be in the range of 35 to 60%.

The technical and the economic data obtained as a result of WP1 and WP2 will be utilised in WP3 for development and parameter estimation of the sub-system models for all the eight technology combinations.

The following **short summary statements** on the activities and findings of **WP2** of the TESBiC project can be made:

- To date, there has been relatively little activity at industrial scale on the application of the carbon capture technologies to dedicated biomass power plants or to coal power plants which are co-firing biomass.
- The industry is now moving to large scale demonstration projects on fossil-fuel based power technologies but because these are relatively expensive and require significant government subsidies, progression to the demonstration phase is relatively slow. The alternative technologies with lower TRL values continue to be developed at laboratory and, in some cases, at pilot plant scales.
- The alternative technologies with lower TRL values continue to be developed at laboratory and, in some cases, at pilot plant scales.
- The TRLs for the eight technology combinations considered here, varied over a wide range from TRL 3 to TRL 8.
- Within the TESBiC project, a number of software platforms were employed, as appropriate, for process assessment, and a harmonised set of boundary conditions and assumptions were implemented across the eight technology combinations, in order to facilitate the comparison of the relevant techno-economic parameters.
- The process engineering calculations for mass and energy flows, equipment lists, high level control philosophy, environmental performance, basic process economics and existing gaps in understanding as well as future development requirements were evaluated for the eight technology combinations
- Wherever direct comparison was feasible, it was observed that the net efficiency decrease due to carbon capture varied in the range of 6 to 13% points, the specific investment costs

increased significantly in the range 45 to 130% and the annualised operating and maintenance costs were between 4% and 58%.

- The WP2 data collected for the seven output variables such as capital cost, non-fuel operating costs, generation efficiencies, CO₂ emissions, SO₂ emissions and NO_x emissions as functions of the four input parameters, namely, co-firing levels, extent of carbon capture, nameplate and operating capacities are being utilised for sub-model parameterisation in Work Package 3. This sensitivity data matrix is presented in the WP3 deliverables.
- The data and the associated uncertainties will be directly utilised to carry out benefits assessments of specific development, demonstration and deployment activities in Work Package 4.

INTRODUCTION

The **Techno-Economic Study of Biomass to Power with CCS (TESBiC)** project, which has been commissioned by ETI, is concerned with the performance of a techno-economic assessment of the current and potential future approaches to the combination of technologies which involve the generation of electricity from biomass materials, and those which involve carbon dioxide capture from the process exhaust gases. The work on this project has been performed by a consortium of commercial and academic organisations, as follows:

- Computational Modelling Cambridge Ltd, who has acted as Project Lead,
- Doosan Babcock, who has acted as Technical Lead,
- Drax Power Ltd,
- EDF,
- E4tech,
- Imperial College, London University, involving three research groups, i.e. the Centre for Process Systems Engineering, the Centre for Environmental Policy, and the Centre for Energy Policy and Technology,
- The University of Cambridge, involving two research groups, i.e. the Combustion and Gasification Research Group and the Computational Modelling Group,
- The University of Leeds, involving two research groups, i.e. the Centre for Computational Fluid Dynamics and the Energy and Resources Research Institute.

The principal objectives of the proposed project can be described as follows:

- To recommend a number of specific Case Studies of the relevant technology combinations, based upon the results of a state-of-the-art technical, economic and environmental review of the current and short-medium term future technologies for power generation from biomass and for the capture of carbon dioxide from thermal power plants,
- To perform the Case Studies, covering the relevant technical, economic and environmental issues, and employing an appropriate and robust methodology for the standardisation of the technical and economic data and for parameter estimation,
- To develop the appropriate sub-system level models with validation against the process data and other information from the Case Studies,
- To integrate the process models to facilitate future coupling with ETI's Biomass Value Chain Model and the overall UK Energy Systems Model, and

- To assess the potential benefits from further development of the relevant technologies and to prioritise them, based on their TRLs and their impact on CO_{2e} reduction, costs and energy efficiency.

The project has four technical work packages, and a fifth work package covering the project management activities. The technical work packages can be described as follows:

WP1: Landscape overview of current developments

Under this Work Package, a high level techno-economic survey of the key biomass utilisation and carbon capture technologies, which are currently available or are considered likely to become available during the relevant timeframe, i.e. over the period up to 2050, has been carried out. The reports and other deliverable items from WP1 have been completed and issued.

WP2: High Level Engineering Study

The performance of a high level engineering study of a number of Case Studies of the application of carbon capture and sequestration technologies to biomass energy conversion systems is proposed under this Work Package.

It is proposed that the Case Studies will cover both the key commercially available or near market technologies, and the more innovative technologies that are under development at laboratory and pilot scale at the present time. It is also proposed that at least one of the technology combinations selected for the Case Study work should be suitable for small scale applications.

WP3: Model Development

The work proposed in WP3 will utilise the relevant information from WP2, as well as the data and models that are available in the technical literature and the relevant knowledge, experience and data available from the project partners. The results of this work will be applied to the development of a consistent set of models of the relevant technology combinations, appropriate for integration with the ETI bioenergy value chain model and the ETI energy system model.

WP4: Technology Benchmarking and Recommendation Report

The technology landscape assessment and modelling work will identify the more promising technologies, based on a number of biomass resource-related, technical, economic and environmental criteria. Under this work package, it is proposed that the project partners will:

- Assess the viability of the relevant combinations of biomass to power technologies, both dedicated and co-firing with fossil fuel, and carbon dioxide capture options, in comparison to the other available biomass utilisation options and non-biomass carbon dioxide capture options.

- Select the most appropriate technologies for further development and deployment, based on the outputs of this analysis work, and
- Further assess the opportunities for the application of the selected technologies, based on the technology readiness level, the technology development needs, the commercial attractiveness of the end use application, the economic viability, the specific GHG savings, and the other key sustainability considerations.

Scope and Context

This report is concerned with the results of the work done under Work Package, WP2, i.e. the performance of the Case Studies of the key technology combinations which were identified under WP 1 for further study. For each of the selected technologies, the following items of engineering and other work have been prepared, viz:

- A technical overview of the overall biomass-CCS process and the relevant engineering standards,
- A preliminary process flow diagram with energy, material and carbon balances,
- A list of the major items of equipment with performance specifications covering the key process factors,
- A high level process control philosophy,
- An environmental performance summary, including the process inputs and estimates of the emissions to air, land and water, including the major prescribed pollutant species and carbon in all forms,
- A budget estimate of the project development and capital costs for new build projects,
- Budget estimates of the major fixed and variable operating and maintenance costs,
- An overview evaluation of the system performance and a critical assessment of the knowledge gaps, technical risks and subject areas that will require further development work.

It was stated in the original proposal that budget estimates of the project development and capital costs will be prepared for both new build and retrofit projects. It has become apparent during the project that the estimation of the costs of retrofit projects is not relevant to a number of the technology combinations and is very difficult to achieve, and that the results are so site and project-specific that comparisons between the different technology combinations are largely meaningless. For these reasons, estimates of the capital and development costs of retrofit projects are not presented in this report.

Under Work Package 1 a total of twenty eight technology combinations were assessed. These are listed in Table 1. One of the outcomes of WP 1 was the identification of a shortlist of eight technology combinations which were selected by the project team, after consultation with ETI, for further study under WP2.

These eight technology combinations are the following:

- Case Study 1: [Cell#1 in Table 1] - The co-firing of biomass materials in a pulverised coal-fired power plant, with post-combustion carbon dioxide capture by amine scrubbing,
- Case Study 2: [Cell#2 in Table 1] - The combustion of biomass materials in a dedicated power plant with post-combustion carbon dioxide capture by amine scrubbing,
- Case Study 3: [Cell#11 in Table 1] - The co-firing of biomass materials in a pulverised coal power plant, with carbon dioxide capture by using oxyfuel firing technology, with cryogenic oxygen separation,
- Case Study 4: [Cell#12 in Table 1] - The combustion of biomass materials in a dedicated coal-fired power plant with carbon dioxide capture by oxyfuel firing technology, with cryogenic oxygen separation,
- Case Study 5: [Cell#9 in Table 1] - The co-firing of biomass in a coal-fired boiler, with carbon dioxide capture by post-combustion carbonate looping,
- Case Study 6: [Cell#14 in Table 1] - Dedicated biomass chemical-looping-combustion using a solid oxygen carrier,
- Case Study 7: [Cell#15 in Table 1] - The co-firing of biomass in a large coal-fired Integrated Gasification Combined Cycle plant, with carbon dioxide capture by physical absorption, and
- Case Study 8: [Cell#16 in Table 1] - A dedicated biomass IGCC, with carbon dioxide capture by physical absorption.

An overall view of the technology combinations selected for progression to Work Package 2 is presented in Table 2. Within this list of selected technology combinations, it is considered that there is a reasonable balance between the co-firing and dedicated biomass power plant options, i.e. four examples of each of these, and between the major categories of CO₂ capture technology, i.e.

- pre-combustion capture with three Case Studies,
- post-combustion capture also with three Case Studies and
- oxyfuel combustion with two Case Studies.

For these Case Studies, a set of suitable standard coal and biomass analyses have been identified to provide a common basis for the process calculations. Four fuels have been selected, viz:

- A British bituminous coal,
- A world traded bituminous coal,
- A world traded wood pellet, and
- A wet wood chip fuel.

The key analysis data comprising moisture content, ash analysis, elemental analysis, etc. for the four fuels is listed in Table 3.

The work in this project has been focussed on the utilisation of biomass materials that are currently available in industrial quantities, and whose properties are relatively well understood. It is recognised that a number of processed biomass fuels are beginning to come on to the market, and may become available in significant quantities over the next few years. In general, these are biomass materials that have been subjected to a mild pyrolysis process or have been treated with steam. The processed materials, generally in pelletised form, may have improved handling and storage properties and may have improved milling properties compared to the untreated biomass materials. These processes also result in modest changes to the chemical composition of the biomass materials, but this will not have a major impact on their behaviour in the energy recovery processes described in the case studies in this report.

Overall, the information gathered from the eight Case Studies forms a basis for the formulation of the techno-economic system/sub-system models which remains the focus of WP3.

The report is organised as follows. Following the “Executive Summary” and the “Scope and Context” sections, each of the 8 technology combinations are presented in individual sections. The “Discussion” section covers the overall assessment based on the boundary conditions and assumptions, and the output in terms of efficiency, and economic parameters.

Table 1: The list of 24 technology combinations evaluated in WP1 (Report Deliverable PM01 D1.1 0.3).

		Post-combustion					Oxy-combustion			Pre-combustion					
		Solvent scrubbing, e.g. MEA, chilled ammonia	Low-temp solid sorbents, e.g. supported amines	Ionic liquids	Enzymes	Membrane separation of CO ₂ from flue gas	High-temp solid sorbents, e.g. carbonate looping	Oxy-fuel boiler with cryogenic O ₂ separation	Oxy-fuel boiler with membrane O ₂ separation	Chemical-looping-combustion using solid oxygen carriers	IGCC with physical absorption e.g. Rectisol, Selexol	Membrane separation of H ₂ from synthesis gases	Membrane production of syngas	Sorbent enhanced reforming using carbonate looping	ZECA concept
Coal IGCC gasification	Direct cofiring	Not feasible					Not feasible			15	17	19	21	23	
	Conversion to 100% biomass	Not feasible					Not feasible			15	17	19	21	23	
Pulverised coal combustion	Direct cofiring	1	3	5	5a	7	9	11	11a	13	Not feasible				
	Conversion to 100% biomass	1	3	5	5a	7	9	11	11a	13					
Dedicated biomass combustion	Fixed grate	2	4	6	6a	8	10	12	12a	14					
	Bubbling fluidised bed	2	4	6	6a	8	10	12	12a						
	Circulating fluidised bed	2	4	6	6a	8	10	12	12a						
Dedicated biomass gasification	Bubbling fluidised bed	Not feasible					Not feasible			14	16	18	20	22	24
	Circulating fluidised bed	Not feasible					Not feasible				16	18	20	22	24
	Dual fluidised bed	Not feasible					Not feasible				16	18	20	22	24
	Entrained flow	Not feasible					Not feasible				16	18	20	22	24

Table 2: An overview of the eight selected technology combinations, studied in WP2.

Criteria	Co-firing amine scrubbing	Dedicated biomass with amine scrubbing	Co-firing oxy-fuel	Dedicated biomass oxy-fuel	Co-firing carbonate looping	Dedicated biomass chemical looping	Co-firing IGCC	Dedicated biomass BIGCC
Current TRL	6 to 7	4	6	5	4 to 5	4	5 to 6	4
Key technical issues	Scale-up, amine degradation,	Scale-up, amine degradation,	Corrosion, O ₂ energy costs, slow response	Corrosion, O ₂ energy costs, slow response	Calcliner firing, solid degradation, large purge of CaO	Loss in activity, reaction rates, dual bed operation	Complex operation, slow response, tar cleaning, retrofit impractical	Complex operation, slow response, tar cleaning, retrofit impractical
Suitability for small scale	Low	High	Low	High	Low	High	Low	High
Plant efficiency with capture	OK	Low	OK	Low	Good	Good	High,	Good
Capital costs with capture	OK	Expensive	OK	High ASU costs	OK	Low cost	OK	Expensive,
UK deployment potential	Immediate capture retrofit opportunities,	retrofit opportunities high long-term potential	retrofit opportunities, long-term doubtful	retrofit opportunities, high long-term potential	capture retrofit opportunities, cement integration	Likely first demos in Europe, UK in ~2020. High long term potential	No current UK plants, several demos by 2020 Long-term doubt	No current UK plants, demo unlikely by 2020. High long-term potential

Table 3: The selected fuel specifications for use in the Case Studies.

Parameter	British bituminous coal	World traded coal	Wood pellets	Raw wood chip
Proximate analysis (% , a.r. ; as received basis)				
Total moisture	12	9.0	7.0	50.0
Volatile matter	28.9	30.5	79.0	39.2
Fixed carbon	43.4	48.8	13.8	9.8
Ash	15.7	11.7	0.2	1.0
GCV (MJ kg ⁻¹ , as received)	24.6	27.2	18.7	9.9
Ultimate analysis (% , a.r)				
Moisture	12.0	9.0	7.0	50.0
C	59.6	65.9	43.5	25.0
H	3.8	4.4	4.5	2.7
N	1.5	1.4	0.2	0.2
O	5.5	7.2	42.6	21.1
S	1.8	0.4	0.01	0.02
Cl	0.2	-	0.01	0.01
Ash	15.6	11.7	0.2	1.0
Ash analysis (%)				
SiO ₂	54.3	74.3	13.7	37.5
Al ₂ O ₃	26.3	17.3	3.3	2.7
Fe ₂ O ₃	7.5	3.7	4.9	2.6
CaO	2.3	1.1	34.4	23.7
MgO	1.7	0.7	6.7	5.2
TiO ₂	1.1	0.8	0.4	0.2
Na ₂ O	1.9	0.3	0.3	2.2
K ₂ O	3.6	0.8	24.0	18.6
P ₂ O ₅	0.3	0.3	5.4	4.9
SO ₃	1.0	0.8	6.8	2.4

CASE STUDY 1

Biomass co-firing in a pulverised coal-fired power plant, with post-combustion amine scrubbing

Biomass co-firing in large coal-fired power plants

Pulverised fuel combustion was developed for the large scale utilisation of coal and other solid fuels for the generation of power and heat. A mixture of pulverised fuel and primary combustion air is injected into the combustion chamber through a number of coal burners located in the walls of the furnace. The combustion process takes place within the furnace volume while the fuel particles are in suspension, and the heat from the flame and hot combustion gases is transferred to the furnace walls and then to the heat exchange tube banks in the convective section, as the flue gases pass through the boiler. This combustion and boiler technology has been the principal means of the generation of electricity from coal since the middle of the twentieth century, and there are many hundreds of these boilers currently in operation worldwide.

It has been predicted that, by 2030, steam coal utilisation worldwide will increase significantly, and that the percentage of total global power production that is generated from coal-fired plants will continue to increase. It is also projected that the world market demand will increasingly be for high efficiency, clean coal power generation, in some cases with biomass co-firing capability and, in the longer term, with the capability to capture and store CO₂.

The more important technical options for the co-firing of biomass materials in large pulverised coal-fired boilers have been summarised in Figure 1, and include the following:

Option 1 involves the milling of biomass materials, usually in the form of pellets, through the existing coal mills, after suitable modification, and the combustion of the milled biomass through the existing pulverised coal firing system, again with modification, if required.

Option 2 involves the pre-mixing of the biomass with coal, in the coal handling system, at modest co-firing ratios, generally less than 10% on a heat input basis, and the milling and firing of the mixed fuel through the existing coal firing system. This has been the most popular approach to biomass co-firing to date, despite the relatively modest co-firing ratios, because of the relatively low capital costs and the short project implementation times associated with this approach.

Options 3, 4 and 5 involve the direct injection of pre-milled biomass into the pulverised coal firing system, i.e.

- into the pulverised coal pipework,
- into modified burners, or
- into new dedicated biomass burners

These options involve higher levels of capital investment, but significantly higher co-firing ratios can be achieved than are possible with Option 2.

Option 6 involves the gasification of biomass in a dedicated unit, normally air blown and at atmospheric pressure, and the co-combustion of the product gas in the pulverised coal boiler. The product gas may or may not be cleaned prior to firing into the coal boiler. Option 6 has been

adopted in only a very small number of plants in Northern Europe, and is not considered further in this report.

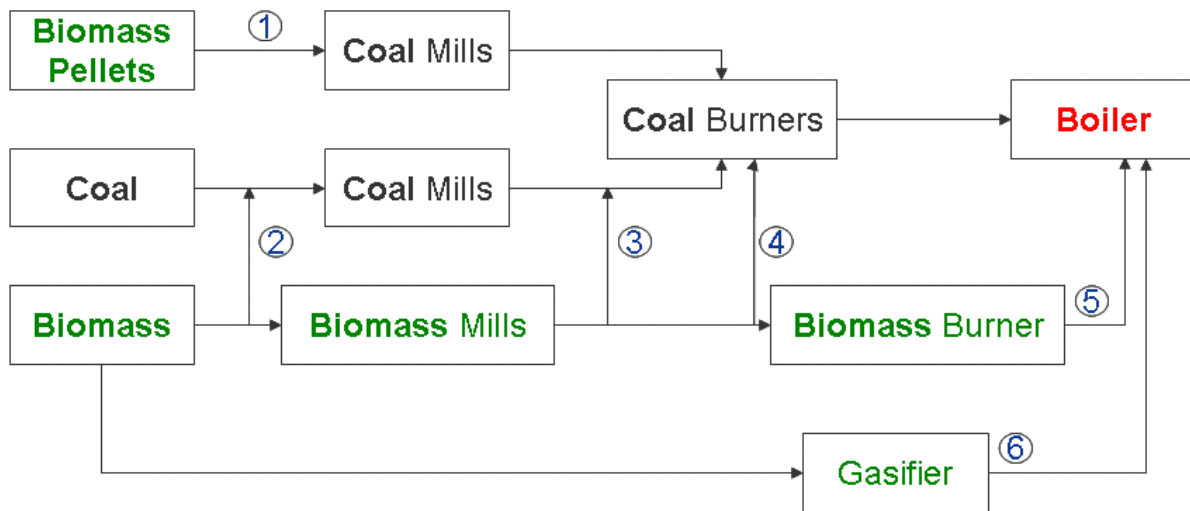


Figure 1: Biomass co-firing options at large pulverised coal-fired power plants.

All of the biomass co-firing scenarios for large coal-fired utility boilers listed below are in current commercial operation, or are being developed by power plant operators in Europe. The interest in Option 1 has increased significantly in recent years both in Europe and North America. It is fair to say, however, that Option 2 has been by far the most popular, principally because it makes maximum use of the existing coal conveying, bunkering, milling and firing systems and hence requires the minimum capital expenditure. This approach has been particularly popular with plant operators who are just starting out on the implementation of biomass co-firing technology, and who may have concerns about the security of supply of the biomass and the economics of co-firing in the longer term. There are, however, limits to the co-firing ratio that can be achieved in this way, principally because of milling constraints.

For operation at higher co-firing ratios, additional milling capacity for the biomass must be installed, generally in the form of hammer mills or the modification of a number of the installed coal mills to process biomass pellets. The pre-milled biomass is then co-fired through a direct injection system, as described under Options 3, 4 and 5 above, or is fired through the existing pulverised coal firing system, after suitable modification. It is clear, therefore, that the selection of the biomass feed material and of the biomass co-firing option has a significant impact on the achievable co-firing ratio, the delivered biomass fuel costs, and the capital and operating costs of the biomass co-firing systems. It should also be noted that practically all of the projects to date have involved the retrofitting of biomass co-firing capability to existing coal power plants. It is likely, however, that a significant number of the future new build coal-fired utility power plants will be designed to have a biomass co-firing capability, probably at around 10-30% on a heat input basis.

Post combustion CO₂ capture by solvent scrubbing

Post-combustion CO₂ capture technology by solvent scrubbing involves the extraction of CO₂ from the combustion flue gases using a specific liquid solvent, normally an aqueous solution of an organic

amine. This technology is applicable to the removal of CO₂ from practically any combustion process, including conventional gas, oil and solid fuel-fired boilers, and gas turbines.

In the current application, the combustion flue gas from the boiler, after suitable flue gas cleaning to remove entrained particulate material, the oxides of nitrogen and sulphur, and HCl is cooled and brought into contact with an aqueous amine solution in the absorber tower, at temperatures typically in the range 40-60°C. The CO₂ is absorbed by the chemical solvent as it passes up through the packed tower, and the cleaned gas at the top of the tower is then washed with water to remove solvent droplets or vapour carry over and return them to the reactor. The cleaned flue gas, with a much reduced CO₂ concentration, then leaves the absorber, and is sent for further processing or is emitted in to the atmosphere.

The 'rich' solvent, containing the chemically-bound CO₂, is then pumped to the stripper or regeneration vessel, via a heat exchanger. The regeneration of the chemical solvent is carried out in the stripper unit at elevated temperatures (100-140°C) and at just above atmospheric pressure. The CO₂ product gas leaves the stripper via the condenser, for further processing.

The 'lean' solvent is pumped back to the absorber tower via the lean/rich amine heat exchanger and a further lean amine cooler unit, to bring it to the absorber temperature level. The amine scrubbing and stripping processes have significant requirements for heat in the form of steam from the steam turbine circuits and or power for the large circulation pumps, fans and other items of equipment.

Status of development and technology readiness level (TRL)

The status of development and commercialisation of biomass co-firing in large boilers depends on the co-firing option under consideration, and these are discussed in the next section of the report.

The application of biomass co-firing employing **Option 1**, as defined above, has been demonstrated in a small number of cases in Northern Europe. It has been shown that large, vertical spindle coal mills can be employed, with fairly modest modifications, to reduce dried and pelletised biomass back to something close to the primary particle size distribution of the sawdust, and that the mill product material can be fired successfully through the existing pulverised coal pipework systems and burners, normally with some modification. This approach to co-firing, or to 100% conversion of coal fired boilers to biomass, is in the full scale demonstration/first commercial project phase, and further projects which involve this approach are planned in the short term future both in Europe and North America.

Option 2 has been by some way the most popular approach to co-firing, as it can be implemented relatively quickly and with modest capital investment, although it should be noted that the co-firing ratio achievable in this way is limited. The great majority of biomass co-firing in the UK and in much of Northern Europe is by pre-mixing biomass with coal, normally in the existing coal handling and conveying system. The mixed fuel is then processed through the installed coal bunkers and mills, and the installed pulverised coal firing equipment. This approach has been well demonstrated and can be regarded as being a well established technology.

Options 3, 4 and 5 involve higher levels of capital investment, but significantly higher co-firing ratios can be achieved than are possible with Option 2. A number of coal-fired power plants in the UK and Northern Europe have installed direct injections systems over the past few years, and this is likely to be one of the favoured options for the provision of biomass co-firing capabilities in new build coal power plant projects. All of the relevant technical approaches to direct injection co-firing involve the

pre-milling of the biomass to a particle size distribution that will provide high combustion efficiency in a pulverised fuel flame, and all involve pneumatic conveying of the pre-milled biomass from the biomass handling/milling facilities to the boilers.

The co-firing of biomass in coal-fired power stations is, therefore, either fully commercial or is in the large scale demonstration phase, depending on the specific co-firing technology being considered. An overall TRL value of 8-9 currently, and TRL 9 by 2020 would appear to be appropriate for this component of the Case Study 1 technology combination.

Post-combustion CO₂ capture using solvent scrubbing is at small-scale demonstration scale (TRL 6) moving to larger scale demonstration (TRL 7), for coal firing, but has not yet been applied commercially to biomass co-firing or 100% biomass firing in converted pulverised coal boilers. A number of the proposed demonstration scale projects at coal plants indicate that they may include biomass co-firing. Given the very high levels of industrial development activity and investment in this subject area, it is not unreasonable to consider that solvent scrubbing technologies for post-combustion CO₂ capture could reach TRL 8 to 9 by 2020.

Overall, therefore, the Technology Readiness Level for the combination of solvent scrubbing applied to biomass co-firing in pulverised coal boilers is currently considered to be 6-7, i.e. at small/large scale demonstration level. There is significant technical and commercial development of both technologies proposed over the next decade and it is considered that the TRL value should increase to 7-8 i.e. full scale demonstration/ first commercial plant stage by 2020.

Process flow diagrams

The summary process flow diagrams and the technical summaries with the key flow rates for the following base cases:

- A 500 MW_e coal-fired power plant with biomass co-firing, and
- A 500 MW_e coal-fired power plant with biomass co-firing, and CO₂ capture by solvent scrubbing.

The basic data have been taken from a fairly detailed report prepared by Foster Wheeler for the IEA Greenhouse Gas R&D Programme in 2009.

It should be noted that the performance and cost information refer to an amine scrubbing process using a monoethanolamine based solvent. Also, ongoing developments in amine solvent formulations and in CO₂ scrubbing process improvements are providing significant advances in process efficiencies, reductions in energy requirements and reductions in capital and operating costs.

In these cases, the coal is a good quality world traded bituminous coal with high calorific value, and the biomass is wet wood chip with a total moisture content around 50%. The co-firing ratio was 10% on a heat input basis.

Looking in the first instance at the flow sheet on Figure 2 and the data in Table 4 for the base case without CO₂ capture, the gross power output is 545.2 MW_e and the total auxiliary power consumption is 26.3 MW_e, or around 4.8% of generation. The net power output is therefore 518.9 MW_e, equivalent to a net electrical efficiency of 44.8%, on an LHV basis.

Turning to the flow sheet for the case with CO₂ capture, on Figure 3 and Table 5, it is clear that flow sheets for the basic fuel handling system, the boiler island and the interface between the boiler and

the turbine island are largely the same, as is expected for a post-combustion technology. It is also clear that the fuel and combustion air flow rates have been held constant. This means that the total flue gas flow rates and the ash production rates, and the reagent usage rates in the flue gas clean-up system are largely the same for both cases.

Table 4: Key process parameters for coal and biomass co-firing without CO₂ capture.

500 MWe PC Boiler Co-firing Coal and Biomass without CO ₂ Capture				
Coal LHV (kJ/kg)		25870		
Biomass LHV (kJ/kg)		7300		
Thermal Energy of Feedstock (based on LHV) (MWt)		1158		
Load				
Gross Electrical Output (Gen) (MWe)		545.2		
Auxiliary Consumptions				
Solid Receiving, Handling and Storage(MWe)		1.4		
Boiler island and flue gas treating (MWe)		15.8		
Power Island (MWe)		3.1		
Utilities (MWe)		6		
Total Electric Power Consumption (MWe)		26.3		4.82% Gen
Net Power Output (MWe)		518.9		95.2% Gen
Gross Plant Efficiency,(based on LHV),(%)		47.1		
Net Electrical Efficiency,(based on LHV),(%)		44.8		
	Flow Rate (t/h)	Temperature (°C)	Pressure (bar a)	
Boiler Island and DeNox System				
Coal	145	ambient	ambient	
Biomass	57	ambient	ambient	
Air	1678	ambient	ambient	
Ammonia	0.5	ambient	ambient	
Fly Ash	14.7	ambient	ambient	
Bottom Ash	3.7	ambient	ambient	
FGD System				
Limestone	4.1	ambient	ambient	
Make-up water	60	ambient	ambient	
By-Product, Gypsum	7.1	ambient	ambient	
Flue Gas to Stack	2046	90	1.005	

Table 5: Key process parameters for coal and biomass co-firing with CO₂ capture by amine scrubbing.

500 MW_e PC Boiler Co-firing Coal and Biomass with CO₂ Capture				
Coal LHV (kJ/kg)		25870		
Biomass LHV (kJ/kg)		7300		
Thermal Energy of Feedstock (based on LHV)(MWT)		1158		
Load				
Gross Electrical Output (Gen) (MWe)		474.1		
Auxiliary Consumptions				
Solid Receiving, Handling and Storage (MWe)		1.4		
Boiler island and flue gas treating (MWe)		16.3		
Power Island (MWe)		2.4		
Utilities (MWe)		8.7		
Auxiliary Consumption excluding CO₂ Capture (MWe)		28.8		6.07% Gen
CO ₂ Capture Plant (MWe)		9.91		2.09% Gen
CO ₂ Compression (MWe)		36.74		7.75% Gen
Total Electric Power Consumption (MWe)		75.5		15.9% Gen
Net Power Output (MWe)		398.9		84.1% Gen
Gross Plant Efficiency, (based on LHV), (%)		41		
Net Electrical Efficiency, (based on LHV), (%)		34.5		
	Flow Rate (t/h)	Temperature (°C)	Pressure (bar a)	
Boiler Island and DeNox System				
Coal	145	ambient	ambient	
Biomass	57	ambient	ambient	
Air	1678	ambient	ambient	
Ammonia	0.5	ambient	ambient	
Fly Ash	14.7	ambient	ambient	
Bottom Ash	3.7	ambient	ambient	
FGD System				
Limestone	4.5	ambient	ambient	
Make-up water	60	ambient	ambient	
By-Product, Gypsum	7.9	ambient	ambient	
Flue Gas to CO ₂ Capture Plant	2046	50	1.01	
CO₂ Capture Plant				
Flue Gas to Atmosphere	1619	100	1.005	
CO ₂ to Compression	356.2	35	1.5	
CO₂ Compression and Drying				
CO ₂ to Storage	350.5	32.7	110	

In the case with CO₂ capture, however, the gross electrical output is significantly lower, at 474.1 MW_e, due to the significant heat requirement of the CO₂ capture plant. This is supplied in the form of a steam take-off from the turbine circuits, as is indicated in Table 5 and Figure 3.

The power consumption of the O₂ capture and compression system is also significant, as indicated in Table 5. The overall effect is that the net electrical efficiency of the power plant with CO₂ capture and compression to 110 bar, is 34.5%, i.e. is more than 10% lower than for the base case without CO₂ capture.

Table 6: Capital costs estimates (2009 €) for Case Study 1, with CO₂ capture by amine scrubbing.

500 MWe PC Boiler Co-Firing Coal and Biomass			
		without CO ₂ Capture	CO ₂ Capture with Solvent Scrubbing
Plant Performance			
Gross Output	MWe	545.2	474.1
Net Output	MWe	518.9	398.9
Efficiency and Emissions			
Gross Electrical Efficiency (LHV)	%	47.1	41.0
Net Electrical Efficiency (LHV)	%	44.8	34.5
Actual CO ₂ Emissions	g/kWh	748.5	973.7
Total CO ₂ Captured	g/kWh	0.0	876.4
Economic Performance			
Capital Costs			
Storage and Handling of Solid Materials	£M	36.3	36.3
Boiler Island and Flue Gas Treating	£M	278.1	278.5
Power Island	£M	117.3	107.9
Utilities & Offsites	£M	77.6	95.0
CO ₂ Capture Plant	£M	0.0	94.8
CO ₂ Compression and Drying	£M	0.0	25.6
Total Installed Costs	£M	509.3	638.2
Land Purchases; Surveys and Fees (10%)	£M	50.9	63.8
Contingency (10%)	£M	50.9	63.8
Project development costs	£M	50.9	63.8
Total Investment Cost	£M	662.1	829.6
Specific Investment Costs	£/MWe	1.276	2.079

High level process control philosophy

The process control of a large pulverised coal power plant in response to grid demand is very complex but is well understood. The control of the co-firing of biomass materials in the boiler furnace will depend on the approach to co-firing adopted.

In the case of Option 2 in Figure 1, which involves mixing the biomass with the coal and milling mixed fuel in the installed coal mills, the biomass feed rate is automatically controlled to add the appropriate mass ratio of biomass to coal, based on a coal flow rate signal from a weigh belt upstream of the biomass mixing point.

In the case of the direct injection systems, Options 3, 4 and 5 in Figure 1, the heat input from the biomass is considered to be additional to that from the coal mills, i.e. the individual coal mill receives a demand signal from the boiler master controller, and in turn will modulate the biomass co-firing system feeder speed. The control of the solvent scrubbing system will also be complex, but the system will maintain the emitted flue gas and CO₂ quality and flow rate by modulation of the flow rate of the solvent solution in the circuit in response to changes in the raw flue gas flow rate and CO₂ concentration. The co-firing of biomass and carbon dioxide capture by solvent scrubbing respond well to changes in biomass and flue gas throughput, and will not have a significant limiting effect on the start-up, shutdown and load following capabilities of the pulverised coal-fired boiler.

Environmental performance

Coal-fired power plants are large stationary sources of pollutions and are subject to a high degree of control over their environmental performance by the regulators. This includes continuous emissions monitoring and reporting of the key gaseous and gas-borne pollutant species, i.e. particulates, NO_x and SO_x, and regular monitoring and reporting of other releases to air, land and water. The emissions standards for power plants are set at European Commission level and are subject to regular review. There may also be specific additional local environmental requirements for individual power plants.

The basic environmental standards for large power plants are described in the Industrial Emissions Directive (2010). The permitted emission levels for the principal gaseous and gas-borne pollutant species are as follow:

Sulphur dioxide	200 mg Nm ⁻³ , dry at 6% O ₂
Nitrogen oxides	200 mg Nm ⁻³ , dry at 6% O ₂
Particulates	20 mg Nm ⁻³ , dry at 6% O ₂

The co-firing of biomass has only a modest impact on the environmental performance of the coal power plant. In general terms, the gaseous and gas-borne emissions from the combustion of biomass are a little lower than for coal, and the total ash production will be lower, depending on the nature of the biomass and the co-firing ratio.

The operation of solvent scrubbing systems will have a positive impact on the environmental performance of the power in terms of the CO₂ emissions. There will be a requirement to control and monitor the releases of any organic solvents to air, land and water.

Capital cost and operating and maintenance cost estimates

The basic parameters for the capital costs as well as the operating and maintenance cost estimates were harmonized across the eight technology combinations studied here. The capital cost estimates cover the Total Installed Costs (TIC) comprising the costs for the power island, storage and handling, boiler island, flue gas treatment, utilities, etc.), and the Total/Specific Investment Costs including contingency, land purchases, civils, project development etc. The operating and maintenance costs include fixed O&M (maintenance and labour, insurance) and variable O&M (fuel/feedstocks,

chemicals, waste disposal etc.). The detailed assumptions are explained in the Discussion section of this Report.

The capital cost estimates for Case Study 1, i.e. for the cases involving the pulverised coal power plant with biomass co-firing, both with and without CO₂ capture by amine scrubbing, are provided in Table 6. The key process parameters and the CO₂ emission and capture levels are listed in the top half of the Table 6. As stated above, the negative impacts of the heat and power requirements of the amine scrubbing and the CO₂ compression and drying processes on the power output from the plant and the power generation efficiency are clear.

The estimated capital cost data for the two cases indicate that the basic capital costs for the fuel handling system, the boiler and the power train are much the same for the cases with and without CO₂ capture, as would be expected from the process flow sheet. The capital costs of the CO₂ capture plant and CO₂ compression/drying equipment are substantial, however, and add around 25% to the overall investment costs of the power plant. The data in Table 6 also indicate that when the reduction in the net electrical output of the power plant, associated with the operation of the carbon dioxide capture system, is taken into account, the specific investment costs increase from £1.276/MW_e to £2.079/MW_e, an increase of 62%.

The annual operating and maintenance (O&M) cost estimates for the two cases, i.e. with and without CO₂ capture, are listed in Table 7. The increases in the costs associated with the basic operation and maintenance of the carbon capture plant are fairly modest. There is a 2% increase in the fixed O&M costs due to the additional maintenance of the amine scrubbing and associated plant, and an 8% increase in the variable O&M costs due to the MEA utilisation. Overall, the increase in the total annual O&M costs due to the amine scrubbing plant is of the order of 9%.

Overview of system performance and further development requirement

As stated above, the current Technology Readiness Level for the combination of solvent scrubbing applied to biomass co-firing in pulverised coal boilers is considered to be 6-7, i.e. at the small/large scale demonstration level, and this is due largely to the development status of the application of amine scrubbing technology to the recovery of CO₂ from combustion flue gases. In this context, it is relevant to note that there is a pilot scale, flue gas side-stream plant which has just been built at Ferrybridge Power Station in Yorkshire. This project, the CCPilot100+ project, involves the capture of 100 tonnes of CO₂ per day from a flue gas side stream, corresponding to approximately 5 MW of electric power. The flue gas side stream is extracted from the ductwork after the newly commissioned flue gas desulphurisation unit. The demonstration plant at Ferrybridge is currently at the commissioning stage, and will be used over the next few years for the development and demonstration of the technology.

There is significant technical and commercial development expected for both biomass co-firing and solvent scrubbing technologies over the next decade, and it is considered that the TRL value should increase to 7-8, i.e. full scale demonstration/ first commercial plant stage by 2020.

It is anticipated that the main thrust of the further development of the biomass co-firing and firing technology will be aimed at two main areas:

- increasing the percentage biomass co-firing with coal and other fossil fuels and of the conversion of coal boilers to 100% biomass firing, and

- Increasing the degree of fuel flexibility to include biomass fuels with higher ash contents, more difficult ash chemistries and higher sulphur and chlorine contents.

This work will involve longer term demonstration projects involving biomass firing and co-firing at selected coal power plants, with extensive monitoring of the boiler plant and environmental performance, and with the appropriate support R&D activities on ash behaviour, fireside corrosion, environmental control, etc.

One of the key areas of concern in any biomass co-firing and biomass conversion project is the potential for significant impacts on the performance and integrity of the boiler. The flue gas and fly ash from dedicated biomass firing are generally significantly more aggressive than those from coal firing with respect to its potential to cause ash deposition on boiler surfaces, and accelerated metal wastage of superheater/reheater surfaces, due to high temperature corrosion. Restrictions on the fuel composition may therefore be a feature of operation at high co-firing ratios and on boilers converted units to 100% biomass. There may also be a requirement to re-tube parts of the boiler and to make use of chemical additives or of protective coatings to control the rate of metal loss by high temperature corrosion processes.

The compliance with existing and future emission limits is an important issue which needs to be considered in the development of the technology for biomass co-firing, and for future CCS applications. Selective Catalytic Reduction (SCR) is an established commercial technology for the control of NO_x emissions from coal fired boilers. The catalyst life and performance can be reduced when firing or co-firing specific biomass fuels, and particularly those with high concentrations potassium and phosphate. This may affect the application of SCR technology to this type of system.

One of the key problem areas with amine scrubbing CO₂ capture systems is that all of the commonly used alkanolamine solvents are subject to degradation processes, and these can occur by three major routes:

- Carbamate polymerisation
- Oxidative degradation
- Thermal degradation

For the scrubbing of combustion flue gases at low temperatures, the amine degradation is largely due to the presence of significant oxygen concentrations in the flue gas. This is a significant problem with the flue gases from coal fired systems.

The acid gases present in coal combustion flue gases, e.g. SO₂ and NO₂, also react with MEA to form heat-stable salts, and this process reduces the CO₂ absorption capacity of the solvent. These acid gas species have to be removed with reasonably high efficiency if amine scrubbing systems are to be used for CO₂ capture. In practical systems, the accumulation of these degradation products in the liquors circulating within the amine scrubbing and stripping units must be controlled.

There is also an issue associated with the relatively corrosive nature of the liquors circulating within the system, and there is significant development work associated with the control of the corrosion rates of the internal surfaces of the key plant items and with the selection of the appropriate materials of construction of the reactors, pipework and other plant components.

Table 7: Operating and maintenance costs estimates (2009 €) for Case Study 1, with CO₂ capture by amine scrubbing.

500 MWe PC Boiler Co-Firing Coal and Biomass			
Operation and Maintenance Costs (O&M) (£M/yr)		without CO ₂ Capture	CO ₂ Capture with Solvent Scrubbing
Fuel Handling, Milling, Boiler Island, Power Island		17.3	16.9
CO ₂ Capture Plant, CO ₂ Compress., and Drying		0.0	3.0
Common Facilities (Utilities, Offsite, etc.)		1.3	1.6
Labour		5.5	5.5
Adm./gen overheads		1.7	1.7
Fixed O&M Costs		25.8	28.7
Fuel			
Coal		100.6	98.3
Biomass (dry basis)		11.2	11.0
Auxiliary Feedstock			
Make-up water		0.0	0.0
Solvents			
MEA		0.0	11.9
Catalyst		1.7	1.6
Chemicals		1.6	1.9
Waste Disposal		5.1	5.0
Variable O&M Costs		120.2	129.7
Total O&M Costs		146.0	158.5

Overall, post-combustion capture technologies are end-of pipe solutions that can both be retrofitted to an existing power plant or applied to new build plants, without a major impact on the flexibility of the power plant. The major disadvantage of these technologies is that they are relatively energy intensive, requiring a significant amount of steam for the regeneration of the solvent. This reduces the electrical output of the power plant and decreases its overall efficiency. Because the CO₂ concentration in the flue gas is relatively low (typically less than 20%) and the CO₂ loading of the solvents is modest, the capture systems also tend to be large, and have high capital cost¹.

Much of the current developments are aimed therefore at finding improvements in the following subject areas:

- The optimisation of the amine formulations and the scrubbing/stripping cycle,
- The reduction of the energy requirement for the stripping of the loaded solvent, and of the system power requirement,
- The minimisation of the costs associated with amine degradation, and

- The control of the corrosion of the internal surfaces of the CO₂ scrubbing and stripping units and the associated equipment.

ⁱ Oxy-Coal Technology for Carbon Dioxide Capture: Worldwide Development Activity Update, EPRI, Palo Alto, CA 2010. 1019673, December 2010.

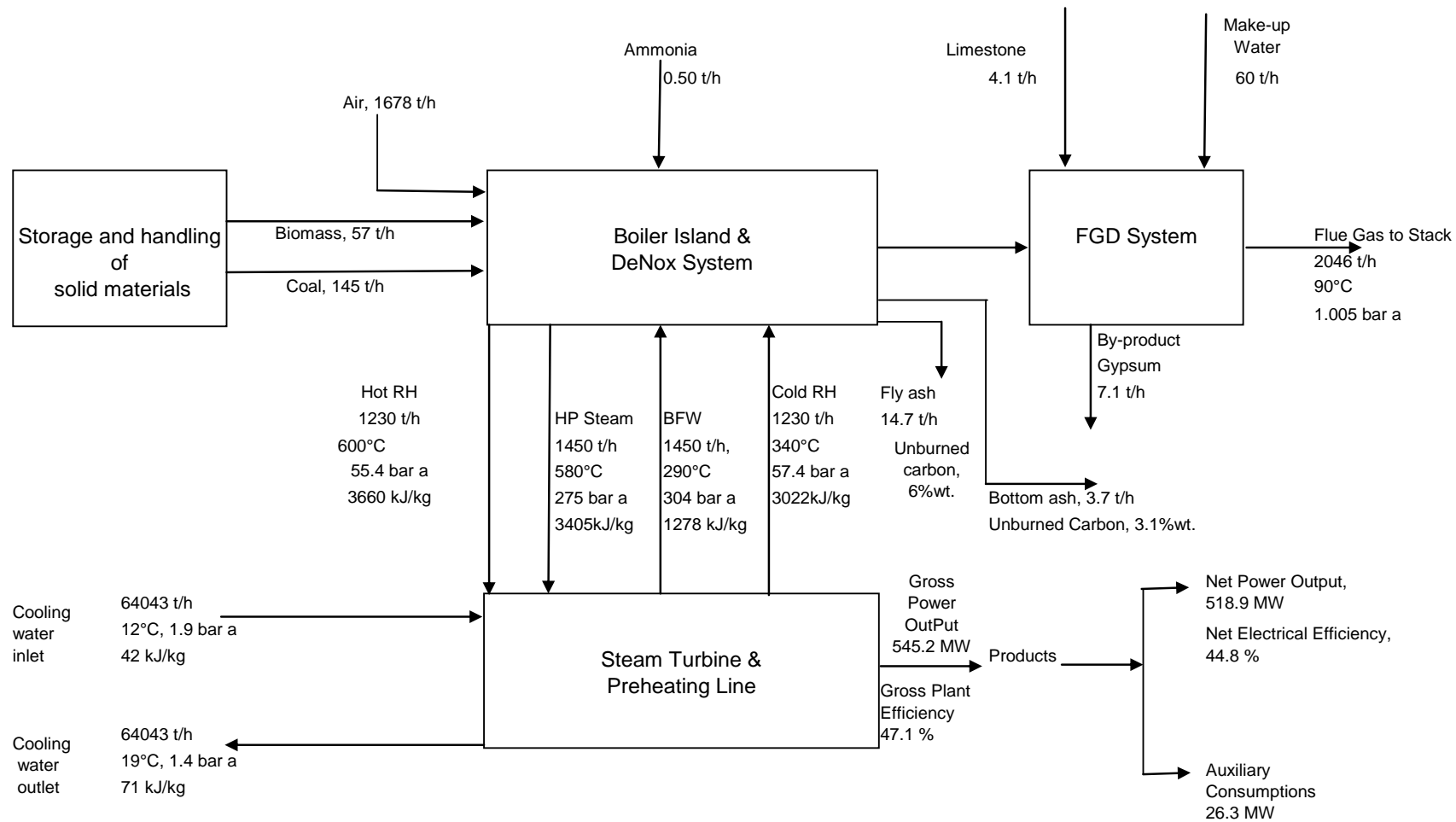


Figure 2: The process flow diagram for a 500MW_e PC boiler co-firing coal & biomass without CO₂ capture.

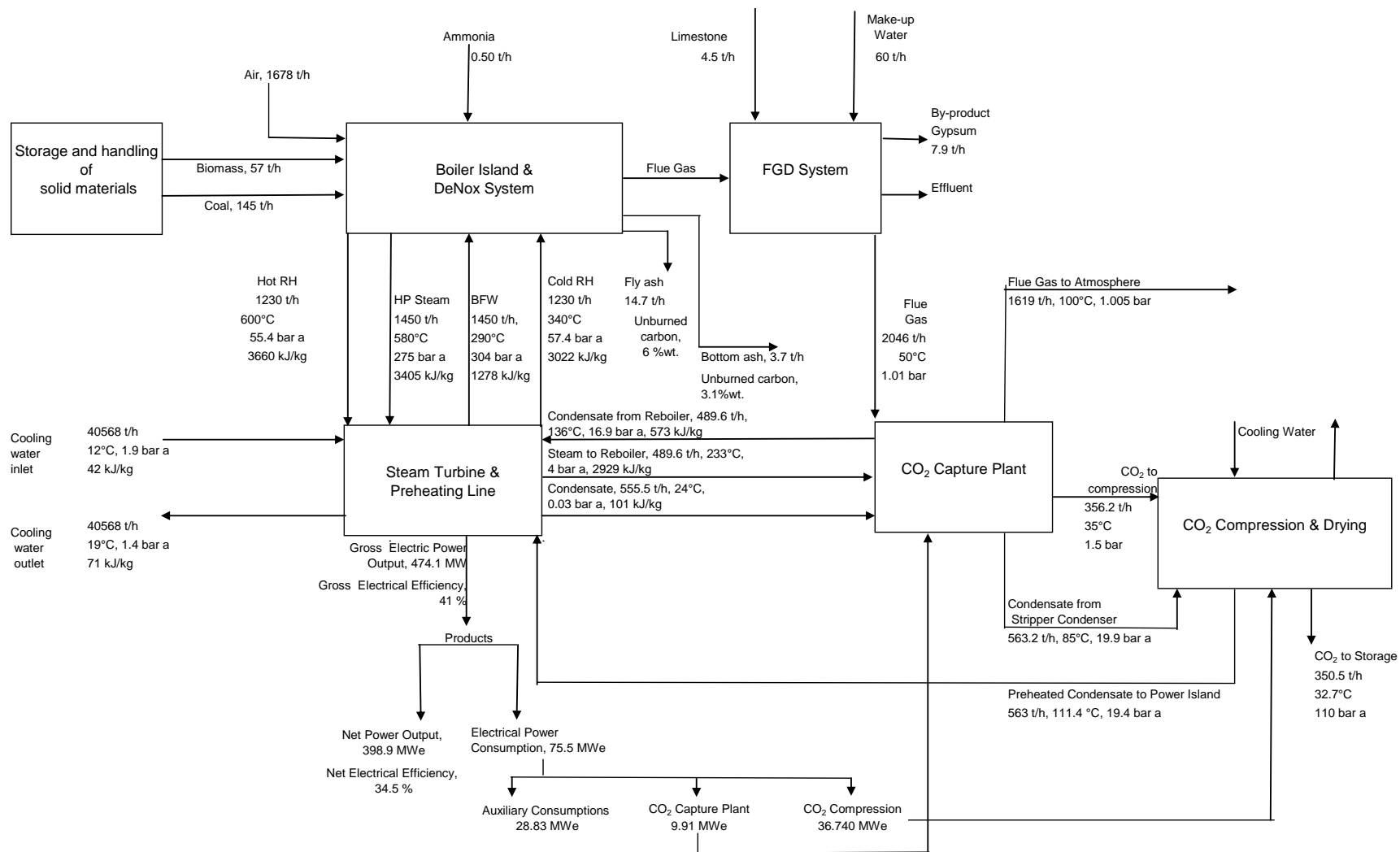


Figure 3: Flow diagram of 500 MW_e PC boiler co-firing coal and biomass with CO₂ capture.

CASE STUDY 2

Biomass combustion in a dedicated power plant with carbon dioxide capture by solvent scrubbing

Biomass combustion in dedicated power plants

For most solid fuels, including solid biomass materials, there are three general types of combustion equipment that are employed for applications at domestic, commercial and industrial scales of operation:

- Grate and retort combustors of a number of different types are employed for domestic fires and stoves, and for the small and medium-sized industrial and commercial applications,
- Fluidised bed combustors, of the bubbling bed (BFB) and circulating bed (CFB) types, are commonly employed for the medium to large scale industrial, commercial and utility applications, with circulating fluidised beds generally being more appropriate at the larger scales of operation, and
- Pulverised fuel combustors, are usually employed for the combustion of coal and other solid fuels for large industrial and utility applications. They have been rarely used for 100% biomass firing, although there has been some interest recently in the conversion of large pulverised coal boilers to 100% biomass in Europe and North America.

For this Case Study, it has been assumed that the combustion system and boiler will be based on a bubbling fluidised bed combustor. In this type of fluidised bed combustion system, the fuel particles, generally up to 10-50 mm or so in size, are suspended in a bed, by a fluidising air stream, along with a size graded bed material, most commonly silica sand. The nominal bed temperatures are generally in the range 750-900°C, depending principally on the behaviour of the ash material generated by the combustion of the solid fuel. The hot flue gases pass up through the furnace chamber and the heat exchangers in the boiler convective section, and then through the flue gas cleaning systems before emitting to the atmosphere. The great majority of the biomass ash leaves fluidised bed combustor furnaces in the form of fly ash particles, generally of up to around 50-100µm in diameter, and are collected in the electrostatic precipitators or bag filters. This fly ash material will also contain fine particles of elutriated bed material.

If limestone is employed as a bed material for sulphur capture, then unreacted limestone, with lime and calcium sulphate/sulphite may also be present in the fly ash materials. The bed material is replaced on a regular basis and a relatively small portion of the bed material and ash may also be removed periodically through bed drains, located underneath the fluidised bed to help maintain the bed inventory. The combustion temperatures in fluidised beds are somewhat lower than those that apply in fixed beds. The bed and freeboard temperatures when burning biomass materials tend to be less than 900°C and the ash particles are largely unfused.

For this Case Study, it has been assumed that the biomass, in a wet chip form, will be burned in a 75 MW_e bubbling fluidised bed boiler with a conventional steam cycle. The arrangement of the boiler and power train will be described in more detail below.

Post combustion CO₂ capture by amine scrubbing

An overview of the key process characteristics of this technology has been presented in Case Study 1 above, and there are no major impacts of a fuel switch from coal to biomass on the performance and operation of the solvent scrubbing system, provided that both the boiler and the solvent scrubbing system are designed appropriately. The combustions flue gases will be cleaned, to reduce the concentrations of particulates and acid gas species, prior to entry to the solvent scrubbing system.

Status of development and technology readiness level (TRL)

The combustion of a wide range of biomass materials in dedicated power stations based on bubbling fluidised bed combustion systems is a commercial technology. There are a large number of plants operating worldwide and a number of experienced equipment suppliers. An overall TRL value of 8 to 9 would appear to be appropriate. Post-combustion CO₂ capture using solvent scrubbing is at small-scale demonstration stage moving to larger scale demonstration, for coal firing. A TRL of 6 would be appropriate. This technology has not as yet been applied commercially to biomass co-firing or 100% biomass firing in converted pulverised coal boilers. Given the very high levels of industrial activity and investment in the area, solvent scrubbing technologies for post-combustion CO₂ capture could reach TRL 8 to 9 by 2020.

Overall, however, the current Technology Readiness Level for the combination of solvent scrubbing applied to dedicated biomass boilers is considered to be 4, since there are not yet any fully integrated pilot plants using dedicated biomass. It is expected, however, that there will be significant technical and commercial development of amine scrubbing technologies (for coal CCS) over the next decade and so the TRL value of a fully integrated biomass combustion plant with amine scrubbing should increase to 6-7 i.e. large/full scale demonstration stage by 2020.

Process flow diagrams

The summary process flow diagrams and the technical summaries with the key flow rates for the following cases, are presented in this section:

- A 75 MW_e dedicated biomass power plant, and
- A 75 MW_e dedicated biomass power plant with CO₂ capture by solvent scrubbing.

The basic data have been taken from a fairly detailed report prepared by Foster Wheeler for the IEA Greenhouse Gas R&D Programme in 2009 [1]. It should be noted that the performance and cost information in this report refer to an amine scrubbing process using an MEA based solvent. The ongoing developments in liquid solvent formulations and process improvements are aimed at providing significant advances in process efficiencies, reductions in energy requirements and reductions in capital and operating costs.

In these cases, the biomass is wet wood chip with a total moisture content around 50%. The process flow diagram for a 75 MW_e power plant based on a bubbling fluidised bed combustor and a conventional steam cycle, and for the same system fitted with post-combustion CO₂ capture by amine scrubbing, have been reproduced in Figures 4 and 5, and the key performance data and process flows are listed in Tables 8 and 9.

The installed equipment includes:

- The fuel reception, storage and handling facilities,

- The bubbling fluidised bed combustor, designed for biomass firing, and the boiler and auxiliary equipment,
- The flue gas clean-up system and chimney,
- The steam turbine/generator and condenser/feed water heating system, and, in Figure 5,

Table 8: The key performance data for the 75 MW_e dedicated biomass power plant with and without CO₂ capture with solvent scrubbing.

75MWe BFB Boiler for Dedicated Biomass Combustion				
	without CO₂ Capture		CO₂ Capture with Solvent Scrubbing	
Biomass LHV (kJ/kg)	7300		7300	
Thermal Energy of Feedstock (based on coal LHV)(MWt)	211		211	
Gross Electrical Output (Gen) (MWe)	83.4		68.5	
Auxiliary Consumption				
Solid Receiving, Handling and Storage (MWe)	0.3		0.3	
Boiler island and flue gas treating (MWe)	3.6		3.8	
Power Island (MWe)	2.13		1.9	
Utilities	1.58		2.6	
Auxiliary Consumption (excluding CO₂ Capture (MWe)	7.61	9.1% Gen	8.6	12.55% Gen
CO₂ Capture Plant (MWe)			2.4	3.50% Gen
CO₂ Compression (MWe)			8.6	12.55% Gen
Total Electric Power Consumption MWe)	7.61	9.1% Gen	19.6	28.61% Gen
Net Power Output (MWe)	75.8	90.88% Gen	48.9	71.4% Gen
Gross Electrical Efficiency, (based on LHV), (%)	39.6		32.5	
Net Electrical Efficiency, (based on LHV),(%)	36		23.2	

In Figure 5, the CO₂ capture plant and the CO₂ compression and drying system have been included. The fuel and air inputs are the same for both cases, and the system is designed for a net power output of 75.8 MW_e from the power plant with no carbon dioxide capture. In this case, the gross power production is 83.4 MW_e and the parasitic power requirement is 7.61 MW_e, around 9% of generation, which is not untypical of a power plant of this type. The total flue gas flow rates and the

ash production rates, and the reagent usage rates in the flue gas clean-up system are the same for both cases, except that there is limestone injection to the plant with the CO₂ scrubbing system to further reduce the flue gas SO_x concentration, and protect the amine scrubber.

It is also clear from the data presented in Tables 8 and 9 that the addition of the carbon dioxide capture plant has a major impact on the power production. The heat requirement of the amine scrubbing plant results in a significant reduction in the gross power output, from 83.4 to 68.5 MW_e, and the additional power requirement of the amine scrubbing and CO₂ processing equipment is significant. The combined effect is a reduction in the net power generation of the plant from 75.8 to 48.9 MW_e, a reduction of around 35.5 %. The overall Gross Electrical Efficiency of the power plant decreases from 39.6% to 32.5% and the Net Electrical Efficiency is reduced from 36% to 23.2%.

Table 9: The major flows and temperature data for the 75 MW_e dedicated biomass power plant with and without CO₂ capture by solvent scrubbing.

75MWe BFB Boiler for Dedicated Biomass Combustion						
	without CO ₂ Capture			CO ₂ Capture with Solvent Scrubbing		
	Flow Rate (t/h)	Temp. (°C)	Pressure (bar a)	Flow Rate (t/h)	Temp. (°C)	Pressure (bar a)
Boiler Island and DeNox System						
Biomass	104	ambient	ambient	104	ambient	ambient
Air	286	ambient	ambient	286	ambient	ambient
Fly Ash	0.9	ambient	ambient	1.1	ambient	ambient
Bottom Ash	0.2	ambient	ambient	0.2	ambient	ambient
FGD System						
Limestone				0.3	ambient	Ambient
Flue Gas to CO ₂ Capture Plant				469	148	1.015
Flue Gas to Stack	469	90	1.005			
CO₂ Capture Plant						
Flue Gas to Atmosphere				313	107	1.005
CO ₂ to Compression				87.3	35	1.5
CO₂ Compression and Drying						
CO ₂ to Storage				85.9	25.9	110

High level process control philosophy

The process control of a fluidised bed biomass power plant in response to grid demand is very complex, however there are many such plants in operation, and the requirements are well understood. In terms of the control of the heat input to the boiler, the biomass feeder speed will respond to a demand from the boiler master controller. The combustion air flow rate is controlled to maintain the flue gas oxygen concentration set point at the boiler outlet, and the forced and induced fans are balanced in such a way as to maintain a small negative pressure in the furnace.

The control of the solvent scrubbing system will also be complex, but the system will maintain the required quality and flow rates of the emitted flue gas and CO₂ streams by modulation of the flow

rate of the solvent solution in the circuit, in response to changes in the raw flue gas flow rate and CO₂ concentration. It is anticipated that the application of carbon dioxide capture by solvent scrubbing does not have a major impact on the start-up, shutdown and load following capabilities of the dedicated biomass power plant.

Environmental performance

Similar to the coal-fired power plants described above, dedicated biomass power plants at industrial scale are large stationary sources of pollution and are subject to a high degree of control over their environmental performance by the regulators. This includes continuous emissions monitoring and reporting of the key gaseous and gas-borne pollutant species, i.e. particulates, NO_x and SO_x, and perhaps others, and regular monitoring and reporting of the releases to land and water.

The emissions standards for power plants are set at a European level, and are subject to regular review. There may also be specific additional local environmental requirements for individual power plants. The specific regulations that apply to a particular biomass boiler are dependent on the nature of the biomass fuel diet. The basic environmental standards for large power plants firing clean biomass materials are described in the Industrial Emissions Directive (IED), published by the EC in 2010. The maximum permitted emission levels for the principal gaseous and gas-borne pollutant species are as follows:

Sulphur dioxide	200 mg Nm ⁻³ , dry at 6% O ₂
Nitrogen oxides	200 mg Nm ⁻³ , dry at 6% O ₂
Particulates	20 mg Nm ⁻³ , dry at 6% O ₂

If waste materials are to be fired, the environmental regulations, and the monitoring and reporting requirements, are much more demanding than for clean biomass materials.

The operation of solvent scrubbing systems will have a positive impact on the environmental performance of the power plant in terms of the reduction in the CO₂ emissions. The levels of the major pollutant species may have to be reduced to concentration levels lower than the IED values to protect the solvent scrubbing system. There will be a specific requirement to control and monitor the releases of any organic solvents to air, land and water from the solvent scrubbing system.

Capital cost and operating and maintenance cost estimates

The capital cost estimates for Case Study 2, i.e. for the cases involving the dedicated biomass power plant, with and without CO₂ capture by amine scrubbing, are provided in Table 10. The key process parameters and the CO₂ emission and capture levels are listed in the top half of Table 10. As was in Case Study 1 above, the negative impacts of the significant heat and power requirements of the amine scrubber and stripper units, and of the CO₂ compression and drying equipment, on the net power output and the power generation efficiency of the plant are clear.

The capital cost estimate data for the dedicated biomass power plants with and without CO₂ capture indicate that the basic capital costs for the fuel handling system, the boiler and the power train are of the same order for the cases with and without CO₂ capture, as would be expected. The capital costs of the CO₂ capture plant and the compression/drying equipment are substantial, however, and add around 38% to the overall investment costs of the power plant.

The data in Table 10 also indicate that when the reduction in the net electrical output of the power plant, associated with the operation of the carbon dioxide capture system, is taken into account, the specific investment costs increase from £2.461 MWe to £5.268 MWe, an increase of 114%.

The estimates of the annual operating and maintenance costs for the dedicated biomass power plant both with and without CO₂ capture, are listed in Table 11. The increases in the operating costs associated with the amine scrubbing plant are fairly modest. There is a 10% increase in the fixed O&M costs due to the additional maintenance of the amine scrubbing and associated plant, and a 12.5% increase in the variable O&M costs, principally due to the MEA utilisation. Overall, the increase in the total annual O&M costs due to the amine scrubbing plant is of the order of 11%.

Table 10: The capital cost estimates (2009 €) for the 75 MWe dedicated biomass power plant, with and without carbon dioxide capture.

75MWe BFB Power Plant			
		without CO ₂ Capture	CO ₂ Capture with Solvent Scrubbing
Plant Performance			
Gross Output	MWe	83.5	68.5
Net Output	MWe	75.8	48.9
Efficiency and Emissions			
Gross Electrical Efficiency (LHV)	%	39.6	32.5
Net Electrical Efficiency (LHV)	%	36.0	23.2
Actual CO ₂ Emissions	g/kWh	1257.3	1948.9
Total CO ₂ Captured	g/kWh	0.0	1754.6
Economic Performance			
Capital Costs			
Storage and Handling of Solid Materials	£M	17.4	17.4
Boiler Island and Flue Gas Treating	£M	80.2	80.7
Power Island	£M	23.8	21.3
Utilities&Offsites	£M	22.2	30.1
CO ₂ Capture Plant	£M	0.0	37.9
CO ₂ Compression and Drying	£M	0.0	10.7
Total Installed Costs	£M	143.5	198.1
Land Purchases; Surveys and Fees (10%)	£M	14.4	19.8
Contingency (10%)	£M	14.4	19.8
Project development costs	£M	14.4	19.8
Total Investment Cost	£M	186.6	257.6
Specific Investment Costs	£/MWe	2.2461	5.267

Overview of system performance and further development requirement

As stated above, the combustion of a wide range of biomass materials in dedicated power stations is now a commercial technology, with a large number of plants operating worldwide and a number of experienced equipment suppliers.

Post combustion CO₂ capture using liquid solvent scrubbing is at small-scale demonstration scale (TRL 6) moving to larger scale demonstration, for coal firing, but has not as yet been applied commercially to 100% biomass firing in converted pulverised coal boilers. Given the very high levels of industrial activity and investment in the area, however, it is considered that solvent scrubbing technologies for post-combustion CO₂ capture could reach TRL 8 to 9 by 2020. Overall, it is expected that the TRL value for fully integrated combination of post combustion capture by amine scrubbing to dedicated biomass combustion plants will be 6-7 by 2020.

There are a few major development issues associated with dedicated biomass combustion and boiler technology. With some biomass materials, there may be concerns about high chlorine and alkali metal contents, which can lead to increased risks of excessive ash deposition on boiler surfaces, increased metal wastage rates of high temperature boiler components due to fireside corrosion and of negative impacts on SCR catalyst performance. These risks are minimised by careful equipment design, close control of the delivered fuel specification, and, in some cases, the use of fireside additives can be of value.

The developments in grate firing and fluidised bed firing in recent years have largely been associated with the application of these technologies to an ever wider range of fuels and, in the case of CFB combustion and boiler technology, with the increase in boiler plant sizes from industrial scale to the utility scale. Otherwise the developments have generally been incremental in nature. The development requirements of the solvent scrubbing technologies have been described in some detail under Case Study 1, which was concerned with their application to large coal boilers co-firing biomass. There are no significant additional development requirements specifically associated with their application to dedicated biomass plants.

Table 11: The operating and maintenance cost estimate (2009 €) for the 75 MW_e biomass power plant with and without carbon dioxide capture.

75MWe BFB Boiler			
		without CO₂ Capture	CO₂ Capture with Solvent Scrubbing
Operation and Maintenance Costs (O&M)			
Fuel Handling, Milling, Boiler Island, Power Island	£M/yr	4.9	4.8
CO ₂ Capture Plant, CO ₂ Compress., and Drying	£M/yr	0.0	1.2
Common Facilities (Utilities, Offsite, etc.)	£M/yr	0.4	0.5
Labour	£M/yr	5.5	5.5
Adm./gen overheads	£M/yr	1.7	1.7
Fixed O&M Costs	£M/yr	12.4	13.7
Fuel			
Coal	£M/yr	0.0	0.0
Biomass (dry basis)	£M/yr	20.5	20.0
Auxiliary Feedstock			
Make-up water	£M/yr	0.0	0.0
Solvents			
MEA	£M/yr	0.0	2.9
Catalyst	£M/yr	0.0	0.0
Chemicals	£M/yr	0.0	0.1
Waste Disposal	£M/yr	0.3	0.4
Variable O&M Costs	£M/yr	20.8	23.4
Total O&M Costs	£M/yr	33.3	37.1

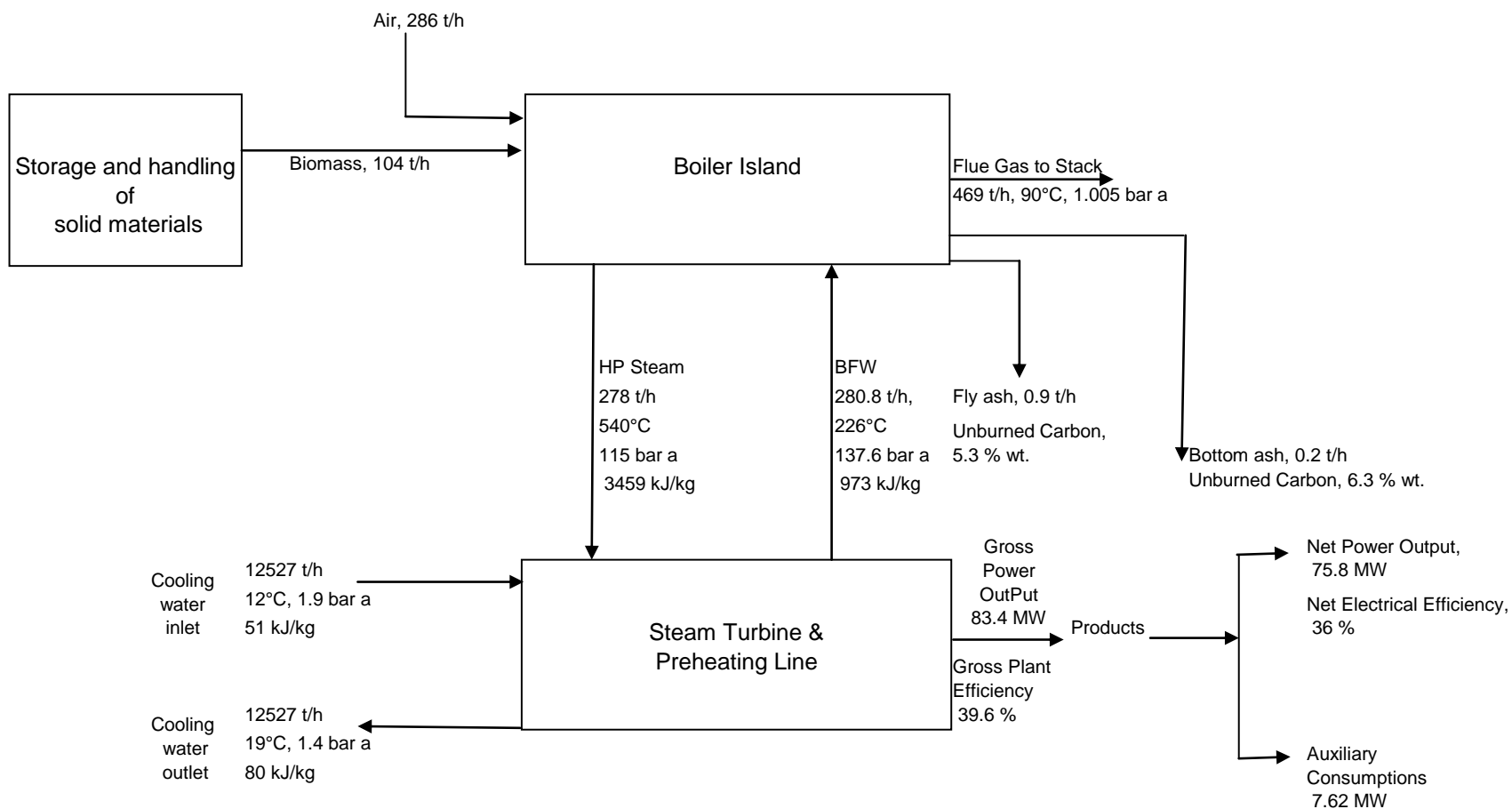


Figure 4: The Process Flow Diagram for a 75MW BFB, dedicated biomass power plant.

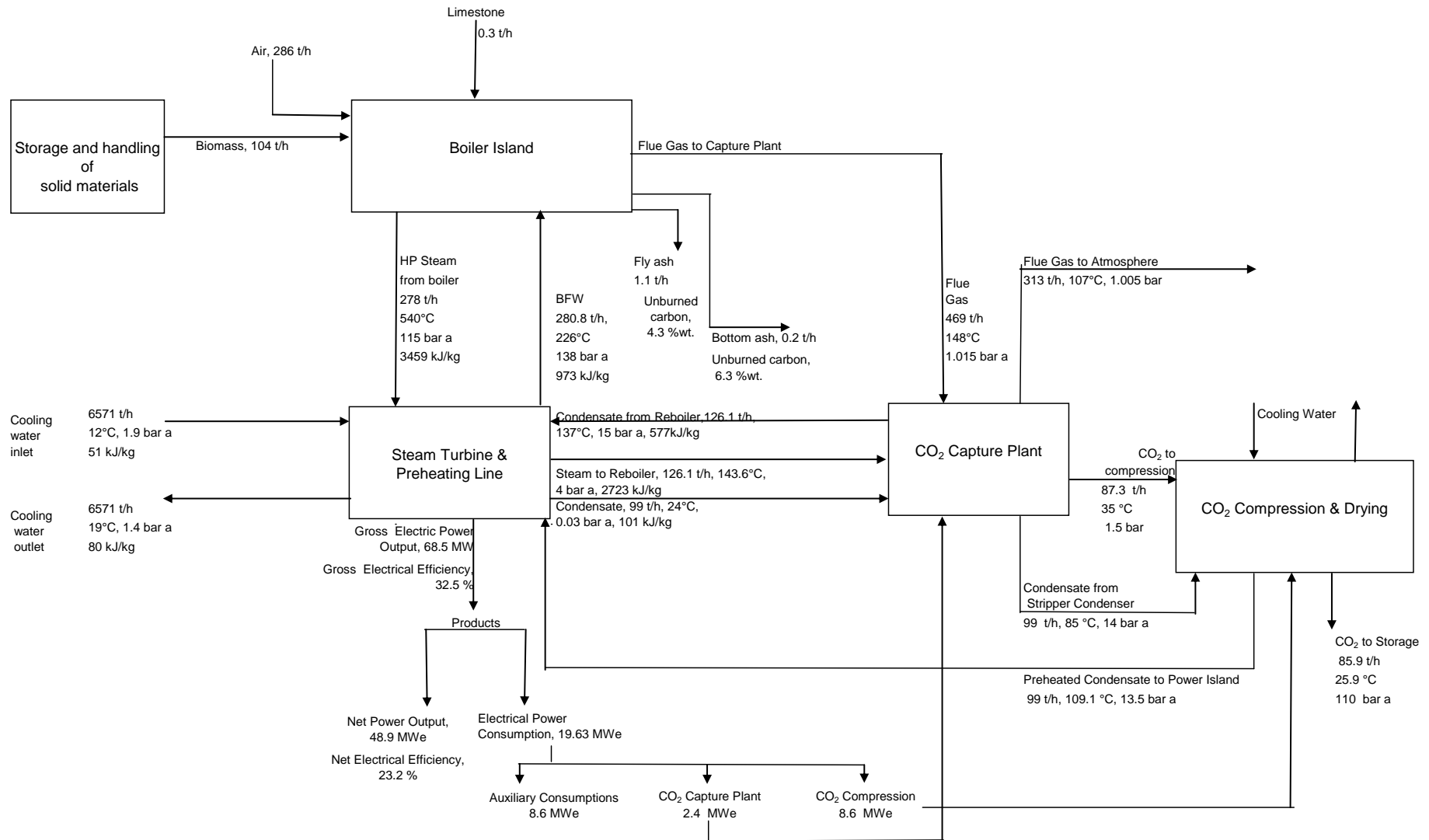


Figure 5: The Process Flow Diagram for a 75MW_e BFB, dedicated biomass combustion with CO₂ capture.

CASE STUDY 3 Biomass co-firing in a large pulverised coal power plant with CO₂ capture by oxyfuel firing

The more important technical options for the co-firing of biomass materials in large pulverised coal-fired boilers are summarised in Figure 1, and the technical and other aspects of this technology have been described under Case Study 1 above.

Oxy-combustion involves the combustion of a fossil fuel in a mixture of oxygen and recycled flue gas, rather than air, to produce a flue gas which comprises mainly CO₂ and water, rather than nitrogen and CO₂. The CO₂ concentration in the flue gas from an oxyfuel combustion firing system is, therefore, significantly higher than in the flue gas from an air firing system, and the CO₂ can be cleaned, compressed and stored with significantly less downstream processing than would be necessary with air firing. The key elements of the process are illustrated in Figure 6. The oxygen required for combustion in the oxyfuel combustion plant, is obtained from a dedicated cryogenic air separation unit. In this case, the oxygen is mixed with the combustion flue gases recycled from the FGD plant exit, to provide the combustion medium for the pulverised coal. There are other oxyfuel firing processes which involve less cleaning of the recycled flue gases. The flue gas exhaust stream from the oxyfuel combustion process is chilled and compressed to separate out gaseous impurities, to provide a CO₂ product of the required quality.

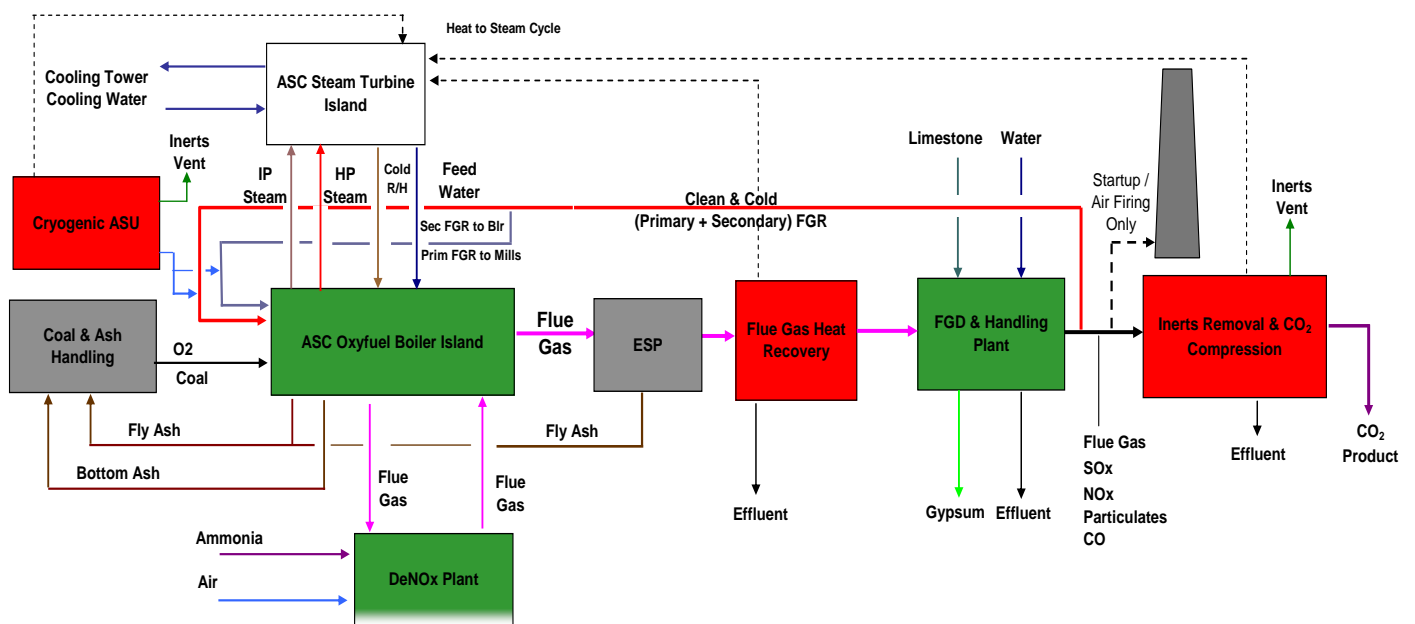


Figure 6: Oxy-combustion CO₂ capture applied to a large pulverised coal-fired power plant.

The status of development and technology readiness level

As stated above, the co-firing of biomass in coal-fired power stations is commercial or is in the large scale demonstration phase, depending on the specific co-firing technology being considered. An

overall TRL value of 8-9 currently, and TRL 9 by 2020 would appear to be appropriate for this component of the overall Case Study 3.

The oxyfuel combustion of fossil fuels with CO₂ capture is not yet fully commercial. A number of integrated pilot plants have been or are currently being built (mainly only using coal), and detailed plans to build commercial power plants and to convert existing thermal power plants to oxy-combustion are being developed at the present time.

The current TRL of oxy-fuel technology is therefore around 5 to 6, and given the high level of international activity, is judged likely to reach TRL 8 to 9 for power applications, including the application to coal plants co-firing biomass, by 2020.

Process flow diagram

The process flow diagrams for coal-biomass co-firing in a 500 MW_e pulverised coal power plant with and without CO₂ capture by oxyfuel firing are presented in Figures 7 and 8. The key plant performance data and the process flows are listed in Tables 12 and 13.

Looking in the first instance at the flow sheet on Figure 9 and the data in Table 12 for the base case without CO₂ capture, the gross power output is 557 MW_e and the total auxiliary power consumption is around 42 MW_e, or around 7.6% of the gross generation. The net power output is therefore 514 MW_e, equivalent to a net electrical efficiency of 44.5%, on a LHV basis.

The flow sheet on Figure 10 for the system with CO₂ capture by oxyfuel firing is significantly more complex, and includes,

- The new interfaces between the air separation unit, the turbine island and the boiler,
- The flue gas recycle loops to the boiler, and
- The interfaces between the CO₂ compression/purification system and the turbine island.

The power consumption of the air separation unit and the CO₂ drying and compression system is also significant, as indicated in Table 12. The overall effect is that the net electrical efficiency of the power plant with CO₂ capture and compression to 110 bar, is 34.4%, around 10% lower than the case without CO₂ capture.

High level process control philosophy

As stated above, the process control of a large pulverised coal power plant in response to grid demand is very complex but is well understood. The control of the co-firing of biomass materials in the boiler furnace will depend on the approach to co-firing adopted, however in most cases the heat input from the biomass is considered to be additional to that from the coal mill, i.e. the coal mill receives a demand signal from the boiler master controller and in turn will modulate the biomass co-firing system feeder speed along with that of the coal feeder to meet this demand.

The operation of the oxyfuel firing system does introduce significant additional complications to the combustion control system, particularly associated with the transition from air firing to oxyfuel firing. These issues are the subject of significant development work at pilot and demonstration scale projects planned over the next few years. The operation of the ASU will impose significant limitations on the flexibility of operation of the power plant. The precise nature of these limitations will vary depending on a number of site specific factors.

Table 12: Performance data of in a PC boiler with coal-biomass co-firing, with and without CO₂ capture by oxyfuel firing.

500 MWe PC Boiler Coal Firing with Biomass				
	without CO ₂ Capture		CO ₂ Capture with Oxyfuel Combustion	
Coal LHV (kJ/kg)	23478		23478	
Biomass LHV (kJ/kg)	17110		17110	
Co-Firing % (based on LHV)	20% Biomass & %80 Coal		20% Biomass & %80 Coal	
Thermal Energy of Feedstock (based on LHV) (MWt)	1158		1158	
Gross Electrical Output (Gen) (MWe)	557.25		559	
Auxiliary Consumption				
Turbine Island (MWe)	18.57		17.99	
Draft & Milling Plant (MWe)	10.33		10.05	
CW System (MWe)	3.81		3.7	
DeNOx Plant (MWe)	2.27		-	
DeSOx Plant (MWe)	3.08		3	
BoP (ESP + Others) (MWe)	4.26		4.76	
Auxiliary Consumption (excluding ASU and CO₂ Capture)(MWe)	42.32	7.6% Gen	39.5	7.06% Gen
Air Separation Unit (MWe)	-		67.3	12.03% Gen
Capture & Compression Plant (MWe)	-		54.3	9.71% Gen
Total Electric Power Consumption (MWe)	42.32	7.6% Gen	161.1	28.82% Gen
Net Power Output (MWe)	514.93	92.41% Gen	397.9	71.2% Gen
Gross Electrical Efficiency, (based on LHV), (%)	48.12		48.27	
Net Electrical Efficiency, (based on LHV),(%)	44.47		34.36	

Environmental performance

As stated above under Case Study 1, coal-fired power plants are large stationary sources of pollution and are subject to a high degree of control over their environmental performance by the regulators. This includes continuous emissions monitoring and reporting of the key gaseous and gas-borne pollutant species, i.e. particulates, NO_x and SO_x, and regular monitoring and reporting of other releases to air, land and water. The emissions standards for power plants are set at European Commission level and are subject to regular review. There may also be specific additional local environmental requirements for individual power plants.

Table 13: Flow and temperature data in a PC with coal-biomass co-firing, with and without CO₂ capture by oxyfuel firing.

500 MWe PC Boiler Coal Firing with Biomass								
	without CO ₂ Capture				CO ₂ Capture with Oxyfuel Combustion			
	Flow Rate (kg/s)	Flow Rate (t/h)	Temperature (°C)	Pressure (bar g)	Flow Rate (kg/s)	Flow Rate (t/h)	Temperature (°C)	Pressure (bar g)
Boiler Island								
Coal	38.67	139.2			38.67	139.2		
Biomass	14.62	52.63			14.62	52.63		
Air	481	1732						
Oxygen to Primary Flue Gas Recycle					19.3	69.48	163.9	1.01
Oxygen to Secondary Flue Gas Recycle					77.7	279.7	163.9	1.01
Primary Flue Gas Recycle to mills (including O ₂ addition)					105.9	381.2	215.7	0.978
Secondary Flue Gas Recycle to Boiler (Including O ₂ addition)					330.3	1189	296.7	0.94
Bottom Ash	1.3	4.68			1.3	4.68		
Flue Gas out of boiler to GGH					415.2	1495	349.9	0.906
					85	306	349.9	0.895
Flue Gas to DeNOx Plant	513.6	1849	360.3	1.005				
Electrostatic Precipitator (ESP)								
Flue Gas to ESP	545.6	1964	141.5	0.997	510.8	1839	214.2	0.901
Flue Gas out of ESP	551.4	1985	139.9	0.995	513.3	1848	212.7	0.899
Fly Ash	5.1	18.36			5.1	18.36		
Flue Gas Heat Recovery								
Flue Gas to FGD-GGH					513.3	1848	105	0.897
FGD-GGH								
Flue Gas to FGD Handling Plant	545.8	1965	83.6	0.995	513.28	1848	87.9	0.893
FGD & Handling Plant								
Make-up water	8.03	28.91			3.04	10.94		
Flue Gas out of FGD	560	2016	46.9	0.995	515.8	1857		0.895
Flue Gas to Stack	565.49	2036	83.6	1.02				
Overall Flue Gas Recycle					349.8	1259		
Direct Contact Cooler (DCC)								
Flue Gas to DCC					166.05	597.8		

The basic environmental standards for large power plants are described in the Industrial Emissions Directive (2010). The permitted emission levels for the principal gaseous and gas-borne pollutant species are as follows:

Sulphur dioxide	200 mg Nm ⁻³ , dry at 6% O ₂
Nitrogen oxides	200 mg Nm ⁻³ , dry at 6% O ₂
Particulates	20 mg Nm ⁻³ , dry at 6% O ₂

The co-firing of biomass has only a modest impact on the environmental performance of the coal power plant. In general terms, the gaseous and gas-borne emissions from the combustion of biomass are a little lower than for coal, and the total ash production will be lower, depending on the nature of the biomass and the co-firing ratio. The operation of the oxyfuel firing system does not involve the use of any additional reagents and will not result in any significant increases in the levels of releases of any of the prescribed pollutant species from the power plant.

Table 14: The capital cost estimate (2009 €) for the 500 MW_e pulverised coal boiler, co-firing biomass, with and without carbon dioxide capture by oxyfuel firing.

500 MWe PC Boiler Co-Firing Coal and Biomass			
		without CO ₂ Capture	CO ₂ Capture with Oxyfuel Combustion
Plant Performance			
Gross Output	MWe	545.2	545.2
Net Output	MWe	518.9	388.7
Efficiency and Emissions			
Gross Electrical Efficiency (LHV)	%	47.1	47.1
Net Electrical Efficiency (LHV)	%	44.8	33.6
Economic Performance			
Capital Costs			
Storage and Handling of Solid Materials	£M	36.3	36.3
Boiler Island and Flue Gas Treating	£M	278.1	278.5
Power Island	£M	117.3	107.9
Utilities & Offsites	£M	77.6	95.0
ASU	£M	0.0	153.9
CO ₂ Compression and Drying	£M	0.0	26.4
Additional Draft Plant, and Duct work	£M	0.0	11.0
Total Installed Costs	£M	509.3	709.1
Land Purchases; Surveys and Fees (10%)	£M	50.9	70.9
Contingency (10%)	£M	50.9	70.9
Project development costs	£M	50.9	70.9
Total Investment Cost	£M	662.1	921.8
Specific Investment Costs	£/MWe	1.276	2.371

Capital and operating cost estimates

The capital cost estimates for Case Study 3, i.e. for the cases involving the pulverised coal power plant with biomass co-firing, with and without CO₂ capture by oxyfuel firing, are presented in Table 14.

As described above, the negative impact of the power requirements of the air separation unit and the CO₂ compression and drying processes on the power output and the overall net power generation efficiency are clear from the data presented in the top half of Table 14.

The estimated capital cost data for the two cases indicate that the basic capital costs for the fuel handling system, the boiler and the power train are of the same order for the cases with and without CO₂ capture, as would be expected. The additional capital costs associated with the oxyfuel firing system, and principally for the ASU, and the CO compression/drying plants, are substantial.

The data on this table also indicate that when the reduction in the net electrical output of the power plant, associated with the operation of the ASU, the oxyfuel firing system and the carbon dioxide drying and compression systems, is taken into account, the specific investment costs increase from £1.276 MWe⁻¹ to £2.371 MWe⁻¹, an increase of 86%.

The annual operating and maintenance (O&M) cost estimates for the two cases, i.e. with and without CO₂ capture, are listed in Table 15. These are very similar, indicating that there are no significant additional operating costs associated with the oxyfuel firing system.

Overview of system performance and further development requirement

As stated above, the current Technology Readiness Level for the fully integrated combination of CO₂ capture by oxyfuel firing applied to biomass co-firing in pulverised coal boilers is considered to be 6, i.e. at small/large scale demonstration level.

There is significant technical and commercial development of both the biomass firing and co-firing technologies and the oxyfuel firing technology proposed over the next decade and it is considered that the TRL value for a fully integrated co-firing oxyfuel combustion power plant should increase to 7, i.e. full scale demonstration stage by 2020.

As stated above it is anticipated that the main technical direction of the further development of the biomass co-firing and firing technology will be in two main areas:

- The increasing of the percentage biomass co-firing that can be achieved without significant additional negative impacts on the plant performance and integrity, and
- The increasing of the degree of fuel flexibility, i.e. to include the capability to co-fire biomass fuels with higher ash contents, more difficult ashes with lower fusion temperatures and higher chlorine contents, etc.

This work will, in the main, involve the establishment of long term demonstration projects involving biomass firing and co-firing at selected coal power plants, with extensive programmes of plant operation and monitoring and with the appropriate supporting R&D activities on ash deposition behaviour, fireside corrosion, environmental control, etc.

Oxyfuel firing technology has now been demonstrated at large pilot scale and the major emphasis of future development is increasingly on full scale boiler demonstration projects.

There is currently significant R&D effort on the risks of boiler tube corrosion issues because of the potential for increasing concentrations of CO₂, H₂O, sulphur oxides and HCl in the flue gases in contact with the high temperature boiler surfaces within the flue gas recirculation loop.

There is also significant development effort on the use of membrane separation techniques for oxygen separation, and on the integration of these systems with the boiler island. It is claimed by a number of the companies involved in these developments that the use of membrane separation technology as an alternative to conventional cryogenic separation, for oxygen production can provide significant savings in capital and operating costs, and significant reductions in the energy requirements. It is anticipated that the membrane separation technology may have the first applications at the scales of operation relevant to the power industry over the next five years.

Table 15: The operating and maintenance cost estimate (2009 €) for the 500 MW_e pulverised coal boiler, co-firing biomass pellets, with and without carbon dioxide capture by oxyfuel firing.

500 MWe PC Boiler Co-Firing Coal and Biomass			
		without CO ₂ Capture	CO ₂ Capture with Oxyfuel Combustion
Operation and Maintenance Costs (O&M)			
Fuel Handling, Milling, Boiler Island, Power Island	£M/yr	17.3	16.9
ASU, CO ₂ Compress., and Drying	£M/yr	0.0	3.0
Common Facilities (Utilities, Offsite, etc.)	£M/yr	1.3	1.6
Labour	£M/yr	5.5	5.5
Adm./gen overheads	£M/yr	1.7	1.7
Fixed O&M Costs	£M/yr	25.8	28.7
Fuel			
Coal	£M/yr	87.6	85.7
Biomass (dry basis)	£M/yr	53.2	52.1
Auxiliary Feedstock			
Make-up water	£M/yr	0.0	0.0
Solvents			
Catalyst	£M/yr	1.7	0.0
Chemicals	£M/yr	1.6	1.9
Waste Disposal	£M/yr	5.1	5.0
Variable O&M Costs	£M/yr	149.2	144.7
Total O&M Costs	£M/yr	175.0	173.4

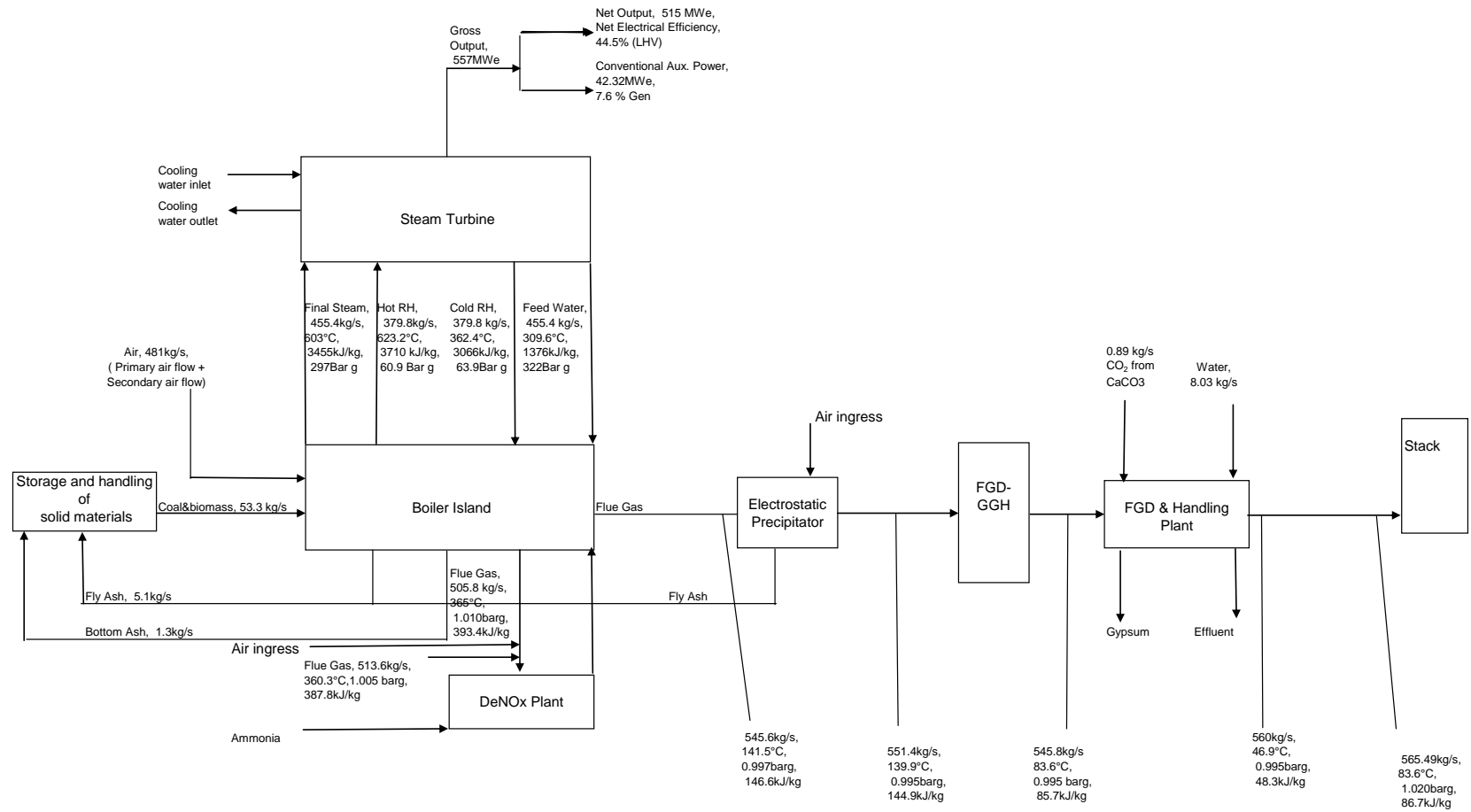


Figure 7: Flow diagram of coal biomass co-firing in a PC boiler without CO₂ capture.

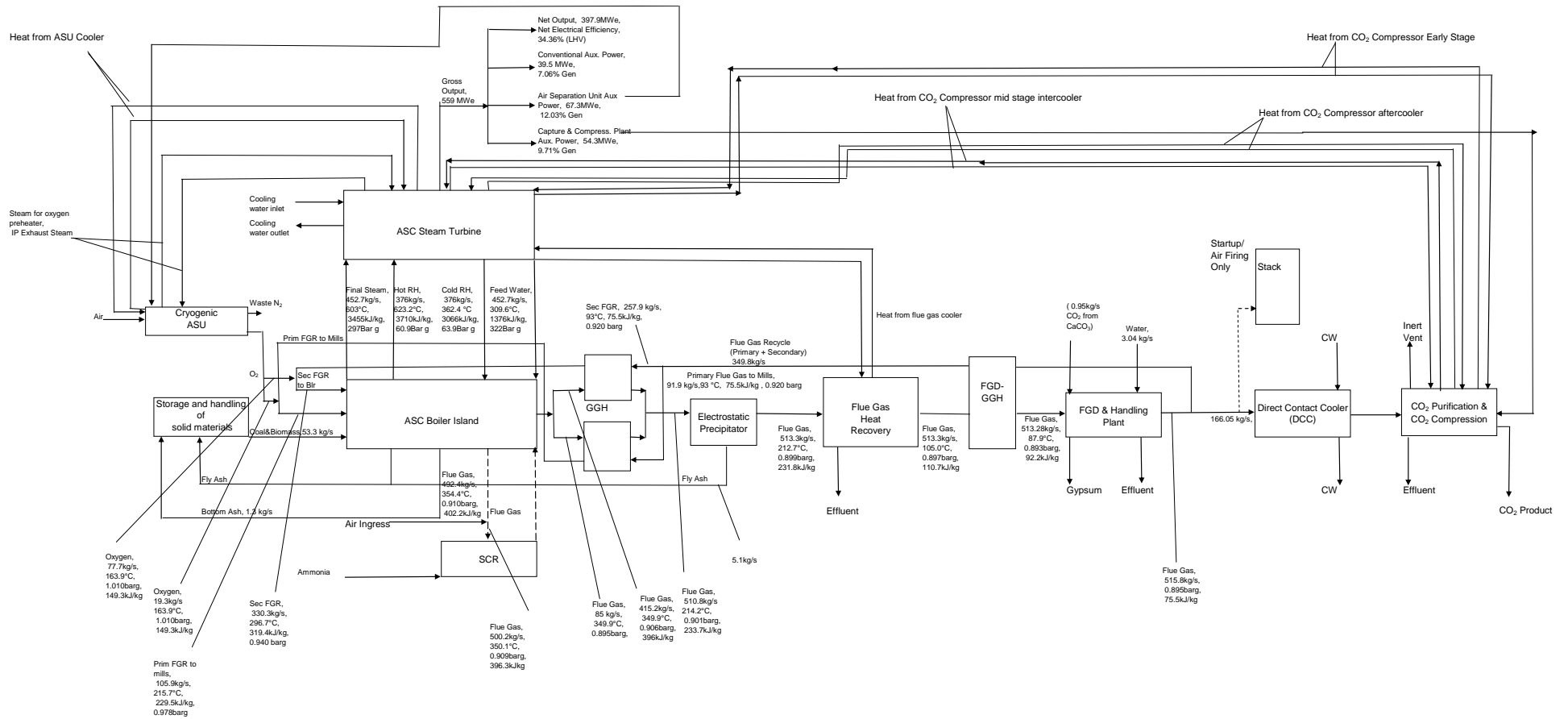


Figure 8: Flow diagram of biomass-coal co-firing in a PC boiler with CO₂ capture by oxyfuel combustion.

CASE STUDY 4

A 75 MW_e dedicated biomass power plant with carbon capture by oxyfuel firing

For this Case Study, it has been assumed that the combustion system and boiler will be of conventional design, and will be based on a bubbling fluidised bed combustor. In fluidised bed combustion systems, the fuel particles, generally up to 10-50 mm or so in size, are suspended in a bed, by a fluidising air stream, along with a relatively coarse-grained bed material. The nominal bed temperatures are generally in the range 750-900°C, depending principally on the fusion behaviour of the ash.

The great majority of the biomass ash material leaves fluidised bed combustor in the form of fly ash particles. This material will also contain fine particles of elutriated bed material, which is most commonly silica sand. If limestone is employed as a bed material for sulphur capture, then unreacted limestone, with lime and calcium sulphate/sulphite may also be present in the bed and fly ash materials. The bed material is replenished on a regular basis and a relatively small portion of the bed material and ash may also be removed periodically through bed drains, located underneath the fluidised bed to help maintain the bed quality and inventory. The combustion temperatures in fluidised beds are somewhat lower than those that apply in fixed beds and in suspension flames. The bed and freeboard temperatures when burning biomass materials tend to be less than 900°C and, at these temperatures, the ash particles are largely unfused. For this Case Study, it has been assumed that the biomass in chip form will be burned in a fairly conventional 75 MW_e bubbling fluidised bed boiler and power train arrangement. These will be described in more detail below.

Oxyfuel combustion processes involve the combustion of the fuel in a mixture of oxygen and recycled flue gas, rather than in air, to produce a flue gas which comprises mainly CO₂ and water, rather than nitrogen and CO₂. The CO₂ concentration in the flue gas from an oxyfuel combustion firing system is, therefore, significantly higher than in the flue gas from air firing, and hence the CO₂ can be cleaned and compressed with significantly less downstream processing than would be necessary with the CO₂/N₂ mixture from air firing.

Status of development and technology readiness level

As stated above, the combustion of a wide range of biomass materials in bubbling fluidised bed boilers with conventional steam turbines is a fully commercial technology with many plants in operation worldwide. An overall TRL value of 8 to 9 currently would appear to be appropriate for this plant component.

The oxyfuel combustion of fossil fuels with CO₂ capture is not yet commercial. A number of integrated pilot plants have been or are currently being built, mainly with coal firing only and detailed plans for both new-build power plants, and to convert existing pulverised coal power plants to oxyfuel combustion are being developed at the present time.

The current TRL of oxy-fuel technology is therefore around TRL 5 to 6, however, given the high level of international activity that is planned over the next few years, it is judged likely to reach TRL 8 for power plant applications by 2020.

Table 16: Performance data of 75 MW_e BFB dedicated biomass power plant with and without CO₂ capture by oxyfuel combustion.

75MWe BFB Boiler for Dedicated Biomass Combustion				
	without CO ₂ Capture		CO ₂ Capture with Oxyfuel Combustion	
Biomass LHV (kJ/kg)	7300		7300	
Thermal Energy of Feedstock (based on LHV)(MWt)	211		211	
Gross Electrical Output (Gen) (MWe)	83.4		84.6	
Auxiliary Consumption				
Boiler island and flue gas treating (MWe)	3.6		3.68	
BoPP Auxiliary Power Consumption (MWe)	4.01		4.95	
Auxiliary Consumption (excluding CO₂ Capture (MWe)	7.61	9.1% Gen	8.63	10.20% Gen
Capture Auxiliary Power Consumption (MWe)	-		27.12	32.05% Gen
Total Electric Power Consumption (MWe)	7.61	9.1% Gen	35.75	42.26% Gen
Net Power Output (MWe)	75.79	90.9% Gen	48.85	57.74% Gen
Gross Electrical Efficiency, (based on LHV), (%)	39.53		40.09	
Net Electrical Efficiency, (based on LHV),(%)	35.9		23.2	

Process flow diagram

The process flow diagram for biomass firing in a 75 MW_e dedicated power plant with CO₂ capture by oxyfuel firing is presented in Figure 9. The key plant performance data and the more important process flows are listed in Tables 16 and 17.

The key additional elements in Figure 11 below, i.e. those that are associated with the oxyfuel firing system are:

- The flue gas recycle loops from the flue gas clean-up train to the combustion system, and
- The oxygen supply pipework from the ASU to the biomass combustion system.

Looking in the first instance at the flow sheet on Figure 9 and the data in Table 16, i.e. for the base case without CO₂ capture, the gross power output is 83.4 MW_e and the total auxiliary power consumption is around 7.6 MW_e, or around 9% of the gross generation. The net power output is therefore 75.8 MW_e, equivalent to a net electrical efficiency of 36%, on a LHV basis.

The power consumption of the air separation unit, and the CO₂ capture and compression system is also significant, at around 27.1 MW_e, as indicated in Table 16. The overall effect is that the net electrical efficiency of the power plant with CO₂ capture and compression is only around 23%, i.e. around 13% lower than the case without CO₂ capture.

Table 17: Flow and temperature data of 75 MWe dedicated biomass (woodchips) combustion without and with CO₂ capture by oxyfuel combustion.

75 MW _e dedicated biomass power plant							
	without CO ₂ Capture			CO ₂ Capture with Oxyfuel Combustion			
	Flow Rate (t/h)	Temperature (°C)	Pressure (bar a)	Flow Rate (kg/s)	Flow Rate (t/h)	Temperature (°C)	Pressure (bar g)
Boiler Island							
Biomass	104	ambient	ambient	28.9	104.04		
Air	286	ambient	ambient				
Oxygen to Primary Flue Gas Recycle				5.5	19.8	150	1.01
Oxygen to Secondary Flue Gas Recycle				15	54	150	1.01
Primary Flue Gas Recycle (including O ₂ addition)				37.5	135		0.978
Secondary Flue Gas Recycle to boiler (including O ₂ addition)				86.7	312.12		0.94
Flue Gas out of boiler to GGH				101	363.6	348.5	0.906
				55.4	199.44	348.5	0.895
Bottom Ash	0.2	ambient	ambient				
Electrostatic Precipitator (ESP)							
Flue Gas to ESP				173.6	624.96	212.7	0.901
Flue Gas out of ESP				176.8	636.48	209.8	0.899
Fly Ash	0.9	ambient	ambient	0.3	1.08		
Flue Gas Heat Recovery							
Flue Gas out of Heat Recovery				176.8	636.48	130	0.897
Secondary Flue Gas Recycle				77.8	280.08	130	0.92
GGH							
Flue Gas to Flue Gas Condensor				99	356.4	95.2	0.893
Flue Gas Condensor							
Flue Gas out of fluegas condensor				78.5	282.6	30	0.895
Primary Flue Gas Recycle				43.1	155.16		0.92
Flue Gas to CO ₂ Purification and Comp.				35.38	127.368		
Flue Gas to Stack	469	90	1.005				

High level process control philosophy

The process control of a fluidised bed biomass power plant in response to grid demand is very complex but is well understood. In terms of the control of the heat input to the boiler, the biomass feeder speed will respond to a demand from the boiler master controller. The combustion air flow

rate is controlled to maintain the flue gas oxygen concentration set point at the boiler outlet, and the forced draft and induced draft fans are linked to ensure a small negative pressure in the furnace.

The operation of the oxyfuel firing system does introduce additional complications to the combustion control system and these are the subject of significant development work at pilot and demonstration scale projects planned over the next few years. This work is particularly concerned with the performance of the combustion equipment under oxyfuel conditions, and the control of the system under transient conditions and when switching between air firing and oxyfuel firing conditions.

The operation of the ASU will impose significant limitations on the flexibility of operation of the power plant. The precise nature of these limitations will vary depending on a number of site specific factors.

Table 18: Capital cost estimates (2009 €) for biomass firing in a BFB with and without carbon dioxide capture by oxyfuel combustion.

75MWe BFB Power Plant			
		without CO₂ Capture	CO₂ Capture with Oxyfuel Combustion
Plant Performance			
Gross Output	MWe	83.4	84.6
Net Output	MWe	75.8	48.9
Efficiency and Emissions			
Gross Electrical Efficiency (LHV)	%	39.6	40.1
Net Electrical Efficiency (LHV)	%	36.0	23.2
Economic Performance			
Capital Costs			
Storage and Handling of Solid Materials	£M	17.4	17.4
Boiler Island and Flue Gas Treating	£M	80.2	80.7
Power Island	£M	23.8	21.3
Utilities & Offsites	£M	22.2	30.1
ASU	£M	0.0	53.0
CO₂ Compression and Drying	£M	0.0	10.7
Total Installed Costs	£M	143.5	213.2
Land Purchases; Surveys and Fees (10 %)	£M	14.4	21.3
Contingency (10%)	£M	14.4	21.3
Project development costs	£M	14.4	21.3
Total Investment Cost	£M	186.6	277.2
Specific Investment Costs	£/kWe	2.461	5.674

Environmental performance

As described in some detail under Case Study 3, dedicated biomass power plants at industrial scale are large stationary sources of a number of important pollutant species, and are subject to a high degree of control over their environmental performance by the regulators. This includes continuous emissions monitoring and reporting of the key gaseous and gas-borne pollutant species, and regular

monitoring and reporting of other releases to air, land and water. The emissions standards for power plants are set at European Commission level and are subject to regular review.

The specific regulations that may apply to a particular biomass boiler are described in general terms under Case Study 3, and these are largely dependent on the nature of the biomass fuel diet. If waste materials are to be fired, then the regulations are much more demanding than for clean biomass materials. The operation of the oxyfuel firing system does not involve the use of any additional reagents and will not result in any significant increases in the levels of releases of any of the prescribed pollutant species from the power plant.

Capital and operating cost estimates

The capital cost estimates for Case Study 4, i.e. for the cases involving the dedicated fluidised bed biomass power plant, with and without CO₂ capture by oxyfuel firing, are provided in Table 18.

It is also clear from the estimated capital cost data for the two cases that the basic capital costs for the fuel handling system, the boiler and the power train are of the same order for the cases with and without CO₂ capture. The additional capital costs associated with the oxyfuel firing and CO₂ processing system, principally for the ASU, and the CO compression/drying plants are substantial, and add about 28% to the total investment costs of the power plant.

Table 18 also shows that, when the reduction in the electrical output of the power plant, associated with the operation of the carbon dioxide capture and compression/drying systems, is taken into account, the specific investment costs increase from £2.462 MW_e⁻¹ to £5.674 MW_e⁻¹, an increase of around 130%. The annual operating and maintenance (O&M) cost estimates for the two cases, i.e. for the 75 MW_e dedicated biomass power plant with and without CO₂ capture, are listed in Table 19. The fixed and variable O&M costs for the two cases are fairly similar. There are some additional maintenance costs associated with the ASU and the CO₂ compressors etc, but this is fairly modest. As for the other Case Studies described above, the O&M costs can also be expressed in terms of the cost per tonne of CO₂ captured, however this will depend on the assumptions that are made about the availability and performance of the power plant and the CO₂ capture systems. There is sufficient process and cost information provided in the tables above to allow these calculations to be carried out.

Overview of system performance and further development requirement

As stated above, the combustion of a wide range of biomass materials in dedicated power stations is a fully commercial technology with a large number of plants operating worldwide and a number of experienced suppliers. An overall TRL value of 8 to 9 would appear to be appropriate for this plant component. Oxyfuel firing technology has now been demonstrated at large pilot scale (TRL 5 to 6). The major emphasis of future development is now on full scale boiler demonstration projects to provide detailed engineering and design information for scale-up to fully commercial industrial plants. Overall, fully integrated, dedicated biomass oxy-fuel combustion plants are only currently at TRL 5. By 2020, it is thought likely that the TRL could have increased to 6-7.

There is significant development effort on the use of membrane separation techniques for oxygen separation. There is also significant R&D effort on boiler tube corrosion issues because of the potential increase of the concentrations of CO₂, H₂O, sulphur oxides and HCl in the flue gases in contact with the high temperature boiler surfaces within the flue gas recirculation loop. Overall, fully

integrated, dedicated biomass oxy-fuel combustion plants are only currently at TRL 5. By 2020, it is thought likely that the TRL could have increased to 6.

Table 19: Operation and maintenance cost estimates (2009 €) for a 75MWe dedicated biomass power plant with and without carbon dioxide capture by oxyfuel combustion.

75MWe BFB Boiler			
		without CO ₂ Capture	CO ₂ Capture with Oxyfuel Combustion
Operation and Maintenance Costs (O&M)			
Fuel Handling, Milling, Boiler Island, Power Island	£M/yr	4.9	4.8
ASU, CO ₂ Compress., and Drying	£M/yr	0.0	1.2
Common Facilities (Utilities, Offsite, etc.)	£M/yr	0.4	0.5
Labour	£M/yr	5.5	5.5
Adm./gen overheads	£M/yr	1.7	1.7
Fixed O&M Costs	£M/yr	12.4	13.7
Fuel			
Biomass (dry basis)	£M/yr	20.5	20.0
Auxiliary Feedstock			
Make-up water	£M/yr	0.0	0.0
Solvents			
Chemicals	£M/yr	0.0	0.1
Waste Disposal	£M/yr	0.3	0.4
Variable O&M Costs	£M/yr	20.8	20.5
Total O&M Costs	£M/yr	33.3	34.3

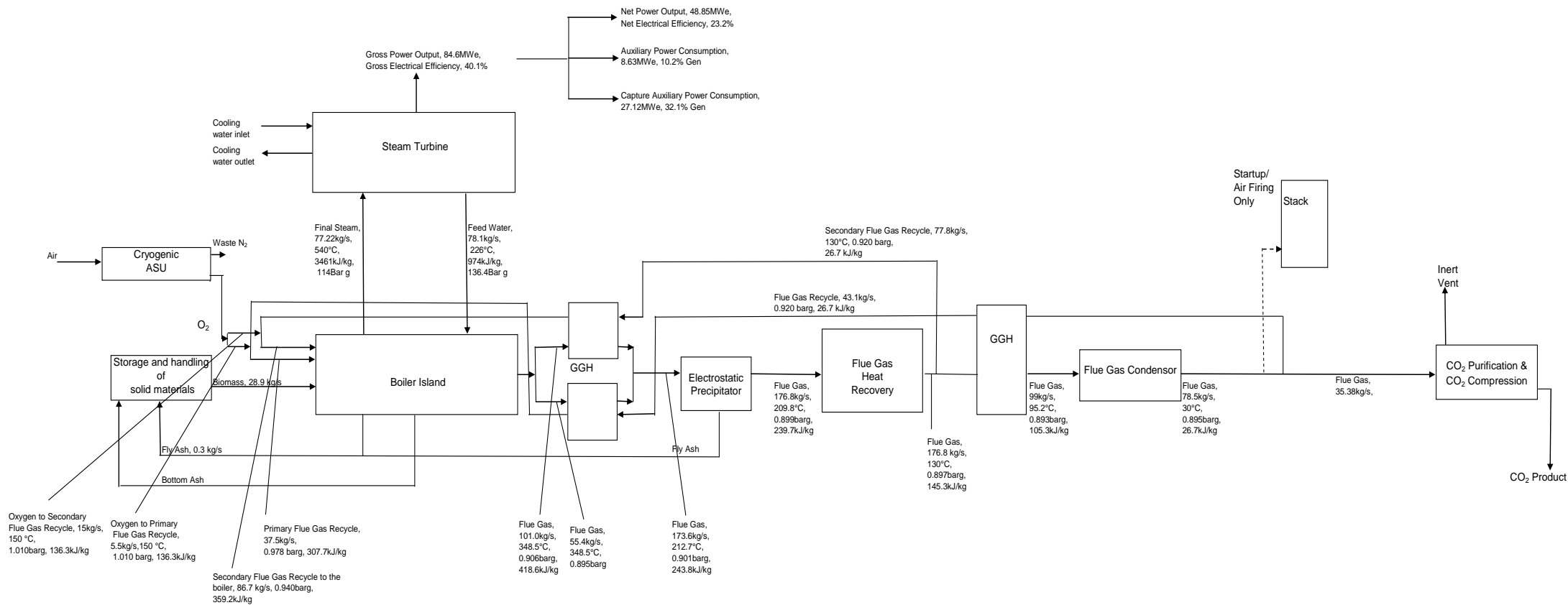
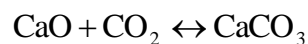


Figure 9: Flow diagram of biomass firing in a 75MWe BFB boiler with CO₂ capture by oxyfuel combustion.

CASE STUDY 5 Biomass combustion, with CO₂ capture by post-combustion carbonate looping

In this Case Study, the calcium carbonate looping plant is fitted to the flue gas outlet of a conventional biomass fired boiler with an associated steam turbine power train. In the carbonation unit, the CO₂ in the boiler flue gas reacts with the Ca-based sorbent to form calcium carbonate. The flue gas is then exhausted to atmosphere after suitable cleaning. The heat generated by the carbonation reaction is recovered in the form of steam and this is supplied to the steam circuit of the biomass boiler.

The Ca-based sorbent is cycled between the oxide and carbonate according to the following chemical reaction:



The CO₂ is released and the CaO regenerated in the calciner unit, which is heated by the combustion of a secondary fuel in oxygen supplied from an Air Separation Unit. The calciner is fluidised by the combustion gases arising from the combustion of the coal and the CO₂ released by the calcination. This moderates the temperature in the calciner from that expected if the coal were burnt in pure oxygen. Coal has been chosen as the fuel for the calciner in this case to permit operation and the appropriate temperature of around 900°C. With some biomass fuels there may be a risk of contamination of the sorbent by alkali metal salts, which can result in defluidisation and slumping of the fluidised bed. The exhaust gases from the calciner unit comprise largely CO₂ and water, which are sent for further processing, after suitable cleaning.

Status of development and technology readiness level

As stated above, the firing of biomass in a dedicated power station is a fully commercial technology. An overall TRL value of 8-9 currently, and TRL 9 by 2020, would appear to be appropriate. The current TRL level of Ca looping for this type of application is between 4 (component and/or partial system validation in a laboratory environment) and 5 (Component and/or partial system validation in a relevant environment).

Some of the key components (e.g. CFB combustion systems and boilers) have been validated in a working environment, though not in precisely the same setting as for Ca looping. Laboratory scale and small rig tests have been ongoing since 2003 in Canada (75 kW_{th}, semi-continuous), and more recently in Germany (10 kW_{th}, continuous) and Spain (30 kW_{th}, continuous dual CFB).

Overall, it is anticipated that the TRL of the Ca looping technology will be 5-6 by 2020, and that the application to biomass combustion systems would be at the same or similar levels.

Process Flow diagram

The outline flow diagram of the carbonate looping plant is presented in Figure 12. The conventional biomass fired side of the plant is based on the design in Figure 4, above. As stated above, the carbonate looping system comprises two separate circulating fluidised bed reactors, i.e. for carbonation and calcination. In the carbonator, the CaO reacts with the CO₂ in the boiler flue gases to form CaCO₃. This reaction is performed at temperatures around 650°C. The carbonation of CaO is

exothermic and the heat released from the carbonation reaction will be recovered in a system of heat exchanger banks both in the reactor vessel and in the exhaust gas cooling system. The steam produced will be supplied to the biomass boiler steam circuit. It has been assumed in the process calculations that the capture efficiency of the CO₂ in the carbonation unit will be around 90 %, which is within the limit imposed by the chemical equilibrium described above, at this temperature.

The level of CO₂ capture was calculated by mass balance using typical literature values for the conversion in the carbonation and calcinations units, under the relatively modest assumption that the reactions are fast. The inclusion of a detailed reaction model would provide a more accurate calculation at the expense of a substantial increase in the complexity of the model. However, this is not currently possible due to the limited availability of relevant kinetic data.

Table 20: Performance parameters for the carbonate looping system.

Overall efficiency before capture (%)	43.9
Overall efficiency after capture (%)	40.4
CO ₂ capture efficiency (%)	90
CO ₂ purity after compression (%)	96.6
Total inventory of calcium carbonate carrier (te)	128
Average rate of circulation between calciner and carbonator (te/s)	0.667
Rate of fuel consumption (kg/s, dry basis) wood/coal	18.3/9.7
Oxygen from ASU (kg/s)	23.0
Residence time of solid for calciner/carbonator (s)	43/120

The carbonated sorbent is then sent to the calciner unit, in which all of the captured CO₂ is released at a temperature of 900°C. This is an endothermic process and the heat required is provided by burning a secondary fuel. This fuel is burned at close to a stoichiometric ratio of 1 with oxygen, provided by an air separation unit (ASU). For this Case Study, the fuel is assumed to be fully combusted to CO₂, H₂O, N₂ and SO₂, and the flow rates of fuel and O₂ are controlled such that the heat release from the combustion process is equal to the heat required for calcination, with both processes operating at a temperature of 900°C. The product CO₂ is then cooled and cleaned before being sent for further processing. The heat recovered is supplied to the biomass boiler steam circuit.

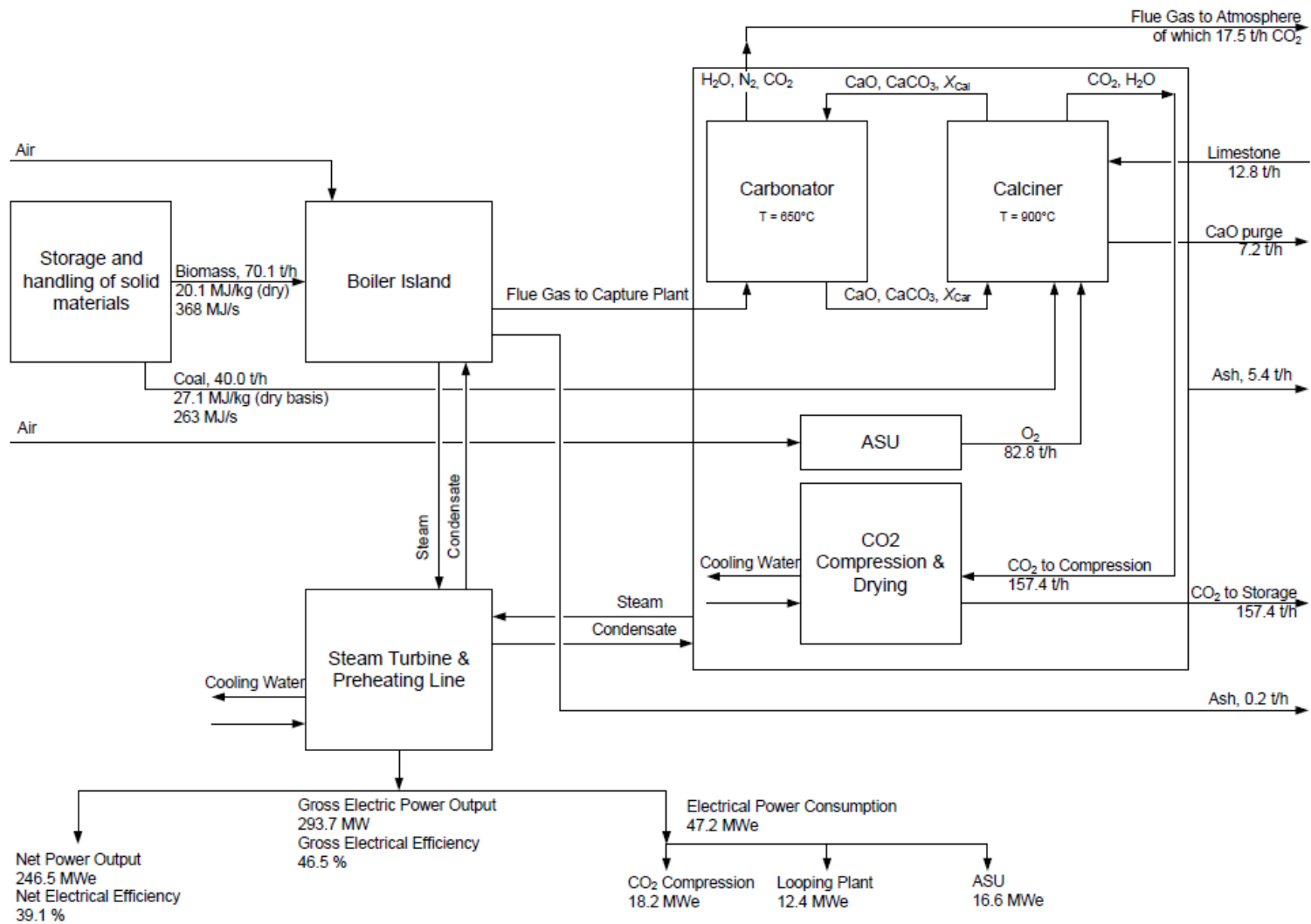


Figure 10: Flow diagram of the Calcium Looping Process.

The selection of the secondary fuel for firing in the calcinations unit is a key decision. This fuel could be gaseous, liquid or solid. The key requirements are that it should be reactive and capable of combustion at low excess oxygen levels. It should have low ash, sulphur and chlorine levels. For the purposes of this Case Study, the fuel used in the calciner is a bituminous coal. Depending on the sulphur contents of the primary and secondary fuels there will be significant formation of CaSO_4 , and a requirement to prevent the accumulation of sulphates in the recycled solids and to replenish the lime reactant. It should be noted, however, that the formation of CaSO_4 is not currently included in the process model. The key process parameters of the carbonate looping system are listed in Table 20, and the specific data for the carbonator and calciner units for the 300 MW_e power plant application are provided in Table 21. Furthermore, for the large scale (~300 MW_e) application considered here, a sensitivity analysis was carried out to investigate the impact of splitting the large carbonator and calciner reactors into four smaller reactors, on the performance and the economic parameters of the looping process. For the conditions studied, this sizing-split was not observed to be a sensitive parameter, hence the calculations were performed with a single reactor each, for carbonation and calcination. Note that the technology may be better suited to relatively small scales (~30-50MW) due to the physical size of the reactors and the high rates of circulation of the solids in the system.

Table 21: The sizing of the calciner and carbonator for a 300 MW_e plant.

	Carbonator		Calciner
U/Umf	6.9		5.2
Solids Inventory (te)	96		31
Circulation rate solids (kg/s as CaO)	667		
Solids mean residence time (s)	120		43
Bed Dia (m)	9.0		6.0
Expanded bed height (m)	5.2		3.6
Particle Dia (mm)	0.25		
Bed Volume (m ³)	833		251
Pumping Energy (MW)	10.2		2.2
Material	Purbeck Limestone		
Conversion CaO at exit	0.26		0.14
Cycles Life (h)	10		

The key process parameters of the steam plant, the CO₂ compression system and the ASU are listed in Table 22.

The steam cycle receives heat from

- (i) The biomass combustor and boiler, and
- (ii) The carbonation unit, where the flue gas from the biomass boiler reacts with CaO in the exothermic carbonation reaction, and
- (iii) The cooling systems for the flue gas and CO₂ streams prior to further processing.

The steam cycle is supercritical, with a boiler pressure of 300 bar, and maximum steam temperature of 600°C. It has a single stage of reheat at 50 bar, with a reheat temperature of 620°C. The IP turbine which follows the reheater has an outlet pressure of 10 bar, and is followed by an LP turbine which expands the steam down to the condenser pressure of 0.048 bar. The fuel used in the boiler was pine chip, which as a LCV (on dry basis) of 21.2 MJ/kg. At the exit from the boiler, the temperature of the flue gas was reduced to 130°C. No desulphurisation stage has been included, because the SO₂ will be captured by the CaO as solid CaSO₄.

The captured CO₂ stream is compressed to 108bar over a four-stage process. Prior to each compression stage, and after the final stage, the gas stream was cooled to 45°C and any water vapour in the stream condensed until the vapour pressure at 45°C was reached. The compression ratio of each stage was assumed to be 4, 3, 3 and 3, respectively, with each compressor having an isentropic efficiency of 0.9.

High Level Control Philosophy

The calculations undertaken in this study suggest the following:

- A calcium looping plant is likely to be most suitable for use as a base load electricity generation system because of the limitations on plant operating flexibility imposed by the ASU and the relatively slow response times for the carbonation/calcination loop to changes in boiler load that are expected. It is envisaged that a turndown level of 50-60% may be possible, depending on the details of the design of the reactors and the associated equipment.
- Monitoring the CO₂ from the carbonator (exothermic but equilibrium controlled) as well as offline sampling of the state of the Ca sorbent are regarded as two important parameters from process control point of view. For example, if the sorbent goes off-spec. then, the circulation rate between the two beds can be used as an effective control parameter. For the calciner, the amount of fuel and oxygen can be used as control parameters depending upon the feedback from the temperature signal.
- The potential to co-fire with coal and biomass in the calciner needs further consideration. Further, the above calculations assume that the carryover of char from the calciner to the carbonator will be small, on the basis that oxyfuel combustion is occurring, so that the bed carbon inventory will be small.
- Modelling is needed to consider the rate of heat transfer to the water and steam tubes and how these are matched into the steam cycle, particularly during start-up, shut down and transient conditions.

- As stated above, the control of the sulphur inventory within the system is important since the sulphur oxides will react irreversibly with the CaO to form gypsum.

Table 22: The process parameters for the steam cycle, CO₂ compressors and the ASU.

Steam Plant	
HP/IP/LP turbine pressure at inlet (bar)	291/47.5/10
Boiler/Reheater temperature (°C)	600/620
Condenser pressure (bar)	0.048
Isentropic efficiency of HP/IP/LP Turbine	0.92/0.94/0.88
Mechanical efficiency	0.996
Generator efficiency	0.985
Isentropic efficiency of Feed pump	0.7
Isentropic efficiency of Cooling Water pump	0.7
Cooling water pressure (bar)	2
Fractional pressure loss, economiser/superheater cold side	0.03
Fractional pressure loss, reheater cold side	0.05
CO ₂ Compression	
Number of compression stages	4
Compression ratio of each stage	4/3/3/3
Temperature of CO ₂ stream after cooling (°C)	45
Isentropic efficiency of each stage	0.8/0.8/0.8/0.8
Air Separation Unit	
Required work input of ASU (kWh/tonne O ₂)	200
Supply temperature of O ₂ (°C)	25

Table 23: Capital cost estimate for the 300 MW_e biomass combustion plant with post combustion CO₂ capture by carbonate looping.

Item	£M, 2011
Storage and handling of solid materials	41.1
Boiler island	220.5
Power island	76.5
CO ₂ compression plant	31.4
Air separation unit	136.2
Calciner	45.7
Carbonator	92.8
Total installed CAPEX	644.2
Operation and utilities (% of TIC)	32.2
Civils and land costs (% of TIC)	64.4
Project Development Costs (% of TIC)	32.2
Contingency (% of TIC)	64.4
Total investment cost	837.4
Specific Investment Cost (£M/MWe)	2.79

Environmental performance

The calcium looping plant is fitted to the flue gas outlet of a conventional biomass fired power plant. The environmental performance would be expected to be equivalent to the biomass fired power plant in Case Study 2, subject to the following key points:

- The calcium looping provides a robust method of CO₂ capture.
- Any SO₂ or acid gases (for example HCl) in the flue gas would react with the CaO. The CaO would be in large excess so this would not be expected to unduly reduce the carbon capture capacity of the system. A purge would be required to prevent accumulation of reacted material, for example gypsum (CaSO₄ formed from the SO₂) in the calcium loop. The purge could remove a mixture of CaO and gypsum. This mixture would be a valuable raw material for cement manufacture and could potentially be sold on.
- The SO₂ emissions calculated using equilibrium assumptions with CaSO₄ for the typical temperature range incurred in the carbonator reveals that the SO₂ concentrations are small, usually < 1 ppm by volume. Equilibrium calculation means that the reactor operation might allow a bigger slip of SO₂, however, most of the sulphur is captured by the carbonate.
- The level of NO_x emissions are a key unknown. They are expected to be lower than for the conventional biomass fired power plant boiler because the temperature in the calcium loop is lower than in conventional boilers. This would result in less NO_x production by nitrogen

fixation. However, the fuel-NO_x chemistry (the interaction with CH type radicals from the fuel) is unknown.

Capital and operating cost estimates

The capital costs of a 300 MW_e plant are listed in Table 23. Of the total installed plant costs of £390 million, the costs of the carbon capture plant and ASU are around £198 million or of the order of 51% of the total.

The specific investment costs of the 300 MW_e biomass power plant with carbon capture by carbonate looping are of the order of 2.79 £M/ MW_e.

The operating and maintenance (O&M) cost estimate for a 300 MWe dedicated biomass power plant with CO₂ capture by carbonate looping is presented in Table 24. The total annualised O&M costs comprising the variable and fixed costs, are of the order of 174.72 £M/year.

The cost calculations are for a new-build plant. In principle, a calcium looping system could be retrofitted to an existing biomass power plant, but the feasibility and costing would have to be assessed on a case-by-case basis. The retrofit project would obviously involve a significant increase in the power output from the power plant, with the additional capital costs of the new water and steam cycles and the additional electricity generation capacity.

Table 24: Operating and maintenance cost estimate for the 300 MW_e biomass powerplant with CO₂ capture by carbonate looping.

Variable Costs	Usage	£M/yr
1. Wood fuel	4.91×10^8 kg/yr	68.69
2. Coal	2.61×10^8 kg/yr	22.41
3. Limestone	9.51×10^7 kg/yr	1.65
3. Spent limestone (credit)	5.33×10^7 kg/yr	-1.65
4. Wood fly ash disposal	1.06×10^6 kg/yr	0.0021
5. Coal ash disposal – assume sold with spent sorbent		0
6. Cooling water make-up	9588 kg/s	51.4
Variable costs		142.51
Maintenance and Labour		25.77
Insurance		6.44
Fixed costs		32.21
Total O&M costs		174.72

Technology readiness level and development requirements

As stated above, it is anticipated that the TRL of the Ca looping technology for CO₂ capture will be at level 5-6 by 2020, and that the application to biomass combustions systems would be at the same or similar levels.

There are a number of areas where further development work on the current process and economic models of the Ca looping systems needs to be carried out, viz:

- More detailed process modelling of the reactors is needed to accompany the pilot and larger scale test work to obtain more accurate solid residence times and reactor volumes. In particular, realistic kinetic results for the sorbent needs to be obtained, using representative fuels at pilot scale. The models also need to consider the rate of heat transfer from the carbonator to the water and steam tubes, and how these are matched into the steam cycle.
- The rate of circulation is high, and careful consideration needs to be given to the design of practical transfer systems for the movement of the hot solids between the reactors.
- The physical strength and abrasion resistance of the limestone and carbonate reagent are poor, and the sulphation of the reagent will be an issue in most systems. The inclusion of purge and replenishment flows of sorbent will need consideration. This will affect the overall average reactivity of the solid reagent, based on the distribution of the age of the cycled solid.
- The design and the control of the heat recovery systems from the carbonate looping reactors and the boiler plant will require careful consideration.
- Much more detailed and reliable costing information will be needed for the fluidised bed reactors, when the process parameters for these are better defined.
- The process and economic models are sensitive to whether or not coal has to be used in the calciner: this is an area needing further experimental investigation.

CASE STUDY 6 Dedicated biomass chemical-looping-combustion using a solid oxygen carrier

The chemical looping combustion plant evaluated in this Case Study uses a Cu-based oxygen carrier, cycled between the states of CuO and Cu₂O. In this case, it has been assumed that the oxygen carrier is not reduced to Cu. One advantage of this choice of oxygen carrier is that under certain conditions, CuO dissociates to Cu₂O and gaseous O₂ via the following “uncoupling” reaction



In this way, a chemical looping oxygen uncoupling (CLOU) system is adopted, whereby the fuel is fed directly into the fuel reactor without prior gasification, and combusted using gaseous oxygen. The net reaction in the fuel reactor is generally exothermic, and so heat is extracted from the reactor to raise steam for power generation.

A further advantage of this system is that conversion of the fuel is rapid, to the point where it can be assumed that CuO reduction reaction is the rate-limiting step within the fuel reactor. The uncoupling of oxygen is favoured at higher temperatures, and a fuel reactor temperature of 950°C was chosen. Laboratory results suggest that when CuO is supported on alumina (~ 20 wt% Al₂O₃, 80 wt% CuO), it is very stable for cycles of release and uptake of oxygen at high temperatures up to ~970°C. The system is also non-toxic and relatively cheap compared to other possible oxides capable of this type of reaction.

After depletion of oxygen in the fuel reactor the oxygen carrier is subsequently re-oxidised in the air reactor. This is always an exothermic reaction, and is carried out at 850°C. This case study is concerned with a 300MW_e power plant firing a wood chip fuel, with a chemical looping combustion system of the type described above.

Status of development and technology readiness level

Recent years have seen significant advances made in both reactor design and scale. The current TRL of chemical looping combustion technology is considered to be 4. Based on the recent progress and the projected pilot scale and demonstration plants, the technology is expected to reach a TRL of 5 to 6 by 2020.

Process Flow Diagram

An overall flow diagram of the combustion arrangement is shown in Figure 11. In the fuel reactor, it is assumed that the fuel is fully converted to CO₂, H₂, SO₂, N₂ and ash, and that the oxygen provided by the oxygen carrier is 5% in excess of the stoichiometric quantity required for complete combustion of the fuel. The concentration of oxygen in the fuel reactor is limited to 40 mol% by the addition of CO₂ as a fluidising gas, as shown in Figure 11. The level of CO₂ capture was calculated by mass balance using estimated values for the CO₂ removal based on (currently) unpublished experimental data. More detailed modelling is not currently possible due to the limited availability of kinetic data.

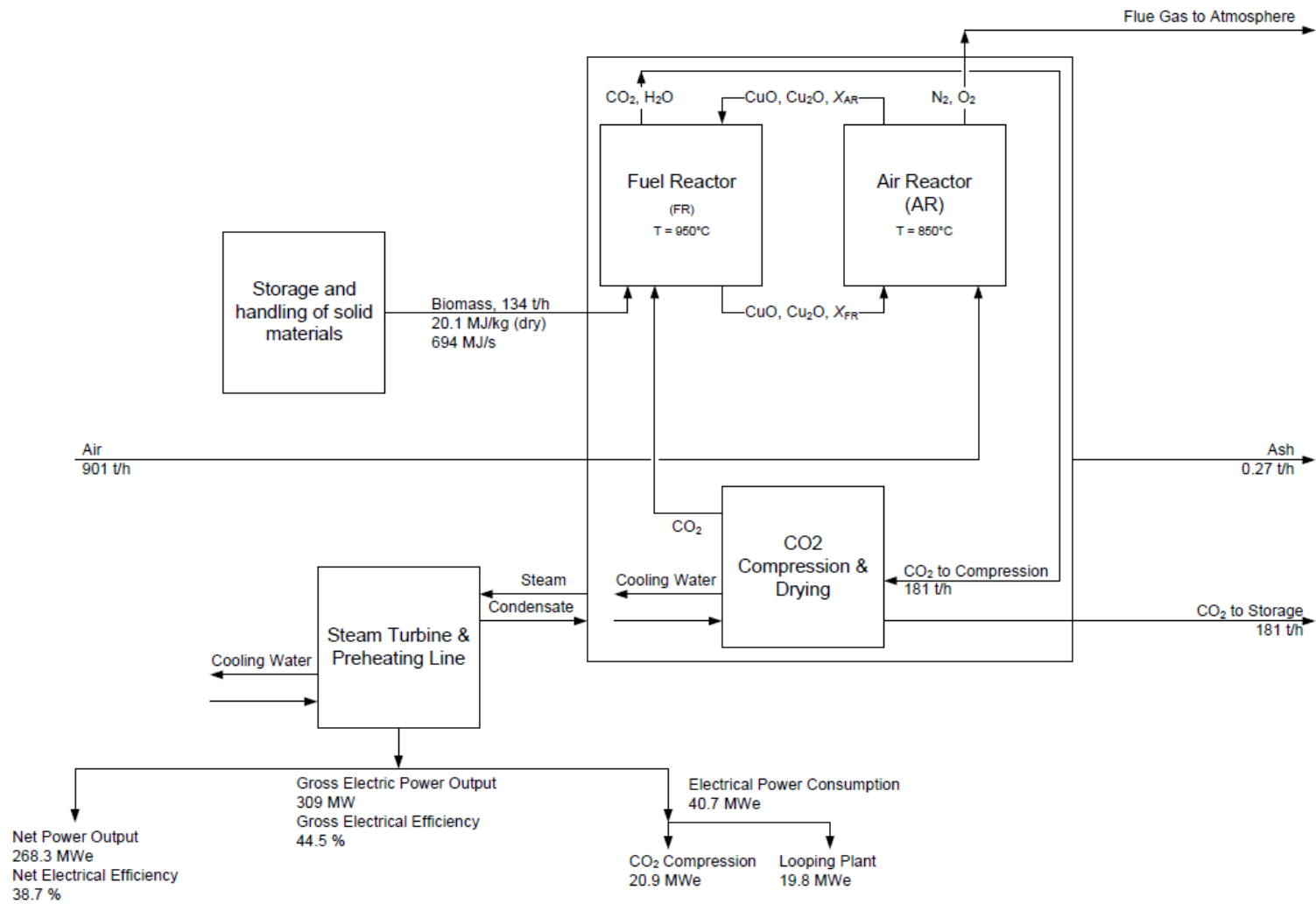


Figure 11: The basic flow diagram of the CuO chemical looping cycle.

In the air reactor, 20 % excess air on a molar basis is supplied to restore the oxygen carrier to the state entering the fuel reactor. In the process model, two key process variables in the system were controlled:

- The mole fraction of Cu contained within CuO leaving the fuel reactor
- The difference in mole fraction between the two reactors.

The fuel used in the boiler/fuel reactor is pine chip, which has a LCV (on a dry basis) of 21.2 MJ/kg. In the CLC plant, combustion of the fuel is carried out within the fuel reactor via a stoichiometric quantity of oxygen released by the uncoupling reaction at a temperature of 950°C. The fuel is assumed to be fully converted to CO₂, H₂O, N₂ and SO₂. At the exit from the boiler, the temperature of the flue gas was constrained to 130°C; currently no desulphurisation stage is included because the experimental evidence suggests that the SO₂ will be largely emitted in the off-gas from the fuel reactor and so be present in the CO₂ stream. The SO₂ will have to be removed, therefore, during the preparation of the CO₂ for sequestration. Further experimental work is, however, needed on the role of sulphur in these chemical looping systems.

The key process parameters and the results of the sizing calculations are presented in Tables 25 and 26. It should be noted that the chemical looping technology may be better suited to smaller scales (order of magnitude smaller ~30-40MW_e) due to the physical size of the reactors and the volume of solids in the looping cycle.

Table 25: Performance parameters for the chemical looping combustion system.

Overall efficiency before capture (%)	44.6
Overall efficiency after capture (%)	41.0
CO ₂ capture efficiency (%)	100
CO ₂ purity after compression (%)	96.6
Carrier composition (fully oxidised form)	17.5 wt% Al ₂ O ₃ / 82.5 wt% CuO
Total inventory of reactive, oxidised solid (tonne CuO basis)	158.5
Carrier oxidised particle density, kg/m ³ (assumed particle dia. 1.8 mm)	3000
Average rate of circulation between air and fuel reactors (te oxidised carrier basis/s)	1.35
Rate of fuel consumption (kg/s, dry basis)	34.5
Rate of air flow into air reactor (kg/s)	250.4
Residence time of solid for air/fuel reactors (s)	42.7/75.4
Residence time of gases for air/fuel reactors (s)	0.6/1.5

A list of the key steam cycle parameters and the CO₂ Compression system are presented in Table 27. The steam boiler was assumed to be supercritical, with a boiler pressure of 300 bar, and maximum temperature of 600°C, and a single stage of reheat at 50 bar, with a reheat temperature of 620°C. The IP turbine which follows the reheater has an outlet pressure of 10 bar, and is followed by an LP turbine which expands the steam down to the condenser pressure of 0.048 bar. The isentropic efficiencies for the HP, IP and LP turbines were assumed to be 0.92, 0.94 and 0.88, respectively, whilst that of the feed pump was taken to be 0.7. Fractional pressure losses of 3 % and 5 % were assumed for the economiser/superheater and reheater, respectively. No feedheating or additional feedpumps have been considered, but should be included at a later stage.

Table 26: Sizing of chemical looping air and fuel reactors for a 300 MW_e Plant.

	Air reactor		Fuel Reactor
U/U _{mf}	9.7		9.5
Bed height minimum fluidisation	0.5		1.6
Bed height, operational	2.4		6.4
Solids Inventory (te)	58		102
Circulation rate solids (kg/s)	1348		
Solids mean residence time (s)	43		75
Bed dia (m)	9.5		7.0
Particle dia (mm)	1.8		
Bed volume (m ³)	458		582
Pumping Energy (MW)	6.8		13
Temperature (K)	1123		1223

The captured CO₂ stream was assumed to be compressed to 108 bar over a four-stage process. Prior to each compression stage, and after the final stage, the gas stream was cooled to 45°C and any water vapour in the stream condensed until the vapour pressure at 45°C was reached. The compression ratio of each stage was assumed to be 4, 3, 3 and 3, respectively, with each compressor having an isentropic efficiency of 0.9.

Table 27: A listing of the key process parameters of the steam cycle and the CO₂ compression system.

Steam Plant	
HP/IP/LP turbine pressure at inlet (bar)	291/47.5/10
Boiler/Reheater temperature (°C)	600/620
Condenser pressure (bar)	0.048
Isentropic efficiency of HP/IP/LP Turbine	0.92/0.94/0.88
Mechanical efficiency	0.996
Generator efficiency	0.985
Isentropic efficiency of Feed pump	0.7
Isentropic efficiency of Cooling Water pump	0.7
Cooling water pressure (bar)	2
Fractional pressure loss, economiser/superheater cold side	0.03
Fractional pressure loss, reheater cold side	0.05
CO ₂ Compression	
Number of compression stages	4
Compression ratio of each stage	4/3/3/3
Temperature of CO ₂ stream after cooling (°C)	45
Isentropic efficiency of each stage	0.8/0.8/0.8/0.8

High Level Control Philosophy

The scale-up of Cu-looping systems is still in its early stages, but the calculations undertaken in this study suggest the following:

- Because of the time required to bring the fluidised bed reactors and the looping cycle to temperature and to steady state operation, it is likely that the chemical looping combustion system is likely to most suitable for baseload operation, with relatively few shutdowns and start-up cycles. A reasonable level of turndown might be possible depending on the design parameters of the dual fluidised bed system.
- The large inert inventory in the form of the CuO support material (e.g. alumina) and the enhanced mixing in the fluidized bed are the two main control parameters that can be used to avoid the hotspots. However, in the event of a thermal runaway, reducing the fuel feed and increasing the flow rate of the CO₂ stream (fluidization gas) will form an important part of the temperature control strategy.
- Significant further consideration of the integration of the heat exchangers in the chemical looping system is required.
- The control of the sulphur inventory in the system represents a key process unknown.

Table 28: The capital cost estimate for a 300 MW_e biomass power plant with CO₂ capture by chemical looping combustion technology.

Item	£M, 2011
Storage and handling of solid materials	41.1
Boiler island	220.5
CO ₂ compression and drying plant	31.4
Power island	76.5
Air reactor (458 m ³)	64.8
Fuel reactor (581 m ³)	74.9
Total installed CAPEX	509.2
Operation and utilities (% of TIC)	25.5
Civils and land costs (% of TIC)	50.9
Project Development Costs (% of TIC)	25.5
Contingency (% of TIC)	50.9
Total investment cost	661.9
Specific investment cost (£M/MW_e)	2.21

Environmental performance

The Cu-looping plant fires biomass and its environmental performance would be expected to be equivalent to the biomass fired power plant in Case Study 2, subject to the following key points:

- The Cu-looping provides a potentially interesting method of CO₂ capture.
- The level of SO₂ emissions would be low because the plant is not co-firing.
- The level of NO_x emissions are a key unknown. They are expected to be lower than for the conventional biomass fired power plant boiler because the temperature in the calcium loop is lower than in conventional boilers. This would result in less NO_x production by nitrogen fixation. However, the fuel-NO_x chemistry (the interaction with CH type radicals from the fuel) is completely unknown.

Capital and operating cost estimates

The capital cost estimate for the 300MW_e biomass power plant with CO₂ capture by chemical looping combustion is presented in Table 28. The specific investment costs of the 300 MW_e biomass power plant with carbon capture by chemical looping technology are of the order of 2.21 £M/ MW_e. The O&M cost estimate for the 300 MW_e power plant are listed in Table 29. The total annualised O&M costs comprising the variable and fixed costs, are of the order of 193.36 £M/year. The cost calculations are for a new-build plant. Copper looping is not retro-fit technology and would require a dedicated plant to be built for the purpose of power generation with CO₂ capture.

Table 29: The O&M cost estimate for the 300MW_e biomass power plant with CO₂ capture by chemical looping combustion technology.

Variable Costs	Usage	£M/yr
1. Wood fuel	8.29 × 10 ⁸ kg/yr	116.0
2. Oxygen carrier (new)	1.19 × 10 ⁶ kg/yr	4.74
3. Spent carrier (credit)		-4.22
4. Fly ash disposal	1.78 × 10 ⁶ kg/yr	0.00356
5. Cooling water make-up	9588 kg/s	51.4
Variable costs		167.9
Maintenance and Labour		20.37
Insurance		5.09
Fixed costs		25.46
Total O&M costs		193.36

Overview of system performance and further development requirement

As stated above, the current TRL of chemical looping combustion technology is considered to be 4, and the technology is expected to reach a TRL of around 5 to 6 by 2020.

There are a number of areas where further work needs to be carried out. The specific development areas associated with the process and economic modelling include:

- More detailed modelling of the air and fuel reactors to obtain more accurate solid residence times and reactor volumes. In particular, realistic kinetic results for the Cu-based carrier are required using representative fuels at pilot scale.
- The model also needs to consider the rate of heat transfer to the water and steam tubes and how these are matched into the steam cycle.
- A further major issue is that the rate of circulation is high and careful consideration needs to be given to the design of a solids transfer system.
- The impact of sulphur on the performance of the carrier needs further research.
- Revised costing information is needed for the air and fuel reactors, when these are better defined. Uncertainties in the current equipment costs are a possibly significant factor for the current specific CAPEX estimate.

- The potential to co-fire coal and biomass in this type of system needs further consideration. It has been assumed that the level of carryover of char from the fuel reactor to the air reactor will be low, on the basis that the reactivity of wood char is high. The bed carbon inventory will be small. With coal, however, the level of unburnt material will be significant, and a carbon stripper with the return of the carbon to the fuel reactor. The final inventory of carbon, and the requirement for a carbon recycling system is a key unknown associated with this technology when applied to biomass-coal co-firing.
- Consideration of different oxygen carrier materials. This may change the underlying assumptions for certain reactors; for example, the uncoupling reaction does not occur for an Fe-based oxygen carrier, and so gasification of char is likely to be the rate-limiting step in the fuel reactor.
- Potential alternative plant designs are of interest, e.g. the CLC-Combined Cycle, whereby the air and fuel reactors are pressurised and the exhaust gas streams are expanded through a turbine. Other considerations for the CLC plant might include separate gasification.
- A more detailed heat integration model, which restricts the components which can exchange heat with one another, is required.
- The degree of dust treatment at the outlet from the fuel reactor requires further investigation as it is a strong function of fuel, carrier formulation, fluidising conditions, rate of circulation *etc.* The attrition tendency of the carrier needs further investigation. Further investigations are also needed on trace sulphur and nitrogenous impurities arising from such a system, because of high cost of gas purification equipment if such systems are required.

Clearly there will also be significant development of the process through testing at pilot scale and demonstration projects to provide the test data and engineering information required for commercialisation of the technology.

CASE STUDY 7 Co-fired Integrated Gasification Combined Cycle (IGCC) plant with CO₂ capture by physical absorption

The Case Study reported in this section is for IGCC plants firing British bituminous coal and biomass in the form of wood pellets. Two plant sizes are considered, viz:

- the smaller system with a single gas turbine with a gross power output of 323 MW_e and
- the larger with two gas turbines with a total gross power output of 646 MW_e.

Two firing scenarios were considered, 100% coal and a mixture of 90% coal with 10% biomass on a lower heating value (LHV) basis. The CCS scenarios investigated were 0% CCS (no CCS), 90% CCS and 95% CCS. Each case was studied at a capacity factor of 85%. For this Case Study, the techno-economic calculations have been performed using the Integrated Environmental Control Model (IECM) developed at Carnegie-Mellon University, developed as part of a DOE project.

Status of development and technology readiness levels

The TRL of coal IGCC could be assessed as at least 8, if not 9. A large number of IGCC power plants already exist at utility scale, with power outputs ranging from 50 to 550 MW_e, although there are some remaining availability problems. The TRL in 2020 of the current coal IGCC designs will likely remain at 9. The co-firing of biomass materials in these plants is also a fairly common practice. In summary; fully integrated, co-firing IGCC plants combined with pre-combustion capture are only currently at TRL 5 to 6. By 2020, it is thought likely that the TRL could have increased to 7.

Process flow diagram

The process diagram for a generic IGCC plant using a cooling tower system and landfill waste disposal is presented in Figure 12a and the same system with the addition of a CCS Selexol system is depicted in Figure 12b. The fuel (i.e. coal, biomass or mixtures) is mixed with water to form a slurry, which is fed to the oxygen-blown gasifier and converted to syngas. The oxygen is produced in a conventional air separation unit. The slag generated in the gasifier, mainly ash and water, is collected and sent to landfill disposal.

If CCS is being employed (Figure 12b), the syngas is transferred to a water gas shift reactor which converts the CO into CO₂ to improve the efficiency of the CO₂ removal process. Sulphur in the form of H₂S and COS is removed from the syngas in the first Selexol unit, and is sent to a Claus plant and a Beavon-Stretford tail gas treatment unit for conversion to elemental sulphur which can be recovered and sold as a by-product.

The sulphur free syngas can then undergo a second stage of the Selexol process for the pre-combustion CO₂ capture. A system of CO₂ stripping is used to recover the solvent and collect the CO₂, which is then compressed and sent to storage.

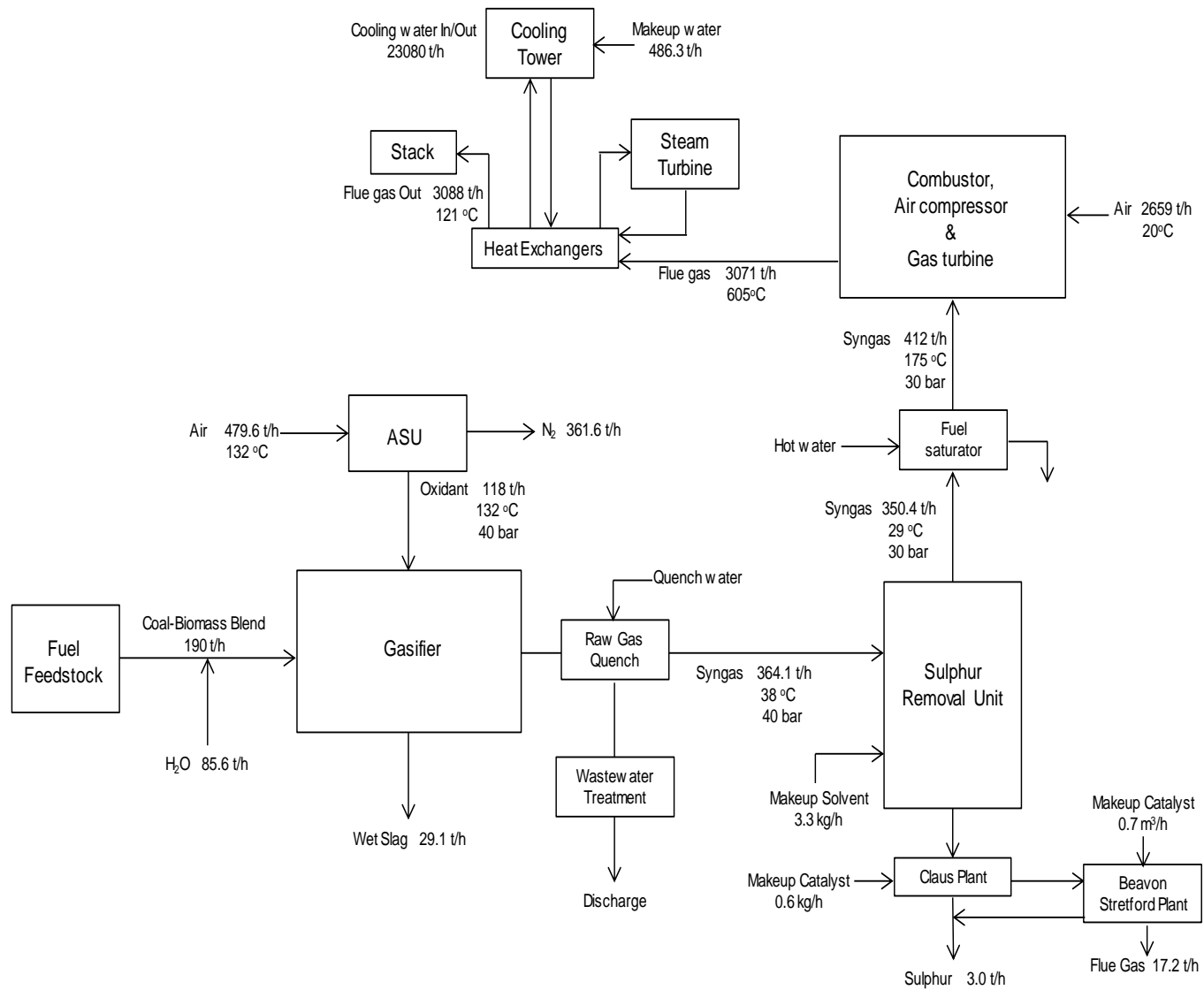


Figure 12a: The process flow diagram for the co-firing of biomass in a large coal-fired IGCC plant, without carbon dioxide capture.

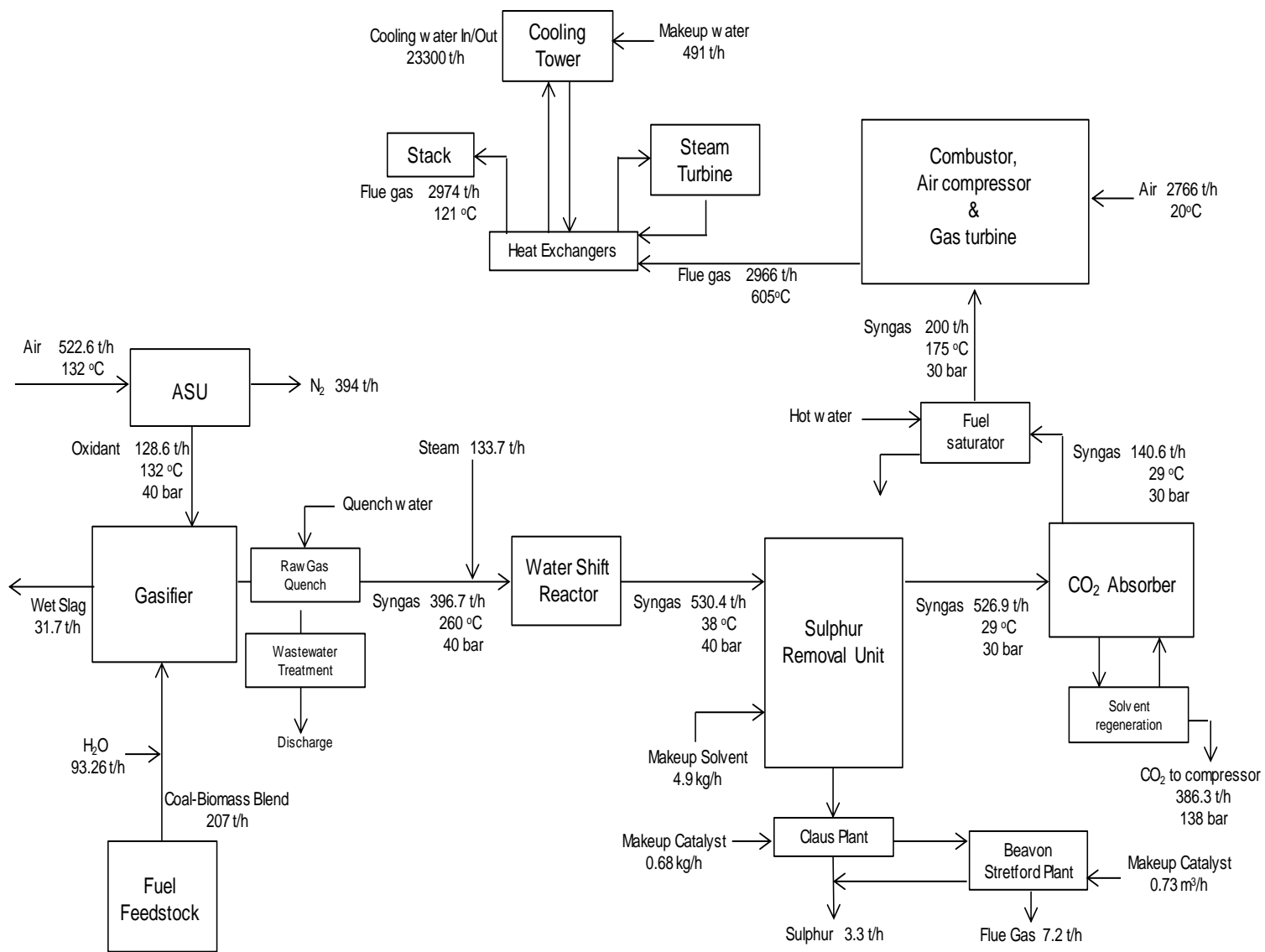


Figure 12b: The process flow diagram for the co-firing of biomass in a large coal-fired IGCC plant, with carbon dioxide capture by physical absorption.

The clean syngas now consisting nominally of pure hydrogen is fed to the combustor where it is mixed with air and burns in the Gas Turbine/ combined cycle unit.

The syngas compositions for coal and blend were calculated with the aid of the FactSage chemical thermodynamics package equilibrium module. The basis of the coal slurry was a mixture containing 65 wt% dry fuel and 35 wt% water; the fuel-oxygen ratio was specified by using a value of $\lambda = 0.33$ which is considered reasonable for oxy-gasification. For the coal / biomass slurry, a higher water content is necessary due to the hygroscopic and hydrophilic nature of biomass (61.1 wt% dry fuel and 38.9 wt% water). The biomass blend contains higher levels of fuel oxygen than coal but oxygen is still required for combustion in the gasifier; in addition, more oxygen is required to maintain the gasification temperature in consideration of the higher water content in the slurry. The result is that the syngas produced when co-firing tends to be more oxidised in comparison to the pure coal case with higher fractions of CO₂ and H₂O. This has an impact of a decrease in net plant efficiency for the co-firing case. The calculated compositions of the syngas are shown in Table 30.

Table 30: Syngas compositions derived from equilibrium calculations and input to the IECM model.

Syngas Component	Coal (vol %)	Blend (vol %)
Carbon Monoxide (CO)	45.32	39.24
Hydrogen (H ₂)	32.78	30.04
Methane (CH ₄)	2.440e-2	1.094e-2
Ethane (C ₂ H ₆)	0.0	0.0
Propane (C ₃ H ₈)	0.0	0.0
Hydrogen Sulfide (H ₂ S)	0.5740	0.4941
Carbonyl Sulfide (COS)	2.690e-2	2.189e-2
Ammonia (NH ₃)	3.000e-3	2.512e-3
Hydrochloric Acid (HCl)	5.700e-2	4.723e-2
Carbon Dioxide (CO ₂)	6.151	8.644
Moisture (H ₂ O)	13.45	20.01
Nitrogen (N ₂)	0.7160	0.6283
Argon (Ar)	0.8930	0.8601

The key process flows for the 323 and 646 MW_e gross IGCC plants with 0, 90 and 95% carbon dioxide capture levels are presented in Tables 31a and 31b, and the key temperatures and pressures are listed in Table 32. The energy balances for the 323 and 646 MW_e plants are listed in Tables 33a and 33b.

Table 31a: The flow rates (tonnes h⁻¹) for the 323 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels, at 85% capacity factor.

Stream	0% CCS		90% CCS		95% CCS	
	Blend	Coal	Blend	Coal	Blend	Coal
ASU air in	239.9	219.7	263.2	243.2	261.3	241.6
ASU oxidant out	59.02	54.08	64.76	59.86	64.3	59.45
Gasifier fuel in	94.98	86.7	104.2	95.98	103.5	95.35
Gasifier water in	42.8	30.7	46.96	33.98	46.64	33.75
Gasifier wet slag out	14.56	15.08	15.98	16.69	15.87	16.57
Gasifier syngas out	182.1	156.2	199.8	172.9	198.3	171.7
Water gas shift reactor steam in	-	-	67.33	68.87	66.86	68.4
Water gas shift reactor syngas out	-	-	267.1	241.8	265.2	240.1
Claus plant sulphur out	1.494	1.528	1.639	1.691	1.628	1.679
Beavon Stretford Plant flue gas out	8.581	6.68	3.623	3.738	3.598	3.711
SRU syngas out	175.2	151.2	265.4	240	263.4	238.3
CO ₂ absorber product	-	-	184.2	174.9	193.1	183.3
CO ₂ absorber syngas out	-	-	81.06	65.07	70.32	54.98
Combustor air in	1329	1342	13.72	1373	1383	1383
Fuel saturator hot water in	30.80	41.60	31.94	46.43	29.78	44.18
Gas turbine flue gas out	1535	1535	1484	1484	1483	1482
Stack flue gas	1541	1544	1488	1488	1487	1486

Table 31b: The flow rates (tonnes h⁻¹) for the 646 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels, at 85% capacity factor.

Stream	0% CCS		90% CCS		95% CCS	
	Blend	Coal	Blend	Coal	Blend	Coal
ASU air in	479.6	439.4	526.3	486.5	522.6	483.2
ASU oxidant out	118	108.1	129.5	119.7	128.6	118.9
Gasifier fuel in	190	173.4	208.5	192	207	190.6
Gasifier water in	85.6	61.39	93.89	67.96	93.26	67.49
Gasifier wet slag out	29.12	30.15	31.95	33.38	31.73	33.15
Gasifier syngas out	364.1	312.3	399.5	345.8	396.7	343.4
Water gas shift reactor steam in	-	-	134.6	137.7	133.7	136.8
Water gas shift reactor syngas out	-	-	534.2	483.5	530.4	480.2
Claus plant sulphur out	2.988	3.055	3.279	3.382	3.256	3.358
Beavon Stretford Plant flue gas out	17.16	13.36	7.247	7.474	7.196	7.423
SRU syngas out	350.4	302.5	530.7	479.9	526.9	476.6
CO ₂ absorber product	-	-	368.6	349.8	386.3	366.7
CO ₂ absorber syngas out	-	-	162.1	130.2	140.6	110
Combustor air in	2659	2683	2742	2744	2766	2766
Fuel saturator hot water in	61.6	83.1	64	92.8	59.4	88.3
Gas turbine flue gas out	3071	3069	2968	2967	2966	2965
Stack	3088	3083	2976	2975	2974	2972

Table 32: Temperatures and pressures for the IGCC-CCS system at 0, 90 and 95% CO₂ capture levels for both fuels and system sizes.

Stream	0% CCS		90%/95% CCS	
	T / °C	P / MPa	T / °C	P / MPa
ASU air in	132.2		132.2	
ASU oxidant out	132.2	3.999	132.2	3.999
Gasifier coal in	132.2		132.2	
Gasifier water in	132.2		132.2	
Gasifier	1343.0		1343.0	4.24
Gasifier syngas out	38.3	3.944	260.0	3.944
SRU syngas out	29.4	2.958	29.4	2.958
CO ₂ absorber product	-	-	-	13.79
CO ₂ absorber syngas out	-	-	29.4	2.889
Combustor air in	20.0	1.592	20.0	1.592
Heated Fuel Gas	175.0	2.889	175.0	2.889
Gas turbine flue gas out	605	1.379e-2	605	1.379e-2
Steam turbine flue gas out	121.1	-	121.1	-
Cooling water in	32.2	-	32.2	-
Cooling water out	21.1	-	21.1	-
Stack	121.1	-	121.1	-

Table 33a: The energy balance for the 323 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels.

	0% CCS		90% CCS		95% CCS	
	Blend	Coal	Blend	Coal	Blend	Coal
Net Electrical Output (MW)	240.7	240	228.9	231.9	225.2	228.4
Total Plant Input (GJ/hr)	2157	2031	2367	2250	2350	2233
Gross Plant Heat Rate, LHV (kJ/kWh)	7853	7480	8613	8194	8637	8215
Net Plant Heat Rate, LHV (kJ/kWh)	8960	8469	10343	9705	10438	9781
Net Plant Efficiency, LHV (%)	40.18	42.51	34.81	37.11	34.49	36.80
Total Generator Output (MW)	442.1	440.7	447.5	447.4	446.3	446.2
Air Compressor Use (MW)	163.7	165.3	169	169.2	170.5	170.6
Turbine Shaft Losses (MW)	5.568	5.507	5.571	5.565	5.515	5.513
Gross Plant Output (MWg)	274.6	271.7	274.8	274.5	272.1	271.9
Misc. Power Block Use (MW)	5.493	5.433	5.496	5.49	5.441	5.439
Air Separation Unit Use (MW)	18.9	17.32	20.74	19.17	20.6	19.04
Gasifier Use (MW)	3.825	3.468	4.197	3.839	4.167	3.813
Sulfur Capture Use (MW)	2.317	2.113	3.118	2.937	3.101	2.922
Water-Gas Shift Reactor Use (MW)	-	-	-10.63	-10.62	-10.53	-10.52
Selexol CO ₂ Capture Use (MW)	-	-	19.62	18.46	20.73	19.51
Cooling Tower Use (MW)	1.617	1.612	1.636	1.635	1.632	1.632

Table 33b: The energy balance for the 646 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels.

	0% CCS		90% CCS		95% CCS	
	Blend	Coal	Blend	Coal	Blend	Coal
Net Electrical Output (MW)	484.3	482.6	460.8	466.7	453.4	459.7
Total Plant Input (GJ/hr)	4313	4065	4733	4498	4701	4469
Gross Plant Heat Rate, LHV (kJ/kWh)	7853	7480	8613	8194	8663	8215
Net Plant Heat Rate, LHV (kJ/kWh)	8908	8420	10267	9638	10371	9724
Net Plant Efficiency, LHV (%)	40.41	42.76	35.05	37.35	34.72	37.03
Total Generator Output (MW)	884.2	881.4	895	894.9	892.5	892.5
Air Compressor Use (MW)	327.4	330.6	338	338.3	341	341.2
Turbine Shaft Losses (MW)	11.14	11.01	11.14	11.13	11.03	11.03
Gross Plant Output (MWg)	549.3	543.3	549.6	549	544.1	543.9
Misc. Power Block Use (MW)	10.99	10.87	10.99	10.98	10.88	10.88
ASU (MW)	37.81	34.64	41.48	38.34	41.19	38.08
Gasifier Use (MW)	7.37	6.682	8.087	7.397	8.03	7.346
Sulfur Capture Use (MW)	3.87	3.505	5.303	4.979	5.273	4.952
WGS use (MW)	-	-	-21.27	-21.25	-21.06	-21.05
Selexol CO ₂ Capture (MW)	-	-	39.18	36.87	41.41	38.97
Cooling Tower Use (MW)	3.233	3.224	3.271	3.271	3.264	3.263

High Level Control Philosophy

Entrained flow gasification involves the process of partial combustion of the fuel at relatively high temperature to produce a mixture of carbon monoxide, hydrogen and carbon dioxide and a range of hydrocarbons under carefully controlled conditions of temperature, reactant composition, and residence time in the gasifier. The water gas shift reactor is a means primarily to react the carbon monoxide with water to produce hydrogen and carbon dioxide via a catalysed chemical reaction, along with its secondary role in converting carbonyl sulphide to hydrogen sulphide. The optimal performance of these systems requires careful control of fuel, oxygen, steam feed rates and temperature. The oxygen is supplied from a conventional cryogenic ASU. The system is best suited, therefore, to a steady state base load mode of operation and has limited flexibility of operation. However, after the CO₂ removal stage, the resulting gas stream is in effect pure hydrogen which with sufficient storage capacity would allow the subsequent power generation plant sections to be in effect decoupled from the fuel generation and thus allow the capability to operate dynamically.

Environmental performance

The predicted pollutant emissions from the IGCC cases studied are presented in this section for the following 2GT cases:

1. Coal, 0% CCS, wet cooling tower (WCT)
2. Coal, 95% CCS, wet cooling tower (WCT)
3. Biomass blend, 0% CCS, wet cooling tower (WCT)
4. Biomass blend, 95% CCS, wet cooling tower (WCT)

The calculated values of the inputs in terms of elemental components and the uncontrolled pollutant species outputs expressed as kg/kWh are summarised in the Tables 34 and Table 35 below.

Table 34: Elemental inputs in the case studies: pure coal 0% CCS and 95% CCS, WCT and biomass blend 0% CCS and 95% CCS, WCT.

kg/kWh	0% CCS		95% CCS	
	Coal	Blend	Coal	Blend
Fuel	0.3592	0.3923	0.4146	0.4566
C	0.2141	0.2261	0.2471	0.2631
H	0.0184	0.0202	0.0212	0.0235
O	0.0581	0.0806	0.0671	0.0938
Cl	7.18E-04	7.06E-04	8.29E-04	8.22E-04
S	6.47E-03	6.16E-03	7.46E-03	7.17E-03
N	5.39E-03	5.22E-03	6.22E-03	6.07E-03

Oxidant	0.2241	0.2437	0.2587	0.2837
N	3.91E-03	4.25E-03	4.51E-03	4.95E-03
O	0.220	0.239	0.254	0.279
Water	1.414	1.413	1.779	1.794
H	0.157	0.157	0.198	0.199
O	1.257	1.256	1.581	1.595

Table 35: The estimates of the uncontrolled emission levels of the key pollutant species for: pure coal 0% CCS and 95% CCS, WCT and biomass blend 0% CCS and 95% CCS, WCT.

kg/kWh	0% CCS		95% CCS	
	Coal	Blend	Coal	Blend
Gaseous				
CO ₂	0.7474	0.7817	0.08103	0.08334
HCl	6.98E-04	6.56E-04	8.05E-04	7.64E-04
SO ₂	2.64E-04	2.57E-04	2.46E-05	2.38E-05
NO	5.70E-05	5.69E-05	6.19E-05	6.28E-05
NO ₂	4.60E-06	4.59E-06	4.99E-06	5.06E-06
Solid/Liquid				
Slag*	6.25E-02	6.01E-02	7.21E-02	7.00E-02
Particulate emissions to air**	5.70E-05	6.03E-05	6.58E-05	7.02E-05

*Carbon in slag 3% suggested by DOE reports as a typical value.

** Exact composition unknown

Due to the lower heating value of the biomass, more fuel is used in the co-firing case. However, because of the lower sulphur content of the biomass the overall SO₂ emission intensity is lower than that for coal. The same trend is observed for the HCl formation and the slag production due to lower levels of Cl and ash contained in the biomass. Levels of nitric oxides are very similar for the different fuels studied, while the higher CO₂ and particulate emissions in the co-firing cases are a direct consequence of the higher fuel consumption.

Process economics

The capital and the operating and maintenance cost estimates for the IGCC plants with biomass co-firing and CO₂ capture by physical absorption are presented in Tables 36-39, for the 323 and 646 MW_e power plants respectively. Only the costs associated with the wet cooling tower technology have been reported. The wet cooling tower is cheaper than the air cooled condenser, because of an initial higher capital required for the construction of the condenser and the higher power consumption of the air system during operation. Specifically the capital cost of the air cooled condenser is double in comparison with the wet cooling tower, and the O&M costs are increased by 20%. The capital expenditure listed in Tables 36 and 38, depends on the plant size, as expected, and the additional costs associated with the CO₂ capture plant are significant, and are dependent on the scale of operation. Costs of material handling are included in the tables and they have been chosen in line with the other technologies. Because of the lower heating value of biomass, more fuel is required for the co-firing case. As a consequence, the effect of replacing coal with coal-biomass blends is an increase in the capital costs of the power plant of about 5%. In this Case study, the net output of the power plant is maintained, and the additional capital costs associated with CO₂ capture are associated with the increase in the size of the ASU, gasifier and power train in addition to the costs of the CO₂ capture plant. The O&M cost estimates for the 323 and 646 MW_e power plants are listed in Tables 37 and 39.

Table 36: The capital cost estimate (£M), WCT for the 323 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels, for a capacity factor of 85%.

Unit	0% CCS		90% CCS		95% CCS	
	Coal	Blend	Coal	Blend	Coal	Blend
Air Separation Unit	63.69	67.01	67.46	70.66	67.20	70.40
Gasifier Area	104.45	115.52	114.56	125.70	113.79	124.86
Sulphur Control	30.62	32.35	33.71	34.72	33.52	34.52
CO2 Capture	-	-	61.92	61.93	62.40	62.43
Power island	87.04	86.98	87.04	86.72	86.91	86.59
Cooling Tower	5.44	5.46	5.52	5.53	5.51	5.52
CO2 compressor	-	-	31.4	31.4	31.4	31.4
Offsites, storage and handling	118.7	118.7	136.14	136.1	136.1	136.1
Total Installed Costs (£M)	409.9	426.0	537.7	552.7	536.8	551.8
Operation and utilities	20.50	21.30	26.89	27.64	26.84	27.59
Civils and land costs	40.99	42.60	53.77	55.27	53.68	55.18
Project development	20.50	21.30	26.89	27.64	26.84	27.59
Contingency	40.99	42.60	53.77	55.27	53.68	55.18
Total investment costs (£M)	532.9	553.8	699.1	718.6	697.9	717.4
Specific CAPEX (£M/MW_e)	2.22	2.30	3.01	3.14	3.06	3.19

Table 37: The O&M cost estimate (£M/yr) for the 323 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels, for a capacity factor of 85%.

Unit	0% CCS		90% CCS		95% CCS	
	Coal	Blend	Coal	Blend	Coal	Blend
Fuel Cost	31.31	45.34	34.66	49.74	34.43	49.40
Disposal cost	1.29	1.25	1.43	1.37	1.42	1.36
Water	0.84	0.84	1.20	1.17	1.19	1.18
Sulphur By-product Credit	0.555	0.543	0.614	0.596	0.610	0.592
Variable O&M (£M/yr)	34.00	47.97	37.91	52.88	37.66	52.53
Maintenance and Labour	16.40	17.04	21.51	22.11	21.47	22.07
Insurance	4.10	4.26	5.38	5.53	5.37	5.52
Fixed O&M (£M/yr)	20.50	21.30	26.89	27.64	26.84	27.59
Total O&M (£M/yr)	54.49	69.27	64.80	80.52	64.50	80.12

Table 38: The capital cost estimate (£M), WCT for the 646 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels, for a capacity factor of 85%.

Unit	0% CCS		90% CCS		95% CCS	
	Coal	Blend	Coal	Blend	Coal	Blend
Air Separation Unit	94.14	99.136	99.78	104.58	99.39	104.128
Gasifier Area	170.75	193.73	191.74	210.82	190.53	209.41
Sulphur Control	53.93	57.42	59.85	61.98	59.51	61.61
CO2 Capture	-	-	119.10	121.98	120.64	123.65
Power island	170.18	170.05	170.18	169.60	169.92	169.34
Cooling Tower	10.97	11.01	11.14	11.16	11.11	11.13
CO2 compressor	-	-	31.4	31.4	31.4	31.4
Offsites, storage and handling	118.7	118.7	136.14	136.1	136.1	136.1
Total Installed Costs (£M)	618.7	650.0	819.3	847.6	818.6	846.8
Operation and utilities	30.93	32.50	40.97	42.38	40.93	42.34
Civils and land costs	61.87	65.00	81.93	84.76	81.86	84.68
Project development	30.93	32.50	40.97	42.38	40.93	42.34
Contingency	61.87	65.00	81.93	84.76	81.86	84.68
Total investment costs (£M)	804.3	845.1	1065.1	1101.9	1064.2	1100.8
Specific CAPEX (£M/MW_e)	1.67	1.74	2.28	2.39	2.31	2.43

Table 39: The O&M cost estimate (£M/yr) for the 646 MW_e IGCC-CCS system at 0, 90 and 95% CO₂ capture levels, for a capacity factor of 85%.

Unit	0% CCS		90% CCS		95% CCS	
	Coal	Blend	Coal	Blend	Coal	Blend
Fuel Cost	62.62	90.69	69.34	99.53	68.83	98.81
Disposal cost	2.58	2.49	2.86	2.74	2.84	2.72
Water	1.69	1.69	2.42	2.40	2.39	2.37
Sulphur By-product Credit	1.11	1.0861	1.23	1.1919	1.22	1.1835
Variable O&M (£M/yr)	68.00	95.97	75.84	105.85	75.28	105.08
Maintenance and Labour	24.75	26.00	32.77	33.90	32.74	33.87
Insurance	6.19	6.50	8.19	8.48	8.19	8.47
Fixed O&M (£M/yr)	30.93	32.50	40.97	42.38	40.93	42.34
Total O&M (£M/yr)	98.93	128.47	116.81	148.23	116.21	147.42

Overview of the system performance and further development requirements

The net power plant electricity generation efficiency levels are listed in Table 40. The basic IGCC systems without CCS are of the order of 42.8% for coal and 40.4% for biomass blend. The estimated efficiency penalty associated with the carbon dioxide capture system is about 5-6%.

Table 40: Overall net power generation efficiency levels on a LHV basis.

Case	Coal			Blend		
	0% CCS	90% CCS	95% CCS	0% CCS	90% CCS	95% CCS
1GT	42.51	37.11	36.80	40.18	34.81	34.49
2GT	42.76	37.35	37.03	40.41	35.05	34.72

One of the key barriers to the wider implementation of the IGCC technology at this stage is the plant capital costs. A number of projects in Europe are being developed and are due to become active in the next 5-6 years. Government funding has been employed in most of these projects (DTI, 1998). A number of projects have also been developed in the U.S. with funding under the DOE's Clean Coal Power Initiative (CCPI).

In the future it is anticipated that the costs of IGCC technology can be reduced by improvement of the design and performance of some of the main components of the system, and in particular the ASU, and gas cleaning system. It is considered that new technologies for air separation using

membranes to separate the oxygen from other air components, with consequent savings in energy consumption may have significant advantages in this regard.

A different approach for the syngas cleaning may also provide a cheaper alternative to the conventional system. The development of gas cleaning systems would potentially enhance the cleaning efficiency and also be employed for the production of H_2 , by using catalysts which would maximise the effectiveness of the water shift reactor.

CASE STUDY 8 A small scale (<50 MWe) dedicated biomass IGCC with CO₂ capture by physical absorption

The Case Study reported in this section focuses on dedicated biomass IGCC, i.e. BIGCC plants using wood chips as fuel.

In the BIGCC system the biomass fuel is fed into the gasifier and converted to syngas, with oxygen produced by an air separation unit (ASU). The gasifier operates at a pressure of 35 bar, which allows for more efficient downstream separation of CO₂. The slag is sent to landfill. The syngas is then sent to the water gas shift reactors used to convert the CO and water into CO₂ and hydrogen. Two water gas shift reactors are needed at different temperatures to achieve a good yield. The Selexol process then removes the CO₂ and any acid gases. The sulphur, predominantly in the form of H₂S and COS is removed from the syngas and sent to a Claus plant and a Beavon-Stretford tail gas treatment unit for conversion to elemental sulphur. This can be recovered and sold as a by-product. The sulphur free syngas can undergo a second stage of the Selexol process for CO₂ capture. A system of CO₂ stripping is used to recover the solvent and collect the CO₂, which is then compressed for further processing.

The clean syngas now consisting primarily of hydrogen is fed in the combustor where it is mixed with air and burns to produce hot combustion products entering the Gas Turbine (Brayton Cycle) and then generating heat for the steam cycle (Rankine Cycle).

A total of five plant configurations were studied in this Case Study:

- 50MWe (40 MW_e net) and 90% carbon capture – overall efficiency found to be 34%
- 50MWe and 70% carbon capture – overall efficiency found to be 37%

- 30MWe (24 MW_e net) and 90% carbon capture – overall efficiency found to be 34%
- 30MWe and 70% carbon capture – overall efficiency found to be 37%
- 30MWe and 0% carbon capture – overall efficiency found to be 41%

Status of development and technology readiness levels

Dedicated biomass IGCC plants combined with pre-combustion capture are only currently at TRL 4, since no fully integrated pilot plants have as yet been built. By 2020, it is considered that the TRL may increase to 5 or 6. Further, owing to constraints on the availability of sufficient biomass, we note that BIGCC plants will be restricted to relatively small scales of operation.

Process flow diagram

The major process units and the overall process flows with and without CCS are presented in Figure 13a and Figure 13b. The individual process units will be discussed in a further technical detail in the next section.

Air Separation Unit

A cryogenic distillation technique is used to separate the oxygen from the nitrogen in air, and the oxygen is compressed from 2 bar to 30 bar prior to injection into the gasifier. The flow rate of oxygen from the ASU depends on the specifics of the case being considered, i.e. the scale of operation and the level of CO₂ capture. In the cases described above, the oxygen flow rate varies between 2.3 and 4 tonnes/hr.

Gasifier Area

A fluidised bed gasifier with bottom cyclone to remove ash has been used for this Case Study. Pyrolysis, gasification and tar conversion (cracking and conversion to gasification products) are assumed to take place in the main reactor. Experimental results at Imperial do indicated two tar conversion mechanisms: homogeneous gas phase cracking and heterogeneous cracking on char surfaces. In this report, tar conversion is assumed, but a tar trap of some kind after the gasifier would be needed in practice. The pressurised oxygen-blown gasifier is used to convert the wood chips into syngas. The syngas at the gasifier exit is at a temperature of about 950°C. The ash is collected separately from the gas as a fused slag, which is sold or sent to landfill disposal. Typical H₂/CO mole ratios in the syngas at the exit of the gasifier would be around 1.63. For the benchmark fuel used in this study, the flow rate varied from 608 tonne/day for 30MW_e plant to 957 tonne/day at 50MW_e.

Water Gas Shift Reactors

Two water gas shift reactors in series are used to convert the CO and water to CO₂ and H₂. The mole ratio of H₂ to CO leaving the High Temperature shift reactor is increased to 7, and that leaving the Low Temperature shift reactor is 27.

Sulphur and carbon removal

A Selexol sulphur capture system is used in the same way as for a coal-based IGCC case. The CO₂ is removed through a second stage of Selexol absorber capture, with an efficiency which varies in this Case Study from 70% or 90%. The energy penalty associated with this process is estimated to be about 3.8%. CO₂ compressors are used to bring the CO₂ to a pressure of around 108 bar prior to transportation for storage in the geological site. The energy penalty of CO₂ compression process is about 5-7%.

Power island

The power island comprises:

- the Gas Turbine section and
- the Steam Cycle.

The number of gas turbines used determines the size of the plant, in particular in this study one gas turbine has been used giving outputs of 17.4 MWg and 29.1 MWg for the 30 and 50 MW cases respectively. Ambient air is compressed to 30 bar and 635C and the hydrogen rich syngas is

combusted in the chamber at a temperature of 1300°C. The hot gases exit the turbine at 549C. The heat is recovered in a HRSG and steam turbine block. The multistage turbine system generates 25MW and 43 MW for the 30 and 50 MW cases.

High Level Control Philosophy

There is relatively little industrial experience of the pressurised gasification of biomass at the 30-50MWe scale. It is clear however that the front end of the process involves the operation of an ASU and, with other factors, this will tend to limit the operational flexibility of the system.

After the CO₂ removal process, however, this part of the plant is essentially a hydrogen CCGT plant. With sufficient storage of hydrogen the upstream and downstream parts can be decoupled and the CCGT part operated in a fairly dynamic fashion. The dynamic control and start-up/shut-down of this type of plant is well-established. This would enable the operator of such a plant to access premium peak-demand payments.

Environmental performance

The environmental performance is expected to be similar to that in Table 41a below. The process of gasification has a large advantage over biomass combustion in that the uncontrolled particulate and other potentially harmful emissions tend to be much lower.

The use of the Selexol process to treat the gases has the combined effect of removing sulphur containing compounds and CO₂ at high efficiency. The amount of ash produced will be somewhat lower than coal but in practice will of course depend on the fuel details. One potential concern is the overall water footprint of the process (see Table 41b); water is needed for the water gas shift reaction in stoichiometric proportions and for cooling if evaporative cooling processes are to be used.

In general term, it is expected that compliance of the dedicated biomass IGCC power plants with the relevant environmental standards will not present any significant difficulties compared to the situation with other comparable processes.

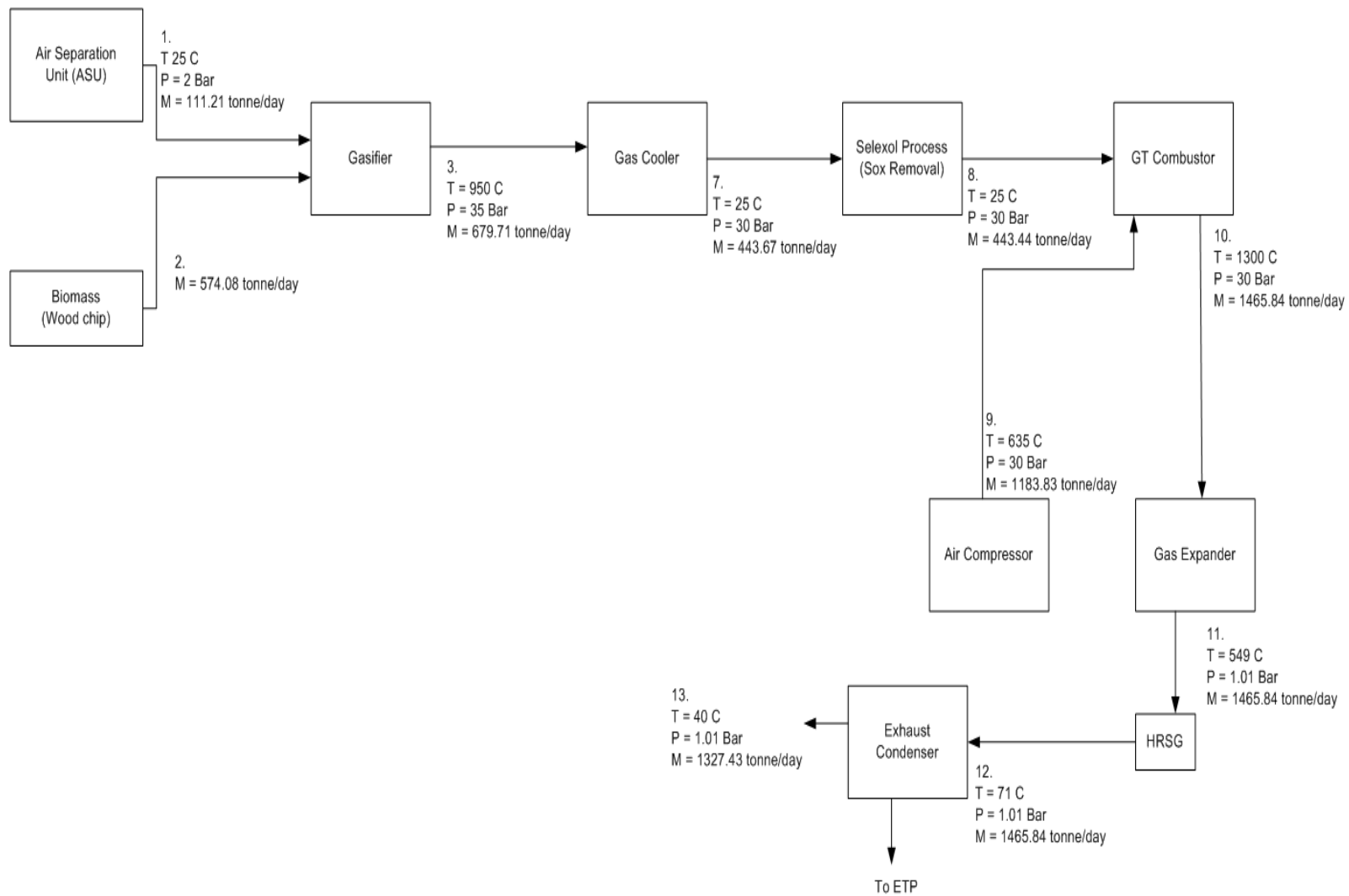


Figure 13a: The overall process flow diagram for BIGCC (without WGS and CCS).

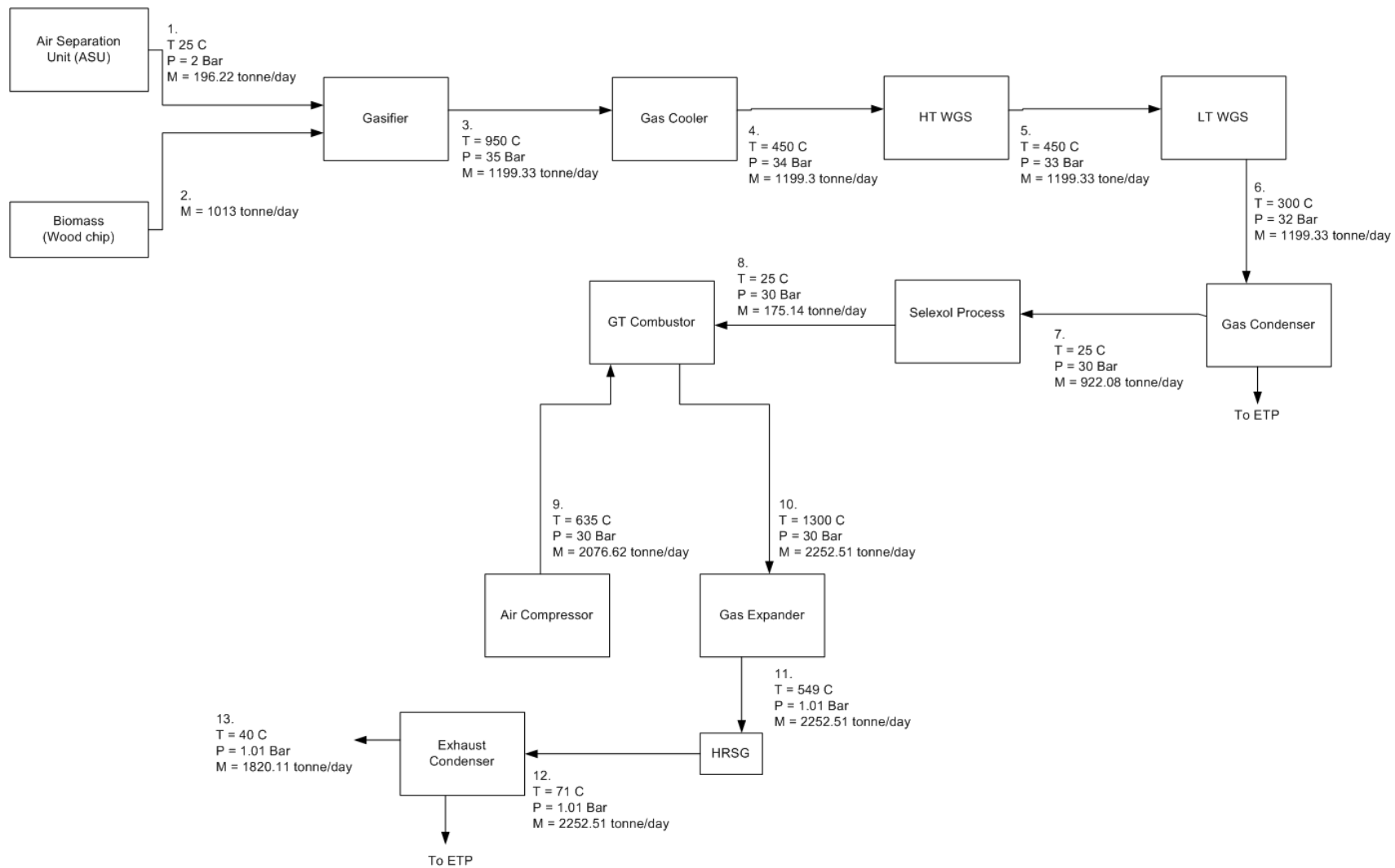


Figure 13b: The overall process flow diagram for BIGCC with CCS.

Table 41: Elemental inputs (a) and emissions outputs (b) for a representative coal and biomass/dedicated IGCC CCS, WCT.

(a)

	1 GT	
kg/kWh	Coal	WP
CO ₂	8.12E-01	6.65E-02
HCl	7.57E-04	4.21E-06
SO ₂	2.87E-04	1.34E-07
NO	5.50E-05	3.59E-05
NO ₂	4.43E-06	2.90E-06
Slag	7.047e-2	5.391e-3
Particulate emissions to air	6.429e-5	4.245e-5

(b)

	1 GT	
kg/kWh	Coal	WP
Fuel	0.4051	0.3518
C	0.232497	0.15628
H	0.0179	1.89E-02
O	0.0462	1.75E-01
Cl	7.80E-04	3.52E-05
S	0.00702	3.52E-05
N	0.00585	7.04E-04
Oxidant	0.2624	4.77E-05
N	0.004579	8.32E-07
O	0.257746	4.69E-05
Water	0.4087	1.694
H	0.0450	0.1863
O	0.3637	1.508

Process Economics

The capital costs estimates for the BIGCC plants described in this Case Study are listed in Table 42. The system capital cost is expected to be in the range of 100 to 158 £M for the 30 and 50 MW_e cases. The energy penalty, including CO₂ compression, is about 30%, i.e. for ASU power + air and CO₂ compressors power + CO₂ capture requirements, as a percentage of the gross system power.

About 1.3 £M more capital is required for the air cooled condenser case, compared to the wet cooling tower case.

The Outside Battery Limit (OSBL) costs including instrumentation, building, grid connections, site operations, civil works, electronics and piping, have been assumed to contribute a further 30% of the ISBL costs. The total direct capital (TDC) is 1.3 times the ISBL cost of BIGCC, 128-205 M£, for 30MW and 70% carbon capture and 50MW and 90% carbon capture cases, respectively.

Table 42: The capital cost estimates for the BIGCC plants. (100% biomass, wet cooling tower).

	50MW and 90% CCS (£M)	50MW and 70% CCS (£M)	30MW and 90% CCS (£M)	30MW and 70% CCS (£M)
Air Separation Unit	38.00	37.00	28.00	28.00
Gasifier Area	55.00	53.00	36.00	34.00
Particulate Control	0.00	0.00	0.00	0.00
Sulfur Control	14.00	14.00	9.00	9.00
CO2 Capture	23.00	22.00	14.00	13.00
Power island	24.00	23.00	15.00	14.00
Cooling Tower	2.00	2.00	1.00	1.00
Storage and handling	11.00	11.00	7.00	7.00
Total installed costs	167.00	162.00	110.00	106.00
Operation and utilities	8.35	8.10	5.50	5.30
Civils and land costs	16.70	16.20	11.00	10.60
Projectdevelopment	8.35	8.10	5.50	5.30
Contingency	16.70	16.20	11.00	10.60
Total investment costs	217.10	210.60	143.00	137.80
Specific investment costs (£M/MW_e)	5.42	5.26	5.95	5.74

The total investment costs for the scales of generation considered here, i.e. 30 to 50 MW_e was observed to lie in the range of about 137 to 217 £M/MW_e. The specific investment costs for BIGCC with CCS varied between 5.2 and 5.95 £M/MWh.

The OPEX cost estimates are presented in Table 43. The costs of insurance (1% of TIC), labour and maintenance (4% of TIC), make-up solvent for Selexol process, waste treatment (Effluent treatment plant, ETP) and wood chip constitute the total operating cost of the BIGCC system. The total OPEX varies between 20-35 M£/year for a 50 £/t wood chip price.

Table 43: The operating cost estimates for the BIGCC plants at wood chip prices of £50/t.

		50MW and 90% CCS	50MW and 70% CCS	30MW and 90% CCS	30MW and 70% CCS
		£M/yr	£M/yr	£M/yr	£M/yr
Make-up solvent (0.5%)	20Euros/t	10.1	7.4	6.1	5.7
Waste treatment	0.063 £/MWh	0	0	0	0
Wood Chip/Fuel	50 £/t	16.8	15.8	10.1	9.5
Variable O&M		26.9	23.2	16.2	15.2
Labour and Maintenance	4% TIC	6.68	6.48	4.4	4.24
Insurance	1% TIC	1.67	1.62	1.1	1.06
Fixed O&M		8.35	8.1	5.5	5.3
Total O&M		35.25	31.3	21.7	20.5

Overview of system performance and further development requirements

The core of these flowsheets is the gasification process; this is the technology which most strongly influences system performance, and with which the least experience has been evident. The other processes, including ASU, shift reactions, Selexol process and power block can be considered to be at TRL 9 (although membrane separation systems for O₂ production which might be ideal for this scale of plant are closer to level 6). In common with all the processes studied here, the CO₂ compression is of course less well developed and this can have a significant impact on the energy recovery and gas quality.

Dedicated biomass IGCC plants combined with pre-combustion capture are only currently at TRL 4, since no fully integrated pilot plants have been built yet. By 2020, it is thought likely that the TRL could have increased, reaching TRL 5 to 6.

There are a number of different technologies available based on this design, including: TPS, Foster Wheeler, Battelle, Andritz Carbona, Repotec, Uhde, Enerkem and Lurgi. There are many examples of operating systems which have worked well for a number of years, as well as others which have had problems associated with gas cleanup, tar formation, mechanical problems, etc. and these have been the subject of significant development programmes. Biomass gasification is a rapidly developing area and a key challenge will be to ensure feedstock flexibility. The other key challenge is the full systems integration with carbon capture, shift reactions, ASU and power train.

DISCUSSION

An assessment of the relative performance and economics of the eight technology combination covered by the case studies is provided in this section of the report. The case studies that are presented above have covered briefly both the technical and economic aspects of the following technology combinations:

1. The co-firing of biomass in a pulverised coal-fired power plant, with post-combustion carbon dioxide capture by amine scrubbing,
2. The combustion of biomass in a dedicated power plant with post-combustion carbon dioxide capture by amine scrubbing,
3. The co-firing of biomass in a pulverised coal power plant, with carbon dioxide capture by using oxyfuel firing technology, with cryogenic O₂ separation,
4. The combustion of biomass in a dedicated coal-fired power plant with carbon dioxide capture by oxyfuel firing technology, with cryogenic O₂ separation,
5. The co-firing of biomass in a pulverised coal-fired power plant, with carbon dioxide capture by post-combustion carbonate looping,
6. Dedicated biomass chemical-looping-combustion using a solid oxygen carrier,
7. The co-firing of biomass in a large coal-fired Integrated Gasification Combined Cycle plant, with carbon dioxide capture by physical absorption, and
8. A dedicated biomass Integrated Gasification Combined Cycle (IGCC), with carbon dioxide capture by physical absorption.

In each case, the basic technology and the current status of development have been described. Outline process flow diagrams with listings of the key flows and process conditions have been presented, and calculations of the overall process efficiencies have been prepared. High level assessments of the control and environmental issues have been performed. Preliminary study estimates of the capital and O&M costs have been prepared.

The level of detail given in each case study varies significantly, depending largely on the status of development of the technology combination. For Case studies 1-4 and 7, the status of development is such that good quality data are available, as has been described above. For the other cases, the technologies are much less well developed, and there are significant shortfalls and gaps in the available technical and economic data. These can only be filled by significant further development work at laboratory and pilot scale, to provide the basic process information on which to base the required engineering and economic studies.

A comparison of the key technical and economic features of the technology combinations will be presented in this section of the report. This will cover:

- the current status of development and the barriers to further implementation, and
- the potential technical advantages and disadvantages of the technologies and the technology combinations.

A direct comparison of the estimated capital and operating costs of the technology combinations will be made (as far as is possible with the economic data available).

Overall, the great majority of the industrial expenditure to date on the development of carbon dioxide capture technology has been, and continues to be, on the more conventional approaches which have been designed for use on large coal-fired power plants, i.e.

- The application of CO₂ capture by solvent scrubbing technologies to the flue gas streams from large pulverised coal-fired boilers and fluidised bed boilers firing coal, and the compression and drying of the CO₂ prior to transportation offsite,
- The application of oxyfuel firing technologies to large coal-fired boilers, and the purification, compression and drying of the product CO₂ prior to transportation off site
- The application of physical CO₂ absorption techniques to the syngas produced by large coal-fired IGCC power plants.

To date, the development of all three of these technologies has been pursued in parallel, with each currently at the pilot- and small demonstration scale. However, it has proven to be difficult for the electricity supply industry and their equipment suppliers to come to a clear decision as to which technology is to be preferred for future application. The industry is now moving to large scale demonstration projects but because these are relatively expensive and require significant government subsidies, the progression to the demonstration phase has been relatively slow. It is also clear that there has been relatively little activity on the application of these carbon capture technologies to dedicated biomass power plants or to coal power plants which are co-firing biomass. For most applications, the biomass fuel is significantly more expensive than coal and the application of the CO₂ capture technology is relatively expensive. At the present time, there is no commercial incentive to apply CO₂ capture technologies to dedicated biomass power plants, and specific incentives, associated with the removal of CO₂ from the atmosphere will need to be introduced to support these activities. This is not as yet being discussed seriously at government level.

It should also be noted that the power industry is historically very conservative in character, and the current electricity generation, distribution and trading systems have become accustomed over many decades to the degree of flexibility of operation provided by the pulverised coal boiler. It may also be the case that in the future, when a higher portion of the UK's power generation will be from intermittent sources, the flexibility of operation provided by thermal power plants will be of even more importance.

As stated above, the eight selected candidate technology combinations for CO₂ capture can be considered under three distinct categories, viz:

- Post-combustion,
- Oxyfuel firing, and
- Pre-combustion.

Post-combustion technologies involve the addition of significant chemical process equipment to the flue gas outlets of the thermal power plant, downstream of the current environmental control equipment. This type of technology is relatively familiar to the electricity supply industry, since the application of liquid or slurry scrubbing of flue gases for the control of gaseous emissions has been in

practice for several decades. In general terms, these technologies will not have a major impact on the flexibility of operation of the power plant. One exception to this will be in cases where there is a requirement for an oxygen supply, e.g. for the carbonate looping system, where there is oxygen blown combustion of a secondary fuel in the calciner. The requirement of an air separation unit (ASU) to supply the combustion system may have a significant impact on the flexibility of operation of the overall power plant.

Oxyfuel technologies conventionally involve a fairly major reconfiguration of the fuel firing systems, the draft plant and the boiler, and the installation of an air separation unit for oxygen production. In the case of chemical looping combustion technologies, the reconfiguration of the firing and heat transfer systems is fairly radical. In this case the ASU will have a significant impact on the flexibility of operation of the plant. This can be reduced by the provision of oxygen storage facilities on site (although is clearly a process and site specific issue).

The pre-combustion technologies normally involve gasification or co-gasification of the biomass fuel, with CO₂ removal from the syngas, and the combustion of the hydrogen-rich fuel gas in a modified gas turbine.

Post-combustion carbon dioxide capture

Looking in the first instance at the more promising post-combustion CO₂ capture technologies which can be applied to biomass combustion or co-firing systems, the **liquid solvent** based technologies, most commonly involving aqueous solutions of organic amines as described in Case Study 1, have proven to be by far the most popular to date, and may well also prove to be so for installations at thermal power plants in the long term.

These technologies are flexible in operation, and are equally suitable as retrofits to existing boilers as they are to new build plants, which may assist with their development and deployment in the short- to medium term future. They can also be applied to new build power plants, at practically any scale of operation and to boiler plants firing coal with or without biomass co-firing or dedicated biomass power plants.

The CO₂ concentrations in most combustion flue gases, however, are relatively low, at less than 20% on a volumetric basis, and the CO₂ loading of the rich solvent solutions is modest. One result of this is that the flue gas and liquid solvent flow rates within the plant are relatively high. As a result, the absorber and stripper reaction vessels and the associated equipment tend to be large, and have high capital costs. The technology is also relatively energy intensive. There is a significant steam requirement for solvent regeneration as well as a power requirement for the compression and drying of the captured CO₂. This reduces the electrical output of the power plant substantially. The overall cycle efficiency can be reduced by up to 10-13% points, depending on the process configuration and the scale of operation.

The majority of the current development in solvent scrubbing technology is at pilot and small demonstration scales, and these efforts are aimed principally at finding process improvements in the following subject areas:

- the optimisation of the chemical formulation of the liquid solvents, and of the scrubbing/stripping cycles,

- the reduction of the heat energy requirement of the solvent scrubbing and stripping systems,
- the minimisation of the solvent replacement and other costs associated with amine losses and degradation, and
- the control of the corrosion rates of the internal surfaces of the scrubbing and stripping vessels and associated equipment, by the use of corrosion inhibitors in the solvent formulations and by the use of improved materials.

It is clear also that the design and operation of the post-combustion CO₂ capture technologies based on liquid solvents will minimise any negative impact on the flexibility of operation of the power plant. It is anticipated that the liquid solvent CO₂ capture technologies will be commercial with a number of demonstration plants in operation, and a number of technology and equipment suppliers by the year 2020.

The use of **Ca looping technology** as an alternative to liquid solvents for post-combustion CO₂ capture, has been described in Case Study 5. This technology relies on relatively well understood gas-solid reactor system designs, based on circulating fluidised beds, with lime/limestone as the bed material. Scale-up of the technology from the laboratory and the small pilot plant scale may proceed rapidly, if the level of technical and commercial interest from the larger industrial companies continues to increase over the next few years. At present this is fairly modest, because of the level of effort on the other technologies.

The demonstrations of Ca looping taking place in Darmstadt, Germany and the La Pereda power plant, Asturias, Spain, as part of the EU CaOling project, are both at the scale of 12 MW_{th}. The future development of this technology will be largely dependent on the success of these demonstrations to provide sufficient confidence in the technical and economic viability of the process, to encourage significant industrial involvement in moving towards the small demonstration scale.

As with most systems involving a solid phase absorption process, the solid circulation rate in the carbonate looping systems is reasonably high, and the key technical issues are with the attrition rate of the solid particles in the CFB environment, and with the practical and control difficulties associated with the movement of large volumes of hot solids between the reactors at industrial scale. Limestone is a relatively fragile material, particularly in the calcined form. The physical degradation and sulphation of the lime sorbent with time will tend to reduce the CO₂ capture efficiency of the system, and there may be increased operating costs associated with spent sorbent separation or recycling. The ability of the carbonate looping process to absorb sulphur may well reduce the power plant's flue gas desulphurisation requirements. As indicated in Case study No. 5, there may be significant synergies with the cement industry – with the potential to sell the spent sorbent, as opposed to paying for its disposal. Therefore, it is important that the link between cement manufacture and Ca looping is adequately developed and demonstrated. This link, with the potential to decarbonise the power industry and substantially decarbonise cement manufacture, is one potential advantage of the Ca looping technology.

It is also clear that in a power plant fitted with CO₂ capture in a Ca looping system, a large amount of additional fuel would need to be burnt in the calciner. The heat would then be recovered by heat exchange tubes, part of the boiler steam cycle, inserted into the calciner operating at temperatures around 650°C. The technical and other implications of incorporating such a heat exchanger into the

steam cycle of a power plant boiler are significant and will have to be considered very carefully. However, provided this is carried out correctly, the additional fuel burnt in the calciner would lead to a corresponding increase in the power plant's gross electrical output – i.e. a likely repowering of some 30-40% in a retrofit situation.

The limestone calciner unit also has a requirement for a supply of oxygen from an air separation unit to burn the fuel, and provide the high temperatures, around 900°C, required for calcination of the limestone. The ASU's power requirement will reduce the net power output of the system and, as indicated above, could have a significant impact on the flexibility of operation of the plant if oxygen storage facilities are not present. However, the size of the ASU will be significantly smaller than for an oxy-fuel firing boiler, hence this power requirement is only likely to be around 3% points.

In summary, therefore, solvent scrubbing techniques applied to coal power plant co-firing biomass or dedicated biomass plants are likely to be the preferred post-combustion technology for CO₂ capture from biomass in the near-medium term. Ca looping was identified as representing the most promising potential alternative candidate, offering significant potential efficiency improvements. There are, however, a number of technical hurdles to overcome before this technology becomes competitive with more advanced carbon capture technologies.

Oxyfuel firing technologies

Normally, the term oxyfuel combustion is employed to describe the combustion of a fuel in a mixture of oxygen and recycled flue gas, rather than in air, to produce a flue gas which comprises mainly carbon dioxide and water, rather than nitrogen and carbon dioxide. The CO₂ concentration in the flue gas from an oxyfuel combustion firing system is, therefore, significantly higher than in the flue gas from an air firing system, and hence the CO₂ can be cleaned, compressed and stored with significantly fewer downstream processing and purification steps than would be necessary with flue gas from air firing.

Oxyfuel firing technology for boiler plants based on pulverised fuel or fluidised bed combustion has now been demonstrated at large pilot scale and the major emphasis of future development work on this technology is on full scale boiler demonstration projects to provide detailed engineering information for the first fully commercial projects. This technology can be applied equally well to biomass co-firing in a pulverised coal boiler and to a dedicated biomass boiler.

The behaviour of the combustion system depends largely on the degree of flue gas recirculation, the composition of the recycled flue gases (RFG) and also on how the gases are mixed with the fuel and the oxygen. Flue gas recirculation in large coal fired boilers is a well established technique, commonly employed in a large number of plants for the control of NO_x emissions and for reheat steam temperature control.

The majority of the current development and demonstration work is therefore concerned with the demonstration of the mixing of the oxygen with the recirculated flue gas stream, with the optimisation of the combustion system, and with the control the combustion system and boiler under oxyfuel conditions at industrial scales of operation, and during start-up, shutdown, boiler and mill trips, restarts, etc. It should be noted in this context that the operation of the ASU has a significant negative impact on the operational flexibility of the power plant. This can be mitigated to

some extent by the provision of large scale oxygen storage; however the practicality of this is site specific.

There is currently significant R&D effort on the boiler tube corrosion issues because of the potential risks of a significant increase in the concentrations of CO₂, H₂O, sulphur oxides and HCl in the flue gases in contact with the high temperature boiler surfaces within the flue gas recirculation loop. This is relatively long term work and this is aimed at the development of suitable predictive corrosion rate models, which can inform the design decisions about boiler tube materials for particular duties, and at the development and demonstration of the appropriate corrosion mitigation techniques.

Chemical looping combustion technology is a potential alternative to CO₂ capture by oxyfuel firing. This technology is in the fairly early stages of development and requires a completely different approach to combustion technology. It involves the use of a solid state oxygen carrier rather than a gaseous oxidant. The application of this technology to a solid fuel i.e. coal or solid biomass will normally require the endothermic gasification of the fuel with air, steam or CO₂, and this has implications for the complexity of the overall process. The required heat for the gasification process may be provided by the exothermic air oxidation reaction, the circulation of hot oxygen carrier into the gasifier/fuel reactor or from the fuel combustion in the main reactor.

The chemical looping combustion plant evaluated in Case Study 6, however, uses a Cu-based oxygen carrier, cycled between the states of CuO and Cu₂O. One advantage of this choice of oxygen carrier is that under certain conditions, CuO dissociates to Cu₂O and gaseous O₂ via its characteristic “uncoupling” reaction. In this way, a chemical looping oxygen uncoupling (CLOU) system is adopted, whereby the fuel is fed directly into the fuel reactor without prior gasification, and combusted using gaseous oxygen.

Again, as with most of these systems, the hot solids circulation rate is reasonably high, and the key technical issues are associated with the attrition rate of the solid particles in the CFB environment, with the practical difficulty of the movement of the hot solids at the required rates and with good control of the circulation rate between the reactors at industrial scale. The physical degradation of the Cu oxygen carrier with time reduces the efficiency of the system, and there may be significant costs associated with sorbent replacement. In this case, all of the heat transfer to the boiler steam circuit is from the heat exchangers in the reduction and oxidation reactors. These operate at temperatures between 600-900°C for the reduction reactor and 700-1000°C for the oxidation reactor.

The key technical challenges facing CLC are the scale-up to commercial utilisation and the optimisation of the key process parameters, i.e. the reactor temperatures, the solids circulation rate, the reactor dimensions, and the air-to-fuel ratio, to ensure a sufficiently high conversion rate of the fuel to CO₂ and H₂O. There has been some industrial interest in this technology. Alstom, along with Air Liquide are partners of the ÉCLAIR project, and have completed construction of a 65 kW_{th} test facility in Windsor, Connecticut in 2003. The facility has subsequently undergone successful testing with a CaSO₄ oxygen carrier, and a 3 MW_{th} pilot facility is currently under development.

Overall, the development of conventional oxyfuel combustion is continuing at small demonstration scale with significant industrial involvement. The further development of chemical looping

combustion technology will very much depend on the results of the pilot scale work planned over the next 5-10 years.

Pre-combustion technologies

The **co-firing of biomass in a large coal-fired Integrated Gasification Combined Cycle plant**, with carbon dioxide capture by physical absorption is by far the most advanced of the pre-combustion technologies. There are many years of operational experience in numerous industrial chemical plants, including natural gas desulfuration and decarbonation processes, as well as processes involved in the production of hydrogen, ammonia and other important process gases.

The syngas from an IGCC plant is similar in many ways to the gases processed in chemical plants, i.e. they are at high pressure, and have low concentrations of dust and of O₂, SO₂, NO₂ and other pollutants. Consequently, there are relatively few technical barriers to full commercialisation in the current context. It should be noted in this context, however, that oxygen blown gasification systems currently employ cryogenic air separation units, and the operation of these units have a significant negative impact on the flexibility of operation of the power plant. As stated above, this can be mitigated to some extent by the provision of large scale oxygen storage, however the practicality of this is site specific.

The majority of pre-combustion capture applications to date have been, however, in the chemical industry, not the power sector, and there is a general lack of successful industrial experience in integrating pre-combustion capture systems within power generation systems at a commercial scale. It is anticipated, however that the relevant experience may be gained as CO₂ capture systems are progressively deployed in the proposed new fleet of coal IGCC demonstrations, with and without biomass co-firing, that are being planned at the present time for implementation across the world over the next few years. The application to biomass co-firing will be largely incidental to these developments, but do represent a significant opportunity for the deployment of biomass-CCS technology.

BIGCC technology, a dedicated biomass IGCC, with carbon dioxide capture by physical absorption, shares many of the advantages of co-firing IGCC plants as discussed above. However, there exists a relative lack of experience of biomass gasification at the 30-50MW_e scale. Further, the requirement for an ASU to supply oxygen for the biomass gasification unit and fairly complex materials handling and metering requirements will mean that it is most likely that the front end of the process will not operate in load-following or flexible modes but rather under baseload conditions.

Downstream of the CO₂ removal process, the plant comprises essentially a hydrogen CCGT plant. Provided that there is sufficient storage capacity for the hydrogen, the upstream and downstream sections of the plant can be decoupled, and the CCGT part can be operated in a fairly flexible fashion. In this context, it is clear that the dynamic control, and start-up and shut-down, of industrial gas turbines is well-established. This mode of operation would enable the plant operator to access premium peak-demand payments. The safe storage of hydrogen at reasonable scales of operation has already been demonstrated.

Economic analysis

During this high-level engineering evaluation of all the eight technology combinations (WP2), two major challenges were faced by the TESBIC team:-

- The first one related to the wide spectrum of technology readiness levels, TRLs of the eight technology combinations studied, in which the TRLs extended from as low as 3-4 to as high as 7-8.
- The second challenge rose from the lack of a “one size fits all” engineering software, i.e. not a single common software platform exists which can be utilised in the course of the project to evaluate all eight technology combinations.

In order to tackle these two issues, the TESBIC consortium implemented the following measures across the consortium:

- In addition to estimating the main techno-economic parameters such as the TRLs, specific CAPEX etc., the uncertainties associated with the estimation of these parameters were also identified and/or specified.
- The estimation of the overall electrical generation efficiency values for the technologies is also clearly affected by the TRL. For Case studies 1-4 and 7, where good quality data for current technology at state of the art steam conditions, are available, these have been employed in the assessment. For Case Studies 5,6 and 8 where good quality data are not available and the level of engineering detail of the configuration of the process plant and heat exchangers are not available, significant speculation about future operating conditions was required.
- The TESBIC team employed a total of four software platforms ranging from commercial flow-sheeting software, validated open-source power plant and carbon capture simulation tools to in-house bespoke academic computer programmes to assess the eight technologies in detail.
- AS far as was possible, the basic economic assumptions and the cost estimates for a number of common items of equipment were harmonised. The methodology is explained in the next section.

Harmonised inputs for the economic calculations:

To help ensure that the overall economic parameters can be compared across the eight technology combinations, cost estimates for a number of the more common items of equipment, etc. have been prepared for use in this work.

The list of values, i.e. capital and operating costs for specific items of plant which were set as inputs for all the eight of the technology combinations studied are listed in Table 41. As far as was possible, these were applied to the economic analyses presented in the Case Studies above and in the economic comparison tables presented below. In addition to the harmonised economic parameters, the following assumptions were made in preparing the overall CAPEX and OPEX estimates for the eight technologies.

Additional capital costs (£m)

- Operation and utilities: 5% of Total Installed CAPEX (TIC)
- Civils and land costs: 10% of TIC
- Project development costs: 5% of TIC
- Contingency: 10% of TIC

Fixed operating costs (£m/yr)

- Maintenance and labour: 4% of TIC/yr
- Insurance: 1% of TIC/yr

Feedstocks

- UK coal: 1.97 £/GJ
- Global bituminous coal: 3.40 £/GJ
- UK forestry wood chip: 2.8 £/GJ [£50/odt]
- Traded wood pellet: 7.5 £/GJ [£135/odt]

Lastly, the plant capacity factor was assumed at 85%.

Table 44: Cost estimates for common equipment etc. used across all 8 technology combinations.

Description	Value	Units
CO ₂ Compressor CAPEX	880,000	£/MW
CO ₂ Compressor OPEX	0.164	MWh/tCO ₂
Air Separation Unit (ASU) CAPEX	250,000	£/MW
Air Separation Unit (ASU) OPEX	0.2319	MWh/tO ₂
Steam turbine system CAPEX	218,000	£/MW
Steam plant (boiler island) CAPEX	500,000	£/MW
Limestone	17.6	£/tonne

Cost comparisons

The key performance and economic parameters for the eight technology combinations studied in WP2, have been summarised in the Tables 42a and 42b below. All the eight combinations are also represented in terms of their three main measures of performance and costing, i.e. overall net efficiencies, CAPEX and OPEX in Figure 14a.

Looking in the first instance, at the gross and net electricity generation efficiency values obtained for all eight technology combinations, the negative impacts of the carbon dioxide capture technologies was evident. In the cases studied, the operation of the CO₂ capture and compression/cleaning systems wherever directly comparable, decreased the net electrical efficiency of the power plant in the range of 6 to 13% points. This also had the effect of increasing the specific investment costs significantly. In the cases where direct comparison was possible, the rise in specific CAPEX was observed in the range of 45 to 130%. The impact of the operation of the CO₂ capture plant on the total annual O&M costs of the plants was observed to be more modest (range, 4% to 58%). The specific O&M cost per MWh of power is clearly much higher for the cases with CO₂ capture, principally because of the significant reduction in the net power output of the plant.

The data in Table 42a also illustrate the difficulties the industry has in identification of the preferred technology between solvent scrubbing and oxyfuel firing. The estimated capital costs of these technologies and the impacts on the net plant efficiencies are generally of the same order. The oxyfuel technology involves the installation and operation of an ASU, with the associated costs penalties and impacts on power plant flexibility of operation. The solvent scrubbing process does not require an ASU but has a similar impact on the net power plant efficiency, and there are additional environmental and other issues associated with the on-site storage handling and use of large quantities of organic chemicals.

The specific OPEX + fuel cost estimates for the technology combinations listed in Table 42a and Table 42b indicate that the specific operating costs for the large coal power plants co-firing biomass, and IGCC plants are significantly lower than those for the smaller dedicated biomass combustion and biomass gasification/BIGCC power plants. The differences between the specific fuel + OPEX costs for the solvent scrubbing and the oxyfuel firing processes are fairly modest.

As the study involves technology combinations, and their associated individual technology components, spanning a wide range of technology readiness levels, from as low as TRL 3 to 4, up to 7 to 8, it was deemed necessary to account for the uncertainties associated with, for example, the TRLs and the specific CAPEX in addition to the data presented in Table 42a and Table 42b. Figure 14b shows the scatter of the eight technology combinations (i.e. with CCS) and the associated uncertainties in the TRLs and the specific CAPEX costs. It is evident from this data, that the uncertainties associated with economic parameters for relatively better-known and understood technology combinations are low, i.e. of the order of 20%, whereas for relatively new technologies, the uncertainties were considered to be in the range of 35 to 60%.

Links with the next work packages

The data collected in WP2 for the key output variables, i.e.

- The capital cost estimate,
- The non-fuel operating cost estimate
- The overall generation efficiencies,
- The CO₂ emissions,

- The SO₂ emissions, and
- The NO_x emissions

as a function of the four inputs, namely,

- The biomass co-firing levels,
- The extent of carbon capture,
- The nameplate capacity, and
- The operating capacity

are being utilised for sub-model parameterisation in work package 3.

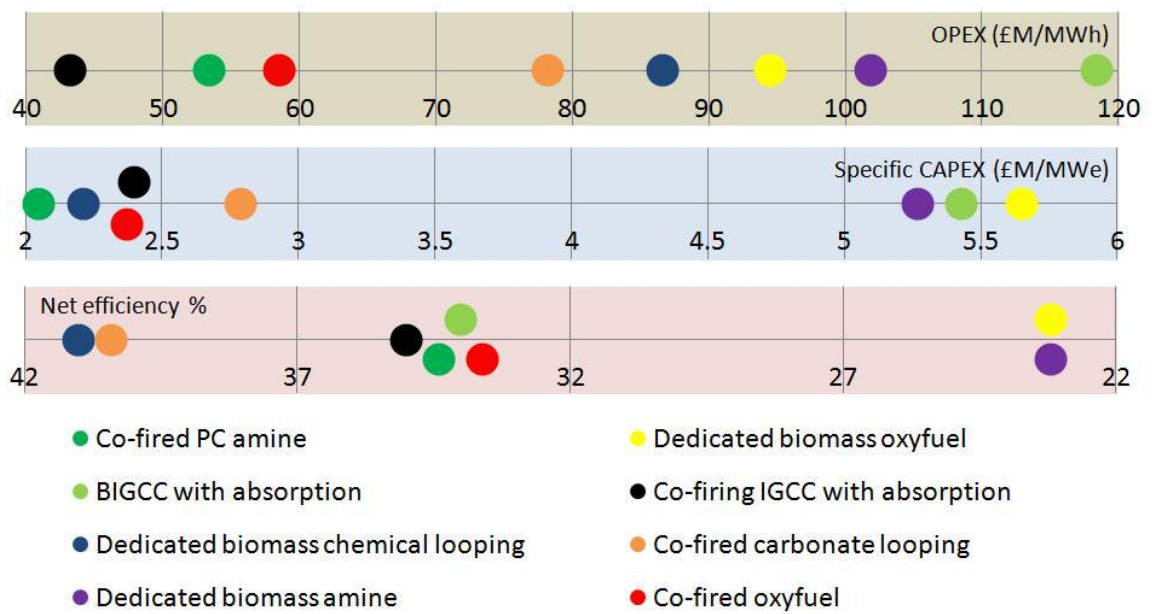
This sensitivity data matrix is being presented in more detail in WP3 deliverables focusing on the modelling strategy, and the user documentation. The data presented including all the information on the associated uncertainties will be directly utilised to carry out benefits assessment of specific development, demonstration and deployment activities in WP4.

Table 45a: Summary of economic parameters for technology combinations 1 to 4 with and without carbon capture.

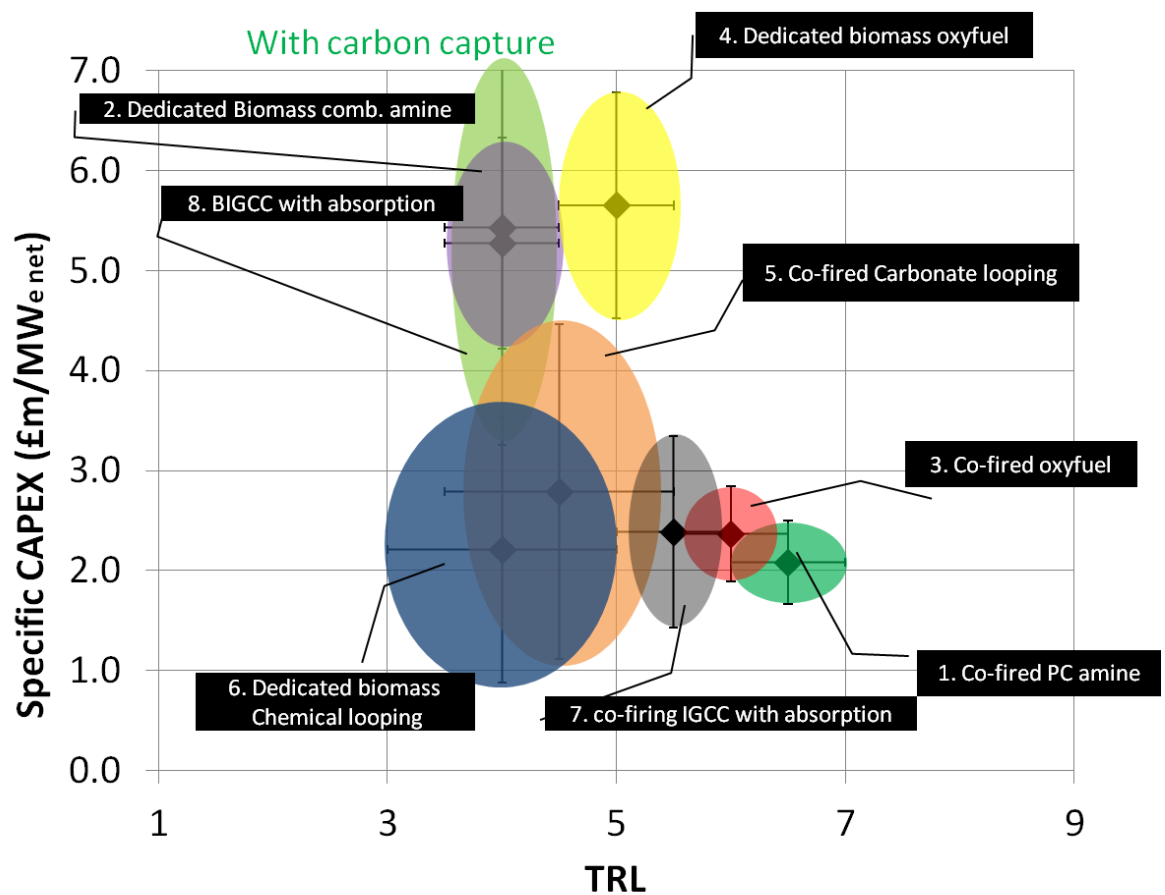
	1 Biomass co-firing in pulverised coal boiler		2 Dedicated biomass combustion		3. Biomass co-firing in pulverised coal boiler		4 Dedicated biomass combustion	
	Without CCS	with solvent scrubbing	Without CCS	With solvent scrubbing	Without CCS	With oxyfuel firing	Without CCS	With oxyfuel firing
Gross power (MW _e)	545.2	474.1	83.5	68.5	545.2	545.2	83.4	84.6
Net power (MW _e)	518.9	398.9	75.8	48.9	518.9	388.7	75.79	48.85
Gross efficiency %	47.1	40.9	39.6	32.5	47.1	47.1	39.5	40.1
Net efficiency %	44.8	34.4	35.9	23.2	44.8	33.6	35.9	23.2
Total installed cost (£M)	509.3	638.2	143.5	198.1	509.3	709.1	143.5	213.2
Total investment cost (£M)	662	830	187	258	662	922	187	277
Specific CAPEX (£M/MW _e)	1.28	2.08	2.46	5.27	1.28	2.37	2.46	5.65
Total O&M costs (£M/yr)	146	158.5	33.3	37.1	175	173.4	33.3	34.3
Energy generated (MWh/yr) x 10 ⁶	3.86	2.97	0.56	0.36	3.86	2.89	0.56	0.36
OPEX & fuel (£/MWh)	37.78	53.36	59	101.9	45.64	58.52	59	94.5

Table 45b: Summary of economic parameters for technologies 5 to 8 with and without carbon capture.

	5 Carbonate looping		6 Chemical looping		7. IGCC co-firing with CO ₂ absorption		8 BIGCC	
	Without CCS	with carbonate looping	Without CCS	With chemical looping	Without CCS	With CO ₂ capture	Without CCS	with CCS
Gross power (MW _e)	N.A.	326	N.A.	325	549	549	64.7	73
Net power (MW _e)		300		300	484	461	47.5	40
Gross efficiency %		43.9		44.5	45.8	41.7	56.0	56.0
Net efficiency %		40.4		41.0	40.4	35.0	41.0	34.0
Total installed cost (£M)		644		509	650	848	109	167
Total investment cost (£M)		837		662	845	1102	142	217
Specific CAPEX (£M/MW _e)		2.79		2.21	1.74	2.40	2.98	5.43
Total O&M costs (£M/yr)		174.70		192.36	128.47	148.45	22.55	35.25
Energy generated (MWh/yr) x 10 ⁶		2.23		2.23	3.61	3.43	0.35	0.29
OPEX & fuel (£/MWh)		78.20		86.60	35.60	43.20	63.80	118.40



(a): Main performance and economic parameters



(b): Capital costs vs. technical risk

Figure 14: Accounting for uncertainties in specific capital costs (£M/MWe net) and technical risk represented in terms of technology readiness level (TRL) for all eight technology combinations with carbon capture.

SUMMARY

Current status of CO₂ capture technologies in the electricity supply industry:

- To date, there has been relatively little activity on the application of the carbon capture technologies to dedicated biomass power plants or to coal power plants which are co-firing biomass.
- The industry is now moving towards large scale demonstration projects on fossil-fuel based power technologies, but because these are relatively expensive and require significant government subsidies, progress on the demonstration phase has been relatively slow.

TESBiC approach:

- Four software platforms were employed and a harmonised set of boundary conditions and assumptions were implemented across the eight technology combination considered here, in order to facilitate the comparison of the techno-economic parameters (e.g. process efficiencies, CAPEX, OPEX).
- Process engineering calculations for mass and energy flows, equipment lists, high level control philosophy, environmental performance, basic process economics and existing gaps in understanding as well as future development requirements were evaluated for the eight technology combinations

Key observations:

- TRLs for the eight technology combinations (including the TRLs for the associated individual components) considered here, varied over a wide range from TRL 3 to TRL 8.
- Wherever direct comparison was feasible, it was observed that the efficiency (net) drop with carbon capture varied in the range of 6 to 13% points., the specific investment costs increased significantly (range of 45 to 130%) and the annualised operating and maintenance costs were relatively modest (between 4% to 58%).

Linking with WP3 and WP4:

- The WP2 data collected for the seven technical and economic output variables as a function of the four process input variables are being utilised for sub-model parameterisation in Work Package 3. The sensitivity data matrix is also being presented to the ETI in one of the WP3 deliverables.
- The technical and economic data and the associated uncertainties presented above will be directly utilised to carry out the proposed benefits assessment of specific development, demonstration and deployment activities in WP4.