EVALUATION OF PROCESS CONTROL STRATEGIES FOR NORMAL, FLEXIBLE AND UPSET OPERATION CONDITIONS OF CO_2 POST COMBUSTION CAPTURE PROCESSES

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EVALUATION OF PROCESS CONTROL STRATEGIES FOR NORMAL, FLEXIBLE AND UPSET OPERATION CONDITIONS OF CO_2 POST COMBUSTION CAPTURE PROCESSES

Key Messages

- Electricity market models suggest power plants with carbon capture and storage (CCS) will need to adopt flexible operation in the future. Appropriate control strategies will be necessary to ensure their ability to operate in such a market and their profitability.

- An evaluation of process control strategies for normal, flexible and upset conditions of CO_2 post-combustion capture (PCC) processes (considered the leading technology for deployment in the power sector) based on amine scrubbing has been undertaken.

- This work used a high-fidelity modelling tool that can describe the dynamic operation of the CCS chain to investigate 3 different process control strategies for both pulverised coal (PCPP) and combined cycle gas turbine power plants (CCGT), with PCC.

- The power plant modelling showed the performance of the CO_2 capture unit can be maintained even during periods of significant load fluctuation, using industry standard control techniques, thus avoiding other more expensive solutions.

- Manipulating the solvent flow rate generally provided better control of the CO_2 capture rate than varying the solvent lean loading, as it results in less oscillation, i.e. more constant hydraulic conditions in the CO_2 capture plant.

- For the PCPP, a control strategy that manipulates the CO_2 capture rate by varying the solvent flowrate is the more profitable option. For the CCGT, all strategies provided the same benefit, due to the dilute nature of the CCGT flue gas.

- The CO_2 capture plant was able to continue operation for a limited amount of time, i.e. 3.5-5 hours, in case of hazardous events, such as injection shutdown or loss of compression.

- In conclusion, this study has shown that simple and well-tuned control strategies can maintain critical operational parameters of a CO_2 capture plant.

- The authors recommend further work in this area could include development of advanced control strategies and fine-tuning of the existing modelling and simulation tools. This is not something IEAGHG would take up at this time but could be pursued by model developers and academia. In addition, IEAGHG recommends evaluation of faster power plant ramp up rates and other systems that provide more flexibility and easier integration into the host plant.

Background to the Study

It is important for power plants to be able to operate flexibly to respond to changes in consumer demand for electricity. Flexibility is also becoming increasingly important due to the greater use of other low carbon generation technologies, particularly variable renewable generators. The issue of operating flexibility of power plants with carbon capture and storage (CCS) has been the subject of a previous technical study by IEAGHG[1]. This report contributes to the knowledge base on flexible operation of power plants with CO_2 capture by focusing on process control issues.

A team from Imperial College London and Process Systems Enterprise has undertaken this work for IEAGHG.

The study focuses on performing an evaluation of process control strategies for normal, flexible and upset operation conditions of CO_2 post-combustion capture (PCC) processes based on solvent scrubbing. PCC is currently the leading near-term technology for large-scale deployment of CO_2 capture in the power generation sector.

Scope of Work

The aim of this study is to develop process control strategies, to select appropriate control variables for a PCC process, and design efficient control structures for operation of a PCC process with minimum energy requirements for coal and natural gas fired power plants.

To this end, the scope of work is:

1. Identify the different operating regions that are relevant to the flexible operation of coal and natural gas fired power plants.
2. Identify sets of controlled and manipulated variables that are commonly used for PCC processes.
3. Develop control strategies for feasible and economically efficient operation of PCC processes under normal and part-load operating conditions.
4. Evaluate the performance of coal fired and natural gas fired power plants with PCC for each of the proposed process control strategies.
5. Evaluate the effect of different process control strategies on the economics of PCC processes.
6. Evaluate the impacts of upset conditions on the full CCS chain.

[1] Operating flexibility of power plants with CCS, IEAGHG report 2012/6, June 2012.

Findings of the Study

Operating regimes

Electricity market models (MOSSI and UCCO) were used to produce a set of scenarios for the future operation of CCS plants. A model was then used to identify the dispatch pattern of individual generating units in the period of 2030s to 2050s. Expansion of nuclear and wind capacity during this period forces thermal plants to adopt more dynamic operating patterns, with reduced running hours, more stop/start cycles and greater ramping rates.

Three coal CCS units were studied, with low, mid and high positions within the merit order. Figure 1 shows their operating profiles in the 2030s, 2040s and 2050s. The increase in dynamic operation is evident as time progresses, with more frequent modulation between maximum and minimum stable operation and more time spent not operating. This emphasises the importance of control strategies to cope with changing plant loads.

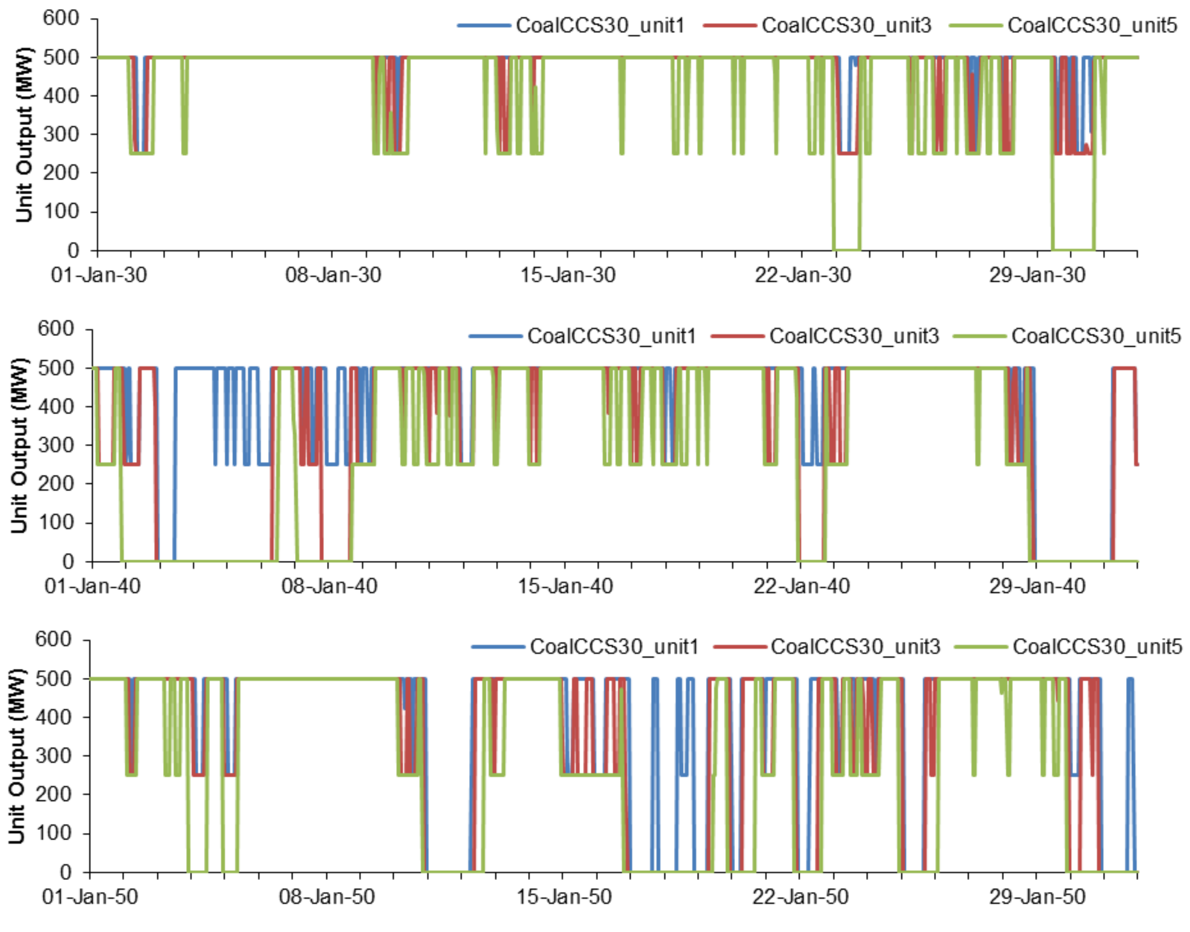

Figure 1 Operating profiles for three coal CCS units in the 2030s, 2040s and 2050s

Similar patterns were observed for gas CCS plants, although they start with a more dynamic operating pattern in the 2040s, when gas CCS plants start to be introduced into the electricity system, due to their being priced as mid-merit as opposed to base-load capacity.

The above analysis shows that coal and natural gas fired power plants will need to be ready for increasing demands of flexibility. These operating scenarios were used to inform the dynamic operating scenarios employed in subsequent sections of this report. Finally, the implication is that emphasis must be placed upon optimising efficiency and profitability during two-shift operations. The remainder of this study focuses on this assumption.

Control strategies

Reference plant models

The reference case models of power plants with CCS for this study were developed using the gCCS toolkit from Process Systems Enterprise. This toolkit includes high-fidelity models that describe the dynamic operation of all the stages of the CCS chain: power generation, post-combustion carbon capture, compression, transportation, and storage.

The reference case supercritical pulverised coal power plant (PCPP) developed for this project has a net electricity output of 779 MW at 100% load without CO_2 capture. Adding an amine-based post combustion capture plant reduces the net output to 621 MW and captures 90% of the CO_2 in the flue gas.

The reference model for the combined-cycle natural gas fired power plant (CCGT) has a net electricity output of 740 MW at 100% load without capture. Capturing 90% of the CO_2 reduces the net output to 643 MW.

Control strategies

The control strategies proposed in this work consist of combinations of controlled and manipulated variables in the carbon capture plant, based on a review of the literature on PCC process control, and on the feedback received from industrial operators and CCS technology providers. Most approaches to PCC process control suggest that the percentage of CO_2 capture should remain approximately constant throughout the power plant operation. This study follows this principle by considering two alternative strategies for controlling the CO_2 capture rate. Additionally, a strategy in which the control loops are dynamically switched during the change in power plant load is considered. The control strategies can be summarised as:

- Control strategy 1: the amount of CO_2 captured is controlled by varying the *lean solvent flowrate*. The reboiler temperature is kept constant by varying the *steam flowrate*.
- Control strategy 2: the amount of CO_2 capture is controlled by varying the *solvent lean loading*. The lean solvent loading is controlled to the required value by varying the *reboiler temperature*.
- Control strategy 3: dynamic switching between strategies 1 and 2.

Performance of control strategies

The performance of PCPP and CCGT plants was modelled for a scenario involving full load operation during the daytime and turndown to 60% capacity factor overnight. Figure 2 shows load factors and electricity prices that are assumed in this study.

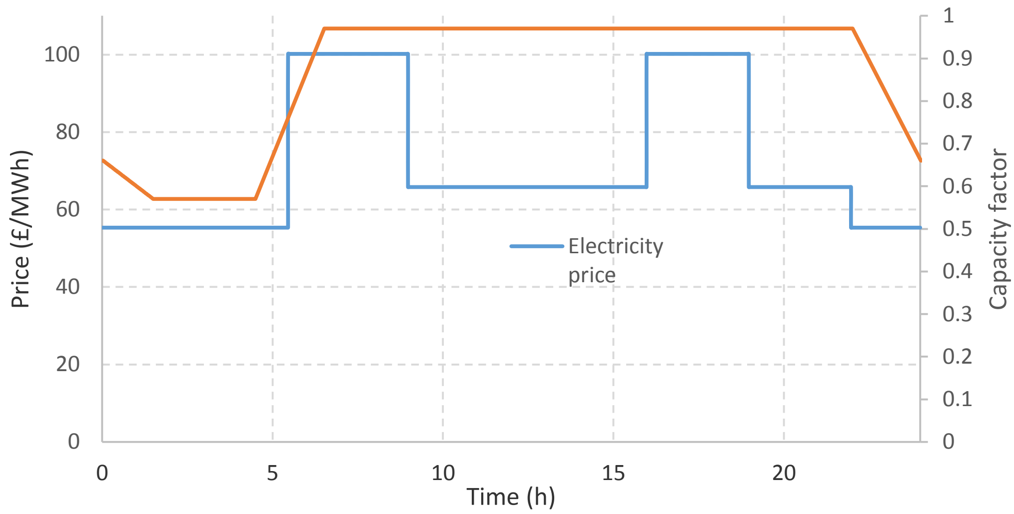

Figure 2 Variation of electricity prices throughout the day, and typical operation profile for two-shifting power plant

Figure 3 to Figure 6 present the behaviour of some key variables for the PCPP. It was observed that employing Control Strategy 2 resulted in a greater degree of oscillatory behaviour and was difficult to tune. Ultimately, Control Strategy 1 was, in this study, easier to employ and better control was observed. In all cases, it is notable that, following the disturbance, i.e., power plant ramping up or down, the CCS plant returns to steady state relatively quickly.

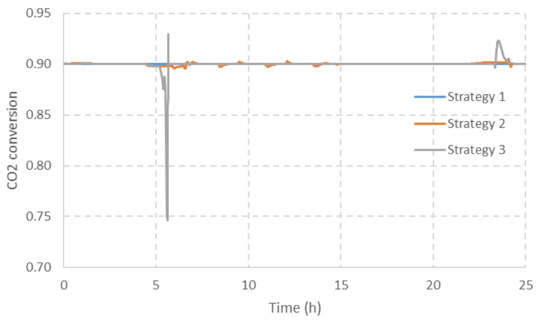

Figure 3 CO_2 capture

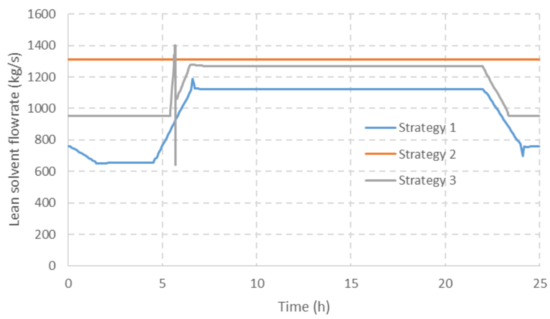

Figure 4 Lean solvent flowrate

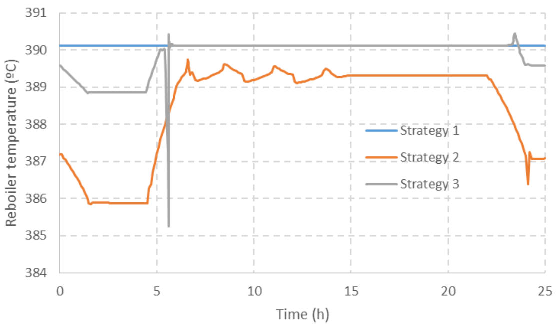

Figure 5 Reboiler temperature

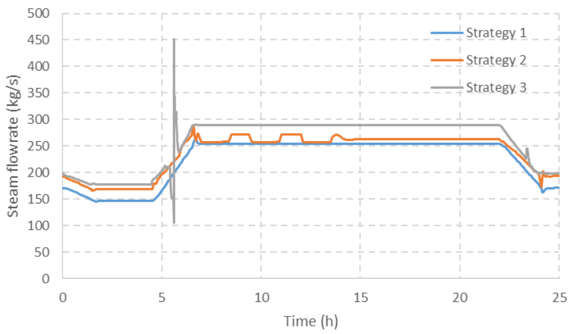

Figure 6 Steam flowrate

The duration of the relaxation time and the amplitude of the oscillations are however a strong function of the control strategy chosen. Strategy 1 appears to provide the most stable operation of the power plant, Strategy 2 follows, but with an appreciably longer relaxation time, and finally Strategy 3 has a relatively short relaxation time but comparatively greater amplitude of oscillation from the set point.

The performance of the CCGT plants follows similar trends but in general the magnitudes of the variations are lower (see full report for figures).

Economic evaluation of control strategies

The short run marginal operating costs and electricity revenues of the plants were calculated to determine their "profitability" over the course of a day. The marginal operating cost is assumed to consist of the fuel cost, the cost associated with emitting CO_2 and the cost of cooling water utilities. No fixed costs are considered, as they will be same for all of the cases. The electricity price is expected to vary during the course of a day. The plant was assumed to operate in a two-shifting mode, with the electricity prices shown in Figure 2.

The economic assessment found that for PCPP Control Strategy 1 is more profitable than either strategy 2 or 3. The analysis for CCGT showed all control strategies to provide the same profit. This is due to the dilute nature of the exhaust gas and the resulting greater solvent circulation rate relative to the PCPP case. Figure 7 and Figure 8 show the profits for both plants during a given day.

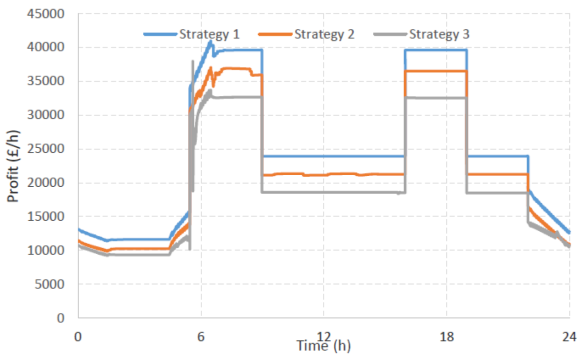

Figure 7 Profit of PCPP

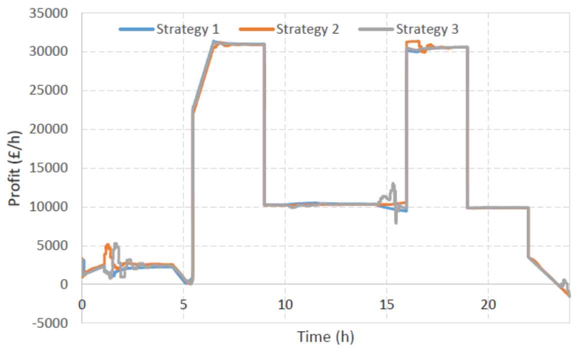

Figure 8 Profit of CCGT

Outages in the CCS chain

The study identifies and analyses potential trip, outage and shutdown scenarios and describes the effects on stakeholders up and down the CCS chain. It demonstrates how the process can be safely managed in the event of a system failure. To assess the impact of these hazards, a CCS chain including a coal-based post-combustion capture plant, pipeline transmission of CO_2 and geological storage was simulated. The system was assumed to include a capture and compression plant, a 20km onshore CO_2 pipeline followed by a 200km offsite pipeline, and 4 injection wells. It should be noted that only the downstream part of the CCS chain was considered. Further work will be needed to assess the impacts on the capture plant of upstream outages in the power plant, FGD etc.

The study considered the following scenarios:

- Unplanned shutdown at injection site

- Loss of upstream compression

Unplanned shutdown at injection site

When the injection site is shut down, the pressure within the pipeline and the CO_2 compressor speed will begin to increase. The capture plant could continue to operate as normal for slightly less than 3.5 hours until the compressor reaches its maximum speed, after which the capture plant would have to be shut down or begin venting CO_2.

Loss of upstream CO_2 compression

The failure of the compression facility will have the following two significant effects:

1. It will cause the pipeline pressure to drop, leading to loss of injection capability and in due course probably a trip of the injection well.
2. The loss of the ability to remove CO_2 from the capture process, leading to either a shutdown of the capture plant and consequent shutdown of the power plant or initiation of CO_2 venting.

As the first situation will presumably occur very quickly following the loss of compression, this was not included in the scenario.

In order to simulate the loss of a compressor, the flow of inlet CO_2 supply to the pipeline was dropped to zero in a step change. This intends to represent the most drastic situation. Figure 9 shows that it takes over 4 hours for the CO_2 supply at the injection site to reduce to 50% of the design flow, following a complete loss of supply.

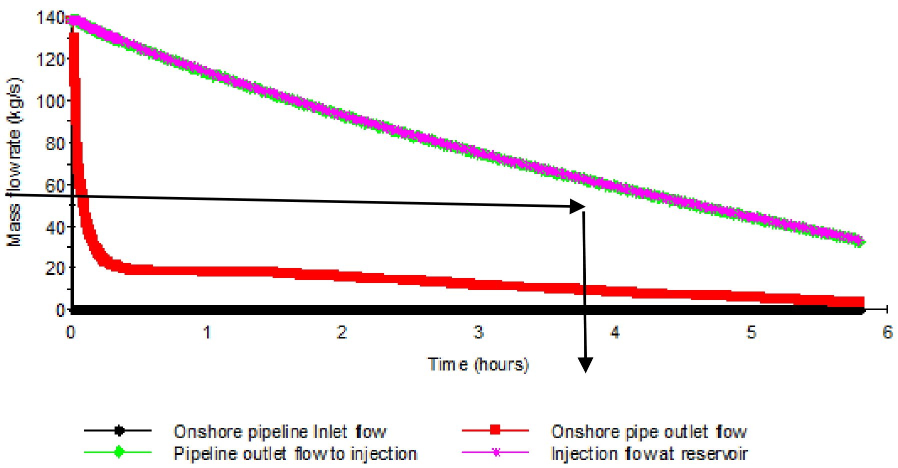

Figure 9 Mass flowrates at locations in the CO_2 transmission and injection sub-systems

During normal operation, it is expected that a CO_2 pipeline would operate above the critical pressure of CO_2, which is 73.8 bar, although the location of the critical point is a function of the composition of the stream, with the critical point of an impure stream typically being at a higher pressure than that of pure CO_2. In the event of a loss of inlet compression, there is the potential for two-phase flow to occur if the pressure of the CO_2 stream drops below the critical pressure. Based on this simulation, it is unlikely that this will occur within 5 hours of loss of supply, as shown in Figure 10.

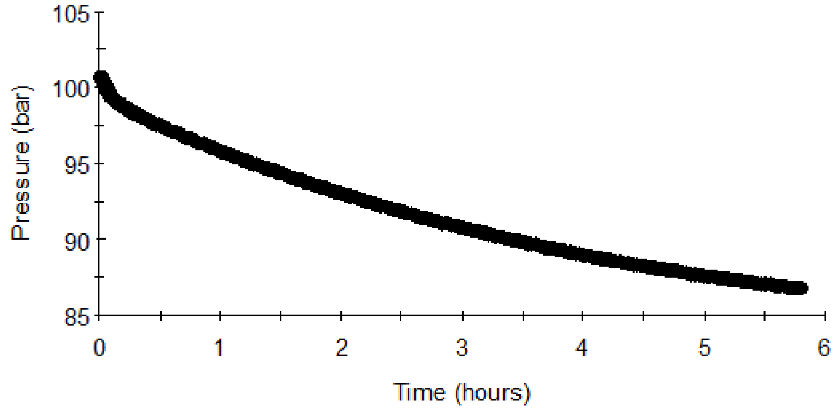

Figure 10 Minimum pressure in the transmission system

Expert Review Comments

Comments on the draft report were received from reviewers working in the field of control and dynamic performance of capture plants. The contribution of all reviewers is gratefully acknowledged.

The reviewers thought the report was in most respects of a high standard. A small number of particular areas of concern were identified. The contractor has provided responses to the reviewers' comments and made appropriate modifications to the report.

Conclusions

Energy market models support the notion that power plants will need to adopt flexible operation patterns in the future. Fluctuations between full and part-load operation will become more frequent during the day, and therefore an appropriate control strategy is necessary to ensure the operability and profitability of the process.

Several control strategies have been proposed in the literature. There is a need to design conceptual strategies and accurately predict their behaviour under flexible operating conditions using high-fidelity modelling tools. This study has made use of such a tool, gCCS.

The study has shown that with an appropriate and well-tuned control strategy it is possible to maintain critical parameters such as the CO_2 capture rate at the desired set-point, even during periods of significant fluctuation in the power plant load.

Using an appropriate control strategy, even if based in simple and well established control techniques, such as PID, avoids the need for more risky solutions such as adding solvent storage tanks to the process.

From a control point of view, using the solvent flowrate as manipulated variable to control the CO_2 capture rate is a better option than manipulating the solvent lean loading, which has more oscillatory behaviour. Maintaining a constant solvent flow rate results in more constant hydraulic conditions in the absorber and stripper columns.

The choice of control strategy can affect the economics of the plant. For the PCPP, a control strategy that manipulates the CO_2 capture rate by varying the solvent flowrate is a more profitable option than a strategy using the solvent lean loading or a dynamic switching between the two. For CCGT, all strategies provided the same benefit, due to the dilute flue gas.

From a hazard management perspective, it was observed that the capacity of the compressor to add to pipeline pressure in the event of an outage at the injection site is a limiting element in the chain. Under the scenarios investigated here, the capture plant can operate for about 3.5-5 hours in case of injection shutdown or loss of compression.

Recommendations

The output from this report and a critical review of the gaps in the literature related to control of power plants with CCS resulted in the following recommendations from the authors for further research and technical studies:

- *Modelling and simulation aspects* – further developments in high-fidelity modelling and simulation tools to include accurate descriptions of all components in the CCS chain. Additionally, testing of the control strategies proposed in this work for different solvents in the post-combustion capture system.

- *Simultaneous process and control design* – development of appropriate algorithms for including process control design decisions at the process design stage.
- *Advanced process control techniques* – development of hierarchical model predictive control formulations for the integrated system, including power plant and capture plant, in order to optimise the performance and economics of the overall process.
- *Optimisation of control strategy switching* – follow-up to this work, where the schedule of dynamic switching between control strategies is determined by an optimisation problem where the costs and benefits of each strategy are encompassed in the formulation.

The above points are not something that IEAGHG would take up at this time but could be pursued by model developers and academia. However, IEAGHG adds the following recommendations:

- It would be interesting to investigate cases that represent the impact of a changing energy sector on power plant ramp up rates.
- Other systems such as membranes should be considered due to their flexibility and "lack of integration" with the power plant.

Evaluation of Process Control Strategies for Normal, Flexible and Upset Operation Conditions of CO_2 Post Combustion Capture Processes

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April, 2016

Contents

List of figures

List of tables

List of Acronyms

CAPEX	-	Capital expenditure
CCGT	-	Combined Cycle Gas Turbine
CCS	-	Carbon Capture and Storage
COMAH	-	Control of Major Accident Hazards
DCC	-	Direct Contact Cooler
ESP	-	Electrostatic Precipitator
FGD	-	Flue Gas Desulphurisation
GGH	-	Gas-Gas Heater
GT	-	Gas Turbine
HP	-	High Pressure
HRSG	-	Heat Recovery Steam Generator
IEAGHG	-	IEA Greenhouse Gas R&D Programme
IP	-	Intermediate Pressure
KO	-	Knock-out
LHV	-	Lower Heating Value
LP	-	Low Pressure
MCR	-	Maximum continuous rating
MEA	-	Monoethanolamine
MOSSI	-	Merit Order Stack with Step Investments
MPC	-	Model Predictive Control
OCGT	-	Open-Cycle Gas Turbine
OPEX	-	Operational expenditure
PCC	-	Post-Combustion Capture
PCPP	-	Pulverised Coal Power Plant
PI		Proportional Integral
PID		Proportional Integral Derivative
SCR	-	Selective Catalytic Reduction
SP-PC	-	Supercritical Pulverised Coal
UCCO	-	Unit Commitment Capacity Optimiser
USC-PC	-	Ultra-Supercritical Pulverised Coal

Executive summary

Project overview

The IEA Greenhouse Gas R&D Programme (IEAGHG) is an international collaborative programme that promotes research and technical studies on different aspects of technologies that can reduce greenhouse gas emissions derived from the use of fossil fuels. The issue of operating flexibility of power plants with CO_2 capture has been the subject of previous technical studies by IEAGHG and received considerable attention in both academia and industry. This report contributes to the knowledge base on flexible operation of power plants with CO_2 capture by focusing on process control issues relevant to carbon capture processes subject to flexible operating conditions. The work has been commissioned by IEAGHG to a team from Imperial College London and Process Systems Enterprise.

This project focuses on performing an evaluation of process control strategies for normal and flexible operation conditions of CO_2 post-combustion capture (PCC) processes. PCC is a promising, near-term technology for large-scale deployment for the decarbonisation of the power generation and other sectors. However, the integration of this technology imposes a well-known efficiency penalty on the power plant with which it is integrated. Once an optimal process design has been identified, this energy penalty can be somewhat reduced via application of an appropriate control strategy to the PCC plant. An appropriate process control strategy is also fundamental to guarantee the safety and feasibility of the process under flexible operating conditions that the power plants may be subject to.

The aim of this project is to develop the process control strategy, to select appropriate control variables for a PCC process, and design efficient control structures for operation of a post-combustion capture process with minimum energy requirements for coal and natural gas power plants. The control structures are developed for power plant operating ranges of around 50% to 100% load.

The aims of this study are to:

1. Identify the different operating regions that are relevant to the flexible operation of coal and natural gas fired power plants.
2. Identify sets of controlled and manipulated variables that are commonly used for PCC processes.
3. Develop control strategies for feasible and economically efficient operation of PCC processes under normal and part-load operating conditions.
4. Evaluate the performance of coal fired and natural gas fired power plants with PCC for each of the proposed process control strategies.
5. Evaluate the effect of different process control strategies on the economics of PCC processes.
6. Hazard management in the CCS chain.

Study of relevant part-load scenarios

The electricity market models (MOSSI [1-4] and UCCO [3, 5, 6]) were used to produce a set of scenarios for the future operation of CCS plants. These models were designed and are employed at Imperial Business School:

- MOSSI (Merit Order Stack with Step Investments) to explore the long-term investment decisions which shape the mix of installed capacity
- UCCO (Unit Commitment Capacity Optimiser) to simulate the short-term operating decisions for individual plants in greater detail, taking account of their technical limitations.

Two technologies are considered: coal with CCS and gas with CCS. The evolving role of each technology over its lifetime is considered by taking snapshots of operations during the 2030s and beyond.

The evolution of the installed capacity in Britain, shown in Figure 1 was projected using MOSSI. The aim of this analysis was to estimate when coal and natural gas fired power plants with CCS become viable. Coal with CCS becomes viable in the 2030s once the capital cost (CAPEX) has fallen below £2,500/kW and the carbon price has risen to £76/T. 14 GW are installed in the 2030s, followed by a further 5 GW in the 2040s. After this, capacity remains stable until the 2070s (which is beyond the period we are interested in). Combined-cycle gas with CCS becomes viable in the 2040s once its CAPEX has fallen below £1,650/kW, and the carbon price has risen further to £138/T. 16 GW are installed in the 2040s, followed by 5 GW in the 2050s.

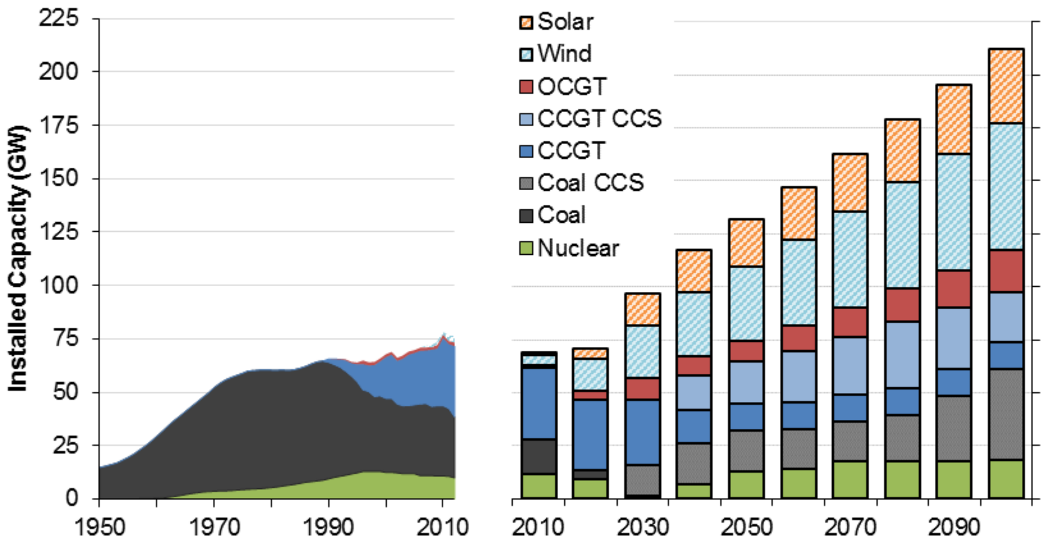

Figure 1: Evolution of installed electricity generation capacity in the UK for the period 2010 to 2100.

The UCCO model was then used to identify the dispatch pattern of individual generating units in the period of interest (2030s to 2050s). Three coal CCS units were studied, with low, mid, and high positions within the merit order. The hourly operating patterns of these units are shown in Figure 2 for the 2030s (top), 2040s (middle), and 2050s (bottom). The increase in dynamism is evident as time progresses, with more frequent modulation between maximum and minimum stable generation and more time spent off.

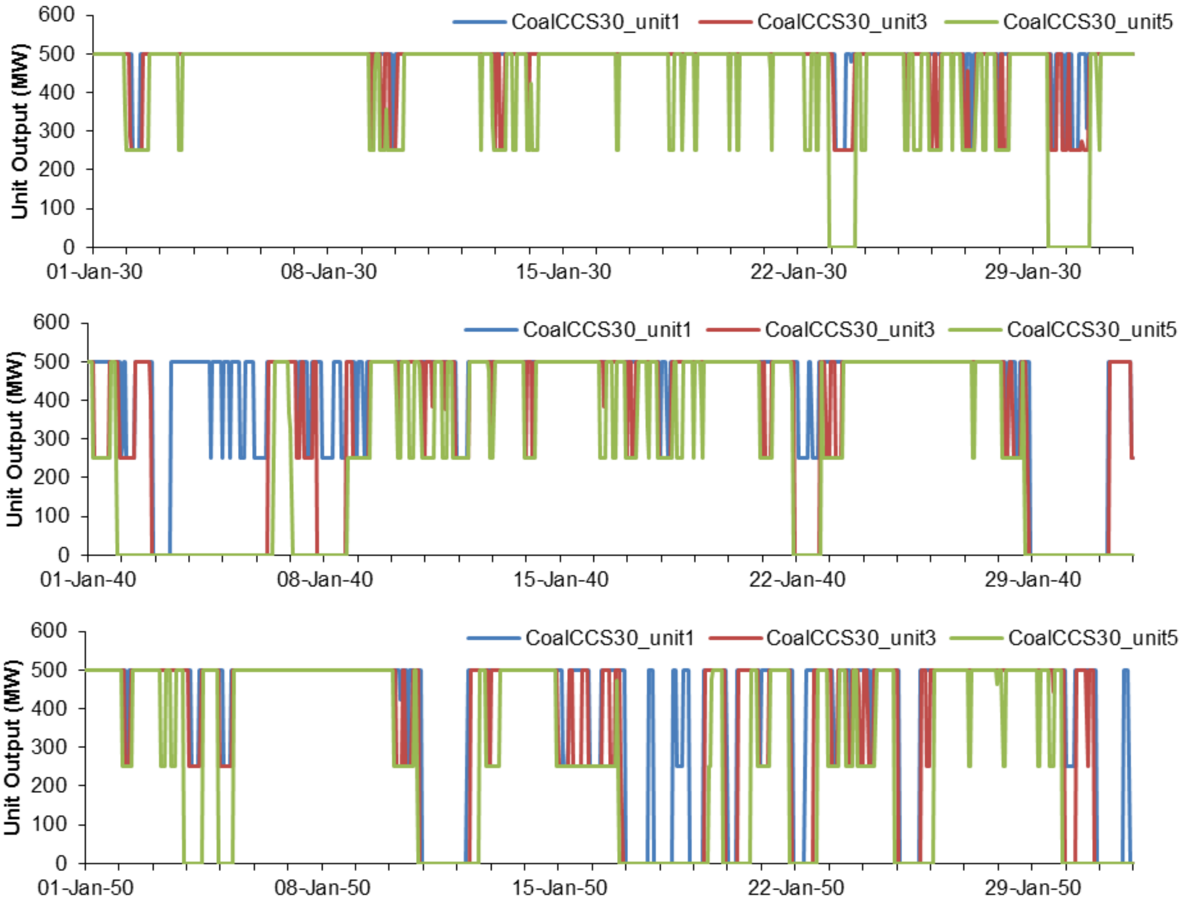

Figure 2: Operating profiles for three coal CCS units in the 2030s (top), 2040s (middle) and 2050s (bottom).

The same pattern was observed for gas CCS plants, although they start with a more dynamic operating pattern in the 2040s due to being priced as mid-merit as opposed to base-load capacity.

The above analysis shows that coal and natural gas fired power plants will need to be ready for increasing demands of flexibility. These operating scenarios were used to inform the dynamic operating scenarios employed in subsequent sections of this report.

Overview of reference case models

The reference case models for this study were developed using the gCCS toolkit from Process Systems Enterprise. This toolkit includes high-fidelity models that describe the dynamic operation of all the stages of the CCS chain: power generation, post-combustion carbon capture, compression, transportation, and storage.

Figure 3 shows a schematic of the reference case supercritical pulverised coal power plant developed for this project. At 100% load the plant produces a gross electricity output of 825MW[1], with 90% of the CO_2 in the flue gas being captured in the amine-based post-combustion capture plant[2] illustrated in Figure 4 below.

[1] Gross power produced, before capture

[2] Using a 30 wt% MEA solvent

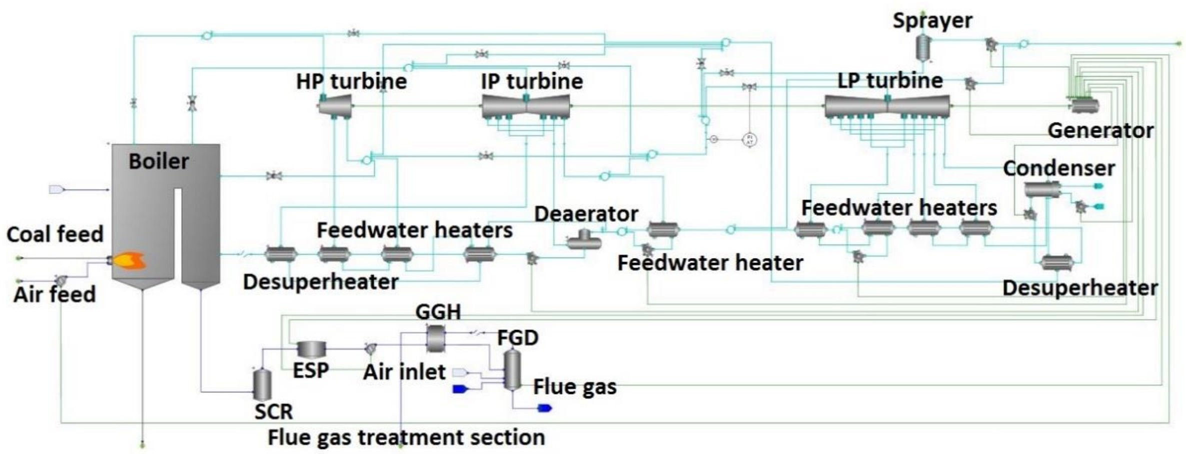

Figure 3: Reference flowsheet for the supercritical pulverised coal-fired power plant simulated in this work

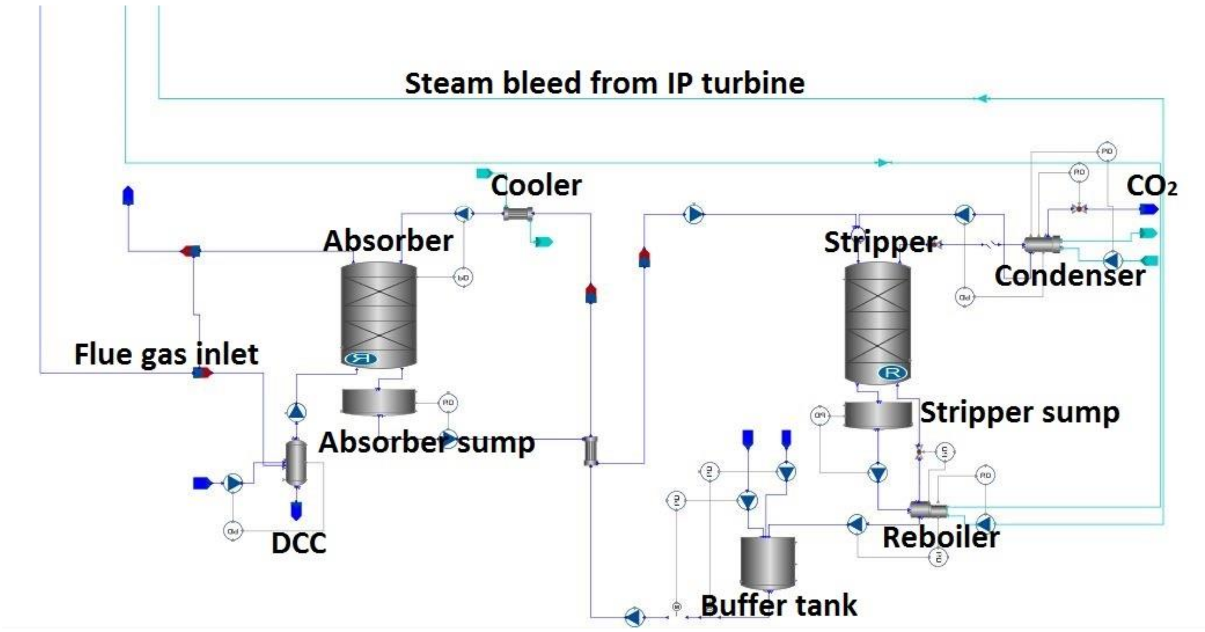

Figure 4: Reference flowsheet for the post-combustion CO_2 capture process simulated in this work

The reference model for the combined-cycle natural gas fired power plant, producing a gross electricity output of 825MW at full-load, is schematically shown in Figure 5. This plant is also fitted with an amine-based post-combustion carbon capture process as in Figure 4

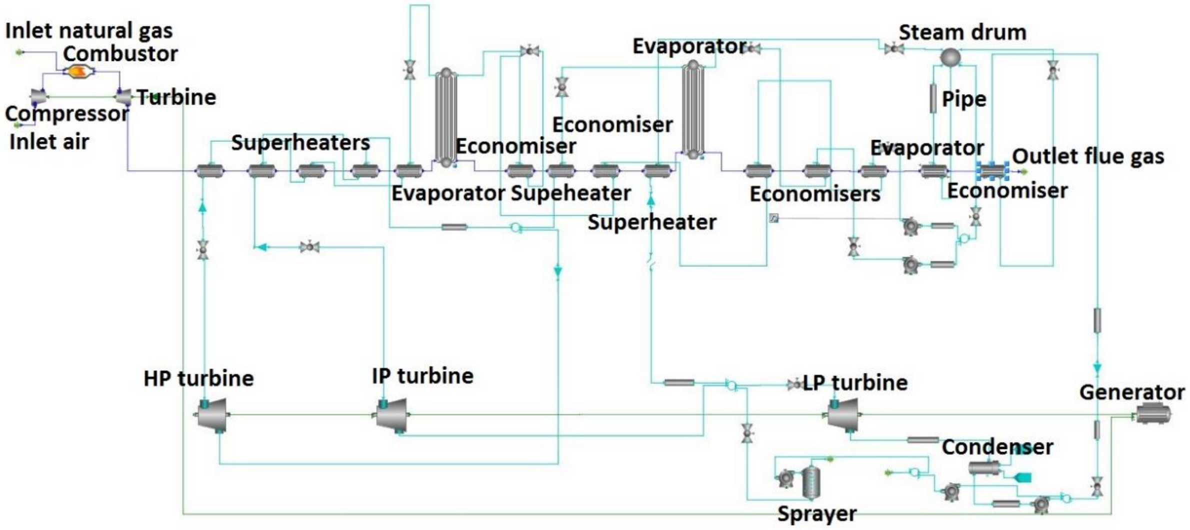

Figure 5: Reference flowsheet for the CCGT power plant simulated in this work

Proposed control strategies

The control strategies proposed in this work consist of combinations of controlled and manipulated variables in the carbon capture plant, based on a review of the literature on PCC process control, and on the feedback received from industrial operators and CCS technology providers. Most approaches to PCC process control suggest that the percentage of CO_2 capture should remain approximately constant throughout the power plant operation. The principle is followed in this study by considering two alternative strategies for controlling the CO_2 capture rate. Additionally, a strategy in which the control loops are dynamically switched during the change in power plant load is considered. The main process variables involved in each control strategy are listed in Table 1, below.

Table 1: Control strategies proposed in the project.

Strategy	*Controlled variable*	*Manipulated variable*
1	CO_2 capture	Lean solvent flowrate (MEA)
	Reboiler temperature	Steam flowrate
2	CO_2 capture	Reboiler temperature
	Lean solvent flowrate	Lean solvent loading
3	Dynamic switching between strategies 1 and 2	

Overall performance of control strategies

It was observed that employing Control Strategy 2 resulted in a greater degree of oscillatory behaviour and was difficult to tune. Ultimately, Control Strategy 1 was, in this study, easier to employ and better performance was observed. The behaviours of some key variables for the PCPP are presented in Figure 6 to Figure 9 below.

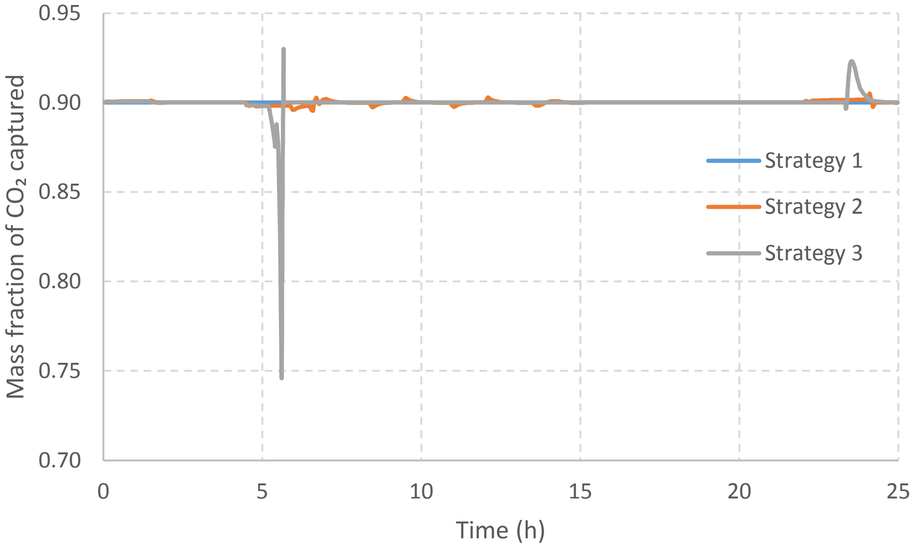

Figure 6: CO_2 capture using different control strategies for the PCPP process.

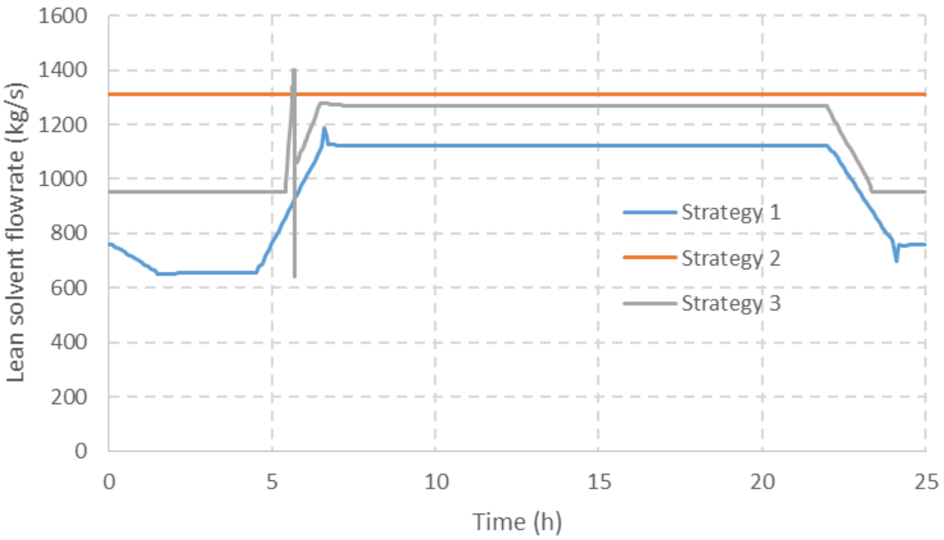

Figure 7: Lean solvent flowrate using different control strategies for the PCPP process.

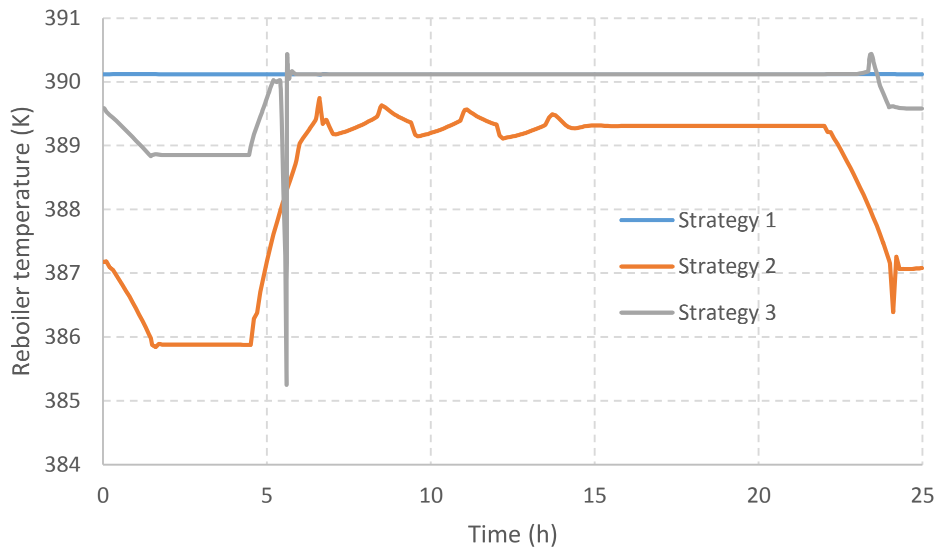

Figure 8: Reboiler temperature using different control strategies for the PCPP process.

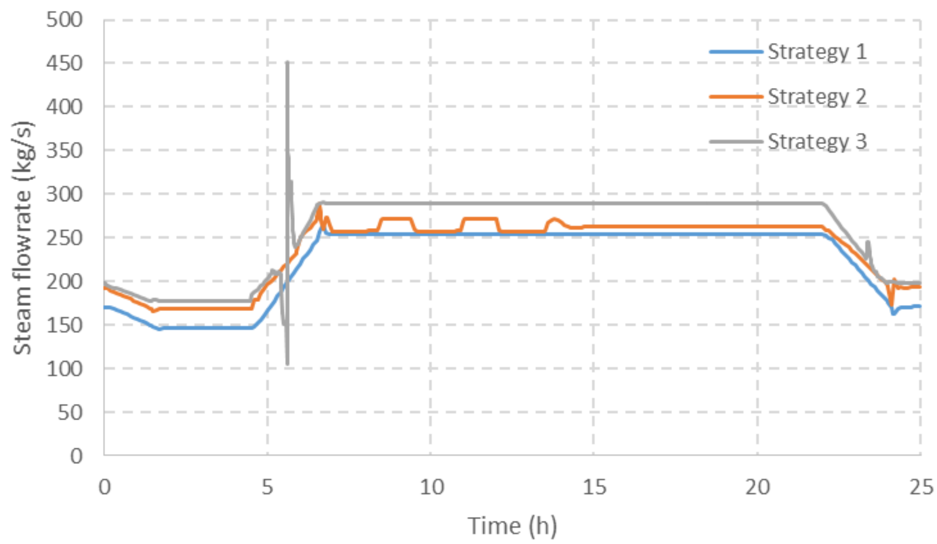

Figure 9: Steam flowrate using different control strategies for the PCPP process.

Similarly, the behaviours of some key variables for the CCGT are presented in Figure 10 - Figure 13 below.

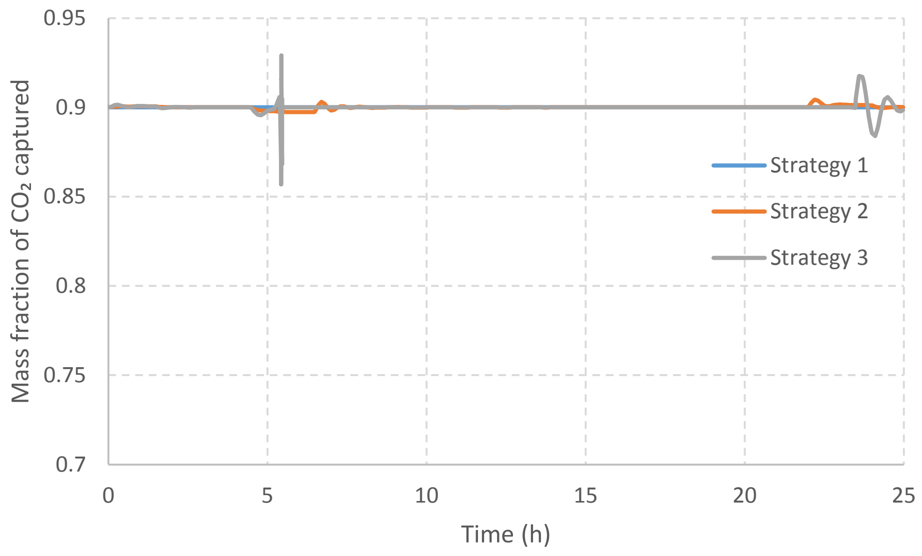

Figure 10: CO_2 capture using different control strategies for the CCGT process.

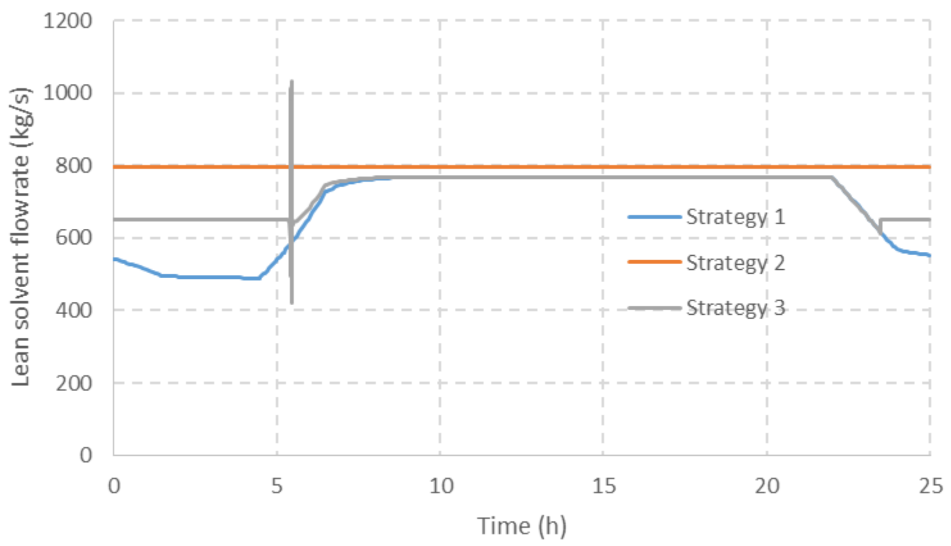

Figure 11: Lean solvent flowrate using different control strategies for the CCGT process.

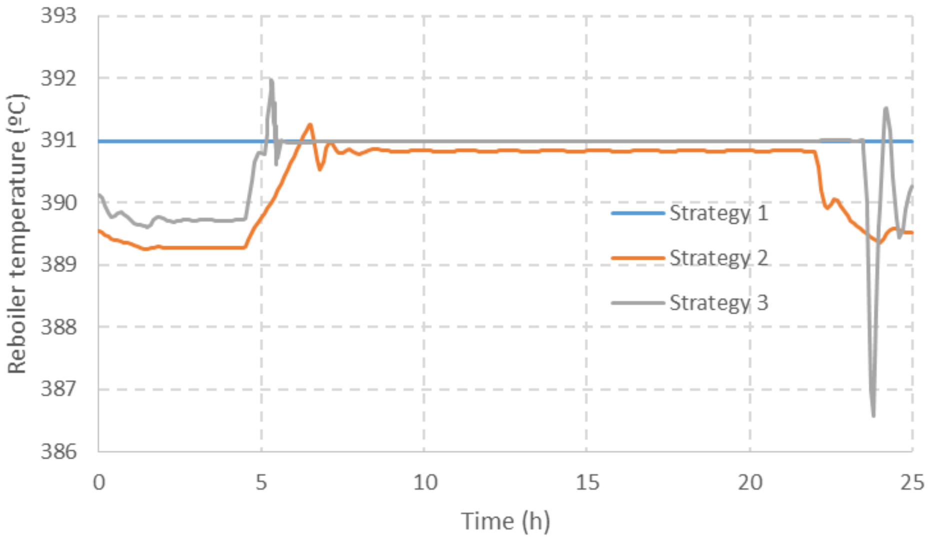

Figure 12: Reboiler temperature using different control strategies for the CCGT process.

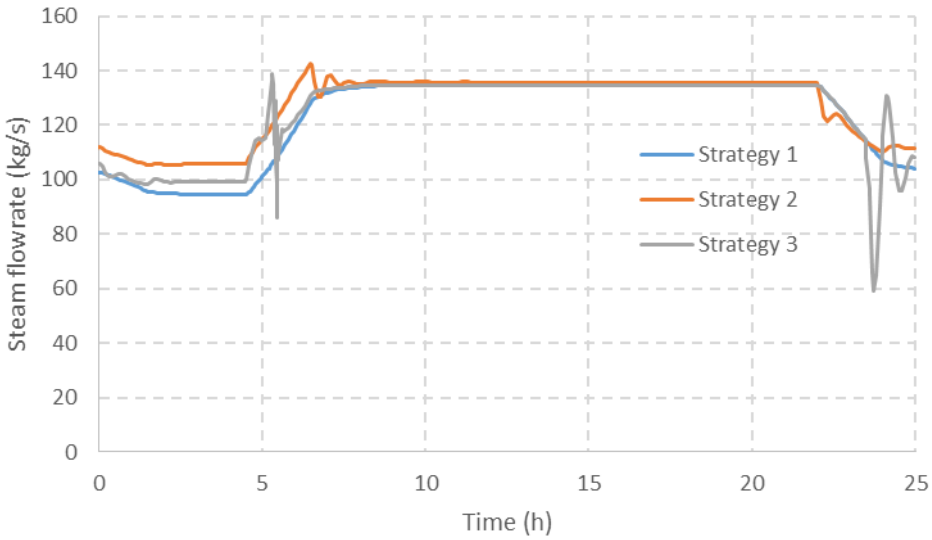

Figure 13: Steam flowrate using different control strategies for the CCGT process.

In all cases, it is notable that, following the disturbance, i.e., power plant ramping up or down, the CCS plant returns to steady state relatively quickly. However, the duration of the relaxation time and the amplitude of the oscillations are observed to be a strong function of the control strategy chosen; Strategy 1 appears to provide the most stable operation of the power plant, Strategy 2 follows, but with an appreciably longer relaxation time and finally Strategy 3 being characterised by a relatively short relaxation time, but commensurately greater amplitude of oscillation from the set point.

Economic evaluation of control strategies

It was then necessary to assess the economic impact of the three Control Strategies, following our previous work [7, 8].

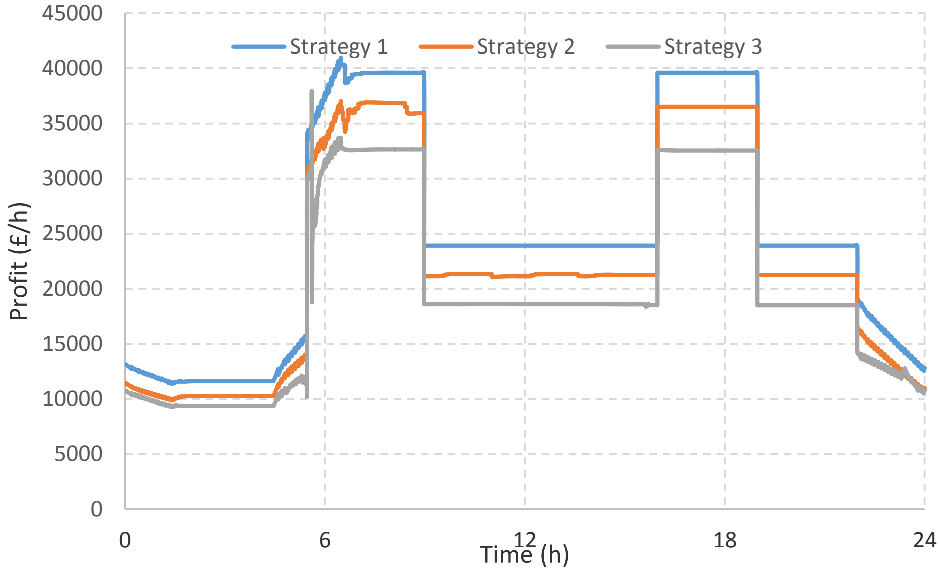

Figure 14: Profit for PCPP plant with different control strategies.

It was observed, see Figure 14, that Control Strategy 1 was more profitable than either of strategy 2 or 3. Thisis quantified in Table 2 below.

Table 2: Total daily profit for each control strategy in the PCPP process.

	Total profit
Strategy 1	£592k
Strategy 2	£533k
Strategy 3	£475k

The same analysis was performed for the CCGT, and is presented in Figure 15 and Table 3.

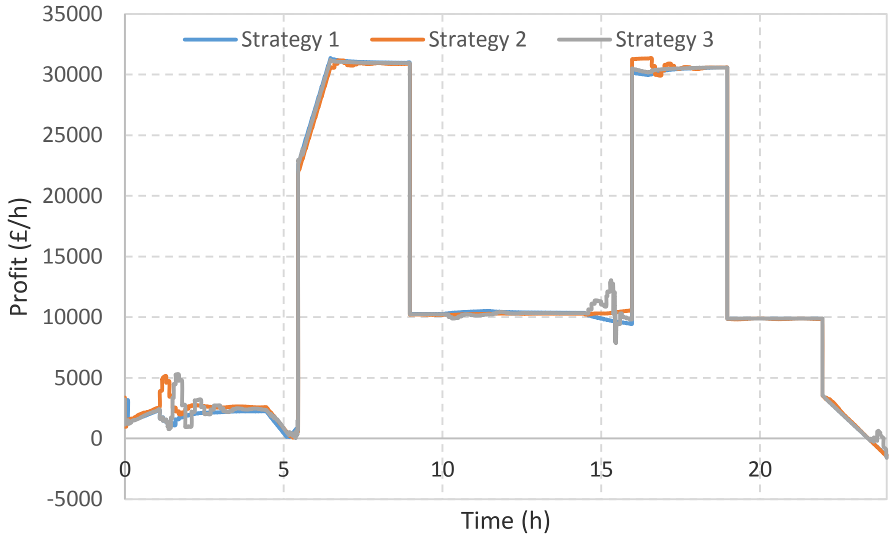

Figure 15: Profit for CCGT plant with different control strategies.

In this instance, all of the control strategies have proven to provide the same profit. This is a function of the dilute nature of the exhaust gas stream and the commensurately greater solvent circulation rate relative to the PCCP case.

Table 3: Total daily profit for each control strategy in the CCGT process.

Total profit	
Strategy 1	£290k
Strategy 2	£287k
Strategy 3	£289k

Hazard management in the CCS chain

A typical CCS project will involve multiple operator companies, e.g. the power station, CO_2 transport pipeline and offshore CO_2 injection operators. Complexity is likely to increase as the incorporation of subsequent CCS projects on the backbone of an initial project will lead to the creation of CO_2 transport networks with multiple CO_2 sources and storage options.

The safe operation of the CCS system will require close co-ordination of the different stakeholders. This section aims to identify and analyse potential hazards by considering potential trip, outage and shutdown scenarios and describing the effects on stakeholders up and down the CCS chain. It will seek to demonstrate how the process can be safely managed in the event of a system failure and suggest a communication procedure based on the best practices currently available across industry. To assess the impact of these hazards, we simulated the relevant sections of the CCS chain based on a coal-fired power plant with post-combustion CCS. This is illustrated in Figure 16 below.

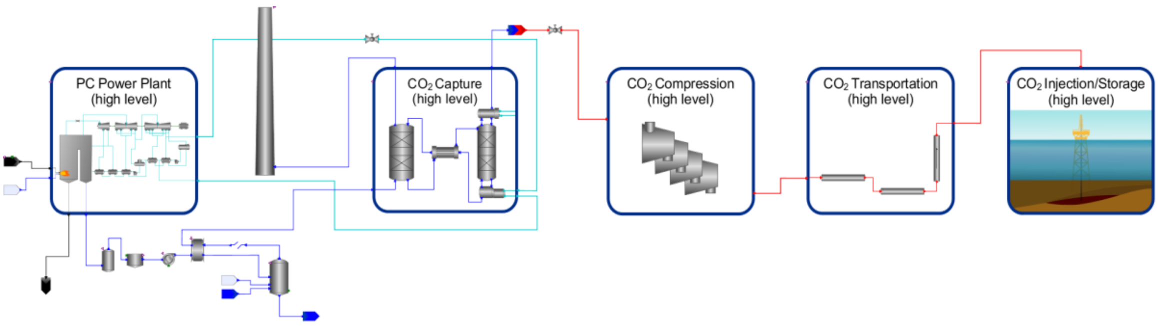

Figure 16: Full chain CCS model, including power plant, capture plant, compression train, pipeline, injection and storage

The following scenarios were considered in this work using the specific details illustrated in Figure 17:

- Unplanned shutdown at injection site / loss of storage in a single chain
- Loss of upstream compression

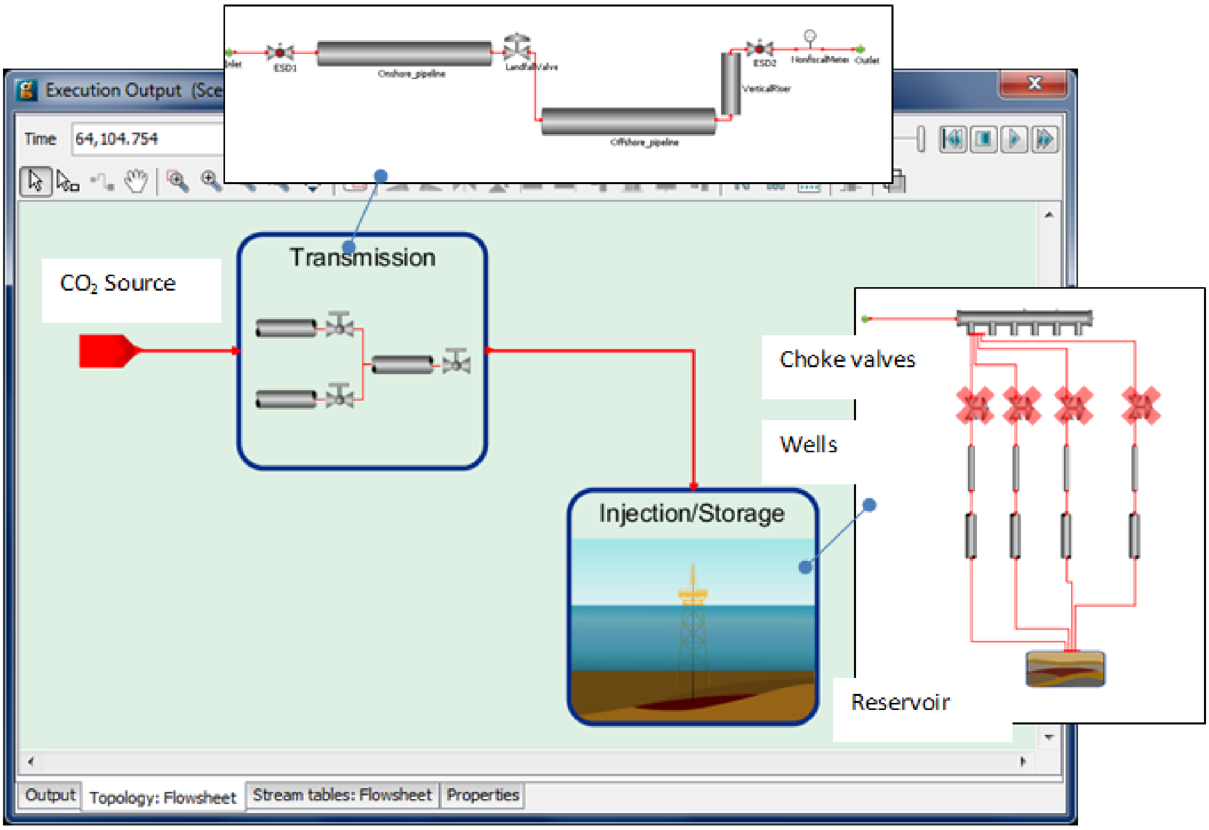

Figure 17: CO_2 transmission, injection and storage as modelled in gCCS – loss of storage scenario

The system was simulated using steady state initial conditions. Then, by evaluating the drive speed, we can evaluate how long it takes this disturbance to have a limiting impact on the CO_2 compressor at the CCS plant gate, as illustrated in Figure 18.

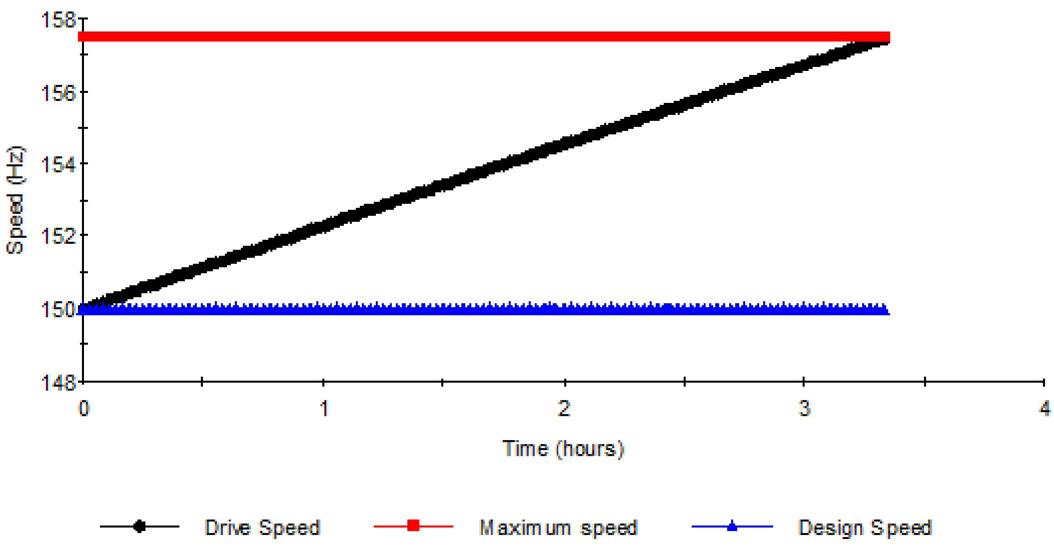

Figure 18: Determining the maximum compressor discharge pressure based on drive speed

On this basis, we observe that the CCS plant can, in this scenario, continue to operate as normal for slightly less than 3.5 hours.

The maximum allowable pressure is 153 bar and it was found that this pressure would be attained (if nothing else changes) after approximately 5.5 hours from the onset of the disturbance. From this perspective, we observe that compressor sizing is likely to be a limiting factor in how the CCS chain can operate in the event of a loss of injection. This is illustrated in Figure 19.

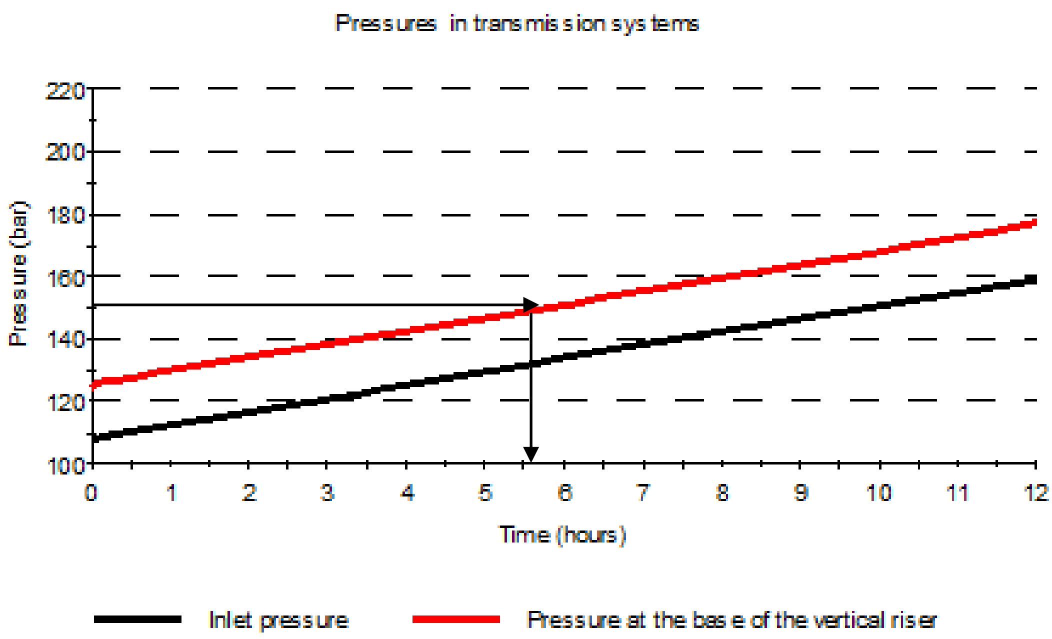

Figure 19: Increase in inlet pressure as a function of time following a storage failure

We then move on to consider the consequence of a loss of upstream compression using the gCCS elements illustrated in Figure 20 below.

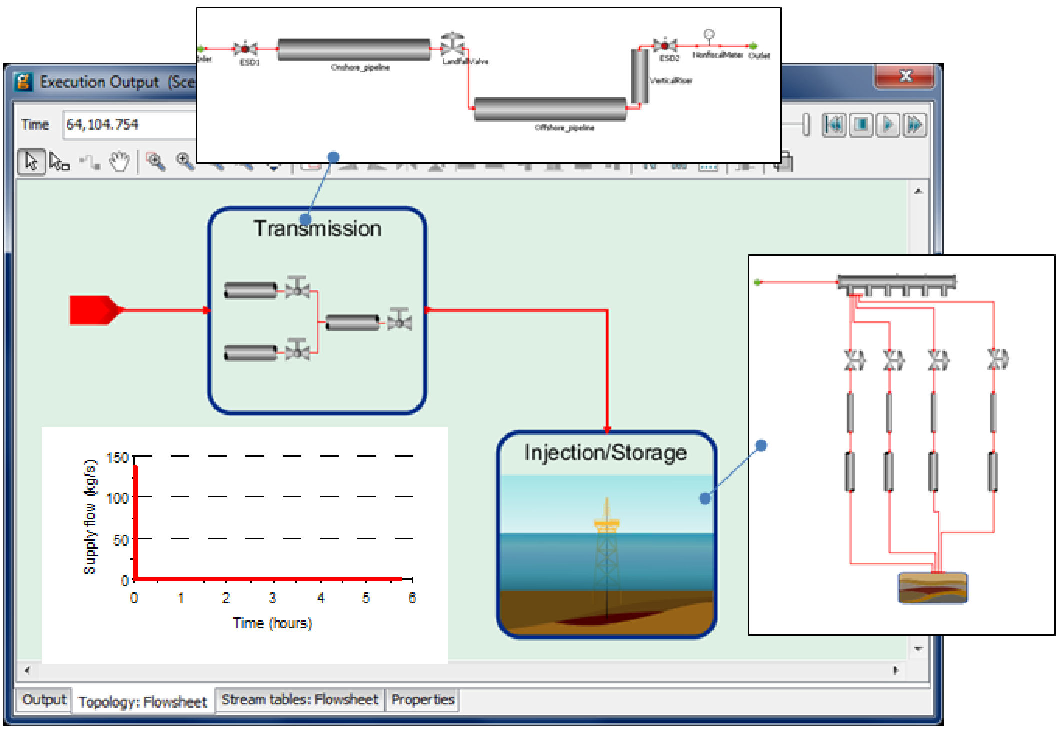

Figure 20: CO_2 transmission, injection and storage as modelled in gCCS- loss of compression scenario

In order to simulate the loss of a compressor, the flow of inlet CO_2 supply is dropped to 0kg/s in a step change. This is intended to represent the most drastic situation, and is illustrated in Figure 21.

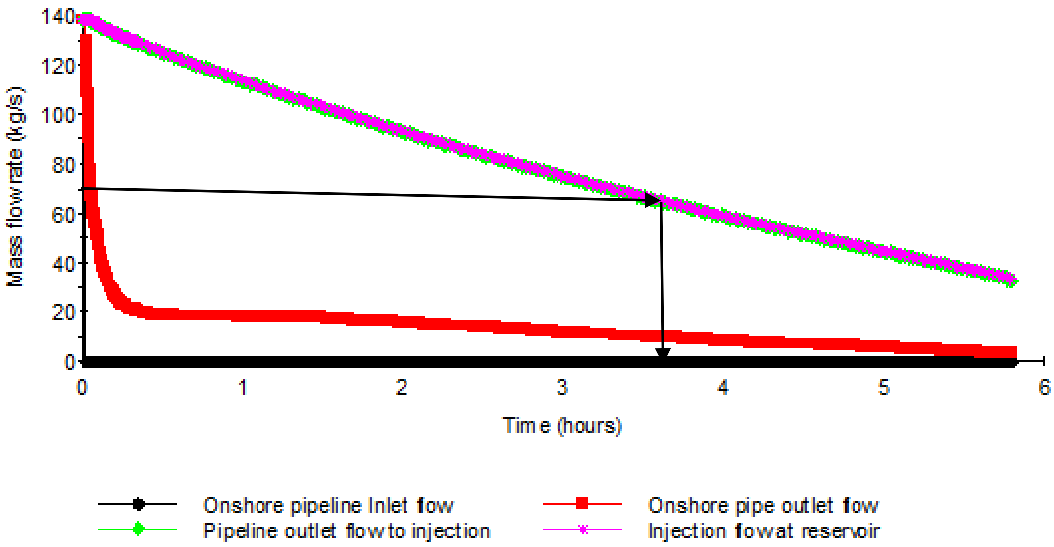

Figure 21: Mass flowrates at different locations in the CO_2 transmission and injection sub-systems

As can be observed from Figure 21, it takes 3.6 hours for the CO_2 supply at the injection site to reduce to 50% of the design flow, following a complete loss of supply

The critical point of CO_2 is 73.8 bar. During normal operation, one would expect the pipeline to operate at supercritical pressures. Thus, in the event of a loss of inlet compression, there is the potential for two phase flow to occur if the pressure of the CO_2 stream were to drop below the critical pressure, as illustrated in Figure 22 below.

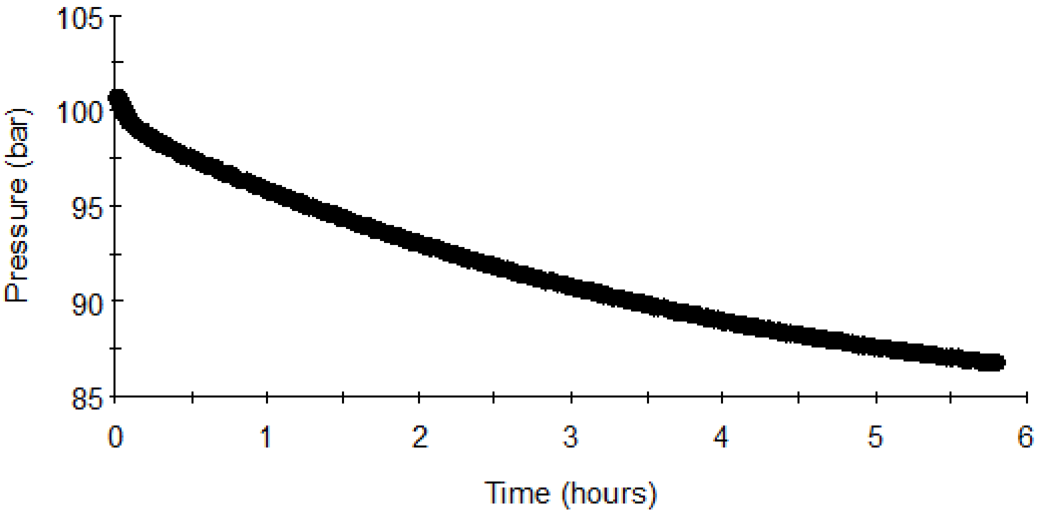

Figure 22: Minimum pressure in the transmission system

Based on this simulation, it is not expected that this will occur within 5 hours of loss of supply. However, the location of the critical point is, of course, a function of the composition of the stream, with the critical point of an impure stream typically being at a higher pressure than a pure stream.

If there is an instance where pressure downstream exceeds that of the upstream pipeline, a reverse flow situation could occur. This evaluation is presented in Figure 23.

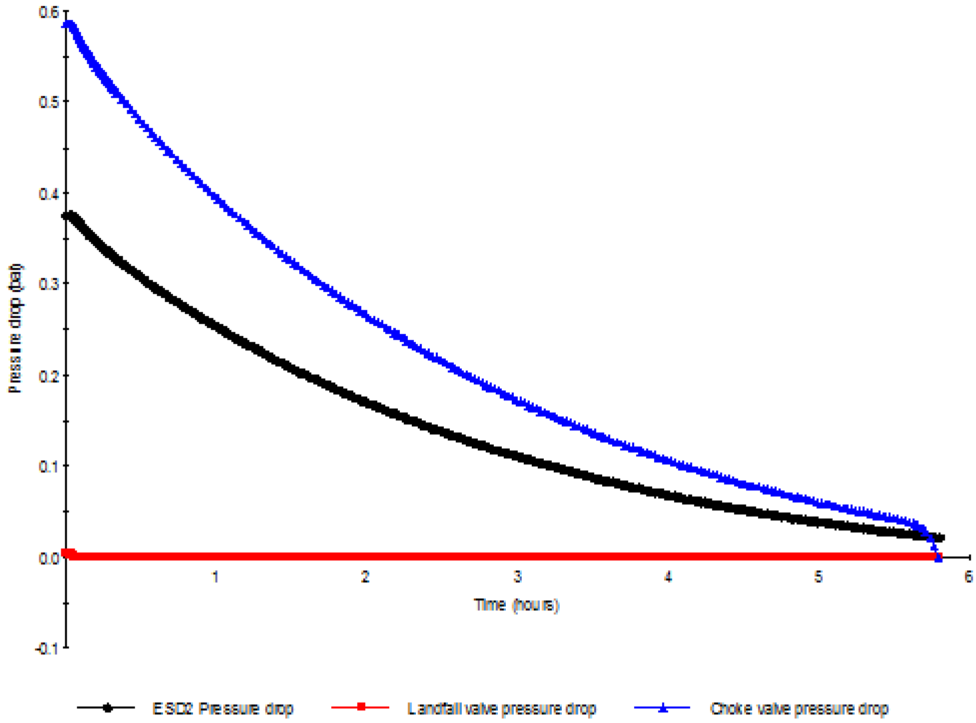

Figure 23: Pressure drop across various valves in the system

It was found that, in the case of a loss of compression, with no intermediate compression along the pipeline, a reverse flow scenario could occur. Reverse flow would occur first at the choke valves. The pressure drop across the choke valve (blue line) drops to a slightly negative value just before the 6-hour mark.

Main conclusions and recommendations for future work

- Electricity market models support the notion that power plants will need to adopt flexible operation patterns in the future. Furthermore, they show that fluctuations between full and part-load operation will become more frequent during the day, and therefore an appropriate control strategy is necessary to ensure the operability and profitability of the process.
- Several control strategies have been proposed in the academic literature and are typically evaluated using reduced order, control-orientated models. However there is a need to implement these conceptual strategies using high-fidelity modelling tools.
- Tools such as gCCS are very useful to accurately predict the behaviour of power plants with carbon capture under flexible operating conditions. Design of suitable control strategies should be carried out using such tools.

- With an appropriate and well-tuned control strategy, it is possible to maintain critical parameters such as CO_2 capture at the desired set-point, even during periods of significant fluctuation in the power plant load.
- Using an appropriate control strategy, even if based in simple and well established control technologies, such as PID, avoids the need for more expensive solutions such as adding solvent storage tanks to the process.
- From the operational point of view, using the solvent flowrate as manipulated variable to control the CO_2 capture rate is a better option than manipulating the reboiler temperature and therefore the lean loading.
- From a hazard management perspective, it was observed that the capacity of the compressor to add to pipeline pressure in the event of an outage at the injection site is a limiting element in the chain

Despite the tuning issues and more oscillatory behaviour, control of solvent lean loading was more profitable for PCPP, whereas for CCGT all strategies have proven to be the same. This is a function of the dilute nature of the exhaust gas stream and the commensurately greater solvent circulation rate relative to the PCCP case.

The output from this report and a critical review of the gaps in the literature related to control of power plants with CCS resulted in the following recommendations for further research and technical studies.

- *Modelling and simulation aspects* – further developments in high-fidelity modelling and simulation tools to include accurate descriptions of all components in the CCS chain. Additionally, testing of the control strategies proposed in this work for different solvents in the post-combustion capture system.
- *Simultaneous process and control design* – development of appropriate algorithms for including process control design decisions at the process design stage.
- *Advanced process control techniques* – development of hierarchical model predictive control formulations for the integrated system, including power plant and capture plant, in order to optimise the performance and economics of the overall process.
- Optimisation of control strategy switching – follow-up to this work, where the schedule of dynamic switching between control strategies is determined by an optimisation problem where the costs and benefits of each strategy are encompassed in the formulation.

1. Introduction

Carbon capture and storage (CCS) is a near-term solution for significantly reducing the CO_2 emissions from coal and natural gas fired power plants. From the different CCS technologies currently under study and development, amine-based post-combustion capture is the most mature option available, having recently been the first carbon capture technology demonstrated at a commercial scale power plant.

In order to meet CO_2 reduction targets, decarbonised power plants operating with CCS will likely play a major role in the energy production mix. The important role of intermittent renewable energies in this energy production mix, combined with the variability of consumer demand for electricity, implies that power plants with CCS need to be able to operate under flexible load conditions.

In this report we employ the gCCS platform, developed by Process Systems Enterprise[3] to study the dynamic behaviour of coal and natural gas fired power plants with amine-based post combustion carbon capture processes. This toolkit includes high-fidelity models that describe the dynamic operation of all the stages of the CCS chain: power generation, post-combustion carbon capture, compression, transportation, and storage. In particular, the aim of this study is to develop and evaluate process control strategies for the operation of a post-combustion capture process for coal and natural gas fired power plants. The design of an appropriate control structure is important to minimise the well-known energy penalty that a carbon capture process imposes on the power plant efficiency.

The remainder of this report is presented as follows.

Section 1.1 consists of a study of part-load operation scenarios that are relevant to the period within the scope of this work – in the 2030s and beyond. This analysis is informed by suitable electricity market models and ultimately aims at defining the part-load scenarios under which the proposed control strategies are evaluated.

The relevant literature on flexible operation of decarbonised power plants, and corresponding process control aspects, is reviewed in Section 2. This review highlights the most common control structures, used in post-combustion carbon capture processes, which we use as the basis for the development of control strategies in this work.

In Section 3, the reference case models are presented, along with the corresponding implementation in gCCS. The section includes information on some of the equations used to model the processes in the CCS chain, the design parameters of the different components, and an overview of the implementation of the integrated system in gCCS.

The control strategies proposed in this work are presented in Section 4, and Section 5 shows the methodology used to tune the control loops in each of the proposed control strategies. Full simulation results are shown for the tuning procedure, in order to establish the base case under which the control strategies are evaluated in Section 6.

The conclusions from this study and recommendations for further research are presented in Section 7 and Section 9, respectively.

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[3] http://www.psenterprise.com/power/ccs/gccs.html

1.1 Part-load scenarios that will be relevant to a CCS plant in the 2030s and beyond

The electricity market models (MOSSI [1-4] and UCCO [3, 5, 6]) were used to produce a set of scenarios for the future operation of CCS plants. These models were designed and are employed at Imperial Business School:

- MOSSI (Merit Order Stack with Step Investments) to explore the long-term investment decisions which shape the mix of installed capacity.
- UCCO (Unit Commitment Capacity Optimiser) to simulate the short-term operating decisions for individual plants in greater detail, taking account of their technical limitations.

Both models simulate investor and operator behaviour within a liberal electricity market (such as the British wholesale market), assuming that firms attempt to maximise profits and that competition between firms maintains electricity prices close to costs.

Two technologies are considered: coal with CCS and gas with CCS; which represent the first generation of each technology to be built in the chosen electricity market scenario. The evolving role of each technology over its lifetime is considered by taking three snapshots of operations, during the 2030s, 2040s and 2050s.

For each technology we evaluate the operating patterns of three units located at different positions within the merit order stack: with low, average, and high variable costs relative to other plants within that vintage. Not all plants built within the decade will have the same variable costs, and so this allows the spread of operating patterns to be evaluated due to the range in factors such as the unit's efficiency and the operator's fuel and maintenance costs.

The following sections detail the workflow of producing these scenarios:

- Projecting the future mix of installed capacity using MOSSI;
- Simulating plant operation in the three study periods (2030s to 2050s);
- Extracting and distilling operating patterns for individual CCS plants.

1.2 Projecting the future mix of installed capacity using MOSSI

The evolution of the installed capacity in Britain is shown in the figure below, with snapshots of installed capacity during each decade out to 2100 shown against the backdrop of historic capacity evolution. As can be observed from Figure 24, in the early decades, existing nuclear and unabated coal plants reach the end of their lives and are not replaced, due respectively to capital costs and the carbon price making them economically unviable. A reasonable amount of unabated CCGT and OCGT capacity remain on the system, both of which are used for peaking capacity.

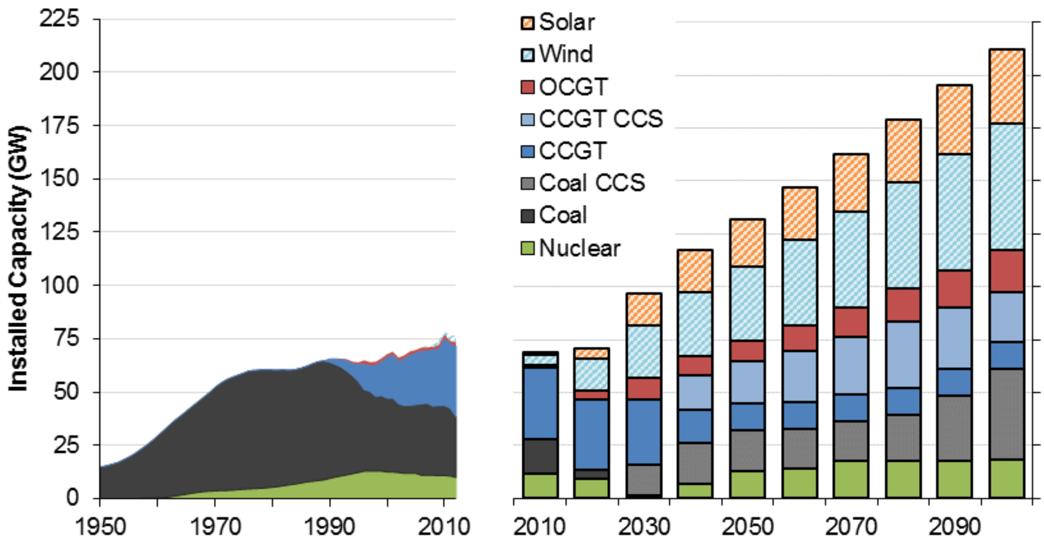


Figure 24: Evolution of installed electricity generation capacity in the UK for the period 2010 to 2100

Coal with CCS becomes viable in the 2030s once the capital cost (CAPEX) has fallen below £2,500/kW and the carbon price has risen to £76/T. 14 GW are installed in the 2030s, followed by a further 5 GW in the 2040s. After this, capacity remains stable until the 2070s (which is beyond the period we are interested in). Combined-cycle gas with CCS becomes viable in the 2040s once its CAPEX has fallen below £1,650/kW, and the carbon price has risen further to £138/T. 16 GW are installed in the 2040s, followed by 5 GW in the 2050s.

1.3 Simulating plant operation in the three study periods (2030s to 2050s)

The hour-by-hour dispatch of the entire power system was optimised using UCCO for each of the three snapshots. Figure 25, presents a sample of the power system operation in the 2030s. Three energy sources dominate: coal with CCS (grey, bottom), unabated gas (dark blue, middle), and wind (hatched, top).

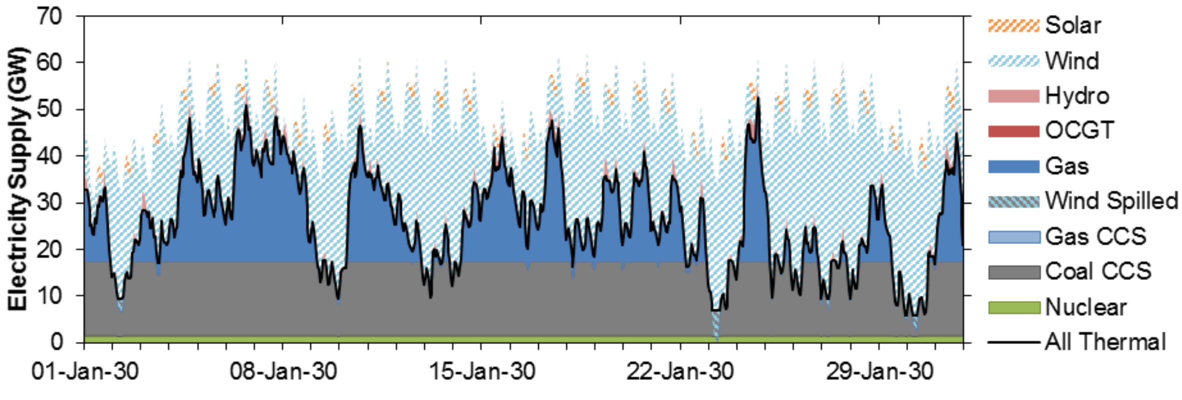


Figure 25: Power station dispatch over the course of January in the 2030s scenario, from UCCO.

Coal-CCS benefits from the lowest variable costs apart from nuclear, and so operates as base-load capacity. The majority of units are able to run at full power for stretches of time, as seen in the first half of January 2030. However, when high wind output coincides with low demand, such as in the fourth week, they are forced to follow the fluctuating net demand (black line) by ramping and shutting down.

During short-lived spikes in wind output, it can be cheaper to keep some CCS units running, rather than shut them down only to restart them soon afterwards. In these cases, the thermal power stations on the system produce more electricity than is required, and so some wind farms must be shut off to maintain system balance. These periods are highlighted in the figure as wind being spilled (blue and grey hatched areas). Based on the current market design in Britain (and across Europe), the electricity price becomes negative during these times (–£50/MWh), so the plants which choose to remain operating (inflexible nuclear and CCS) pay for the surplus electricity they produce, and so run at a loss (albeit a smaller one than would be incurred by stopping and restarting soon afterwards).

Moving forwards to the 2040s (Figure 26), gas with CCS becomes part of the installed mix and displaces much of the output from unabated gas plant. Expansion in nuclear and wind capacity forces thermal plants to adopt more dynamic running patterns, with reduced running hours, more start/stop cycles and greater ramping rates.

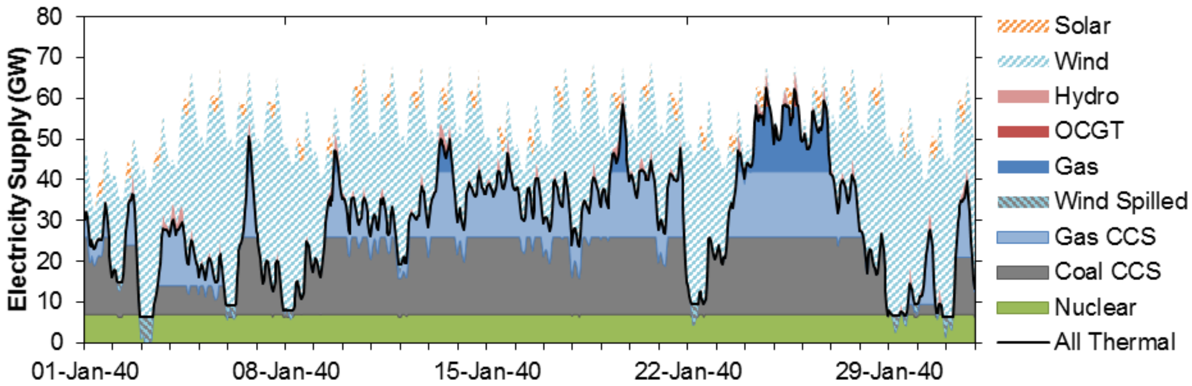


Figure 26: Power station dispatch over the course of January in the 2040s scenario.

These trends continue to intensify in the 2050s, as it illustrated in Figure 27 .

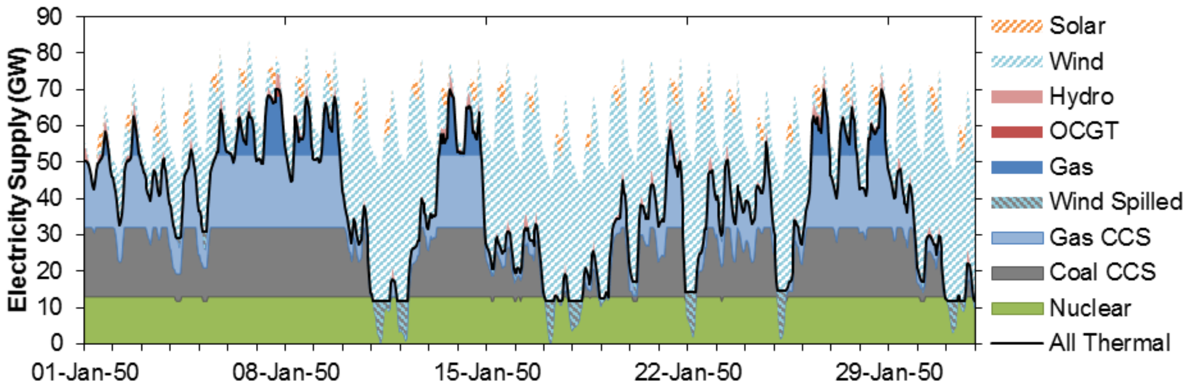


Figure 27: Power station dispatch over the course of January in the 2050s scenario.

1.4 Extracting and distilling operating patterns for individual CCS plants

The UCCO model reports on the dispatch pattern of individual generating units. Three coal CCS units were studied, with low (Unit 1), mid (Unit 3), and high (Unit 5) positions within the merit order. The hourly operating patterns of these units are shown in Figure 28 for the 2030s (top), 2040s (middle), and 2050s (bottom). The increase in requirements for dynamic operation is evident as time progresses, with more frequent modulation between maximum and minimum stable generation and more time spent off.

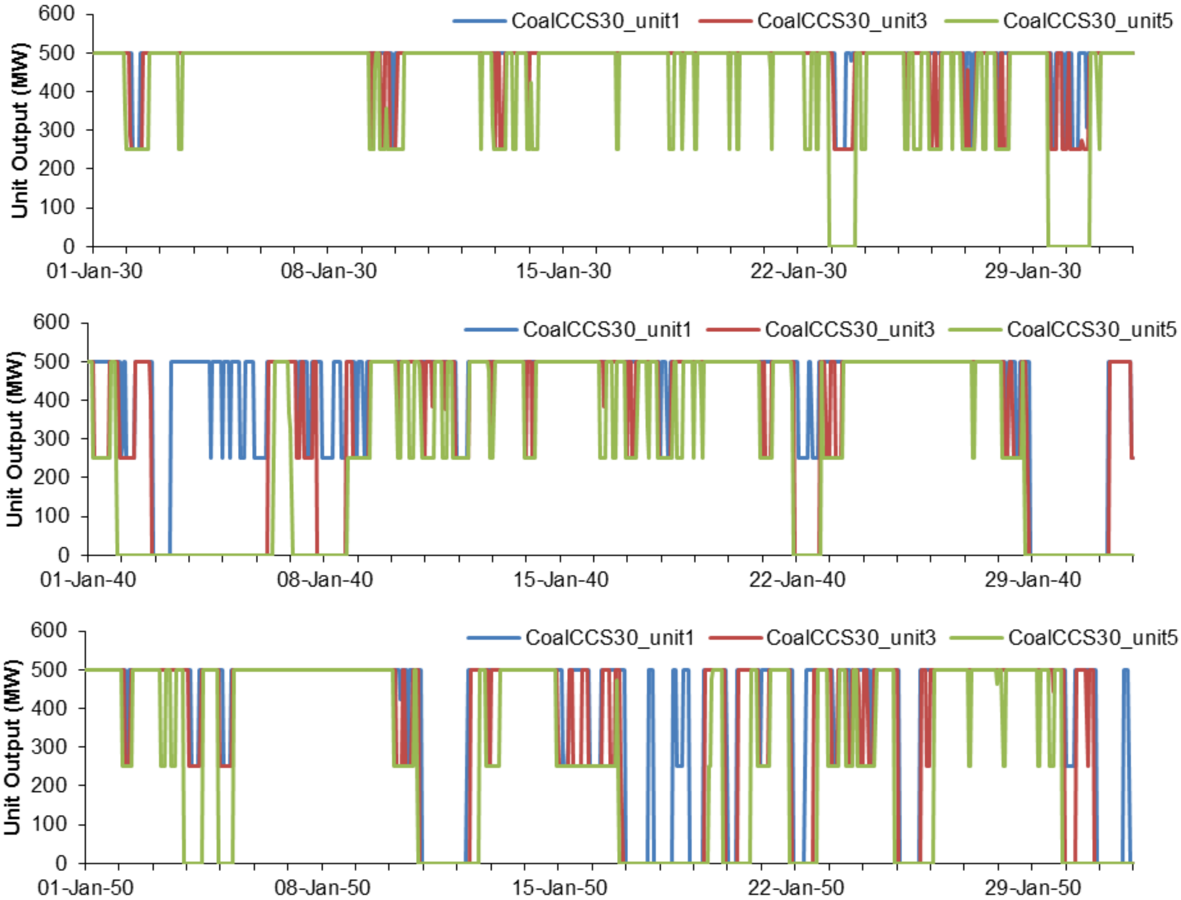


Figure 28: Operating profiles for three coal CCS units in the 2030s (top), 2040s (middle) and 2050s (bottom).

The same pattern is evident with gas CCS plants, although they start with a more dynamic operating pattern in the 2040s (Figure 29) due to being priced as mid-merit as opposed to base-load capacity.

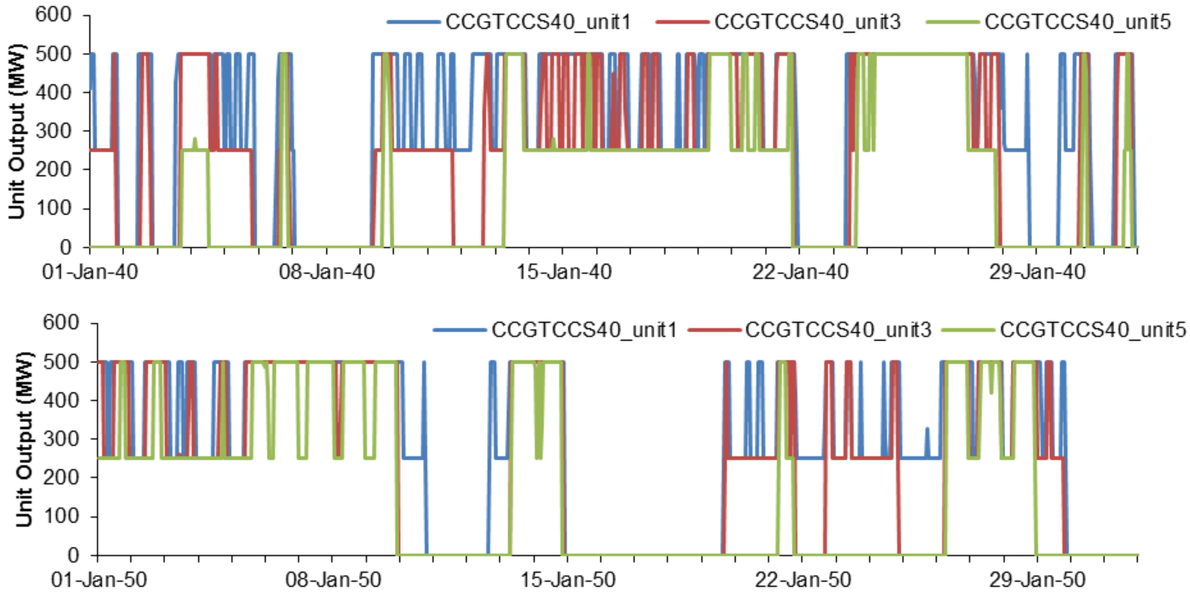


Figure 29: Operating profiles for three gas CCS units in the 2040s (top) and 2050s (bottom).

On the basis of the above calculation, the operation of an individual unit can be divided into four states:

- Operating at maximum power
- Floating between maximum and minimum power (i.e. being the marginal unit on the system)
- Operating at minimum stable generation
- Not generating

These categorisations allow the evolution of CCS unit behaviour over the decades to be seen more clearly. The operating states for coal and gas CCS units are shown in Figure 30 and Figure 31.

Within each panel, the range of behaviour for CCS units is shown by the three bars for units at different positions within the merit order. For coal in the 2030s, all units are operating as typical base-load units, running at full load for 80–96% of the year (ignoring periods of downtime due to maintenance), and the range in behaviour between units is relatively narrow. Annual average capacity factors range from 87% to 96% (again, excluding downtime).

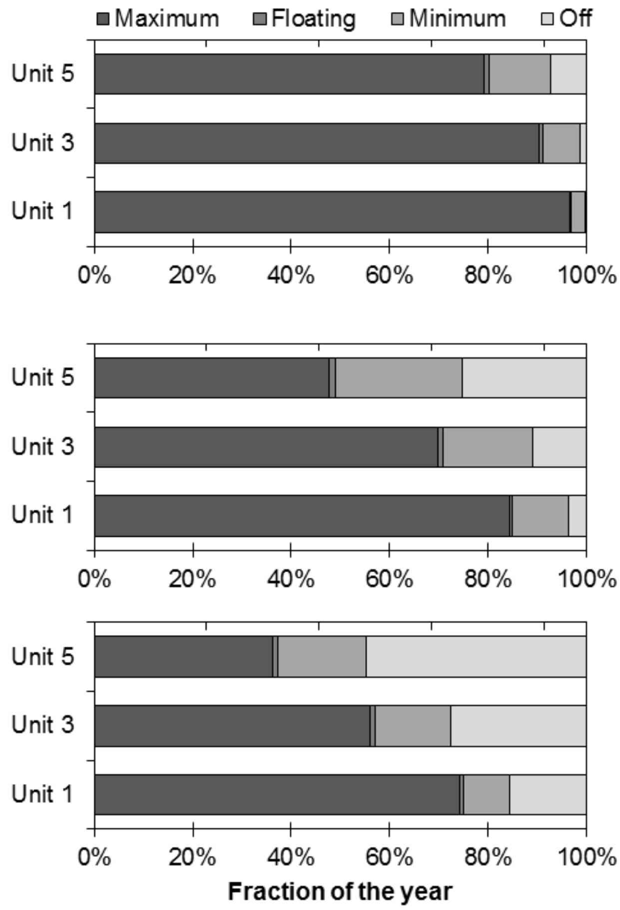


Figure 30: Summary of operating states for three coal CCS units during the 2030s (top), 2040s (middle) and 2050s (bottom).

In the 2040s, the operation of individual units is more diverse, and the range of behaviour between units is far wider. The unit with the lowest variable costs (i.e. the most efficient) is reduced to running at full load for 84% of the year, while the highest cost unit is only able to run for half of the year at full load, and spends a quarter of the year sat idle. Annual average capacity factors fall and broaden out to between 60% and 90%. These trends continue into the 2050s, with capacity factors reaching the range of 45% to 80%.

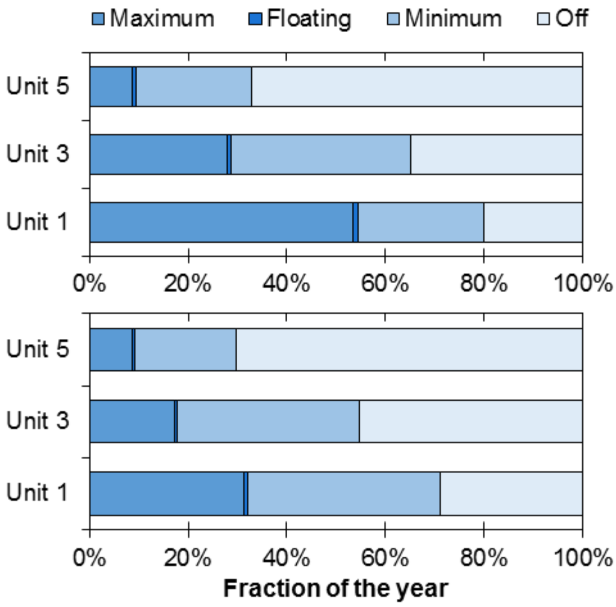


Figure 31: Summary of operating states for three gas CCS units during the 2040s (top) and 2050s (bottom).

Capacity factors for gas CCS plants start much lower, and the units experience a broader range of operating states because they are priced as mid-merit plant. The range of behaviour between different gas CCS units narrows, rather than broadens, as they converge on load capacity factor operation. In the 2040s, the range is from 21% to 67%, and by the 2050s this contracts to 19% to 51%.

2. Review of literature on operation and control of power plants with post-combustion CO_2 capture

This section presents a review of the available literature on operation and control of power plants with post-combustion CO_2 capture. The first deals with the recent literature pertaining to flexible operation of decarbonised power plant and the second addresses the development of a control strategy for these plants, with explicit consideration of how the control of a flexible plant may need to differ from that of a base-load plant.

2.1. Flexible operation of decarbonised power plants

The concept of "flexible CCS" is not new, with some of the key concepts being considered in academic circles at least a decade ago [9, 10]. In many of these papers, the assumption has been that the capture process would simply be partially or wholly bypassed and a portion of the exhaust gas vented to atmosphere [10] or that the solvent regeneration process would be wholly or partially bypassed and the CO_2 rich solvent stored at times of high electricity demand and then subsequently regenerated at a later time when electricity demand was reduced, i.e., there was a surplus of steam generation capacity in the power plant [9].

However, the viability of the capture by-pass option is an important function of the prevailing CO_2 price and the regulatory framework under which the plant is operating. It has been recently shown that at CO_2 prices sufficient to incentivise the large scale deployment of CO_2 capture technology, the cost associated with capture by-pass (i.e., exhaust gas venting) was such that very little was gained by the power plant operators in this mode of operation [11]. Similarly, the concept of storing large quantities of CO_2-rich solvent has always been viewed with some scepticism as it would then mean that the power plant was more akin to a large chemical plant and may need to comply with the COMAH regulations[4].

It is also worth noting that much of the work performed in the context of flexible CCS has generally considered standard CCS retrofit configurations. Therefore, to date, it would appear that the concept of flexibility has meant choosing *when* to emit CO_2. Recently an approach has been proposed in the literature wherein post-combustion CCS was optimised for flexible operation via the accumulation of CO_2 in the working solvent during periods of high electricity demand and the subsequent thorough regeneration of that solvent during off-peak periods [7, 12].

However, there are important distinctions to bear in mind if one is to consider new-build sites where the power plant is decarbonised from its conception.

A recent study [13] evaluated a range of strategies for improving the operating flexibility of a range of different power plants with CCS. They considered combined cycle gas turbine (CCGT) power plants, integrated gasification combined cycle (IGCC) plants and ultra-super-critical pulverised coal (USC-PC) plants with post-combustion capture (PCC) in addition to USC-PC with oxy-combustion and cryogenic CO_2 capture. It is worth bearing in mind that this work did not consider the sub-critical pulverised coal-fired power plants in their study. This is somewhat unfortunate as, owing to the steam drum in the thermodynamic cycle, this class of plant is recognised as being more flexible that USC-PC plants. This is important as it has been recently reported that over half of new coal-fired power plants deployed in 2012 were sub-

[4] www.hse.gov.uk/comah/

critical plants [14]. Moreover, as these plants are overwhelmingly deployed in regions of the world with the largest emissions reductions targets, it will be important to understand how they will operate in flexible way in synergy with renewable energy and in conjunction with their eventual decarbonisation. An important counterfactual to this is that in the US and EU, subcritical coal plants are being closed at a rapid rate and therefore SCPC plants are becoming more important, at least in these regions.

As CCGT and USC-PC with PCC are the technologies which are relevant to this project, only these will be commented upon further. A particularly useful output of Domenichini's paper [13] was the comparison of key operating flexibility features of power plants with and without CCS. Some of the key features are summarised in Table 4.

Table 4: Flexibility features of power plants with and without CCS [13].

	Turndown	**Cycling capability Start-up to full load**	**Ramp rates**	**Part load efficiency**
CCGT	Low load operation: 15-25% CC load (10-20% GT load) Min. environmental Load: 40-50% CC Net Power Output (NPO) (30-40% GT load)	Hot start-up: 45-55 min Warm start-up: 120 min Cold start-up: 180 min	5 - 50 MW/minute max Hot start-up load change rate: - 0-40% GT load: 3-5%/min - HRSG pressure: 1-2%/min - 40-85% GT load: 4-6%/min - 85-100% GT load: 2-3%/min	Approx. constant efficiency down to 85% GT load 2-3 percentage points less @ 60% CC load
CCGT + CCS	Post-combustion unit min. load: 30% CO_2 compressor min. efficient load: 70%	Regenerator preheating: - hot start-up: 1-2 h - warm start-up: 3-4 h	Same as plant w/o CCS	Same as plant w/o CCS
USC-PC	Min. boiler load: 25- 30%	Very hot start-up: < 1h Hot start-up: 1.5-2.5 h Warm start-up: 3-5 h Cold start-up: 6-7 h	30-50% load: 2-3%/min 50-90% load: 4-8%/min 90-100% load: 3-5%/min	Subcritical boiler: -4 perc. point @ 75% load Supercritical boiler:- 2 perc. point @ 75% load
USC-PC + CCS	Post-combustion unit min. load: 30% CO_2 compressor min. efficient load: 70%	Regenerator preheating: - hot start-up: 1-2 h - warm start-up: 3-4 h	Same as plant w/o CCS	Same as plant w/o CCS

It was observed that one of the largest constraints on the part-load operation of decarbonised power plants is the operability of the CO_2 compression train. This leads one to consider using multiple or parallel compressors.

Domenichini et al. [13] also considered the use of solvent storage with a view to increasing the flexibility of the decarbonised power plant. However, this was found to have important design implications, including the potential need to oversize the solvent regeneration and compression to cope with regeneration of the solvent stored during peak hours whereas if the plant was expected to operate regularly at reduced load, then under-sizing of the regenerator and compressor would be preferable. There would also be design implications associated with the rich-lean heat exchanger and other non-fuel operational costs associated with the trim coolers, condensers and potentially the water-wash process for the cleaned exhaust gas, again giving rise to obvious implications to decarbonised power plants operating in water stressed areas.

2.2 Control of flexible decarbonised power plants

From the discussion in the previous section, it is clear that the nature of power plants is such that large disturbances over short and long time scales are part of the normal daily operation and expected by design. For this reason, the development of control strategies for power plants with carbon capture should be carried out with flexible operation in mind, rather than simply with the objective of having a regulatory control system capable of handling occasional disturbances.

The major concerns related to the flexible operation of a post-combustion capture process are related to changes in hydraulic conditions in the absorber and stripper columns that occur if the plant is bypassed or if significant changes in the flue gas and solvent flow occur [15]. When the gas flow rate is significantly higher than the design value, flooding will occur in the column; a low liquid-gas ratio will also result in poor wetting and reduce the efficiency of the absorption column.

The first works on control of PCC focused on standalone systems where control strategies were developed for single units, such as the absorber column. In this context, Ziaii et al. [16] developed a strategy that involved controlling the capture rate by using the ratio between the rich amine flow rate and the reboiler heat duty. In this strategy, the lean loading, i.e. the molar ratio of CO_2 to MEA in the lean solvent, is kept constant when the stripper load changes.

From 2010, several works started focusing on the dynamic simulation and control design of the entire PCC system, with some works also including the simulation of the power plant. Lin et al. [15] proposed a plant-wide control strategy for PCC in which the lean solvent feed rate to the absorber column was used to maintain the CO_2 removal level at 90%. This work identified the importance of maintaining the water inventory in the system, by controlling the liquid level in the reboiler of the stripping column using a make-up stream. The temperature at the bottom of the stripper was controlled by varying the rate at which steam is supplied to the reboiler.

The work of Lin et al. [15] was updated in 2012 [17] to include the dynamic simulation of a wider and more representative portion of the power plant, including the low pressure turbine of the steam cycle, the post-combustion capture system, and a compression train. This work considers control design with the explicit objective of addressing the above-mentioned concerns related to the hydraulic conditions in the absorber and stripper columns. In particular, the CO_2 capture rate targets are met by changing the lean solvent loading, while keeping the lean solvent flow rate to the absorber at a fixed value. The stability of the stripper column is maintained by recycling part of the CO_2 product to the bottom of the stripper.

The compressor train is controlled by an anti-surge system through which the flow rate of CO_2 for compression is kept above a minimum by recycling part of the CO_2 outlet from the compressor train. It is noted that if anti-surge control is used in the compression train, the energy penalty will not be proportionately reduced when the CO_2 capture rate decreases, due to the energy needs involved in recycling the CO_2 outlet stream. A common approach to increasing the flexibility of the compression train is to use asymmetric parallel compressor trains, typically with a 30/70 split of the overall duty, introducing another key control variable.

Another control strategy, proposed by Lawal et al. [18], involves maintaining the lean solvent load at a specified set-point, by controlling the temperature in the bottom of the reboiler of the regeneration column, using the reboiler heat duty as manipulated variable. The temperature at the top of the column is also controlled, by using the condenser duty. In contrast to the approach of Lin et al. [17], the water make-up is located at the top of the absorber column. This make-up stream is used to maintain the water balance in the system, by controlling the water mass fraction in the lean solvent. This control strategy has been used in the context of an integrated dynamic model, comprising a sub-critical coal power plant and the PCC system under industrial operating conditions [19]. An interesting observation from this study is the interaction between the control loops in the capture plant and the control loops in the power plant.

The choice of control loops in the control strategies discussed above was carried out based on heuristics or by trying different combinations of controlled variables and manipulated inputs. A more systematic approach to control design, using the self-optimising control method, has been developed by Skogestad [20]. This method has been applied in the context of PCC systems by Panahi and Skogestad [21]. In this study, three different operating regions, corresponding to different flue gas flow rates, are identified and control loops are designed for each of the regions separately. A follow-up work [22] identifies an alternative set of control loops that provides close to optimal control performance in all operating regions.

A different control strategy consists of using model predictive control (MPC), in which a central controller uses a mathematical model of the process to determine the control actions, which, for a given set of manipulated variables, achieves optimal operation according to a specified objective function. This method has been used by Bedelbayev et al. [23], for a standalone absorber column, and by Arce et al. [24], in the context of a full PCC system. As shown in the latter work, the use of MPC allows the design of multi-level control structures in which a high-level controller determines the cost-optimal operating conditions, based on energy and CO_2 prices, and a low-level controller implements these decisions at the process level.

Despite the different choice of algorithms, variables, and control objectives, all strategies presented in this section show that the current research trend in PCC systems control is to achieve flexible operation without the need for solvent storage facilities, and avoiding the need to bypass the capture system.

3 Reference case models

The sections below present the reference case dynamic models developed using the gCCS toolkit. Larger versions of the flowsheets shown below are available in Appendix A at the end of this report.

3.1 Supercritical pulverised coal power plant model

Figure 32 shows the flowsheet developed for a supercritical pulverised coal power plant producing a gross electricity output of 825MW at full-load. The following sections present the specifications for the main units and model inputs.

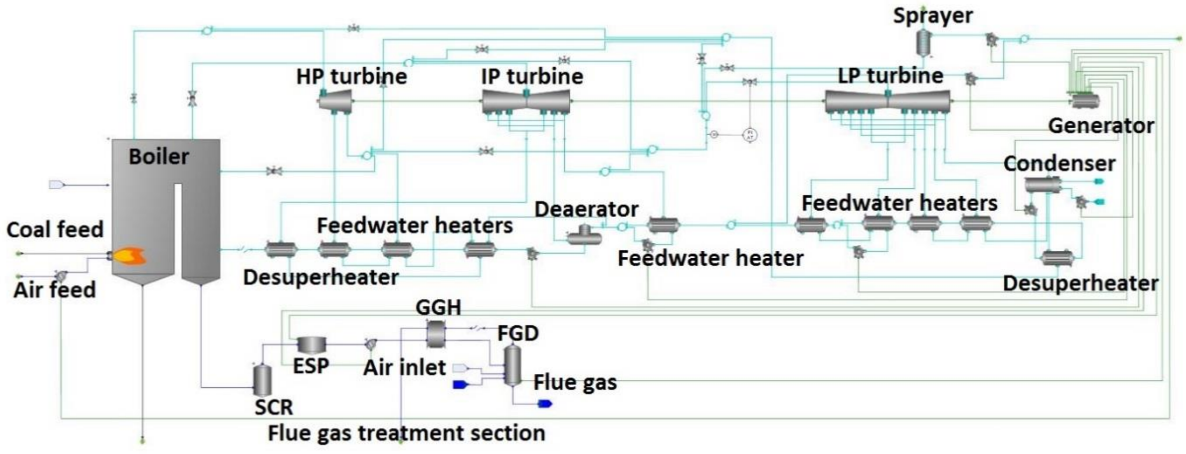


Figure 32: Supercritical pulverised coal reference flowsheet.

3.1.1 Coal feed

The specifications of the coal used as fuel for the PCPP illustrated in Figure 32 are presented in Table 5.

Table 5: Coal feed specification.

Coal ultimate analysis (mass fraction)	**C**	0.665
	H	0.038
	O	0.054
	N	0.016
	S	0.005
	Water	0.080
	Ash	0.142
Milling energy requirement		40 kJ/kg
Specific heat capacity		1 kJ/kg/K
Coal temperature		30 ºC
LHV		25 MJ/kg

3.1.2 Air feed specification

The assumptions for the thermodynamic state of the inlet air feed are presented in Table 6 below.

Table 6: Inlet air specification.

Temperature		15 ºC
Pressure		1.01e5 Pa
Dry composition (molar basis)	**N_2**	78.09%
	O_2	20.95%
	Ar	0.93%
	CO_2	0.03%
Relative humidity		60%

3.1.3 Boiler specification

To account for the variation of boiler efficiency with reheat temperature and load point, a correction factor is introduced, according to the following expression.

$$\mathrm{LP}^{'}(\mathrm{Efficiency})_{\mathrm{correction}} = \mathrm{CE}_1 \times \mathrm{RHT}^{'} + \mathrm{CE}_2 \times \mathrm{RHT}^{'2} + \mathrm{CE}_3 \times \mathrm{LP}^{'} + \mathrm{CE}_4 \times \mathrm{LP}^{'2} + \mathrm{CE}_5 \times \mathrm{RHT}^{'} \times \mathrm{LP}^{'} \quad (1)$$

The normalised reheated temperature, RHT'RHT', and load point, LP'LP', in (1) are determined as follows.

$$\mathrm{RHT}^{'} = \frac{\mathrm{RHT}}{(\mathrm{RHT})_{\mathrm{design}}} - 1 \quad (2)$$

$$\mathrm{LP}' = \frac{\mathrm{LP}}{100} - 1 \quad (3)$$

Similarly, the reheat temperature is corrected according to the following expression.

$$\mathrm{RHT} = (\mathrm{RHT})_{\mathrm{design}} + \min\left(0, \mathrm{CR}_1 \times \mathrm{LP}^{'} + \mathrm{CR}_2\right) \quad (4)$$

Table 7 presents the operational and design parameters related to the supercritical pulverised coal boiler.

Table 7: Boiler specification.

Maximum continuous rating (MCR) thermal		1576.6 MW
Design efficiency		94.98%
Coefficients (CECE) of efficiency curve (1)	**1**	0
	2	0
	3	-2.13754
	4	-2.00792
	5	0
Design superheat temperature		600 ºC
Design reheat temperature		620.1 º C
Coefficients (CR)CR) of reheat correction curve (4)	**1**	94.379 K
	2	22.059 K
Design oxygen at furnace top		2.5% (m/m)
Design feedwater temperature		581.9 K
Mass fraction of ash in flue gas		0.77
Mass fraction of carbon in ash		5E-09
Boiler flue gas outlet pressure		1.0412e5 Pa
Temperature of bottom ash		600 ºC
Excess oxygen at 100% load		0.0305
Excess oxygen at minimum load		0.0305
Minimum load		50%
Superheater flowrate at MCR		604.28 kg/s
Superheater outlet pressure		285e5 Pa
Reheater flowrate at MCR		493.81 kg/s
Reheater flow conductance		43.82 bar m^3/kg
Typical pressure drop		1e5 Pa
Air heater flue gas inlet temperature		430.7 ºC
Flue gas pressure drop		792.3 Pa
Air heater leakage fraction		4.75 %
Inlet tramp air fraction		0 %

3.1.4 Turbine specification

Power is generated by expansion of steam in three sets of turbines, as shown in Figure 32: high-pressure (HP) turbine, intermediate pressure (IP) turbine, and low-pressure (LP) turbine.

The high-pressure turbine is composed of 2 steam turbines, as shown in Figure 33.

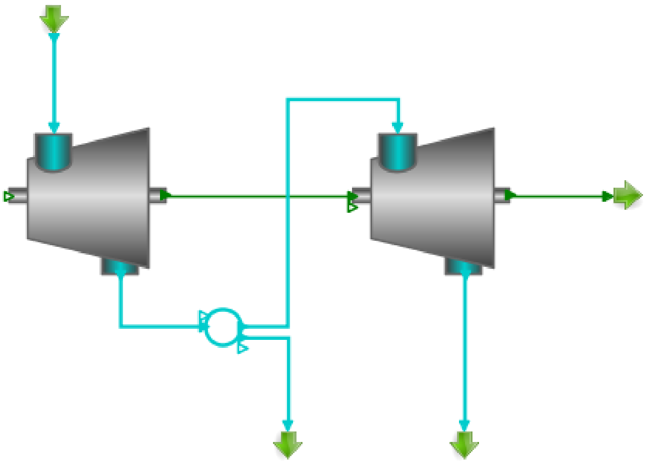


Figure 33: High-pressure steam turbine train composed of 2 turbines. Here, the inlet stream from the boiler section of the power plant is shown entering from the top. The bleed stream is shown to be in an intermediate position between both turbine stages and finally, the stream leaving the bottom of the second turbine stage is sent back to the reheater.

The specifications of the two turbines in Figure 33 are shown in Table 8

Table 8: High-pressure turbine specification.

	Stodola constant	Isentropic efficiency
Turbine 1	5088.3	91.73 %
Turbine 2	567.48	91.97 %

The intermediate-pressure and low-pressure turbines are composed of 3 and 5 steam turbines, respectively. The corresponding specifications are shown in

Table 9 and Table 10.

Table 9: Intermediate-pressure turbine specification.

	Stodola constant	Isentropic efficiency
Turbine 1	291.37	93.98 %
Turbine 2	90.17	93.24 %
Turbine 3	14.49	93.72 %

Table 10: Low-pressure turbine specification.

	Stodola constant	Isentropic efficiency
Turbine 1	10.52	91.32 %
Turbine 2	5.04	91.18 %
Turbine 3	0.78	91.59 %
Turbine 4	0.50	88.58 %
Turbine 5	0.13	96.75 %

3.1.5 Flue gas treatment section specification

The flue gas outlet from the boiler is subject to a treatment section, prior to being directed to the capture system. This section includes a selective catalytic reduction unit (SCR) (Table 11), an electrostatic precipitator (ESP) (

Table 12), a gas-gas heater (GGH) (

Table 13), and a flue gas desulphurisation unit (FGD) (Table 14 and Table 15), as shown in Figure 34.

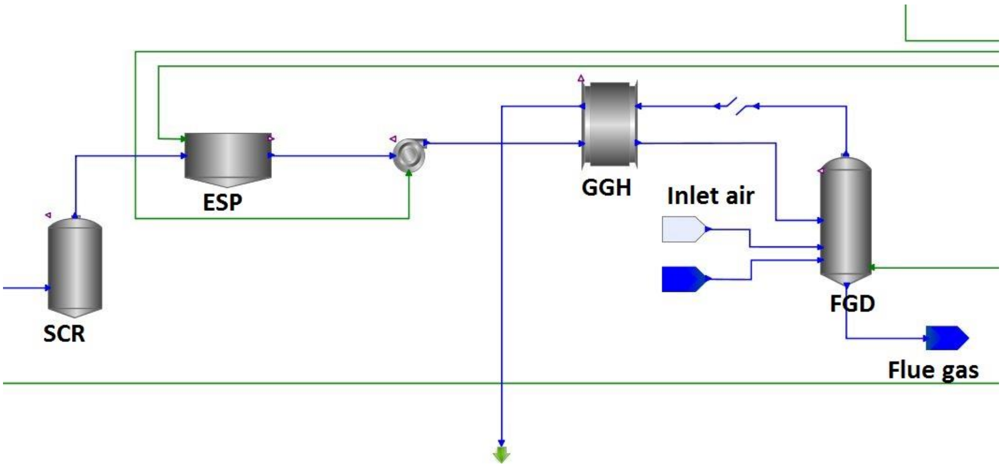


Figure 34: Schematic of flue gas treatment section.

Table 11: Selective catalytic reduction (SCR) unit specification.

Pressure drop	0 Pa
NOx removal efficiency	90 %

Table 12: Electrostatic precipitator (ESP) specification.

Pressure drop	0.015e5 Pa
Efficiency	99.5 %

Table 13: Gas-gas heater (GGH) specification.

Pressure drop on hot-side	0.005e5 Pa
Hot gas outlet temperature	90 ºC
Pressure drop on cold-side	0.005e5 Pa

Table 14: Flue gas desulphurisation (FGD) unit specification.

Efficiency	98 %
Pressure drop	0.030e5 Pa
O_2 ratio	1.4 mol/mol
CaCO3/SO2 ratio	1.05 mol/mol
Limestone purity	0.95 kg/kg
Limestone slurry temperature	45 ºC
Solids in limestone slurry	0.1 kg/kg
Gypsum in slurry before dewatering	0.14 kg/kg
Outlet gypsum content	0.1 kg/kg

Table 15: Flue gas desulphurisation air inlet specification.

Temperature		25 ºC
Pressure		1.013e5 Pa
Dry composition (molar basis)	**N_2**	78.084 %
	O_2	20.946 %
	Ar	0.934 %
	CO_2	0 %
Moisture content		0 %

3.2 Reference capture process for SC-PC power plant

The reference capture process chosen for this study is an amine-based system using MEA (30 wt%) as solvent and comprising an absorber section and a solvent regeneration section, as shown in Figure 35. The solvent circulation rate is 3340 kg/s (L/G ratio ~4kg/kg) and the lean loading 0.28 mol CO_2/mol MEA. However, in order to reduce simulation time, the flows were split between two identical post-combustion capture plants, and this is what is reflected in the results section of this report. The steam used to provide heat to the reboiler is extracted from the outlet of the intermediate pressure turbine of the power plant. The flue gas is cooled down by adding water in the direct contact cooler before it enters the absorber. The flue gas composition at the inlet of the absorber is shown in Table 16.

Table 16: Flue gas composition at the absorber (mass fraction)

N_2	0.75
H_2O	0.04
CO_2	0.21

In the reference case, the capture rate is controlled to achieve a value of 90% by manipulating the solvent flowrate pump located at the top of the absorber. From this unit, the solvent is sent to the cross heat exchanger where it is heated up and before it is sent to the stripper, it is mixed with the condensate of the condenser unit. Both absorber and stripper are modelled with a rate-based column model, using gSAFT for the physical property calculations and solving the equilibrium at the vapour-liquid interface. This model is discretised along the height to achieve more accuracy and to provide insight of the column's performance across its whole length, both in terms of CO_2 flux and pressure/temperature profiles. The pressure in the reboiler is controlled by a valve located at the top of this unit. The pressure of this unit is controlled by a valve which sets the CO_2 flowrate sent to compression.

The solvent exiting the reboiler is cooled down in the cross heat exchanger and it is stored in the lean solvent tank. In this unit, MEA and water make-ups are also added. The solvent exiting the lean solvent tank is cooled down and sent again to the absorber. The specifications of the main units in the reference capture process are presented in sections 3.2.1-3.2.3.

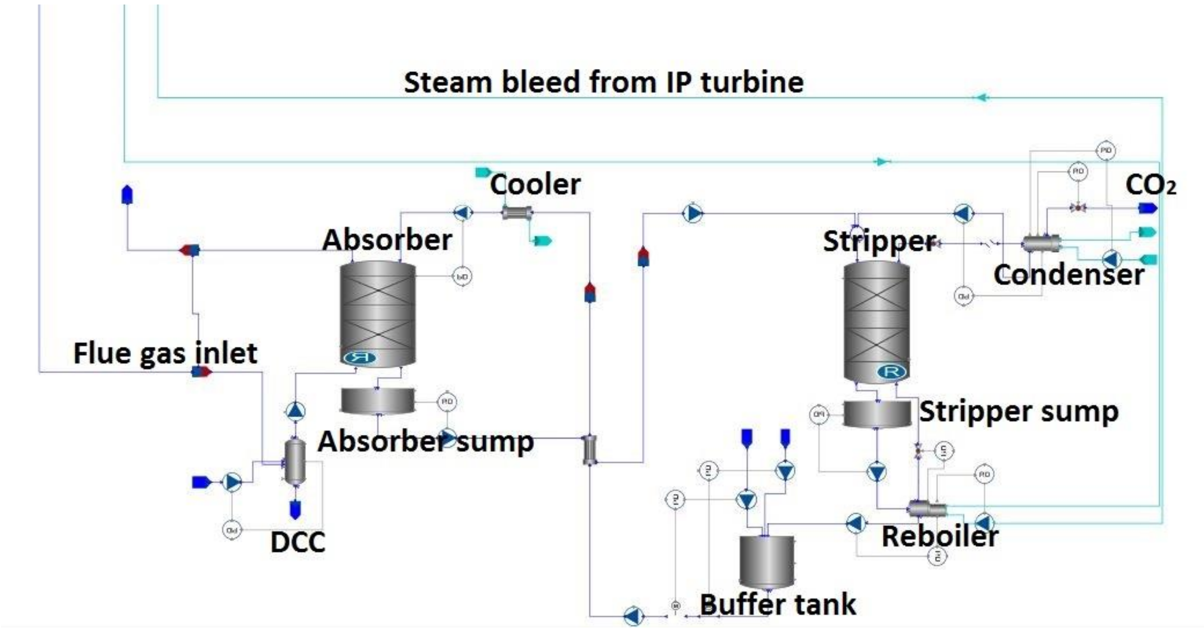


Figure 35: Reference capture process flowsheet.

3.2.1 Absorber section

The absorber column and absorber sump specifications are presented in Table 17 and Table 18, respectively. Absorber specifications were determined on the following basis: first the diameter was calculated based on the volumetric gas flow rate – here the aim was to ensure that at the design point the superficial gas velocity was 2.8 m/s, consistent with operating at 70% of column flood point. Then the solvent flow rate and lean loading were selected following the parameterisation procedure laid out in Mac Dowell and Shah [25]. Finally, the height of packing was increased to achieve a 90% capture rate, assuming a structured packing.

Table 17: Absorber column specification.

Packing height	55 m
Diameter	20 m
Mass transfer model	Fick
Pressure drop correlation	Dry bed friction factor correlation
Mass transfer coefficient correlation	Onda
Dry bed packing factor	66
Capture rate	90%

Table 18: Absorber sump specification.

Packing height	2 m
Diameter	20 m

3.2.2 Stripper section

The stripper column and stripper sump specifications are presented in Table 19 and Table 20, respectively.

Table 19: Stripper column specification.

Packing height	20 m
Diameter	10 m
Mass transfer model	Fick
Pressure drop correlation	Dry bed friction factor correlation
Mass transfer coefficient correlation	Onda
Dry bed packing factor	55

Table 20: Stripper sump specification.

Packing height	2 m
Diameter	10 m

It must be noted that after a sensitivity analysis conducted on the model, the stripper's size showed limited influence on the degree of solvent regeneration. However, it is also important to note that these models do not describe maldistribution or wall effects on flow, so the influence of stripper sizing may be underestimated. The reboiler temperature (or rate of steam condensation) was much more influential and this variable was manipulated to achieve the optimal lean loading for the capture model.

3.2.3 Heat exchanger

The rich solvent coming from the absorber sump and the lean solvent from the top of the stripper column transfer heat in a heat exchanger as shown in Figure 35. In designing this unit, a 10K approach temperature was used. The heat exchanger specifications are presented in Table 21.

Table 21: Heat exchanger specification.

Heat transfer coefficient	600 W/(m^2 K)
Pressure drop in cold stream	1e4 Pa
Pressure drop in hot stream	1e4 Pa
Heat transfer area	71,078 m^2

3.3 Reference compression train for SC-PC power plant

This section describes the compression system used for the decarbonised SCPC power plant. As we were strictly considering the integration of the power and capture plants from a control perspective, this model is not discussed further in the control section. However, it was important to simulate the compression system in order to accurately calculate the efficiency penalty imposed on the power plant by the capture plant.

Two compressor trains are used to increase the CO_2 pressure to 108bar. As both trains are identical, flow multipliers are used to reduce simulation times. Each train comprises four compressor sections. The first three are connected to one drive and the fourth to a second drive. The latter is a variable frequency drive which controls the pressure after the dehydration unit by manipulating the speed of the last compressor. After each compressor section the CO_2 is cooled down by using a water-cooler model. Following this model, a knock-out drum removes any condensed water from the CO_2 stream. After two compression stages, a molecular sieve unit is used to further reduce the water content of the CO_2 stream prior to the final two compression stages.

The compressor section models are based on head and polytropic efficiency maps. These maps are created by the same model using design heuristics given design point conditions.

The train of 4 compressors is shown in Figure 36.

The first two compressors are separated from the last two by a dehydration unit. The outlet pressure for each of these sections is controlled by separate PI controllers that manipulate the gear speed of the compressors in the corresponding section.

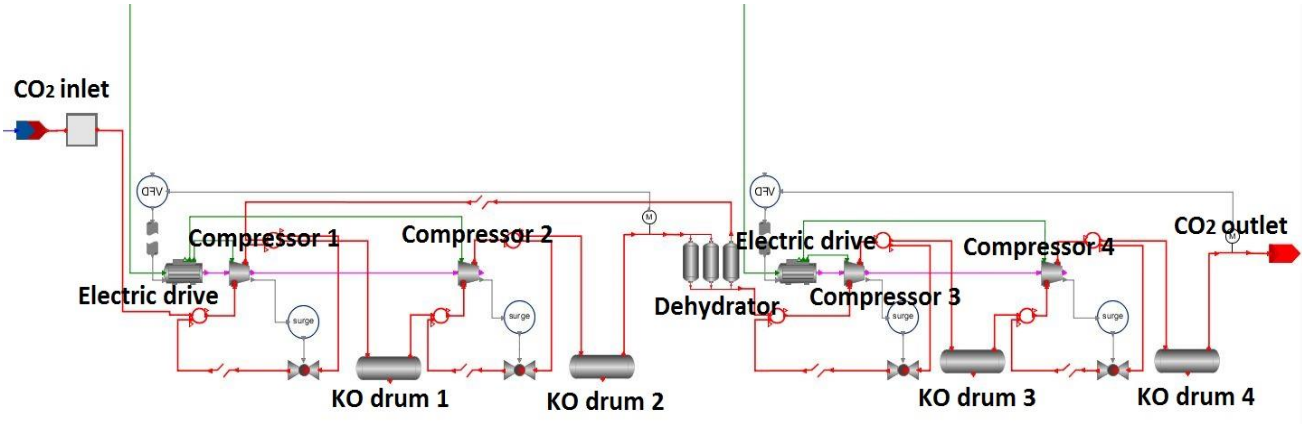


Figure 36: Compression train flowsheet.

Table 22 - Table 25Table 24 show the specification of each of the compressors, the cooler knock-out drums between compressors, the dehydration unit, and the tuning parameters of the two PI controllers.

Table 22: Compression train specification.

	Compressor 1	Compressor 2	Compressor 3	Compressor 4
Impeller diameter (m)	1.22	0.938	0.334	0.254

Table 23: Cooler knock-out drums specification.

	KO drum 1	KO drum 2	KO drum 3	KO drum 4
Outlet temperature (°C)	40	40	40	73
Pressure drop (Pa)	4e4	4e4	4e4	2e4

Table 24: Dehydration unit specification.[26]

Outlet moisture content	50 ppm [27]
Pressure drop per bed	0.9e5 Pa
Heat of adsorption	4220223 J/kg H_2O
Recycle fraction	0.1 kg/kg
Heating efficiency	90 %
Regeneration temperature	250 ºC

Table 25: Pressure controller specification and tuning parameters.

	Section 1	Section 2
Set-point	3.463E6	1.252e07 Pa
Minimum input	3e6 Pa	3e6 Pa
Maximum input	4e6 Pa	1.5e7 Pa
Gain	5	10
Reset action	10	10

3.4 Combined cycle gas turbine power plant model

Figure 37 shows the flowsheet developed for a combined cycle gas turbine power plant producing a gross electricity output of 750MW at full-load.

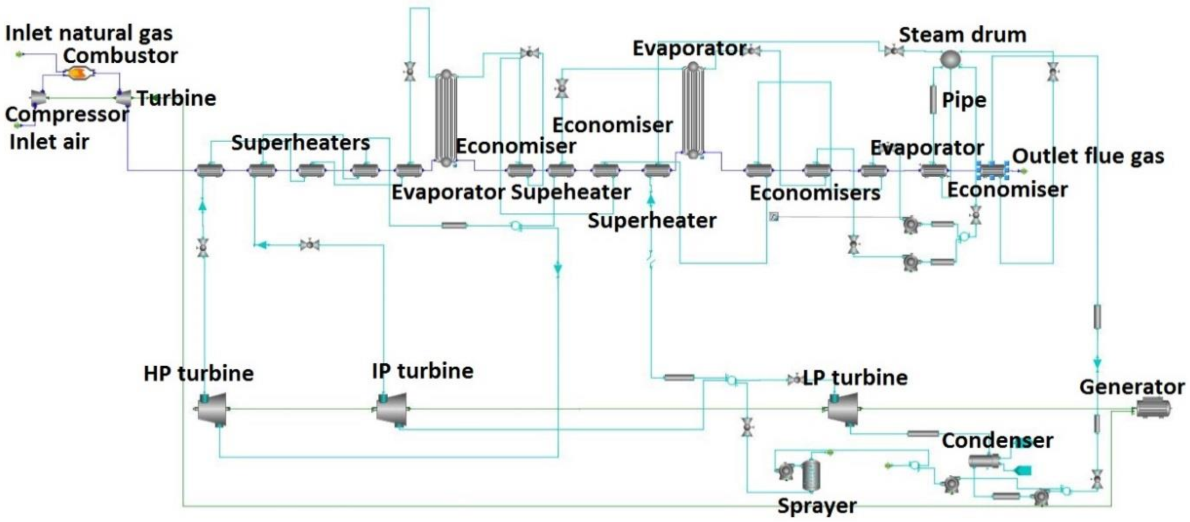


Figure 37: Combined cycle gas turbine power plant reference flowsheet.

The Combined Cycle Gas Turbine flowsheet comprises two Gas Turbines (GT) connected to two Heat Recovery Steam Generators (HRSG). The steam generated in the two HRSG is expanded in the High Pressure (HP), Intermediate Pressure (IP) and Low Pressure (LP) turbines.

The GT is simulated by modelling the air compressor, combustor and gas turbine units individually. The natural gas is fully combusted in presence of air in the combustor unit model. Both the gas turbine and compressor units contain polynomials for calculating the isentropic efficiency based on data at part-load operation. The gas turbine is assumed to be choked in this study. Two identical trains consisting of a gas-turbine, one HRSG unit and one steam-cycle have been considered.

The HRSG comprises three pressure levels with natural circulation evaporators. The low pressure pump increases the pressure of the condensate and sends it to the low pressure economiser. The low pressure feedwater is sent to the low pressure evaporator. The natural circulation process is simulated by modelling the downcomer, riser and drum units individually. Feedwater from the low pressure drum is extracted and pumped using the IP and HP pumps. Both pumps send the IP and HP feedwater to a series of economisers and then to the IP and HP evaporators. In this case, the drum, downcomer and riser are modelled individually. The steam produced in the low pressure drum is heated-up in a superheater and it is mixed with the IP turbine exhaust before it is expanded in the low pressure turbine. The exhaust is condensed using a water-cooled condenser. The HP steam is sent to a series of superheaters and it is expanded in the HP turbine. The exhaust steam from this turbine is then mixed with the IP steam and reheated before it is expanded again in the IP turbine.

The steam required by the capture plant is extracted at the IP/LP cross over. This steam is desuperheated using condensate from the reboiler of the capture plant. A control valve ensures the pressure at the extraction point does not fall below 3.5bar.

The following sections present the specifications for the main units and model inputs.

3.4.1 Inlet air specification

Table 26: Inlet air specification.

Temperature		25 ºC
Pressure		1.01e5 Pa
Dry composition (molar basis)	**N_2**	78.09%
	O_2	20.95%
	Ar	0.93%
	CO_2	0.03%
Relative humidity		60%

3.4.2 Natural gas inlet specification

Table 27: Natural gas inlet specification.

Temperature		25 ºC
Flowrate		13.4 kg/s
Composition (% m/m)	**CH_4**	96.109
	N_2	3.891

3.4.3 Natural gas combustor specification

The pressure drop in the natural gas combustor is specified using the polynomial function (5), where F_{out} is the flowrate of exhaust gas leaving the combustion chamber. The coefficients of function (5) are given in

Table 28.

$$\Delta P = a_0 F_{out} + a_1 F_{out}^2 + a_2 F_{out}^3 + a_3 F_{out}^4 + a_4 F_{out}^5 \tag{5}$$

Table 28: Natural gas combustor specification.

Heat losses	1e6 W
$\boldsymbol{a_0}$	-8399.4
$\boldsymbol{a_1}$	92.853
$\boldsymbol{a_2}$	0
$\boldsymbol{a_3}$	0
$\boldsymbol{a_4}$	0

3.4.4 Turbine specification

The flowsheet includes one gas turbine, for which the specification is given in Table 29, and three steam turbines, specified in Table 30. The isentropic efficiencies of the turbines are specified as per equation (6).

$$\eta_{is} = d_0 + d_1 \times \frac{F}{F_{max}} + d_2 \times \left(\frac{F}{F_{max}}\right)^2 + d_3 \times \left(\frac{F}{F_{max}}\right)^3 \quad (6)$$

Table 29: Gas turbine specification.

Choke coefficient	0.017905487 kg $K^{0.5}$ / (Pa.s)
Outlet pressure	1.013e5 Pa
Maximum flowrate	607kg/s
d_0	39.438
d_1	50.056
d_2	0
d_3	0

Table 30: Steam turbine specification.

	HP steam turbine	MP steam turbine	LP steam turbine
Stodola coefficient	18967.863	559.02563	24.071047
Maximum flowrate	151 kg/s	176kg/s	194kg/s
d_0	41.069	28.399	40.674
d_1	47.042	67.755	51.138
d_2	0	0	0
d_3	0	0	0

3.5 Reference capture process for CCGT power plant

The capture process integrated with the CCGT power plant is analogous to the capture process shown in Section 3.2. Table 31 shows the flue gas inlet composition after the DCC unit at the absorber.

Table 31: Flue gas composition at the absorber (mass fraction)

N_2	0.89
H_2O	0.05
CO_2	0.06

3.5.1 Absorber section

Absorber specifications were determined on the following basis: first the diameter was calculated based on the volumetric gas flow rate – here the aim was to ensure that at the design point the superficial gas velocity was 2.8 $m.s^{-1}$, consistent with operating at 70% of column flood point. Then the solvent flow rate and lean loading were selected following the parameterisation procedure laid out in Mac Dowell and Shah [25]. Finally, the height of packing was increased to achieve a 90% capture rate.

Table 32: Absorber column specification.

Packing height	50 m
Diameter	20 m
Mass transfer model	Fick
Pressure drop correlation	Dry bed friction factor correlation
Mass transfer coefficient correlation	Onda
Dry bed packing factor	66
Capture rate	90%

Table 33: Absorber sump specification.

Length	2 m
Diameter	20 m

It is noted that the data presented in Table 17 (for the PCCP) and Table 32 (for the CCGT) imply the use of relatively large columns. The same effect could be achieved by the deployment of two or more smaller units. However, it should also be noted that the absorber unit that was designed for the now-cancelled Peterhead CCS project was intended to have dimensions of 20x15x60m.

3.5.2 Stripper section

Table 34: Stripper column specification.

Length	20 m
Diameter	10 m
Mass transfer model	Fick
Pressure drop correlation	Dry bed friction factor correlation
Mass transfer coefficient correlation	Onda
Dry bed packing factor	55

Table 35: Stripper sump specification.

Length	2 m
Diameter	10m

3.5.3 Heat exchanger

Table 36: Heat exchanger specification.

Heat transfer coefficient	600 W/(m^2 K)
Pressure drop in cold stream	1e4 Pa
Pressure drop in hot stream	1e4 Pa
Heat transfer area	10,5742.66 m^2

3.6 Reference compression train for CCGT power plant

As in the flowsheet described in Section 3.3 the compression train for the CCGT power plant comprises two sections, separated by a dehydration unit. As previously, the dehydration unit is modelled as a molecular sieve. In this case, the first section included three compressors and the second section one compressor. As before, the outlet pressure for each of these sections is controlled by separate PI controllers that manipulate the gear speed of the compressors in the corresponding section. The following tables show the specification of each of the compressors, the dehydration unit, and the tuning parameters of the PI controller.

Table 37: Compression train specification.

	Compressor 1	Compressor 2	Compressor 3	Compressor 4
Impeller diameter (m)	0.952	0.642	0.444	0.254

Table 38: Dehydration unit specification.

Outlet moisture content	41 ppm
Pressure drop per bed	1e5 Pa
Heat of adsorption	4220223 J/kg H_2O
Recycle fraction	0.06 kg/kg
Heating efficiency	90 %
Regeneration temperature	327 ºC

Table 39: Pressure controller specification and tuning parameters.

Set-point	4432695.5 Pa
Minimum input	10e5 Pa
Maximum input	50e6 Pa
Gain	5
Reset action	10

3.7 Integrated reference cases

3.7.1 Supercritical pulverised coal power plant model

An integrated model has been developed to allow the dynamic simulation of the entire system, comprising power plant, amine-based capture process, and compression train, as depicted in Figure 38.

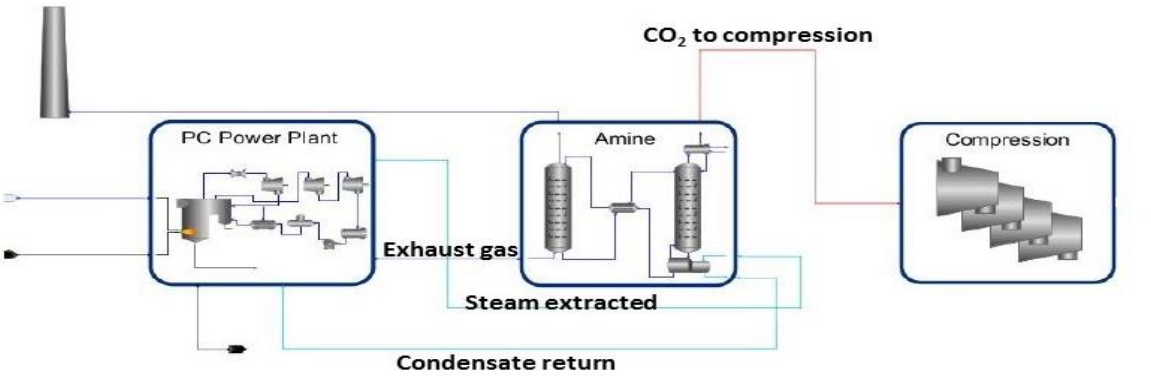


Figure 38: Integrated supercritical pulverised coal power plant flowsheet with capture and compression.

It is important to note that the image in Figure 38 is intended for illustrative purposes, only. The models describing the PCPP, capture plant and compression train are those presented previously. The intention of this image is to illustrate the connection between the power, capture and compression plants.

The carbon capture process and the compression train impose a penalty on the electricity generated in the power plant, as shown in Table 40. The decrease in gross electricity generated is due to the steam extraction.

Table 40: Electricity generated in the coal-fired power plant with and without capture and compression.

	Power plant only	With capture
Gross electricity generated	825 MW	700 MW
Net electricity generated	779 MW	621 MW

The breakdown of power consumption in the power plant is presented in Table 41.

Table 41: Breakdown of power consumption in coal-fired power plant with capture and compression.

CO_2 compression train (section 1)	26 MW
CO_2 compression train (section 2)	7 MW
Pumps and utilities	46 MW

Note that this excludes the steam extraction from the turbine train for solvent regeneration, as this is not a power consumption, per se, rather it corresponds to a reduction in power generated. Results of preliminary simulations of the integrated process are shown in Appendix A.

3.7.2 Combined cycle gas turbine power plant model

Similarly to the supercritical pulverised coal power plant, the combined cycle gas turbine power plant has been integrated with an amine-based capture system and a compression train as it is illustrated in Figure 39.

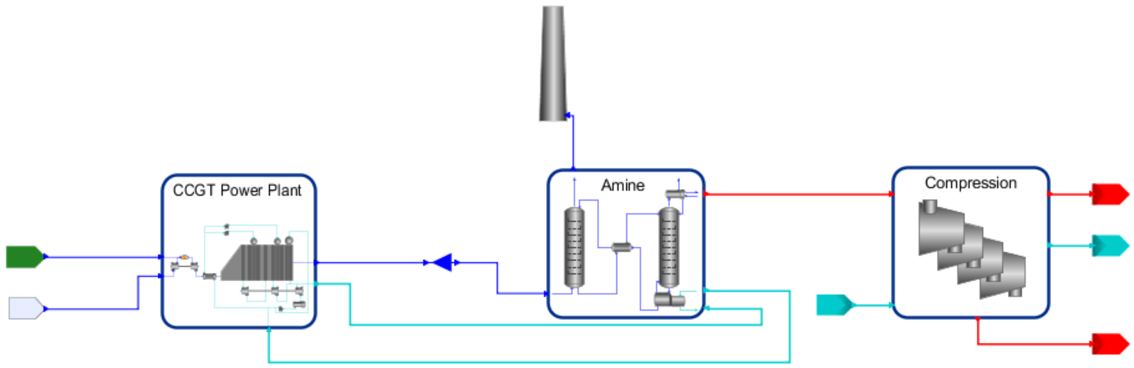


Figure 39: Integrated combined cycle power plant flowsheet with capture and compression.

Table 42: Electricity generated in the combined cycle power plant with and without capture and compression.

	Power plant only	With capture
Gross electricity generated	746 MW	660 MW
Net electricity generated	740 MW	643 MW

Table 43: Breakdown of power consumption in combined-cycle power plant with capture and compression.

CO_2 compression train (section 1)	9 MW
CO_2 compression train (section 2)	2 MW
Pumps and utilities	6 MW

4 Overview of control strategies

Studies on process control strategies for post-combustion carbon capture plants in the literature often suggest that the percentage of CO_2 capture should remain approximately constant throughout the power plant operation [15-17]. In this study we follow this principle, and consider two alternative strategies for controlling the CO_2 capture rate. Additionally, we consider a strategy in which the control loops are dynamically switched during the change in power plant load. Some works in the literature suggest that the best control strategy for a post-combustion carbon capture plant depends on the operating region [20]. The turndown range of the absorption column is largely dictated by the column internals used. For example, whilst it is highly likely that packed columns will be preferred to tray columns, the distinction between dumped (random) packing and structured packing is important. Moreover, the type of liquid inlet system and the number and type of liquid and vapour distributors used will play an important role in determining the flexibility of the absorption column of the CCS plant. Finally, the availability or otherwise or liquid collectors will also exert an important influence here. It may be important to consider these factors when the detailed design of CO_2 capture plants is being performed in the future; it will behove the designers to consider the additional requirements placed upon them to satisfy requirements for flexible operation, as opposed to merely steady-state operation at the column's design point.

It is also important to realise that the expected performance of the CCS plant will change and evolve with time, in sympathy with the role of the power plant in the energy market. Where one would initially, rationally, design the CCS plant for steady state operation at, or near, its original design point, the required operation may subsequently change to encompass more operation in off-design conditions. It will further be important to ensure that the CCS plant can operate efficiently in off-design conditions without the requirement for major retrofit works. Therefore, one could envision a scenario where initially, the solvent flowrate is reduced until just before the onset of either underwetting or entrainment. Thereafter, the lean loading of the solvent is varied – likely increased, thereby giving the effect of a reduced solvent flow – allowing the power plant to be turned down further. These effects have been examined in detail in some of our recent publications [12, 28].

These options, listed in Table 44, are discussed in the sections below.

Table 44: Control strategies proposed in this study.

Strategy	*Controlled variable*	*Manipulated variable*
1	CO_2 capture	Lean solvent flowrate
	Reboiler temperature	Steam flowrate
2	CO_2 capture	Reboiler temperature
	Lean solvent flowrate	Lean solvent loading
3	Dynamic switching between strategies 1 and 2	

In addition to the CO_2 capture rate control loop, there are a number of other loops that need to be included to guarantee the operation of the post-combustion plant during the load change.

4.1 Strategy 1: Control of capture rate using lean solvent flow

In this strategy, a controller maintains the carbon capture rate at a value of approximately 90%, by regulating the flow of lean solvent entering the absorber column. This control loop is depicted in Figure 40. Note that the control signal originating from the absorber column – the CO_2 capture rate – is calculated based on the flue gas streams entering and exiting the column.

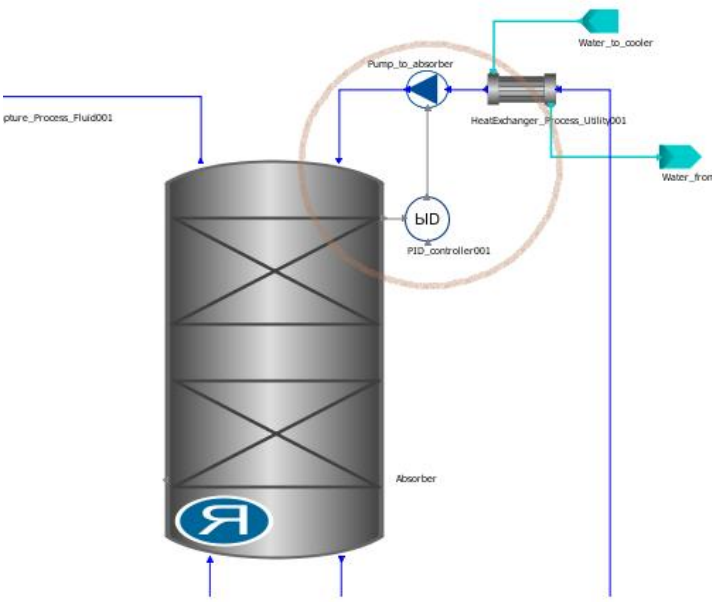


Figure 40: Control strategy 1: the flow of lean solvent entering the absorber column is manipulated in order to control the CO2 capture rate.

The CO_2 loading of the lean solvent is determined by the temperature in the reboiler of the solvent regeneration column. Therefore, in this strategy, the lean solvent loading is kept approximately constant by manipulating the flow rate of steam extracted from the intermediate pressure turbine of the power plant, as shown in Figure 41.

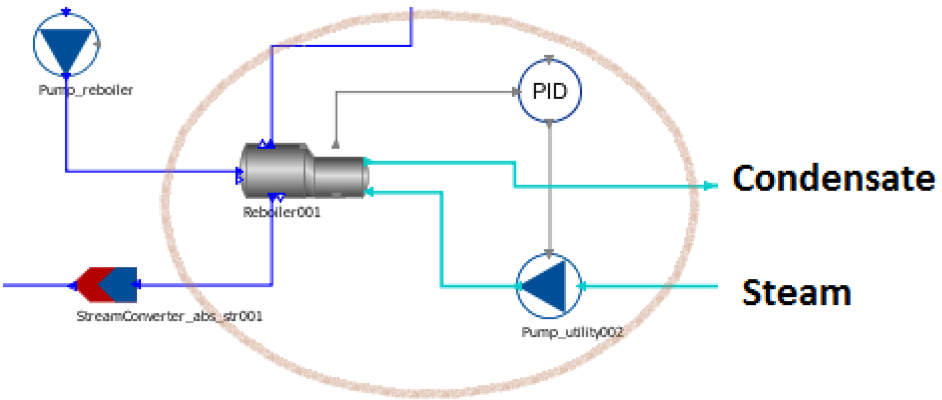


Figure 41: Control strategy 2: The lean solvent CO_2 loading is kept approximately constant by manipulating the steam extracted from the power plant

4.2 Strategy 2: Control of capture rate using lean solvent loading

As in strategy 1, the objective is to maintain the CO_2 capture rate constant throughout the power plant load change. In this case, the flow of solvent circulating between the absorber and regeneration columns is kept at a constant value, throughout the operation. The flow of steam extracted from the power plant is manipulated in order to provide a variable heat duty to the reboiler of the regeneration column. In this manner, the temperature of the reboiler, and therefore the lean solvent CO_2 loading, can be manipulated to reach the CO_2 capture rate target set by the controller. This strategy is depicted in

Figure 42.

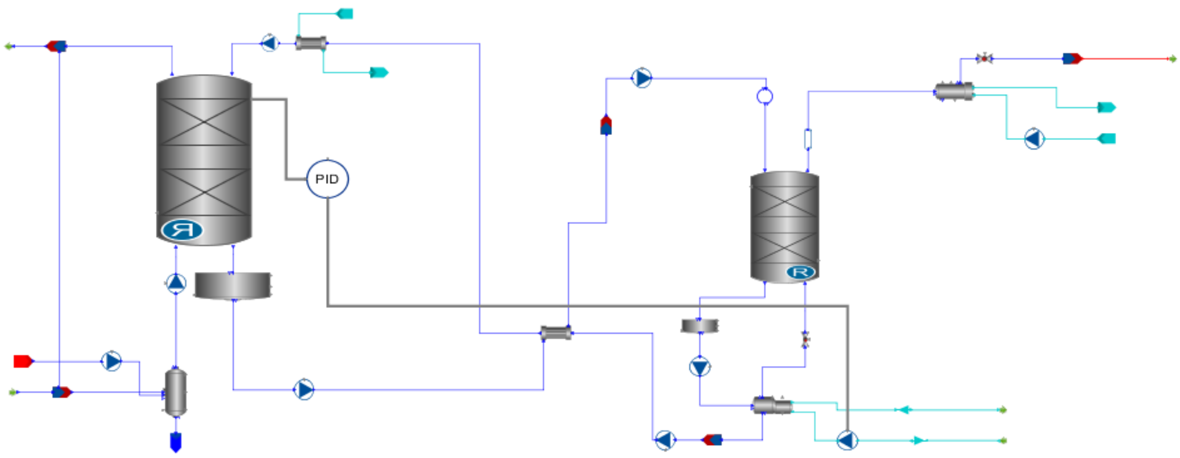


Figure 42: Control strategy 2: the lean solvent CO_2 loading is manipulated in order to control the CO_2 capture rate.

This strategy has the benefit of maintaining the flow conditions in both columns, since the flow rate of circulating solvent is kept constant. Although the gas flow in the regeneration column is changed, the flow conditions can be maintained, if needed, by recirculating part of the CO_2 produced stream, as proposed by Lin et al. [15, 17].

4.3 Strategy 3: Dynamic switching between strategies 1 and 2

In this strategy we propose a dynamic switching approach that alternates between control strategies 1 and 2 depending on how far the power plant is from 100% load.

As mentioned above, the control strategies include a set of common control loops required to guarantee the operation of the carbon capture plant during the power plant load change.

Table 45: Additional control loops in the process.

Type of controller	Controlled variables	Manipulated variables	Set-point
Level controller	Reboiler liquid level	Lean MEA from reboiler mass flow	0.41m
Temperature controller	Condenser temperature	Cooling water mass flow	313.15 K
Level controller	Condenser level	Lean MEA from condenser mass flow	0.71m
Level controller	Absorber sump liquid level	Rich MEA mass flow	0.63m
Quality controller	Concentration of MEA	Mass flowrate of make-up MEA	30.9%
Quality controller	Concentration of water	Mass flowrate of make-up water	63%
Temperature controller	DCC temperature	Cooling water mass flow	314.1 K
Pressure controller	Reboiler pressure	Stem position of the gas stream valve	1.69 bar
Pressure controller	Condenser pressure	Stem position of the product stream valve	1.66 bar
Level controller	Stripper sump liquid level	Outlet stream mass flow	0.5m

The balance of solvent (MEA) and water in the process needs to be maintained by including two make-up streams, given that part of these components is inevitably lost with the clean flue gas stream emitted through the stack and with the CO_2 product, as it is illustrated in Figure 43. To achieve this, a tank is included in the flowsheet, in which the solvent from the regeneration section is mixed with the make-up streams. The concentration of water and MEA is measured at the outlet of the tank and controlled by manipulating the flow of the inlet streams. Note that this measurement is effective in the case of control strategy 1, where the CO_2 loading of the lean solvent is maintained constant. In the case of control strategy 2, the set-point of these controllers will have to be calculated based on the amount of water and MEA lost in the outlet streams of the process, since the lean loading is used as a manipulated variable and varies to control the capture rate throughout the operation of the plant.

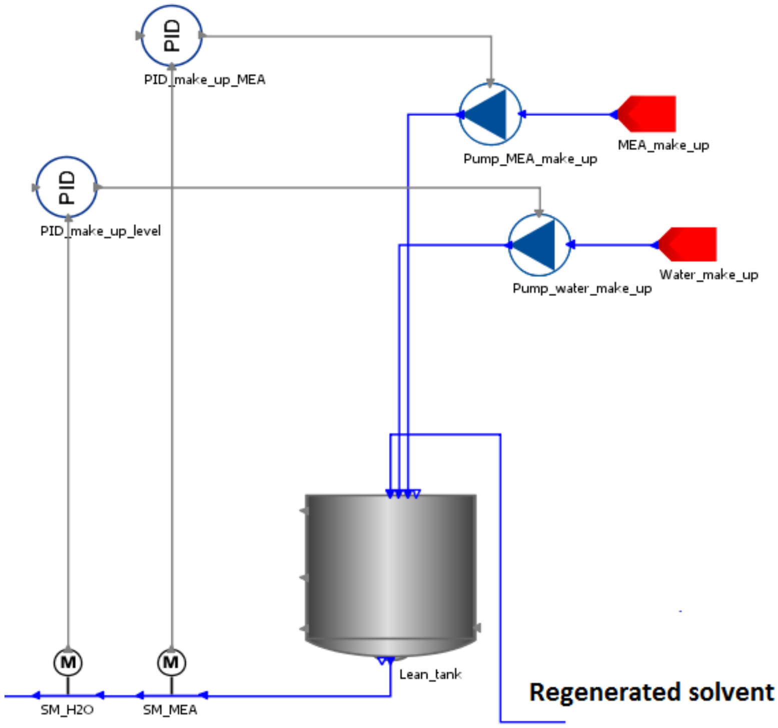


Figure 43: PI controllers regulating the flow of the make-up streams of water and MEA.

Another control loop, illustrated in Figure 44, maintains the temperature of the direct contact cooler (DCC unit) used to reduce the temperature of the flue gas entering the absorber column. The control is achieved by manipulating the flow rate of cooling water.

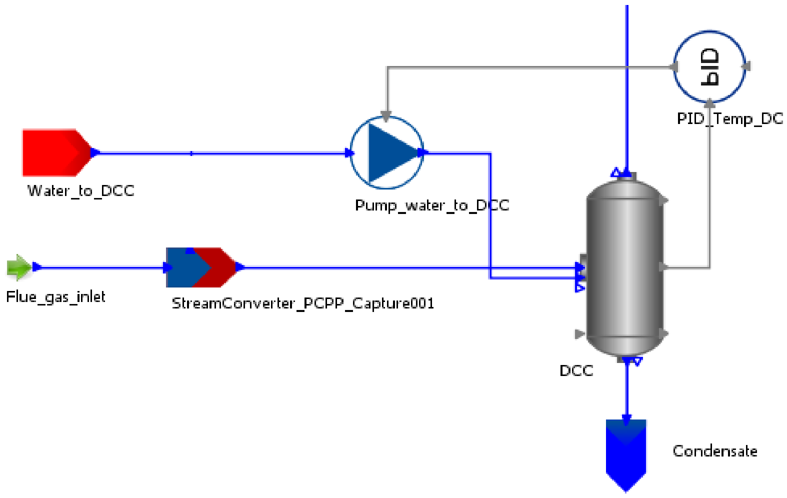


Figure 44: Control of the temperature in the DCC unit.

The level of liquid in the sump of the absorber and regeneration columns is maintained approximately constant by manipulating the flow rate of the outlet stream, as shown in Figure 45

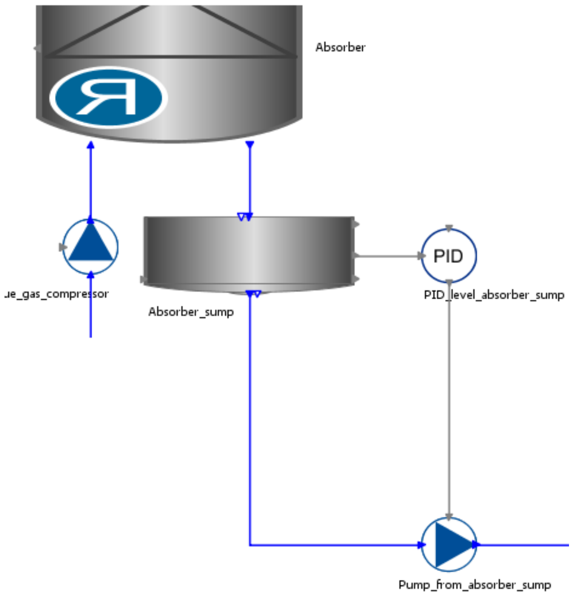


Figure 45: Level control in the absorber column sump.

The stripper section of the process includes control loops for the reboiler and for the condenser of the regeneration column.

As shown in Figure 46, the level of liquid in the reboiler is controlled by adjusting the flow rate of the bottom stream, while the pressure in the reboiler is controlled by manipulating the stem position of the gas stream valve.

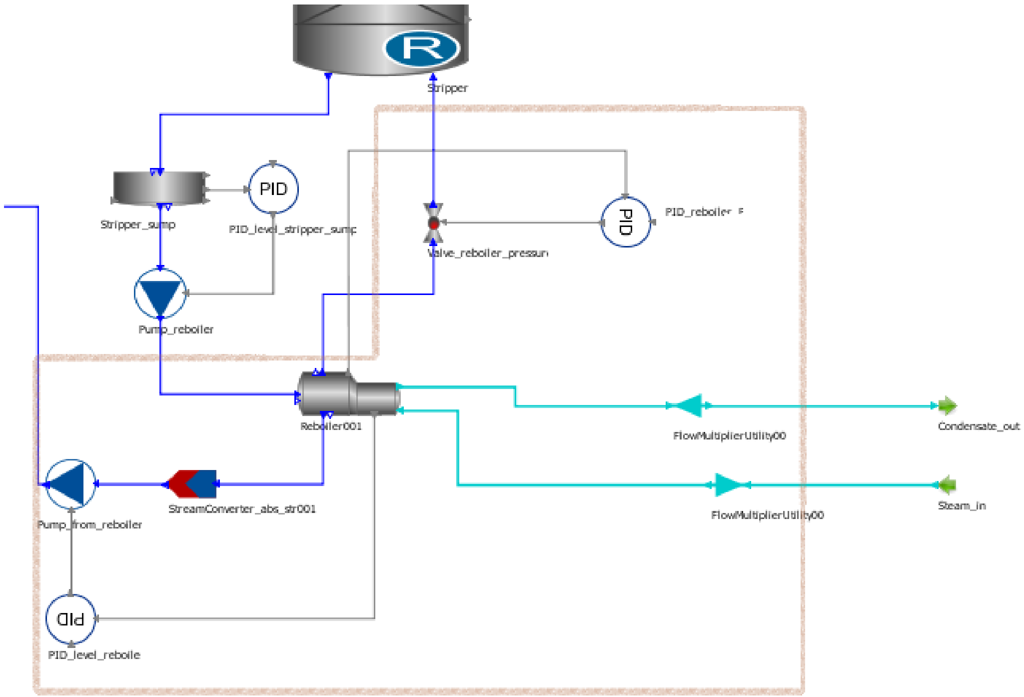


Figure 46: Pressure and level controller in the reboiler of the regeneration column.

In the condenser of the regeneration column, the pressure and temperature are controlled by manipulating the stem position of the product stream valve and the flow rate of cooling water, respectively. This section is shown in Figure 47.

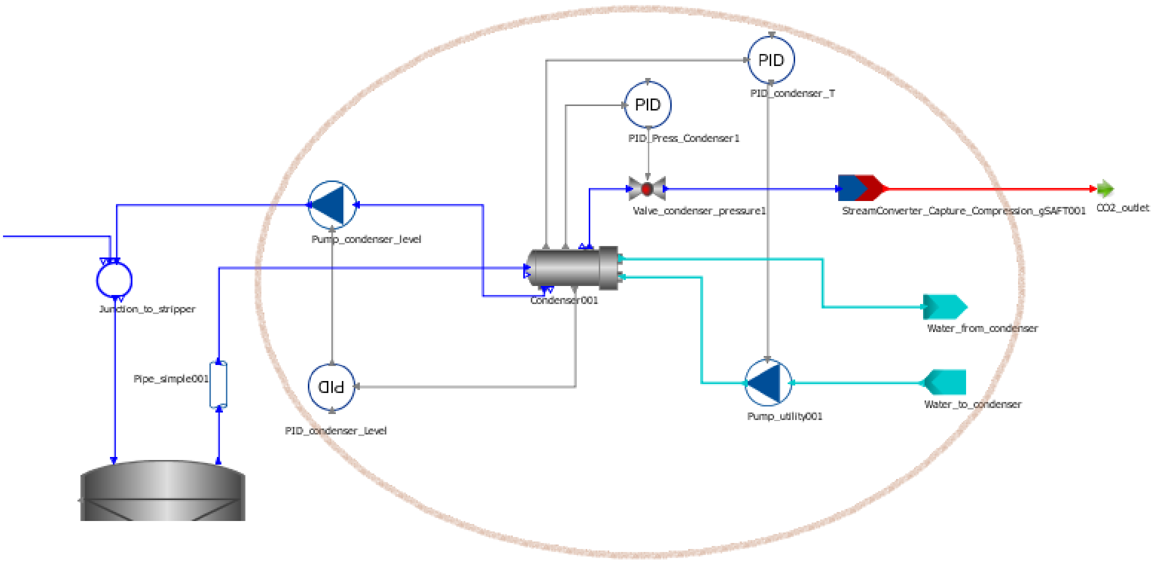


Figure 47: Controllers for the stripper column condenser: Pressure, temperature, and level.

5 Tuning of control strategies

The tuning of control loops may generally be accomplished by applying certain systematic methods, such as the Ziegler-Nichols method. In that method, the procedure starts by setting the controllers to proportional (P) mode. A step-disturbance is then simulated, and the controller gain increased until the controlled variable shows a sustained oscillatory response, from which certain values may be extracted and used to calculate the tuning controllers of a proportional-integral (PI) or proportional-integral-derivative (PID) controller.

However, when using gCCS it was found that varying the controller gain in such way often resulted in numerical errors due to the complexity and high number of equations involved in the process. Therefore, the method adopted to tune the controllers consisted of manually increasing the controller gain until such numerical errors were encountered, and then slowly increasing the integral action of the controller to remove the off-set from the set-point and to obtain a fast and non-oscillatory response. It was found that good results could be obtained using a proportional-integral (PI) controller which follows the following control equation.

$$u(t) = K\big(SP - Mv(t)\big) + \frac{1}{\tau_i}\int_0^t \big(SP - Mv(\tau)\big)d\tau \tag{7}$$

In equation (7), K is the proportional gain, τ_i is the integral action, SP is the set-point, u is calculated control action, and Mv is the measured variable.

The following sections show the results of this procedure for control strategies 1 and 2 for the capture processes in the pulverised coal and natural gas fired plants. The tuning parameters thus obtained were also used for control strategy 3, since it consists of a combination of strategies 1 and 2.

5.1 Pulverised-coal power plant (PCPP)

To tune the controllers, a step disturbance of +20% and -20% in the flowrate of the flue gas inlet of the capture plant was used. A period of 5 hours is allowed between the two step-disturbances, as shown in Figure 48.

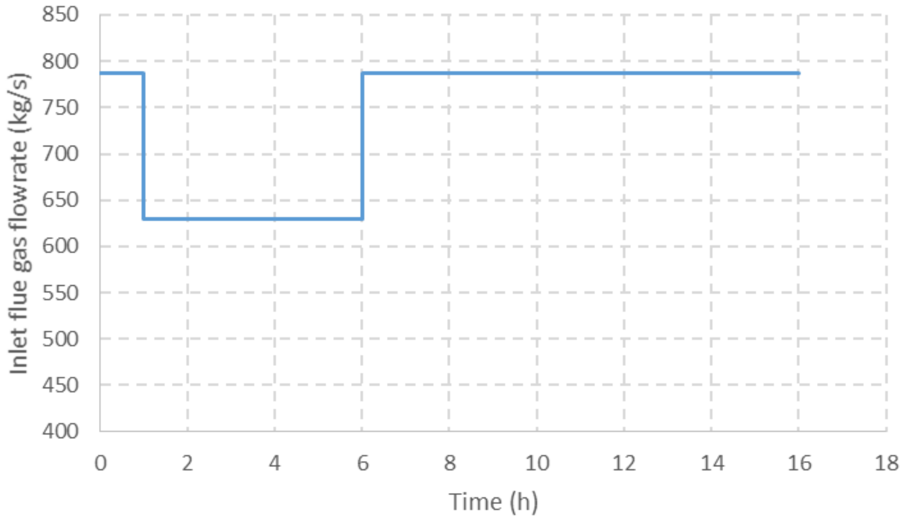


Figure 48: Step disturbances in flowrate of flue gas used for controller tuning in the PCPP process.

5.1.1 Strategy 1

For control strategy 1, it was found that good disturbance rejection and settling time could be achieved within a wide range of tuning parameters in which no significant differences in the control responses were observed. Therefore, the parameters were chosen to allow fast convergence of the simulation, while maintaining good controller performance. The chosen parameters are shown in Table 46. Figure 49 and Figure 50 show the trajectories of the controlled and manipulated variables of the first control loop, respectively.

Table 46: Tuning parameters for strategy 1 of PCPP process.

	Controlled variable	Manipulated variable	K	τ_i
Loop 1	CO_2 capture	Lean solvent flowrate	1.5	0.1
Loop 2	Reboiler temperature	Steam flowrate	0.05	30

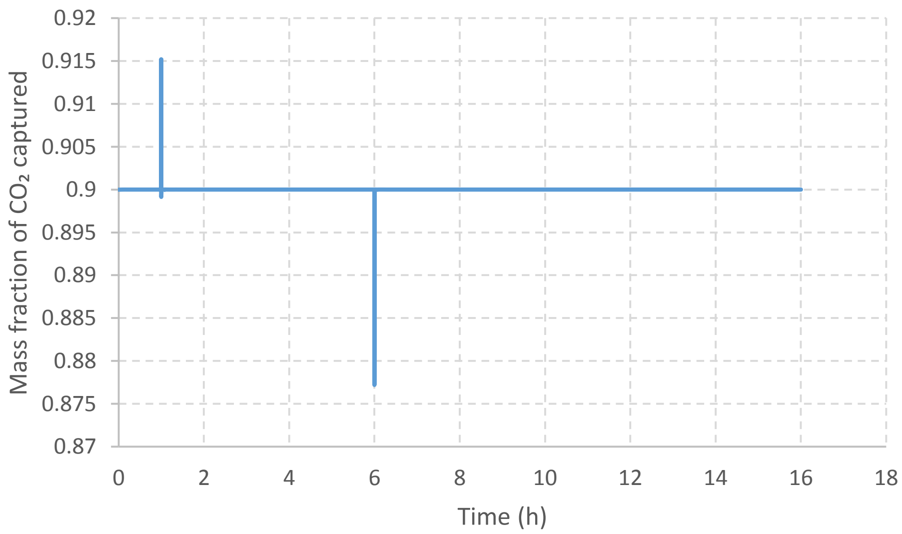


Figure 49: CO_2 capture (controlled variable) during tuning of control strategy 1 - PCPP plant. The set-point is 90% CO_2 capture.

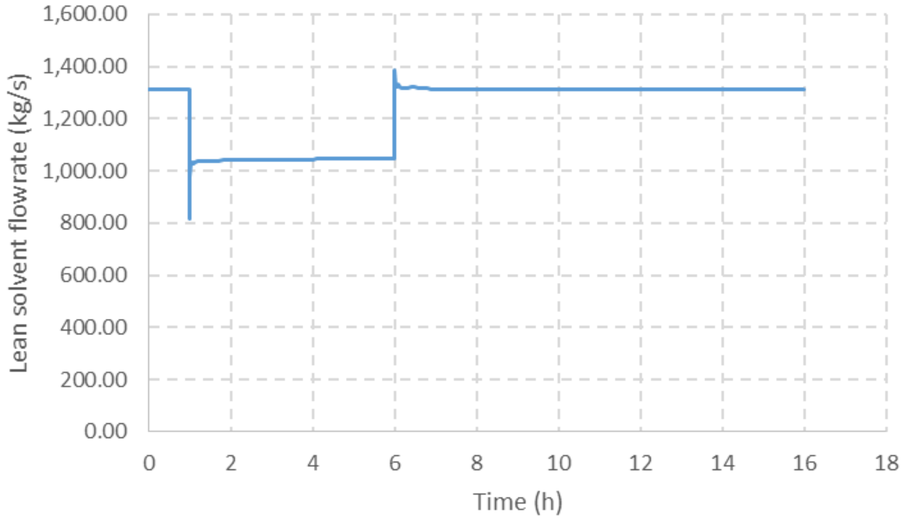


Figure 50: Lean solvent flowrate (manipulated variable) during tuning of control strategy 1 (PCPP plant).

Figure 51 and Figure 52 show that similarly good results were obtained for the second control loop.

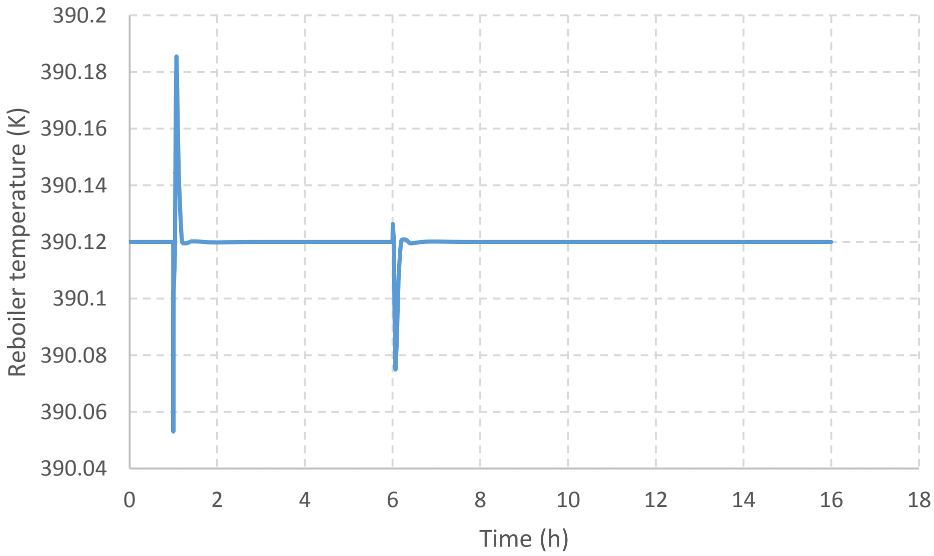


Figure 51: Reboiler temperature (controlled variable) during tuning of control strategy 1 - PCPP plant. The set-point is a reboiler temperature of 390.12 K.

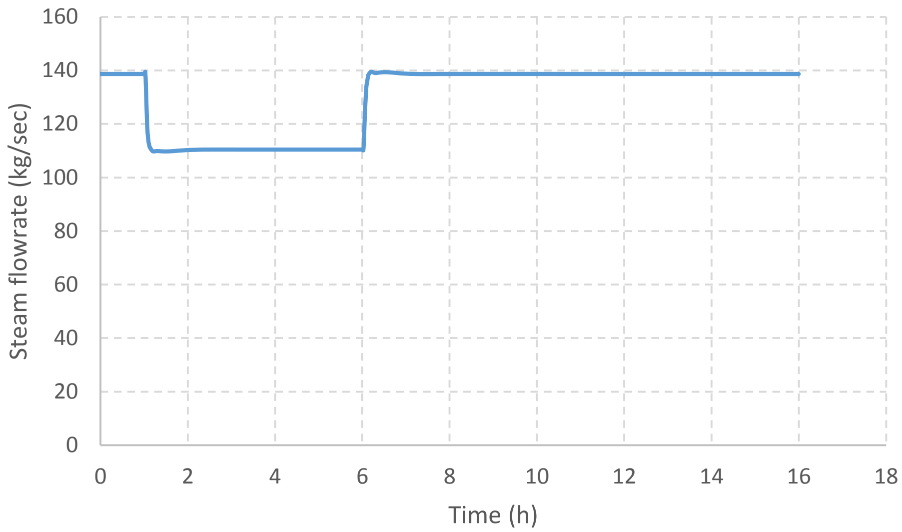


Figure 52: Steam flowrate (manipulated variable) during tuning of control strategy 1 - PCPP plant.

5.1.2 Strategy 2

For control strategy 2, it was found that, contrary to control strategy 1, the choice of tuning parameters had a significant impact on the controller performance. Therefore, in this case, the above mentioned methodology was followed, by testing different values of proportional gain and integral action until the best possible control performance was achieved. The simulation results for the first control loop (CO_2 capture as controlled variable, and steam flowrate as manipulated variable) are shown in Figure 53 and Figure 54.

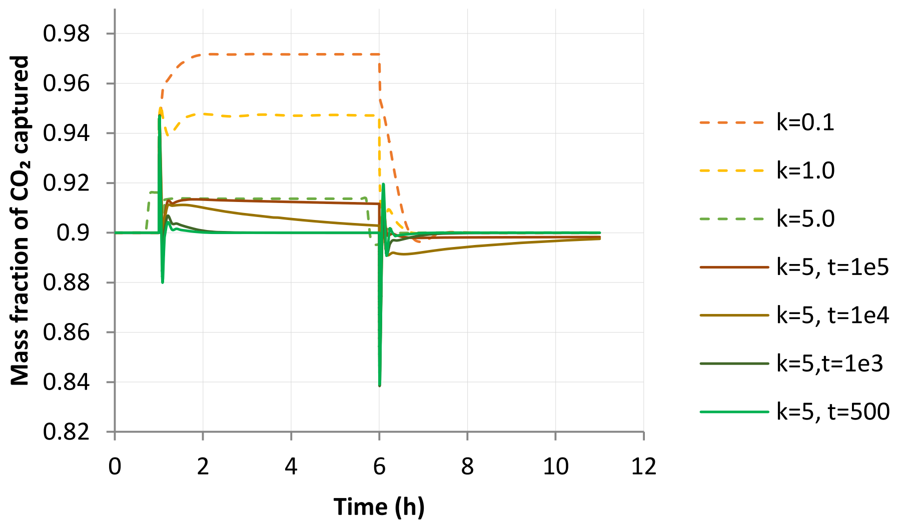


Figure 53: CO_2 capture (controlled variable) during tuning of control strategy 2 - PCPP plant.

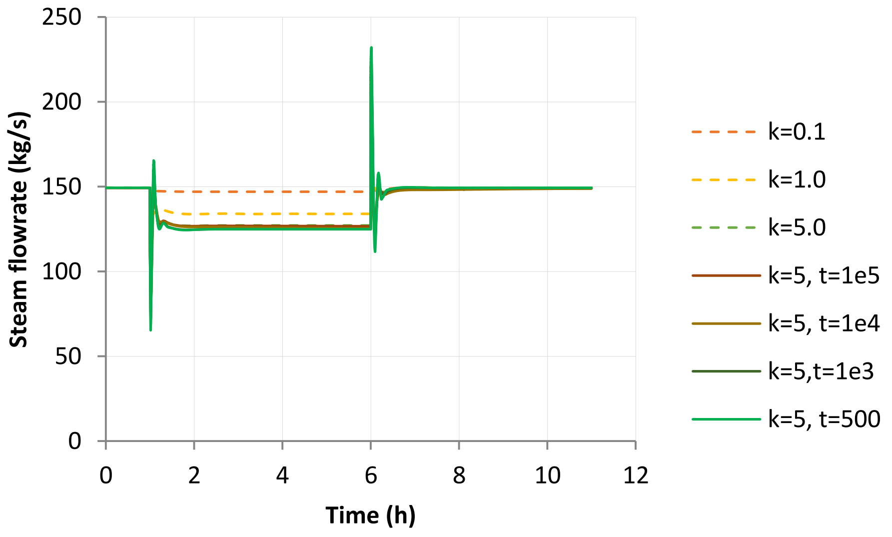


Figure 54: Steam flowrate (manipulated variable) during tuning of control strategy 2 - PCPP plant.

The figures above show how the off-set from set-point can be eliminated by progressively increasing the integral action. The second control loop in strategy 2 consists of simply maintaining the lean solvent flowrate at a controlled value. A good performance was achieved for a wide range of tuning parameters and, as shown in Figure 55, the response was not affected by the choice of tuning parameters in the first control loop.

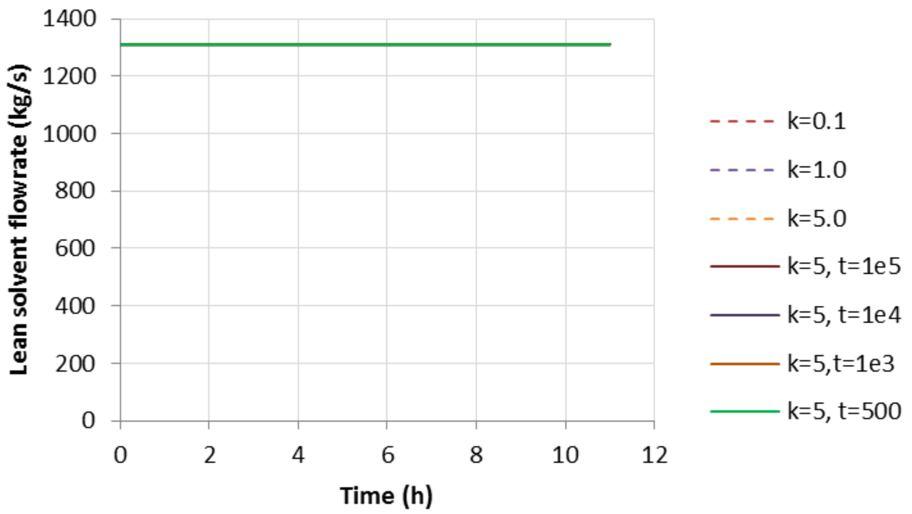


Figure 55: Lean solvent flowrate (manipulated variable) during tuning of control strategy 2 - PCPP plant. The values of tuning parameters refer to the first control loop.

In contrast to strategy 1, in this case the reboiler temperature is not part of the control loops. The corresponding variation is shown in Figure 56.

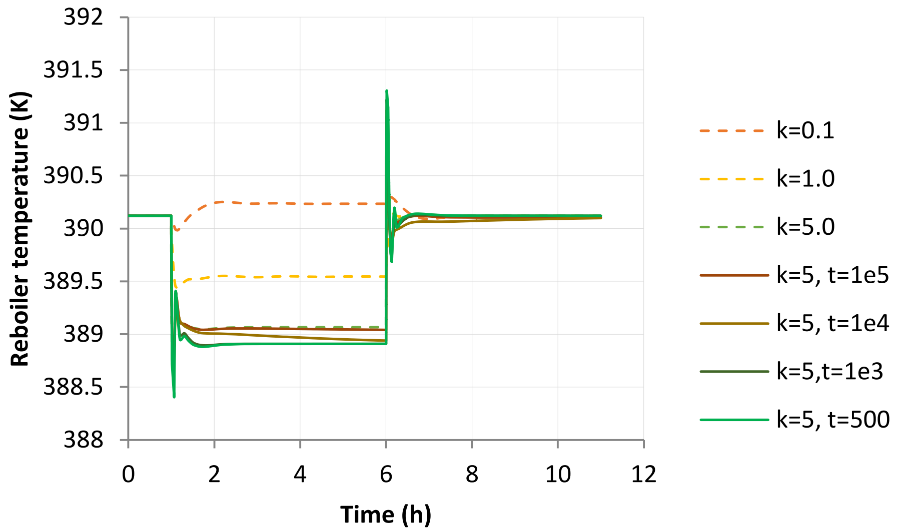


Figure 56: Reboiler temperature (controlled variable) during tuning of control strategy 1 - PCPP plant. The values of tuning parameters refer to the first control loop.

The tuning parameters chosen for this control strategy are presented in Table 47.

Table 47: Tuning parameters for strategy 2 of PCPP process.

	Controlled variable	Manipulated variable	K	τ_i
Loop 1	CO_2 capture	Reboiler temperature	5	500
Loop 2	Lean solvent flowrate	Lean solvent loading	1	10

5.1.3 Strategy 3

As mentioned above, for controller strategy 3, the parameters obtained for strategies 1 and 2 have been used.

5.2 Natural gas fired plant (CCGT)

The same methodology used for tuning the controllers presented in the previous section has been used to tune the control loops in the capture process of the natural gas fired power plant.

Figure 57 shows the disturbance in the flowrate of flue gas entering the process which was used to tune the performance of the controllers.

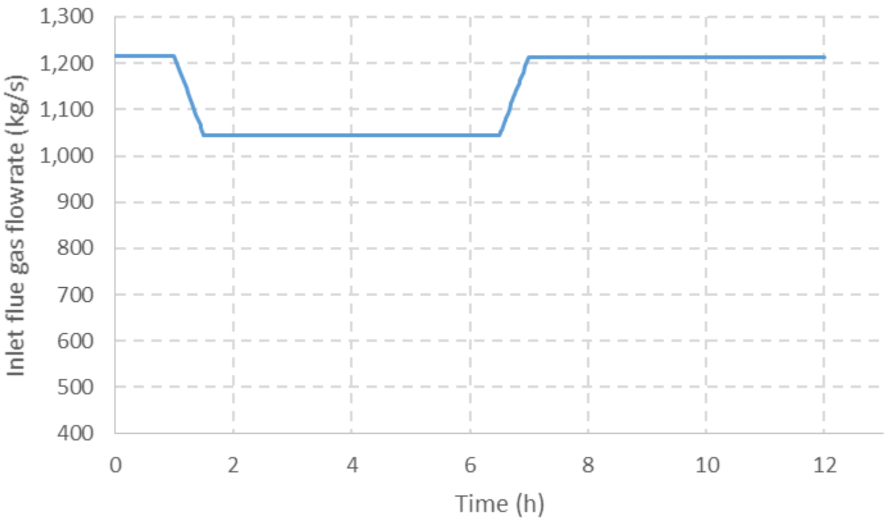


Figure 57: Step disturbances in flowrate of flue gas used for controller tuning in the CCGT process.

5.2.1 Strategy 1

As in the case of the PCPP process, the tuning of control strategy 1 was straightforward, as the tuning parameters did not affect the performance significantly. Good control performance has been achieved with the tuning parameters listed in Table 48.

Table 48: Tuning parameters for strategy 1 of CCGT process.

	Controlled variable	Manipulated variable	K	τ_i
Loop 1	CO_2 capture	Lean solvent flowrate	1	0.1
Loop 2	Reboiler temperature	Steam flowrate	1	10

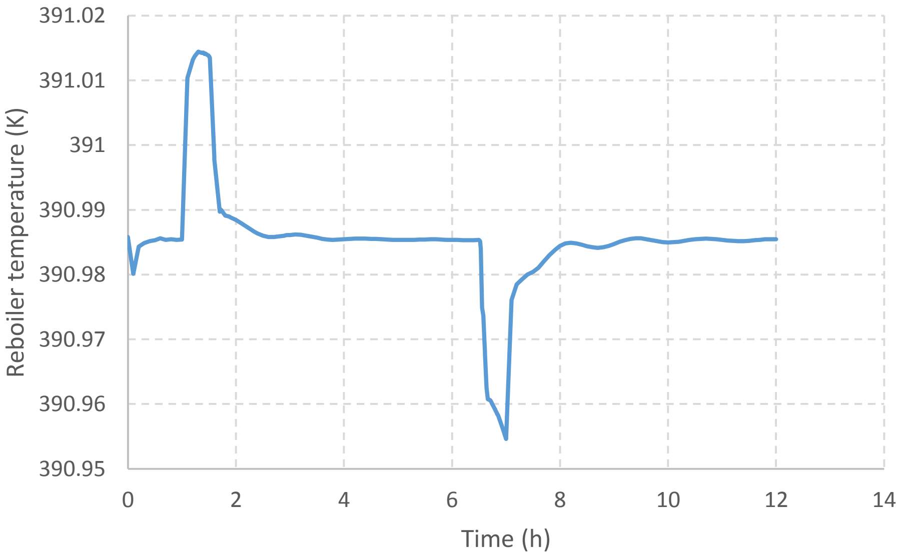


Figure 58: Reboiler temperature (controlled variable) during tuning of control strategy 1 - CCGT plant.

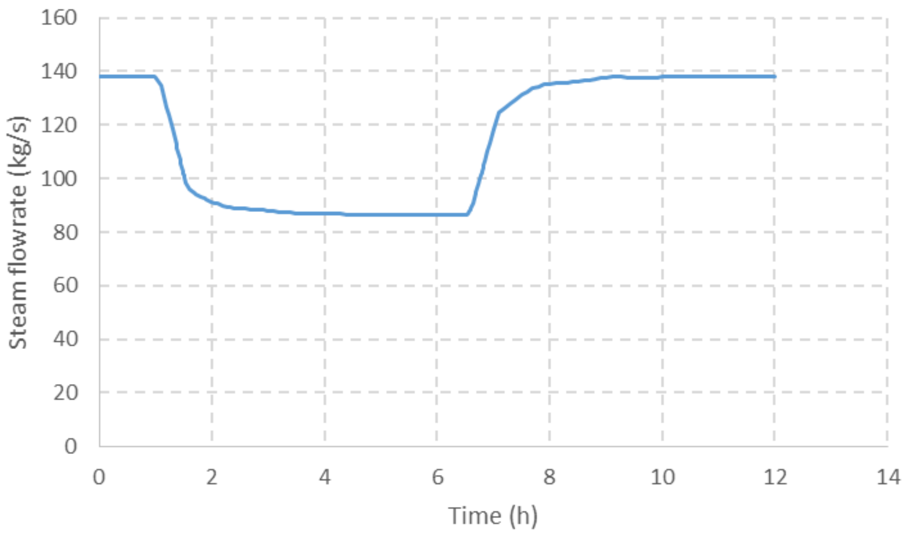


Figure 59: Steam flowrate (manipulated variable) during tuning of control strategy 1 - CCGT plant.

5.2.2 Strategy 2

As in the case of the PCPP process, the tuning of control strategy 2, the choice of the tuning parameters had a significant effect on the control performance. The trajectories of the controlled and manipulated variables for strategy 2 are shown in Figure 60 - Figure 62. It should be noted that, the upper limit on the steam flowrate has been reached by the controller. In comparison with control strategy 1, the results below show a more aggressive and oscillatory control response.

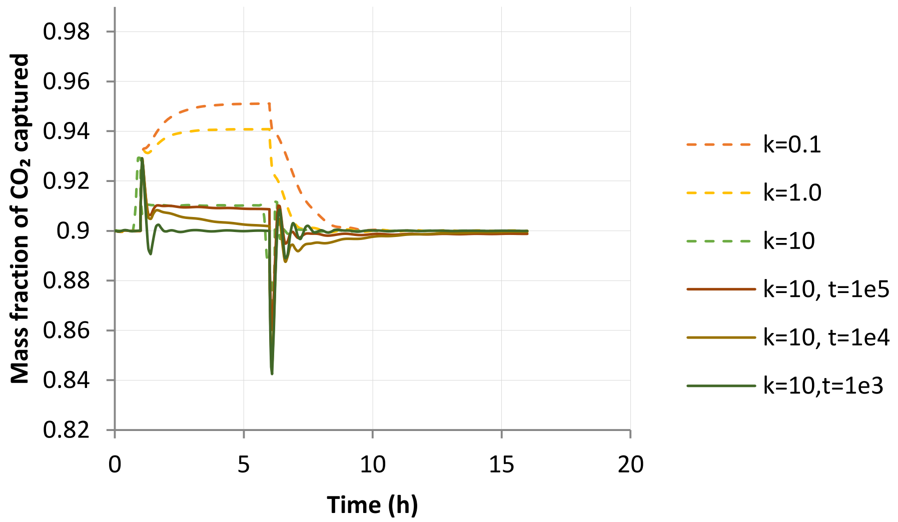


Figure 60: CO_2 capture (controlled variable) during tuning of control strategy 2 - CCGT plant.

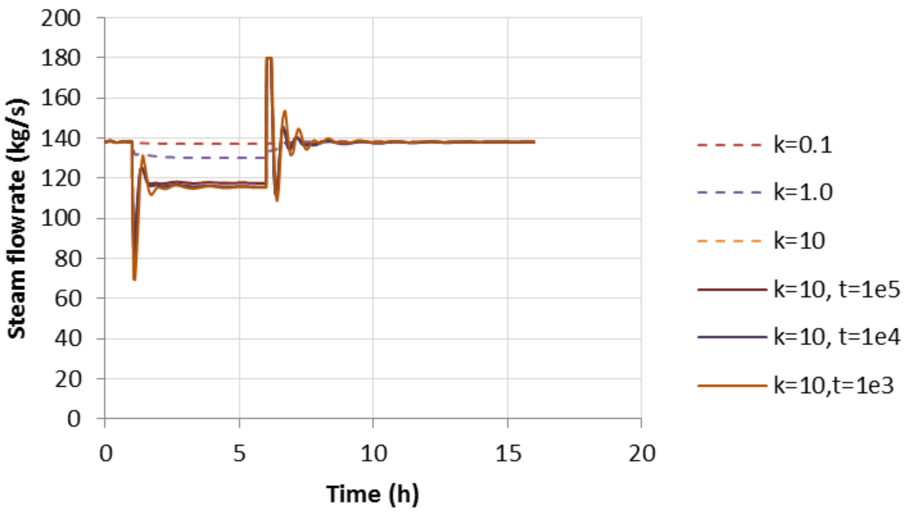


Figure 61: Steam flowrate (manipulated variable) during tuning of control strategy 2 - CCGT plant.

Figure 62 shows the trajectory of the reboiler temperature throughout the tuning of control strategy 2.

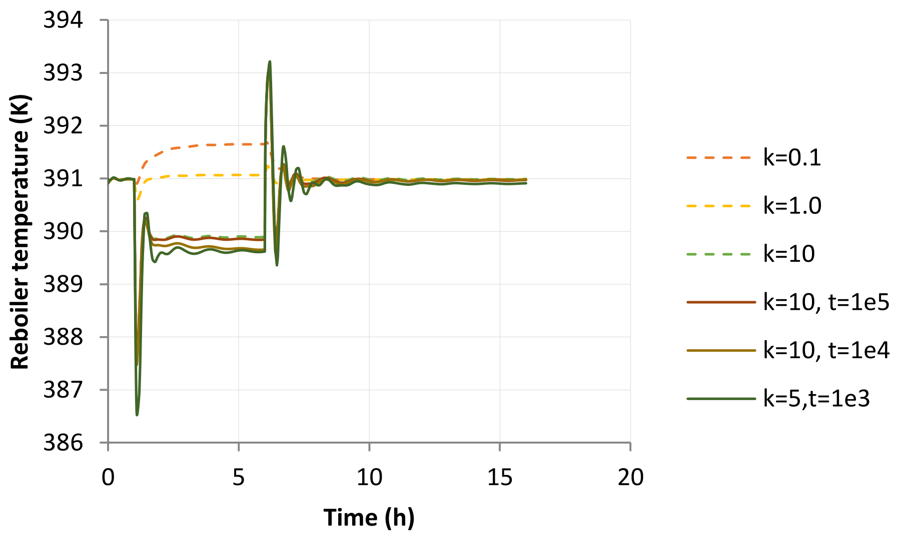


Figure 62: Reboiler temperature (controlled variable) during tuning of control strategy 2 - CCGT plant.

Based on the results shown above, the best tuning parameters for control strategy 2 are listed in Table 49.

Table 49: Tuning parameters for strategy 2 of CCGT process.

	Controlled variable	**Manipulated variable**	**K**	τ_i
Loop 1	CO_2 capture	Reboiler temperature	10	1000
Loop 2	Lean solvent flowrate	Lean solvent loading	0.001	0.5

5.2.3 Strategy 3

As mentioned above, for controller strategy 3, the parameters obtained for strategies 1 and 2 have been used.

6 Techno-economic evaluation of control strategies for flexible power plant operation

This section presents an economic evaluation of the three proposed control strategies for a scenario of flexible operation of the power plant. Power plants are required to operate flexibly, i.e. under a varying capacity factor throughout the day, to accommodate different electricity demands from the grid, and to compensate for the intermittent nature of renewable energy sources. Flexible operation also provides an opportunity to maximise the power plant profitability, by increasing the power output when electricity prices are at the highest, and decreasing during off-peak times.

6.1 Flexible power plant operation during 24 hour period

In order to compare the performance of the three control strategies proposed, a period of 24 hours was simulated for the pulverised coal and natural gas fired plants coupled with the post-combustion amine-based capture process. The first step towards the calculation of electricity cost is the calculation of the short run marginal costs (SRMC) for different fossil fuel plants, such as super-critical pulverised coal (SCPC), combined cycle gas turbines (CCGT) and open cycle gas turbines (OCGT) [12]. We envision that this kind of generator will be situated in the mid-merit/peak market with base load electricity generation provided by inflexible nuclear power. We consider that peak energy demand is met by a combination of fossil-fuels and renewable generators. We assume that fossil fuel derived electricity will be provided by super-critical pulverised coal (SCPC), combined cycle gas turbines (CCGT) and open cycle gas turbines (OCGT). Clearly, each of these generators have different short run marginal costs (SRMC). Data from the UK's Department of Energy and Climate Change (DECC) were used to specify fossil fuel and carbon prices. Similarly, data from the Joint Research Centre of the European Commission was used to specify a range of probable generator efficiencies and carbon intensities. Using these data, we calculate possible short run marginal cost (SRMC) prices for SCPC, CCGT and OCGT in the 2030s using the following equation:

$$\frac{£^{SRMC}}{MWhr} = \frac{£^{MWhr}_{Fuel}}{n_{Plant}} + \left(£^{CO_2}_{Tonne} \cdot CI^{Tonnes\ CO_2}_{MWhr}\right) + £_{VarO\&M} + £^{CO_2}_{T\&S} \qquad (8)$$

where $£^{\mathrm{SRMC}}$ is the SRMC of the electricity generated by a given plant which includes the variable operating and maintenance costs, $£_{\mathrm{VarO\&M}}$ and also the fixed cost of CO_2 transport and storage $£^{\mathrm{CO_2}}_{\mathrm{T\&S}}$. These values are provided in Mac Dowell et al. [8].

The prices for the different types of plants were £55.31/MWh, £65.79/MWh and £100.18 /MWh, respectively. On this basis we assume that the over-night (off-peak) electricity process are set by the SCPC plants, the day-time prices by the CCGT with morning and evening peaks serviced by the OGCT plants. The capacity factors are representative of the behaviour of a load following plant in the UK in 2012. The variation of electricity prices considered in this study is shown in Figure 63. The figure also shows a typical variation of the power plant capacity factor.

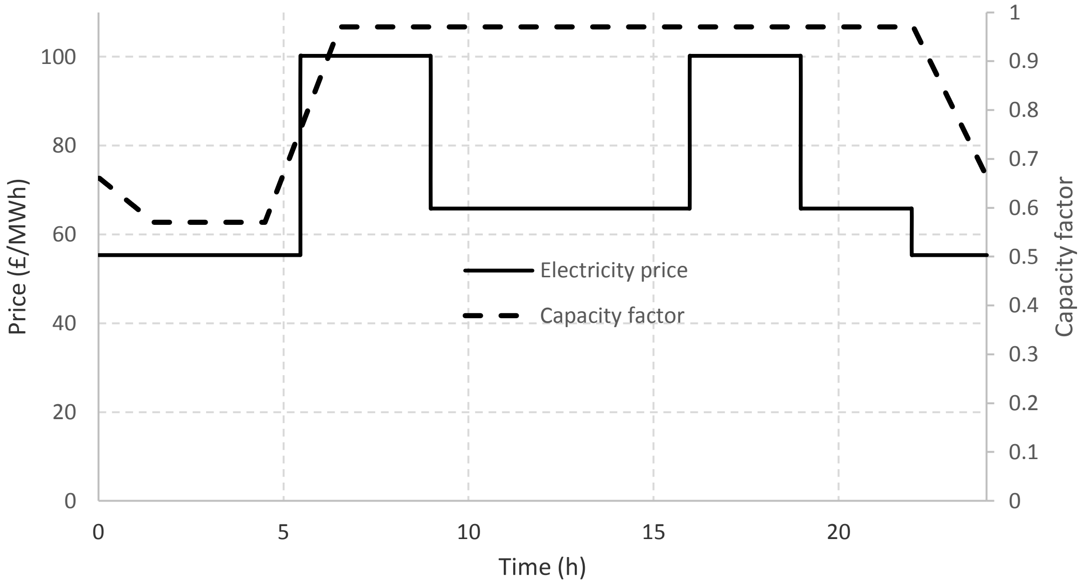


Figure 63: Variation of electricity prices throughout the day, and typical operation profile for two-shifting power plant.

The variation in electricity prices, as well as the cost of utility water and the assumed value of CO_2 are shown in Table 50 using data from [7, 12].

Table 50: Electricity prices, assumed CO_2 cost, and utility water cost used in this study, data taken from [7, 12].

Electricity prices	
05:00 – 09:00	£100.18 / MWh
09:00 – 16:00	£65.79 / MWh
16:00 – 19:00	£100.18 / MWh
19:00 – 05:00	£55.31 / MWh
CO_2 value	£70 / ton
Utility water cost	£0.025 / ton

6.2 PCPP

The behaviour of individual control variables for the PCPP case are detailed in the following section.

6.2.1 Trajectory of key variables

In Figure 64, the exhaust gas flowrate is observed to vary as the power plant is ramped up and down over the course of the simulation. The exhaust gas composition varied from 19-21 wt%, while the degree of CO_2 capture was observed to remain approximately constant throughout the simulation.

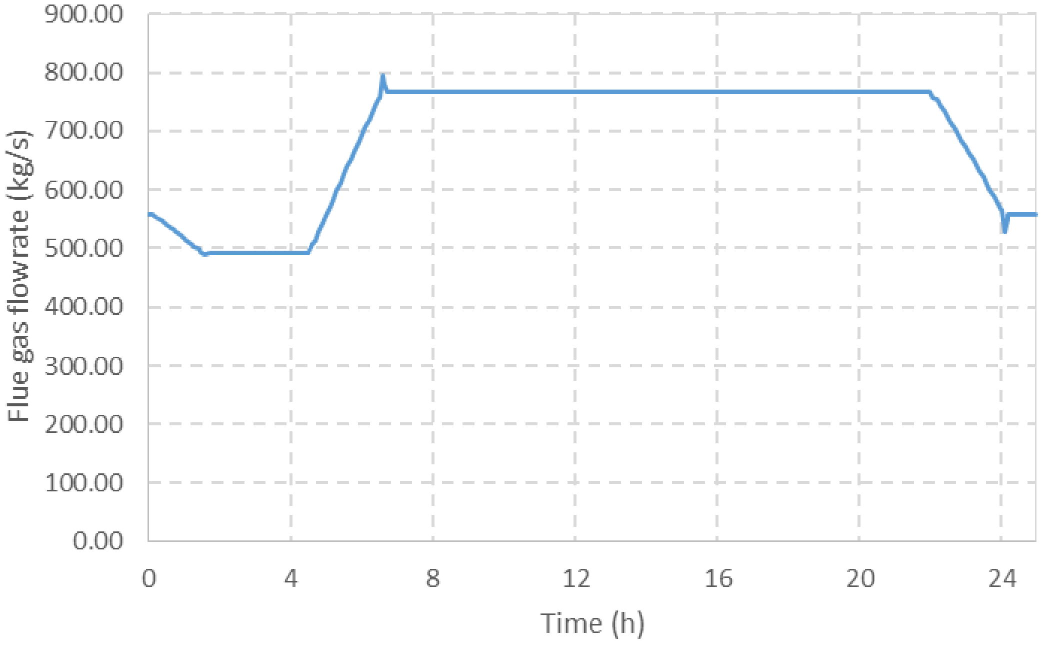


Figure 64: Flue gas flowrate variation during 24 hour operation in the PCPP process.

In Strategy 2, the manipulated variable is the reboiler temperature, with the aim of using the lean loading to control the degree of CO_2 capture. As may be observed in Figure 65 during the period from t=6 to t=14 there is a small but persistent fluctuation in the degree of CO_2 capture, indicating the sensitivity of the process to this variable. The reverse is observed for strategy 3 – the amplitude of the departure from the set point is relatively large, but short-lived.

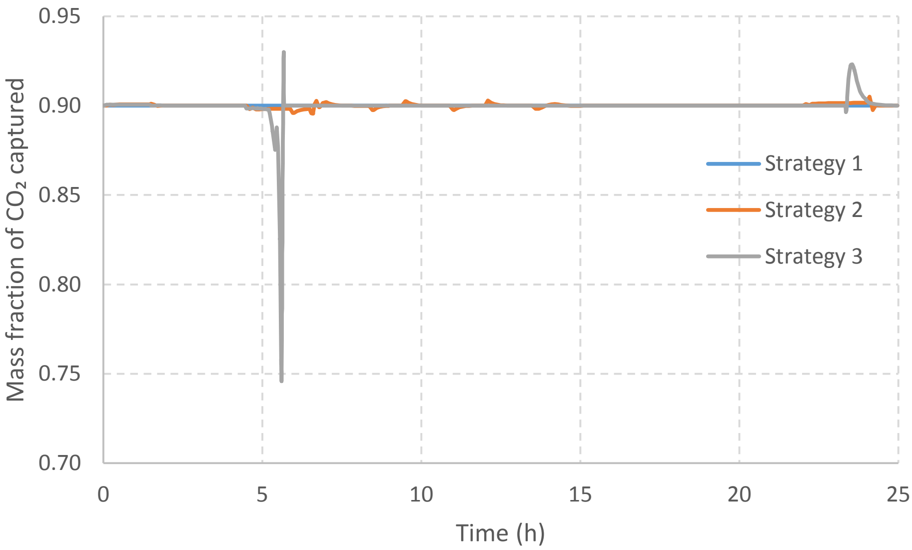


Figure 65: CO_2 capture using different control strategies for the PCPP process.

The reason for the nature of the behaviour exhibited by Strategy 2 is evident from Figure 66.

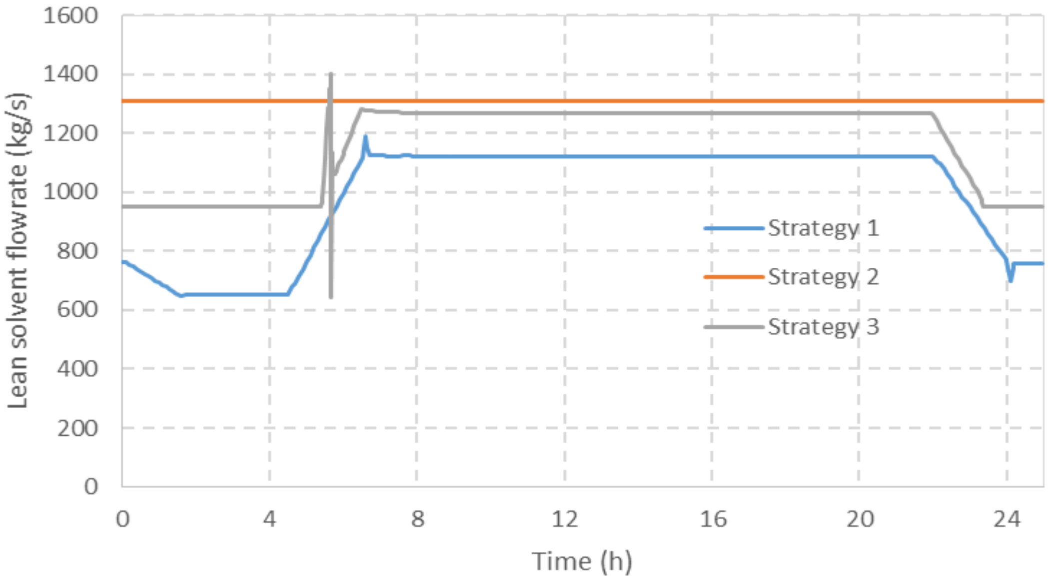


Figure 66: Lean solvent flowrate using different control strategies for the PCPP process.

The redistribution of the solvent over the column internals gives rise to a short-term fluctuation in the degree of CO_2 capture. Some further insight into the long term oscillatory behaviour exhibited by Strategy 2 is provided by Figure 67 and Figure 68.

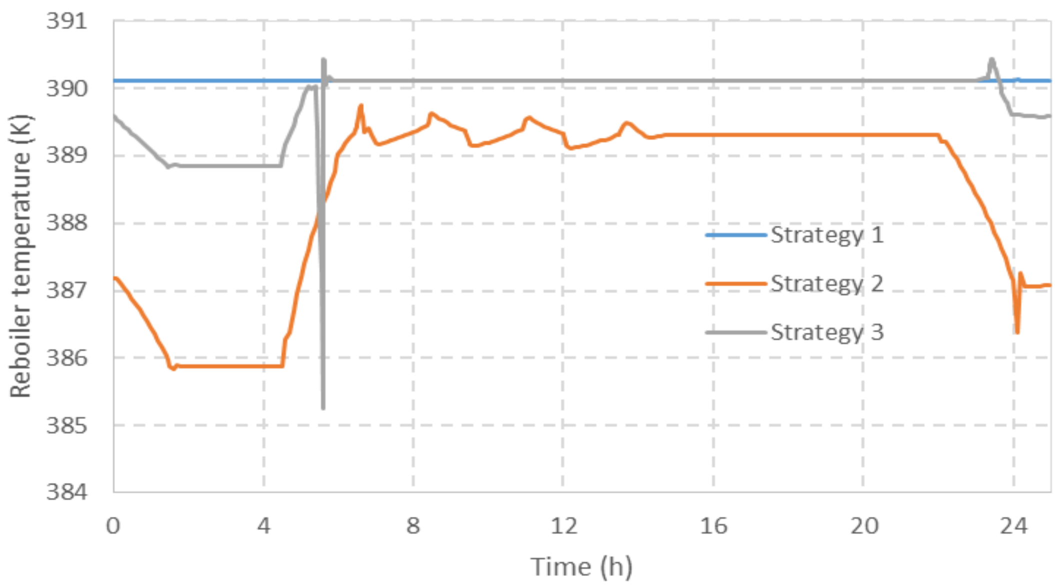


Figure 67: Reboiler temperature using different control strategies for the PCPP process.

The tight coupling between solvent temperature and lean loading is illustrated from these results and the high sensitivity of the process model to this control variable is evident here.

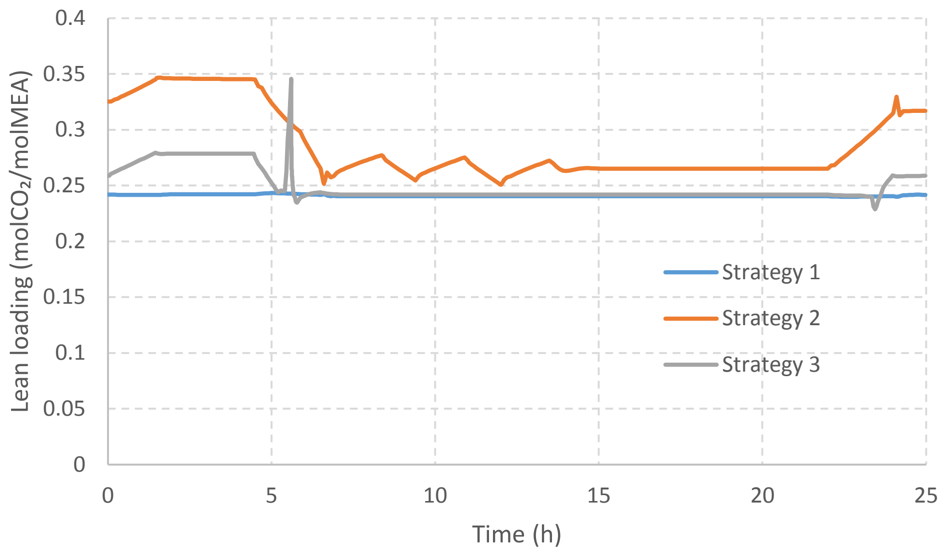


Figure 68: Lean loading at reboiler using different control strategies for the PCPP process.

Arguably, owing to the sensitivity of the process to the solvent lean loading, using this variable as a means of controlling the plant is not a good idea.

6.2.2 Economic evaluation

A simple approximation to the profitability of the plant was evaluated. Here, profit was calculated considering revenue generated by selling electricity and the costs associated with emitting CO_2, burning fuel and cooling water utilities. No fixed costs are considered, and thus this is an approximation to the short-run marginal cost profit of the plant. In Figure 69, we present the total cost for the three different strategies. The different lines (low, central, high) correspond to different coal prices (£51/t, £62/t and £79/t), respectively, which correspond to the DECC prices for 2030.

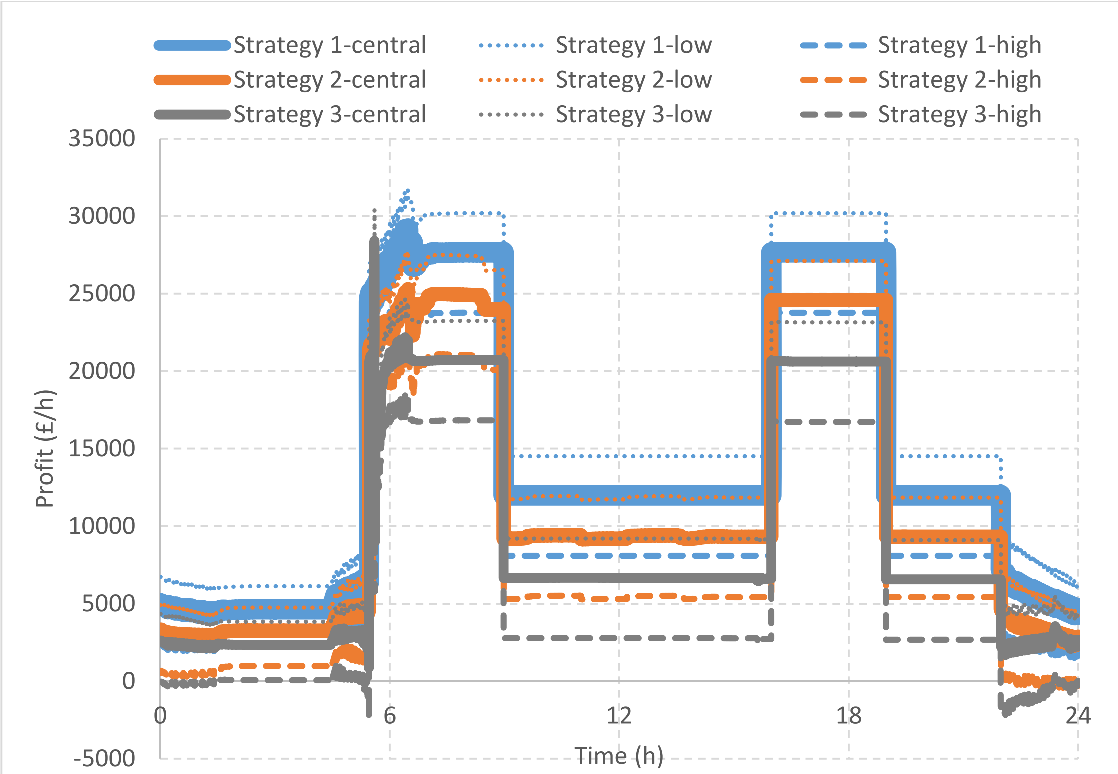


Figure 69: Profit for PCPP plant with different control strategies.

As can be observed from Figure 69, for the central scenario, Strategy 1 is more profitable than Strategy 2 due to the higher electricity revenue. Strategy 3 appears to be the least economic option due to the lower electricity revenue. This can be explained by the fact that the steam requirement for Strategy 3 appears to be higher compared to the other two strategies. The reason for this is that switching from one strategy to another can have some delay which in turn affects the control variables, in our case the reboiler temperature which leads to higher steam requirement to keep the control variables in certain set points. More detail is provided in Figure 70, wherein the breakdown of costs for the three strategies is presented.

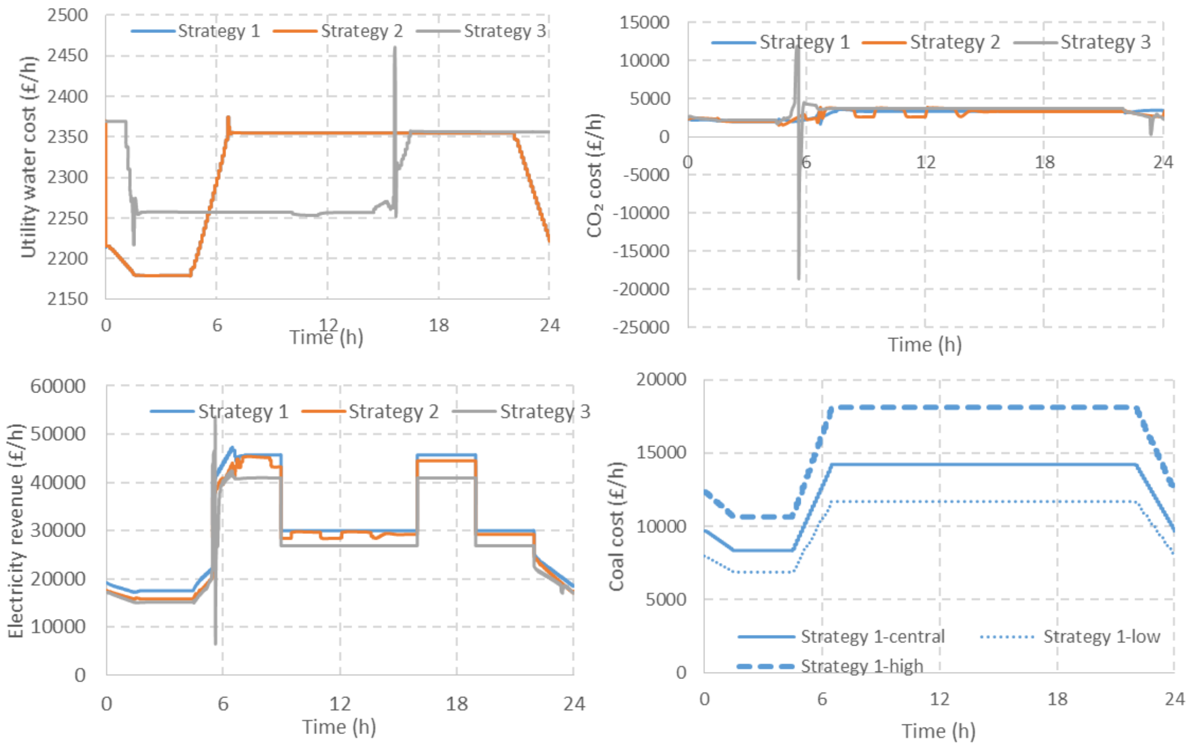


Figure 70: Cost breakdown for PCPP plant with different control strategies.

In Figure 71, we present the cumulative profit attained by the plant for each mode of operation and in Table 51 we present the total profit for the PCPP for the three control strategies. The cumulative profit obtained by following Strategy 1 resulted in an increase in profits of approximately 18% compared to Strategy 2 with a total profit of £336k. Where the electricity revenue is higher for Strategy 1. The power plant is ramping up and down following the same pattern for the three control strategies, so the coal flowrate and therefore the coal cost remains the same, as illustrated in Figure 75.

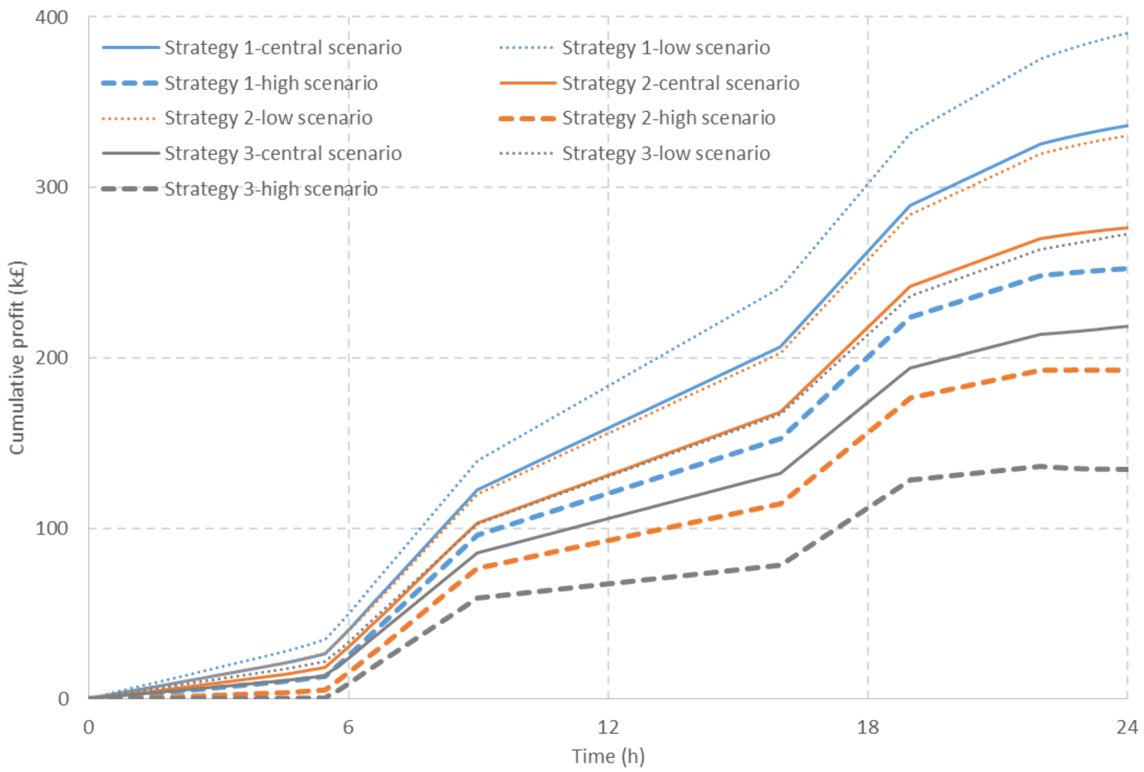


Figure 71: Cumulative profit for PCPP plant with different control strategies.

As can be observed from Figure 71 and Table 51, the fuel cost exerts a significant impact on the overall profitability. The general ranking and conclusion is preserved, however. There is one aspect of this figure that is non-intuitive. Given that Strategy 3 is intermediate to Strategies 1 and 2, one would expect the profitability of Strategy 3 to also be intermediate. However, owing to the relatively long duration required to switch between Strategy 1 and 2, there was an undue accumulation of solvent in the system, resulting in increased steam demand, leading to a reduction in profitability.

Table 51: Total profit for each control strategy in the PCPP process.

	Total profit
Strategy 1	£336k
Strategy 2	£276k
Strategy 3	£219k

6.3 CCGT

6.3.1 Trajectory of key variables

The behaviour of the CCGT was broadly similar to that of the PCCP, and consequently this analysis is not repeated here. However, there is one point of distinction; whereas in the case of the PCCP, the economic evaluation showed a better performance of Strategy 1 as illustrated in Table 51, in the case of CCGT the three strategies had an equivalent. This is due to the more dilute CO_2 concentration within the capture process, and therefore the absolute amount of CO_2 emitted in the absorber outlet gas was less sensitive to fluctuations in solvent distribution within the absorption column. As a consequence, regardless of the control strategy employed, barely any deviation from the 90% CO_2 capture set point was observed, as illustrated in Figure 72.

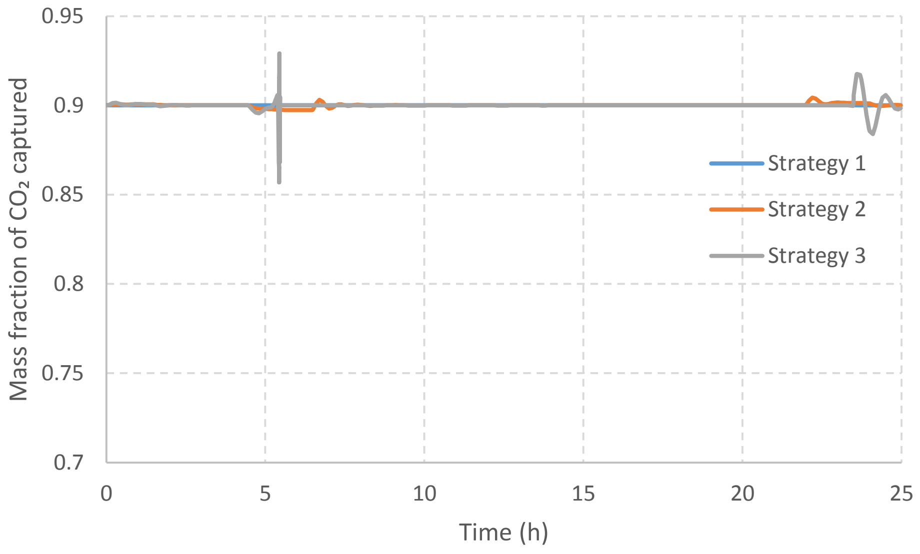


Figure 72: CO_2 capture using different control strategies for the CCGT process.

The distinction in amplitude and relaxation time between the various control strategies is still evident here, however, as illustrated in Figure 73.

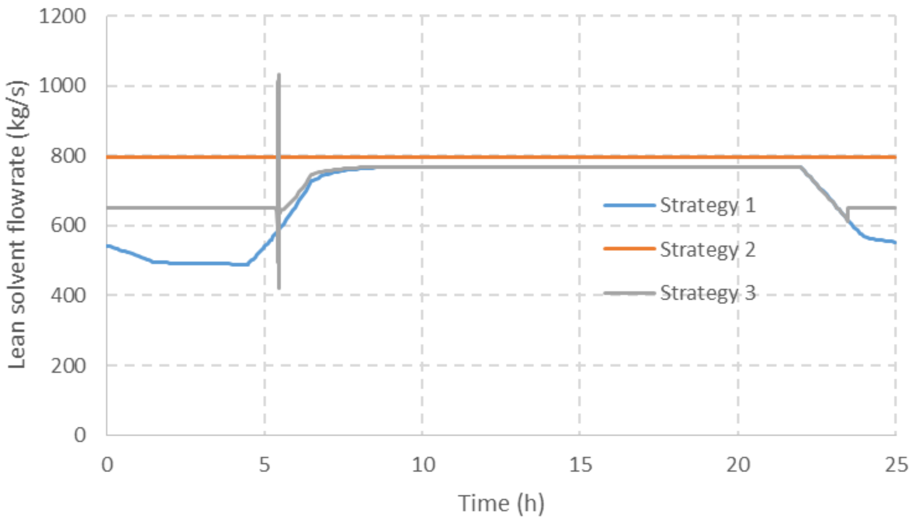


Figure 73: Lean solvent flowrate using different control strategies for the CCGT process.

The following figures illustrate the fluctuation of the different process variables during the three different control strategies.

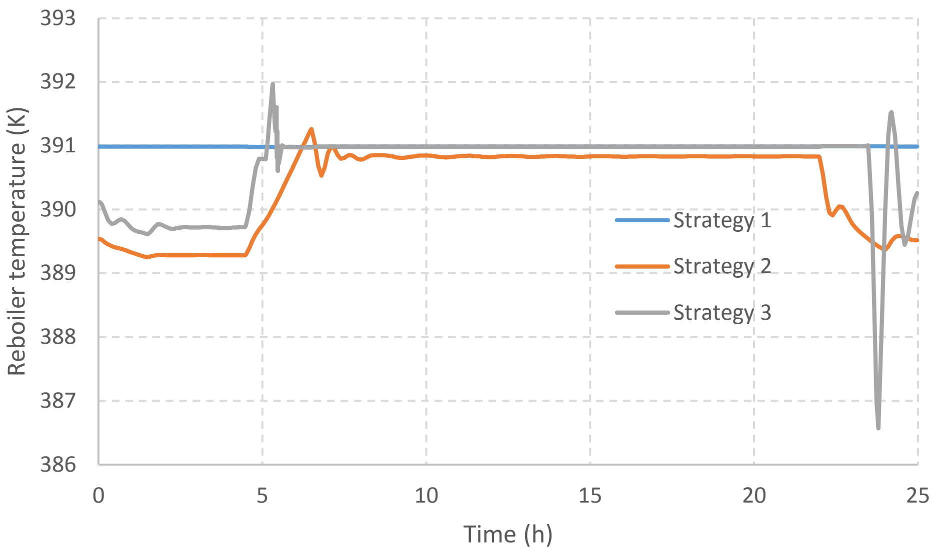


Figure 74: Reboiler temperature using different control strategies for the CCGT process.

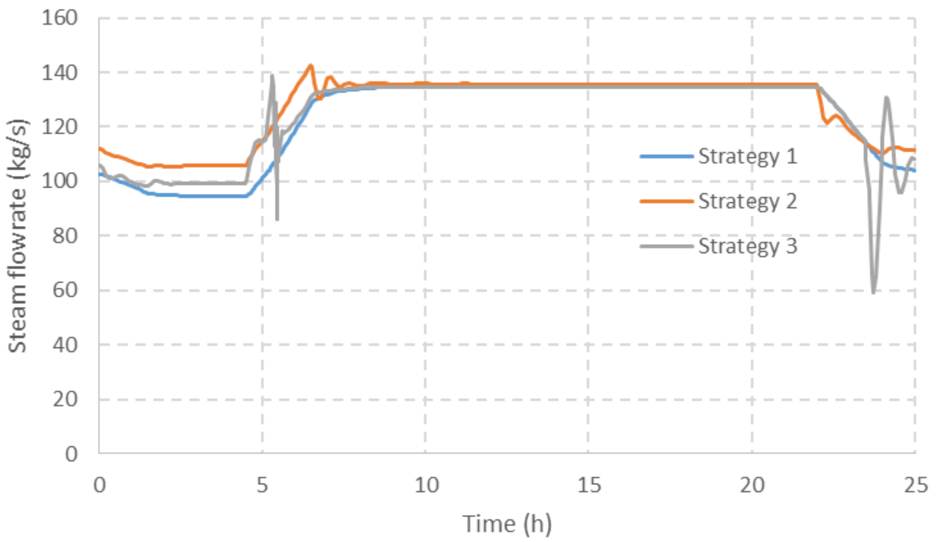


Figure 75: Steam flowrate using different control strategies for the CCGT process.

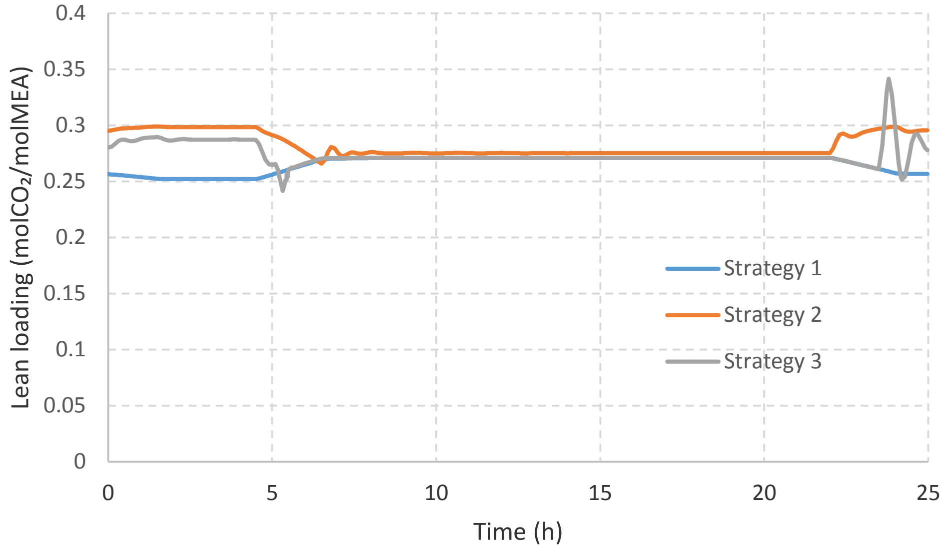


Figure 76: Lean loading at reboiler using different control strategies for the CCGT process.

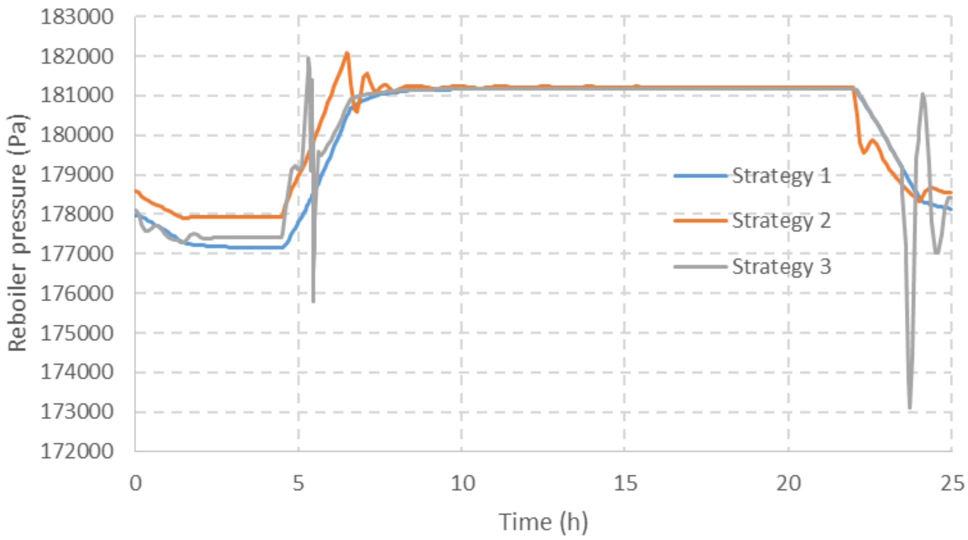


Figure 77: Reboiler absolute pressure using different control strategies for the CCGT process

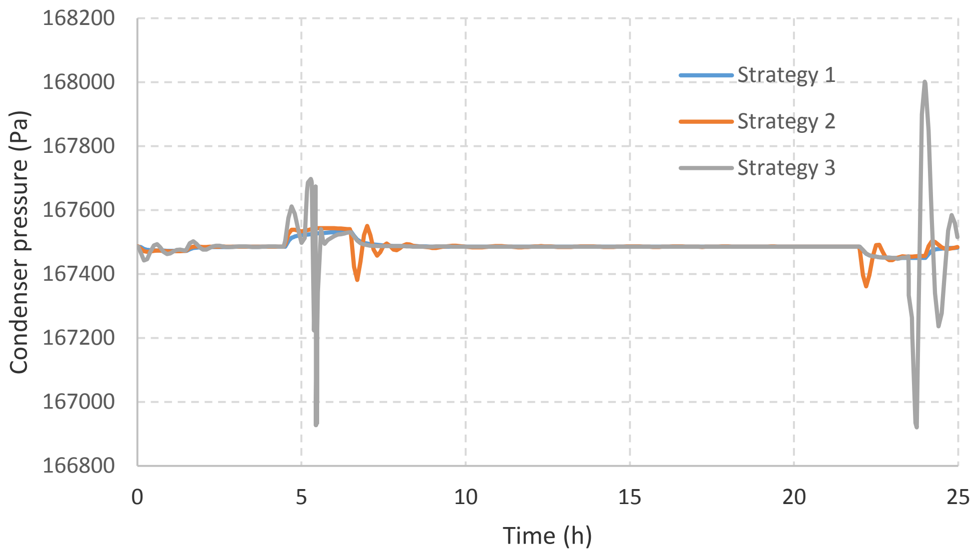


Figure 78: Condenser absolute pressure using different control strategies for the CCGT process.

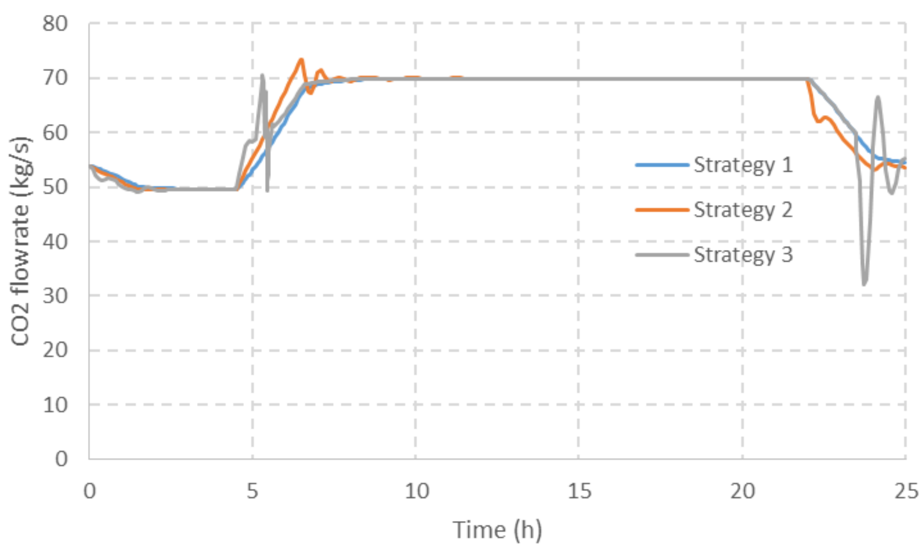


Figure 79: Condenser absolute pressure using different control strategies for the CCGT process.

6.3.2 Economic evaluation

In this section we present the economic evaluation for the CCGT process. Noting that the key distinguishing feature between the different control strategies was how much CO_2 was captured and given that the power plant ramping behaviour was identical for all scenarios, it follows that the profitability of the CCGT was approximately constant, regardless of the control strategy employed, as all three of them provided essentially steady state behaviour from the perspective of the degree of CO_2 capture. This is evident from Figure 80.

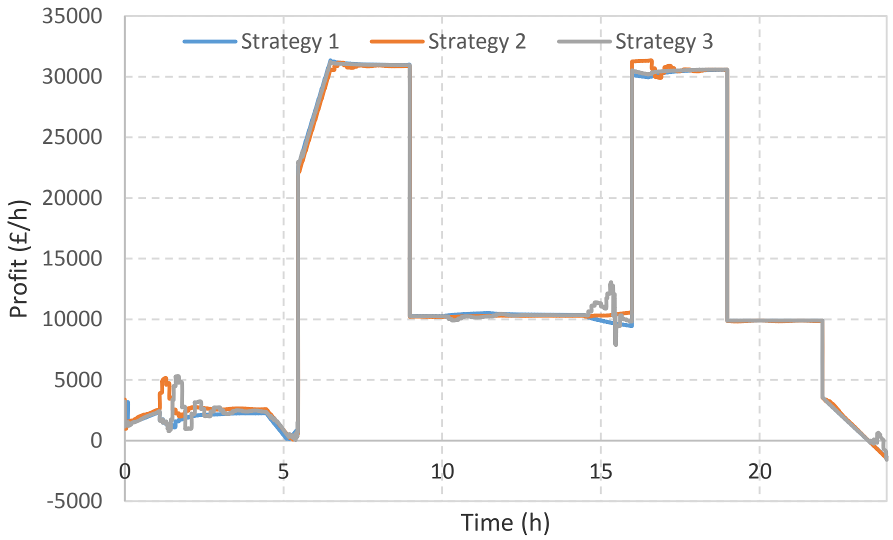


Figure 80: Profit for CCGT plant with different control strategies.

In Figure 81, we present the cumulative profit for the three different control strategies. The total profit for the three strategies is approximately £300k for one day of operation of the plant.

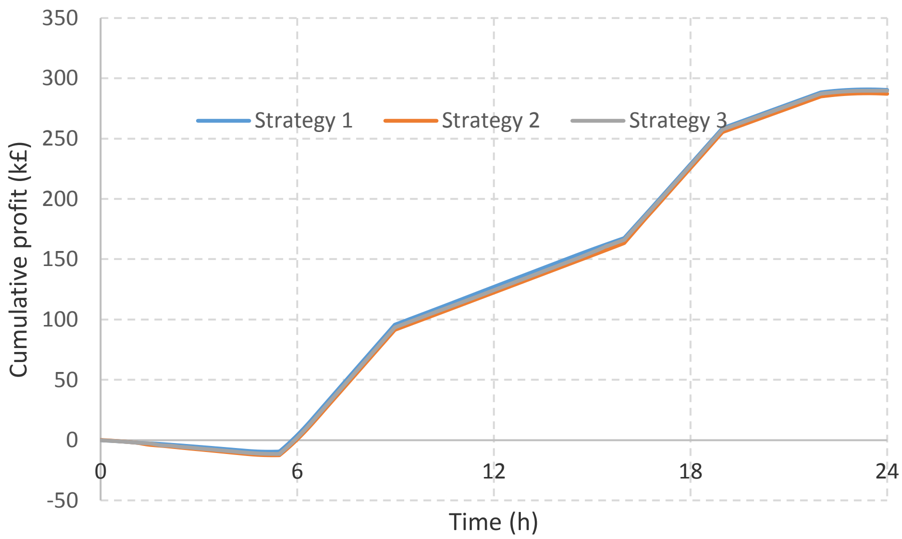


Figure 81: Cumulative profit for CCGT **plant with different control strategies.**

The total profit for the period of 24 hours is presented in

Table 52.

Table 52: Total profit for each control strategy in the CCGT process.

Total profit	
Strategy 1	£290k
Strategy 2	£287k
Strategy 3	£289k

A key contributing factor to the profitability in these scenarios was the cost attributed to CO_2 emissions in this scenario (£70/t_{CO2}). Owing to the fact that the exhaust gas stream arising from the gas-CCS plant is significantly more dilute than the coal plant, the variation in profitability with control strategy is near zero in the case of the CCGT (barely any CO_2 is emitted) whereas this is not the case in the coal plant. Similarly, the variation in reboiler temperature in the case of the PCCP is much greater than in the case of the CCGT, which also affects operating costs, and thus profitability.

7 Hazard management in CCS chains

A typical CCS project will involve multiple operator companies, e.g. the power station, CO_2 transport pipeline and offshore CO_2 injection operators. Complexity is likely to increase as the incorporation of subsequent CCS projects on the backbone of an initial project will lead to the creation of CO_2 transport networks with multiple CO_2 sources and storage options.

The safe operation of the CCS system will require close co-ordination of the different stakeholders. This section aims to identify and analyse potential hazards by considering potential trip, outage and shutdown scenarios and describing the effects on stakeholders up and down the CCS chain. It will seek to demonstrate how the process can be safely managed in the event of a system failure and suggest a communication procedure based on the best practices currently available across industry.

7.1 Modelling overview

7.1.1 Summary of scenarios

The scenarios are designed to cover a range of different consequences from a small number of events, and have been designed with input from the Energy Institute (EI), Health and Safety Laboratories (HSL) and the Carbon capture and Storage Association (CCSA).

The categories used in the definitions below have the following meanings:

7.1.1.1 Anticipated impact

'Anticipated impact' is an attempt to understand the scope of what needs to be explored and hence which models to include a priori, rather than an attempt to prejudice the outcome of the analysis. The anticipated impact reflects what would happen in an 'open loop' scenario. In reality many of the outcomes would be pre-empted by operator intervention or pre-defined operating policy; it is part of the purpose of this study to identify where the latter would be of value.

7.1.1.2 Included gCCS flowsheet components

This describes the CCS elements that will be simulated in gCCS, including the upstream and downstream boundaries.

7.1.1.3 Scenarios and implementation in gCCS

This describes the scenarios that will be executed and the action that will be taken in gCCS to implement the each scenario.

7.1.1.4 Key analysis

This will list the key items that will be explored in the scenario.

7.1.2 Model configuration

We have considered a whole-chain system that is disposing of CO_2 generated by a conventional power plant (pulverised-coal or CCGT) with an integrated amine-based post-combustion CO_2 capture plant, compression station, downstream transmission line and injection to a depleted gas reservoir. It assumes dense-phase transmission.

7.1.3 Modelling assumptions

The following assumptions and limitations will apply:

- On loss of the capture plant capacity, CO_2 will be vented upstream of the capture plant.
- 'Shutdown' of a component will be modelled by reducing the flowrate through the component to a small, non-zero value
- There will be no control schemes on the intermediate CO_2 storage capacity; this will be modelled simply as a volume
- No attempt will be made to anticipate control responses; for example, in the case of a compressor trip it might be possible to maintain pressure in the pipeline by closing the well choke valve; however the scenario analysis will model the 'open loop' responses
- Except where explicitly stated, power and capture plant will not be integrated

7.1.4 Key results

Analysis of the scenarios will include, but not be restricted to, identifying and quantifying the following, as appropriate:

- Key trends – for example, rising pressures
- Duration of key disturbances or time to key events – for example, to maximum pipeline pressure; compressor trip; capture plant shutdown
- Phase change in transmission lines
- The effectiveness of some proposed passive control strategies such as buffer storage

7.1.5 Model configuration

The configuration of the system model, in addition to those described in previous sections of this report are outlined in Table 53 and Figure 82 below.

Table 53: Model configuration for hazard identification study

Description	
Power plant output (without capture) (MW)	819
CO_2 capture plant process technology	MEA, 30wt% solution
Onshore pipeline length (km) /nominal diameter (m)	20 / 0.81
Offshore pipeline length (km) /nominal diameter (m)	200 / 0.81
Injection well tubing depth (km) /nominal diameter (m)	2 / 0.18
Maximum allowable pipeline pressure (bar)	153
Number of injection wells	4
Initial reservoir pressure (bar_a)	252
Nominal CO_2 flowrate (kg/s) / (ton/day)	138 / ~12000
Ambient temperature (air / water) (°C)	15 / 9

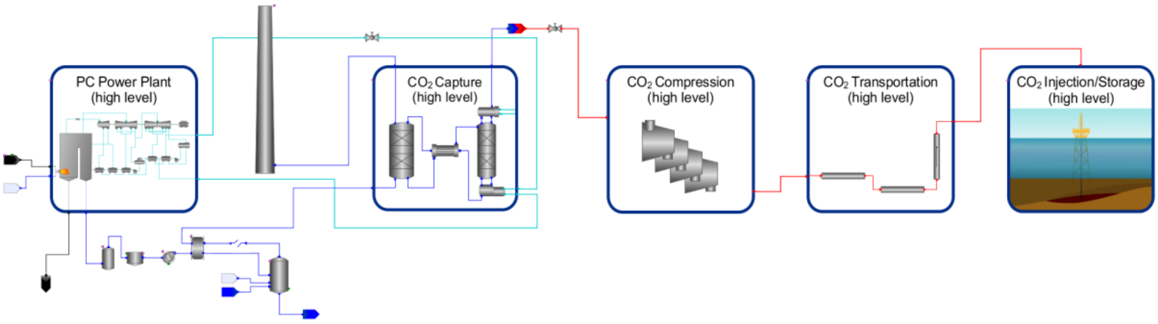


Figure 82: Full chain CCS model, including power plant, capture plant, compression train, pipeline, injection and storage

7.2 Scenario 1: Unplanned shutdown at injection site / loss of storage in a single chain

This situation might arise from the following scenarios:

- Where an instrumentation error, control system error or similar eventuality causes the well choke valve to fail shut
- The operator shuts the choke valve intentionally because, for example, continued operation would result in damaging the geological formation

This scenario could also represent a pipeline blockage – for example, where fluctuations in operating conditions have created conditions for formation of hydrates.

7.2.1 Anticipated impact

The failure of downstream injection activity will cause the pipeline pressure to rise. The upstream compressors will continue to operate against an increasing discharge pressure, with recycles opening where necessary, until (a) the pipeline pressure or (b) the capture plant operating pressure (represented by the compressor suction pressure) is sufficient to cause a trip.

7.2.2 gCCS flowsheet components

All components downstream from the capture process outlet (i.e. compressor suction drum) as far as the choke valve at the injection point.

7.2.3 Scenarios and implementation in gCCS

Scenario 1.1 Trip of base-case chain

Scenario 1.2 Trip of chain with intermediate buffer storage of specified volume added

The scenarios will be implemented by closing the choke valve at the inlet to the storage reservoir. The case will run until the safe operating pressure for the capture plant (as represented by the compressor suction pressure) is exceeded, at which point the scenario will terminate. The process flow diagram in gCCS used for this analysis is illustrated in Figure 83, with each of the constituent components comprising each of sub-units are displayed therein. The CO_2 Source represents the CO_2 flow from the compression system, the flow is 138.4 kg/s. The CO_2 is assumed to be in the supercritical phase and its conditions are assumed to be $\boldsymbol{T}$ = 50°C and $\boldsymbol{P}$ = 108.1 bar_a.

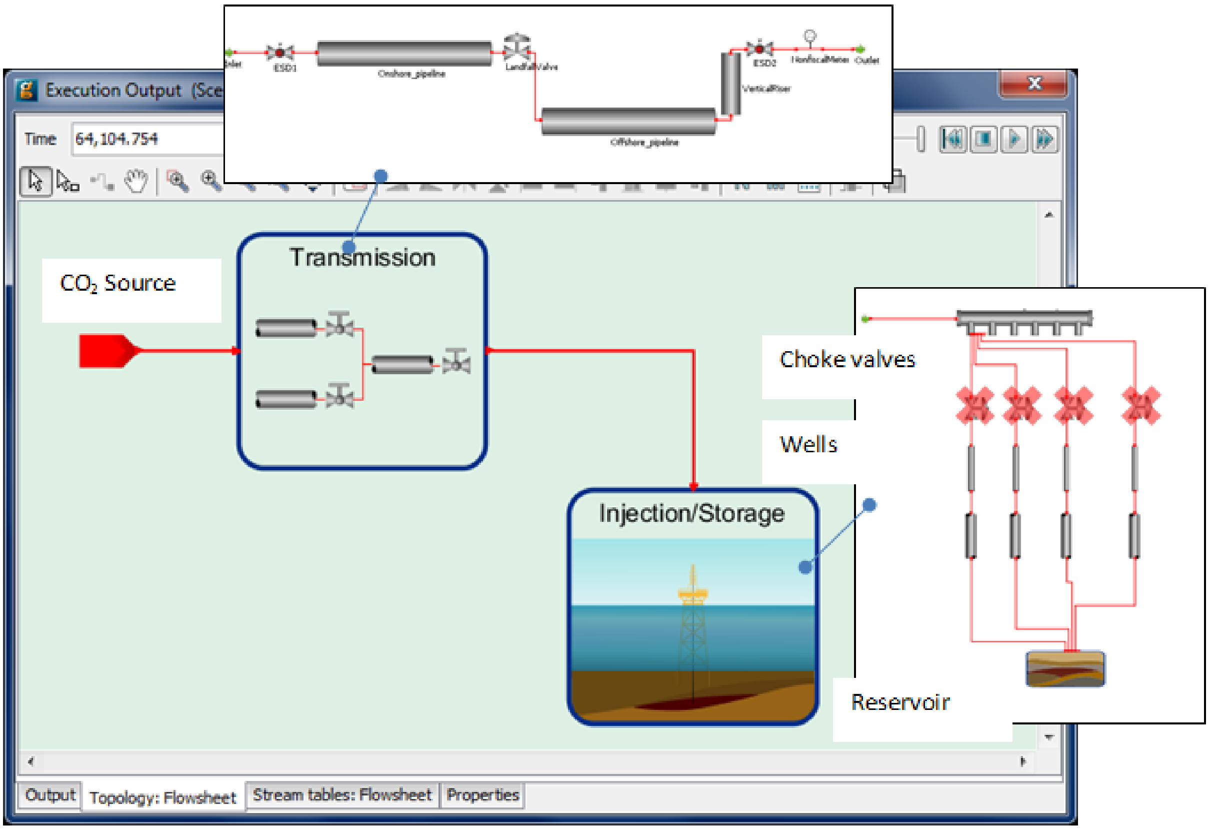


Figure 83: CO_2 transmission, injection and storage as modelled in gCCS – loss of storage scenario

7.2.4 Key analysis: How long it would take before this disturbance significantly affects the operation of the compressor (i.e. reaches maximum pressure / minimum flow condition and begins recycling).

The system was simulated using steady state initial conditions. The choke valve stem position setting was set to zero after 1 minute of steady state simulation. The valves have their inherent dynamics and the actual stem position is fully closed within 5 seconds. This dynamic is illustrated in Figure 84.

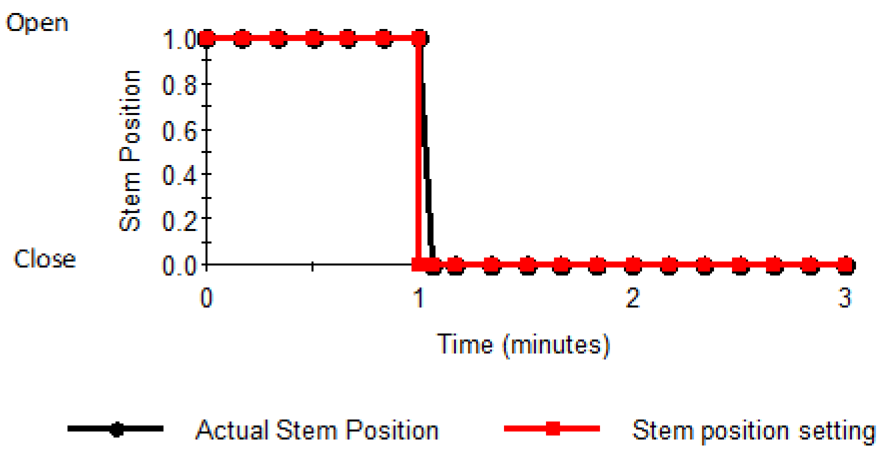


Figure 84: Actual stem position vs Stem position setting

Then, by evaluating the drive speed, we can evaluate how long it takes this disturbance to have a limiting impact on the CO_2 compressor at the CCS plant gate. From Figure 85 it can be observed that the compressor reaches its maximum speed after approximately 3.5 hours.

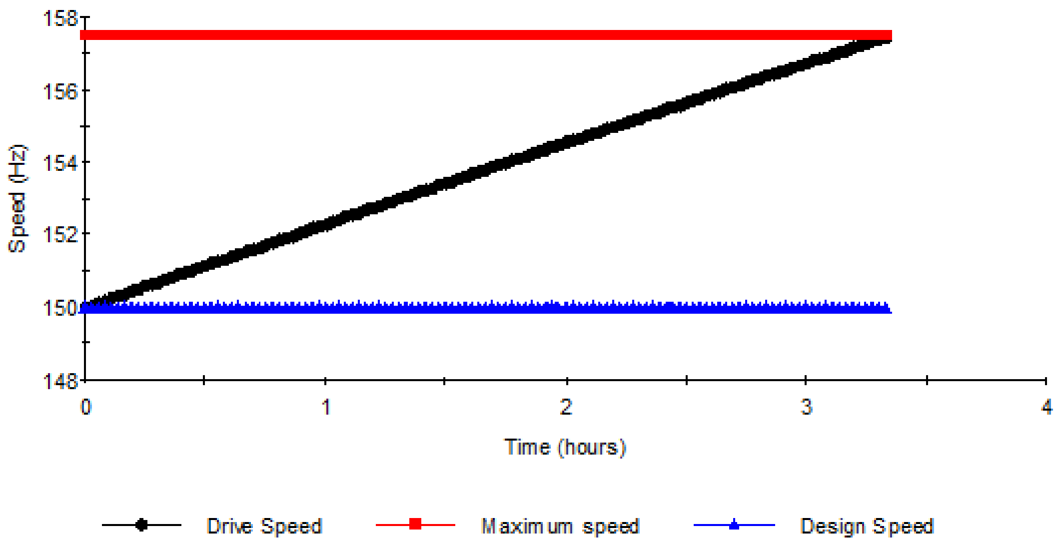


Figure 85: Determining the maximum compressor discharge pressure based on drive speed

7.2.5 Key analysis: How long before pipeline maximum pressure is reached?

In this scenario, the maximum allowable pressure is 153 bar. The pressure at the base of the vertical riser is typically the highest pressure in the transmission pipeline system. Based on the conditions outlined in Table 53, this pressure would be attained (if nothing else changes) after approximately 5.5 hours from the onset of the disturbance, as it is illustrated in Figure 86.

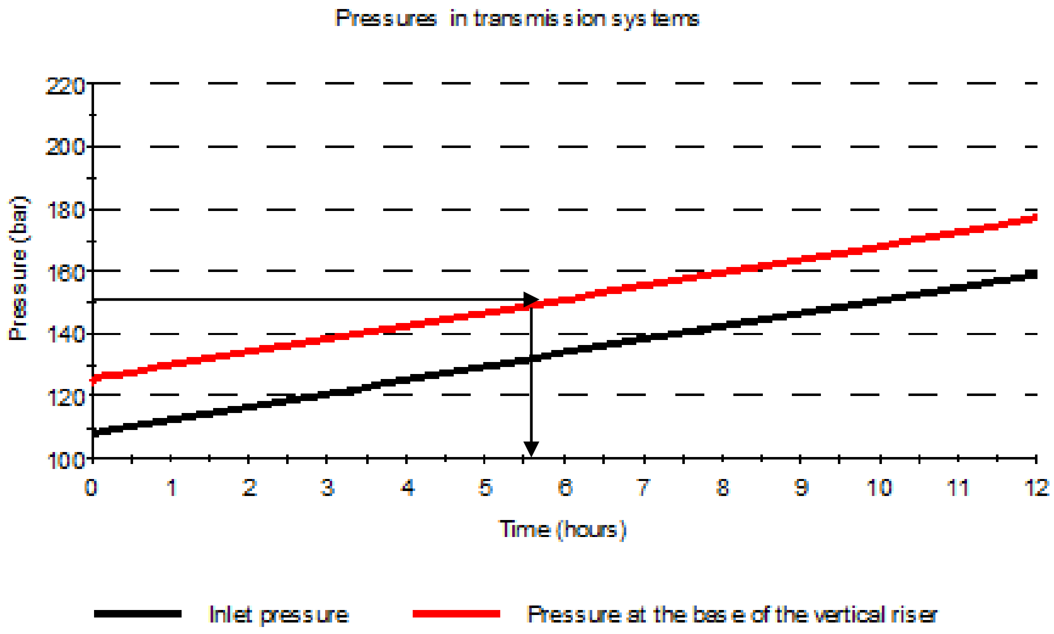


Figure 86: Increase in inlet pressure as a function of time following a storage failure

7.2.6 Key analysis: Impact on the capture plant – how much time is available before the capture plant needs to be shut down?

The capture plant would need to be shut down, or begin venting CO_2 once the compression plant shuts down, which in this scenario is after 3.5 hours.

7.2.7 Key analysis: Would any units on the chain require an emergency shutdown?

This would appear to depend on whether all relevant units can be shut down within the 3.5 hour time frame before the compressor fails to deliver CO_2 flow. The dynamics of the build-up of pressure appear gradual.

7.3 Scenario 2: Loss of upstream compression

This scenario could arise if the CO_2 compression train were manually or automatically tripped.

7.3.1 Anticipated impact

It is anticipated that the failure of the compression facility will have two significant effects:

a) It will cause the pipeline pressure to drop, leading to loss of injection capability (and probably a trip of the injection well)

b) The loss of the ability to remove CO_2 from the capture process, leading to either a shutdown of the capture plant and consequent shutdown of the power plant or initiation of CO_2 venting.

It can be assumed that b) will occur very quickly following the loss of compression, so this will not be included in the scenario.

7.3.2 Included gCCS flowsheet components

- The compressor would be represented by a CO_2 flow source, using a predetermined flow curve
- All components downstream of the compressor.

7.3.3 Scenarios and implementation in gCCS

The scenarios will be implemented by applying the aforementioned CO_2 flow curve to the flowsheet illustrated in Figure 87. As before, the Source represents the CO_2 flow from the compression system, the flow is 138.4 kg/s. The CO_2 is supercritical and its conditions are assumed to be ***T*** = 50ºC and ***P*** = 108.1 bar_a.

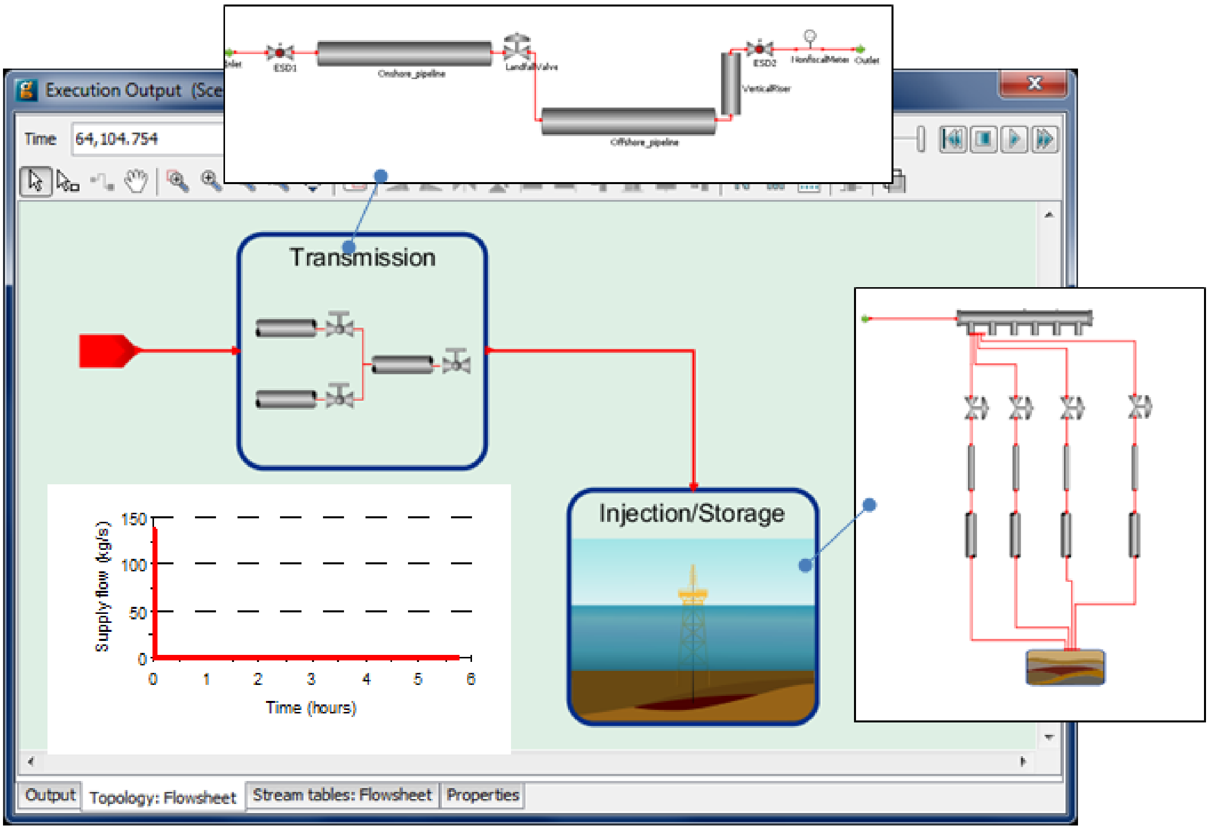


Figure 87: CO_2 transmission, injection and storage as modelled in gCCS- loss of compression scenario

In order to simulate the loss of a compressor, the flow of inlet CO_2 supply is dropped to 0kg/s in a step change. This is intended to represent the most drastic situation, and is illustrated in Figure 88. As can be observed, it takes 3.6 hours for the CO_2 supply at the injection site to reduce to 50% of the design flow, following a complete loss of supply

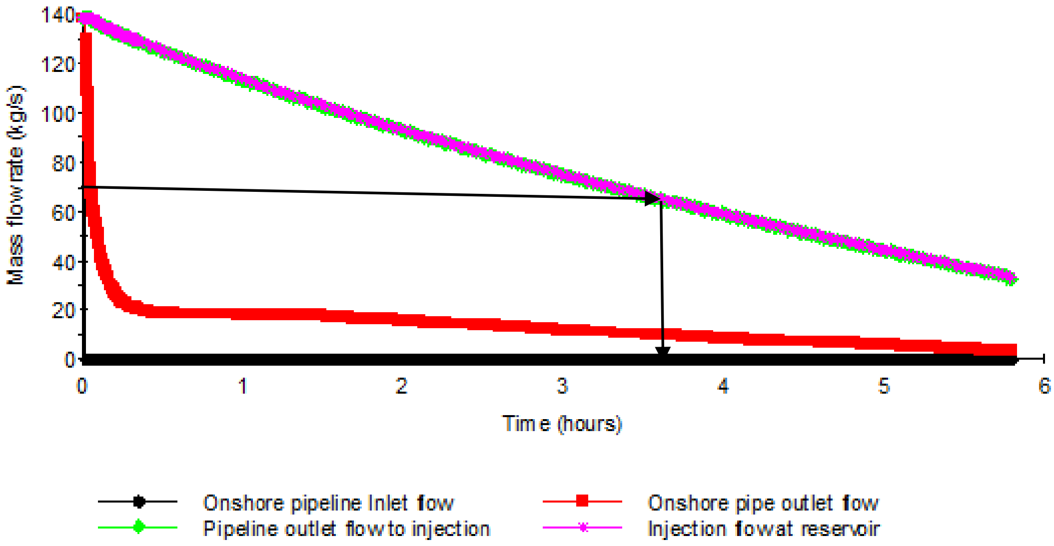


Figure 88: Mass flowrates at different locations in the CO2 transmission and injection sub-systems

7.3.5 Key analysis: Risk of two-phase flow in the pipeline

The original phase is dense phase. Two phase flow could occur with a drop in pressure below the critical pressure. The lowest pressure point in the system is the highest elevation in the onshore pipeline. If this drops below 73.8 bar, two-phase flow could develop.

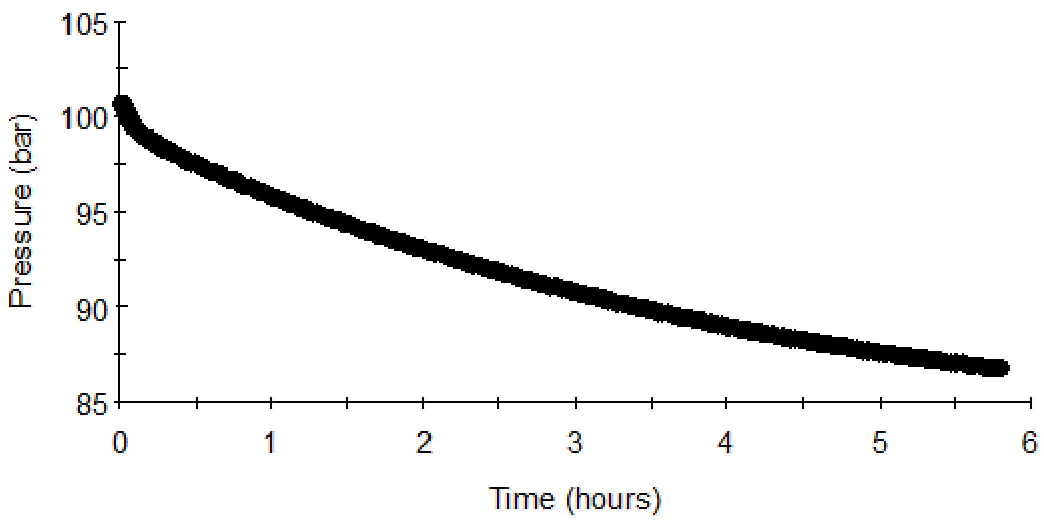


Figure 89: Minimum gauge pressure in the transmission system

Based on this simulation, it is not expected that this will occur within 5 hours of loss of supply.

7.3.6 Key analysis: Risk of reverse flow in the pipeline

If there is an instance where pressure downstream exceeds that of the upstream pipeline, a reverse flow situation could occur. It was found that, in the case of a loss of compression, with no intermediate compression along the pipeline, a reverse flow scenario could occur. Reverse flow would occur first at the choke valves. This is illustrated in Figure 90. The pressure drop across the Choke valve (blue line) drops to a slightly negative value just before the 6-hour mark.

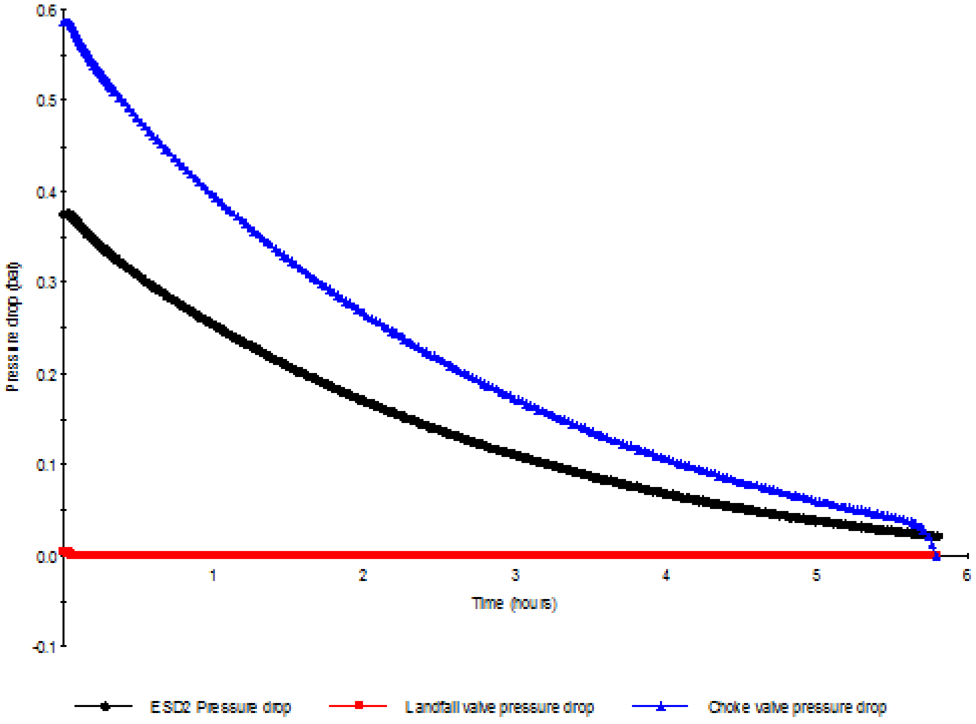


Figure 90: Pressure drop across various valves in the system

8 Conclusions of this study

- Energy market models support the notion that power plants will need to adopt flexible operation patterns in the future. Furthermore, they show that fluctuations between full and part-load operation will become more frequent during the day, and therefore an appropriate control strategy is necessary to guarantee the safety and profitability of the process.
- Several control strategies have been proposed in the literature. However there is a need to implement these conceptual strategies using high-fidelity modelling tools.
- Tools such as gCCS are are very useful to accurately predict the behaviour of power plants with carbon capture under flexible operating conditions. Design of suitable control strategies should be carried out using such tools.
- With an appropriate and well-tuned control strategy, it is possible to maintain critical parameters such as CO_2 capture at the desired set-point, even during periods of significant fluctuation in the power plant load
- Using an appropriate control strategy, even if based in simple and well established control technologies, such as PID, avoids the need for more expensive solutions such as adding solvent storage tanks to the process.
- From the operational point of view, using the solvent flowrate as manipulated variable to control the CO_2 capture rate is a better option than manipulating the reboiler temperature and therefore the lean loading.
- Despite the tuning issues and more oscillator behaviour, strategy 2 is more profitable for the PCCP case whereas for CCGT there was no obvious difference between the different strategies.
- The controllers in each case study can be tuned, but some oscillations can occur during the real time operation of the plant.
- Strategy 2 which was more profitable in PCPP has an advantage over the others, since it keeps the flows in the two columns steady and is able to maintain stable hydraulic conditions.
- There is a significant degree of interconnectivity between the various elements of the CCS chain – each of these elements has the potential to impose significant constraints on the operation of the remainder of the system. In this study, we have explored some of the boundaries of the transport and storage elements, but it is clear that a further and more detailed analysis is warranted.

9 Recommendations for future research

The recommendations below provide what we believe to be the basis for further progress in the area of modelling, simulation, and control of carbon capture plants subject to flexible operation.

9.1 Modelling and simulation aspects

The results of this study demonstrate the benefits of high-fidelity models that allow testing a number of scenarios without the need for costly and time-consuming experiments that could be prohibitive in real power plants and capture plants. The response to a survey, carried out with industrial operators and technology providers, shows that the importance of modelling and simulation tools is equally acknowledged by industry.

Tools such as gCCS provide an important contribution to the availability of models that integrate the different components of the CCS chain: power generation, carbon capture, transportation, and storage. This study benefited from the use of well-established dynamic models for post-combustion capture through which it was possible to study the performance of control systems under different flexible operation scenarios. It is suggested to further take advantage of these capabilities by performing studies related to critical operations such as start-up and shut-down. Additionally, it is suggested that more modelling efforts should be carried out to bring the additional elements in the chain to the same level of detail as the post-combustion plant; for example, the pulverised coal and natural gas fired power plant models in gCCS have some components which are modelled as steady-state. It would be particularly interesting to have a more detailed dynamic model of the pulverised coal boiler, in which the components - economiser, steam drum, superheater, steam condenser, reheater, and air circuit air path - are currently lumped into a single model.

As a follow-up to this work, it is also suggested for the proposed control strategies to be evaluated in the presence of different solvents and process alterations in the post-combustion capture plant. This would involve developing flowsheets in gCCS that include common modifications such as intercooling in the absorber column and heat integration with the power plant and compression system.

9.2 Simultaneous process and control design

In this study, it was assumed that the design of the power plant and post-combustion capture plant were fixed, and that the implementation of different control strategies corresponded to a retrofit exercise where only operational variables, such as temperatures or pressures could be varied. However, for the case of new plants, better overall performance may be achieved by considering the design of the process and control systems as a single problem. This approach has been long established [29] and may be carried out using different methodologies based on controllability analysis or optimisation. While high-fidelity dynamic models, such as those in gCCS, are available, the major challenge is the choice of an appropriate algorithm that can handle the computational burden arising from the optimisation problem formulation.

9.3 Advanced process control techniques

9.3.1 Centralised vs distributed MPC

Classical process control systems, such as proportional-integral-derivative (PID) control, utilise measurements of a single process output variable (e.g., temperature, pressure, level) to compute the control action needed to be implemented by a control actuator so that this output variable can be regulated at a desired set-point value. PID controllers have a long history of success in the context of chemical process control and will undoubtedly continue to play an important role in the process industries. In addition to relative ease of implementation, maintenance and organisation of a process control system that uses multiple single-loop PID controllers, an additional advantage is the inherent fault-tolerance of such a decentralised control architecture since failure (or poor tuning) of one PID controller or of a control loop does not necessarily imply failure of the entire process control system. On the other hand, decentralised control systems, like the ones based on multiple single-loop PID controllers, do not account for the occurrence of interactions between plant components (subsystems) and control loops, and this may severely limit the best achievable closed loop performance.

While there are very powerful methods for quantifying decentralised control loop interactions and optimizing their performance, the lack of directly accounting for multivariable interactions has certainly been one of the main factors that motivated early on the development of model-based centralised control architectures, ranging from linear pole-placement and linear optimal control to linear model predictive control (MPC). In the centralised approach to control system design, a single multivariable control system is designed that computes in each sampling time the control actions of all the control actuators accounting explicitly for multivariable input/output interactions as captured by the process model.

However, the substantial increase of the number of decision variables, state variables and measurements, may increase significantly the computational time needed for the solution of the centralised control problem and may impede the ability of centralised control systems (particularly when nonlinear constrained optimisation-based control systems such as MPC are used), to carry out real-time calculations within the limits set by process dynamics and operating conditions. Furthermore, this increased dimension and complexity of the centralised control problem may cause organisational and maintenance problems as well as reduced fault-tolerance of the centralized control systems to actuator and sensor faults [30]. These considerations motivate the development of distributed control systems that utilise an array of controllers that carry out their calculations in separate processors yet they communicate to efficiently cooperate in achieving the closed-loop plant objectives.

When referring to a post-combustion plant integrated with a power plant (coal or natural gas), advanced model predictive control formulations can be utilised as opposed to decentralised PID and a very interesting study would be to compare the two in terms of closed-loop settling time, integral square error and compliance of operational and environmental constraints. Moreover a centralised MPC can be also compared to distributed MPC formulations (DMPC) to see which of the formulations exhibits the best performance.

9.3.2 Economic MPC

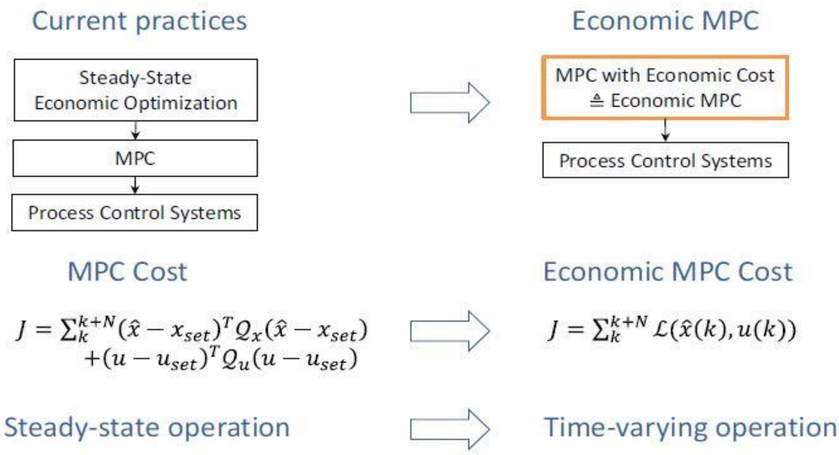
As previously stated, model predictive control (MPC) has been highlighted and studied for the last few decades. It is a feedback design technique which determines input and output variables by taking into account the current state of a plant and all sorts of constraints to be fulfilled.

More closely, it is a model-based method that through constrained optimisation addresses the trade-off between cost and performance, typical of all engineering designs. On one hand its goal is to find the appropriate control action which minimises the given objective functional, i.e., the control effort or input energy. At the same time the controllers should select a control action steering the states of the plants to the desired operating condition within reasonable amount of time. In both levels the controller should consider the constraints on states, input and output variables of systems assumed for the operation of the process. In fact, one of the main advantages of MPC that has allowed wide adoption in industrial processes is this capability of explicitly dealing with constraints. Standard MPC, also often referred as tracking MPC, adopts the distance (or a related quadratic form) between the current state and the set-point as its cost function to minimize. Therefore, in order to minimize the sum of stage costs over any time horizon directly implies that the control action steers the states to the set-point. Some systems, however, require to optimize some economic cost rather than just drive the state to a specific destination, such as the design of controllers for large-scale systems which includes economic issues such as cost margins, cost reduction or operation efficiency [31]. The peculiarity of economic MPC is to change the stage cost functional from the distance between a set-point and the current state to a cost with economic consideration [32].

Economic MPC (eMPC), is a variant of MPC to optimise objective functions acquired in actual process operations directly. Therefore the objective function in economic MPC has a physical meaning in itself or a specific value to evaluate the performance or profitability of the plant's operation.

The model predictive control (MPC) strategies, that employ an economic-related cost function for real-time control, have lately proved a numerically efficient approach to managing the portfolio of energy usage in various residential and industrial projects. They are designated as economic MPCs, whose main endeavour is to cope with regularly changing energy prices. Unlike the traditional MPCs, economic MPCs optimise the process operations in a time-varying fashion, rather than maintain the process variables around a few desired steady states. The process may thus totally operate in the transient state with economic MPCs.

The economic MPC has been found effective in which the real-time optimisation (RTO) layer is not required for computing targets to the lower layer MPC as usually be the case in the process industry. Figure 91 depicts the differences between the typical hierarchical RTO plus MPC structure and the economic MPC. The objective function of EMPC consists of the economic objective and the MPC target tracking objective functions. An EMPC "directly and dynamically optimises the economic operating cost of the process, doing so without reference to any steady state" [33].



$\hat{x}$: the value of the controlled variable, $x_{set:}$the value of the set-point, $Q_{x:}$the weight of the controlled variables reflecting the importance of the controlled variable, u: the value of the manipulated variable, u_{set}:the nominal value of the manipulated variable, Q_u: the weight of the manipulated variables reflecting the importance of the manipulated variable, N: is the prediction horizon, $L(\hat{x}(k), u(k))$: economic cost function.

Figure 91 – RTO plus MPC vs. Economic MPC

Rawlings et al. have first described the so-called unreachable set-point in MPC implementations in their paper [34], and shown that the set-point tracking MPC exhibit some advantages over the traditional target-tracking MPC [35, 36], when the set-points are not reachable. It is recommended that, the approach “should also prove useful in applications where optimisation of a system’s economic performance is a more desirable goal than simple target tracking”.

Feasibility, performance and optimality have been thoroughly discussed issues in the research on economic MPC. The optimality analysis, however, does not focus on the speed of convergence to a certain set-point, or tracking ability of specific trajectories, and this is why the stage costs adopted are typically different from the positive definite quadratic form of standard MPC. To enhance the economic performance of the systems, the optimization process is approached by decomposing it into two levels of optimization. The first level which determines the best region of operation is obtained by optimising the cost functional with respect to the steady state [37]. The solution in the first level, which is called the best steady-state, is then delivered to the second real-time optimization (RTO) level, which is implemented with MPC [31]. This layer will operate with the same stage cost of the previous layer. Therefore, unlike standard MPC, the stage cost function for economic MPC is not necessarily minimum at steady-state since the best steady-state is only the minimum among steady states but not necessarily among all feasible states.

As described above the formulation of an economic MPC model in large systems is quite important, especially when the economics are a key constraint in the deployment and operation of the process. Bui et al. [38], in their review paper about the dynamic modelling and optimisation of flexible operation in post-combustion CO_2 capture plants, highlighted that dynamic process models coupled with economic analysis will play a crucial role in process control and optimisation. They suggested that a key point when studying the flexible operation of capture plants, is to couple economic models with dynamic process models to study the sensitivity of electricity prices during flexible PCC operation.

In order to do this we suggest the implementation of the economic model predictive control (MPC). To the authors knowledge there is no study in the literature related to EMPC on post-combustion processes.

More specifically, when the power plant (either natural gas or coal) is integrated with the post-combustion plant, for the regeneration of the solvent there is a steam input provided by the crossover between the IP and LP turbines. This leads to a drop in the power plant efficiency due to less steam in the turbines and therefore less electricity generation. A suggested EMPC model, would take into account the maximisation of the profit from selling electricity by taking into account the operational constraints of the process. This means that the EMPC would use electricity price signals to control the integrated system operating not at certain set-points but at physical constraints optimised each time by the economic objective function. By using price signals, both current and future prices, the optimal control signal is calculated at every time step over the prediction horizon to obtain a closed loop profile. Thus the operation of the post-combustion system is specified by the varying electricity prices. Another important aspect of this formulation is the cost of emitting CO_2 in the atmosphere which is one of the most important constraints in the formulation of the EMPC. If an emission price is considered, then the EMPC single optimisation problem is transformed to a multi-objective optimisation problem. A multi-objective optimisation task involving multiple conflicting objectives ideally demands finding a multi-dimensional Pareto-optimal front, i.e. a representative set of Pareto-optimal solutions.

9.4 Optimisation of control strategy switching

In this study, two control strategies with different objectives were introduced. It is clear that each strategy has its own merits, which lead to the implementation of third control strategy where the previous two are switched according to the load point in the power plant. However, the switching was defined to occur at a somewhat arbitrary threshold between full-load and part-load. An interesting study would be to encompass the costs and benefits of each strategy within an optimisation formulation, in order to determine the switching schedule that provides the best trade-off.

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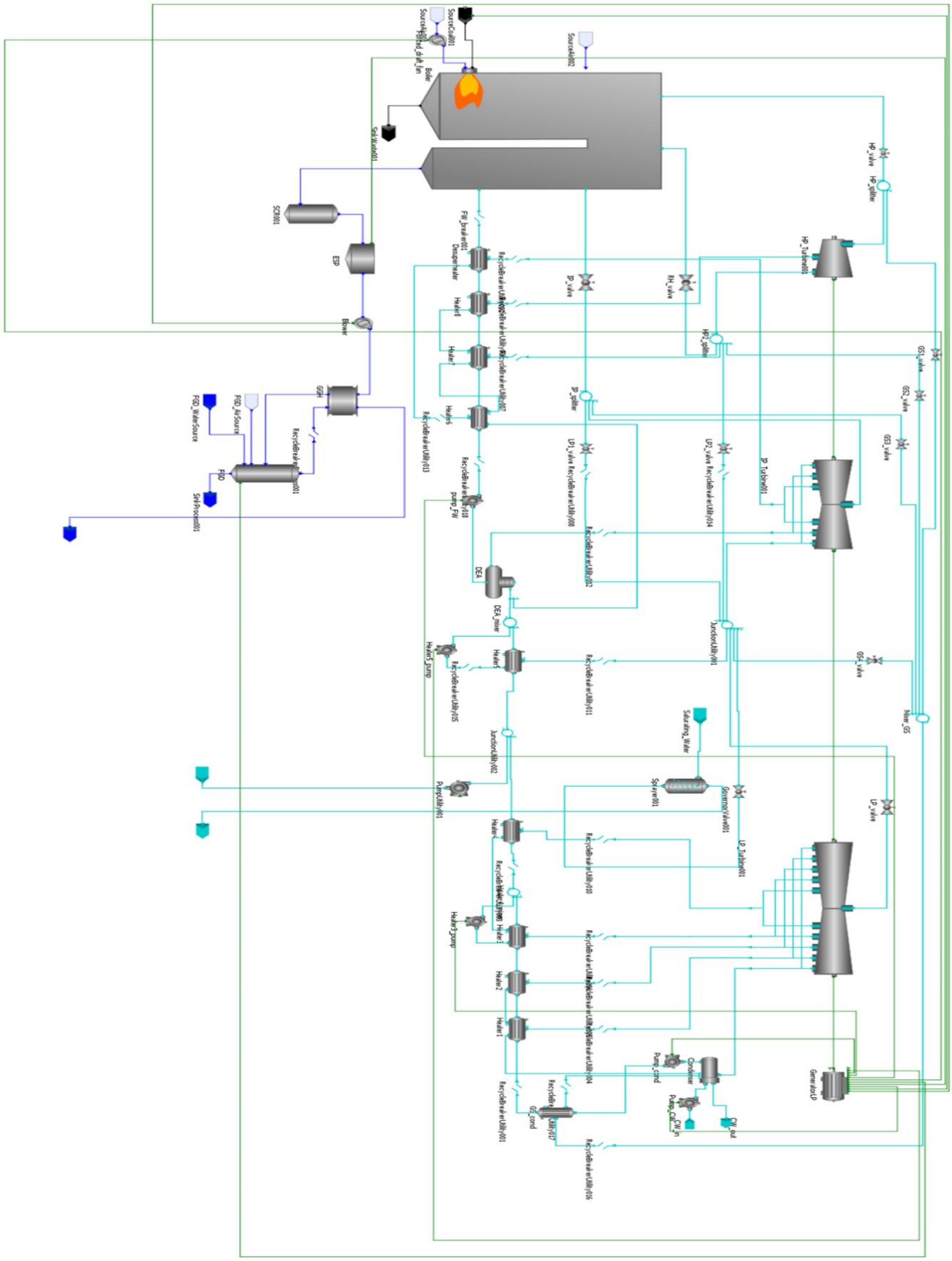
Appendices

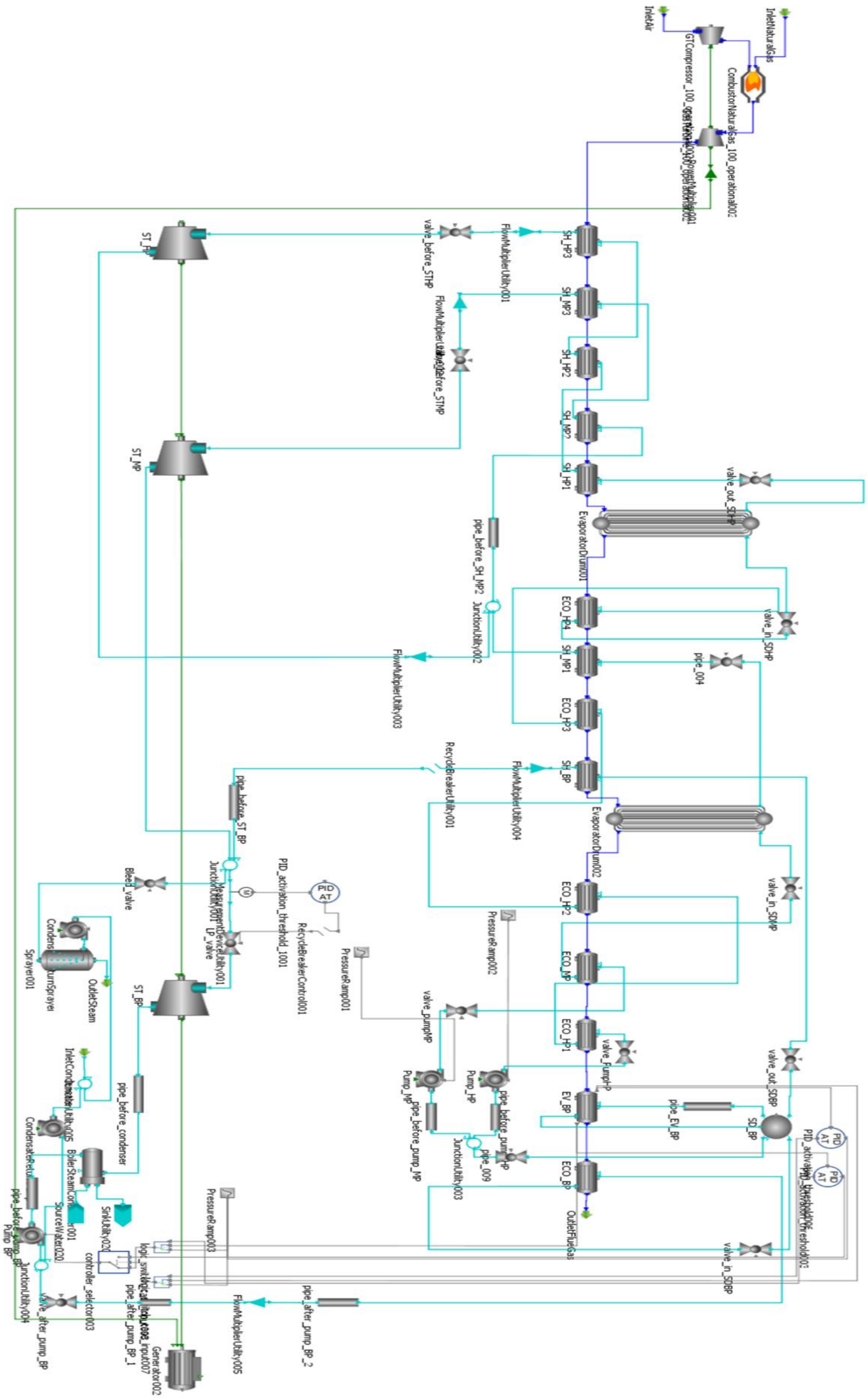
The various process flowsheets developed in gCCS are presented in this section.

Appendix A. Process flowsheets

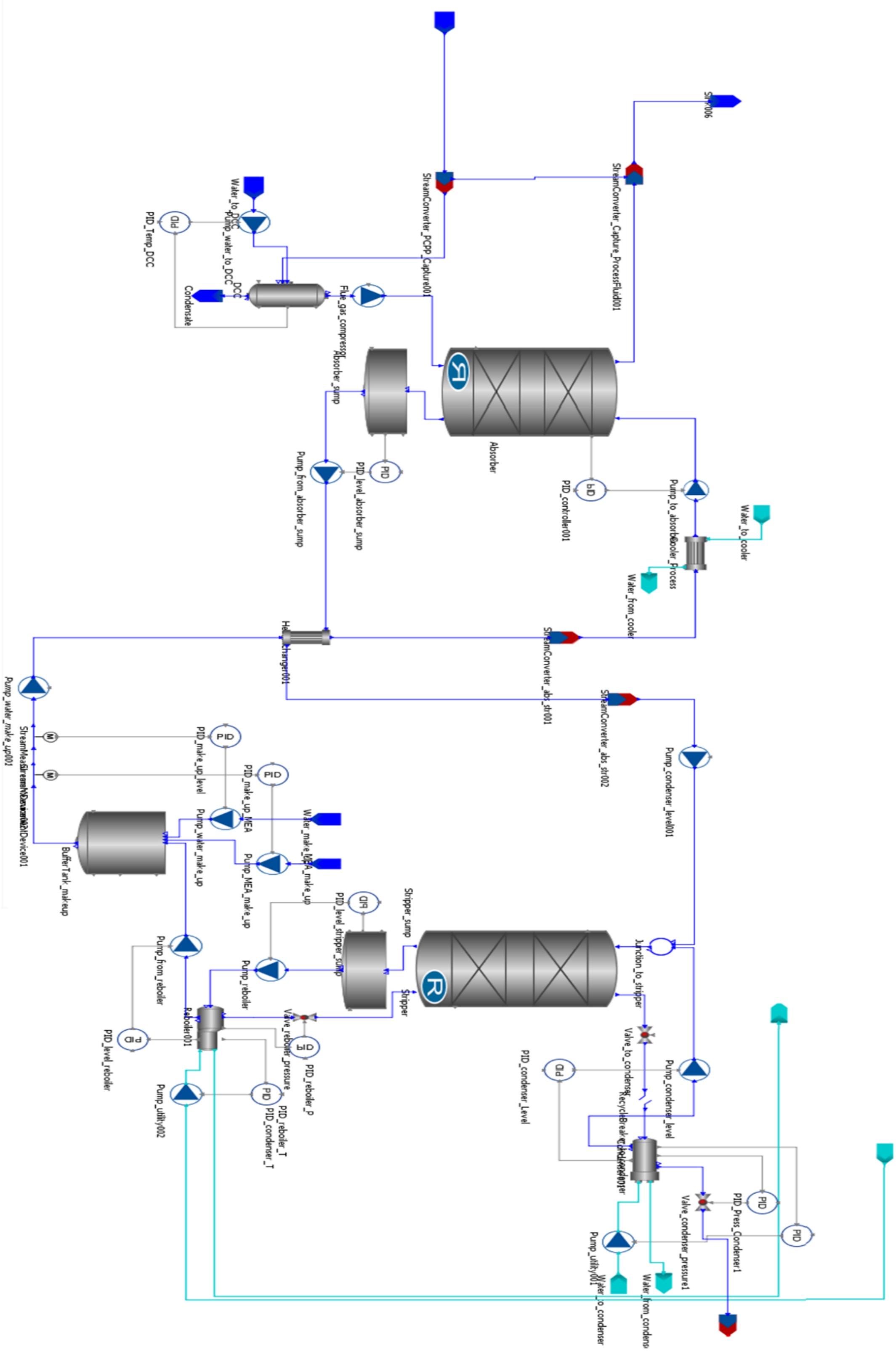
A.1 Pulverised-coal power plant flowsheet



A.2 Combined cycle gas turbine plant flowsheet



A.3 Amine-base post-combustion CO_2 capture plant flowsheet



A.4 Compression train flowsheet