Process control strategies for flexible operation of post-combustion CO_2 capture plants

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Abstract

With increasing penetration of intermittent renewable energy into the electricity grid, one can expect thermal power plants to be required to operate in a more dynamic fashion, with more frequent departures from design point operation. However, the application of optimal control strategies can offer solutions to these operational challenges, associated with the integration of the power plant with the capture plant. In this paper a process control strategy is developed in order to select the optimal control variables for a PCC process. In addition, economically efficient control structures for operation of a post-combustion capture process with minimum energy requirements for coal and natural gas power plant are designed. The results have shown that with an appropriate and well-tuned control strategy, it is possible to maintain critical parameters, such as the degree of CO_2 capture, at the desired set-point, even during periods of significant fluctuation in the power plant load and even if based on simple and well established control technologies, such as PID, avoids the need for more risky solutions such as adding solvent storage tanks to the process.

Keywords: `post-combustion`, flexible PCC, decentralised control

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1. Introduction

In order to mitigate the adverse effects of climate change and keep the global temperature increase below 1.5 C, an 80% reduction of greenhouse gas(GHG) emissions will be required by 2050 [1]. This implies a near complete decarbonisation of the electricity sector by 2050, with the requirement for the generation of carbon negative electricity in the second half of the century [2]. 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 commercial scale [3], [4].

In order to meet CO_2 reduction targets, decarbonised power plants operating with CCS will likely play a major role in the energy system. The important role of intermittent renewable energies in this energy portfolio implies that power plants with CCS need to be able to operate in a flexible, load following manner. The European Commission foresees that the electricity mix will be dominated by three generator types: 1) renewable sources with a share of 59-83 % of generated electricity, of which 42%-65% by Intermittent Renewable Energy Sources (IRES), 2) Carbon Capture and Storage (CCS) with a share of 7-32% and 3) nuclear energy with a share of 3-19% [1]. The IEA predicts similar trends for other OECD countries in its 450-scenario [5]. This implies that, despite the fact that the details of the solution will be different in different regions, the capacity of CCS plants to operate flexibly will be an important contribution to the energy system in many countries.

The low variable costs of intermittent renewable generators gives them an early position in the merit order, shifting the thermal power plants position from base-load toward mid-merit. The power plants will have to balance variations in intermittent renewable power production and provide sufficient balancing reserves, resulting in more changes in operation with more frequent ramping and start-stop cycles. As a result of these structural changes, the flexibility of the electricity system may become an important issue. The dynamic modelling and controllability analysis of the integrated power and capture plant is therefore highly important in this context. Most studies in the literature consider individually the capture plant, or even discrete parts of the capture plant such as the absorber and stripper for control, while in this study the integrated coal and gas power plants with post-combustion are considered [6], [7], [8], [9], [10], [11].

The concept of flexible CCS is not new, with some of the key concepts being considered in academic circles at least a decade ago [12], [13]. 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 [13] 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 [14], [13]. We have recently proposed a new approach where 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 [15], [16].

In their work, Ziaii et al, [9], proposed a ratio controller to maintain a 90 % CO_2 capture rate. The disturbance in the system was the steam flowrate in the reboiler, while the manipulated variable was the liquid to gas flow ratio. The operation of the plant without controller showed a settling time of 30 min compared to the ratio controller where the settling time was decreased in 1 min. Lin et al. [7], presented two different control strategies to control the CO_2 capture rate. The manipulated variables were the lean solvent flowrate and the reboiler heat duty. They observed that the strategy where the manipulated variable was the reboiler steam flowrate provided more stable hydraulic conditions in both the absorber and the stripper.

Panahi et al.[17], designed a control structure for the CO_2 capture process with the aim of achieving the most economically optimal CO_2 removal by using the self-optimising method [18]. To validate the proposed control structures, they considered dynamic modelling of the process along with different control configurations using decentralised and model predictive control (MPC) [10]. Here, the CO_2 recovery in the absorber was controlled by the reboiler duty and the temperature in the stripper was controlled by the recycle amine flowrate with the manipulator at the outflow of the absorber. The decentralised control model, has dynamic performance comparable to MPC, and is proposed as the best alternative because of its much simpler implementation.

Recently, Nittaya et al.[8], proposed three control strategies applied in a post-combustion plant. The design of those control strategies was based on the relative gain array (RGA) analysis, and heuristic approaches. Loop pairing between carbon capture and lean solvent flow rate and temperature reboiler and reboiler duty were implemented in all three controllers. However, control structure B featured faster responses to reject disturbances and track the changes in set points compared to the other control structures. Arce et al. [19], proposed and developed a multilevel model predictive control architecture for the solvent regeneration system of an amine-based CO_2 capture process. Focusing on minimising the operational cost associated with the post-combustion CO_2 capture, the two level control architecture maximises the flexibility of the solvent regeneration process whilst simultaneously ensuring cost-optimal performance and safety. In their work, Sandoval et al. [20], presented an advanced, centralised, multivariable, model predictive control (MPC) technique to address the controllability of a post-combustion CO_2 capture process from a coal-based power plant and compared it to decentralised Proportional-Integral (PI) control schemes. The CO_2 capture model was developed on the Aspen Hysys dynamic simulator while the MPC was implemented in Matlab. The MPC performed significantly better than the PI controllers in terms of closed-loop settling time, integral square error and compliance of operational and environmental constraints. Luu et al. [21] , presented three control schemes for a post-combustion process; *i.e.*, a standard feedback control scheme, a cascade PID scheme and a model-based strategy in the form of a model predictive control. The closed-loop simulation results showed that the MPC handles the control problems very well without violating economic, operational and environmental constraints. Manaf et al. [22], developed a mathematical black box model to predict the dynamic responses of a PCC plant, including the absorber, stripper and heat exchanger unit operations. Based on a sensitivity analysis, the reboiler heat duty and the lean solvent flowrate were selected as manipulated variables to control the energy performance (EP) and the CO_2 capture efficiency, respectively. The results showed that this model was proven to be capable of representing the responses of the capture plant and provided insights into the operational flexibility of the PCC process. Qadir et al. [23], examined the flexible operation of the PCC process while minimising the energy loss due to reduced power plant electricity output, by considering a capture plant and a solar thermal plant. The results have shown that the solar energy input increases the profitability of the system when used for power plant repowering. He et al. [24], presented a dynamic flexibility analysis of a post-combustion CO_2 capture process using model predictive control (MPC). They have investigated the dynamic performance of two key variables (CO_2 capture rate and CO_2 composition in the product stream) in open and closed loop. They have found that the closed loop system shows better performance but it is crucial to tune the control systems under high-frequency variations of the flue gas disturbance in order to avoid large oscillations under critical operating conditions in the load. They have also performed a simultaneous control and scheduling approach in order to examine the economic benefits in terms of CO_2 emissions and process energy demand. They have shown that dynamic optimisation can lead to reduction in the CO_2 emission penalty costs without additional energy consumption from the power plant compared to keeping a stable CO_2 capture rate.

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.

The novelty of this work compared to previous studies is the integrated approach of the power and capture plant and also the economic evaluation of the different control strategies which are proposed in this work. The different control strategies have been evaluated in terms of the economical and operational point if view, based on real operating profiles of the integrated gas and coal with post-combustion capture plants 25.

The remainder of this paper is laid out as follows: Section 2 presents the reference case models. Section 3 explains the systematic approach to conduct model-based PCC controllability studies and the proposed structure of control schemes. Section 4 performs closed loop analyses for stepwise set point tracking and disturbance rejection in order to evaluate the proposed control schemes. Finally, conclusions from this study are presented in Section 5.

2. Model development and implementation

This section presents the power plant and capture plant models used in this work. In this study, all models were developed using the gCCS toolkit [26]. gCCS, based on the gPROMS platform, provides an environment for modelling the behaviour of complex systems. gCCS is a state-of-the-art platform, developed by Process Systems Enterprise 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. We have provided selective information on the design and process parameters in order to keep a concise paper, however more details can be found in Mechleri et al. [27].

2.1. Coal fired power plant

This section presents the supercritical pulverised coal fired power plant used in this work producing a gross electricity output of 825 MWe and a net electricity output of 779 MWe at full load. The process flowsheet considered for this study is illustrated in Figure 1.

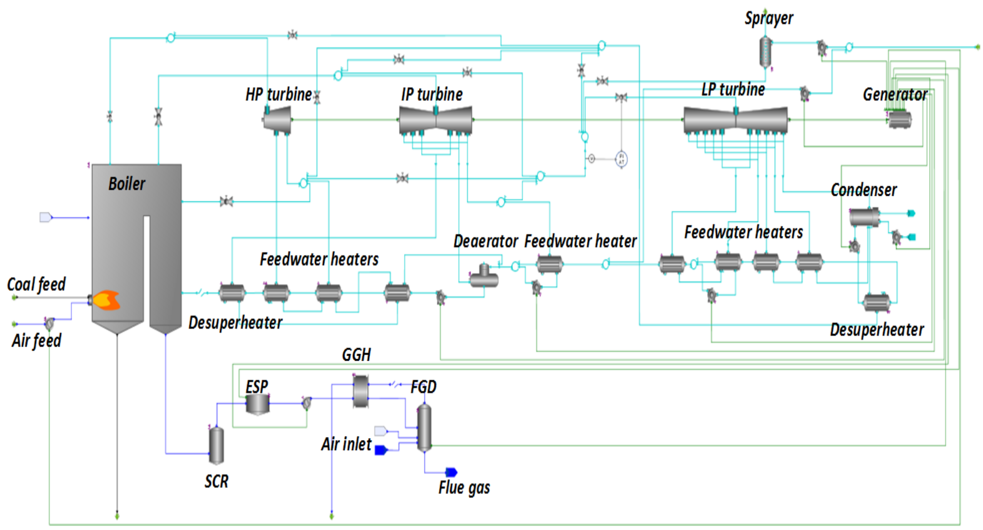

Figure 1: Supercritical pulverised coal reference flowsheet

Coal and air are fed in the boiler where the coal is burnt and produces heat. The exhaust flue gas passes through a selective catalytic reduction (SCR) where the NO and NO_2 are removed. It then passes through an Electrostatic Precipitator (ESP) in order to remove the particulate matter from the inlet stream. In the Gas-gas Heater (GGH) the two gas streams exchange heat, while in the flue gas desulphurisation unit (FGD) the sulphur dioxide is removed prior to final emission of the exhaust gas via the stack. The power needed in the blower and FGD is provided by the generator with an efficiency of 98.8 %. As is typical, the turbine train in power plant is considered to be composed of three distinct steam turbines, operating at high pressure (HP), intermediate pressure (IP) and low pressure (LP). The high pressure turbine is composed of two steam turbines with 91.73 % and 91.97 % isentropic efficiency, respectively. The IP turbine is composed of three steam turbines with isentropic efficiencies 93.98 %, 93.24 % and 93.72 %. The LP turbine composes of five steam turbines with efficiencies 91.32 %, 91.18 %, 91.59 %, 88.58 % and 96.75 %.

2.2. Combined cycle gas turbine power plant model

Figure 2 shows the flowsheet developed for a combined cycle gas turbine power plant producing a gross electricity output of 746 MWe and a net electricity output of 740 MWe at full load.

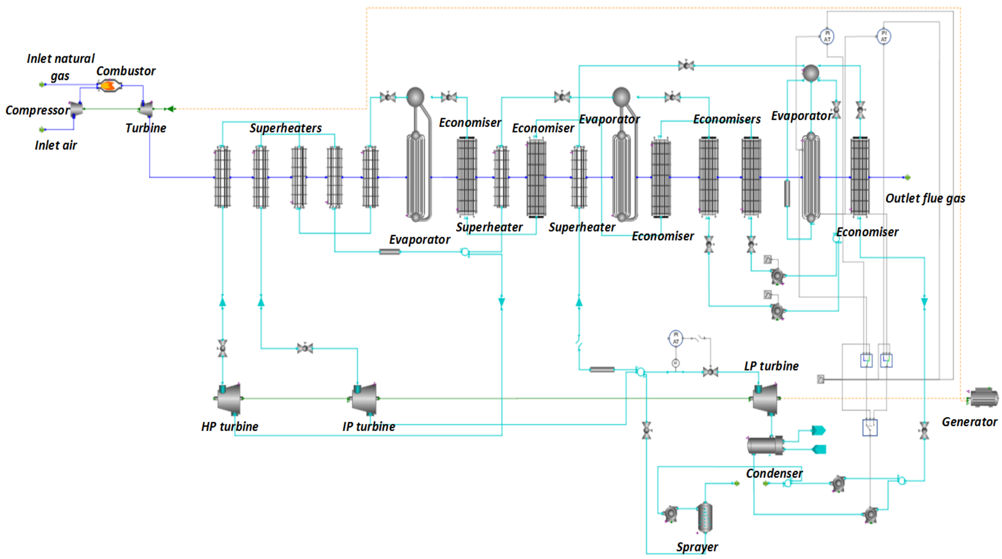

Figure 2: Combined cycle gas turbine power plant reference flowsheet

The Combined Cycle Gas Turbine flowsheet comprises two Gas Turbines (GT) of axial type 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, with 88.3 %, 96.7 % and 60.1 % isentropic efficiencies, respectively. 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 HRSG comprises three pressure levels with natural circulation evaporators. The low pressure pump increases the pressure of the condensate at 17 bar 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 with outlet pressures, 34 and 135 bar, respectively. 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 as an individual model. The steam produced in the low pressure drum is heated-up in a superheater at 495 K and it is mixed with the IP turbine exhaust before 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 with a temperature from 602 K to 807 K to the last superheater and it is expanded in the HP turbine. The exhaust steam from this turbine is then mixed with the IP steam and reheated to 796 K before 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 PI controller ensures the pressure at the extraction point does not fall below 3.5 bar by closing a valve located at this point.

2.3. Capture plant process

This section presents the CO_2 capture model used in this work. The process flowsheet considered for this study is presented in Figure 3.

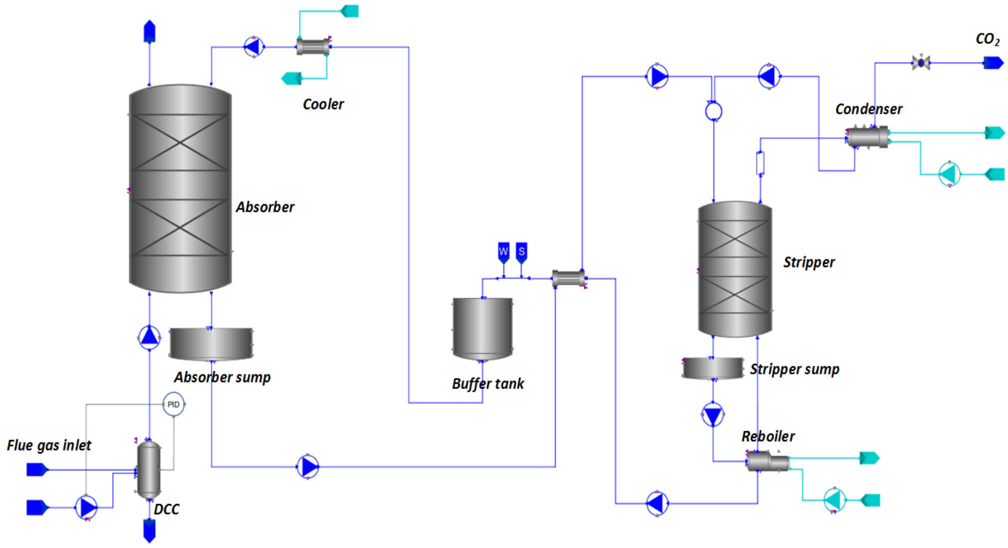

Figure 3: Reference capture process flowsheet

The reference capture process chosen for this study is a standard amine-based system using MEA 30 wt% as solvent and comprising an absorber section and a solvent regeneration section. The L/G ratio is 4 for the pulverised coal-capture plant and 1.3 for the combined cycle-capture plant and the lean loading 0.28 and 0.27 mol CO_2/mol MEA, respectively. As shown in that figure, the flue gas enters the absorber at the bottom and is contacted with the counter current flow of the amine solution. The two make-up streams of water and MEA are maintaining the solvent's stream concentration in the entire system. The thermophysical properties of the complex fluids used in this process are described using the statistical associating fluid theory [28], [29] for potentials of variable range [30] (gSAFT). gSAFT was also used to describe the effect of the reactions on the fluid-phase properties and phase behaviour of the MEA-H_2O-CO_2-N_2 fluid mixture in the CO_2 capture process. A detailed description of the thermodynamic modelling approach used and molecular models developed [31], [32] in addition to how the molecular models are integrated with the process models [33], [34], [16] is provided in separate contributions and is not repeated here. The capture plant model integrated with the coal and gas fired power plants both have the same configuration. However, the size of the different units of the process changes as presented in Table 1. Absorber specifications were determined following the parameterisation procedure described in Mac Dowell and Shah [34]. Finally, the height of packing was increased to achieve a 90% capture rate. It must be noted that after a sensitivity analysis conducted in the model the strippers size showed limited influence on the degree of solvent regeneration. 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.

Table 1: Unit specifications for post-combustion plant in coal and gas fired power plants

	Coal-PCC	Gas-PCC
Absorber packing height(m)	55	50
Absorber diameter (m)	20	20
Absorber sump height (m)	2	2
Absorber sump diameter (m)	20	20
Stripper packing height (m)	20	20
Stripper diameter (m)	10	10
Stripper sump height (m)	2	2
Stripper sump diameter (m)	10	10

2.4. Compression train

The compression train after the capture plant process is presented in Figure 4. Two compressor trains are used to increase the CO_2 pressure to 96 bar. The three first 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. Before the last compressor section a dehydrator further reduces the water content of the CO_2 stream. 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 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.

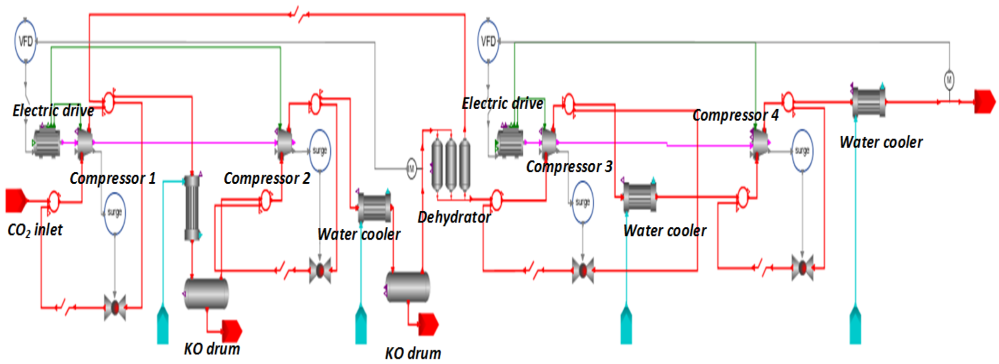

Figure 4: Compression train flowsheet

2.5. Integrated model of power plant with capture plant model

In order to gain insight into the transient behaviour of the capture plant in actual operation, it is necessary to study its interactions with the power plant. In Figure 5 the coal fired power plant with post-combustion and compression is presented. The flowsheet with the gas fired power plant is equivalent and is not presented here for economy of space. The connection points between the two units are three; the steam for the regeneration of the rich solvent which is extracted from the IP-LP crossover, the condensate returned to the feedwater heating train from the reboiler and the exhaust gas entering the capture plant from the power plant.

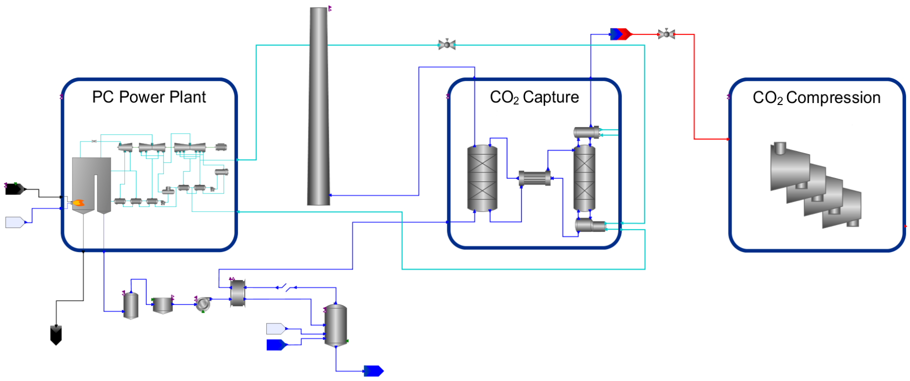

Figure 5: Integrated supercritical pulverised coal power plant flowsheet with capture and compression

It must be noted that the capture process and compression train impose a penalty on the electricity generated in the power plant. Tables 2 and 3 present the electricity generated in the coal and gas power plants respectively with and without capture and compression.

Table 2: Electricity generated in the coal-fired power plant with and without capture and compression

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

Table 3: Electricity generated in the gas-fired power plant with and without capture and compression

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

3. Control strategies for post-combustion capture plants

This section describes the methodology to develop and implement PID control strategies to be embedded into the PCC system. 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 [9], [35]. 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. These options, listed in Table 4, are discussed in the sections below.

Table 4: 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 flowrate
3	Dynamic switching betweeen strategies 1 and 2	

In Strategy 1, a controller maintains the carbon capture rate at a value of 90% by regulating the flow of lean solvent entering the absorber column. 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. 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 constant by manipulating the flow rate of steam extracted from the intermediate pressure turbine of the power plant. As in Strategy 1, the objective for Strategy 2 is to maintain the CO_2 capture rate constant throughout the power plant load change. In this case, the total 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 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. [7], [35]. Some studies suggest that the best control strategy for a post-combustion carbon capture plant depends on the operating region [18]. 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. The turndown of the absorption column, and therefore the flexibility of the capture plant, is largely dictated by the column intervals used, packed (structured or dumped) or tray columns, the type of inlet liquid stream and the type and number of liquid and vapour distributors and liquid collectors. The efficient operation of the CCS plant will ensure that in off-design conditions the capture plant can operate flexibly. 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 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 [15], [33].

In addition to the aforementioned control objectives of the study, there are some additional control variables that need to be taken into account to maintain the feasible operation of the system in the presence of critical realisations in the disturbances, e.g., sustained oscillations in the flue gas flowrate with high and low frequency content. There are four liquid levels that need to be controlled to maintain the inventories within the system, i.e., the liquid levels in the absorber sump tank, stripper sump tank, reboiler and condenser. The additional control variables in the system are the composition of the lean solvent, the pressure and temperature in the condenser, the pressure and temperature in the reboiler, and the temperature in the direct contact cooler (DCC). The additional control loops in the process which guarantee the operation of the capture plant during the power plant load change are presented in table 5.

Table 5: 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	0.41 m
Temperature controller	Condenser temperature	Cooling water mass flow	313.15 K
Level controller	Condenser level	Lean MEA from condenser mass flow	0.71 m
Level controller	Absorber sump liquid level	Rich MEA mass flow	0.63 m
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	0.5 m

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 3. 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 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.

3.1. Tuning of controllers

When using gCCS it was found that varying the controller gain as it is widely recommended often resulted in numerical errors due to the complexity and large 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:

$$z^{calc} = K(P + I)(z^{max} - z^{min}) + B \tag{1}$$

where P and I are the proportional and integral terms calculated by equations 2 and 3, respectively.

$$P = \epsilon \tag{2}$$

$$\tau_I dI/dt = \epsilon \tag{3}$$

where ϵ is the error calculated by equation 4.

$$\epsilon = (y^{SP} - y)/(y^{max} - y^{min}) \tag{4}$$

In equation 1, K is the proportional gain and B the controller bias (the controller output when all errors are zero) in equation 3 τ_i is the integral action and in equation 4 y^{SP} is the controller setpoint, y is the measured variable and y^{max} and y^{min} are the maximum and minimum values of the controller input respectively.

The tuning parameters for the PC-CP and CC-CP plants for the different control strategies, are presented in tables 6, 7 , respectively.

Table 6: Tuning parameters for PC-CP process

	Controlled variable	Manipulated variable	K	τ_i
Strategy 1				
Loop 1	CO_2 conversion	Lean solvent flowrate	1.5	0.1
Loop 2	Reboiler temperature	Steam flowrate	0.05	30
Strategy 2				
Loop 1	CO_2 conversion	Lean loading	5	500
Loop 2	Lean Loading	Steam flowrate	1	10
Strategy 3				
Parameters from Strategies 1 and 2				

Table 7: Tuning parameters for CC-CP process

	Controlled variable	Manipulated variable	K	τ_i
		Strategy 1		
Loop 1	CO_2 conversion	Lean solvent flowrate	1	0.1
Loop 2	Reboiler temperature	Steam flowrate	1	10
		Strategy 2		
Loop 1	CO_2 conversion	Lean loading	10	1000
Loop 2	Lean loading	Steam flowrate	0.001	0.5
		Strategy 3		
		Parameters from Strategies 1 and 2		

More information about the detailed procedure followed for the tuning and also for the additional control schemes of the post-combustion plant can be found in Mechleri et al. [27].

Moreover in order to limit the behaviour of the controllers, we have set upper and lower bounds at the input and output variables of each controller, applied for all the different control strategies. For the capture rate we have set a lower bound of 0.1 and upper bound of 0.99, while the lean solvent flowrate varies from 0 to 2000 kg/s. The reboiler temperature varies from 300-400K with a steam solvent flowrate from 10-300 kg/s. Accordingly the other additional control schemes have been set.

4. Results and discussion

The three different control strategies, as described in the previous section, were evaluated in terms of dynamic performance and economic benefit, for the pulverised coal-capture plant and combined cycle-capture plant. For the dynamic performance of the different control strategies, a variation in the flue gas flowrate has been applied, as the main disturbance in the process and the response of the different control strategies has been observed. The economic evaluation is representative of the flexible operation of the power plant as described in our previous work [15], [16] and is illustrated in Figure 6.

The 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 using the following equation:

$$\frac{£^{SRMC}}{MWhr} = \frac{£^{MWhr}_{Fuel}}{n_{plant}} + (£^{CO_2}_{Tonne} \cdot CI^{TonnesCO_2}_{MWhr}) + £^{Var}_{O\&M} + £^{CO_2}_{T\&S} \tag{5}$$

where $£^{SRMC}$ is the SRMC of the electricity generated by a given plant. In this calculation the variable operating and maintenance costs ($£^{Var}_{O\&M}$) and fixed cost ($£^{CO_2}_{T\&S}$) for transport and storage are also considered.

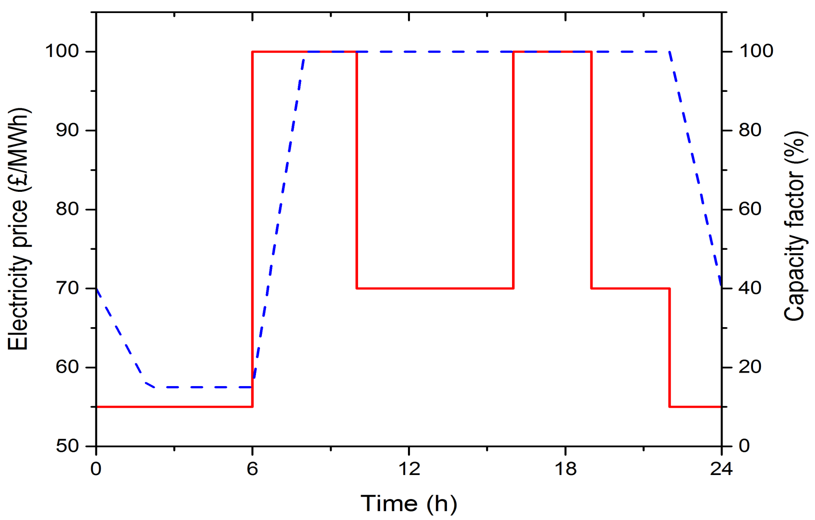

Figure 6: Variation of electricity prices throughout the day, and typical operation profile for two-shifting power plant

The variation in electricity prices, as well as the cost of utility water and the assumed value of CO_2 are shown in Table 8.

Table 8: Electricity prices, assumed CO_2 value, utility water cost and fuel prices used in this study

Electricity prices	
06:00-09:00	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
Coal price	9.86 £/MWh
Natural gas price	24.53 £/MWh

4.1. Coal fired power plant

4.1.1. Dynamic performance of the different control strategies

In this section the behaviour of individual control variables for the PCPP case are detailed.The main disturbance in the process is the flue gas flowrate variation, which is representative of the operation of the power plant, as illustrated in Figure 7. 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.

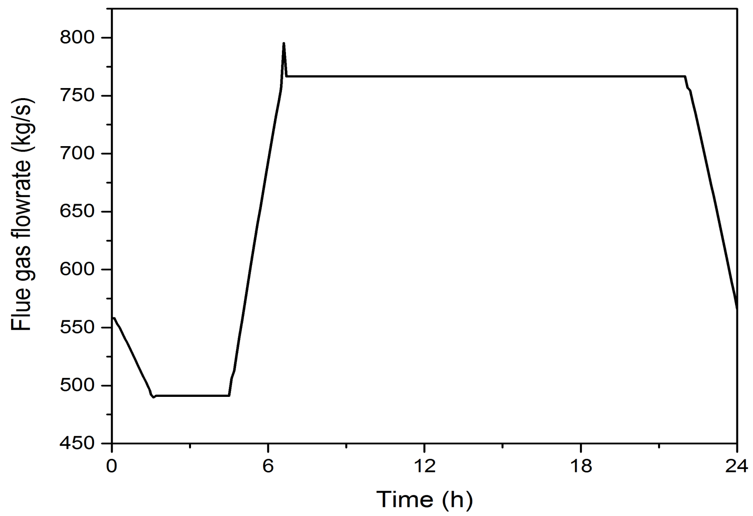

Figure 7: Flue gas flowrate variation during 24 hour operation in the PC-CP process

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 8 during the period between hours t=6 and t=14 there is a small but persistent fluctuation in the degree of CO_2 capture. This is explained by the fact that manipulating the reboiler temperature, has a much slower impact on capture rate considering the huge holdups of solvent in the system. As a result, there is a higher risk of overshoot with such a control strategy. Furthermore, manipulating the reboiler temperature implies changing the steam demand which affects the performance/efficiency of the power plant and ultimately introduces further disturbances. These coupling impacts are really only observed when using truly integrated process models. On the other hand, manipulating the solvent circulation rate will have a much quicker impact on the capture rate and thus provides more effective control and therefore no deviation of the control variable as illustrated in Figure 8. The reverse is observed for Strategy 3; the amplitude of the departure from the set point is relatively large, but short-lived. The dynamic switching between the two strategies can cause instability to the whole process due to the time delay between the two control schemes as it is observed from Figure 8. Therefore it is better to chose a single control strategy in order to overcome these large oscillations. When considering the power plant the net power output is affected by the control schemes applied to the capture process and more specifically for Strategy 2 where we observe large oscillations from the set point. The consideration of a power controller is the way that someone would undergo these transients.

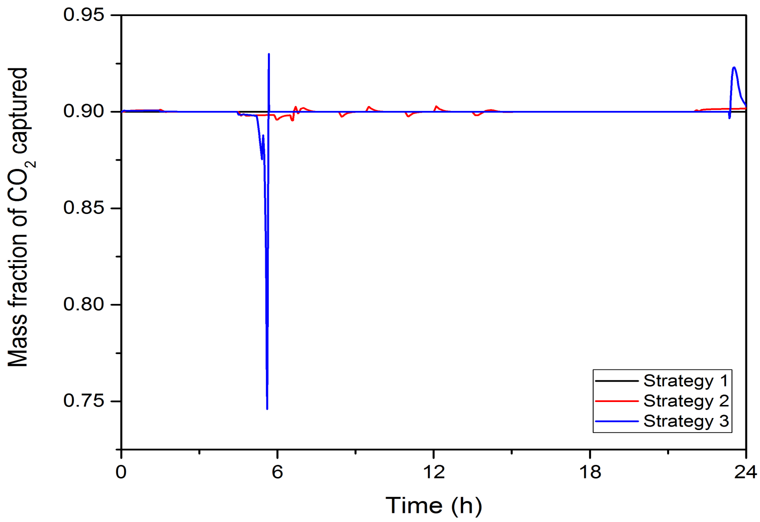

Figure 8: CO_2 conversion using different control strategies for the PC-CP process

The reason for the nature of the behaviour exhibited by Strategy 2 is evident from Figure 9, where the lean solvent flowrate variation is illustrated for the three strategies. 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 10 and Figure 11.

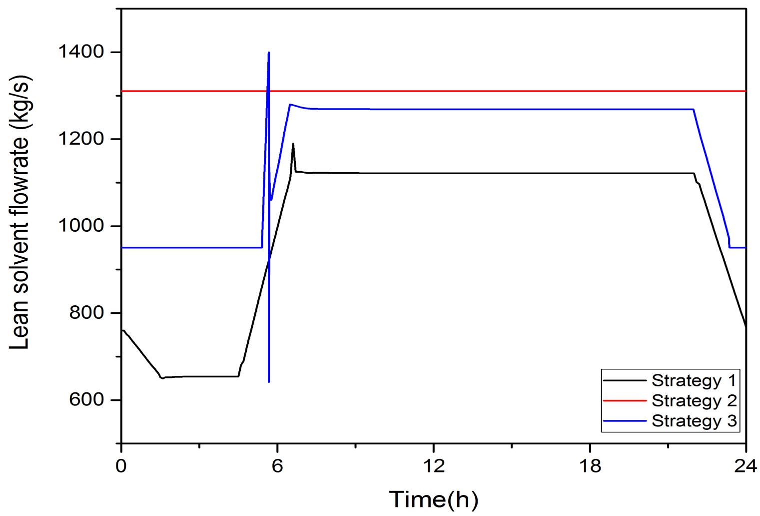

Figure 9: Lean solvent flowrate using different control strategies for the PC-CP process

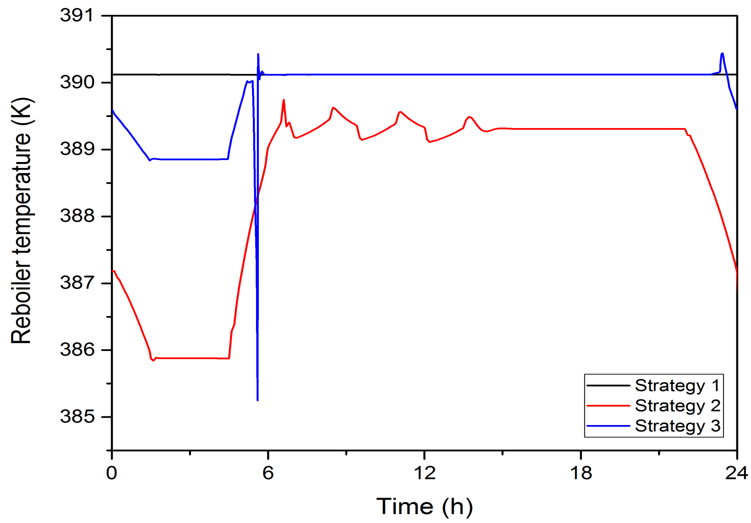

Figure 10: Reboiler temperature using different control strategies for the PC-CP process

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.

The lean loading variation, as illustrated in Figure 11, provides the capture plant to operate with more flexibility, without violating the columns specifications, and leading to conditions of underwetting or entrainment.

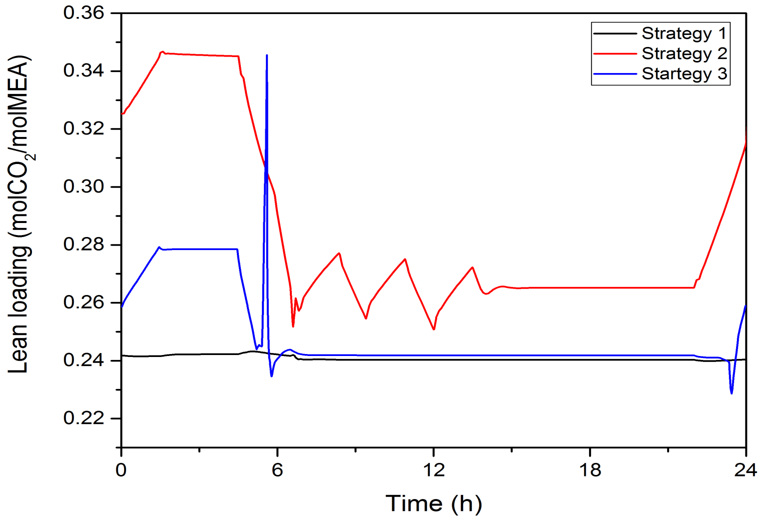

Figure 11: Lean loading variation using different control strategies for the PC-CP process

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.

4.1.2. Economic evaluation

A simple approximation to the profitability of the plant was evaluated.

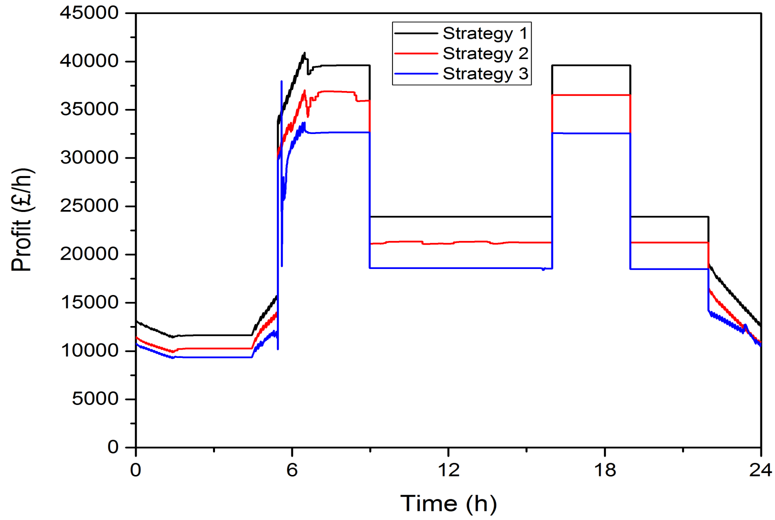

Figure 12: Profit for Pulverised coal power plant for different control strategies

As can be observed from Figure 12, Strategy 1 is more profitable than Strategy 2 due to the lower CO_2 emission cost, even if Strategy 2 had a higher electricity revenue due to the lower steam penalty. Strategy 3 appears to be the least economic option due to the higher CO_2 emission cost. This can be explained in more detail from Figure 13, where the breakdown of costs for the three strategies, is presented. These findings are in accordance with the work by He et al. [24], where they have concluded that there can be an economic advantage related to dynamic optimisation of the process with MPC due to the reduced CO_2 emission cost even if the process energy penalty remains the same as in Strategy 1.

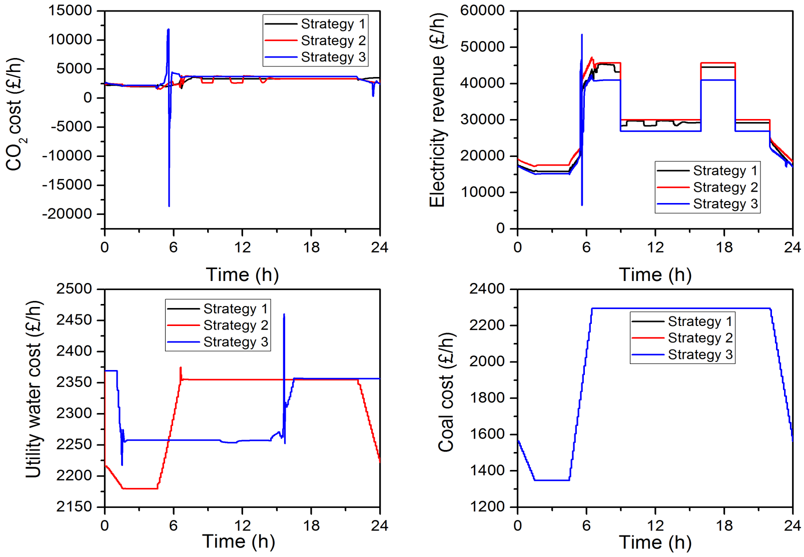

Figure 13: Cost breakdown for Pulverised coal power plant for different control strategies

In Figure 14, we present the cumulative profit attained by the plant for each mode of operation and in Table 9 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 10% compared to Strategy 2 with a total profit of £592k. This can be explained by Figure 13 , where the cost of emitting CO_2 is lower even if the steam loss for Strategy 2 is lower leading to higher electricity revenue.

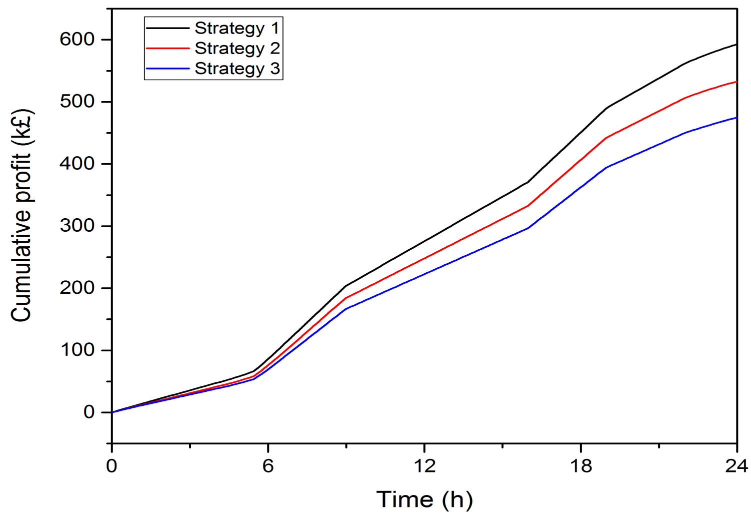

Figure 14: Cumulative profit for PCPP plant with different control strategies.

Table 9: Total profit for the three control strategies

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

4.2. Gas-Fired power plant

4.2.1. Dynamic performance of the different control strategies

The behaviour of the capture plant when integrated with the CCGT plant 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 2 as illustrated in Table 9, in the case of CCGT the three strategies had the same profit as presented in Table 10. It is speculated that this is due to the more dilute CO_2 concentration within the capture process, and therefore the exhaust 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 15.

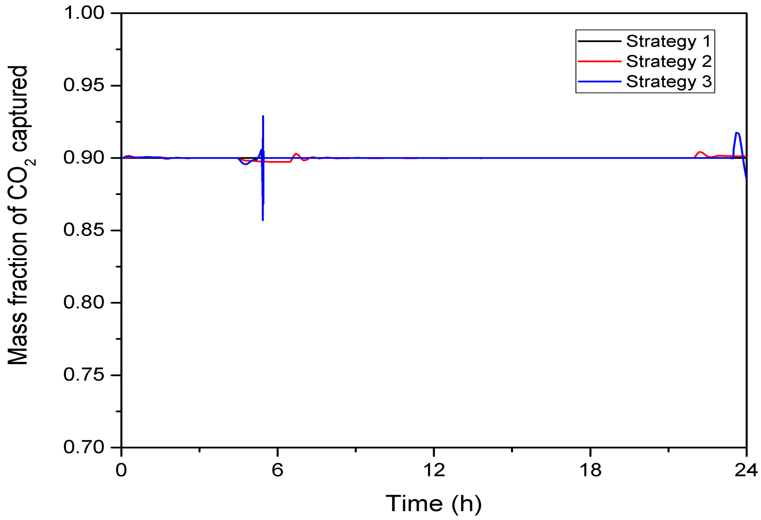

Figure 15: CO_2 conversion using different control strategies for the CCGT process

This deviation of the mass fraction of the CO_2 captured is explained in Figure16, where the steam flowrate for the three strategies, is illustrated. As it can observed, for the first strategy the steam flowrate change in compliance with the disturbance so as the CO_2 captured remains stable at 90 %. For the other two strategies, we observe a fluctuation of the steam flowrate which mirrors the fluctuations of the CO_2 captured and also the temperature in the reboiler, as illustrated in Figure17.

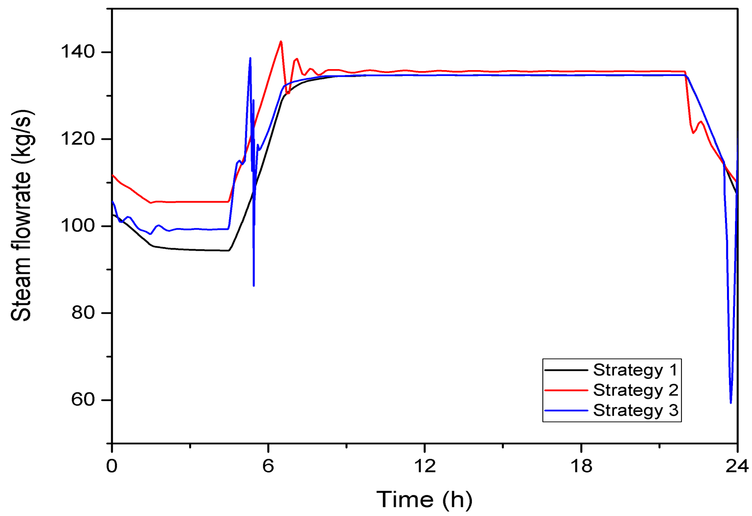

Figure 16: Steam flowrate using different control strategies for the CC-CP process

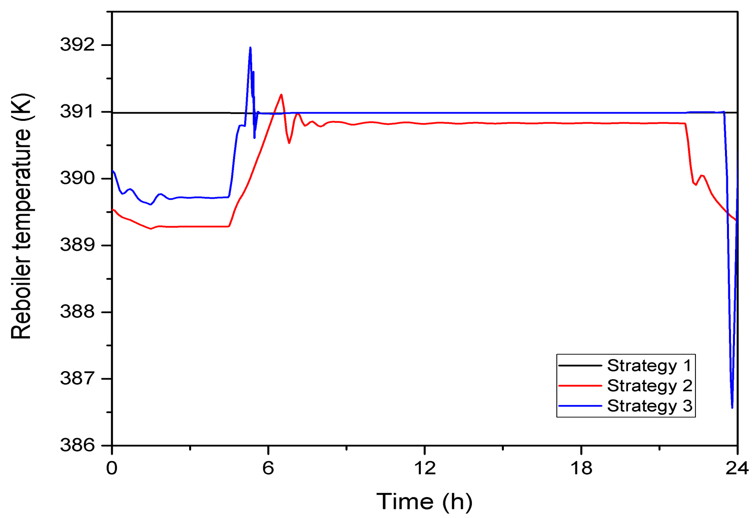

Figure 17: Reboiler temperature using different control strategies for the CC-CP process

4.2.2. 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 18.

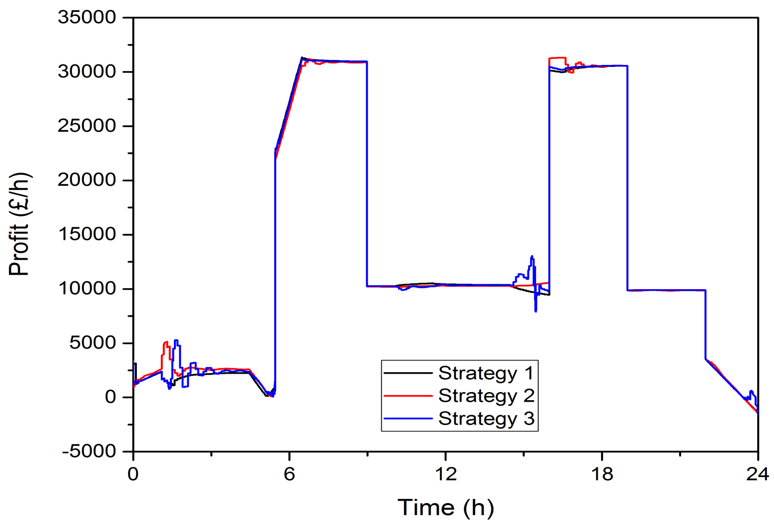

Figure 18: Profit for CCGT plant with different control strategies.

Table 10: Total profit for the three control strategies

	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 emission in this scenario (£$70/t_{CO_2}$). 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.

5. Conclusions

In this study an evaluation of process control strategies for normal, flexible and upset operation conditions of CO_2 post-combustion capture (PCC) processes has been performed. The aim was 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 and higher profit under normal and part-load operating conditions for coal and natural gas power plants. The control structures are developed for power plant operating ranges of around 50% to 100% load. Three different control strategies have been studied. In Strategy 1, the CO_2 conversion which is the main control variable of the plant is manipulated by the lean amine flowrate, while for keeping the reboiler temperature at its set-point the steam flowrate in the reboiler is adjusted. In Strategy 2, the lean amine flowrate is kept constant, while the CO_2 conversion is controlled by the lean loading variation by manipulating the steam flowrate to the reboiler and therefore variating the reboiler temperature. In Strategy 3, there is a switch between the two aforementioned control strategies depending on the loading of the power plant. The main conclusions of this work are:

1. 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.

2. 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.

3. 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.

4. From the economical point of view, 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.

To summarise, although the solvent circulation rate is the best strategy for control in load following operations, the other strategies may be useful during start-up and shut down where there are significant constraints on steam supply, which can be studied in future work.

6. Acknowledgements

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