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Control Strategies for Flexible Operation of Power Plant with CO_2 Capture Plant

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About 20% power output penalties will be incurred for implementing CO_2 capture from power plant. This loss can be partially compensated by flexible operation of capture plant. However, daily large variations of liquid and gas flows may cause operation problems to packed columns. Control schemes were proposed to improve the flexibility of power output without causing substantial hydraulic disturbances in capture plant is presented. Simulations were implemented using ASPEN Plus. In varying lean solvent flow strategy, the flow rate of recycling solvent was manipulated to control the CO_2 capture rate. The liquid flow of the absorber and gas flow of the stripper will vary substantially. In an alternative strategy, the lean solvent loading will be varied. Variation of gas throughput in the stripper is avoided by recycling part of CO_2 vapor to stripper. This strategy provided more stable hydraulics condition in both columns and is recommended for flexible operation. *AIChE J,* 58: 2697–2704, 2012
Keywords: CO_2 scrubbing, process control, plantwide control

Introduction

In recent years, global warming and climate change caused by greenhouse gases have received widespread concern. To mitigate the greenhouse effect, developed countries aim at reducing anthropogenic emissions of CO_2 to 20–40% below 1990 levels by 2020.[1] Majority of the CO_2 emission comes from flue gas emitted from electricity generation, coal-fired power plant especially. The United States emits about 1.9 billion metric tons of CO_2 annually from coal-fired power plants. This constitutes 33% of total energy-related CO_2 emissions and 81% of CO_2 emissions from electricity generation.[2] If the CO_2 emitted from existing coal-fired power plant can be successfully reduced, it will be a great help for reducing CO_2 emissions. Therefore, postcombustion carbon dioxide capture and sequestration (CCS) systems have been recognized and researched as an important strategy for sustainable development. The most mature technology for post combustion CO_2 capture is amine scrubbing using monoethanolamine (MEA) as absorbent.[3] Pilot-scale plants of various sizes have been constructed and operated to investigate the design and operability of such processes.[1]

Implementing CO_2 capture incurs penalties on the electric power output. First, regenerating lean solvent after scrubbing requires a large amount of heat. The thermal energy is usually acquired from low-pressure steam extracted from the power plant. Second, the stripped CO_2 vapor must be compressed to above 100 bar for transportation and sequestration. Reboiler steam and compression work consumed by the capture plant will reduce electricity output of power plant by about 20–30%.[1]

Coal-fired power plant usually performs as a base load power plant that produces a steady electricity output. However, power demands fluctuate on daily and seasonal basis.[4] Electricity has higher prices during peak load periods. It was suggested that electric power output can be increased to meet higher electricity demand by turning off CO_2 capture plant in peak hours[5] Chalmers et al.[6] showed that if CO_2 trading price is included, bypassing CO_2 capture is valuable when \$/MWh electricity selling price is 2–3 times higher than \$/ton CO_2. Thus, flexibility added to the power plant was proclaimed as one of the advantages of postcombustion CCS by amine scrubbing.

However, a continuous process such as the amine scrubbing and regeneration process commonly used in CCS cannot be shut down and turned on at will. There must be limits to flexibility constrained by design, operation and control of the CCS plant. Furthermore, the flexible operation is not only seasonal throughput variations that are commonly in chemical plant seen to meet product requirement but large daily variations of throughput between peak load and off-peak load periods. It needs to be recognized that when flue gas bypasses the CO_2 capture plant, hydraulics conditions of the absorber and the stripper will change substantially. A column will operate normally only when the gas flow velocity and liquid gas ratio are within range.[7] Flooding will occur when the gas velocity is too high. Flooding will impede liquid flow to get down the column lead to high

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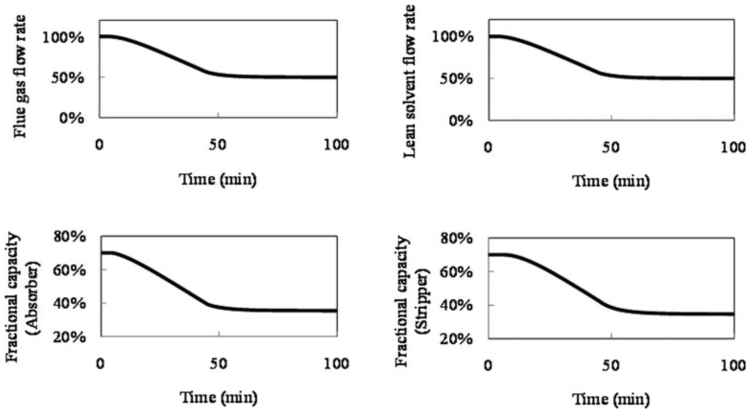

Figure 1. The hydraulic performance of absorber and stripper when 50% of flue gas bypasses CO_2 capture plant.

pressure drop. Poor wetting will occur when the liquid gas ratio is too low that liquid film cannot be maintained on the surface of packing. Both phenomena will result in poor efficiency of column and should be avoided. Tray hydraulics are characterized by two key indices, the reduced vapor velocity and flow parameter. The reduced vapor velocity is a dimensionless gas velocity, and the flow parameter is a normalized ratio of gas and liquid mass flow rates. At a flow parameter, the maximum gas velocity, which depends on the type of trays or packings used, is the flooding velocity. At a given operation, the ratio of the operating gas velocity to the flooding gas velocity is known as the ''fractional capacity.'' Fractional capacity is commonly used as a parameter that indicates hydraulic conditions in packed column. The common practice is to design a column so that the nominal fractional capacity of the plant is about 70–80%. Using the simulation example described in this work, Figure 1 showed that if 50% of the flue gas is bypassed from the capture plant, fractional capacity of the absorber and stripper is reduced to less than 40%. Normal operation cannot be maintained if the throughput is turned down much from designed operating conditions because of poor wetting of packing, which causes poor efficiency and unstable operation.

It was also suggested that rich solvent can be stored during peak load period and can be regenerated later in off-peak period.[8,9] This strategy could avoid CO_2 emission penalty because CO_2 is captured all the time. However, this strategy requires huge additional tanks and solvent inventory for buffering between peak and off-peak load period.[10] The additional cost and safety hazards can be staggering. Using this strategy, a normal gas and liquid throughput can be maintained in the absorber, but large changes in throughput can still be found in the stripper. Ziaii et al.[11] proposed an operating strategy implemented by recycling part of rich solvent to absorber instead of delivering all rich solvent to stripper when reboiler steam is reduced in peak load period. In this method, ''solvent loading,'' which is defined as the mole flow ratio of CO_2 to MEA in the lean solvent coming out from stripper, can be maintained at constant when stripper load changes. The effects on the absorbers due to mixing of regenerated and unregenerated solvents have not been considered. Moreover, the gas and liquid throughput in the stripper will change.

The operability of the plant is also limited by throughput variations in the compressors. Compressor surge will occur if the gas flow rate is too low compared with the design value. To prevent surge, anti-surge control is commonly applied. Anti-surge control is implemented by recycling a portion of flow to maintain the lower limit of gas flow throughput rate. Once anti-surge control is turned on, the energy penalty will not be proportionately reduced when the CO_2 capture rate decreases (Figure 2).

There has been very little literature on the dynamics and control strategies of the power plant with capture unit. Lin et al.[12] studied the plant wide control strategies of a capture plant for maintaining water inventory, minimizing energy as maintaining removal targets. However, only relative modest ($\pm$20%) and long-term disturbances in throughput rate, CO_2 concentrations etc., were considered. Robinson and Luyben[13] proposed a plant-wide control scheme for the operation of methanol plant that was coupled with the Integrated Gasification Combined Cycle (IGCC) plant to give it flexibility between peak and off-peak hours. Regulatory control strategies allowed the methanol plant to turn-down to 20% capacity. However, changes of tray hydraulics of the methanol purification column during transition were not reported.

In this work, we suggest that flexible operation can be achieved by implementation of proper control strategies. The feasibility of this approach is verified using dynamic simulation of an integrated system with power generation and heat recovery sections of a power plant and the corresponding CO_2 capture plant by ASPEN Plus and ASPEN Dynamics. Using this approach, no additional large storage tanks and solvent inventory will be needed, nor will large variations in

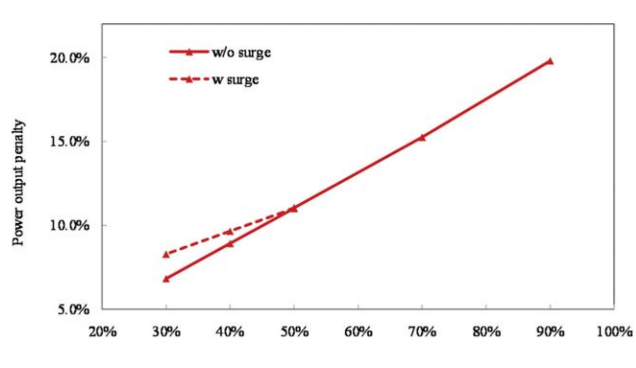

Figure 2. The relationship between CO_2 capture rate and power output penalty with and without compressor surge.

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gas and liquid throughputs be introduced in the absorber and stripper.

Process Description

Power plant

A power plant model is required so that interaction between multistage compressor, turbine output and the CCS capture plant can be simulated. In this work, a reference 580 MW power plant burning bituminous coal with 36.3% net efficiency (Higher Heating Value (HHV))[14] was selected. Boiler supplies high pressure steam at 170 bar/560°C and reheated steam at 38 bar/560°C for power generation. Power is generated by high, intermediate, and low pressure turbines. The low-pressure steam coming out of the low-pressure turbines are cooled into condensate water in the condenser. High pressure, intermediate, and low pressure steams are extracted for preheating the condensate before it reenters the boiler. In this reference plant, steams are seven-staged preheater is applied, including a deaerator and four low-pressure steam preheaters, one intermediate steam preheaters, and one high-pressure steam preheater. Flue gas out from boiler is then sent to flue gas desulfurization process for removing sulfur dioxide. After that, about 2300 ton/h flue gas containing 13 mol % CO_2 will head to CO_2 capture plant. As the plant is regarded as a base load plant, instead of simulating boiler, constant high pressure, and intermediate pressure, steam condition out from the boiler is used as input to the power plant model.

Power extractions from steam are simulated by series of turbines with different outlet pressures. The outlet steam of each turbine is available for preheating condensate or providing heat for the reboiler in the CO_2 capture plant. To facilitate heat transfer, saturate temperature of heating steam should be at least 10°C approach above reboiler temperature.[15] In this power plant, the most suitable extraction point is the steam at 2.9 bar from low-pressure turbine, which saturates at 130°C.

CO_2 capture plant

The CO_2 capture plant includes two columns, absorber and stripper, and one lean/rich solvent cross heat exchanger. Flue gas carrying CO_2 generated from power plant is delivered into bottom of packed absorber to contact with lean solvent, an aqueous solution containing 30 wt % MEA. Treated gas is vented to atmosphere from top of absorber. After absorption, the rich solvent is preheated to 100°C by heat exchanger before being sent to stripper. In the stripper, low-pressure steam from power plant is injected into reboiler for CO_2 desorption. Then, hot lean solvent out from stripper is reused after being cooled to 40°C by heat exchanger and cooler. Hot stripped vapor with CO_2 and H_2O is cooled and compressed.

Steam usage in reboiler relies on extracting steam at 2.9 bar from the low-pressure turbine. About 50% of the low pressure superheated steam over 200°C leaves low-pressure turbine to the reboiler. To avoid amine degradation caused by intensive heat stress in the reboiler, the superheated steam should be cooled to near saturated temperature before injecting to reboiler. After releasing latent heat in the reboiler, reboiler condensate then is sent back to steam-condensate cycle in power plant.

A process model, similar to the one used by Lin et al.,[12] was used for the capture plant. The thermodynamic model used was the electrolyte nonrandom-two liquid property method and CO_2-MEA-H_2O system chemistry that includes the following five equilibrium reactions.

$$RNH_2 + H_3O^+ \leftrightarrow RNH_3^+ + H_2O$$

$$CO_2 + 2H_2O \leftrightarrow H_3O + HCO_3^-$$

$$HCO_3^- + H_2O \leftrightarrow H_3O^+ + CO_3^{2-}$$

$$RNH_2 + HCO_3^- \leftrightarrow RNHCOO^- + H_2O$$

$$2H_2O \leftrightarrow H_3O^+ + OH^-$$

where R− stands for the monoethanol group $CH_2(OH)CH_2-$. Details of this thermodynamic model were reported by Freguia.[16] Five equilibrium stages were assumed for both absorption and stripper.

In this work, absorber and stripper column diameters are sized by gas and liquid flow rate in each column at 90% capture rate. Pressure drop for packing section (16-mm Pall rings) is calculated by Eckert's method[17] in ASPEN. Fractional capacity is taken as 70% so that absorber's diameter will be 18.7 m, and stripper is 11.7 m. In practice, to avoid building such a large column, four parallel sets of absorption/stripping process are suitable applied so that absorber's diameter will reduce to 9.3 m in each set.

Multistage compressor

CO_2 product accompanying with water vapor is at about 100°C as being stripped out from top of stripper. Before being compressed, CO_2 product is cooled to 40°C by overhead condenser and part of water is condensed. Further, the CO_2 is compressed from 2 to 110 bar through a multistage compressor, which includes intercoolers to cool the exhaust gas back to 50°C before entering next stage. Three stages are used in this case and compression ratio is four in each stage. Each stage is simulated by compressor, heat exchanger, and condenser, responsible for compressing, cooling, and knocking water out, respectively.

Heat integration

Several previous studies have investigated the possibility of providing heat by drawing low pressure steam from power plant[15,18,19] and the potential of implementing heat integration.[20-23] Romeo et al.[23] showed that heat integration of intercoolers can save about 2% of electricity output. To implement heat integration, first, heat acquired from cooling CO_2 in overhead condenser and intercoolers of the CCS plant is used to preheat condensate coming out of the condenser of the power plant at 36°C. Part of condensate is delivered to intercoolers to cool CO_2 vapor to 50°C before entering subsequent compressor. Rest of the condensate is delivered to overhead condenser of the CCS plant to recover waste heat. Even though large amount of latent heat is recovered by preheating condensate of the steam cycle, additional cooling water is required to cool the CO_2 vapor before entering first compressor. Figure 3 shows the process flow sheet with heat integration.

Flexible Operation Strategies

In this study, we assume that an average capture rate 70% should be attained and that transition to 50% capture or 90% capture rates can be achieved within 1 h. This would allow a peak electricity output period of about 11 h per day with CO_2 capture rate reduced to 50%. To balance overall capture rate to 70%, CO_2 removal rate has to increase to 90% in the

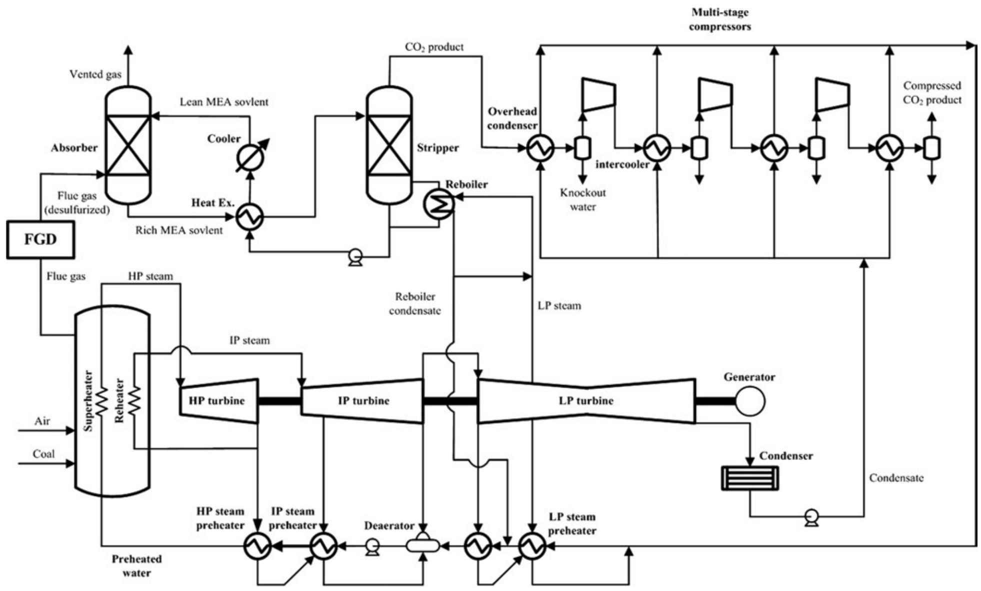

Figure 3. Integrated system flowsheet with steam-condensate cycle, multistage compressor and CO_2 capture plant.

next 11 off-peak load hours with another transition period of 1 h. For more stringent requirement of 90% removal, large variation of flow is still possible if the peak hour is short. For example, a 4-h peak load period, the CO_2 removal rate can be reduced to $\sim$ 45% if an off-peak load of near 100% is possible. The actual optimal flexible operating policy is dependent on the duration of peak loads, price of electricity and penalty for carbon dioxide emission. The purpose of this work is to show that we can adjust the CO_2 capture rate without causing large flow disturbances in the capture plant. Given more detailed information, an optimal control policy can always be implemented as a supervisory scheme on top of the base control scheme suggested in this work.

Variation of lean solvent flow

For an amine-based CO_2 capture plant, the plant-wide control strategy proposed by Lin et al.[2] pointed out that CO_2 capture rate can be adjusted to meet flexible operation as Figure 4 shows CO_2 capture rate is controlled by variation of lean solvent flow (VLSF), and reboiler temperature is controlled by manipulating reboiler steam flow rate. In VLSF control structure, flexible operation can be implemented by adjusting set-point of CO_2 capture rate controller. Lean solvent circulating rate is varied to meet the capture rate target. In the VLSF control scheme, reboiler temperature is controlled at a fixed value. Reboiler temperature is an indicator of lean loading. Hence, the residual loading of CO_2 in the recirculating solvent is approximately constant during flexible operation.

The VLSF strategy delivers all the flue gas into absorber. In this way, variation of gas flow rate in absorber is avoided when flexible capture targets are pursued. However, liquid flow rate will vary substantially in the absorber since the capture rate target is achieved by changing the solvent flow rate. Furthermore, because the net amount of CO_2 captured and stripped from the stripper will change, liquid flow in the stripper will also vary substantially.

Variation of lean solvent loading (VLSL)

To avoid the potential fluctuations in liquid flow in the absorber and gas and liquid flow in stripper, we propose an alternative control strategy that stabilizes the hydraulic conditions of both columns during flexible operation.

First, if the circulating lean solvent rate is fixed, lean solvent loading can be used to meet different CO_2 capture rate. The lean solvent loading can be reduced so that more CO_2

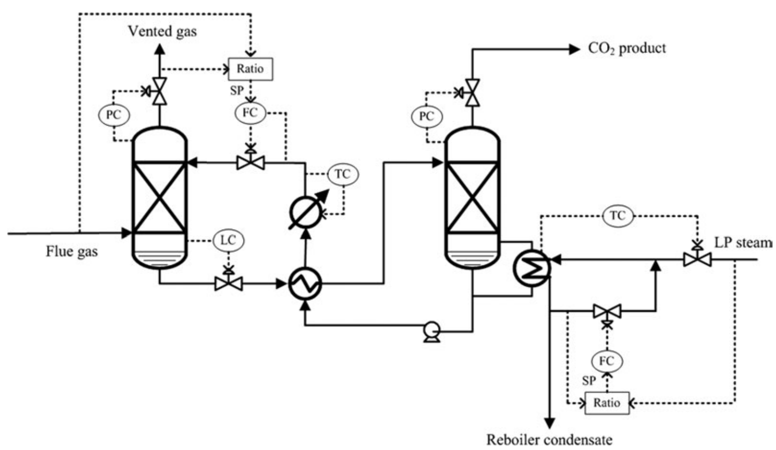

Figure 4. Control structure of CO_2 capture plant in variation of lean solvent flow strategy.

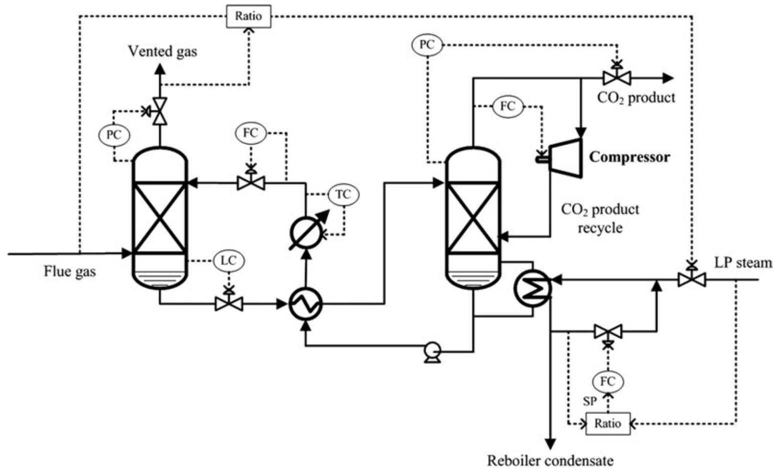

Figure 5. Control structure of CO_2 capture plant in variation of lean solvent loading strategy.

can be captured in the absorber with a steady lean solvent flow. Conversely, if we wish to reduce CO_2 capture rate, a higher lean solvent loading can be allowed, thus reducing the load of the reboiler. Hence, the scheme is based on VLSL.

By changing the loading of the lean solvent, the gas and liquid flow in the absorber and liquid flow will be stabilized. However, the gas flow rate in stripper will change as the quantity of captured CO_2 produced at the top of the stripper changes. Hence, we propose to recycle part of CO_2 product vapor to bottom of stripper so that gas flow instability in stripper can be avoided by adjusting the recycle rate back to stripper. The recycle of CO_2 product has little effect compared with the case without recycling. If 30% of CO_2 product is recycled, mole fraction of CO_2 in vapor increases from 0.476 to 0.480 at the bottom. The composition of CO_2 vapor at top of stripper and the reboiler duty required are almost unchanged.

The control scheme of VLSL is shown in Figure 5. Lean solvent flow rate is controlled at a given value by a flow controller. CO_2 capture rate is controlled by manipulating reboiler steam flow rate.

Dynamic Simulation

To understand dynamic behaviors while implementing flexible operation, the integrated system modeled in ASPEN Plus is exported to ASPEN Dynamics and then simulated dynamically.

In this work, only equilibrium model can be applied due to the restrictions in dynamic simulation in ASPEN. However, Peng et al.[24] have compared dynamic simulation results of equilibrium and rate-based model. Although there were some differences in steady-state value, dynamic responses were similar. It is not our purpose to match plant performance with simulations but to investigate the variation of tray-hydraulics, which is largely determined by liquid and gas

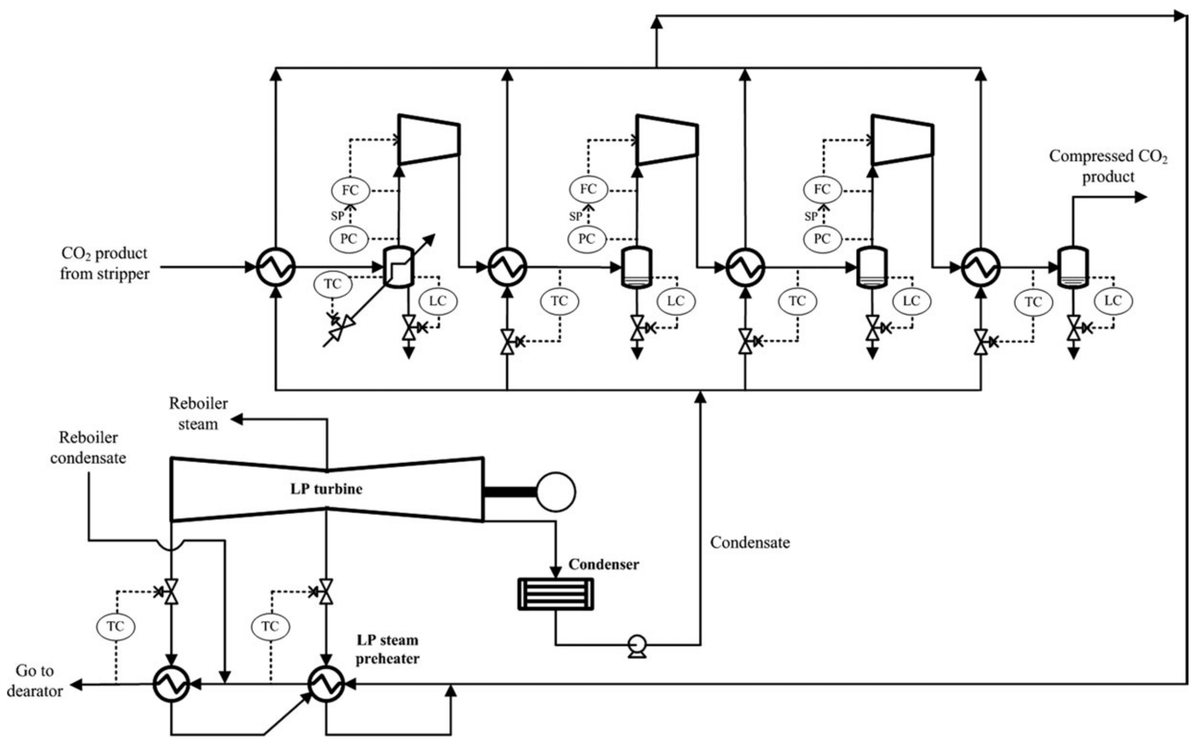

Figure 6. Control scheme of multistage compressor and steam cycle.

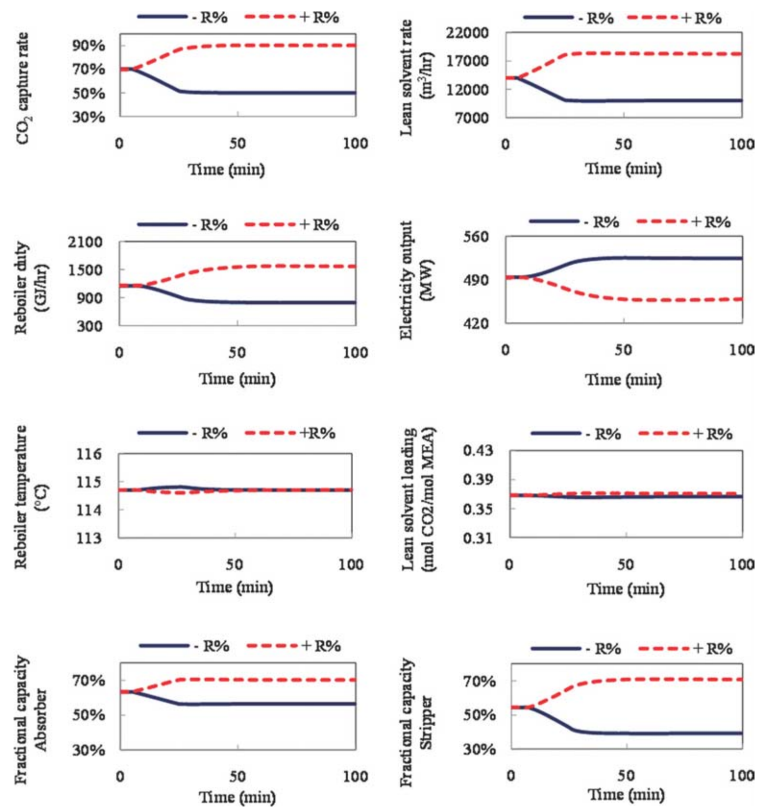

Figure 7. Dynamic responses of flexible operation adjusting capture rate from 70% to 50% and 70% to 90% implemented by VLSF strategy.

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flow rates, on transition between operating states. Dynamic simulation based on equilibrium stage model can be adequately applied.

After being exported to ASPEN Dynamics, basic controllers that maintain steady operation are installed. There are several pressure controllers and level controllers in columns and vessels. Figure 6 shows control scheme of multistage compressor and steam cycle. Varying speed control method is applied in compressors' control to meet correct gas flow rate. To implement heat integration between intercoolers when CO_2 vapor flow rate is changing, condensate should be adequately distributed to overhead condenser and intercoolers. So, temperature controllers are installed to manipulate condensate flow rate to each intercooler and rest condensate is sent to overhead condenser. After the intercoolers, condensate is back to steam cycle and preheated by low pressure steam to 140°C before going to deaerator. All Proportional Integral Derivative (PID) controllers' parameters are obtained from tuning tools in ASPEN Dynamics applied Ziegler-Nichols method.

A base case with 70% capture is used to demonstrate flexible operation decreasing capture rate to 90% and decreasing to 50% in two operating strategies. The base case is running at 70% CO_2 capture rate with 14,000 m^3/h lean solvent flow rate and 1160 GJ/h reboiler duty. Lean loading is at 0.37 mol CO_2/mol MEA, corresponding to reboiler temperature at 114.6°C. Fractional capacity is 63% in absorber and 54% stripper due to less liquid and vapor rate compared with 90% capture rate case.

Results and Discussions

Results of variation of lean solvent flow

To demonstrate operability, set-point of capture controller is changed in ramp rate of 1% capture rate/min. Set-point given is from 70% to 50% or 90% in 20 min to increase or decrease power output.

Figure 7 shows dynamic responses of flexible operation implemented by VLSF strategy. CO_2 capture rate starts to change at 5th min. When CO_2 capture rate is changing, lean solvent rate is manipulated to track correct capture rate. To cope with the changing circulating solvent rate, reboiler steam is manipulated to maintain reboiler temperature. We can see that lean solvent loading is keeping nearly at 0.37 mol CO_2/mol MEA.

Table 1. System Conditions After Flexile Operation Implemented by Two Strategies

	50%			90%		
	Rich Solvent T (°C)	Reboiler Duty (GJ/h)	Power Output (MW)	Rich Solvent T (°C)	Reboiler Duty (GJ/h)	Power Output (MW)
VLSF	109.7	795	523	107.8	1559	459
VLSL	104.3	861	518	112.4	1502	463

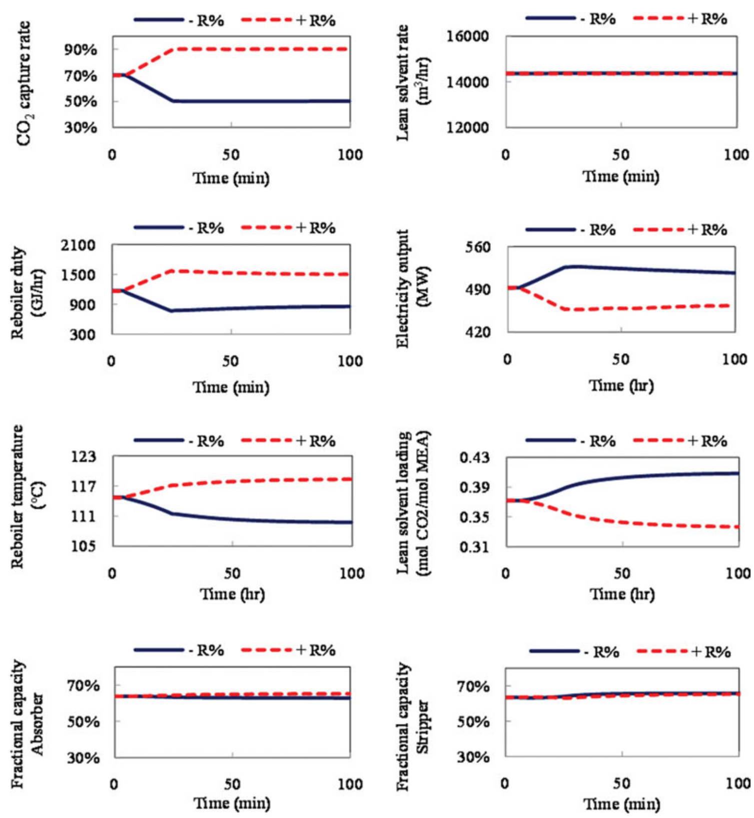

Figure 8. Dynamic responses of flexible operation adjusting capture rate from 70% to 50% and 70% to 90% implemented by VLSL strategy.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

Using this operation strategy, fractional capacity in absorber varied in a relatively smaller range compared with simple bypass, between 55% in peak hours and 70% in off-peak hours. However, the fluctuation in stripper still exists. Fractional capacity decreases to below 40% when CO_2 capture rate is reduced to 50%. New steady-state value of reboiler duty and power output are listed in Table 1. Power output increases to 523 MW in peak load period and decreases to 459 MW in off-peak load period.

Results of variation of lean solvent loading

Figure 8 shows dynamic responses of flexible operation implemented by VLSL strategy. In this operating strategy, system successfully attains CO_2 capture targets of 50% and 90%. Lean solvent flow rate is fixed at 14,000 m^3/h. The reboiler temperature was increased to 118°C to meet a higher CO_2 capture rate target of 90% and reduced to 110°C to meet a lower CO_2 capture rate target of 50%. Lean solvent loadings also change from 0.37 to 0.41 and 0.33 mol CO_2/mol MEA at 50% and 90% capture rate, respectively.

When comparing new steady-state value of reboiler duty obtained by two operating strategies, VLSL has slightly lower energy requirement at 90% capture and slightly higher at 50% capture rate. The main reason contributing the difference is the temperature of rich solvent before entering the stripper. In VLSL, when reboiler temperature increases to meet 90% capture rate in off-peak load period, it also heats rich solvent to a higher temperature in lean/rich cross heat exchanger so that less reboiler duty is required. Conversely, at 50% capture rate, more reboiler duty is required due to lower rich solvent temperature. However, the difference of power output between two strategies is relatively small and the difference will be offset when operating hours of peak load and off-peak load are close.

In VLSL strategy, 30% of CO_2 product is recycled to the stripper initially and fractional capacity increases to 63%. Recycle rate is manipulated to maintain the gas flow rate out from top of stripper constant. By stabilizing the throughput of absorber and stripper in constant value, we can see that the fractional capacities in both columns are almost unchanged. Fluctuations in both columns due to large variations of liquid and vapor rate during flexible operation are avoided in this control strategy.

It is interesting to note, in Table 1, that the VLSL strategy actually produces more power, 463 MW, during the off peak hours compared with 459 MW for VLSF strategy. However, the net power output (518 MW) is inferior to that of VLSF strategy (523 MW). This is because the lean solvent loading chosen is preferable for 90% capture. The differences in power output are about 1%. Substantial changes in fractional capacity can still be observed using the VLSF strategy, whereas such variations are negligible in VLSL strategy. Therefore, from the viewpoint of flexible operation, VLSL strategy is recommended.

Conclusions

CO_2 capture incurs about 20% penalty of electricity output to a fossil fuel based power plant. However, certain degree of flexibility is also built in by manipulating the target of CO_2 capture in peak-load and off peak-load periods. However, large changes in hydraulic conditions absorber and stripper will occur if the amount of flue gas entering the capture plant and the amount of circulating solvents change substantially. Such changes are not desirable from the operation point of view. To implement flexible operation but avoid potential fluctuations in packed columns due to large variation of liquid and gas flow rate, two operating strategy are proposed. In VLSF strategy, instability in absorber is partly reduced by delivering all flue gas to absorber. Capture rate is controlled by manipulating lean solvent rate and reboiler temperature is controlled at constant to maintain a nominal lean loading. In VLSL strategy, lean circulating solvent rate is unchanged to reduce fluctuations in both packed columns. Further, part of CO_2 product is recycled to stabilize stripper's operation. This strategy will be more preferable because it is able to maintain stable hydraulic conditions in both the absorber and stripper during both peak and off-peak load hours.

Acknowledgments

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