

Selection and design of post-combustion CO₂ capture process for 600 MW natural gas fueled thermal power plant based on operability

Rohan Dutta, Lars O. Nord, Olav Bolland

Department of Energy and Process Engineering, NTNU-Norwegian University of Science and Technology, Trondheim, Norway

Abstract

Post-combustion CO₂ capture (PCC) plant for a typical 600 MW natural gas fueled thermal power plant was designed as a trade-off between operability and mitigation of the efficiency penalty. Two modified PCC plant configurations with low efficiency penalty were selected. The methodology for designing PCC plants was adapted by incorporating design constraints based on operability and the construction of absorbers. This was applied in sizing the equipment of the plants. Two configurations of absorbers were analyzed based on flue gas flow rate at full-load condition and at time-average of an assumed load variation of a power plant operating flexibly. It was found that the absorber designed at time-average load provided a reduction of approximately 4% in the purchased cost of absorbers. The performance of the designed plants under power plant load variation, flow maldistribution and variable capture ratio was analyzed using off-design condition simulations. The absorber designed at full-load condition was found to lead to lower reboiler duty in order to maintain a similar capture rate to that of the other absorber during part-load operation. Dynamic simulations of the plants with the existing control structure were performed under similar power plant load variations to confirm their operability, and suggestions for selecting one of them were presented.

Keywords Design and rating; Post-combustion CO₂ capture; MEA; Dynamic simulation

1. Introduction

Global warming is evident from the current weather statistics [1]. Initiatives have been taken in order to reduce the emission of CO₂, mostly from large point sources such as fossil fuel based thermal power plants, to mitigate the effects on climate change [1, 2].

One of the methods of CO₂ capture from large point sources, such as fossil fuel based power plants, is the post-combustion CO₂ capture (PCC) process, using chemical absorption by amines [1]. This process is similar to that which has been commercially used for CO₂ recovery from natural gas treating facilities and for CO₂ production for fertilizer industries on a small and medium scale [3]. As those plants are dedicated to CO₂ production, the energy requirements to regenerate amines are not commonly a factor in the commercial implementation of the technology. There are a few disadvantages in using alkanolamines as they are degradable, require very high energy for regeneration, become corrosive in the presence of oxygen and high vapor pressure, leading to solvent loss in the process [4]. However, for the sake of process efficiency, reduced corrosion and reduced cost of production of CO₂, such plants are constantly being improved via the development of novel solvents or process modifications [3]. Similar approaches have been taken by researchers for power plants with PCC through the development of novel solvents and the optimization of plant configuration and process conditions [5]. A number of novel solvents has been developed to reduce reboiler duty, and researchers have proposed modifications in PCC configurations by addressing sources of irreversibilities in the cycle [6-8], by increasing the CO₂ content in the flue gas [9], and by improving the performance of both absorbers and regenerators [10]. Being one of the most mature

1 technologies, with known scale-up and design issues and having the capability to be retrofitted with
2 existing power plants, PCC processes are being preferred as near-commercial technology for capture
3 processes and are already being implemented in the commercial coal-based power plant of SaskPower at
4 Boundary Dam in Canada [11].

5 The use of capture processes integrated with power plants, which themselves are thermodynamically quite
6 inefficient, further reduces the overall power plant efficiencies. It has been reported that the inclusion of
7 PCC processes in the power plant might lead to a reduction in the overall power generation efficiency by
8 8-10%-points for the case of natural gas based power plants [12]. This is mostly due to steam extraction
9 from the power plants, along with CO₂ compressor duty, cooling water pumps and auxiliary power [13,
10 14]. Although the incorporation of modifications in PCC plants may mitigate this reduction to some
11 extent, the inclusion of such modifications in the plant increases the complexity of the overall cycle. For
12 that reason, the operability of such plants needs to be considered before deciding on the configuration. In
13 addition, a control structure needs to be designed for such modified plants during flexible, upset
14 conditions of the power plants [15]. A few such modifications in PCC plants, such as split-stream and
15 vapor recompression, have been analyzed in dynamics to investigate their operability, and they have been
16 found to be capable of handling disturbances in a certain range of variation (10%) [16]. *Therefore, it is*
17 *important to analyze the operability of modified plants in respect of different disturbances that can be*
18 *envisaged for futuristic power plants. As a trade-off between operability and the mitigation of power*
19 *generation efficiency reduction, any such modified configurations in PCC plants may be selected.* Upon
20 investigation of the trade-offs, this paper presents the design and operation of a suitable PCC process for a
21 MW-class natural gas based power plant.

22 The objective of this study is to *select and design a suitable PCC process for a 600 MW natural gas based*
23 *power plant.* In order to achieve the objective, the following tasks were performed.

- 24 1. Identification of optimum process conditions for the power plant with PCC.
- 25 2. Adaptation of design methodology for PCC plants by applying design constraints based on
26 operability and the construction of absorbers/regenerators, and by utilizing the methodology for
27 designing suitable PCC processes, after identifying the designed flue gas flow rate for power
28 plants operating flexibly.
- 29 3. Investigation of the operability of the designed plants with existing control structure during
30 assumed power plant load variation using dynamic simulation.
- 31 4. Trade-off between operability and the efficiency penalty to select a specific modified
32 configuration.

1 **2. Methodology**

2 **2.1 Process simulator and solution methodology**

3 The power plant was analyzed at design and off-design conditions using GT PRO[®] and GT MASTER[®] by
4 Thermoflow Inc. to identify its part-load characteristics. For steady state simulation of the PCC plants,
5 Aspen HYSYS[®] V8.6 was used. At steady state simulation, either equilibrium-based or rate-based models
6 have been used [17, 18]. As it was found that rate-based models were more accurate than their
7 counterparts, during this study, the rate-based models in HYSYS were used. The acid gas property
8 package was employed in the simulator to calculate the thermodynamic properties of all the streams. In
9 steady state simulations, the absorber and regenerator models are rate-based, using any of the methods
10 proposed by Bravo et al. (1985,1992) [19], Billet and Schultes (1993) [20], or Hanley and Chen (2012)
11 [21] for the calculation of the interface area and mass transfer for structured packing. The method
12 proposed by Bravo et al. (1992) was used in this study. The following assumptions were made, while
13 building the steady state flow sheet in the process simulator.

- 14 1. No piping was considered. Pressure drops were added with the neighboring equipment such as
15 heat exchangers, valves, etc. The contribution of the thermal mass of piping in the overall thermal
16 inertia was ignored.
- 17 2. The absorber sump was not considered.
- 18 3. An equal distribution of lean amine and flue gas in the absorbers was considered. Maldistribution
19 in liquid or gas was ignored in both absorbers and regenerators.
- 20 4. A minimum temperature approach of 5⁰C was considered in the rich/lean heat exchanger.
- 21 5. Pumps were considered to have an adiabatic efficiency of 75%.
- 22 6. Outlet gas pressures in both the absorber and condenser of the regenerator were considered as at
23 atmospheric condition.
- 24 7. The regenerator was considered to be working at 2.1 bar pressure.
- 25 8. No water-wash section was considered. Water balance in the absorbers in the cycles was obtained
26 by adding make-up water.
- 27 9. The presence of argon in flue gas was ignored.

28 For the dynamic simulations, equilibrium-based absorber/regenerator models in UniSim Design[®] R430
29 were used with reduced stage efficiency for improved accuracy [22, 23]. Reduced stage efficiency was
30 estimated by comparing the dynamic simulation results with those of the steady state simulation in Aspen
31 HYSYS. Detailed verification of dynamic modeling and validation of the simulation results of both the
32 processes simulators have been presented in [22, 24]. As rate-based column models predict the
33 performance of such equipment most accurately [25], it is necessary to mention the deviation achieved

1 while using equilibrium-based models. Efforts were made to discover such deviations, and they are
2 presented in corresponding sections later in this paper. A built-in amine package in the simulator was used
3 for calculation of the property data. In building the flow sheet for dynamic simulation, the following
4 assumptions were made.

- 5 1. Piping were ignored as in the steady state simulations.
- 6 2. The presence of argon in flue gas was ignored.
- 7 3. Changes in flue gas flow rate and composition were considered as instantaneous with the change
8 in power plant load. The relatively low dead time involved in the gas turbine was ignored.
- 9 4. No water-wash section was considered. Water balance in the absorbers was obtained by adding
10 make-up water.
- 11 5. Constant lean loading at the inlet of the absorber was considered.
- 12 6. Steam extracted always at 150⁰C and 3.5 bar.
- 13 7. Treated gas from the absorber and separated CO₂ in the regenerator were considered to be at
14 atmospheric pressure.
- 15 8. Pumps adiabatic efficiencies were considered to be constant, allowing the control of the work
16 requirement to vary instead of using performance curves.
- 17 9. Constant flow fractions, equal to full-load condition, were used in modified cycles, even during
18 part-load, ramp-up and ramp-down operations.

19 The implicit Euler method was used as a numerical method with a time step size of 0.1 s. Results of the
20 simulation for selected parameters were recorded after every 20 s. Additionally, as the flow sheet needed
21 to be initialized properly to avoid any divergence at the start of the simulation, initialization was
22 performed by specifying each inlet stream, outlet stream and vessel wall of all the equipment at
23 atmospheric conditions. Pressure specifications in boundary streams of the flow sheet and flow
24 specifications at the intermediate streams in the regenerator were used. Solutions obtained from steady
25 state simulations for streams at steady state conditions were used as the initial values during the dynamic
26 simulations.

27 **2.2 Assumed power plant load and flue gas characteristics**

28 Due to the high penetration of variable renewable energy sources, thermal power plants may operate in a
29 cyclic mode, considering the demand and generation variability [26]. Therefore, the operational scenario
30 of futuristic power plants will depend on the electricity price and demand. The assumed scenario of the
31 variation of daily power plant load, depending upon the electricity price and demand as presented by [25],

1 was considered here to analyze the performance of the designed PCC plants using the dynamic models;
2 refer to Figure 1.

3 The power plant was considered to be working at full-load condition for 10 hours; ramp-down started with
4 a rate of reduction in power generation of 7.5% load change per hour for three hours. For the next three
5 hours, the plant worked with a load of 70%, following a two-hour ramp-up operation, during which the
6 rate was increased by 15% every hour. After the plant had reached full load, it operated at full load for
7 another five hours. Flue gas flow rate variation was estimated according to the variation presented later. It
8 is noteworthy that, while analyzing the performance of the designed PCC plants in dynamic simulation,
9 the ambient temperature was considered at 15⁰C.

10 **2.3 Methods for estimating equipment purchase cost and efficiency penalty in power plants**

11 Although a thorough capital cost analysis was not performed, only the purchase cost of the equipment was
12 calculated based on the capacity of the equipment. The six-tenths rule was used to estimate the percentage
13 increase in purchase cost. The relationship between the purchase cost and equipment capacity is [27]:

$$14 \quad \frac{C_a}{C_b} = \left(\frac{A_a}{A_b} \right)^n \quad (1)$$

15 where C is purchase cost, A is equipment cost attribute, and here it is volume of equipment, and n is the
16 cost exponent (0.6). Subscripts a and b refer to equipment with the required attribute and the base
17 attribute, respectively.

18 The efficiency penalty in the power plant due to the inclusion of PCC processes was calculated using the
19 following method for estimating the efficiency of the power plant with PCC, discussed in [28].

$$20 \quad \% - \text{penalty} = \left(\frac{E_{AUX} C}{LHV} + \frac{E_R \alpha C_f}{LHV} + \frac{E_{CMP} C_f}{LHV} \right) \quad (2)$$

21 where E_{AUX} is the specific energy requirement for pumps and blower. The power requirement in
22 circulation pumps was found to vary in a negligible range and was therefore considered constant unless
23 otherwise stated. The blower power requirement was calculated considering an efficiency of 75% and a
24 compression ratio of 1.05. The inlet temperature of the blower was considered constant, even during part-
25 load conditions. The specific CO₂ compressor power required (E_{CMP}) was considered to be 0.33 MJ/kg
26 CO₂, as found in [12]. A constant lower heating value (LHV) of 9 MJ/kg of fuel was used [28]. It was
27 assumed that the steam from the power plant was extracted at 3.5 bar and 150⁰C during both full-load and

1 part-load operations. Therefore, the ratio of incremental power reduction to incremental heat output (α)
2 was obtained from [28] as 0.24. During the operation of the power plant with PCC, under part-load
3 conditions, α was considered as constant. f is the %-capture rate and C is the ratio between formed CO₂
4 and fuel.

5 **3. Power plant and PCC plant configurations**

6 **3.1 Power plant specifications and flue gas conditions during part load**

7 The natural gas based power plant considered in this study, as shown in Figure 2, has a rated capacity and
8 efficiency of 592 MW and 56% on a LHV basis. The plant is a combined cycle, having 1×1 configuration
9 with air-cooled gas turbine. GT-PRO[®] was used to identify the required process conditions for designing
10 the PCC plant for it. Table 1, below, presents the flue gas conditions obtained for different ambient
11 temperatures (0°C, 15°C and 30°C) and generated power at full load at those ambient temperatures with
12 steam extraction at the conditions previously mentioned.

13 The power plant was analyzed during part-load conditions, down to 50% load condition, in GT-
14 MASTER[®] to identify the variation in flue gas flow rate and its CO₂ content. The results of the analyses
15 are presented in Figure 3. These results were used as input to the PCC processes, while analyzing them for
16 their operability under power plant load variations. It may be observed that the flue gas flow rate varies
17 linearly with changes in the power plant load. With a 50% reduction in the power plant load, the flue gas
18 flow rate decreases by 34-40% from the value at full-load condition. Also, with subsequent increases in
19 ambient temperature, the power plant's capacity reduces drastically, along with the flue gas flow rate.

20 In addition, the CO₂ mole-fraction was also found to reduce with the decreasing power plant load. A 10%
21 reduction in the mole-fraction of CO₂ in the flue gas can be observed at 50% load in the power plant. This
22 was envisaged as having a significant impact on the liquid to gas ratio and the reboiler duty of the PCC
23 plant. Therefore, while analyzing the operability of the designed plants, this reduction in CO₂ mole-
24 fraction was considered, together with a variation in ambient conditions.

25 Two more important parameters, flue gas pressure and temperature, are also significant in the design of
26 PCC plants. However, varying the coolant flow rate at the direct contact cooler, it is possible to maintain
27 the inlet temperature to the absorbers without greatly affecting the overall power generation efficiency.

28 The flue gas pressure has an impact on work requirement at the blower and thereby contributes to the
29 efficiency penalty of the integrated power plant. This pressure is the sum of exhaust pressure loss in the
30 GT and the ambient pressure. As a constant ambient pressure was considered in this paper, the change in
31 exhaust pressure loss in the GT and stack temperature were determined in order to identify their range of

1 variation, as shown in Figure 4. A reduction in exhaust pressure loss to around 15 millibar can be
2 observed from the figure for a 50% change in load condition. The stack temperature also reduces by 12⁰C
3 at similar load conditions. This can significantly change the cooling duty in the direct contact cooler.
4 While analyzing the PCC plant during part-load operation, these variations were considered in
5 determining the efficiency penalty due to the inclusion of the PCC plant. It is noteworthy that the gas
6 turbine exhibits a comparatively fast response to changes in fuel and air flow rate; therefore, the variation
7 in flue gas flow rate and composition follows the trends found in off-design simulation with a certain ramp
8 rate, which is similar to dynamic simulation except with small dead time. This dead time is in the range of
9 a few minutes and was ignored in the current work.

10 **3.2 Identification of process conditions and control structure for PCC plant**

11 Typical PCC cycle configurations comprise of amine based absorbers and regenerator. Figure 5 shows the
12 cycle configuration. One of the important issues in designing PCC plants is the identification of process
13 conditions such as:

- 14 1. CO₂ capture rate,
- 15 2. lean/rich loading,
- 16 3. amine weight-% in lean liquid,
- 17 4. liquid to gas ratio,
- 18 5. absorber/regenerator operating pressures, temperatures,
- 19 6. rich/lean heat exchanger minimum approach.

20 The most important parameter among all those listed above is the CO₂ capture rate. Dictated by future
21 emission regulations, CO₂ tax, trading values, etc., this rate may not be constant throughout the year or
22 even for a month or a week. Based on the operational policy of PCC plants, commercial requirements,
23 etc., suitable economic considerations will determine the rate of capture required for a particular period of
24 operation of the plants. Therefore, to determine the required rate of CO₂ capture, when designing the PCC
25 plant, there is a need for suitable techno-economic studies with a consideration of future emission
26 regulations. As this is beyond the scope of this paper, a constant capture rate of 90% was considered to be
27 the requirement for the PCC plants. It was also envisaged at this point that changing the rate of capture
28 might change the process conditions selected, based on this assumption of the capture rate.

29 The rest of the parameters are dependent on the capture rate as well as correlated to each other in such a
30 way that changing one may: 1. increase the required specific reboiler duty in the regenerator, thereby
31 leading to a higher energy penalty in the overall plant; 2. increase the size of all the equipment in the
32 plant, leading to increased capital cost. Adding to these, as power plants may work under flexible load

1 conditions, operation under variable flue gas and CO₂ content may lead to deviation from the optimum
2 conditions of some process parameters such as:

- 3 1. operating pressure of the absorber;
- 4 2. overall heat transfer coefficient in the rich/lean heat exchanger;
- 5 3. reboiler temperature, leading to further variation in the energy penalty.

6 Therefore, when deciding on these parameters, they should also be least sensitive on and around the
7 selected values.

8 According to an early study by Abu-Zahra et al. [29], the optimum lean loading has been found to be 0.3
9 using 40 wt.-% MEA solution. In a recent study by Alhajaj et al. [30], this value has also been reported to
10 be optimum at 0.31 with 30.4 wt.-% MEA. In another study, with 30 wt.-% MEA, the minimum reboiler
11 duty has been found to be at lean loading of 0.25 for flue gas with 3% CO₂ mole-fraction for 85% capture
12 rate [31]. In most of the recent studies, this value has been considered as 0.2 to 0.23 with 30 wt.-% MEA
13 [12, 32-34].

14 As the use of higher than 30 wt.-% MEA entails the disadvantage of corrosion, the use of lower than 30
15 wt.-% MEA has been recommended [35]. Therefore, in this work, 30 wt.-% MEA was used as solvent.
16 Rich loading has been found to be optimum below 0.5 [36]. It was found that, with lean loading of 0.22,
17 the minimum reboiler duty was 3.77 MJ/kg CO₂, which was 8% higher than the minimum possible
18 reboiler duty of 3.5 MJ/kg CO₂ with 30 wt.-% MEA and rich loading also found to be 0.47, which was
19 within the limit [29]. Therefore, 0.22 was used as lean loading. Optimal lean loading may vary for
20 different designs and configurations as well as for different power plant loads. However, in order to
21 maintain uniform conditions for all the plants, constant lean loading for all the cycle configurations during
22 all the power plant load conditions was used.

23 Upon deciding the lean amine condition, the liquid to gas ratio was calculated using the scale-up equation
24 given by [37] and manipulated for 90% CO₂ capture rate. The mass-basis liquid to gas ratio was found to
25 be 1.08, and in molar-basis it was 1.28.

26 Another important process parameter in optimizing plant conditions is regenerator operating pressure.
27 Using a single regenerator and one pressure configuration, the optimum operating pressure has been found
28 to be 2.1 bar [12, 38]. The absorber was considered to be working near ambient condition plus required
29 pressure drop in packing. The lean/rich heat exchanger optimum minimum approach has been found to be
30 5⁰C, and this was used in our study.

1 The process conditions adopted for this study are shown in Table 2.

2 The control structure for the adopted process was obtained from the literature. Reboiler pressure and
3 temperature, and condenser pressure and temperature, were required to control for maintaining the
4 operating pressure of the regenerator, the CO₂ mole fraction in the recovered stream to compression and
5 the lean loading at the outlet from the regenerator. The reboiler and condenser liquid level was controlled
6 to maintain the operating pressure in the regenerator. Also, the make-up water required controlling to
7 maintain lean loading at the inlet of the absorber. Therefore, these parameters were controlled using linear
8 PI-controllers.

9 In order to maintain the desired CO₂ capture rate in the absorber, the flow rate of lean amine was
10 controlled. Other parameters, such as liquid and gas flow distribution at the absorber, were also controlled
11 using PI-controllers. Although a few other controllers were also used for the stable operation of the plant,
12 such as controlling the temperature at the outlet of the cooler, the rich amine flow to the regenerator etc.,
13 they are not shown in Figure 5. Table 3 lists the controllers and the controlled, manipulated and measured
14 variables. Control structures for modified plants are discussed in the corresponding sections.

15 This configuration was used as the base case for the current study, and equipment was designed for both
16 off-design and dynamic simulations, following the method discussed in a later section. The specifications
17 of the equipment are presented in the corresponding sections.

18 **3.3 Selection of modified PCC cycle configuration**

19 A number of modifications in PCC plant configuration have been proposed in the literature, such as lean
20 vapor compression, multi-pressure regenerator, heat integrated stripping column, split flow process,
21 overhead condenser heat integration etc.; see [13, 39]. In a study on the techno economic evaluation of
22 various PCC plant modifications in [39], it has been reported that, with a natural gas based power plant,
23 the highest mitigation of energy penalty could be obtained using PCC configurations with split flow with
24 overhead condenser heat integration (OHC) as low as 5.28% with futuristic solvents. Also, the efficiency
25 penalties using improved split flow process (SF) (5.46%) and vapor recompression (5.86%), while using
26 futuristic solvents, have been found to be similar to what could be achieved using OHC. It is to be noted
27 that, with MEA, these efficiency penalty values would be higher due to the high energy requirement in the
28 regeneration of the solvent, as compared to the futuristic solvents mentioned in [39]. In another
29 comparative study on different improved stripper configurations [40], it has been shown that the
30 performance of a modified cycle with OHC and warm rich bypass provides the lowest energy penalty.
31 However, this is a combination of two configurations, namely OHC and bypassing part of the rich amine
32 stream from additional lean/rich heat exchanger. It has also been shown that a temperature approach of

1 5⁰C for those configurations, individually, does not provide sufficient reduction in the required reboiler
2 duty [40]. However, when considered individually, the two processes of OHC and SF are simpler
3 compared to other options with only static equipment providing sufficient thermal mass that may provide
4 sufficient damping of unwanted disturbances. It is to be noted that a better reduction in the efficiency
5 penalty can be obtained using a combination of some of the above-mentioned configurations, as also
6 shown in [40]. However, this is beyond the scope of this paper. Therefore, for designing the PCC plant,
7 these two configurations, namely OHC and SF, were chosen as examples, and individual flow sheets were
8 developed for this study.

9 The improved split stream configuration was adapted from [41] upon performing a parametric study on
10 deciding the split ratio for the process conditions at full-load power plant condition selected in this work.
11 In order to keep uniformity among all the configurations considered in the paper, the minimum
12 approaches at the lean/rich heat exchangers were considered to be 5⁰C. However, reduced minimum
13 approaches in both the modified configurations may lead to lower efficiency penalties than those achieved
14 in this paper. This issue will be addressed in future publications.

15 It was found that, with reduced rich amine stream pressure, the reboiler duty reduced for the split flow
16 configuration. The lowest possible reboiler duty can be obtained with a split fraction of 0.19 with 3 bar
17 rich amine stream pressure, as evident from Figure 8. Therefore, in the modified cycle with split stream,
18 the rich amine pressure was reduced from 6.8 bar to 3 bar with split fraction of 0.19. However, regenerator
19 pressure was not altered and kept constant at 2.1 bar. It may be noted that this split fraction obtained here
20 may not be optimal during part-load conditions; however, a study to identify the optimal fraction was not
21 performed, as it was beyond the scope of this study.

22 The simplified process configuration for incorporating OHC in the PCC plant is shown in Figure 9 as
23 adopted from [39]. The modified cycle configuration with OHC requires a high liquid to gas ratio for the
24 natural gas plant. The value that has been reported in [39] is 4.6 (mass basis). However, for this study the
25 liquid to gas ratio was selected to be optimum at 1.075. Therefore, a parametric study was performed to
26 obtain the flow fraction to OHC and the corresponding reduction of reboiler duty; refer to Figure 9. It was
27 found that a minimum reboiler duty requirement of 3.36 MJ/kg CO₂ could be obtained using a flow
28 fraction of 0.19 to OHC with rich amine pressure of 3 bar. Therefore, this flow fraction was used in the
29 modified cycle.

30 **4. Designing PCC plant**

31 **4.1 Design methodology of PCC plants**

1 Designing PCC mostly focused on rating the columns. The exchangers in the cycle are generally based on
2 plate and fin heat exchangers, and tools for rating this equipment are available and well established. These
3 exchangers were designed using one such tool available with the current process simulator. It is to be
4 noted that, in order to calculate the overall heat transfer coefficient and the overall heat transfer area,
5 flashing at the rich amine stream was considered. However, constraining the exchangers with similar
6 pressure drops at both the streams led to minor modifications in the overall heat transfer coefficients, and
7 they were allowed to be recalculated as a function of mass flow rates of the streams in the exchangers.

8 There are a few different approaches for designing absorbers for PCC processes, such as the empirical
9 design method, the theoretical design method, the laboratory method and the pilot plant technique [42].
10 All of these methods are used for determining the absorber height after calculating its diameter. The steps
11 involved in these methods are discussed in [42]. The empirical design method, laboratory method and
12 pilot plant technique require either experimental data or experimental set-up for accurate enhancement
13 factors. It is always difficult to obtain such data for either a full-scale plant or a pilot plant, due to
14 unavailability. Therefore, the theoretical method is the only choice when such data are not available.

15 The theoretical method requires numerical methods to solve the rigorous calculations for overall material
16 balances, liquid and gas side mass transfer, interfacial equilibrium and heat balances in both liquid and
17 gas. Process simulators can be used as an alternative to modeling the absorbers for solving all those
18 equations. Therefore, for the calculation of column height, process simulators those provide rate-based
19 column models and reaction kinetics can be used.

20 The methodology, including the calculations for column diameter and the use of simulation for
21 determining the column height, is presented in Figure 10. Upon deciding on the overall process
22 conditions, the required number of absorbers was decided based on their calculated diameter being within
23 the state-of-the-art range. In the calculation of the absorber diameter (D), the property data for each inlet
24 and outlet stream were generated using the process simulator. Parameters for packing materials, such as
25 void fraction (f), packing factor, etc., were obtained either from the simulator or from the data sheet
26 provided by the manufacturers. The superficial gas velocity was determined by applying the property data
27 for fluid and packing material in Equations (3), (4) and (5). Here G and G' are the gas mass velocity and
28 corresponding gas mass velocity per unit area. Gas density is represented by ρ , and A is the required area
29 of the absorber.

30

$$A = \left(\frac{1}{f} \right) \left(\frac{G}{G'} \right) \quad (3)$$

1
$$v_g = \left(\frac{G}{A \cdot \rho} \right) \quad (4)$$

2 Upon confirming the limiting value of the superficial gas velocity (v_g), the following correlation was used
3 to determine the diameter of the column.

4
$$D = \sqrt{\left(\frac{4}{\pi} \right) A} \quad (5)$$

5 Finally, the absorber height was determined using the process simulator, where rate-based column models
6 were used for the required capture ratio. The method was repeated until the pressure drop per meter of the
7 absorber was found to be within the desired range.

8 In this iterative design methodology, a few design constraints for columns in PCC processes were
9 imposed, as found in the literature [34, 43-45]. They are as follows.

- 10 1. The superficial gas velocity should be lower than 2.5 m/s for natural gas based power plants and
11 lower than 2 m/s for coal power plants [43].
12 2. The column diameter should be lower than 18.2 m, as it was found to be the highest column
13 diameter constructed for the SO₂ stripper used in the power plant [46].
14 3. The maximum pressure drop per meter of the column should be lower than 4.1 millibar [34].

15 The regenerator was also designed in a similar way to that of the absorber. As vapor flow rate in the
16 regenerator was far lower than in the absorber, the deciding factor was the rich amine flow rate.

17 **4.2 Identification of designed flue gas condition**

18 **4.2.1 Influence of flue gas flow rate on absorber diameter**

19 Considering the flue gas condition from the power plant operating at full load with an ambient
20 temperature of 15⁰C, the design methodology was applied to determine the absorber and regenerator
21 dimensions. Both items of equipment were designed at 75% of flooding velocities for gas.

22 While calculating the absorber diameter for a single absorber, it was found to be higher than the state-of-
23 the-art value of 18.2 m. Therefore, the number of absorbers needed to be more than a single train. It was
24 found that, with two trains of absorber, the required diameter was approximately 15.8 m with equal
25 distribution of mass flow rate of both flue gas and lean amine streams; see Figure 11. With the increased
26 number of absorbers, the required diameter reduced significantly; however, in order to reduce the
27 complexity of the plant and the possibility of flow maldistribution in the process, a two-absorber
28 configuration was selected for this study. Further, the heights of both absorbers were determined using the
29 process simulator so that the pressure drop per meter of absorber remained below 4.1 millibar.

1 As was discussed earlier, the power plants may have to work under flexible operating conditions including
2 operating under periodic part-load conditions; the flue gas flow rate will also vary periodically.

3 Therefore, it raises a design question as to what the designed flue gas condition should be: whether the
4 full-load condition will be optimum or the plant can be designed at lower load conditions. In order to
5 answer this question, a parametric study was performed to determine the benefits of designing the plants at
6 lower flue gas flow rate. It was found that the absorber diameter reduced with the reduction of designed
7 flue gas flow rate expressed in terms of power plant load; see Figure 12.

8 Designing the absorbers at a flue gas flow rate corresponding to 92% power plant load was found to
9 reduce the diameters from 15.8 m to 15.3 m; that led to a reduction of 4% in purchase cost of those
10 absorbers, as can be seen in Figure 13, considering equal flue gas flow rate distribution among the
11 absorbers. However, as high as 12% reduction in purchase cost can be achieved when the absorbers are
12 designed at as low as 75% power plant-load condition. This is the limit for reducing the flue gas flow rate,
13 as, below this, operation at full-load condition is not possible with a limit of 80% flooding.

14 Again, a parametric study was performed in order to discover the optimum flow distribution among the
15 absorbers. It was found that, with uneven distribution of flow among the absorbers, the purchase cost
16 increased, compared to equal distribution. This can be calculated using data from Figure 14 and Equation
17 (1). Therefore, only equal distribution of the flow rate in the absorbers was considered in this study.

18 The time-average value of the adopted assumed power plant load (Figure 1) is 92%, at which it was found
19 that, at most, a 4% reduction in the purchase cost of the absorbers could be achieved with equal flow
20 distribution among the absorbers. Therefore, the base case was analyzed for its performance under off-
21 design condition to decide on the designed value of the flue gas flow rate with absorbers rated at: 1. full
22 load condition with equal flow distribution, 2. time-average of assumed load of power plant with equal
23 flow distribution. Two cases with two different absorber sizes were analyzed:

- 24 1. Case 1: Base case without any process modifications; absorber designed at full-load condition
- 25 2. Case 2: Base case without any process modifications; absorber designed at 92% load condition

26 **4.2.2 Trade-off between performance and purchase cost: Off-design analysis**

27 Both cases were analyzed for three different off-design conditions: 1. part-load with constant liquid to gas
28 ratio, 2. variable capture rate, and 3. change in flow distribution. The results are presented in Figures 15-
29 17. The increased capture rate for Case 2 compared to Case 1, while operating under part-load condition,
30 can be observed from Figure 15(a). This is due to the increased liquid to gas ratio that was used in Case 2
31 for matching the capture ratio during full-load condition. However, this increased the efficiency penalty in

1 the process, as can be observed in Figure 15(b), by increasing the reboiler duty requirement. It can also be
2 observed that the lower is the power plant load, the higher is the increase in the efficiency penalty in Case
3 2 compared to Case 1. Maintaining a constant capture ratio during part load will reduce the liquid to gas
4 ratio and thereby will also cut this increase in efficiency penalty. Therefore, it can be inferred that, when
5 power plants with PCC operate under part-load condition, maintaining a constant capture ratio can keep
6 the efficiency penalty at a similar level; however, maintaining a constant liquid to gas ratio can lead to an
7 increased efficiency penalty, although with a higher capture rate. In the scenario of increased demand of
8 CO₂ capture rate, both the plants were found to have a similar trend of variation in the efficiency penalty
9 with changes in the capture rate; see Figure 16. However, Case 2 has a higher efficiency penalty than Case
10 1. Also, when the flow distribution was varied, it was found in Case 2 that, with less than 1% change in
11 flue gas flow rate, one of the absorbers reached its flooding condition (80% of maximum gas superficial
12 velocity); see Figure 17. However, for Case 1, this condition was not found to be reached until 4-5%
13 variation in flue gas flow rate. This calls for the absorbers to be designed for lower %-flooding conditions;
14 however, this will increase the required absorber diameter and thereby diminish the benefit of purchase
15 cost as found for using Case 2. Therefore, it may be inferred that, when operating under variable capture
16 rates, the PCC plant designed at lower flue gas condition will lead to a higher efficiency penalty with a
17 very low operating flue gas flow rate range. This led to the decision to discard the plant based on Case 2
18 that was designed on time-average power plant load condition. Two modified PCC configurations with
19 two absorbers designed at full-load condition with ambient temperature of 15⁰C were then subjected to
20 dynamic simulation with the capture ratio controlled. The equipment specifications are presented in Table
21 4.

22 **5. Dynamic simulation of the PCC plant for investigation for operability**

23 Three PCC cycles were subjected to dynamic simulation to investigate their operability under flexible
24 power plant operation. It was assumed that the power plant load varies according to the characteristics
25 presented in Figure 1. It was also assumed that the change in power plant load instantaneously changed
26 the flue gas flow rate. Therefore, corresponding changes in flue gas flow rate were determined using
27 Figure 3 and applied at the boundary stream in all cycles. Two operational scenarios can be applied for the
28 PCC plants to follow, while operating under such flue gas flow rate variations: 1. Constant lean amine
29 flow rate and 2. Controlling capture rate at some desired condition by manipulating the flow rate of the
30 lean amine. In the first scenario, it is expected that, during low flue gas flow rate, a higher capture rate can
31 be obtained than in the steady state design condition. Also, it can help to limit the disturbances due to the
32 variation of liquid in the absorber sump to propagate to the regenerator and from there upstream in the
33 power plant via the reboiler. However, the requirement of steam extraction will remain almost the same as

1 in the high flue gas condition, and this will lead to an increase in the efficiency penalty, as the reboiler
2 duty per kg CO₂ will be higher than desired. In the second scenario, the net capture rate can be kept
3 constant and the specific reboiler duty can be reduced, thereby limiting a further increase in the efficiency
4 penalty. Therefore, the second scenario was considered in this study, and all the plants were controlled to
5 provide a constant capture rate of 90%. Therefore, PID controllers were used after tuning them using the
6 simulator. It is to be noted that the discharge pressures of the rich pump in both the modified cycles were
7 controlled in order to maintain the flow rate distribution between two paths, i.e. one to the rich/lean heat
8 exchanger and the other direct to the regenerator or to the overhead condenser. Based on the results
9 obtained from the dynamic simulation and scaled fan, pump and compressor work, the efficiency penalties
10 were calculated using Equation (2) and used for trade-off.

11 **5.1 Operation under flexible power plant conditions**

12 The results of the dynamic simulation of all the cycles are presented in Figure 18. It can be observed that
13 almost all of them showed a stable operational scenario during variation in the flue gas flow rate with
14 regular characteristics, as in Figure 1. In all the cycles, an average capture rate of 90% was maintained
15 during the simulation. However, large overshoots of more than 0.5% during both ramp-up and ramp-down
16 operations were observed for the cycle with split flow configuration. Although the part of the absorber
17 was kept similar for all the configurations, the large overshoot was not expected there. In order to
18 understand the reason behind it, the variations in rich and lean loading were studied and it was found that
19 delayed variation of lean loading in the cycle led to a high capture rate during ramp-down and vice-versa
20 for ramp-down operation. In the cycle with split flow configuration, the settling time for new steady state
21 lean loading was found to be much higher compared to the other two configurations. Therefore, a higher
22 capture rate was obtained for this configuration. Moreover, the following were also observed:

- 23 1. As expected, the base case is the most stable and has a lower settling time compared to the other
24 two configurations.
- 25 2. After the ramp-down was over, reboiler duties in all the cycles were found to be increased with a
26 step change.
- 27 3. All the cycles reached new steady state conditions faster after ramp-down, compared to the
28 sharper ramp-up operation.
- 29 4. Oscillations were found for cycles with split flow and overhead condenser after ramp-up
30 operation; however, there were no such oscillations exhibited in the base-case cycle.
- 31 5. Higher specific reboiler duties for the base case and OHC were obtained compared to the split
32 flow configuration during the operation of the cycles under part-load condition. In addition, the
33 values of specific reboiler duties for both the modified cycles were predicted at most 7% higher

1 than those estimated using steady state rate-based models using HYSYS. However, the cycle with
2 OHC was found to have least reboiler duty as estimated in steady state.

3 6. Similar temperature profiles for all the absorbers were obtained from the simulation. However,
4 regenerators were found to have different temperature profiles for different cycles.

5 Large and continuous oscillation after ramp-up in the cycle with split flow configuration was due to the
6 variation in the inlet pressure of the rich pump. It can be observed in Figure 19 that the inlet pressure at
7 the pump varies continuously due to control action to maintain the pressure at the absorber sump.

8 The controller was tuned during steady state operational conditions. However, reconfiguration was
9 required to avoid such oscillation in the control action. It is important to mention here that a large
10 overshoot in the capture rate was found to induce large oscillations in the downstream operation and led to
11 unstable operation of the plant for both the modified configurations. However, it may be observed that the
12 cycle with OHC eventually dumped such oscillations and led the plant operate at steady state. Such
13 runaway operations in split flow configuration may be eliminated by reconfiguring the controllers used for
14 maintaining the pressure at the absorber sumps. This requires further analysis to identify the proper
15 control structure of such plants.

16 In addition, the discontinuities observed in the variation of reboiler duty for most of the cycles were found
17 to be due to the sudden and unexpected change in some parameters in the cycle and are purely numerical.

18 Moreover, the differences in rich and lean loading for all the configurations were found to have similar
19 trends to those of the base case; see Figure 20. However, rich loading for the base case was found to
20 decrease more compared to that of the two other configurations. This may be due to the variation in the
21 performance of the absorber sump. As it was found that, in the base case, the reduction in absorber sump
22 pressure was higher than in the other two configurations, that reduced the rich liquid hold-up in the base
23 case and thereby also reduced the rich loading by a greater amount than in the other two configurations.
24 However, this small reduction in rich loading did not greatly affect the reboiler duty; however, for the base
25 case, one can observe that the value of the reboiler duty is a little higher than that of the others.

26 Based on the above discussion, the following may be inferred.

- 27 1. The base case exhibits the most stable operation during flexible operation of the power plant.
- 28 2. The lowest specific reboiler duty during part-load operation may be obtained when using the split
29 flow configuration.
- 30 3. An increase in the rate of change of the flue gas may induce unwanted oscillation in the operation
31 of the PCC plants. This is more evident in the split flow configuration.

1 4. The cycle with OHC exhibits the ability to dump the unwanted oscillations that occur due to
2 variation in absorber sump pressure, as it uses static equipment (heat exchangers). Therefore, it
3 may return the reboiler to a stable operating condition without hampering the overall plant.

4 5. The overall control structure needs further analysis to utilize the available extra degrees of
5 freedom in order to eliminate the effect of control action leading to unwanted oscillations in the
6 operation of the reboiler.

7 **5.2 Trade-off between operability and efficiency penalty of PCC plant**

8 It was observed that the reboiler duty for both the modified cycles deviated from the values obtained from
9 steady state simulation using rate-based models in HYSYS. This is due to numerical errors generated due
10 to the use of equilibrium-based column models in dynamic simulation employing UniSim. It was found
11 that a maximum deviation in the calculation of such values was 7% for the case of the cycle with OHC.
12 Therefore, in order to calculate the average efficiency penalties for all the cycles during flexible operation
13 of the power plant, the reboiler duties were scaled based on their values obtained utilizing steady state and
14 steady state off-design condition simulations using equilibrium-based models. However, accurate values
15 of those can easily be obtained by using more accurate steady state rate-based models, and the exact
16 values of efficiency penalties (less than that obtained here) can be obtained. Table 5 presents a comparison
17 of efficiency penalties obtained during the simulation.

18 It can be noted that, during part-load operation and just after the end of ramp-down of flue gas flow rate,
19 the reboiler duty increased suddenly with a step change. This also led to an increase in the average
20 efficiency penalty in all the configurations. However, it was found that, although the cycle with OHC
21 provided the lowest efficiency penalty during full-load condition, the split flow configuration exhibited
22 the lowest average efficiency penalty of 6.98% during the entire operation of 24 hr. Meanwhile, the lowest
23 efficiency penalty of 6.82% can be observed in the cycle with OHC during its operation under full-load
24 condition, and that provides a 6% improvement in the performance of the PCC plant compared to that of
25 the base-case configuration. Therefore, based on the lowest average efficiency penalty, the split flow
26 configuration is the most suitable for implementation.

27 However, where the operation of that configuration is concerned, as previously discussed, the following
28 disadvantages exist.

- 29 1. High overshoot during ramp-down operation.
- 30 2. Runaway process during smaller fluctuation in the inlet pressure of the rich amine.

31 Therefore, the configuration with OHC can be the most suitable choice for using it as an alternative to be
32 used for PCC plant with the following advantages.

- 1 1. Lowest reboiler duty during full-load condition, which is the highest lasting operational scenario
- 2 during flexible operation of the power plant.
- 3 2. At least a 6% reduction in efficiency penalty may be obtained using this configuration as
- 4 compared to the base case.
- 5 3. It provides a high level of damping by using multiple heat exchangers to avoid the runaway
- 6 processes that may happen due to changes in inlet pressure of the rich pump or due to sharp
- 7 changes in capture rate and the corresponding variation in lean amine flow rate to the absorber.

8 However, issues such as designing a proper and optimal control structure by means of the extra degrees of
9 freedom provided by the modified cycles, the effect of flow maldistribution in the regenerator, the
10 comparison with other configurations in use of vapor compression, advance split stripper, etc. together
11 with OHC, need to be addressed further.

12 **6. Conclusions**

13 PCC processes using amine based solvents are considered to be near-term technology for use along with
14 power plants. The basic process is being used commercially in small- and medium-scale applications.
15 However, scaling up such processes cannot be done by just increasing the equipment size. With an
16 increase in the capacity of such processes, multiple design constraints are being imposed, as presented in
17 this paper. Therefore, it was necessary to evaluate a thorough method for selecting and designing a post-
18 combustion CO₂ capture process for a 600 MW range natural gas based power plant. In this paper, one
19 such method was discussed and evaluated, using two alternative PCC plant configurations: with split flow
20 with overhead condenser, and improved split flow without overhead condenser, besides the base-case
21 cycle. Optimum process conditions were identified from the literature for some of the process state points,
22 and parametric studies in steady state at design and off-design conditions were performed to find the rest.
23 Along with the trade-off between capital cost and energy performance in designing equipment like the
24 absorber and regenerator, a methodology was adapted and applied to the present cases, introducing design
25 constraints for superficial gas velocities for the equipment (e.g. 2.5 m/s for natural gas based plants) and
26 feasible limit of diameter (18.2 m).

27 Futuristic PCC processes need to be operated along with power plants that may have to work flexibly.
28 This will change the operation of PCC plants from steady operating mode to periodic operation in order to
29 cope with the change in flue gas conditions. Therefore, those configurations were further analyzed in
30 dynamic simulation to investigate their operability while subjected to flexible power plant conditions. The
31 cycle with overhead condenser was found to be the optimum, compared to the two other configurations, in
32 terms of stability of operation, capability of damping unwanted disturbances and average efficiency

1 penalty of 7%-points during the operation of 24 hrs. In addition, the following operational limits were also
2 identified:

- 3 1. Designing the absorbers at lower flue gas conditions than the full load may lead to a reduction of
4 up to 12% in purchase cost of the equipment. However, operation under full-load flue gas flow
5 rate will increase the required reboiler duty and thereby increase the efficiency penalty. Also, in
6 some cases (designed flue gas flow rate < 10% of full-load flow rate) operation at the full-load
7 flue gas flow rate will lead to flooding conditions in the absorbers.
- 8 2. Flow maldistribution among multiple absorbers needs to be limited to 4-5% to avoid flooding
9 conditions during part-load operation.
- 10 3. Oscillations due to control action in the absorbers need to be mitigated before reaching the
11 regenerator to avoid unwanted periodic disturbances upstream in power plants.

12 These identified operational limits, therefore, need to be considered when designing PCC plants for large-
13 scale power plants.

14 During analysis, the control structure used, however, was found to be insufficient for such integrated
15 cycles. Therefore, further studies need to be performed to identify a suitable and optimal control structure
16 for such plants that provide extra degrees of freedom. In addition, the following issues need to be
17 addressed.

- 18 1. Optimization of the average efficiency penalty during periodic flue gas flow rate variation needs
19 to be performed by suitable operational sequences when the flue gas flow rate changes during
20 ramp-up, ramp-down or part load.
- 21 2. The performance of other cycle alternatives, such as use of vapor compression in the reboiler etc.,
22 needs to be evaluated to identify the most suitable configuration, providing the least average
23 efficiency penalty during operation with a flexible power plant.
- 24 3. The effects of variation in ambient conditions need to be investigated in the cycles to find the
25 variation in average efficiency penalty.
- 26 4. Operability issues when different cycle alternatives are combined need to be studied for further
27 reduction in the efficiency penalty, as well as the elimination of transfer of disturbances upstream
28 in the power plants.

29 **Acknowledgement**

30 The authors gratefully acknowledge the support of the Norwegian University of Science and Technology,
31 Trondheim, Norway, and the Research Council of Norway.

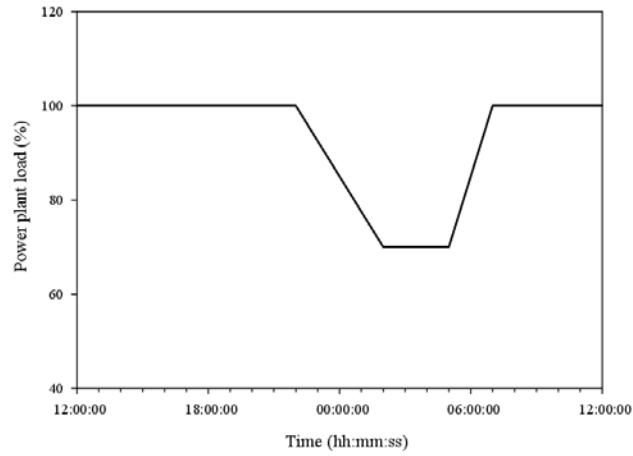
1 References

- 2 [1] IPCC, "IPCC Special Report on Carbon Dioxide Capture and Storage," New York, USA2005.
- 3 [2] A. A. Olajire, "CO₂ capture and separation technologies for end-of-pipe applications – A review," *Energy*, vol. 35, pp.
- 4 2610-2628, 6// 2010.
- 5 [3] T. N. MASAKI IJIMA, TAKASHI KAMIJYO, SHINSUKE NAKATANI, "MHI's Energy Efficient Flue Gas CO₂
- 6 Capture Technology and Large Scale CCS Demonstration Test at Coal-fired Power Plants in USA," *Mitsubishi Heavy Industries*
- 7 *Technical Review*, vol. 48, pp. 26-32, 2011.
- 8 [4] C. Saiwan, T. Supap, R. O. Idem, and P. Tontiwachwuthikul, "Part 3: Corrosion and prevention in post-combustion
- 9 CO₂ capture systems," *Carbon Management*, vol. 2, pp. 659-675, 2011/12/01 2011.
- 10 [5] S. S. Warudkar, K. R. Cox, M. S. Wong, and G. J. Hirasaki, "Influence of stripper operating parameters on the
- 11 performance of amine absorption systems for post-combustion carbon capture: Part I. High pressure strippers," *International*
- 12 *Journal of Greenhouse Gas Control*, vol. 16, pp. 342-350, 8// 2013.
- 13 [6] Z. Amrollahi, I. S. Ertesvåg, and O. Bolland, "Optimized process configurations of post-combustion CO₂ capture for
- 14 natural-gas-fired power plant—Exergy analysis," *International Journal of Greenhouse Gas Control*, vol. 5, pp. 1393-1405, 11//
- 15 2011.
- 16 [7] N. Mac Dowell and N. Shah, "Dynamic modelling and analysis of a coal-fired power plant integrated with a novel split-
- 17 flow configuration post-combustion CO₂ capture process," *International Journal of Greenhouse Gas Control*, vol. 27, pp. 103-
- 18 119, 8// 2014.
- 19 [8] L. E. Øi and S. H. P. Kvam, "Comparison of Energy Consumption for Different CO₂ Absorption Configurations Using
- 20 Different Simulation Tools," *Energy Procedia*, vol. 63, pp. 1186-1195, // 2014.
- 21 [9] X. Luo, M. Wang, and J. Chen, "Heat integration of natural gas combined cycle power plant integrated with post-
- 22 combustion CO₂ capture and compression," *Fuel*, vol. 151, pp. 110-117, 7/1/ 2015.
- 23 [10] Y. Le Moullec, T. Neveux, A. A. Azki, A. Chikukwa, and K. A. Hoff, "Process Modifications for Solvent-based Post
- 24 Combustion CO₂ Capture," *Energy Procedia*, vol. 63, pp. 1470-1477, // 2014.
- 25 [11] S. CCS. (2016, 25.10.2016). *Boundary Dam Carbon Capture Project*. Available: [http://www.saskpowerccs.com/ccs-](http://www.saskpowerccs.com/ccs-projects/boundary-dam-carbon-capture-project/)
- 26 [projects/boundary-dam-carbon-capture-project/](http://www.saskpowerccs.com/ccs-projects/boundary-dam-carbon-capture-project/)
- 27 [12] Z. Amrollahi, P. A. M. Ystad, I. S. Ertesvåg, and O. Bolland, "Optimized process configurations of post-combustion
- 28 CO₂ capture for natural-gas-fired power plant – Power plant efficiency analysis," *International Journal of Greenhouse Gas*
- 29 *Control*, vol. 8, pp. 1-11, 5// 2012.
- 30 [13] S. Ehlers, V. Roeder, A. Kather, P. Singh, and J. Davison, "Energetic Evaluation of Different Flow Sheet Modifications
- 31 of Post-Combustion CO₂ Capture Plant at Coal and Natural Gas Fired Power Plant for a Generic Improved Solvent," *Energy*
- 32 *Procedia*, vol. 63, pp. 1029-1039, // 2014.
- 33 [14] H. Hagi, T. Neveux, and Y. Le Moullec, "Efficiency evaluation procedure of coal-fired power plants with CO₂ capture,
- 34 cogeneration and hybridization," *Energy*, vol. 91, pp. 306-323, 11// 2015.
- 35 [15] IEAGHG, "Evaluation of process control strategies for normal, flexible and upset operation conditions of CO₂ post
- 36 combustion capture processes," September, 2016 2016.
- 37 [16] M. Karimi, M. Hillestad, and H. F. Svendsen, "Investigation of the dynamic behavior of different stripper
- 38 configurations for post-combustion CO₂ capture," *International Journal of Greenhouse Gas Control*, vol. 7, pp. 230-239, 3//
- 39 2012.
- 40 [17] M. Mofarahi, Y. Khojasteh, H. Khaledi, and A. Farahnak, "Design of CO₂ absorption plant for recovery of CO₂ from
- 41 flue gases of gas turbine," *Energy*, vol. 33, pp. 1311-1319, 8// 2008.
- 42 [18] L. E. Øi, "Aspen HYSYS simulation of CO₂ removal by amine absorption from a gas based power plant," 2007.
- 43 [19] J. L. Bravo, J. A. Rocha, and J. R. Fair, "MASS TRANSFER IN GAUZE PACKINGS," *Hydrocarbon Processing*, vol.
- 44 64, pp. 91-95, 1985.
- 45 [20] R. Billet and M. Schultes, "Predicting mass transfer in packed columns," *Chemical Engineering and Technology*, vol.
- 46 16, pp. 1-9, 1993.
- 47 [21] B. Hanley and C. C. Chen, "New mass-transfer correlations for packed towers," *AIChE Journal*, vol. 58, pp. 132-152,
- 48 2012.
- 49 [22] R. Dutta, Nord, L.O. and Bolland, O., "Prospects of using equilibrium-based column models in dynamic process
- 50 simulation of post-combustion CO₂ capture for coal-fired power plant," *Fuel*, vol. Under review, 04/11/2016 2016.
- 51 [23] P. Mores, N. Scenna, and S. Mussati, "CO₂ capture using monoethanolamine (MEA) aqueous solution: Modeling and
- 52 optimization of the solvent regeneration and CO₂ desorption process," *Energy*, vol. 45, pp. 1042-1058, 9// 2012.
- 53 [24] R. Dutta, Nord, L.O., Bolland, O., "Selection and validation of commercial process simulator for dynamic simulation of
- 54 post combustion CO₂ capture processes," presented at the 8th Trondheim conference on CO₂ capture and sequestration: Accepted
- 55 for presentation, Trondheim, Norway, 2015.
- 56 [25] N. E. Flø, "Post-combustion absorption-based CO₂ capture: modeling, validation and analysis of process dynamics,"
- 57 PhD-thesis, Chemical Engineering, NTNU, Trondheim, Trondheim, Norway, 2015.
- 58 [26] R. M. Montañés, M. Korpås, L. O. Nord, and S. Jaehert, "Identifying Operational Requirements for Flexible CCS
- 59 Power Plant in Future Energy Systems," *Energy Procedia*, vol. 86, pp. 22-31, 1// 2016.
- 60 [27] R. Turton, R. C. Bailie, W. B. Whiting, and J. A. Shaeiwitz, *Analysis, Synthesis and Design of Chemical Processes*:
- 61 Pearson Education, 2008.

- 1 [28] O. Bolland and H. Undrum, "A novel methodology for comparing CO₂ capture options for natural gas-fired combined
2 cycle plants," *Advances in Environmental Research*, vol. 7, pp. 901-911, 6// 2003.
- 3 [29] M. R. M. Abu-Zahra, J. P. M. Niederer, P. H. M. Feron, and G. F. Versteeg, "CO₂ capture from power plants: Part II. A
4 parametric study of the economical performance based on mono-ethanolamine," *International Journal of Greenhouse Gas
5 Control*, vol. 1, pp. 135-142, 4// 2007.
- 6 [30] A. Alhajaj, N. Mac Dowell, and N. Shah, "A techno-economic analysis of post-combustion CO₂ capture and
7 compression applied to a combined cycle gas turbine: Part I. A parametric study of the key technical performance indicators,"
8 *International Journal of Greenhouse Gas Control*, vol. 44, pp. 26-41, 1// 2016.
- 9 [31] C. Alie, L. Backham, E. Croiset, and P. L. Douglas, "Simulation of CO₂ capture using MEA scrubbing: a flowsheet
10 decomposition method," *Energy Conversion and Management*, vol. 46, pp. 475-487, 2// 2005.
- 11 [32] Z. Amrollahi, I. S. Ertesvåg, and O. Bolland, "Thermodynamic analysis on post-combustion CO₂ capture of natural-
12 gas-fired power plant," *International Journal of Greenhouse Gas Control*, vol. 5, pp. 422-426, 5// 2011.
- 13 [33] K. Jordal, P. A. M. Ystad, R. Anantharaman, A. Chikukwa, and O. Bolland, "Design-point and part-load considerations
14 for natural gas combined cycle plants with post combustion capture," *International Journal of Greenhouse Gas Control*, vol. 11,
15 pp. 271-282, 11// 2012.
- 16 [34] A. Lawal, M. Wang, P. Stephenson, and O. Obi, "Demonstrating full-scale post-combustion CO₂ capture for coal-fired
17 power plants through dynamic modelling and simulation," *Fuel*, vol. 101, pp. 115-128, 11// 2012.
- 18 [35] A. L. Kohl and R. B. Nielsen, "Gas Purification (5th Edition)," ed: Elsevier.
- 19 [36] P. Galindo, A. Schäffer, K. Brechtel, S. Unterberger, and G. Scheffknecht, "Experimental research on the performance
20 of CO₂-loaded solutions of MEA and DEA at regeneration conditions," *Fuel*, vol. 101, pp. 2-8, 11// 2012.
- 21 [37] E. O. Agbonghae, K. J. Hughes, D. B. Ingham, L. Ma, and M. Pourkashanian, "Optimal Process Design of Commercial-
22 Scale Amine-Based CO₂ Capture Plants," *Industrial & Engineering Chemistry Research*, vol. 53, pp. 14815-14829, 2014/09/24
23 2014.
- 24 [38] S. Ziaii, G. T. Rochelle, and T. F. Edgar, "Optimum design and control of amine scrubbing in response to electricity and
25 CO₂ prices," *Energy Procedia*, vol. 4, pp. 1683-1690, // 2011.
- 26 [39] IEAGHG, "Techno Economic evaluation of different Post Combustion CO₂ Capture Process Flow Sheet
27 Modifications," 2014.
- 28 [40] Y.-J. Lin, T. Madan, and G. T. Rochelle, "Regeneration with Rich Bypass of Aqueous Piperazine and
29 Monoethanolamine for CO₂ Capture," *Industrial & Engineering Chemistry Research*, vol. 53, pp. 4067-4074, 2014/03/12 2014.
- 30 [41] B. Eisenberg and R. Johnson, "Amine regeneration process," 1979.
- 31 [42] Z. H. Liang, T. Sanpasertparnich, P. P. T. Tontiwachwuthikul, D. Gelowitz, and R. Idem, "Part 1: Design, modeling and
32 simulation of post-combustion CO₂ capture systems using reactive solvents," *Carbon Management*, vol. 2, pp. 265-288,
33 2011/06/01 2011.
- 34 [43] N. Razi, "Rate-based modeling of CO₂ absorption in amine solutions, evaluation of mass transfer, kinetics and scale-
35 up," *Doctoral thesis, NTNU, Trondheim, Norway*, 2013.
- 36 [44] T. Nittaya, P. L. Douglas, E. Croiset, and L. A. Ricardez-Sandoval, "Dynamic Modelling and Controllability Studies of
37 a Commercial-scale MEA Absorption Processes for CO₂ Capture from Coal-fired Power Plants," *Energy Procedia*, vol. 63, pp.
38 1595-1600, // 2014.
- 39 [45] T. Nittaya, P. L. Douglas, E. Croiset, and L. A. Ricardez-Sandoval, "Dynamic Modeling and Evaluation of an
40 Industrial-Scale CO₂ Capture Plant Using Monoethanolamine Absorption Processes," *Industrial & Engineering Chemistry
41 Research*, vol. 53, pp. 11411-11426, 2014/07/16 2014.
- 42 [46] IEA-GHG, "IEA Greenhouse Gas R&D Programme (IEA GHG), CO₂ capture in low rank coal power plants," 2006/2,
43 January 2006.

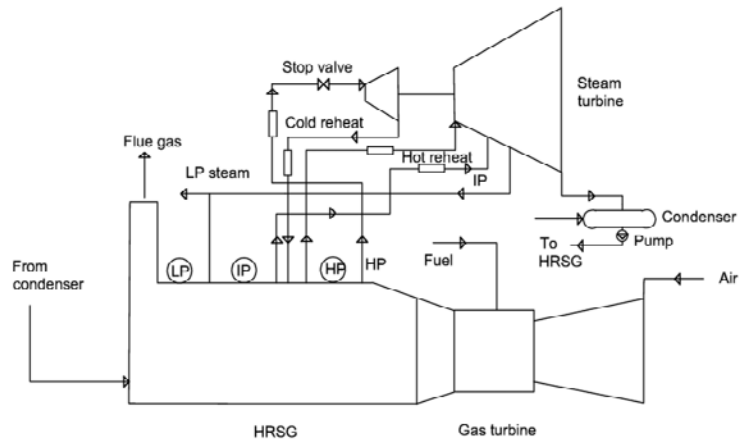
44

45



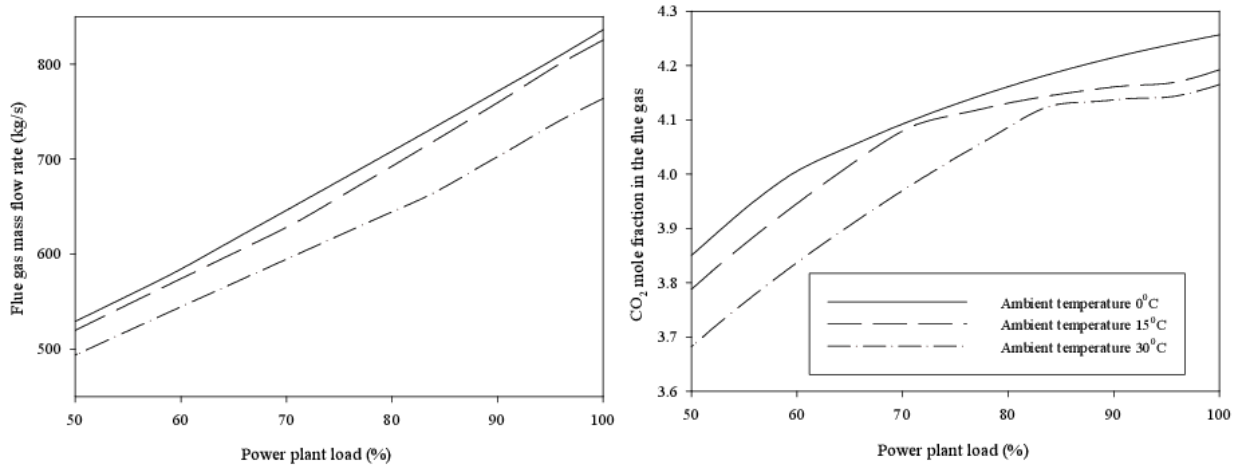
1
2
3

Figure 1: Assumed daily power plant load characteristics based on demand and pricing as adopted from [25]

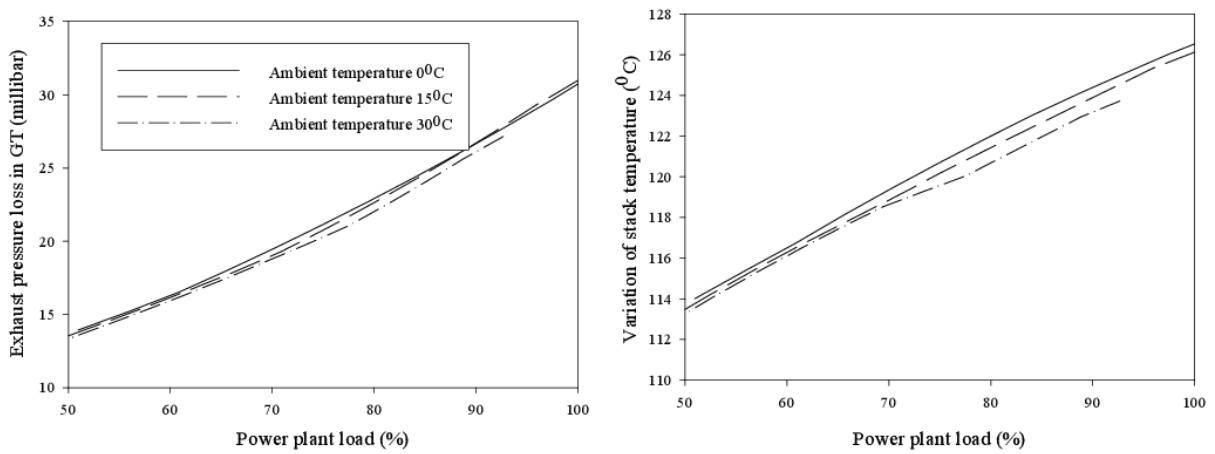


4
5
6

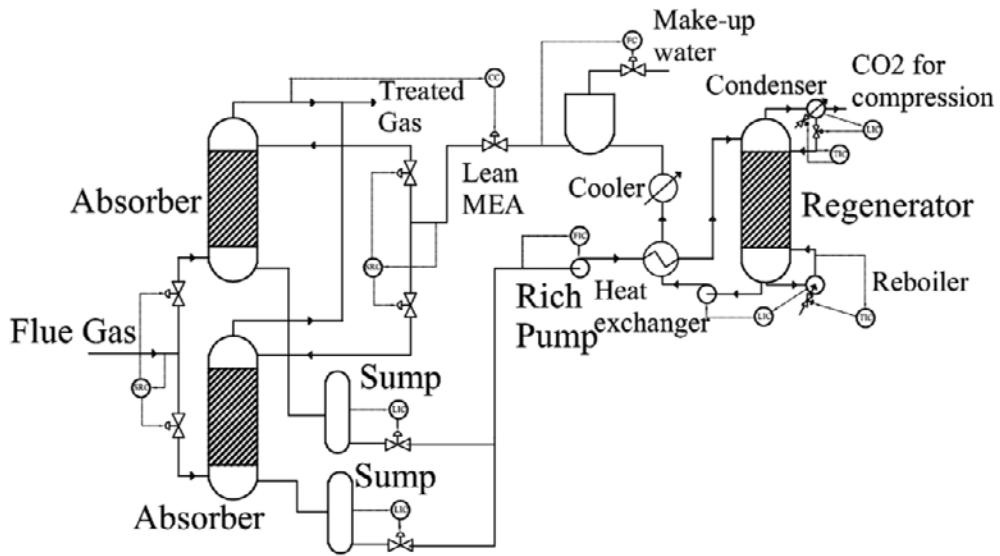
Figure 2: Schematic diagram of a typical natural gas based combined cycle power plant analyzed in this paper



1
2
3
Figure 3: Variation of flue gas conditions with part-load operation of power plant for different ambient temperatures

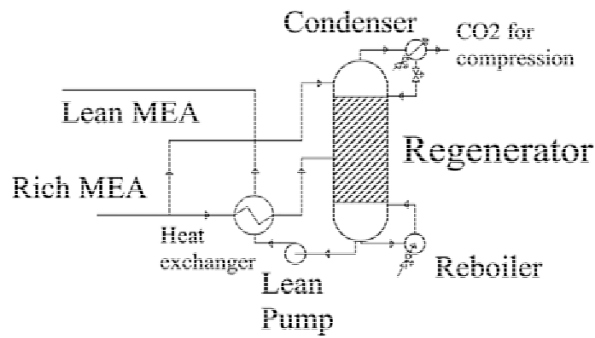


4
5
6
Figure 4: Variation of exhaust pressure loss and stack temperature with changing power plant load condition



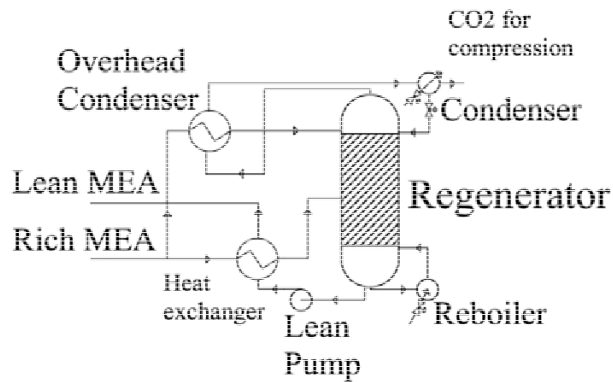
1
2
3

Figure 5: PCC plant configuration with controllers modeled for dynamic simulation as the base case



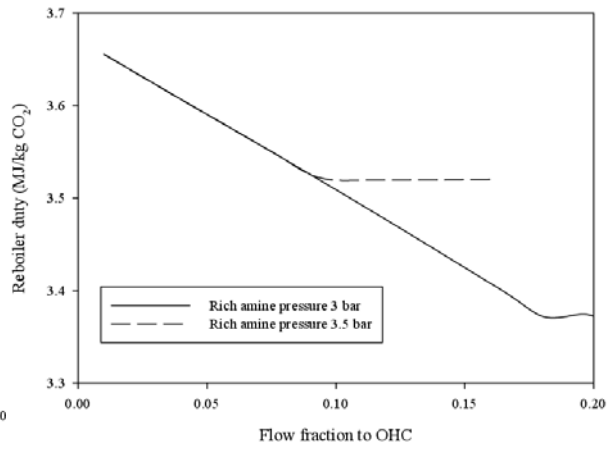
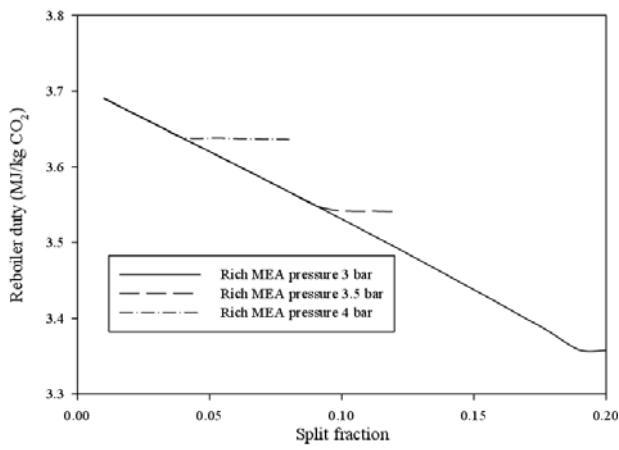
4
5
6

Figure 6: Improved split stream configuration for PCC plant [41]



7
8

Figure 7: Modified cycle configuration with overhead condenser heat integration [39]

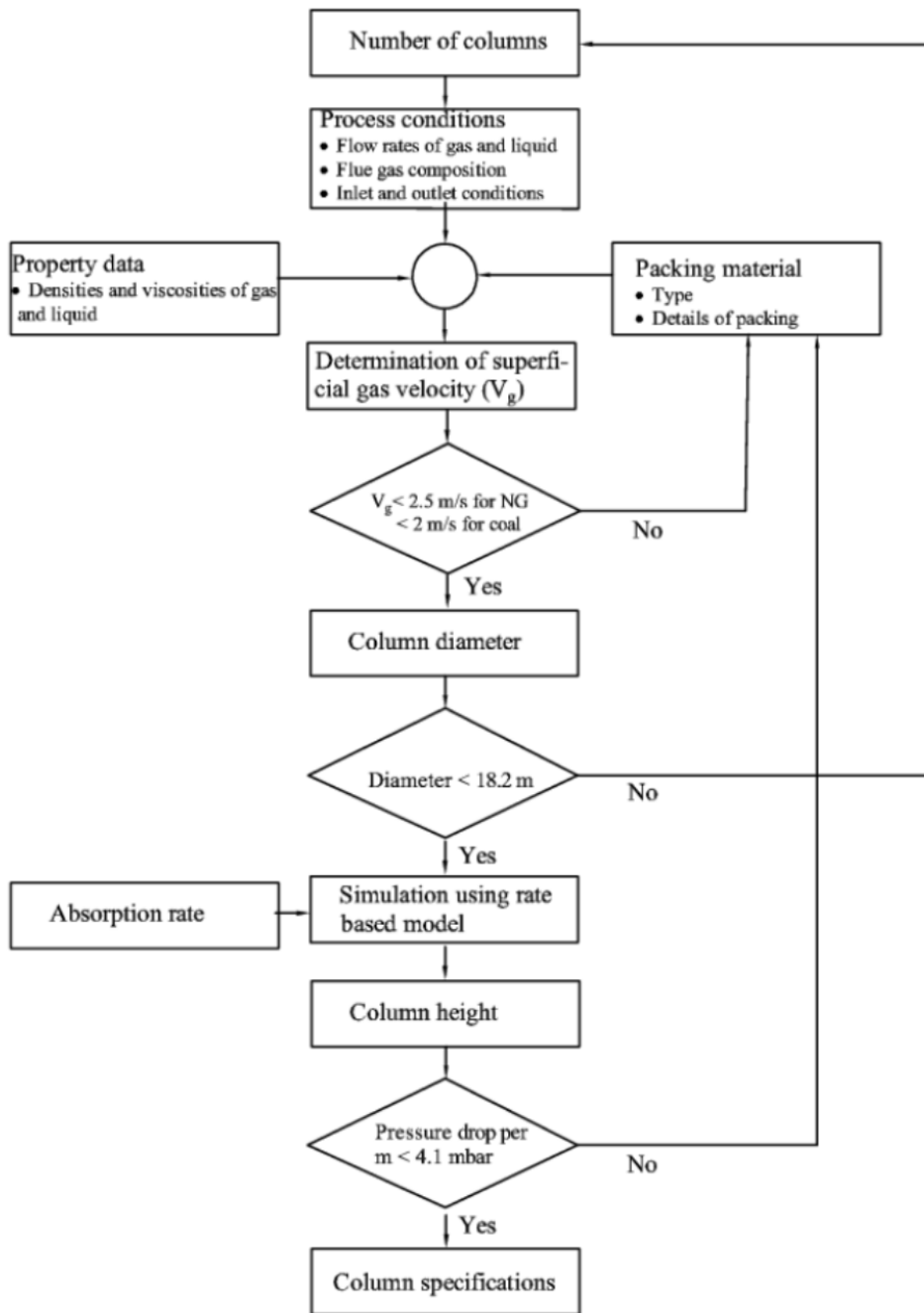


1

2 Figure 8: Results of parametric study for determination of
3 optimum split fraction for current process conditions

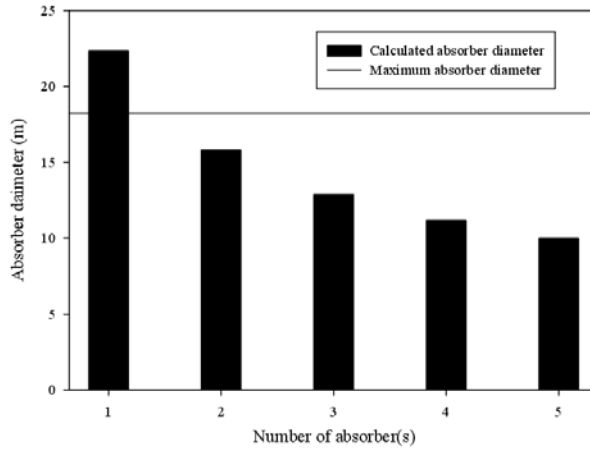
4 Figure 9: Results of parametric study for identification of flow
5 fraction to OHC

6



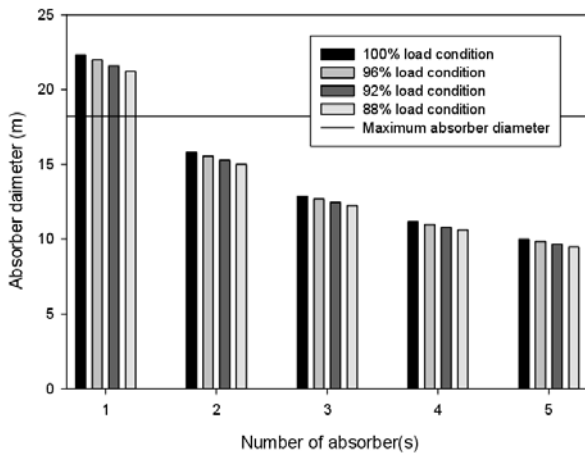
1
2
3

Figure 10: Steps to design the columns for PCC process based on simulation using rate-based column models

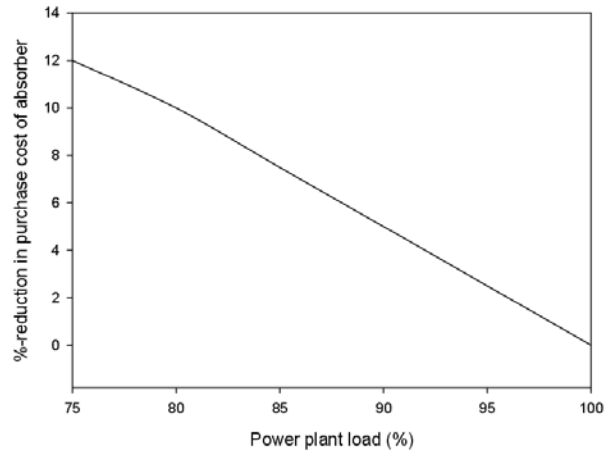


1
2

Figure 11: Calculated absorber diameter with number of absorber trains

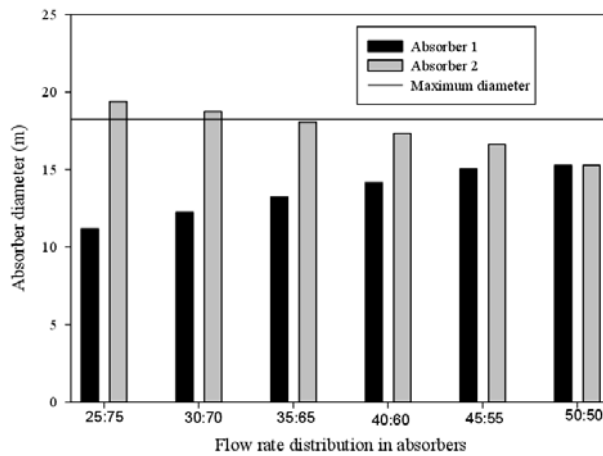


3



4 Figure 12: Calculated absorber diameter for different designed
5 power plant load conditions with number of absorber trains 7

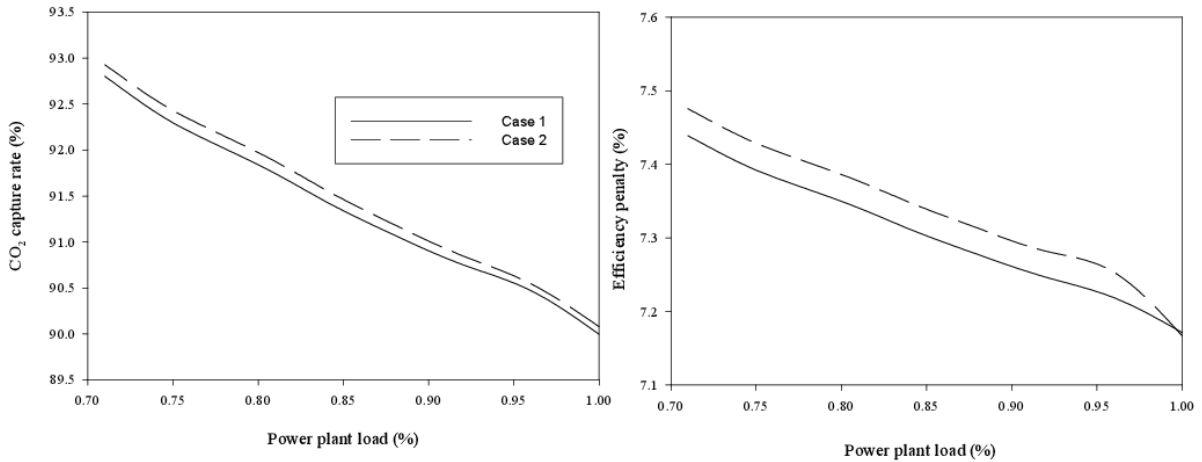
Figure 13: Calculated purchased cost of absorber with
designed power plant load condition



8
9

Figure 14: Calculated absorber diameters for different flow rate distribution in two absorbers

1



2

(a)

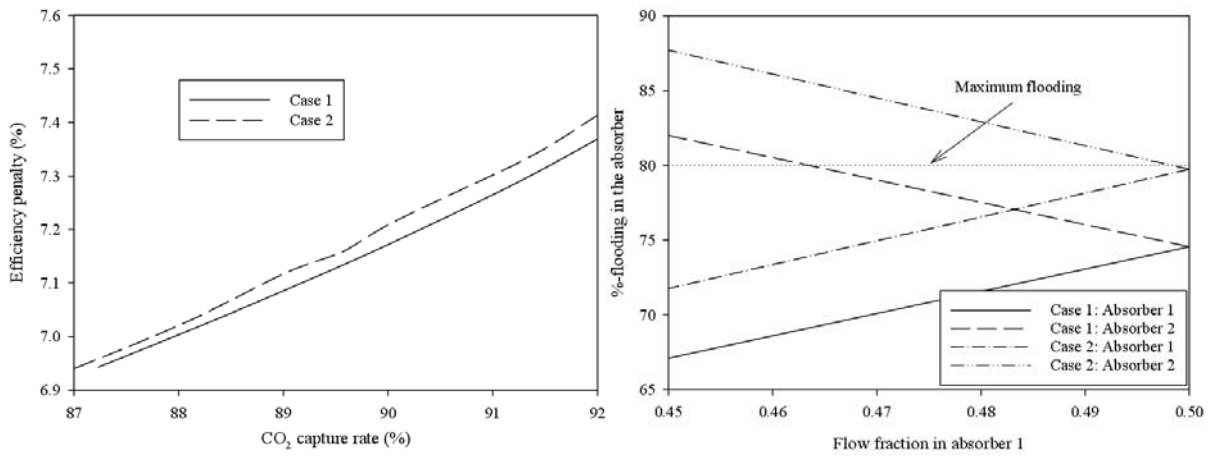
(b)

3

4

Figure 15: Variation of (a) capture ratio, (b) efficiency penalty during operation of the power plant under part-load conditions

5



6

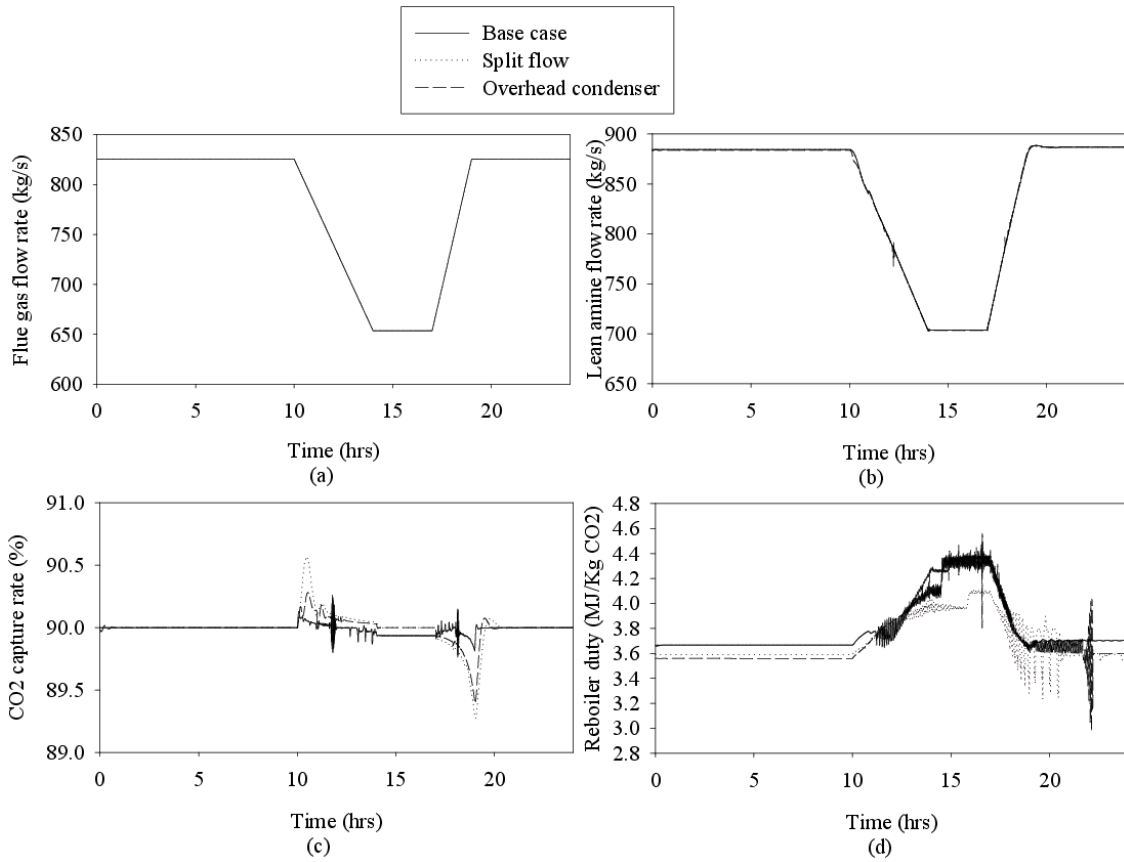
7

Figure 16: Change in efficiency penalty with variable CO₂ capture rate

8

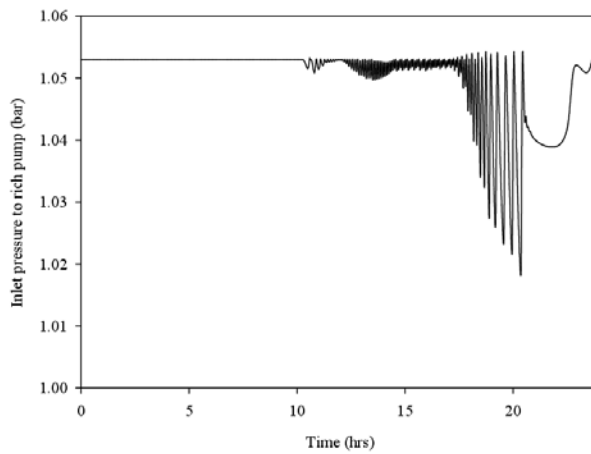
Figure 17: Variation in %-flooding with change in flow fraction among absorbers

11



1
 2 Figure 18: Results of dynamic simulations of cycles as base case and with split flow and overhead condenser: a) Flue gas flow
 3 rate, b) Lean amine flow rate, c) CO₂ capture rate, d) Reboiler duty

4

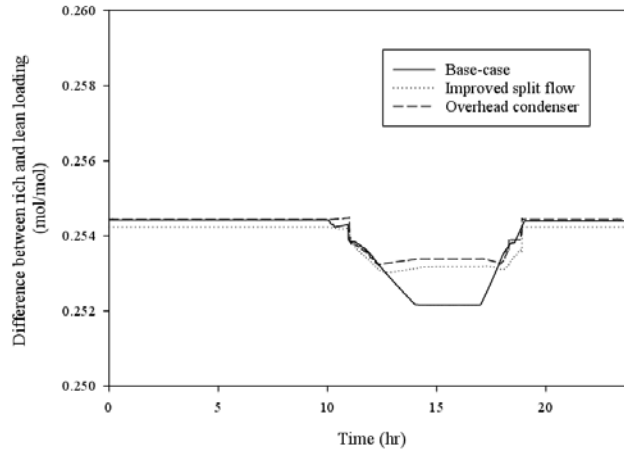


5

6

Figure 19: Variation in inlet pressure of rich pump in the cycle with split flow configuration

7



1
2 Figure 20: Difference between rich and lean loading in PCC plants during operation under flexible power plant

3 Table 1: Power plant operational data for designing PCC plant under different ambient conditions (GT: Gas Turbine)

Parameter	Ambient temperature 0°C	Ambient temperature 15°C	Ambient temperature 30°C
Gross power output (MW)	558.31	547.05	505.59
Fuel flow rate (kg/s)	19.99	19.46	17.99
Flue gas flow rate (kg/s)	836.14	825.31	763.85
Flue gas composition-%-mole-fraction			
N ₂	74.51	74.08	72.97
O ₂	11.54	11.55	11.35
CO ₂	4.26	4.19	4.16
H ₂ O	8.80	9.29	10.64
Ar	0.89	0.89	0.88
Flue gas temperature at stack (°C)	126.93	126.13	123.73
Total exhaust pressure loss in GT (millibar)	31.58	30.98	27.14
Extracted steam temperature (°C)	150.00	150.00	150.00
Extracted steam pressure (bar)	3.50	3.50	3.50

4
5 Table 2: Process conditions specified for development of the flow sheet in the simulator

Equipment	Parameter	Value
Absorber	Flue gas flow rate	825 kg/s
	CO ₂ mole-fraction in flue gas	4.19%
	Lean amine flow rate	891 kg/s
	Lean loading	21%
	Amine wt-% in lean stream	30%
	Rich loading at the outlet	47%
	CO ₂ capture rate	90%
Rich/Lean heat exchanger	Liquid to gas ratio	1.28
	Minimum temperature approach	5°C
	Pressure drop in each side	Shell: 70 kPa, Tube: 70 kPa
Regenerator	Rich amine pressure	2.1 bar
	Reboiler temperature	119°C
	Condenser temperature	30°C
	Reboiler liquid level	50%
	Condenser liquid level	50%
	Reboiler duty	3.7 MJ/kg CO ₂
	CO ₂ mole fraction at top outlet	95.6%

1

Table 3: Details of control structure for the PCC plant

Controlled variable	Manipulated variable	Measurement
CO ₂ capture rate	Lean amine flow rate	CO ₂ capture rate (Difference between CO ₂ flow rate in flue gas and treated gas)
Flue gas flow rate to absorbers	Valves at the inlet to absorbers	Flue gas flow rate to the absorber
Lean amine flow rate to absorbers	Valves at the inlet to absorbers	Lean amine flow rate to the absorber
Absorber sump pressure	Valves at the outlet of sump	Absorber sump pressure
Rich pump outlet pressure	Pump duty	Rich pump outlet pressure
Make-up water flow rate	Valve at the inlet to lean amine buffer	Lean loading
Rich amine temperature at the inlet of regenerator	Bypass valve for lean amine stream at the inlet of rich/lean heat exchanger	Rich amine temperature
Condenser temperature	Cooling water flow rate	Condenser temperature
Condenser liquid level	Reflux flow rate	Condenser liquid level
Reboiler temperature	Reboiler temperature	Steam flow rate
Reboiler liquid level	Reboiler liquid level	Lean pump speed

2

3

Table 4: Equipment specifications for modified PCC plants

Equipment	Parameter	Values
Absorber	Number of absorbers	2
	Diameter	15.8 m
	Height	27.2 m
	Packing type	Flexipac-1X
Regenerator	Number of regenerators	1
	Diameter	10.4 m
	Height	20 m
	Packing type	Mellapak-350Y
Rich/Lean heat exchanger	Overall heat transfer coefficient	7.5 kJ/C-hr
	Type	Compact plate-fin
Rich and lean pump	Efficiency	75%

4

Table 5: Average efficiency penalties obtained during dynamic simulation of all the PCC plant configurations with their corresponding corrected and scaled values for comparison

Configuration	Average efficiency penalty at full-load (%)		Average efficiency penalty at ramp-down (%)		Average efficiency penalty at part-load (%)		Average efficiency penalty at ramp-up (%)	
	Simulation	Scaled	Simulation	Scaled	Simulation	Scaled	Simulation	Scaled
Base case	7.27	-	7.53	-	8.11	-	7.58	-
Split flow	7.17	6.85	7.45	7.11	7.71	7.35	7.44	7.11
OHC	7.13	6.82	7.49	7.16	8.06	7.68	7.59	7.24

7

8