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# Optimal Process Design of Commercial-scale Amine-based $CO_2$ Capture Plants

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**ABSTRACT**

Reactive absorption with an aqueous solution of amines in an absorber/stripper loop is the most mature technology for post combustion $CO_2$ capture (PCC). However, most of the commercial-scale $CO_2$ capture plant designs that have been reported in the open literature are based on values of $CO_2$ loadings and/or solvent circulation rates without an openly available techno-economic consideration. As a consequence, most of the reported designs may be sub-optimal, and some of them appear to be unrealistic from practical and operational viewpoints. In this paper, four MEA-based $CO_2$ capture plants have been optimally designed for both gas-fired and coal-fired power plants based on process and economic analyses. We have found that the optimum lean $CO_2$ loading for MEA-based $CO_2$ capture plants

that can service commercial-scale power plants, whether natural gas-fired or coal-fired, is about 0.2 mol/mol for absorber and stripper columns packed with Sulzer Mellapak 250Y$^{TM}$ structured packing. Also, the optimum liquid/gas ratio for a natural gas combined cycle (NGCC) power plant with a flue gas composition of approximately 4 mol% $CO_2$ is about 0.96, while the optimum liquid/gas ratio for a pulverized coal-fired (PC) power plant can range from 2.68 to 2.93 for a flue gas having a $CO_2$ composition that ranges from 12.38 mol% to 13.50 mol%.

***Keywords***: Optimal, Process Design, Commercial-scale, $CO_2$ Capture; Amine

## 1. INTRODUCTION

Reactive absorption with aqueous solutions of amines in an absorber/stripper loop is the most mature technology for post combustion $CO_2$ capture (PCC).[1] The main barrier that remains unresolved is the huge energy requirement for solvent regeneration in the stripper. In fact, the reduction of solvent regeneration energy is the focus of most of the amine-based PCC research currently being performed globally. From the view point of current research and development (R&D) activities worldwide, three main areas are being investigated in order to reduce the regeneration energy requirement of amine-based PCC, namely: (a) development of new solvents with better overall performance than 30 wt% monoethanolamine (MEA) aqueous solution, which is generally considered as the base-line solvent for solvent-based PCC, (b) PCC process optimization, including modifications of PCC plant configuration, and (c) optimal integration of the PCC Plant, including the associated $CO_2$ compression system, to the upstream power plant.

In recent years, research activities aimed at testing new solvents, as well as the optimisation of solvent-based PCC, have resulted in several projects with the setting up of pilot plants globally.[2-7] In most of the studies that have been reported, aqueous MEA solution is usually taken as the reference solvent to which new solvents are compared. Among the pilot-scale studies that have been reported for MEA, Notz et al.[4] have reported a very comprehensive set of results based on systematic studies of $CO_2$ capture with aqueous MEA solutions in a pilot plant and they also gave a fairly detailed description of the pilot plant with sufficient information and data to permit successful modelling of it.

Process modelling is critical in the scale-up of a pilot plant to a commercial-scale plant during design. There are several rigorous process modelling studies of the MEA-based $CO_2$ capture process at pilot-scale in the open literature, with many of them focusing on the absorber as a stand-alone unit,[8-11] or the stripper as a stand-alone unit,[12-14] and some of them have considered the absorber and the stripper in a closed loop.[15-17] However, in spite of the numerous process modelling and simulation studies of the MEA-based $CO_2$ capture process at pilot-scale that have been reported in the open literature, there is nothing freely available with complete information on the optimal design of amine-based $CO_2$ capture plants that can service commercial-scale coal-fired power plants, as well as onshore-based commercial-scale gas-fired power plants. It is important to state that some work on the design of commercial-scale MEA-based $CO_2$ capture plants for coal-fired and gas-fired power plants have been reported in the open literature[18-31] with minimal or incomplete information on the design process, and/or with partial or non-disclosure of the design results by most of them. However, the paper by Kvamsdal et al.,[22] which reported an optimised design of an MEA-based $CO_2$ capture plant for a 540 MWe (gross) NGCC power plant in an offshore application, is an exception because complete information on the design process was given in addition to full disclosure of the optimised design results. Apart from the paper by Kvamsdal et al.[22] most of the commercial-scale $CO_2$ capture plant designs that have been reported in the open literature are based on values of $CO_2$ loadings and/or solvent circulation rates without an openly available techno-economic consideration. As a consequence, most of the reported designs may be sub-optimal, and some of them appear to be unrealistic from practical and operational viewpoints when compared with the design data in the non-confidential report of the front end engineering design (FEED) study undertaken by Fluor® for the "ROAD project" (**R**otterdam **O**pslag en **A**fvang **D**emonstratieproject'; Rotterdam Capture and Storage Demonstration Project),[32] as well as statements in process licensor reports.[33-35]

### 1.1. Motivation

As earlier stated, most of the commercial-scale $CO_2$ capture plant designs that have been reported in the open literature are based on values of $CO_2$ loadings and/or solvent circulation rates without techno-economic consideration and, as a consequence, most of the designs may be sub-optimal, and some of them even appear unrealistic from practical and operational viewpoints. For example, Lawal et al.,[26] while adopting $CO_2$ loadings of 0.29 mol/mol and 0.47 mol/mol for the lean MEA and the rich MEA solutions, respectively, used the generalized pressure drop correlation (GPDC) for packed columns to scale-up the pilot plant model that they developed with gPROMS®, and they ended up with a commercial-scale $CO_2$ capture plant that can service a 500 MWe (net power without $CO_2$ capture) subcritical coal-fired power plant. Their design comprised of two absorbers, each with a diameter of 9 m based on an assumed pressured drop of 42 mm-$H_2O$/m, and a single stripper with a diameter of 9 m. The pressure drop they assumed as a basis for diameter sizing is about two times the maximum pressure drop that is recommended for amine systems, which are known to be moderately foaming.[28,29] Furthermore, based on the work by Cifre et al.[20], Lawal et al.[26] assumed a preliminary height of 17 m for each of the absorbers and ended up with a packed height of 27m by varying the absorber height, but they did not report the stripper height. Their design results appear to be sub-optimal when compared with Fluor's design data for the ROAD demonstration project, and there may be operational issues with their design because of the large pressure drop they assumed for the absorber diameter sizing. Similarly, Sipocz and Tobiesen[27] adopted 0.132 mol/mol and 0.473 mol/mol for the $CO_2$ loadings of the lean MEA and rich MEA solutions as the basis for scaling up the pilot plant model that they developed with the CO2SIM software to a commercial-scale $CO_2$ capture plant that can service a

410.6 MWe (gross power without $CO_2$ capture) natural gas combined-cycle (NGCC) power plant. Their design comprised of a single absorber with a 9.13 m diameter and a height of 26.9 m, and a single stripper with a diameter of 5.5 m and a height of 23.5 m. Their design results appear to be unrealistic because it is very unlikely that a single absorber with 9.13 m diameter can handle the amount of flue gas they used as the basis for their design. Furthermore, Biliyok and Yeung[28] adopted $CO_2$ loadings of 0.234 mol/mol and 0.4945 mol/mol for the lean MEA and rich MEA solutions, respectively, and they used the method used by Lawal et al.[26] to design a commercial-scale $CO_2$ capture plant that can service a 440 MWe (gross power without $CO_2$ capture) NGCC power plant. They ended up with four absorbers, each having a diameter of 10 m and a height of 15 m, and a single stripper having a diameter of 9 m and a height of 15 m. The choice of four absorbers by Biliyok and Yeung[28] most likely followed the design reported by Hetland et al.,[21] and/or Kvamsdal et al.,[22] which was a special design case for an offshore application where balanced distribution of structural weight is an important design factor since local concentration of dead weight could affect the stability of an offshore platform. Also, both Hetland et al.[21] and Kvamsdal et al.[22] noted that operational flexibility informed their choice of four absorbers and a single stripper in their design, with one absorber servicing each of the four trains that make up the offshore NGCC power plant they used as a basis for their design. Therefore, the design by Biliyok and Yeung[28] is unlikely to be adopted in an onshore application because a 440 MWe (gross) NGCC power plant should require no more than two absorbers if optimally designed. **Table S.1** in the Supporting Information summarizes and compares the commercial-scale designs reported by various authors in the open literature. It is clear from **Table S.1** that, with the exception of the special design case by Kvamsdal et al.[22] for an offshore application, complete information on the optimal design of absorption and stripping columns for commercial-scale amine-based $CO_2$ capture plants is still lacking in the open literature, and it is this lack of information on the optimal design of the absorption and stripping columns for amine-based $CO_2$ capture plants in the open literature that motivated the work discussed in this paper.

### 1.2. Novelty

The design method and philosophy in this work is novel in the way Aspen Plus® has been used in the design of the amine-based $CO_2$ capture plants considered in this paper and, in the spirit of transparency, we have detailed how we used Aspen Plus® so that researchers and process design engineers can easily adopt the design philosophy and methodology for their design work. In addition to using recommended rules for the absorber and stripper column diameter sizing, the column heights needed for 90% $CO_2$ capture were arrived at systematically based on rate-based calculations. Furthermore, to the best of our knowledge, this is likely to be the first work in the open literature on the optimal process design of amine-based $CO_2$ capture plants that can service commercial-scale coal-fired power plants and onshore-based commercial-scale gas-fired power plants, with a complete explanation of the design method and philosophy and full disclosure of the complete design results. Also, it is likely to be the first work in the open literature to optimize both the lean amine $CO_2$ loading and the solvent circulation rate based on process and economic analyses. It is pertinent to add that, in the open literature, this is likely to be the first work to optimally design integrated commercial-scale absorber-stripper systems for amine-based $CO_2$ plants that can service coal-fired power plants and onshore-based gas-fired power plants with full disclosure of the design results, taking into consideration the capital and operating costs of the lean amine solution pump, the rich amine solution pump, the lean/rich cross heat exchanger, the lean amine solution cooler, the stripper condenser, the stripper reflux drum and reflux pump, and the stripper reboiler.

## 2. PROCESS DESCRIPTION AND MODELLING FRAMEWORK

### 2.1. Process Description

The basic flowsheet of an amine-based $CO_2$ capture process is shown in **Figure 1**. The process consists of countercurrent contact of the flue gas coming from a direct contact cooler (DCC) unit with an amine solution in a packed absorber. The rich flue gas enters the absorber at the bottom while the lean amine solution is introduced into the top of the absorber. The treated flue gas leaves the top of the absorber and is normally washed in a water-wash section (not shown in **Figure 1**) so as to remove entrained solvent droplets and, in turn, limit the loss of valuable solvents in addition to meeting environmental regulations on solvent emissions into the atmosphere. The rich solvent from the bottom of the absorber is sent to the stripper for $CO_2$ stripping after absorbing some of the heat in the lean solvent exiting the stripper bottom in a cross heat-exchanger. In the stripper, the downward flowing rich solvent is stripped of its absorbed $CO_2$ by the upward flowing steam generated by the reboiler. The vapour stream from the top of the stripper, which is essentially a mixture of $CO_2$, steam and some traces of the amine used, is partially condensed in a condenser and a fraction or all of the condensed liquid is returned to the top of the stripper as reflux. The uncondensed stream, which is mainly $CO_2$, is sent for compression, transportation and sequestration.

**2.2. Modelling Framework**

Aspen Plus® RadFrac model, a second generation rate-based model for multistage separation operations, was used for the modelling of the absorption and stripping columns in the MEA-based $CO_2$ capture plants as discussed in this paper. Being a pre-requisite for accurate process modelling of the $CO_2$ capture plants, validated high fidelity models were used for thermodynamic and transport properties.

**2.2.1. Thermodynamic Model**

The model adopted for the thermodynamic properties is based on the work by Zhang et al.[36] The model uses the electrolyte-NRTL activity coefficient model for the liquid phase properties and PC-SAFT equation of state for vapour phase properties. The model has been validated by Zhang et al.[36] against experimental data in the open literature. The equilibrium reactions describing the solution chemistry of $CO_2$ absorption with MEA, which are integral components of the thermodynamic model, include:[36]

$$2H_2O \leftrightarrow H_3O^+ + OH^- \tag{1}$$

$$CO_2 + 2H_2O \leftrightarrow HCO_3{}^- + H_3O^+ \tag{2}$$

$$HCO_3{}^- + 2H_2O \leftrightarrow CO_3{}^{2-} + H_3O^+ \tag{3}$$

$$MEACOO^- + H_2O \leftrightarrow MEA + HCO_3{}^- \tag{4}$$

$$MEAH^+ + H_2O \leftrightarrow MEA + H_3O^+ \tag{5}$$

### 2.2.2. Reaction Kinetics Model

The formation of carbamate and bicarbonate are kinetically limited and the forward and reverse reactions are given as follows:[15]

$$MEA + CO_2 + H_2O \rightarrow MEACOO^- + H_3O^+ \tag{6}$$

$$MEACOO^- + H_3O^+ \rightarrow MEA + CO_2 + H_2O \tag{7}$$

$$CO_2 + OH^- \rightarrow HCO_3^- \tag{8}$$

$$HCO_3^- \rightarrow CO_2 + OH^- \tag{9}$$

In Aspen Plus, the reaction rates for the above kinetically limited reactions are described by power law expressions as follows:[15]

$$r_j = k_j^0 exp\left(-\frac{\varepsilon_j}{R}\left[\frac{1}{T} - \frac{1}{298.15}\right]\right)\prod_{i=1}^{N} a_i^{\alpha_{ij}} \tag{10}$$

where $r_j$ is the reaction rate for reaction $j$, $k_j^0$ is the pre-exponential factor, $\varepsilon_j$ is the activation energy, $R$ is the gas constant, $T$ is the system temperature in Kelvin, $a_i$ is the activity of species $i$, and $\alpha_{ij}$ is the reaction order of species $i$ in reaction $j$. The kinetic expressions for the carbamate and bicarbonate reactions, including the rate constant parameters, were obtained from the work by Zhang and Chen[15] and they are summarized in **Table 1**.

### 2.2.3. Transport Property Models

Aspen Plus® RadFrac model requires quantitative values of the transport properties that are part of the correlations for heat transfer, mass transfer, interfacial area, liquid holdup, pressure drop, etc. The transport properties include density, viscosity, surface tension, thermal conductivity, and binary diffusivity[9]. A summary of the models in Aspen Plus that were adopted for the transport properties calculations is given in **Table 2**.

## 3. PROCESS DESIGN OF ABSORPTION AND SRIPPING COLUMNS

The process design of packed absorber and stripper columns entails the determination of the column diameter and the packed height needed to achieve a given separation, having chosen the solvent and packing type to be used. The design process is not a clear cut science but more of a combination of science and art based on experience. The column diameter for a given gas flowrate and liquid flowrate is usually determined based on two criteria: (i) the maximum pressure drop that can be tolerated and (ii) the approach to maximum capacity. The approach to maximum capacity can range from 70 to 86 percent of the flooding point velocity,[37,38] but packed columns are more usually designed within 70 to 80 percent of the flood point velocity[38]. The column height needed to achieve a given separation is determined using the concept of height of transfer unit (HTU) or the height equivalent to a theoretical plate (HETP), but the use of HETP is usually the preferred approach.[38]

### 3.1. Column Diameter Sizing

The column diameter ($D$) is related to the superficial velocity of the gas stream as follows:

$$D = \sqrt{\frac{4G}{\pi U_s}} \tag{11}$$

where $G$ is the gas flowrate and $U_s$ is the superficial velocity of the gas.

The superficial velocity of the gas stream is related to the packed column capacity factor by the following equation:[37-39]

$$C_0 = U_s \left(\frac{\rho_G}{\rho_L - \rho_G}\right)^{0.5} F_p^{0.5} \nu^{0.05} \tag{12}$$

where $C_0$ is the capacity factor; $\rho_G$ and $\rho_L$ are the gas density and the liquid density, respectively; $F_P$ is the packing factor of the packing in the column, and $\nu$ is the kinematic viscosity of the liquid.

The capacity factor for a packed column is a function of the flow parameter ($X$) and the pressured drop per unit height of the packing ($\Delta P$). The flow parameter is defined by the following equation:[37-39]

$$X = \frac{L}{G}\left(\frac{\rho_G}{\rho_L}\right)^{0.5} \tag{13}$$

where $L$ is the liquid flowrate.

Although generalized pressure drop correlation (GPDC) charts have been developed for both random and structured packings,[37-39] the more accurate vendor-developed pressure drop correlation for each specific packing is considered proprietary and is usually not disclosed by vendors. However, Aspen Tech has a special arrangement with packing vendors and, as a consequence, vendor correlations for pressure drop are built into Aspen Plus for several packings.

### 3.2. Packed Height based on HETP

The height equivalent to a theoretical plate (HETP) in a packed column for a stage designated by subscript $j$ is given by:[40,41]

$$HETP_j = \frac{ln\lambda_j}{\lambda_j - 1}\left(HTU_{G,j} + \lambda_j HTU_{L,j}\right) = \frac{ln\lambda_j}{\lambda_j - 1}\left(\frac{u_{Gs}}{k_{G,j}a_{e,j}} + \lambda_j \frac{u_{Ls}}{k_{L,j}a_{e,j}}\right) \tag{14}$$

with

$$\lambda_j = \frac{m_j G_j}{L_j} \tag{15}$$

where $HTU_{G.j}$ and $HTU_{L.j}$ are, respectively, the heights of transfer units for the gas and liquid phases in stage $j$; $\lambda_j$ is the stripping factor for stage $j$; $k_{G,j}$ and $k_{L,j}$ are, respectively, the local mass-transfer coefficients for the gas and liquid phases; $a_{e,j}$ is the effective interfacial area per unit volume of the packed section in stage $j$; $u_{Gs}$ and $u_{Ls}$ are, respectively, the superficial velocities for the gas and liquid phases; $m_j$ is the local slope of the equilibrium line for stage $j$; $G_j$ and $L_j$ are, respectively, the local flowrates of the gas and liquid streams to stage $j$. It is clear that the accuracy of the HETP calculated by eq. (14) is a function of the accuracy of the correlations used for the mass-transfer coefficients, the effective interfacial area, the pressure drop, as well as the model for vapour-liquid-equilibrium (VLE).

The packed height required for a given separation is the summation of the HETPs of the stages in the packed column. Thus, for a column with $N$ number of stages, the packed heights for an absorber (without condenser and reboiler) and a stripper (with a condenser and reboiler) are given as follows:

$$Z_{Absorber} = \sum_{j=1}^{N} HETP_j \tag{16a}$$

$$Z_{Stripper} = \sum_{j=2}^{N-1} HETP_j \tag{16b}$$

# 4. MODEL VALIDATION AT PILOT-SCALE AND DESIGN PHILOSOPHY

## 4.1. Aspen Plus Rate-based Model Validation at Pilot-scale

As previously stated, the Aspen Plus rate-based model was used to model the absorber and the stripper columns in the $CO_2$ capture plants. Although the model had previously been validated by Zhang et al.,[9,15] there was a need to revalidate the rate-based model for the Sulzer Mellapak 250Y™ structured packing used in the scale-up design cases considered in this paper. This was accomplished using the comprehensive pilot plant results reported by Notz et al.[4] The model validation strategy targeted the lean $CO_2$ loading by varying the stripper reboiler duty. Figures 2(a) to 2(c) show the parity plots for the $CO_2$ capture level, the stripper reboiler duty and the rich $CO_2$ loading, respectively, while Figure 2(d) shows the variation of the specific reboiler duty with liquid/gas ratio. The average percent absolute deviations of the model results for the $CO_2$ capture level, the stripper reboiler duty, and the rich $CO_2$ loading, when compared with the 47 experimental cases reported by Notz et al.,[4] are 3.75%, 5.08%, and 2.68%, respectively. The percent absolute deviations of the model results are in good agreement with the maximum uncertainties (5% for the $CO_2$ capture level, 2% for the $CO_2$ loading, and 6% for the reboiler duty) in the pilot plant results reported by Notz et al.[4] Also, Figures 3(a) to 3(d) show how the temperature profiles in the absorber and stripper, as well as the $CO_2$ composition profiles in the absorber and stripper, compare with the experimental values reported by Notz et al.[4] for the set of experiments with a constant liquid/gas ratio. It is clear from Figures 2 and 3 that the model predictions are in very good agreement with the experimental pilot plant results and hence the model may be confidently used as a basis for scale-up design within a conservative margin of $\pm 10\%$.

### 4.2. Design Philosophy Implementation in Aspen Plus

The design philosophy for the commercial-scale plants uses two criteria to determine the diameters of the absorber and stripper columns for different liquid flowrates and lean amine $CO_2$ loadings, while eqs (16a) and (16b) are, respectively, used for the absorber height and the stripper height needed for 90% $CO_2$ capture. A capture rate of 90% was adopted for the design cases in this paper because it is a commonly used basis for amine-base capture design and evaluation in publications the open literature, including special and FEED study reports. The optimum designs were arrived at based on economic analysis using Aspen Plus® Economic Analyzer, V8.4, which is based on the industry-standard Icarus Systems.[42]

The design philosophy was first implemented at pilot-scale, using the Mellapak 250Y structured packing in the absorber and stripper, and the pilot-scale design results were compared with the openly available design information for the pilot plant used by Notz et al.[4] Having validated the design philosophy at pilot-scale, it was then used directly for the commercial-scale design cases discussed in **Section 5**.

The column diameter for a given liquid flowrate was determined based on two recommended criteria for the design of aqueous amine systems, which are known to be moderately foaming. The criteria are a maximum fractional approach to flooding (or maximum operational capacity, MOC) of 0.8, and a maximum pressure drop per unit height of 20.83 mm-$H_2O$/m.[37,38] The vendor correlation for Mellapak 250Y structured packing was used for pressure drop calculation. Further, the 1985 correlation of Bravo et al.[43] was used to calculate mass transfer coefficients and interfacial area for Mellapak 250Y structured packing, while the 1992 correlation of Bravo et al.[44] was used to calculate liquid holdup. The Chilton and Colburn correlation[45] was used to predict the heat transfer coefficient for the Mellapak 250Y structured packing. The correlations used for the pressure drop, mass transfer coefficients, liquid holdup, and heat transfer coefficient calculations are built into Aspen Plus. Furthermore, with a rate-based calculation approach, the HETPs of the stages are calculated directly based on mass transfer theory. The calculated HETPs are the heights of the stages if they were to be assumed as equilibrium stages; thus, the summation of the HETPs for the stages gives the packed height of the column. An alternative way of determining the packed height is to multiply the average value of the HETPs of the stages in the packed section by the number of stages in the packed section.

The packed height needed to achieve a given degree of separation is the sum of the HETPs of the stages that will achieve the given separation, starting from the top stage (stage 1 for the absorber or stage 2 for the stripper) and ending at the stage corresponding to the extent of separation specified. However, Aspen Plus requires that the total number of stages and the inlet stream stages be specified a priori before any calculation can be executed. In order to overcome this unavoidable limitation, a calculator block was used to automatically adjust the ending stage number of the packed section to the number of stages while fixing the starting stage of the packed section at 1 for the absorber or 2 for the stripper. Furthermore, the calculator block automatically adjusts the flue gas (feed) stage, the ending stage number for the reactions, and the ending stage number for the reaction holdup. Starting with a total stage number of 2, the number of stages in the absorber was automatically increased in steps of 1, using a sensitivity block until the desired $CO_2$ capture level was achieved, which was taken as 90%. Data logging of the calculated results of interest in each "pass" was realized using the same sensitivity block that increased the number of stages. Also, with the lean $CO_2$ loading specified as a design specification for the stripper and starting with a total stage number of 10, the number of stages in the stripper was automatically stepped by 1, using another sensitivity block. In each pass, the reboiler duty was manipulated to achieve the specified lean $CO_2$ loading and the optimum stripper height was arrived at when there was negligible (less than 0.001%) or no change in the reboiler duty with further increase in the number of stages. As for the absorber, data logging of the calculated results of interest in each pass was realized using the same sensitivity block that increased the number of stages.

### 4.3. Design Philosophy Validation at Pilot-Scale

The design philosophy validation at pilot-scale followed the explanation given in the previous section and, in contrast to the model validation with explicit specification of the absorber and stripper heights, the absorber and stripper heights needed to achieve the experimentally reported $CO_2$ capture rate were determined and compared with the actual heights of the absorber and stripper. Figures 4(a) and 4(b), respectively, show how the calculated absorber and stripper heights compare with the actual heights of the absorber and stripper. The calculated heights are within $\pm 5\%$ accuracy when compared with the actual heights; thus, the validated mode is deemed to be sufficiently accurate for scale-up design, especially if the calculated results are interpreted with respect to the uncertainties in the experimental values.

## 5. SCALE-UP APPLICATIONS

The equation relating the lean amine solution mass flowrate to the amount of $CO_2$ recovered from the flue gas stream, the mass fraction of the amine in the unloaded solution ($\omega_{Amine}$), and the lean amine solution $CO_2$ loading is given by:

$$F_{Lean} = \frac{F_{FG} x_{CO_2} \Psi_{CO_2}}{100z(\alpha_{Rich} - \alpha_{Lean})} \left( \frac{M_{Amine}}{44.009} \left\{ 1 + \frac{1 - \omega_{Amine}}{\omega_{Amine}} \right\} + z\alpha_{Lean} \right) \qquad (7)$$

where $F_{Lean}$ is the mass flowrate of the lean amine solution, $F_{FG}$ is the mass flowrate of the flue gas, $x_{CO_2}$ is the mass fraction of $CO_2$ in the flue gas, $\Psi_{CO_2}$ is the percentage of $CO_2$ in the flue gas that is recovered, $M_{Amine}$ is the molar mass of the amine, $\alpha_{Lean}$ and $\alpha_{Rich}$ are, respectively, the lean amine solution $CO_2$ loading and the rich amine solution $CO_2$ loading, and $z$ is the number of equivalents per mole of the amine ($z$ is one for MEA).

The stripper reboiler duty needed for $CO_2$ stripping consists of four parts, namely: (i) the heat of $CO_2$ desorption, (ii) the heat needed for stripping steam generation, (iii) the heat needed for solvent heating, and (iv) the heat needed for condensate reflux heating, which is often neglected. Their relative contributions to the stripper reboiler duty needed for a given $CO_2$ capture rate depend on the amine flowrate and the lean amine $CO_2$ loading.

The scale-up and optimisation question that requires an answer is what combination of lean amine flowrate and lean amine $CO_2$ loading will optimize the absorber and stripper sizes as well as the stripper reboiler duty at 90% $CO_2$ capture rate? In order to answer this question for the benchmark amine for solvent-based post-combustion $CO_2$ capture, which is 30 wt% aqueous solution of monoethanolamine (MEA), a total of four commercial-scale $CO_2$ capture plants, each of which can service a 400MWe (gross) natural gas combined cycle (NGCC) power plant, a 450MWe (gross) natural gas combined cycle (NGCC) power plant, a 673 MWe (gross) subcritcal pulverized coal (PC) power plant, and a 827 MWe (gross) ultrasupercritcal pulverized coal (PC) power plant, were optimally designed in this paper. The flue gas composition and flowrate for the 673 MWe (gross) subcritical PC power plant were obtained from a 2010 report by the US Department of Energy (DOE),[46] while the composition and flowrate for the 827 MWe (gross) ultra-supercritical PC power plant were obtained from a 2004 report by the International Energy Agency Greenhouse Gas R&D Programme (IEAGHG).[47] The composition for the two NGCC cases was taken to be the same as the NGCC case in the 2010 US DOE report,[46] and it has a $CO_2$ composition of approximately 4%, which is essentially the same as the $CO_2$ composition of the flue gas used by Sipocz and Tobiesen,[27] and Biliyok and Yeung[28]. The flue gas flowrates for the two NGCC cases were estimated based on the values reported for different MWe (gross) NGCC power plants. It is important to state that the composition of the flue gas from a NGCC power plant will normally depend on the composition of the natural gas that is used, while the flowrate of the flue gas from a NGCC power plant will depend on the composition of the natural gas used, the pressure ratio of the air compressor, the temperature and pressure conditions of the main steam and reheat steam in the steam cycle, etc. The two NGCC cases considered in this paper provide meaningful comparisons with the designs reported by Sipocz and Tobiesen[27] and Biliyok and Yeung.[28] **Table 3** summarizes the conditions and compositions of the flue gas used as bases for the four design cases in this paper, while **Table 4** summarizes the basic design and economic assumptions adopted for the four design cases.

The optimum design of the absorber and stripper columns for the four cases considered in this work are summarized in **Table 5**, and they were arrived at based on process and economic analyses. It is important to note that, in line with what can be delivered by the state-of-the-art technology as documented in the publications by Reddy et al,[33-35] a maximum diameter of 18 m was used as the criterion for arriving at the number of columns needed. However, the choice of two absorbers for the 400 MWe NGCC case was arrived at based on the need for operational flexibility. The complete optimum design data, which include data for the pumps and heat exchangers, can be found in **Table S.2** in the Supporting Information.

The capital cost of the plant (CAPEX) and the operating cost of the plant (OPEX) were calculated using the Aspen Plus Economic Analyser®. The basic flowsheet shown in **Figure 1** and the Costing Template for the UK in the Aspen Plus Economic Analyser®, with default values, were adopted for the economic analyses performed. It is important to note that the CAPEX and OPEX will be higher for an actual plant because of the other equipment (including spares) that must be installed based on a hazard and operability (HAZOP) study. Furthermore, it is important to add that the costing of commercial-scale $CO_2$ capture plants could be associated with high uncertainty since there is no currently operating $CO_2$ capture plant with the same capacity as the ones considered in this paper. However, in line with a recent publication by Rubin et al.,[48] if costing assumptions are applied consistently and systematically in screening technologies for $CO_2$ capture or in screening design and operation parameters for a given $CO_2$ capture technology, we are very likely to arrive at a valid conclusion. That explains the reason for adopting default values in Aspen Plus Economic Analyser® since our primary focus is not really the accuracy of the cost values but rather, the variations of the cost values with important design and operation parameters such as the absorber and stripper sizes, the lean $CO_2$ loading, and the solvent circulation rate.

In each of the four cases considered in this work, the optimum design was taken to be the one with the least OPEX. In order to confirm the validity of using the least OPEX as a basis for the optimum design selection, further economic comparisons were performed based on annualized total cost (TOTEX), which takes both the CAPEX and the OPEX into consideration. The annualized total cost (TOTEX) is given by the following equation:

$$TOTEX = C_1(OPEX) + C_2(CAPEX)\left(\frac{i(1+i)^n}{(1+i)^n - 1}\right) \tag{18}$$

where $C_1$ and $C_2$ are scaling factors.

The annualized total cost (TOTEX) for each of the four $CO_2$ capture plants considered in this paper was calculated by assuming 20 years ( $n = 20$) of plant service life and 10% interest rate ($i = 0.1$) for three different scenarios as follows:

- TOTEX calculated without scaling CAPEX and OPEX ($C_1 = 1.0$ and $C_2 = 1.0$)
- TOTEX calculated with CAPEX scaled up by 50% without scaling the OPEX ($C_1 = 1.0$ and $C_2 = 1.5$)
- TOTEX calculated with OPEX reduced by 50% without scaling the CAPEX ($C_1 = 0.5$ and $C_2 = 1.0$)

The 50% CAPEX scale-up in the second scenario is assumed to be sufficient to account for the other equipment that needs to be installed based on a hazard and operability (HAZOP) study, as well as the uncertainty that may be present in the CAPEX value calculated by the Aspen Plus Economic Analyzer. Also it is assumed that the reduction of the OPEX by 50% in the third scenario will reduce the weight of the OPEX on the TOTEX, especially if the $CO_2$ capture plant is to operate in a location where utilities are relatively cheaper than the values used in this paper.

### 5.1. Commercial-scale MEA-based $CO_2$ Capture plants for Natural Gas Combined Cycle (NGCC) Power Plants

Figures 5(a) and 5(b) summarize the design results for an MEA-based $CO_2$ capture plant that can service a 400 MWe (gross) NGCC power plant, while

Figures 6(a) and 6(b) summarize the design results for an MEA-based $CO_2$ capture plant that can service a 450 MWe (gross) NGCC power plant. The design results for the two cases cover lean $CO_2$ loadings ranging from 0.1 mol/mol to 0.3 mol/mol, and liquid/gas ratios ranging from 0.69 to 2.68. The absorber and stripper heights, as well as the specific reboiler duties, are shown in Figures 5(a) and 6(a), respectively, for the 400 MWe plant and the 450 MWe plant. From Figures 5(a) and 6(a), it is clear that the absorber height required for 90% $CO_2$ capture increases sharply with liquid/gas ratio when the liquid/gas ratio is reduced below a certain optimum value for each lean $CO_2$ loading, and the absorber height decreases gradually if the liquid/gas ratio is increased beyond the optimum value. As the liquid/gas ratio is reduced below the optimum value it becomes increasingly difficult to achieve the rich $CO_2$ loading required for 90% $CO_2$ capture; hence the reason for the sharp increase in the absorber height for liquid/gas ratio below the optimum value. Also, the change in the absorber height with liquid/gas ratio is less pronounced as the lean $CO_2$ loading increases. These observations clearly show that arbitrary assumption of liquid/gas ratio, directly or indirectly, will most likely lead to a sub-optimal design. On the other hand, the stripper height is relatively unaffected by the liquid/gas ratio but the stripper height increases as the lean $CO_2$ loading of the MEA solution decreases, which will have an implication for the overall capital cost of the plant.

The variations of the steam required by the stripper reboiler and the cooling water (C.W.) required by the stripper condenser and lean amine cooler with liquid/gas ratio are shown in Figures 5(b) and 6(b), respectively, for the 400 MWe NGCC plant and the 450 MWe NGCC plant. From Figures 5(b) and 6(b), it is clear that the both the steam and cooling water required for each lean $CO_2$ loading decreases only marginally if the liquid/gas ratio reduces beyond the optimum liquid/gas ratio. The marginal decrease in the steam and cooling water required cannot compensate for the sharp increase in the absorber height; thus, the optimum design is not given by the liquid/gas ratio that has the minimum steam and cooling water requirement. Since there is a direct relationship between the steam requirement and the specific reboiler duty, it follows therefore that the optimum design for a given lean $CO_2$ loading does not correspond with the liquid/gas ratio that has the minimum specific reboiler duty.

The economics of the plant, which includes the overnight capital cost (CAPEX) and the operating cost (OPEX) are shown in Figures 7(a) and 8(a), respectively, for the 400 MWe and the 450MWe NGCC plants. From Figures 7(a) and 8(a), it is clear that CAPEX increases sharply when the liquid/gas ratio is reduced below an optimum value, and the sharp increase in CAPEX is due to the increase in the cost of the absorbers. Also, the OPEX increases slightly as the liquid/gas ratio is reduced below the optimum value as a result of the increase in maintenance costs, which is tied to CAPEX, despite the decrease in the total cost of the utilities (steam, cooling water and electricity) consumed. On the other hand, as the liquid/gas ratio is increased beyond the optimum value, the CAPEX decreases slightly while the OPEX increases sharply because of the sharp increase in the cost of the utilities consumed. The sharp increase in the costs of the utilities is because more electricity is consumed by the rich and lean pumps with increasing liquid/gas ratio, more steam is consumed in heating up the increasing mass of the solvent from the top inlet temperature to the bottom reboiler temperature in addition to the generation of more stripping steam as the solvent mass flow rate increases, and more cooling water is consumed in the stripper condenser and lean amine cooler. These observations clearly show that there is a trade-off between CAPEX and OPEX in the design of amine-based $CO_2$ capture plant and hence there is a need for economic analysis before the optimum design of amine-based $CO_2$ capture plant can be arrived at. Furthermore, from Figures 7(a) and 8(a), it is clear that the overall optimum design depends on the value of the lean $CO_2$ loading in addition to the value liquid/gas ratio, and that both values must be carefully chosen since the optimum liquid/gas ratios are quite different for the different lean $CO_2$ loadings.

From Figures 7(a) and 8(a), the optimum design with minimum OPEX is given by 0.2 lean $CO_2$ loading and 0.96 liquid/gas ratio for both plants. To further confirm the optimum selection based on minimum OPEX, economic evaluations that take both CAPEX and OPEX into consideration were used. Figures 7(b) and 8(b) show the annualized total cost (TOTEX), which is a combination of the OPEX and an annualized cost for the CAPEX. The results of the three different scenarios that were considered for the TOTEX are shown in Figures 7(b) and 8(b), respectively, for the 400 MWe NGCC plant and the 450 MWe NGCC plant.

Interestingly, the three scenarios follow the same trend as the OPEX and they confirm the optimum design arrived at on the basis of least OPEX.

The total cost of the plants, which include both CAPEX and OPEX, per gross MWh are 16.21 £/MWh and 16.81 £/MWh for the 400MWe NGCC plant and the 450 MWe NGCC plant, respectively, which are more or less the same. Additionally, the total cost of the plants per ton of $CO_2$ captured are 51.35 £/ton $CO_2$ and 51.44 £/ton $CO_2$ for the 400MWe NGCC plant and the 450 MWe NGCC plant, respectively, which are also more or less the same.

### 5.2. Commercial-scale MEA-based CO2 Capture plants for Pulverized Coal (PC) Power Plants

Figures 9(a) and 9(b) summarize the design results for an MEA-based $CO_2$ capture plant that can service a 673 MWe (gross) subcritical PC power plant, while Figures 10(a) and 10(b) summarize the design results for an MEA-based $CO_2$ capture plant that can service a 827 MWe (gross) ultra-supercritical PC power plant. The design results for the two cases cover lean $CO_2$ loadings ranging from 0.1 mol/mol to 0.3 mol/mol. The liquid/gas ratios range from 2.09 to 5.29 for the subcritcal plant, and 1.91 to 5.06 for the ultra-supercritical plant. The absorber and stripper heights, as well as the specific reboiler duties, are shown in Figures 9(a) and 10(a), respectively, for the subcritical plant and the ultra-supercritical plant. From Figures 9(a) and 10(a), it is clear that the absorber height required for 90% $CO_2$ capture increases sharply with liquid/gas ratio when the liquid/gas ratio is reduced below a certain optimum value for each lean $CO_2$ loading, and the absorber height decreases gradually if the liquid/gas ratio increases beyond the optimum value. Also, as for the NGCC cases, the change in the absorber height with liquid/gas ratio is less pronounced as the lean $CO_2$ loading increases. Again, these observations clearly show that the arbitrary assumption of liquid/gas ratio, directly or indirectly, will most likely lead to a sub-optimal design. On the other hand, as for the NGCC cases, the stripper height is relatively unaffected by the liquid/gas ratio but the stripper height increases as the lean $CO_2$ loading decreases, which will have an implication on the overall capital cost of the plant.

The variations of the steam required by the stripper reboiler and the cooling water (C.W.) required by the stripper condenser and lean amine cooler with liquid/gas ratio are shown in Figures 9(b) and 10(b), respectively, for the subcritical plant and the ultra-supercritical plant. From Figures 9(b) and 10(b), it is clear that both the steam and cooling water required for each lean $CO_2$ loading decreases only marginally if the liquid/gas ratio is reduced beyond the optimum liquid/gas ratio. The marginal decrease in the steam and cooling water required cannot compensate for the increase in the absorber height, especially if the large diameter of the absorber is taken into consideration; thus, the optimum design is not given by the liquid/gas ratio that has the minimum steam and cooling water requirement. As for the NGCC cases, since there is a direct relationship between the steam requirement and the specific reboiler duty it follows therefore that the optimum design for a given lean $CO_2$ loading does not correspond with the liquid/gas ratio that has the minimum specific reboiler duty.

The economics of the plant, which includes the overnight capital cost (CAPEX) and the operating cost (OPEX) are shown in Figures 11(a) and 12(a), respectively, for the subcritical plant and the ultra-supercritical plant. As for the NGCC cases, there is trade-off between the CAPEX and OPEX and the explanations previously given for the NGCC cases are equally applicable to the coal-fired cases and it will not be repeated here.

From Figures 11(a) and 12(a), the optimum design with minimum OPEX is given by 0.2 lean $CO_2$ loading and 2.93 liquid/gas ratio for the subcritical plant, and 0.2 $CO_2$ loading and 2.68 liquid/gas ratio for the ultra-supercritical plant. The higher liquid/gas ratio for the subcritical plant is because of the higher $CO_2$ captured when compared with the ultra-supercritical plant. In order to confirm the optimum selection based on minimum OPEX, further economic evaluations that take both CAPEX and OPEX into consideration were used. Figures 11(b) and 12(b) show the annualized total cost (TOTEX), which is a combination of the OPEX and an annualized cost for the CAPEX. The results of the three different scenarios that were considered for the TOTEX are shown in Figures 11(b) and 12(b), respectively, for the subcritical plant and the ultra-supercritical plant. The three scenarios followed the same trend as the OPEX, though not exactly, and they give credence to the optimum design arrived at on the basis of least OPEX.

The total cost of the plants, which include both CAPEX and OPEX, per gross MWh are 39.60 £/MWh and 30.90 £/MWh for the subcritcal plant and the ultra-supercritical plant, respectively. However, the total cost of the plants per ton of

$CO_2$ captured are 44.71 £/ton $CO_2$ and 44.19 £/ton $CO_2$ for the subcritical and ultra-supercritical plants, respectively, which are more or less the same.

## 6. CONCLUSIONS

A comparison of the optimal design results in this paper with some of the previously published designs shows that design based on values of $CO_2$ loadings and/or solvent circulation rates without techno-economic consideration may lead to a sub-optimal design for an amine-based $CO_2$ capture plant.

The optimum lean $CO_2$ loading for MEA-based $CO_2$ capture plants that can service commercial-scale power plants, whether natural gas-fired or coal-fired, is about 0.2 mol/mol for absorber and stripper columns packed with Sulzer Mellapak 250Y$^{TM}$ structured packing. Also, the optimum liquid/gas ratio for a natural gas combined cycle (NGCC) power plant with a flue gas composition of approximately 4 mol% $CO_2$ is about 0.96, while the optimum liquid/gas ratio for a coal-fired power plant can range from 2.68 to 2.93 for a flue gas having a $CO_2$ composition that range from 12.38 mol% to 13.5 mol%.

## ▪ ASSOCIATED CONTENT

### Supporting Information

**Tables S.1:** Summary of design data, including the key assumptions made, as reported by different authors in the open literature.

**Tables S.2:** Complete optimum design data for an MEA-based $CO_2$ capture plants needed by four different power plants.

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**Notes**

The authors declare no competing financial interest.

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**Table 1.** Kinetic expressions for MEA carbamate and bicarbonate reactions in the absorber and stripper.[15]

| Related Specie | Reaction Direction | Reaction Kinetics[a] |
|---|---|---|
| $MEACOO^-$ | Forward | $r_6 = 3.02 \times 10^{14} exp\left(-\frac{41.20}{R}\left[\frac{1}{T}-\frac{1}{298.15}\right]\right) a_{MEA} a_{CO_2}$ |
| $MEACOO^-$ | Reverse (Absorber) | $r_7 = 5.52 \times 10^{23} exp\left(-\frac{69.05}{R}\left[\frac{1}{T}-\frac{1}{298.15}\right]\right) \frac{a_{MEACOO^-}\, a_{H_3O^+}}{a_{H_2O}}$ |
| $MEACOO^-$ | Reverse (stripper) | $r_7 = 6.56 \times 10^{27} exp\left(-\frac{95.24}{R}\left[\frac{1}{T}-\frac{1}{298.15}\right]\right) \frac{a_{MEACOO^-}\, a_{H_3O^+}}{a_{H_2O}}$ |
| $HCO_3^-$ | Forward | $r_8 = 1.33 \times 10^{17} exp\left(-\frac{55.38}{R}\left[\frac{1}{T}-\frac{1}{298.15}\right]\right) a_{CO_2} a_{OH^-}$ |
| $HCO_3^-$ | Reverse | $r_9 = 6.63 \times 10^{16} exp\left(-\frac{107.24}{R}\left[\frac{1}{T}-\frac{1}{298.15}\right]\right) a_{HCO_3^-}$ |

[a]The reaction rate and the pre-exponential factor are in kmol/($m^3$ s), while the activation energy is in kJ/mol

**Table 2.** Summary of the models in Aspen Plus® that were used for transport properties calculations.[49]

| Property | Gas phase | Liquid phase |
|---|---|---|
| Density | PC-SAFT equation of state model | Clarke density model |
| Viscosity | Chapman-Enskog model with Wilke approximation | Jones-Dole model |
| Surface tension | - | Onsager-Samaras model |
| Thermal conductivity | Stiel-Thodos model with Wassiljewa-Mason-Saxena mixing rule | Reidel model |
| Binary diffusivity | Chapman-Enskog Wilke-Lee model | Nernst-Hartley model |

Table **3.** Flue gas conditions and compositions adopted for the design cases

| | Gas-fired (NGCC) | Gas-fired (NGCC) | Coal-fired (Subcritical) | Coal-fired (Ultra-supercritical) |
|---|---|---|---|---|
| Flue Gas Pressure (bara), absorber inlet | 1.2 | 1.2 | 1.2 | 1.2 |
| Flue Gas Temperature (°C), absorber inlet | 40 | 40 | 40 | 40 |
| | | Flue Gas Composition | | |
| $CO_2$ (mol/mol) | 0.0404 | 0.04 | 0.1350 | 0.1238 |
| $H_2O$ (mol/mol) | 0.0867 | 0.0867 | 0.1537 | 0.1221 |
| $N_2$ (mol/mol) | 0.7432 | 0.7432 | 0.6793 | 0.7108 |
| $O_2$ (mol/mol) | 0.1209 | 0.1209 | 0.0238 | 0.0433 |
| Ar (mol/mol) | 0.0089 | 0.0089 | 0.0081 | 0.0000 |

Table **4.** Design and Economic Analysis Assumptions used for the design cases in this paper

| | |
|---|---|
| **Design Assumptions** | |
| Lean MEA inlet temperature (°C) | 40 |
| MEA Concentration (kg/kg), without $CO_2$ | 0.30 |
| Striper Condenser temperature (°C) | 35 |
| Stripper Condenser Pressure (bara) | 1.62 |
| $CO_2$ Capture Rate (%) | 90 |
| Cross Heat Exchanger Temperature Approach (°C), hot end | 10 |
| Cross Heat Exchanger pressure drop (bar) | 0.1 |
| Lean Amine Cooler Pressure drop (bar) | 0.1 |
| Lean Amine Pump Discharge Pressure (bara) | 3.0 |
| Lean Amine Pump Efficiency (%) | 75 |
| Rich Amine Pump Discharge Pressure (bara) | 3.0 |
| Rich Amine Pump Efficiency (%) | 75 |
| **Economic Analysis Assumptions**[b] | |
| Steam Cost (£/ton) | 17.91 |
| Cooling Water Cost (£/$m^3$) | 0.0317 |
| Electricity Cost (£/MWh) | 77.5 |
| Plant equipment metallurgy | 316L stainless steel |

[b]Economic analysis was done using Aspen Plus Economic Analyzer®, V.8.4, and the costing template for the UK was adopted.

Table **5.** Summary of the key design results for the absorber and stripper columns

| | Gas-fired (NGCC) | Gas-fired (NGCC) | Coal-fired (Subcritical) | Coal-fired (Ultra-supercritical) |
|---|---|---|---|---|
| Gross Power plant size (MWe) | 400 | 450 | 673 | 827 |
| Flue Gas Flowrate (kg/s) | 622.2 | 725 | 892.57 | 932.42 |
| Liquid/Gas Ratio (kg/kg) | 0.96 | 0.96 | 2.93 | 2.68 |
| Absorber | | | | |
| Number of Absorber[c] | 2 | 2 | 2 | 2 |
| Absorber Packing | Mellapak 250Y[d] | Mellapak 250Y[d] | Mellapak 250Y | Mellapak 250Y |
| Diameter (m) | 11.93 | 12.88 | 16.67 | 16.92 |
| Optimum Height (m) | 19.06 | 19.99 | 23.04 | 23.74 |
| Stripper | | | | |
| Number of Stripper | 1 | 1 | 1 | 1 |
| Packing | Mellapak 250Y | Mellapak 250Y | Mellapak 250Y | Mellapak 250Y |
| Diameter (m) | 6.76 | 7.74 | 14.25 | 13.89 |
| Optimum Height (m) | 28.15 | 28.15 | 25.62 | 25.36 |
| Specific Reboiler Duty (MJ/kg $CO_2$) | 3.96 | 3.96 | 3.69 | 3.72 |

[c]A single absorber will results in diameter sizes of 16.92 m, 18.26 m, 23.08 m, and 23.91 m for the 400 MWe NGCC case, the 450 MWe NGCC case, the subcritical PC case, and the ultra-supercritical PC case, respectively.

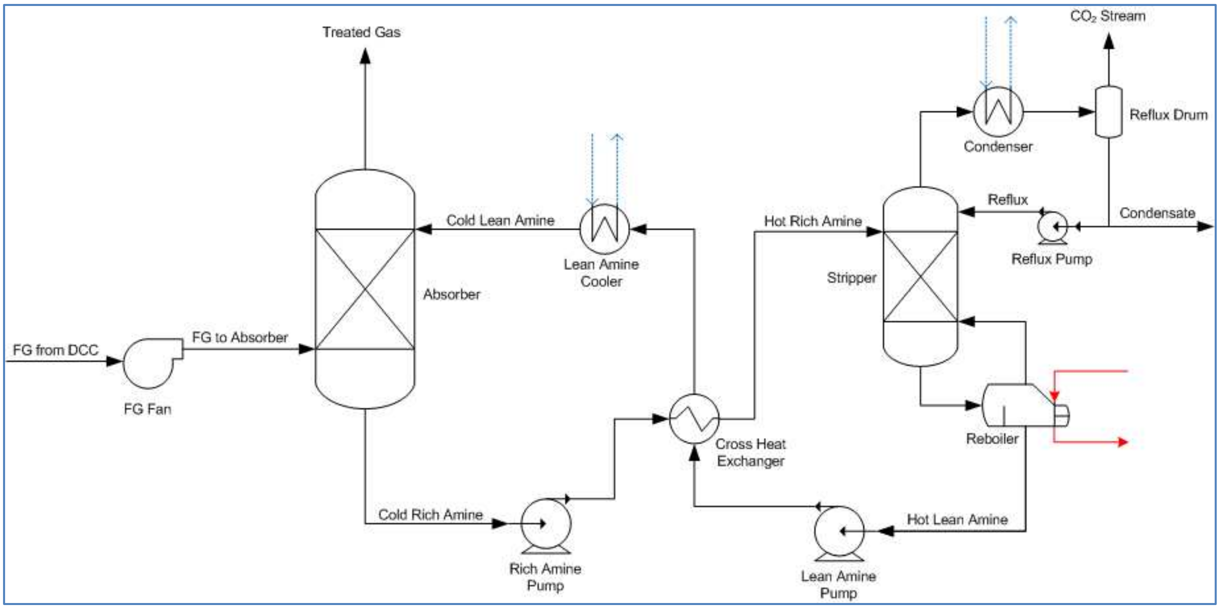


**Figure 1.** The basic flowsheet for an amine-based $CO_2$ capture process.

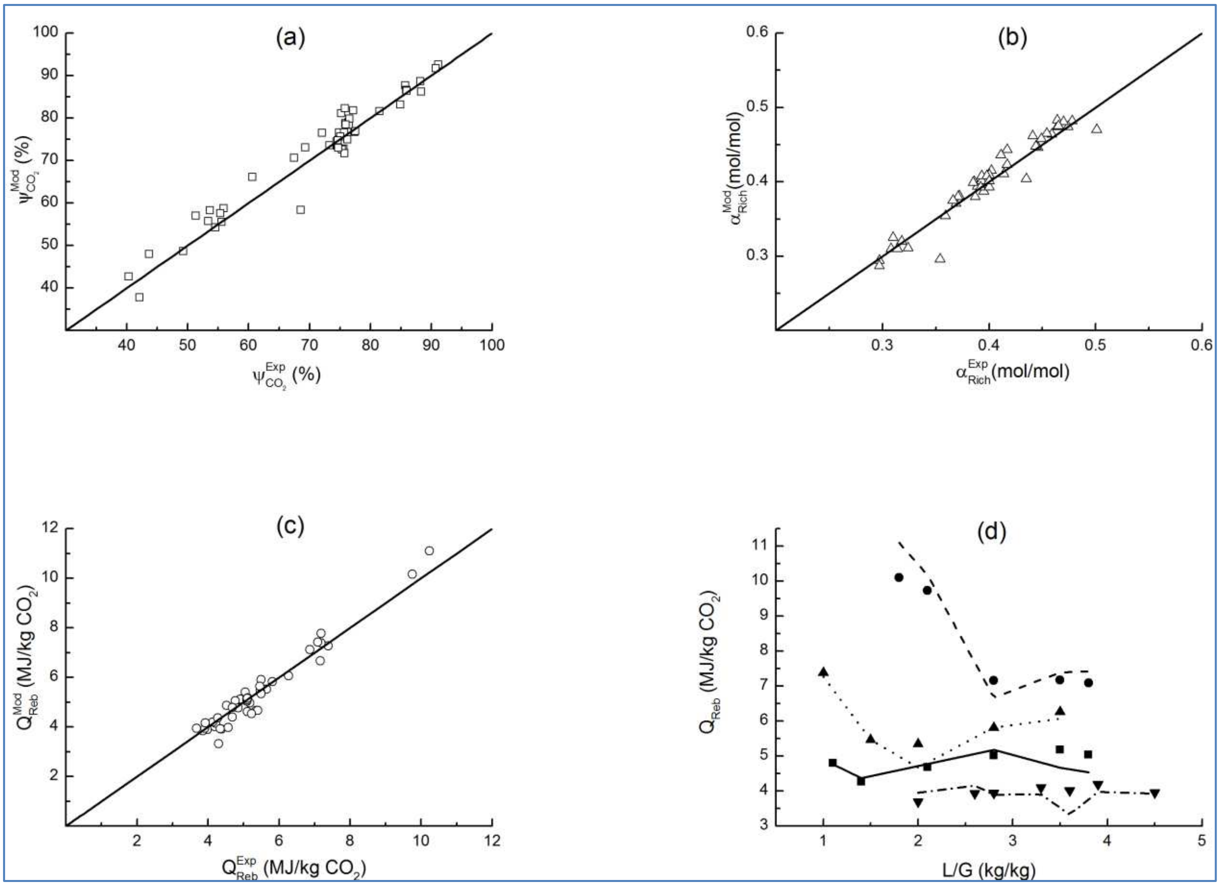


**Figure 2.** Comparison of key simulation results with the pilot plant results reported by Notz et al.[4] **(a)** $CO_2$ capture rate parity plot. **(b)** Rich $CO_2$ loading parity plot. **(c)** Specific reboiler duty parity plot. **(d)** Variations of specific reboiler duty with liquid/gas ratio for the sets of experiments designated as A.1 (G = 71.2 kg/h, $P_{CO2}$ = 54.7 mbar, and $\Psi_{CO2}$ = 76%), A.2 (G = 70.8 kg/h, $P_{CO2}$ = 53.7 mbar, and $\Psi_{CO2}$ = 88%), A.3 (G = 99.6 kg/h, $P_{CO2}$ = 57.1 mbar, and $\Psi_{CO2}$ = 75%) , and A.4 (G = 75.5 kg/h, $P_{CO2}$ = 107.5 mbar, and $\Psi_{CO2}$ = 54%). ■, (A.1); ●, (A.2); ▲, (A.3); ▼, (A.4). Lines: —, Model (A.1); − −, Model (A.2); ···, Model (A.3); − • −, Model (A.4).

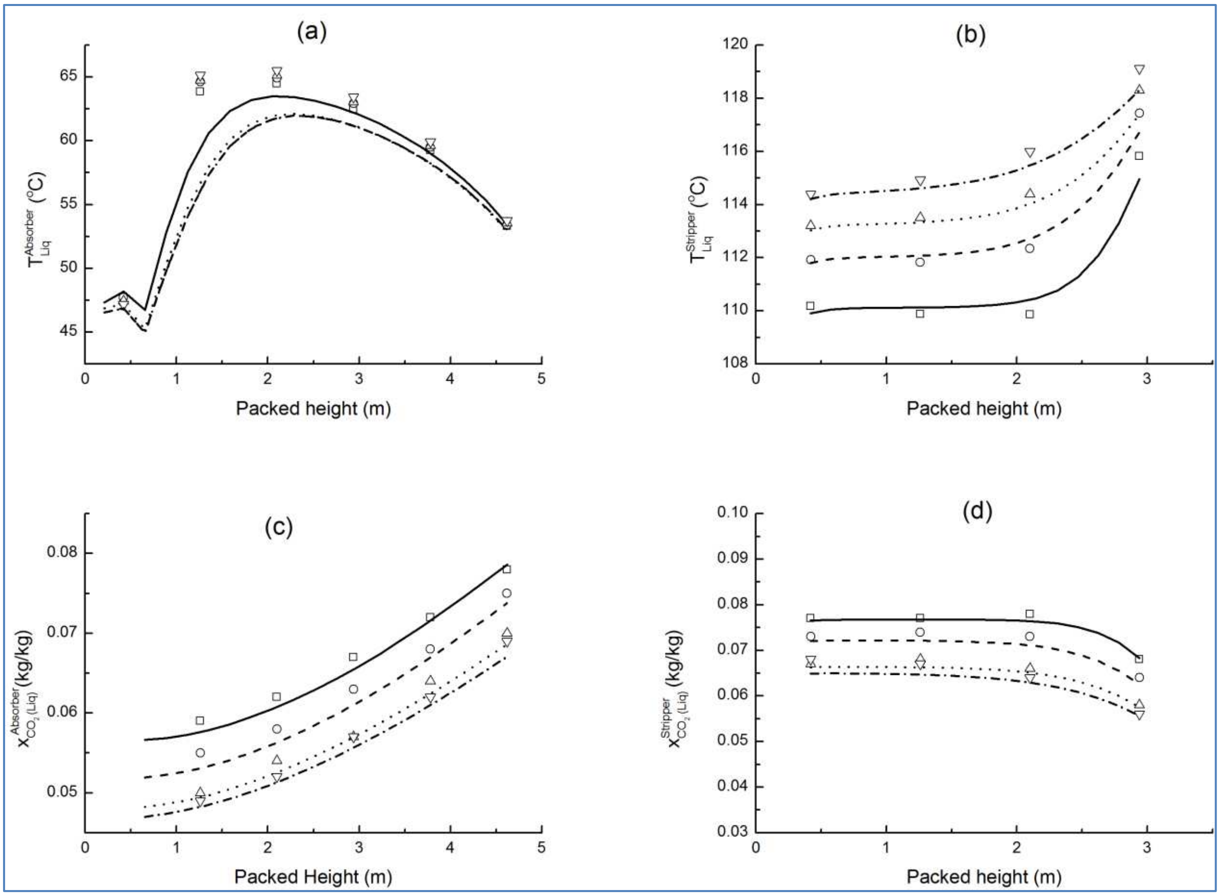


**Figure 3.** Comparison of absorber and stripper profile results at constant liquid/gas ratio (L/G = 2.8) with the pilot plant results reported by Notz et al. **(a)** Liquid phase temperature profile in the absorber. **(b)** Liquid phase temperature profile in the stripper. **(c)** Liquid phase apparent $CO_2$ mass fraction in the absorber. **(d)** Liquid phase apparent $CO_2$ mass fraction in the stripper. □, G = 55.5 kg/hr; ○, G = 72.0 kg/hr; Δ, G = 85.4 kg/hr; ∇, G = 100.0 kg/hr. Lines: —, Model (G = 55.5 kg/hr); − −, Model (G = 72.0 kg/hr); ···, Model (G = 85.4 kg/hr); − • −, Model (G = 100.0 kg/hr).

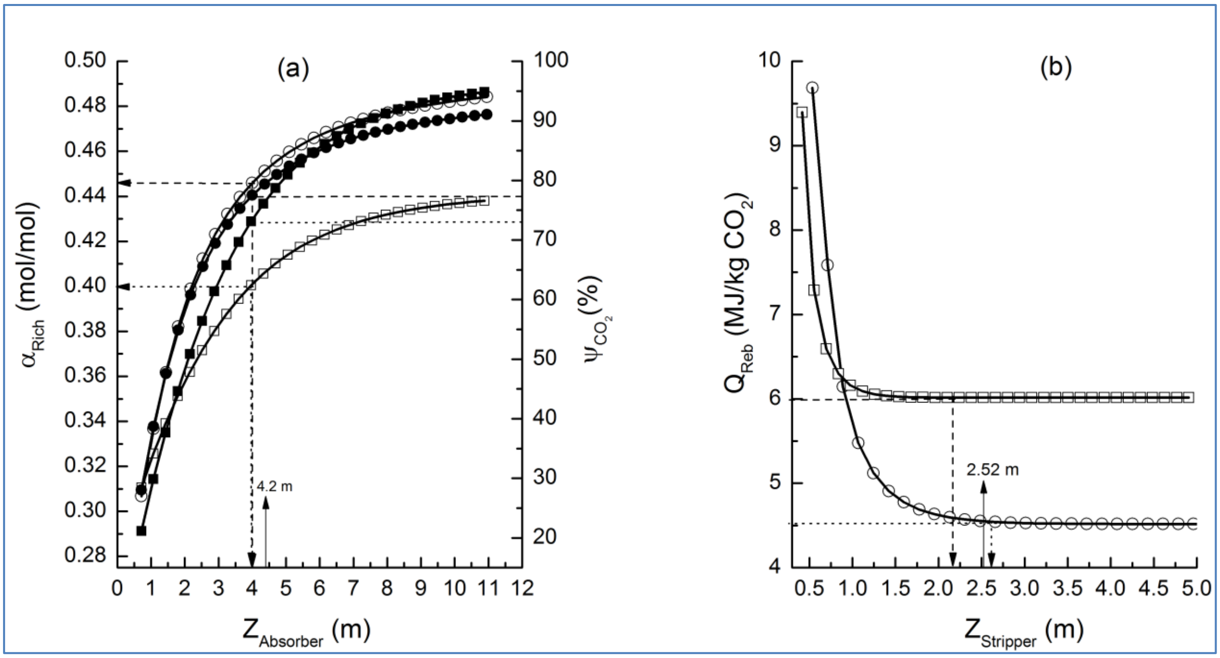


**Figure 4.** Design philosophy validation at pilot scale. **(a)** Comparison of the calculated absorber height needed for a given $CO_2$ capture level, as well as the corresponding lean $CO_2$ loading, with the actual absorber height of the pilot plant. [Symbol: □, rich $CO_2$ loading for a gas-fired case (Exp 23 in Notz et al.[4]); ○, rich $CO_2$ loading for a coal-fired case (Exp 8 in Notz et al.[4]); ■, $CO_2$ captured level for Exp 23; ●, $CO_2$ captured level for Exp 8] **(b)** Comparison of the calculated stripper height with the actual stripper height of the pilot plant. [Symbol: □, Exp 23; ○, Exp 8]

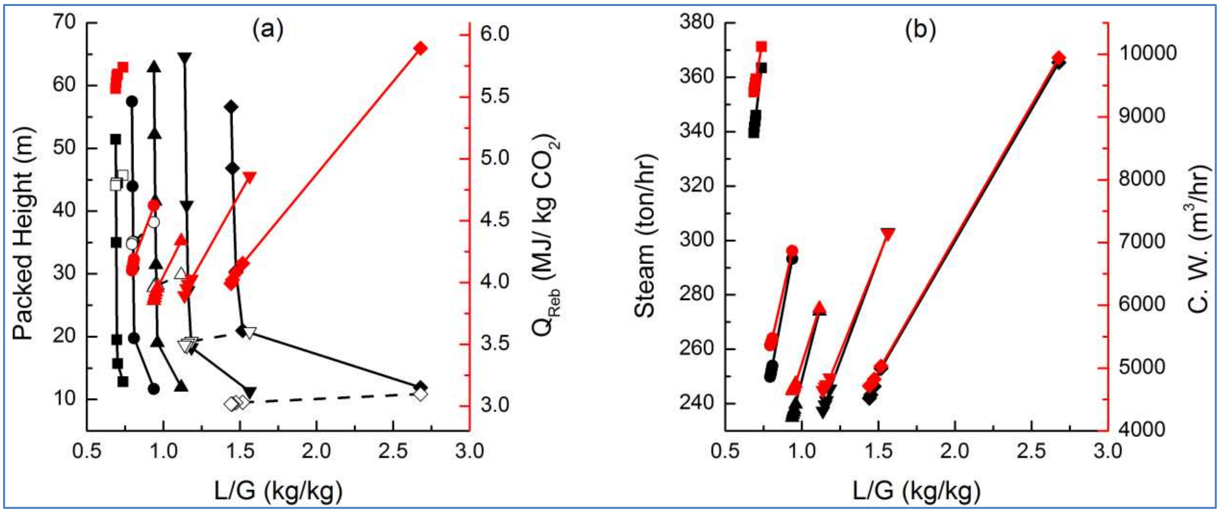


**Figure 5.** Design results for an MEA-based $CO_2$ capture plant that can service a 400 MWe (gross) NGCC power plant at 90% $CO_2$ capture rate. **(a)** Variations of absorber height (black solid lines), stripper height (black dash lines) and specific reboiler duty (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of steam requirement (black lines) and cooling water requirement (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. [Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, Δ, ▲), 0.2 $CO_2$ loading; (▼, ∇, ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].

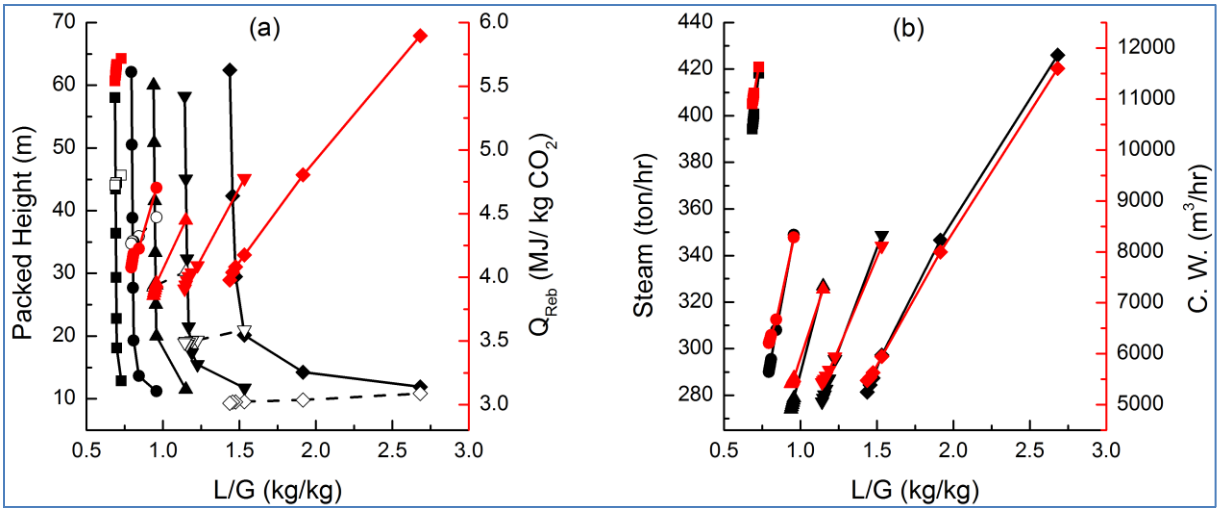


**Figure 6.** Design results for an MEA-based $CO_2$ capture plant that can service a 450 MWe (gross) NGCC power plant at 90% $CO_2$ capture rate. **(a)** Variations of absorber height (black solid lines), stripper height (black dash lines) and specific reboiler duty (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of steam requirement (black lines) and cooling water requirement (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. [Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, Δ, ▲), 0.2 $CO_2$ loading; (▼, ∇, ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].

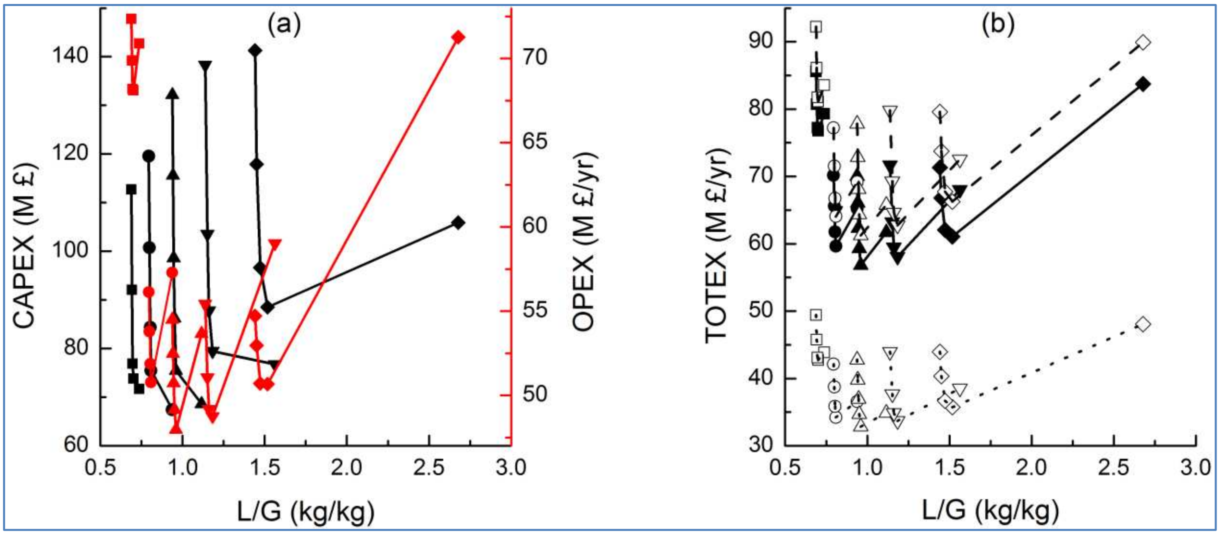


**Figure 7.** Economic results for an MEA-based $CO_2$ capture plant that can service a 400 MWe (gross) NGCC power plant at 90% $CO_2$ capture rate. **(a)** Variations of overnight capital expenditure (black lines) and annual operating expenditure (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of annualized total expenditure with liquid/gas ratio for different lean $CO_2$ loadings: solid line, OPEX + A. CAPEX; dash line, OPEX + 1.5(A. CAPEX); dotted line, 0.5(OPEX) + A.CAPEX.[ Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, Δ, ▲), 0.2 $CO_2$ loading; (▼, ∇ , ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].

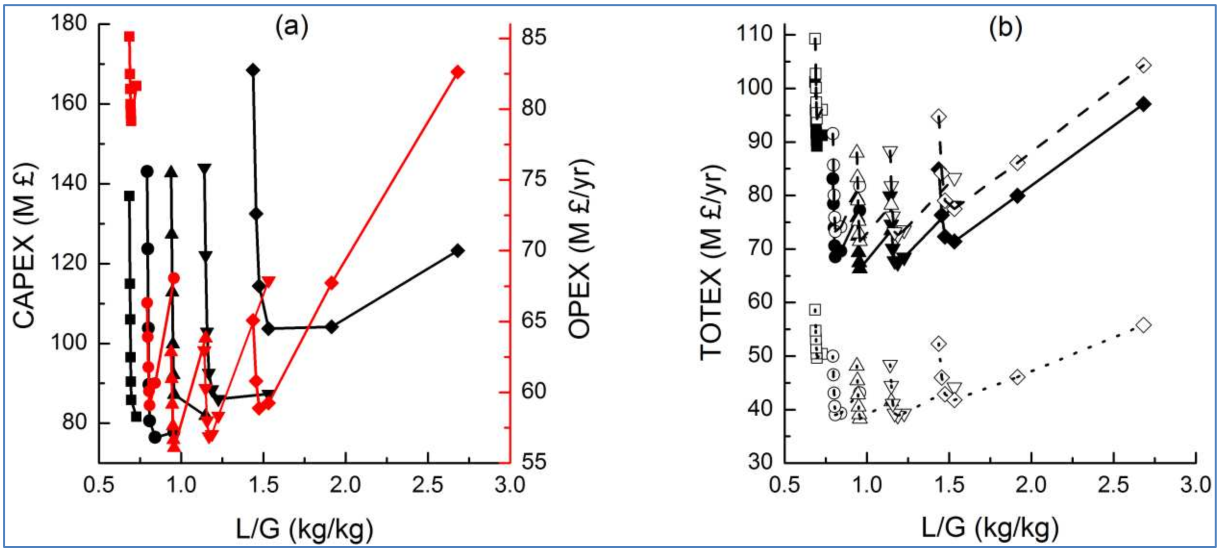


**Figure 8.** Economics results for an MEA-based $CO_2$ capture plant that can service a 450 MWe (gross) NGCC power plant at 90% $CO_2$ capture rate. **(a)** Variations of overnight capital expenditure (black lines) and annual operating expenditure (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of annualized total expenditure with liquid/gas ratio for different lean $CO_2$ loadings: solid line, OPEX + A. CAPEX; dash line, OPEX + 1.5(A. CAPEX); dotted line, 0.5(OPEX) + A.CAPEX.[ Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, Δ, ▲), 0.2 $CO_2$ loading; (▼, ∇ , ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].

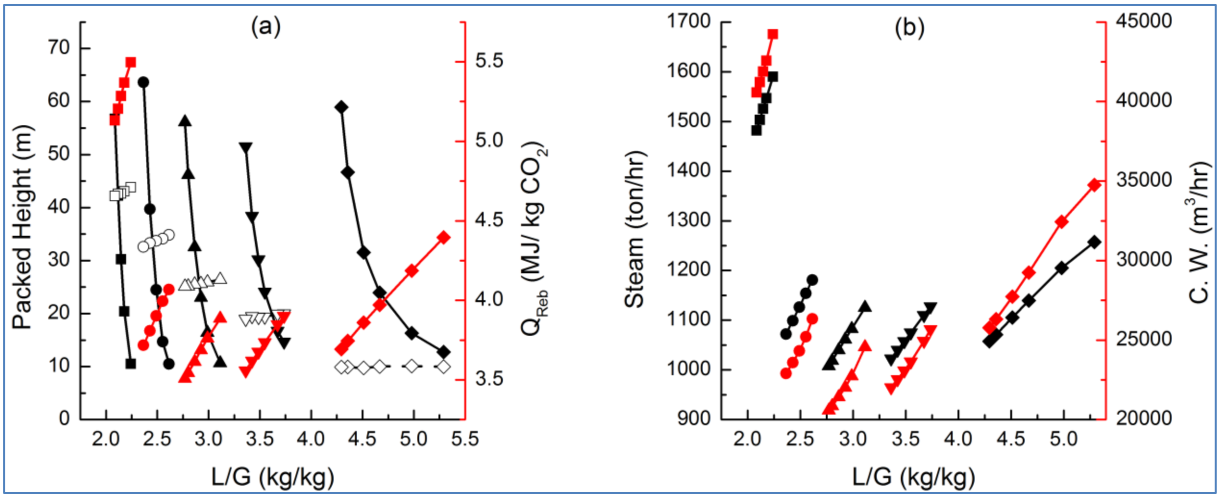


**Figure 9.** Design results for an MEA-based $CO_2$ capture plant that can service a 673 MWe (gross) subcritical PC power plant at 90% $CO_2$ capture rate. **(a)** Variations of absorber height (black solid lines), stripper height (black dash lines) and specific reboiler duty (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of steam requirement (black lines) and cooling water requirement (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. [Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, Δ, ▲), 0.2 $CO_2$ loading; (▼, ∇, ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].

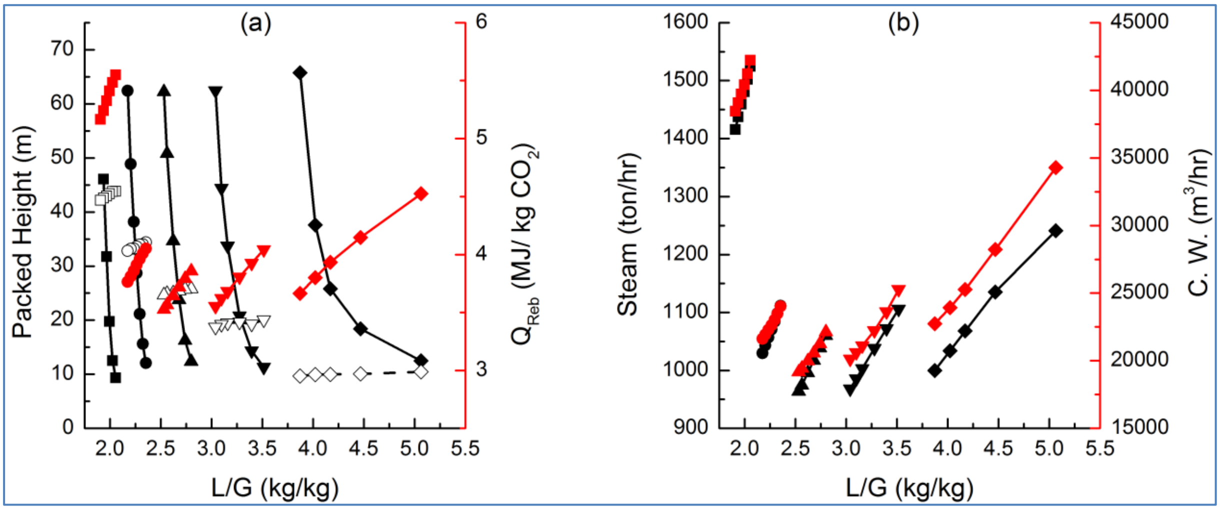


**Figure 10.** Design results for an MEA-based $CO_2$ capture plant that can service an 827 MWe (gross) ultra-supercritical PC power plant at 90% $CO_2$ capture rate. **(a)** Variations of absorber height (black solid lines), stripper height (black dash lines) and specific reboiler duty (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of steam requirement (black lines) and cooling water requirement (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. [Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, Δ, ▲), 0.2 $CO_2$ loading; (▼, ∇, ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].

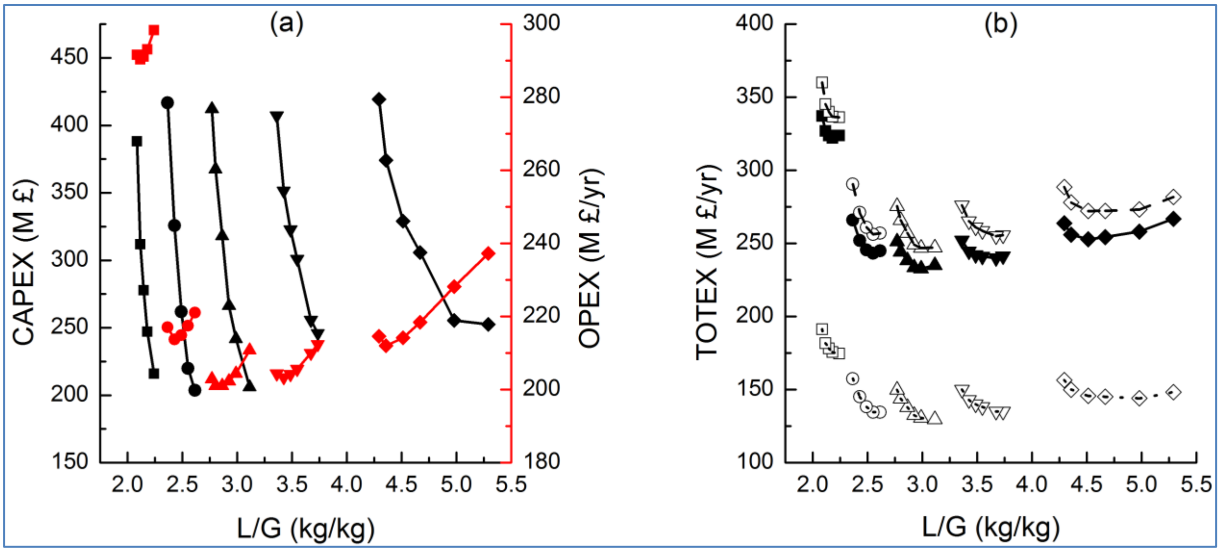


**Figure 11.** Design results for an MEA-based $CO_2$ capture plant that can service a 673 MWe (gross) subcritical PC power plant at 90% $CO_2$ capture rate. **(a)** Variations of overnight capital expenditure (black lines) and annual operating expenditure (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of annualized total expenditure with liquid/gas ratio for different lean $CO_2$ loadings: solid line, OPEX + A. CAPEX; dash line, OPEX + 1.5(A. CAPEX); dotted line, 0.5(OPEX) + A.CAPEX.[ Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, ∆, ▲), 0.2 $CO_2$ loading; (▼, ∇ , ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].

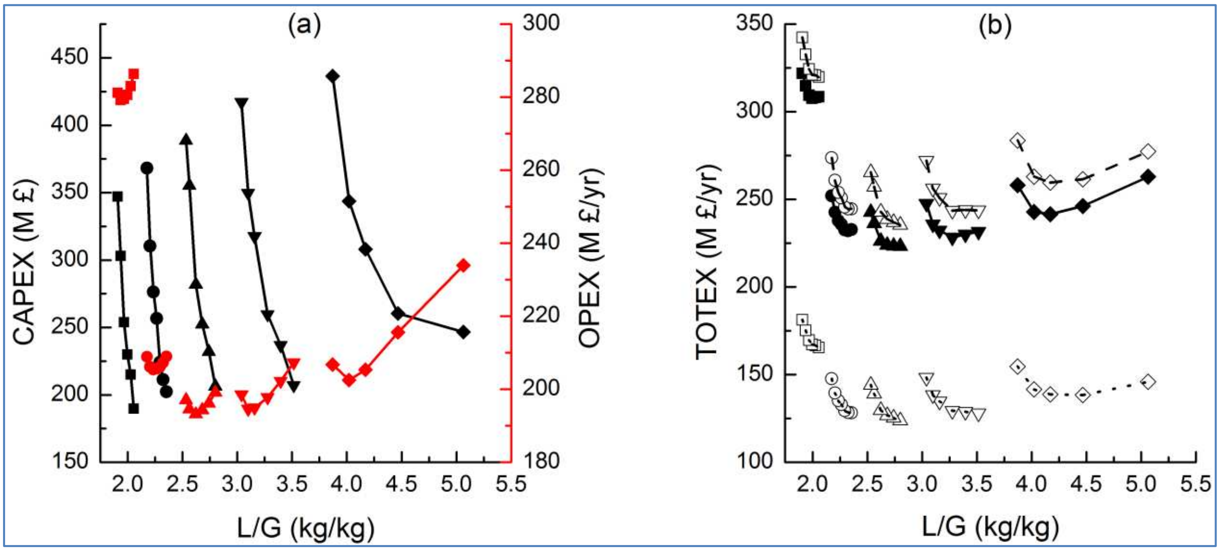


**Figure 12.** Design results for an MEA-based $CO_2$ capture plant that can service an 827 MWe (gross) ultra-supercritical PC power plant at 90% $CO_2$ capture rate. **(a)** Variations of overnight capital expenditure (black lines) and annual operating expenditure (red lines) with liquid/gas ratio for different lean $CO_2$ loadings. **(b)** Variations of annualized total expenditure with liquid/gas ratio for different lean $CO_2$ loadings: solid line, OPEX + A. CAPEX; dash line, OPEX + 1.5(A. CAPEX); dotted line, 0.5(OPEX) + A.CAPEX.[ Symbols: (■, □, ■), 0.1 $CO_2$ loading; (●, ○, ●), 0.15 $CO_2$ loading; (▲, ∆, ▲), 0.2 $CO_2$ loading; (▼, ∇ , ▼), 0.25 $CO_2$ loading; (♦, ◊, ♦), 0.3 $CO_2$ loading].