Techno-economic analysis of a CO_2 capture plant integrated with a commercial scale combined cycle gas turbine (CCGT) power plant

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Abstract

In this study, a combined cycle gas turbine (CCGT) power plant and a CO_2 capture plant have been modeled in GateCycle® and in Aspen Plus® environments respectively. The capture plant model is validated with experimental data from the pilot plant at the University of Texas at Austin and then has been scaled up to meet the requirement of the 427 MWe CCGT power plant. A techno-economical evaluation study has been performed with the capture plant model integrated with flue gas pre-processing and CO_2 compression sections. Sensitivity analysis was carried out to assess capture plant response to changes in key operating parameters and equipment design. The study indicates which parameters are the most relevant (namely absorber packing height and regenerator operating pressure) and how, with a proper choice of the operating conditions, both the energy requirement for solvent regeneration and the cost of electricity may be reduced.

Keywords: process modeling, process simulation, gas turbine, combined cycle, CCGT, post-combustion, carbon capture, techno-economic analysis

1. Introduction

1.1 Background and Motivations

Carbon Capture and Storage (CCS) is regarded as an essential technology to meet greenhouse gases reduction goals [1]. CO_2 capture with chemical absorption using amine solvent is a proven and well established technology. Despite this, CO_2 capture from exhaust gas coming from a power plant poses many technical and economical challenges. Current CO_2 capture projects involve pilot plants on a scale much smaller than required to capture CO_2 from a commercially available power plant. In September 2012, Global CCS Institute has identified 75 large-scale integrated CCS projects (LSIP) running globally [2]. An LSIP is defined by Global CCS institute as a project involving the capture, transport and storage of CO_2 at a scale of at least 800,000 tonnes of CO_2 annually for coal-based power plants or at least 400,000 tonnes of CO_2 annually for other emission-intensive industrial facilities (including natural gas-based power generation). More than half of all projects started during 2012 are located in China, and all of these are investigating enhanced oil recovery (EOR) options as an additional source of revenue. Among these LSIPs only 16 are, however, currently operating or in construction, for a global capture capacity of around 36 million tonnes per annum. These projects require investments of the order of dozens millions of Euros. It is expected that a full scale demonstration project for CO_2 capture would require over a billion dollars [3]. Accurate modeling of CO_2 capture plant, for the insight it can provide, is therefore a necessary intermediate step towards demonstrating full scale CO_2 capture. Both technical performance and costs are determinant factors to select optimal operating conditions.

1.2 Novel contributions of this paper

CO_2 capture process is, with current available technology, a very expensive and energy-intensive process. Despite this, it is gaining attention among researchers and policymakers as a short-mid term solution to contain carbon emissions from existing or yet to be built fossil fuelled power plant. However, as usual for a substantially new technology (at least at the scale required for capturing CO_2 from power plants), much resistance remains, mainly due to the uncertainty connected to actual performance and costs. Therefore, accurate modeling constitute a stepping-stone to increase confidence about CO_2 capture process. In this perspective, rate-based modeling procedure adopted by the Authors constitute, when compared to equilibrium based calculations, a superior solution in terms of accuracy and sensitivity to changes in the operating parameters. In addition to this pilot plants currently existing or, even more, large scale demonstration projects currently being built, are limited in the range of parameters that can be changed. Capture plant modeling, if based on a rigorous and trusted modeling procedure, can overcome this intrinsic limitation and following this idea a wide sensitivity analysis has been conducted on the main operating and design parameters in order to identify optimal working conditions, thus reducing uncertainty in thermal and economics characteristics of the process. Furthermore, combining capture plant commercial scale modeling with an extensive validation campaign (over a wide range of L/G ratio and process conditions) constitute an emblematic element of novelty. Summarizing, the main novelties of this article are:

a. extensive validation campaign of capture plant at pilot plant scale combined with commercial scale modeling and simulation
b. capture plant operating conditions and design parameters sensitivity analysis

2. Modeling of CCGT power plant

A commercial Combined Cycle Gas Turbine (CCGT) power plant, targeted to 427 MWe production (before capture) is modelled in GE's GateCycle® software. GateCycle® software allows an accurate modeling of design but also off-design power plant components operation. The performance of the steam cycle sections, sized for the reference non-capture case, is automatically scaled to take into account the modified pressure and temperature they will face after retrofitting to capture CO_2.

The reference commercial CCGT power plant employs a heavy duty single shaft Ansaldo Energia AE94.3A gas turbine from which exhaust gas is led to an unfired heat recovery steam generator (HRSG). The steam cycle consists of three pressure levels (124, 28 and 4.5 bar respectively) with a reheat loop. The steam is condensed in a condenser with outer water at 15°C. Deaeration is attained in the deaerator, which operates at 4.5 bar, by using low pressure steam. The condensate from the condenser is heated by means of a closed cycle loop in order to increase heat utilization from flue gas as much as possible. All the parameters required for the calculation comes from various sources: Ansaldo company private communications, GateCycle® software library and common practice for large combined cycle power plants.

3. Integration between CCGT Power Plant and Capture Plant

An exhaust gas with mass flow rate of 702 kg/s is delivered to flue gas pre-processing and consequently capture sections. Applying post-combustion CO_2 capture to a CCGT power plant requires minimal structural changes to the original cycle and is therefore regarded as the best capture option for existing power plants. Enough space should be provided for flue gas pipeline and capture related sections (notably flue gas pre-processing, CO_2 capture and compression) which should be located in the vicinity of the power plant. The main connections between power plant and the capture plant are as follows:

a. flue gas pre-processing;
b. steam draw-off from the steam turbine in CCGT power plant to feed the reboiler of the regenerator in the CO_2 capture plant;
c. condensate return from capture plant to the power plant.

The first two processes result in a reduction of electricity output from the CCGT power plant.

3.1 Flue gas pre-processing

Exhaust gases coming from the HRSG, before being sent to the capture plant, need to be cooled down to 40-50°C in order to improve absorption and reduce solvent losses due to evaporation [4]. The cooling system consists of a direct contact cooler (DCC) in which a spray of water cools down flue gases to the desired temperature level.
This process has been modelled in Aspen Plus® environment by using RadFrac block for the DCC, regarded as a two theoretical stages tower with Rashig rings packing. Flue gases are cooled down to 40°C by direct contact with a spray of water at 25°C. During the cooling process water is recovered from the flue gas because of condensation. Finally, a blower increases the pressure of the cooled flue gases to a pressure above the atmospheric level, to balance the pressure losses in the capture plant. In Figure 1, the entire Aspen Plus® flowsheet for flue gas pre-processing is presented. Assuming a blower isentropic efficiency equal to 88.5 %, compression power requirement has been found to be equal to 8,896 kW.

3.2 Steam draw-off

The steam required by solvent regeneration in the reboiler is provided by means of a steam bled from the IP/LP crossover. As a result, the LP steam turbine will see a major reduction of steam flow rate, which will result in the reduction of both its efficiency and power output. A throttled pressure configuration is used in this study. Given the reduced mass flow rate going through the LP steam turbine, its inlet pressure would drop. To guarantee a sufficient temperature (and thus pressure) for extraction, a valve has been added at IP/LP crossover. This adds pressure throttling losses to the efficiency penalty connected to reduced LP steam turbine mass flow rate and efficiency.
To avoid solvent degradation due to high temperature, the steam has to be cooled down to a temperature just above saturation with a water spray. The waste heat resulting from this process has been partially recovered by combining the steam with some of the condensate coming from the reboiler. In this way steam draw-off is also reduced. The remaining condensate is then returned to the condenser.

4. Modeling of CO_2 transportation and compression

At ambient condition, CO_2 is a gas. At a temperature between −56.5 and 31.1 °C, it may be turned into a liquid by compressing it up to the corresponding liquefaction pressure. The critical point occurs at 73.825 bar and 31.4°C. Above this critical pressure (and at temperatures higher than 60°C), only supercritical or dense-phase liquid conditions exist. If the temperature and pressure are both above the critical point, supercritical conditions are attained: CO_2 no longer exists in distinct gaseous and liquid phases, but as a dense-phase with the density of a liquid but the viscosity of a gas. At pressures above, but temperatures below critical conditions, CO_2 is a dense liquid, whose density increases with decreasing temperature. Captured CO_2 has to be transported to a suitable storage site: pipeline is the most economical method of transport in the context of CCS ([5],[6]). To allow an efficient transportation, CO_2 flow coming from capture plants has to be compressed above critical pressure. Indeed, managing of a two-phase system might be technically very hard to achieve and, moreover, a greater density allows a smaller, and thus cheaper, pipeline. A very important requirement is that pressure, all along the pipeline, should never drop below critical value. Thus CO_2 pipeline inlet pressure needs to be chosen according to a proper pressure drop estimation along the pipeline. To cover great distances, as it likely might be required for many power plants, an additional intermediate boosting station should be provided. In literature [7] a lower limit pressure of 80 bar is usually suggested to guarantee a certain margin.

Pressure loss depends on many factors, such as the pipeline diameter and length, the roughness of the pipe and CO_2 flow velocity. It might be calculated using the Darcy-Weisbach equation:

$$\Delta p = 10^{-5} f \frac{L}{D_i} \rho_{avg} \frac{v_{avg}^2}{2} \qquad (1)$$

Assuming a capacity factor, that is the ratio of nameplate power functioning over the year, equal to 75 % and 8766 hours per year (averaging over leap years), a sizing requirement of 0.918 Mtonne/yr for the pipeline has been found. CO_2 density is very dependent on pressure and temperature. Considering an average pipeline pressure equal to 95 bar and a ground temperature of 10°C, CO_2 density is equal to 810.8 kg/m^3.

Carbon steel pipes can be adopted thanks to the high degree of control over the water content of the CO_2 being transported. The pipe has been chosen from existing tabled pipes [8]: a 10 inches (0.254 m) pipe with a 0.307 inches (7.80 mm) thickness has been adopted. This gives an internal diameter equal to 0.238 m and a velocity of 1.07 m/s. With this assumptions, pressure loss over the total assumed pipeline length (100 km), is calculated equal to 17.0 bar. An inlet pressure equal to 110 bar gives a final discharge pressure of 93 bar and is thus sufficient to guarantee desired transport condition all along pipeline. If a greater distance had to be covered, pumping stations should be provided to raise CO_2 pressure as needed.

The compression to 110 bar is achieved with a four stage intercooled centrifugal compressor modelled with Aspen Plus®. Water is removed during cooling process. Peng-Robinson equation of state is used as a thermodynamic base model. In Table 1 compressor assumptions and performance are given.

5. Modeling and simulation of Capture plant

5.1 Methodology

Capture plant section model has been developed in Aspen Plus® starting from [9], to which it can be referred for more details. Absorber and regenerator columns have been modelled using a rate-based approach, which ensures higher reliability over equilibrium based one [10]. Electrolyte NRTL activity coefficient model is used to account for liquid phase non-ideality, while the Redlich-Kwong equation of state is employed for vapour. Both vapour-liquid equilibrium (VLE) and kinetic reactions are accounted for in the columns. The following set of rate-controlled reactions has been defined to represent monoethanolamine (MEA) reaction with CO_2:

$$CO_2 + OH^- \rightarrow HCO_3^- \quad (2)$$

$$HCO_3^- \rightarrow CO_2 + OH^- \quad (3)$$

$$MEA + CO_2 + H_2O \rightarrow MEACOO^- + H_3O^+ \quad (4)$$

$$MEACOO^- + H_3O^+ \rightarrow MEA + CO_2 + H_2O \quad (5)$$

They are governed by power law expressions which kinetic coefficients are given in Table 2. On the other hand, equilibrium constants for equilibrium reactions are calculated from the standard Gibbs free energy change. To reduce the solvent regeneration heat requirement, a cross heat exchanger is used to pre-heat the rich solvent stream entering the regenerator column by using cascade heat from the hot lean solvent stream coming from the regenerator itself. This would constitute a closed loop within the Aspen Plus® flowsheet and therefore would require, rigorously, to provide a tear stream for the cross heat exchanger. This has been avoided by using two different heat exchanger for the hot (lean) and the cold (rich) side of the actual heat exchanger respectively. These heat exchangers are then linked together by a heat stream, to ensure the same heat duty on the two sides. Being absorption process favoured by low temperature, a cooler is needed to further decrease lean solvent temperature. Solvent and water makeup are required to close the loop due to losses in vapour streams leaving both the absorber and regenerator columns.

An extensive validation campaign has been conducted on capture section. For this purpose, performance data from Separation Research Program (SRP) at the University of Texas at Austin pilot plant have been employed [11]. Then, capture plant has been scaled-up to meet the requirement of the commercial scale CCGT power plant. Aspen Plus® flowsheet for pilot plant is given in Figure 2.

5.2 Model validation

The pilot plant is a closed loop absorption and stripping (regeneration) facility for CO_2 removal from flue gas with 32.5 wt% aqueous MEA solution [11]. Two different kinds of packing have been adopted for the columns: Flexipac 1Y, a structured packing with a specific area of 420 m^2/m^3 and IMTP no. 40, a random metal packing with a specific area of 145 m^2/m^3. Out of the 48 experimental cases carried out in the test campaign, 12 ?? cases were chosen for validation to account for different liquid solvent to gas (L/G) molar flows ratios. Table 3 shows the process conditions for the considered cases. With reference to it, solvent loading is defined as the (molar) ratio of CO_2 to MEA. Therefore the lean loading is the loading of the (stripped) solvent stream entering the top of the absorber column and the rich loading is the one of the solvent (in which CO_2 has been captured) coming from the bottom of the absorber column. In Table 4, the overall performance of the CO_2 capture plant model are reported. With reference to it, simulation results are compared with the experimental results and with those obtained by Zhang et al. [10] from their Aspen Plus® model. Lean loading has been controlled by a design specification set in the regenerator column, and is not therefore a validation parameter. All CO_2 loadings available from pilot plant test campaign have been obtained by means of an empirical equation resulting in a 10% uncertainty level. Taking this in consideration, all the rich loading predictions of the model can be considered satisfactory. Cases 47 and 48 have the largest deviations among the selected cases, with an underestimation of the rich loading when compared with the experimental results which is however within the uncertainty range. Interestingly, this is similarly found by Zhang et al. CO_2 capture level is always lower than the experimental results, which is generally observed also in the model by Zhang et al. Only for cases 43 and 44 Zhang et al. report an overestimation of the CO_2 removal rate. Reboiler duty is slightly underestimated by the Aspen Plus® rate based model, but this is in accordance with CO_2 capture rate estimation results. Zhang et al. (2009) studied stand-alone absorber performance only, so no reboiler duty estimation is given by them.

Temperature profile estimation is probably the most important validation parameter. From temperature depends the kinetics of absorption, equilibrium phase, and fluid transport properties. The balance between the heat released from CO_2 and MEA reaction and the heat consumed in the process from various sources (like CO_2 stripping, water evaporation, heating of the streams and heat losses) produces a bulge in absorber temperature profile [15]. The position of this bulge is mainly connected to L/G ratio. When the liquid to gas ratio is low the bulge will be located near the top of the absorber and, conversely, when it is high it will be located near the bottom. In the case of CO_2 capture for CCGT power plant, given the low CO_2 concentration in flue gases, the needed L/G ratio will likely be low. In [10] three types of temperature profiles are identified. Type A is a result of low L/G ratios and is represented, in the current analysis, by Cases 43, 44, 47 and 48. Type B profile, resulting from medium L/G ratios, is represented by Cases 29, 30, 41 and 42. Finally, in the case of high L/G ratios (Cases 28, 31, 32 and 39), type C temperature profile is obtained. In Figures 3, 4 and 5 absorbers and regenerators temperature profiles provided by the rate-based model are compared to experimental results for, respectively, case 48 (low L/G ratio), Case 42 (medium L/G) and Case 39 (high L/G). An excellent match between calculation results and experimental data has been obtained and, consequently, the Aspen Plus® model reliability is thoroughly proven.

5.3 Scale-up and sensitivity analysis

Aspen Plus® is able to size packed column diameters based on the desired approach to flooding on a specified stage, starting from a (user specified) fist-guess diameter. First-guess needed solution has been obtained using the procedure described by Lawal et al. [16]. This has provided the required number of columns and their (first-guess) diameters. The obtained results are presented in Figure 6, in which absorber and regenerator diameters are represented as a function of the columns number. Due to structural limitations, columns diameter should not exceed 12.2 m (i.e. 40 feet) ([16],[17]). According to this limitation, a three-column absorber and one-column regenerator configuration was selected. A greater number of absorber would require larger capital costs and footprint without any major benefit. On the other hand one regenerator column is sufficient to strip all the rich solvent flow coming from the absorbers.

In order to lower the computational time, only one absorber column has been modelled. Assuming the same performance for all the absorber columns the output streams from the column (the vented stream and the rich solvent flow) have been opportunely multiplied to take into account the actual number of columns. In this way regenerator performance can be taken in proper account and the same happens for makeup calculation.

As a result of scale-up procedure columns number and (first-guess) diameter, as well as a reasonable solvent flow rate (and thus the corresponding L/G ratio) have been obtained.

However, to model commercial scale capture plant many other parameters are needed. Various design specifications or Calculator block have been assigned in order to ensure good capture plant performance:

a. lean solvent loading and temperature are user input;
b. lean solvent flow rate (and thus L/G ratio) is evaluated in order to obtain, for the actual operating conditions the desired (90%) capture rate. It will mainly depends on user input lean loading. Scale-up procedure result is used to provide the needed first-guess solution;

c. reboiler duty is defined in order to obtain the user input lean loading at regenerator outlet. This ensures that the solvent coming from the regenerator has, taking into account solvent and water losses, the same loading than the one entering the absorber;
d. cross heat exchanger outlet temperature is calculated to ensure user input heat exchanger approach temperature. An higher approach temperature will correspond to a lower outlet temperature of the cold stream from the heat exchanger. As a consequence, for a given lean loading target, an higher reboiler duty will be needed;
e. lean solvent cooler heat duty is evaluated in order to ensure, at absorber inlet, the user input lean solvent temperature;
f. columns diameters are sized to obtain the desired flooding percentage. However, to allow model convergence, a first-guess reasonable diameter has to be given. The other sizing-relevant parameter, the packing heights, are user input;
g. columns pressure is user input. While absorber usually operates at atmospheric pressure regeneration process is favoured by greater pressures and its operating pressure will have to be chosen accordingly. Pressure drops have been assumed for both the columns (5 kPa for the absorber and 20 kPa in the regenerator).

Considering pilot plant solvent concentration (32.5% aqueous solutions of MEA) and columns packing, a baseline commercial-scale capture plant model have been developed, scaling it up from the pilot plant model. Absorber and regenerator column heights have been set equal to, respectively, 15 and 10 m. Cross heat exchanger approach temperature has been defined equal to 10 K and regenerator pressure is 160 kPa. Most relevant user input parameters have been undergone a sensitivity analysis (varying them from baseline case) in order to highlight their influence on capture plant performance and, notably, reboiler duty requirement. Reboiler duty, despite not being the only relevant parameter, is securely the greatest contributor on techno-economic performance of CCGT power plants with CO_2 capture. In Table 5 the main results of this analysis are given. When sizing and designing the operational parameters of a capture plant, absorber and regenerator columns, on which mostly depend process economy and efficiency, need to be considered contextually to avoid to operate at a region requiring higher capital and operational expenditure. For this reason, in Table 5 the influence of the lean loading has been also investigated to determine reasonable columns design in terms of capital costs and operational performance. When a low lean loading is specified (that is, a low CO_2 concentration in the regenerated solvent has been targeted) reboiler duty (and, consequently, steam draw off) required to strip the rich solvent of CO_2 to the desired lean loading will be larger but, on the other hand, given the increased absorption capacity of the lean stream, the quantity of solvent required to attain the desired capture level is decreased. The contrary happens when an high lean loading is targeted. For lean loading greater then approximately 0.250, reboiler duty requirement will increase slightly due to the sensible heat demand of the increased rich solvent flow rate. Therefore, while L/G ratio always increases with increasing lean loading, reboiler duty will typically present a saddle tendency with respect to this parameter. In the already mentioned Table 5, considering base-case, the impact of lean loading specification on sizing requirement, can be assessed. When L/G ratio is increased the absorber diameter, in order to obtain the desired flooding percentage, is increased as well. On the other hand regenerator flooding (and thus diameter) is mainly connected to reboiler duty requirement. However, from a process optimization point of view, absorber columns will be among the two, given their larger number and size, the most determinant factor.

Absorber packing height was increased from base-case capture plant (15 m). This would require greater capital and also O&M costs connected to capture. By increasing the packing height and, as a consequence, packing volume, absorption capacity is expected to increase. From Table 5 it is clear the beneficial effect of an increase of absorber packing height on thermal energy requirement. Interesting is the fact that the optimal lean loading (0.25 $molCO_2/molMEA$) is almost unchanged from base-case. With 25 m of absorber packing height reboiler duty requirement is decreased, considering the optimal lean loading, by 31%. A further increase of absorber packing height to 30 m would only marginally increase this benefit (-33% in reboiler duty, if compared to base-case capture plant) and, on the other hand, would increase further capital and O&M costs. From column sizing point of view an increase of absorber packing height is very relevant on, as easily expected, absorber requirement but, as well, it is a determinant factor on regenerator sizing. This is mainly due to the lower reboiler duty requirement granted by an increase in the absorber height, on which, as previously shown, regenerator diameter is mainly related. A packing height equal to 25 m ensures a reduction, compared to base-case, of 5% in absorber diameter and of 20% in regenerator one.

Regenerator operating pressure was changed from base-case capture plant. Notably three different values of this quantity where considered: 160 kPa (base-case), 185 kPa and 210 kPa, assuming for everyone of them a pressure drop through the regenerator column equal to 20 kPa. Increasing operating pressure would require greater pumping equipment along major pumping operating costs. From thermal point of view, an increase in regenerator operating pressure corresponds to an increase in driving force and thus a beneficial effect on CO_2 mass transfer rate through the regenerator column is expected. Thermal energy requirement has been proven to decrease linearly with regenerator operating pressure. Notably with a pressure equal to 210 kPa reboiler duty is lower by 9% if compared to base-case capture plant. While no effect on absorber diameter has been found, a regenerator pressure equal to 210 kPa led to a 7% reduction in regenerator diameter requirement. Again, the optimal lean loading is not significantly changed from base-case capture plant.

Cross heat exchanger provides pre-heating to the cold rich solvent by means of cascade heat from the hot lean solvent. So, a decrease in reboiler duty is expected when cross heat exchanger approach temperature is decreased. On the other hand this would require larger heat exchanger surfaces and thus equipment costs. The beneficial effect on thermal energy requirement is however in percent terms very limited. In correspondence to the optimal lean loading (0.250 $molCO_2/molMEA$) reboiler duty decreases by approximately 2% as the approach temperature decreases from 10 K (base case) to 5 K. Interesting is the fact that, even slightly, a decrease of this temperature difference led to an increase (2%) in regenerator diameter requirement. This is believed to be related to the fact that, by decreasing approach temperature, the temperature of the solvent entering the regenerator column will be affected (notably increased) and thus will be affected flooding capacity on base (entering) stage on which column sizing depends.

As for absorber one, by increasing regenerator packing height column performance will be improved and equipment costs increased. Reboiler duty is decreased by 1.5% from base-case by the adoption of 15 meters of packing. A further increase in column packing height doesn't bring any significant benefits. Regenerator diameter, in correspondence with the optimal lean loading (from thermal energy requirement point of view), is only slightly decreased (0.6%) from base-case. No effect on absorber diameter has been proven.

6. Thermo-economic performance

Considering the sensitivity analysis previously shown it has been possible to identify an improved set of capture plant operating parameters. In Table 6 the main equipment design parameters are given. Columns packing heights and regenerator pressure have been increased, given the great influence they have on reboiler duty requirement, while heat exchanger approach temperature has been left unchanged from base case capture plant, given

its minor influence on thermal energy requirement. Such a configuration securely allows a reduction on reboiler duty requirement if compared to base case capture plant. As for the previously analyzed cases reboiler duty requirement shows a saddle tendency with respect to lean loading (see Figure 7). The minimal energy requirement condition (which was 0.250 $molCO_2/molMEA$ for base case) is here between 0.200 and 0.250 $molCO_2/molMEA$ of solvent lean loading. However, despite reboiler duty is securely the most important thermal energy requirement of capture plant (to which correspond the greatest efficiency penalty), net power production is not the only parameter which affects the cost of electricity. As already shown when the lean loading is increased solvent flow rate is increased too. Thus, equipment capital cost and O&M costs are expected to be greater. For this reason, in the already mentioned Figure 7, the cost of electricity is also depicted. Cost of electricity depends on both thermodynamic performance and total annual revenue requirement (TRR) of the integrated power plant and CO_2 related sections (flue gas pre-processing, capture section and CO_2 compression). Major details on the integrated modeling and cost related assumptions are given in [18]. Considering the cost of electricity too as a decision parameter the optimal lean loading (see Table 6) has been identified equal to 0.200 $molCO_2/molMEA$. To better understand this, in Figure 8 net power production and capture plant related total annual revenue requirement as a function of lean loading are shown. It can be noted as, while net power plant power production is mainly dependent on reboiler duty (blower and CO_2 compression power requirement are unchanged), capture plant capital annualized and O&M costs, represented by the TRR parameter, increase almost linearly with lean loading.
As a result, reboiler duty has been identified to be equal to 4.1 $GJ/tonneCO_2$. To satisfy this requirement a large amount of steam need to be extracted from power plant, considerably reducing its thermal efficiency. Power plant net power production is reduced, from 427 MW of no-capture case, to 368 MW, mainly due to steam draw off. This, in combination with the capital and O&M costs connected to capture related sections, gives for the integrated CCGT and capture plants a levelized cost of the electricity equal to 68 €/MWh, increased by 47% when compared to reference (no capture) power plant.

7. Conclusions

While the technology to capture CO_2 from exhaust gas exists and is viable, it still needs to be deployed on commercial scale power plant. This makes modeling and simulation an invaluable tool for investigating CO_2 capture process integration at commercial scale. Very often capture process is simulated in a simplified way or not properly validated. Rate-based approach, as adopted for CO_2 capture section, ensures a higher reliability over traditional equilibrium based one. Experimental results from pilot plant facility at the University of Texas at Austin represent an invaluable source of insight on post-combustion capture by means of MEA solvent. An extensive validation campaign based on these data has allowed to thoroughly assess model validity. A sensitivity analysis has been conducted to prove how, with a proper choice of capture plant key operating parameters and equipment design, capture plant thermal energy requirement for solvent regeneration might be reduced to approximately 4 $GJ/tonneCO_2$. It has been demonstrated as the most economical solution is not the one with the lowest thermal energy requirement, being capture plant related costs also very relevant on the cost of electricity, and thus proving the need to optimize capture plant from both the thermal and economical point of view.

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Symbols

Δp	Pressure loss [bar]
f	Darcy friction factor [-]
L	Pipeline length [m]
D_i	Pipeline internal diameter [m]
ρ_{avg}	average CO_2 mass density [kg/m3]
v_{avg}	average CO_2 velocity [m/s]