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An Automated CO_2 Capture Pilot Plant at ULiège: A Platform for the Validation of Process Models and Advanced Control

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ABSTRACT

The deployment of CO_2 capture technologies at a large scale will largely benefit from the knowledge acquired during pilot testing. A mobile CO_2 capture pilot unit is currently being designed at the University of Liège. Here, the pilot plant is introduced, and the column sizing results are presented. The sizing was performed with a process model built in Aspen Plus. Overall, the pilot installation is expected to serve for process model validation, data collection and technology de-risking while assisting Belgian industries in their transition towards carbon neutrality.

Keywords: CO2 capture, MEA, pilot, ASPEN PLUS

INTRODUCTION

As the European Union accelerates its efforts to decarbonize society, pilot installations play a pivotal role in validating both emerging and established technologies within the field of carbon capture (CC) [1]. Numerous CC pilot projects have been carried out in European countries such as the Netherlands, UK, Norway, Poland, Germany, Denmark, and Switzerland. This extensive testing has yielded parametric studies, validation of steady state and dynamic models, characterization of process emissions, determination of solvent corrosion behavior and degradation, comparison of different solvents, testing of different analytical tools, improved process modifications, and studies on control strategies [2,3]. Nevertheless, the high uncertainty across the whole CO_2 value chain prevents the large-scale deployment of such technologies.

Currently, supported by the European Union's Recovery and Resilience Facility, the Products, Environment, and Processes (PEPs) group at the University of Liège (ULiège) in Belgium is leading the construction of several key pilot installations, including a mobile CO_2 capture pilot plant, a CO_2-to-kerosene conversion unit, and a direct air capture (DAC) test bench.

The mobile CO_2 capture pilot plant will utilize chemical absorption to capture about 1 ton of CO_2 per day. The unit has been designed to operate with the benchmark solvent monoethanolamine (MEA). However, other solvents such as carbonates can be employed too. As shown in Figure 1, the installation is housed within three 20-foot shipping containers to support easy transport and deployment at various industrial locations. Two adjacent maritime containers enclose most of the process equipment as well as part of the electric cabinets. From these containers protrude the main columns. To render the unit more flexible, it comprises a gas pretreatment section that includes dust, NO_x, and SO_x removal by bag filters, selective catalytic reduction, and alkaline scrubbing, respectively. These pretreatment steps are crucial to ensure the long-term performance of the unit by mitigating solvent degradation [1]. The plant is fully electrified, and its maximal power consumption is estimated to be 100 kW. Moreover, a heat pump is integrated into the system to enhance energy efficiency by recovering heat from the condenser and upgrading it for reboiler use. The automation and control system are centralized in the third container. The unit features a high degree of automation for remote and extended operation. This level of automation supports continuous data collection, which is essential for validating computational models and applying advanced control strategies, such as

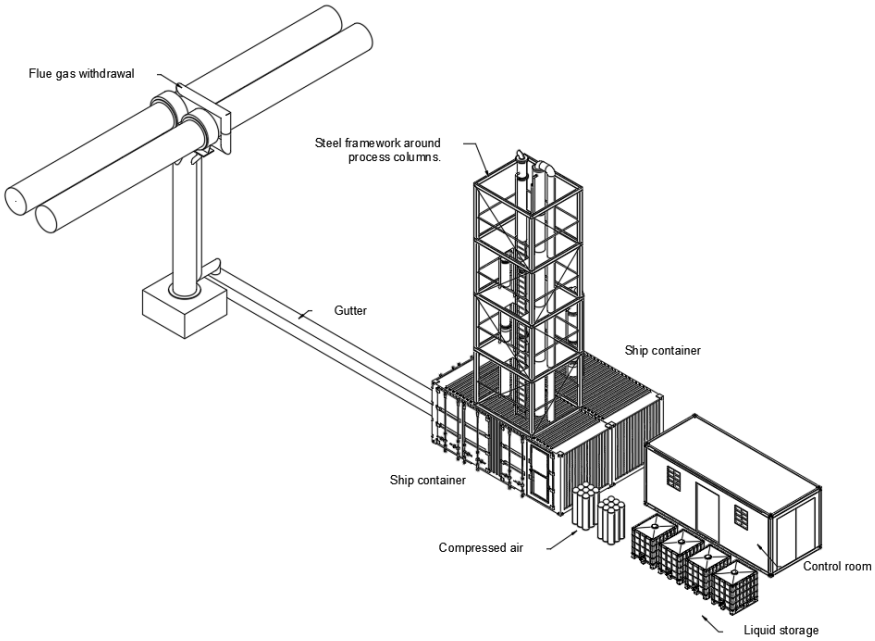

Figure 1. 3D plan of the CO_2 capture unit and its connection to the flue gas pipelines of the central heating station of the Sart-Tilman campus.

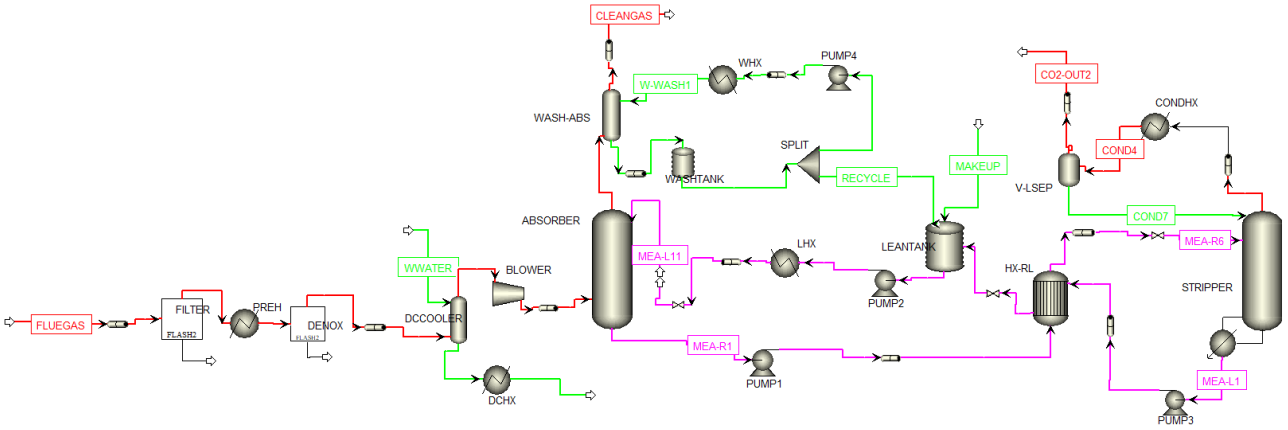

Figure 2. CO_2 capture pilot model in Aspen Plus.

machine learning algorithms [4,5]. Extended operation provides critical information to test solvent stability, understanding corrosion behavior, and refining steady state and dynamic process models—key factors for scaling up CC technology.

Commissioning and the first experimental campaigns are expected to take place at the central heating station at the Sart Tilman campus of ULiège in early 2026. Here, the pilot plant will treat a slipstream of the flue gases produced by a 7 MW_{th} biomass cogeneration unit. All details of the unit, including the simulation files, PFD, P&ID, 3D plans, HAZOP tables and equipment, accessories, and instrumentation specifications will be made available after the commissioning and first experimental tests to facilitate the quick deployment of similar installations.

This work aims at introducing the model used to design the pilot and the main sizing results.

MODEL DESCRIPTION

A model was built in Aspen Plus V11 to perform the basic engineering, see Figure 2. The thermodynamic model was set as the standard ENRTL-RK method [6]. Direct contact cooler, absorber (absorption and washing sections), and stripper (desorption and washing sections) were simulated using Rate-based modeling [6]. To simulate the reaction of CO_2 with the solvent, the following set of reactions were considered [6].

$$CO_2 + MEA + H_2O \rightarrow MEACOO^- + H_3O^+ \quad (1)$$

$$MEACOO^- + H_3O^+ \rightarrow CO_2 + MEA + H_2O \quad (2)$$

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

$$HCO_3^- \rightarrow CO_2 + OH^- \quad (4)$$

$$MEA + H_3O^+ \leftrightarrow MEAH^+ + H_2O \quad (5)$$

$$HCO_3^- + H_2O \leftrightarrow CO_3^{2-} + H_2O^+ \quad (6)$$

$$2H_2O \leftrightarrow H_3O^+ + OH^- \quad (7)$$

Reactions 1-4 are kinetic reactions whereas the rest are fast enough to consider them equilibrium reactions. Two different sets of values of the pre-exponential factors and activation energies were used for the absorber and stripper [6]. Some parameters of the rate-based model were tuned following the recommendations from [7] while the rest were left with the default values. A discretized film was only considered for the liquid phase. The number of discretization points was 5, whose thicknesses were fixed with a discretization ratio (ratio of film thickness between two adjacent discretization regions) of 10. Segments closer to the gas-liquid interface are thinner. The selected type of discretization option is geometric as it is expected to provide a better reproduction of CO_2 concentration profiles in the reactive film with fewer points. The reaction condition factor was 0.9 because of the fast reaction rate between CO_2 and MEA. The flow model was set to mixed to improve conversion.

The washing section of the absorber was modeled separately. Despite being inside the same vessel, this is consistent with the actual process as the washing section features a liquid collector with total draw-off to avoid mixing washing water and solvent inside of the absorber. On the other hand, the washing section of the stripper is modeled in the same block because of the presence of a collector-distributor above the first desorption bed. Both absorber and stripper were modeled using 20 stages while the direct contact cooler and the absorber washing section with only 10.

To facilitate conversion, the solvent loop is broken in the model. Consequently, the lean solvent loading fed to the absorber was fixed to the optimal value (determined by a sensitivity analysis using Equilibrium-based modeling) and a Design Spec (DS) was utilized to vary the lean solvent flow rate to achieve a given capture rate (usually 90%). Another DS was introduced to vary the reboiler duty of the stripper so that the amount of CO_2 recovered matches that captured in the absorber. Similarly to [8], the water balance in the solvent loop was kept by varying the temperature of the water fed to the washing section of the absorber. Thanks to these three DS, the composition of the lean solvent at both sides of the broken loop was very similar.

The model also includes most of the process pumps, piping, and gas filters. Due to the limited temperature of the gases fed to the pilot, an electric gas heater is going to be installed to ensure the effectiveness of the SCR filters. Finally, the CO_2 purification and compression sections are excluded from the model because they are not currently present in the pilot design.

MODEL VALIDATION

Firstly, the reliability of the thermodynamic model was challenged by examining the prediction of the CO_2-MEA-H_2O equilibrium at temperatures relevant for the absorber and desorber. As shown in Figure 3, the model provided an accurate determination of the CO_2 loading in MEA solutions in the required temperature range.

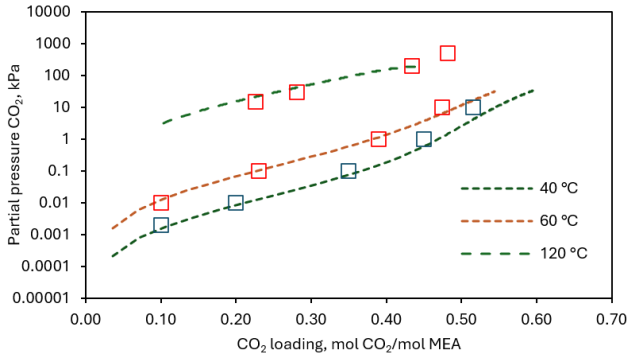

Figure 3. Validation of CO_2-MEA-H_2O equilibrium. Experimental data (□) was obtained from [9].

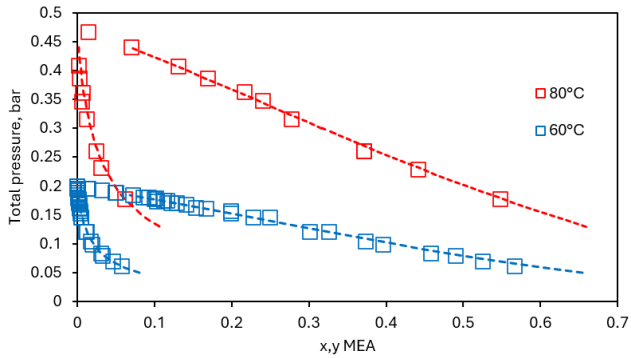

Figure 4. Validation of the MEA-H_2O VLE equilibrium. Experimental data (□) was retrieved from [10,11].

MEA emissions are a concern from both an economic and safety point of view. Amine losses must be compensated by an MEA make-up stream. Additionally, MEA is a hazardous chemical for human health. The main points of solvent emission in the plant are the gaseous streams leaving the top of the absorber and stripper. A common solution to tackle the emissions is to install a water-wash section and a demister on the top of both columns. Experimental data from the MEA-water vapor-liquid equilibrium (VLE) was validated with the simulation tool before addressing the sizing of washing sections. Two temperatures were examined, 60 and 80°C. For both, there is a good agreement between the

experimental data and the model, see Figure 4. To further evaluate the capacity of the simulator to predict the emissions of the absorber and stripper, a literature survey has been conducted to identify MEA emission data from pilot plant installations. Measurements of total emitted MEA (in the form of vapor MEA, entrained droplets, and aerosols) were performed at the TNO's CO_2 capture pilot plant [12]. The MEA emission values were around 407-466 mg/STP m^3 (dry gas) right after the absorber. A water wash step and a conventional demister can reduce the total emissions to about 46%. A Brownian Demister Unit was necessary to further diminish the emissions to ~99%. From the total emissions encountered at the absorber outlet (before the water washing), it is expected that about 200 mg/STP m^3 are due to vapor MEA (gas temperature 30-40°C). This value is in the same range as the experimental volatility data when applied to the absorber top conditions, i.e., a concentration of vapor MEA of 285 mg/Nm^3 (50°C, 0.25 mol CO_2/mol MEA for the lean solvent loading) [13]. It is difficult to validate the data from the TNO's CO_2 capture pilot plant as the lean solvent loading is not provided and certain data must be traced from different documents. Nevertheless, a few simulations indicated that the predicted vapor MEA emissions on top of the absorber are in the same range.

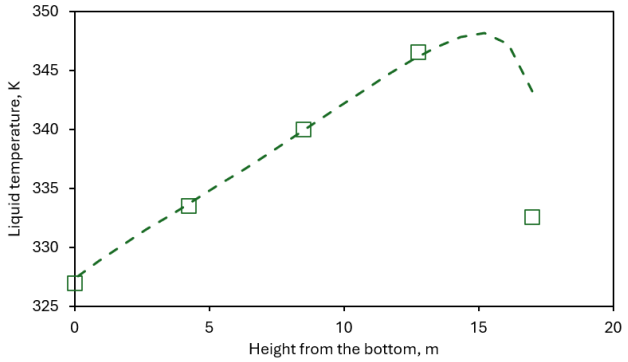

Figure 5. Validation of temperature profile in the absorber. Experimental data (□) was obtained from [15].

Table 1: Validation of absorber outputs. Data from [14].

	Experimental	Model
pCO_2 in, kPa	2.69	2.62
pCO_2 out, kPa	1.13	1.22
Rich loading, mol/mol	0.40	0.40
T gas out, °C	54.4	54.5

The capacity of the model to predict the behavior of the absorber was evaluated by comparing experimental data from pilot installations to simulated results. Two sets of data were utilized for the validation: Run 12 by Tobiesen et al. at Gløshaugen pilot (SINTEF-NTNU), Norway [14] and Run 1-A2 by Razi et al. at the large-scale Esbjerg CESAR pilot plant, Denmark [15]. Particularly, the absorber outputs and the temperature profiles along the column. The data shows the typical temperature bulb in the absorber temperature profiles caused by the exothermic CO_2 absorption.

Similarly, for the desorption unit, the simulation was validated employing data from two different installations: Run 1 by Tobiesen et al. at Gløshaugen pilot (SINTEF-NTNU), Norway [4], and Run 47 by Dugas at the pilot plant from the University of Texas at Austin [16]. The data from the latter pilot was retrieved from [7].

As portrayed in Figures 5 and 6 and Tables 1 and 2, the model was able to reproduce the experimental results to a certain extent. Increasing the number of column segments, the number of liquid thin film discretization points or changing the flow model did not improve the results significantly, indicating that the uncertainty could be coming mainly from the reliability of the kinetic data or the quality of the inputs for the modeling validation.

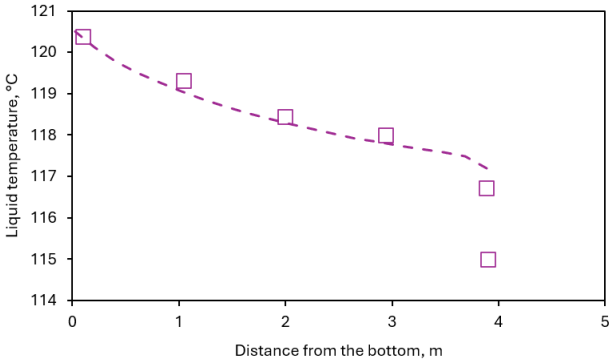

Figure 6. Validation of temperature profile in the stripper. Experimental data (□) was obtained from [4].

Table 2: Validation of stripper outputs. Data from [7].

	Experimental	Model
CO_2 product, kg/h	92	101
Lean loading mol/mol	0.29	0.24

Because of the limitations on data availability, validation results were obtained by modeling standalone equipment instead of the complete flowsheet. However, the latter was employed for the results discussed in the following sections.

COLUMN SIZING

Packed columns typically operate in the range between 30-80% of the gas flooding velocity. The range of flue gas flow rates fed to the absorber was determined based on the desired capture scale (1 ton CO_2/day) and the CO_2 concentration encountered in the fumes produced by the biomass boiler of the Sart Tilman campus (about 10%$_{V\ dry}$). Considering the design (maximum) flue gas flow rate, 360 kg/h, and the available ASME pipes,

the absorber diameter has been fixed to 0.25 m allowing the column to operate at a flooding capacity between 50-85% in a broad range of solvent flow rates, 300-1000 kg/h. The gas flow rate could be varied between 180-360 kg/h without flooding the column (780 Kg/h solvent). Note that the flooding conditions are default parameters in Aspen Plus. The diameter of the stripper is equal to the absorber. In this regard, the stripper is oversized. However, this guarantees a larger flexibility of the vessel and the possibility of operating it under a slight vacuum.

To determine the required absorber height, the lean solvent concentration has been set to 5%$_w$ CO_2 (~0.23 mol CO_2/mol MEA) which is close to the optimum observed in the literature [17,18]. With the specified lean solvent conditions, a smaller column height can deliver the same performance at the expense of higher solvent flow rates than a taller one. Since the amount of solvent affects the reboiler duty, larger columns would lead to lower energy requirements in the stripper. Thus, there is a trade-off between CAPEX and OPEX. Similarly, a shorter stripper would require more energy to regenerate the solvent. The required packing height in both absorber and stripper was estimated by a sensitivity study. As shown in Figure 7, increasing the packing height reduces the specific reboiler duty. However, once a certain value is attained, further increases only lead to marginal improvement. Although it was not considered in the model, in practice, to limit the overall height of the pilot, the required packing in the absorber was distributed in two vessels. Thus, there are two absorbers connected in series in the pilot unit. The desorber height influences the required reboiler duty mildly, see Figure 8. As indicated in the figure, the diminution in energy consumption becomes marginal after 5 m of height.

In the model, the packing of both columns was the commonly employed Sulzer Mellapak™ 250.Y [6].

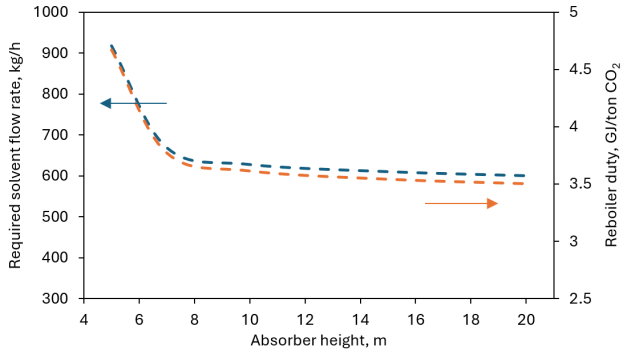

Figure 7. Effect of packing height in the absorber column on the specific reboiler duty.

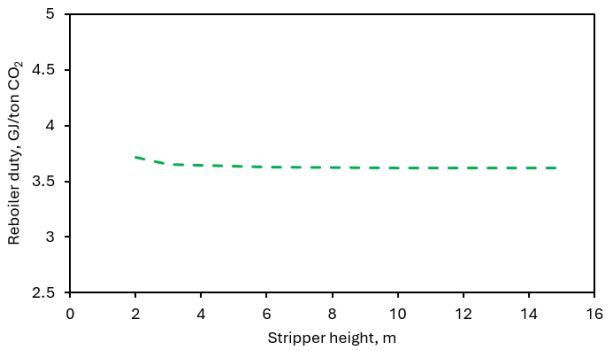

Figure 8. Effect of packing height in the stripper column on the specific reboiler duty.

At the top of the second absorber is the washing section which has the same diameter. The washing section fulfills two objectives: to control the water balance and to reduce MEA volatile emissions. As shown in Figure 9, both washing water temperature and flow rate can be used to keep the water balance. On the other hand, the packing height was determined using a sensitivity analysis on the recovery of volatile MEA, see Table 3. As indicated in the table, the washing water temperature and flow rate have a minor effect in comparison to the packing height.

For the stripper washing section, since the amount of volatile amine in the CO_2 product was fairly small after the condenser, the packing height was kept to 1 m.

Table 3: Sensitivity analysis on the recovery of volatile amine (%) by the water washing section on top of the absorber.

		Packing height, m			
Water temperature, °C	**Water flow rate, kg/h**	0.5	1	1.5	2
30	1000	75.9	87.1	92.0	95.4
40	1000	75.7	87.0	91.9	94.5
40	500	75.0	96.5	91.6	94.2

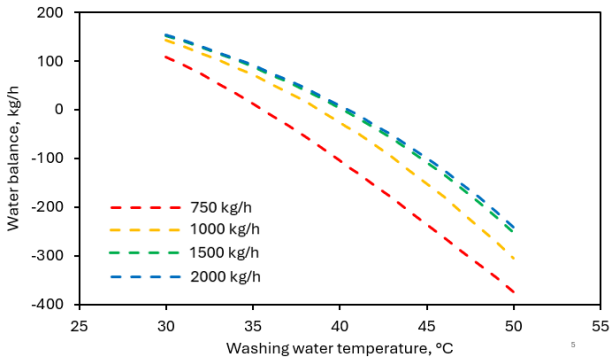

Figure 9. Water balance as a function of washing water flow rate and temperature in the washing section of the absorber.

CONCLUSIONS

A 1 ton/day CO_2 mobile capture pilot is being designed at ULiège, Belgium. The main process equipment, notably the columns, have been sized utilizing a model built in Aspen Plus. From a sensitivity study, it was determined that the required absorber and stripper packed bed heights. Beyond those points, further increments in height did not cause a significant reduction in the reboiler duty. The effect of increasing the height is more marked for the absorber. The washing section of the absorber was set to ensure substantial recovery of volatile amines. Cooling washing water flow rate and temperature had minor effects on the separation. Being less relevant, the washing section of the stripper was shorter. Ongoing work will provide more details about the basic and detailed engineering of this project. Moreover, the models will be validated after the first experimental campaigns.

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