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A new method for scale-up of solvent-based post-combustion carbon capture process with packed columns

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Abstract

Solvent-based post-combustion carbon capture (PCC) with packed column is the most commercially ready CO₂ capture technology. To study commercial-scale PCC processes, validated pilot scale models are often scaled up to commercial-scale using the generalized pressure drop correlation (GPDC) chart which requires assuming the column pressure drop. The GPDC method may lead to either over-estimation or under-estimation of the column diameter. In this paper, a new method for estimating the packed column diameter without assuming the pressure drop has been proposed and used for model scale-up. The method was validated by scaling between two existing pilot plant sizes. The CO₂ capture process was simulated in Aspen Plus[®] and validated at pilot scale. The validated model was scaled up to commercial CO₂ capture plant capable of serving a 250 MW_e combined cycle gas turbine power plant using the new method proposed in this study. The results obtained from the scale-up study were compared to those obtained when the GPDC method was used to design the same commercial CO₂ capture plant. The results showed that the GPDC method overestimated the absorber and stripper diameter by 1.6% and 8.5% respectively. Process simulation results for the commercial-scale plant showed about 2.12% and 5.63% lower solvent flow rate and reboiler duty with the proposed method. Therefore, the capital and operating costs for the process using the newly proposed scale-up method could be lower based on our estimates of the column dimensions, solvent flow rate and specific reboiler duty.

Keywords: *post-combustion CO₂ capture, chemical absorption, process modelling and simulation, model validation, scale-up, combined cycle gas turbine power plant*

Highlights

- Generalized pressure drop correlation (GPDC) commonly used for scale-up in carbon capture.
- New method to estimate packed column diameter proposed.

- 32 • Rate-based model developed and validated at pilot scale for MEA-based PCC in Aspen
- 33 Plus® V8.4.
- 34 • New scale-up method validated using two existing pilot plants.
- 35 • Scale-up of MEA-based PCC process based on the proposed method carried out.

36 Nomenclature

- 37 a specific surface area of packing (m^2/m^3)
- 38 C_i concentration of component i . (kmol/m^3)
- 39 CP capacity parameter
- 40 D diameter (m)
- 41 E_j activation energy (kJ/mol)
- 42 F_{LV} flow parameter
- 43 F_p packing factor (m^{-1})
- 44 G gas mass flow rate (kg/s)
- 45 G_i gas molar flow rate per cross-sectional area ($\text{kmol}/\text{m}^2 \text{ s}$)
- 46 H_{OG} height of the transfer unit (m)
- 47 K_G overall gas-phase mass transfer coefficient ($\text{kmol}/\text{m}^3 \text{ s bar}$)
- 48 k_j^o pre-exponential factor ($\text{m}^3/\text{kmol.s}$)
- 49 L solvent mass flow rate (kg/s)
- 50 M_{MEA} molar mass of MEA (kg/kmol)
- 51 n Temperature factor
- 52 N_{OG} overall number of the transfer unit
- 53 P pressure (bar)
- 54 ΔP_{fl} flooding pressure drop (in.H₂O/ft)
- 55 R ideal gas constant (J/K mol)
- 56 R_j Reaction rate for reaction j , ($\text{m}^3/\text{kmol.s}$)
- 57 T Temperature (K)
- 58 $V_{G,fl}$ flooding velocity (m/s)
- 59 V_G superficial gas velocity (m/s)
- 60 y mole fraction of CO₂ in the gas phase at any point in the column
- 61 $y_{CO_2,in}$ mole fraction of CO₂ in the inlet gas

62	$y_{CO_2,out}$	mole fraction of CO ₂ in the outlet gas
63	y^*	gas-phase mole fraction of CO ₂ in equilibrium with CO ₂ concentration in the liquid
64	z	number of equivalents/moles of amine (1 for MEA)
65	Z_T	packing height (m)
66		
67		<i>Greek letters</i>
68	α_{Lean}	lean loading (mol CO ₂ /mol MEA)
69	α_{Rich}	rich loading (mol CO ₂ /mol MEA)
70	$\Delta\alpha$	absorption capacity (mol CO ₂ /mol MEA)
71	α_{ij}	specie i reaction order in reaction j
72	ρ_G	gas density (kg/m ³)
73	ρ_L	liquid density (kg/m ³)
74	ε	porosity
75	ν	kinematic viscosity (cst)
76	φ_{CO_2}	percentage of CO ₂ captured
77	ω_{MEA}	MEA concentration (wt%)

78

79 *Abbreviations*

80	CCGT	Combined Cycle Gas Turbine
81	GPDC	Generalized pressure drop correlation
82	HETP	Height Equivalent to the Theoretical Plate
83	PCC	Post-Combustion Carbon Capture
84	PRE	Percentage relative error
85	SRP	Separation Research Programme

86

87 **1. Introduction**

88 **1.1. Background**

89 There is an increasing concern about global warming effect arising from the emission of
90 greenhouse gases (GHGs). Anthropogenic CO₂ emissions from different sources constitute
91 about 80% of the total GHG emissions (Sreedhar et al., 2017), and CO₂ emissions from fossil

92 fuel-fired power plants are responsible for approximately about 25% of the total GHG (Soltani
93 et al., 2017; EPA, 2017). This indicates that efforts at reducing GHG emissions must be targeted
94 at cutting down CO₂ emissions from these facilities. One way to achieve this is through the
95 deployment of cost-effective CO₂ capture technologies in fossil fuel-fired power plants.

96 There are three technological options for CO₂ capture: pre-combustion, oxy-fuel combustion
97 and post-combustion. Among these capture technologies, post-combustion CO₂ capture
98 through chemical absorption with amines is the most mature technology to be used to cut down
99 CO₂ emissions from power plants (Wang et al., 2011). In addition to this, the technology is
100 considered the best option for retrofit as its implementation in an existing power plant requires
101 very little modifications (Rezazadeh et al., 2017). Despite these advantages, the commercial
102 implementation of the solvent-based PCC process is faced with a number of challenges such
103 as high capital cost and high energy consumption.

104 **1.2 Previous studies**

105 Process modelling and simulation is critical to the design and operation of the PCC plant, and
106 several studies with focus on model development for the plant have been carried out (Awoyomi
107 et al., 2019; Bui et al., 2018; Enaasen et al., 2015; Garcia et al., 2017; Khan et al., 2011; Lawal
108 et al., 2009; Soltani et al., 2017; Zhang et al., 2009). Earlier studies focussed on model
109 development for the standalone absorber (Khan et al., 2011; Kvamsdal et al., 2009; Lawal et
110 al., 2009; Zhang et al., 2009) and the standalone stripper (Greer et al., 2010; Ziaii et al., 2009).
111 This was followed by model development for the whole solvent-based PCC plant (Gaspar and
112 Cormos, 2012; Harun et al., 2012; Lawal et al., 2010; Warudkar et al., 2013; Zhang and Chen,
113 2013). The reliability of the models' predictions was validated using published experimental
114 data collected from various pilot plants around the world. Experimental data to which model
115 predictions are commonly compared in the literature are the CO₂ capture level (Errico et al.,
116 2016; Harun et al., 2012; Lawal et al., 2009; Razi et al., 2013; Zhang et al., 2009), rich solvent
117 CO₂ loading (Enaasen Flø et al., 2015; Khan et al., 2011; Luo and Wang, 2017), temperature
118 profile (Bui et al., 2014; Canepa et al., 2013; Garcia et al., 2017; Khan et al., 2011; Lawal et
119 al., 2009; Razi et al., 2013), CO₂ concentration profiles (Khan et al., 2011; Luo and Wang,
120 2017; Razi et al., 2013), desorbed CO₂ (Garcia et al., 2017) and specific heat duty (Agbonghae
121 et al., 2014; Luo and Wang, 2017).

122 Zhang et al. (2009) validated the rate-based absorber model developed in Aspen Plus[®] with a
123 pilot plant data by the Separations Research Programmes (SRP) at the University of Texas.

124 The model was validated against the following parameters: CO₂ capture level, CO₂ loadings,
125 and temperature profiles. The model predictions showed excellent agreement with the pilot
126 plant data for each of the parameters. Khan et al. (2011) validated their rate-based model with
127 the pilot and industrial-scale experimental data collected from the studies of Pintola et al.
128 (1993), Tontiwachwuthikul et al. (1992) and Aroonwilas et al. (2001). The model predictions
129 matched experimental measurements for the liquid phase MEA and gas-phase CO₂
130 concentrations and the liquid phase temperature profiles.

131 In order to design and study the possible requirements of a commercial-scale MEA-based CO₂
132 capture process, the validated pilot-scale models are often scaled to commercial scale. Several
133 researchers (Agbonghae et al., 2014; Awoyomi et al., 2019; Biliyok and Yeung, 2013; Canepa
134 et al., 2013; Dutta et al., 2017; Lawal et al., 2012; Nittaya et al., 2014) have performed model
135 scale-up of the process from pilot scale to commercial scale. Lawal et al. (2012) designed a
136 commercial CO₂ capture plant that is capable of capturing 90% of CO₂ from the flue gas stream
137 of a 500 MW_e subcritical coal-fired power plant by scaling-up the validated CO₂ capture pilot
138 plant model developed in gPROMS. Using the generalized pressure drop correlation chart
139 (GPDC), they developed a capture plant with two absorbers each of diameter 9 m and height
140 27 m and a stripper having the same diameter as the absorber. Similarly, Nittaya et al. (2014)
141 scaled up the CO₂ capture pilot plant model developed in gPROMS to a commercial CO₂
142 capture plant capable of capturing 87% of CO₂ from the flue gas of a 700 MW_e supercritical
143 coal-fired power plant. Their scale-up resulted in a commercial CO₂ capture plant with three
144 absorbers, each with a diameter of 11.8 m and height of 34 m and two strippers each having a
145 diameter of 10.4 m and height of 16 m. Agbonghae et al. (2014) scaled up a validated CO₂
146 capture pilot plant model developed in Aspen Plus[®] to a commercial CO₂ capture plant capable
147 of capturing 90% of CO₂ from the flue gas of a 400 MW_e CCGT power plant. They came up
148 with a CO₂ capture plant with two absorbers, each with a diameter of 11.93 m and height of
149 19.06 m and a stripper with a diameter of 6.76 m and height of 28.15 m. In all the studies above,
150 the commercial-scale designs of the absorber and the stripper are based on the GDPC method—
151 which involves assuming the column pressure drop. This study is focussed on developing an
152 alternative method to estimate the diameter of the packed column for solvent-based PCC
153 process using an empirical correlation that estimates the flooding gas velocity. This allows the
154 diameter of the packed column to be calculated without assuming the pressure drop.

155 **1.3 Aims and novelty of this study**

156 Model scale-up from pilot scale to commercial scale for the solvent-based PCC will not only
157 help in providing insights into plant operations but also foresee any commercial-scale
158 development and operational bottlenecks. For the solvent-based PCC process, the packed bed
159 absorber and the stripper are the two largest components in terms of size (Agbonghae et al.,
160 2014; Lawal et al., 2012) and cost (Abu-Zahra et al., 2007). Their design as reported in the
161 literature is based on chemical engineering principles using the GPDC method. Sinnott (2005)
162 recommended a pressure drop range of 147 to 490 Pa/m of packing for packed column design
163 at commercial scale. Within this pressure drop range, experimental data are only available at
164 206 and 412 Pa/m of packing on the GPDC chart thereby limiting the choice of pressure drop
165 that can be assumed within this range. Furthermore, data interpolation for pressure drop are
166 difficult and could lead to inaccurate estimates. In existing studies (Agbonghae et al., 2014;
167 Awoyomi et al., 2019; Canepa et al., 2013; Dutta et al., 2017; Lawal et al., 2012; Luo and
168 Wang, 2017; Nittaya et al., 2014), pressure drop of either 206 or 412 Pa/m of packing has been
169 assumed. To address this limitation, this study aims to propose an alternative method to
170 estimate the packed column diameter that does not require assuming the column pressure drop.
171 The method involves an algebraic equation derived for the flooding velocity from flooding
172 point experimental correlations reported in the literature. As far as open literature is concern,
173 this attempt is first of its kind. In addition, this approach has been validated in this study by
174 scaling between two existing pilot plants sizes, a similar demonstration could not be found in
175 literature for reported scale-up studies of the process. The method developed in this study is
176 used to scale up the pilot plant model developed in Aspen Plus[®] to a commercial CO₂ capture
177 plant. And the results compared to scale-up study results obtained with the GPDC method.

178 **2. Methodology**

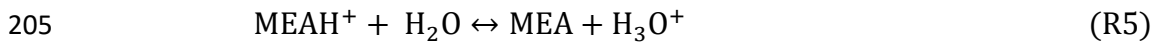
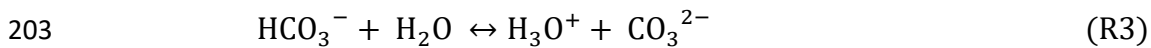
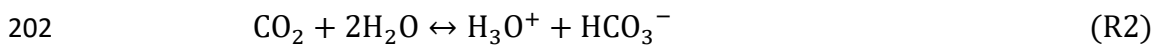
179 **2.1 Model development**

180 The closed-loop model of the CO₂ absorption and stripping process was developed in Aspen
181 Plus[®] V8.4. The absorber and stripper model were developed using the RadFrac rate-based
182 model. The rate-based calculations give more reliable results in comparison to the equilibrium-
183 based model counterpart (Lawal et al., 2009). This is because, in the rate-based model,
184 equilibrium is assumed to be achieved only at the vapour-liquid interface and separation is
185 caused by the mass transfer of component between the contacting phases. On the other hand,
186 the equilibrium-based model assumed that each theoretical stage is made up of a well-mixed
187 vapour and liquid phases in equilibrium with each other. This assumption is an approximation
188 because, in real column, the contacting phases are never in equilibrium (Zhang et al., 2009).

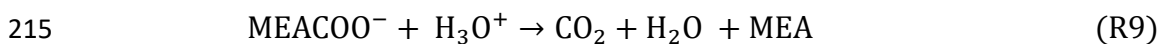
189 The dimensions of the RadFrac columns were specified to be the same as those of the pilot
190 plants as shown in Tables 4 and 6.

191 **2.1.1 Thermodynamic and kinetic models**

192 The liquid phase of the MEA-H₂O-CO₂ system is an electrolyte solution whose accurate
193 modelling requires the selection of a base method that can account for the electrolytes in Aspen
194 Plus[®]. The Electrolyte Non-Random-Two-Liquid (eNRTL) activity coefficient model (Chen
195 and Evans, 1986) was used to calculate the activity coefficient and the SRK equation of state
196 (Soave, 1972) was used to calculate the fugacity coefficient. Other important thermodynamic
197 properties such as Henry's constant, vapour pressure, the heat of absorption and specific heat
198 capacity are calculated using correlations within the eNRTL thermodynamic method in Aspen
199 properties[®]. The equations describing the equilibrium reactions are defined as follows (Aspen
200 Technology, 2008):



206 The equilibrium constants for reactions R1 to R5 are calculated from the Gibbs free energy
207 change, and the equilibrium reactions are assumed to occur in the liquid film. In the rate-based
208 model, the reactions R6 and R7 representing the forward and backward reactions for the
209 formation of bicarbonate and the reactions R8 and R9 representing the forward and backward
210 reactions for the formation of carbamate are considered as kinetics-controlled reactions (Zhang
211 and Chen, 2013).



216 The reaction rates for reactions R6 to R9 can be calculated by the power law which is described
 217 in Aspen Plus® by the following equation.

$$218 \quad R_j = k_j^o T^n \exp \left[-\frac{E_j}{R} \left(\frac{1}{T} - \frac{1}{298.15} \right) \right] \prod_{i=1}^N C_i^{\alpha_{ij}} \quad (1)$$

219 The values of k_j^o and E_j in equation 1 used for reactions R6 to R9 are shown in Table 1.

220 **Table 1**

221 Parameter of the pre-exponential factor and activation energy (Aspen Technology, 2008).

Reactions	Reaction direction	k_j^o (kmol/m ³ s)	E_j (kJ/mol)
R6	Forward	4.32 x 10 ¹³	55.43
R7	Reverse	2.38 x 10 ¹⁷	123.22
R8	Forward	9.77 x 10 ¹⁰	41.24
R9	Reverse	2.18 x 10 ¹⁹	59.19

222

223 2.1.2 Transport property models

224 Transport property models, namely density, viscosity, thermal conductivity, surface tension,
 225 and diffusivity have been calculated using the correlations summarised in Table 2.

226 **Table 2**

227 Summary of models for calculating transport properties (Aspen Technology, 2001).

Property	Gas-phase	Liquid phase
Density	COSTALD model by Hankinson and Thomson	Clark density model
Viscosity	Chapman-Enskog-Brokaw model	Jones-Dole model
Thermal conductivity	Wassiljewa-Mason-Sexena model	Riedel model
Surface tension		Onsager-samaras model
Diffusivity	Chapman-Enskog-Wilke-Lee model	Wilke-Chang model

228

229 2.1.3 Heat and mass transfer calculations

230 Heat and mass transfer calculations have been performed using correlations for the mass
 231 transfer coefficient, heat transfer coefficient, interfacial area, and the liquid holdup. A summary
 232 of the correlations is given in Table 3.

233 **Table 3**
 234 Summary of correlations used for mass and heat transfers.

Correlations	References	
	Absorber	Stripper
Liquid and gas film mass transfer coefficient	Onda et al. (1968)	Bravo et al. (1985)
Heat transfer coefficient	Chilton and Colburn (1934)	Chilton and Colburn (1934)
Liquid holdup	Stichlmair et al. (1989)	Bravo et al. (1992)
Effective Interfacial area	Onda et al. (1968)	Bravo et al. (1985)

235

236 **3. Model validation**

237 In this study, pilot plant data from the Separation Research Programme (SRP) at the University
 238 of Texas at Austin, USA (Dugas, 2006) and the Brindisi CO₂ capture plant located in Brindisi,
 239 Italy (Enaasen, 2015) were used to validate the performance of the rate-based model presented
 240 in the previous section.

241 **3.1 Model validation using the SRP pilot plant data**

242 Experimental data collected at the SRP pilot plant which is a multifunctional test facility were
 243 used to validate the rate-based model. The SRP pilot plant uses synthetic flue gas produced by
 244 mixing air and CO₂ gas. The absorber and the stripper both have internal diameter of 0.427m
 245 and a total height of 11.1 m. The columns are each made up of two 3.05 m bed of packing with
 246 plate collector and liquid redistributor between them. It is capable of handling flue gas flow
 247 rate ranging from 330-830 m³/h and can capture between 125 and 250 kg of CO₂/h. The main
 248 process conditions, dimensions of the absorbers and the strippers, and the type of packings used
 249 in the pilot plant for the three selected cases are summarized in Tables 4.

250 **Table 4**
 251 Pilot plant data from the SRP CO₂ capture plant (Dugas, 2006)

Cases	28	32	47
Flue gas flow rate (m ³ /min)	11.00	5.48	8.22
Flue gas CO ₂ concentration (mol%)	16.54	17.66	18.41
Flue gas temperature (°C)	47.98	46.56	59.23
Flue gas pressure (bar)	1.05	1.05	1.03
Lean solvent flow rate (m ³ /min)	0.08	0.04	0.03
Lean solvent temperature (°C)	40.00	40.56	40.07

Absorber pressure (bar)	1.00	1.00	1.00
Regenerator pressure (bar)	1.62	1.62	0.68
	Absorber	Stripper	
Diameter (m)	0.43	0.43	
Packing height (m)	6.10	6.10	
Packing type	IMTP 40	Flexipac 1Y	

252

253 The three experimental cases selected for model validation from the 48 experimental runs
 254 conducted at the SRP facility have different CO₂ concentrations and L/G ratios. Model
 255 validation was performed by comparing the model predictions for the CO₂ capture level and
 256 CO₂ loadings against experimental data for different feed conditions. The percentage CO₂
 257 capture level and the CO₂ loading in the MEA solvent are calculated using Eqs. 2 and 3.

$$\text{Capture level (\%)} = \left(\frac{y_{CO_2,in} - y_{CO_2,out}}{y_{CO_2,in}} \right) * 100 \quad (2)$$

258

$$\text{Loading} = \frac{[CO_2] + [HCO_3^-] + [CO_3^{2-}] + [MEACOO^-]}{[MEA] + [MEA^+] + [MEACOO^-]} \quad (3)$$

259 Table 5 shows the model performance results against the experiment. There is a good
 260 agreement between the model predictions and the experiment data for all the variables outlined
 261 in Table 5. The percentage relative errors (PRE) of the model prediction against the
 262 experimental data are calculated as follows:

$$\text{PRE} = \frac{|i_{\text{experiment}} - i_{\text{model}}|}{i_{\text{experiment}}} * 100 \quad (4)$$

263 Fig.1 presents the comparison between the measured and predicted liquid phase temperature
 264 profiles along the height of the absorber and the stripper. The model generally gives a good
 265 prediction of the temperature profiles in the absorber and the stripper for the three selected
 266 cases. Also, the model accurately predicted the location of the temperature bulge (maximum
 267 temperature) in the absorber for the three cases as illustrated by curves a, c and e). The location
 268 and magnitude of the temperature bulge depend on L/G ratio (Plaza and Rochelle, 2011).
 269 Dugas (2006) found that the temperature bulge was located at the top of the absorber with L/G

270 less than 5 kg/kg and at the bottom with L/G greater than 6 kg/kg. This explains the location
 271 of the temperature bulge close to the bottom of the absorber packing for curves *a* and *c*
 272 (L/G=6.6 kg/kg) and close to the top of the absorber packing for curve *e* (L/G=3.4 kg/kg).

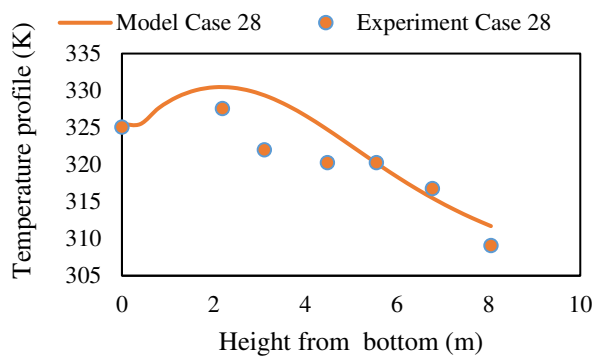
273 **Table 5**

274 Model performance against experimental data for the SRP CO₂ capture pilot plant

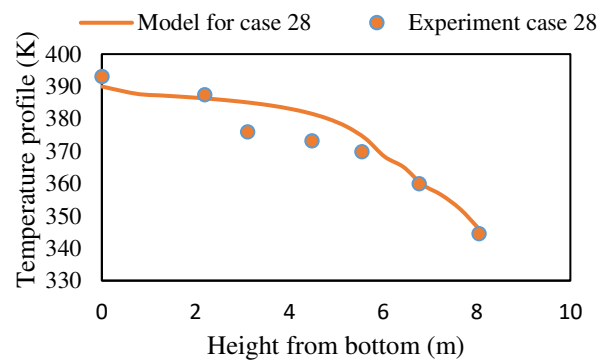
Cases	Lean loading (mol CO ₂ /mol MEA)			Rich loading (mol CO ₂ /mol MEA)			CO ₂ capture level (%)		
	Exp.	Model	PRE (%)	Exp.	Model	PRE (%)	Exp.	Model	PRE (%)
28	0.28	0.28	0.00	0.41	0.41	0.00	86	85	1.16
32	0.27	0.27	0.00	0.43	0.43	0.00	95	90	5.26
47	0.28	0.30	-6.60	0.53	0.48	9.43	69	69	0.00

275

(a)

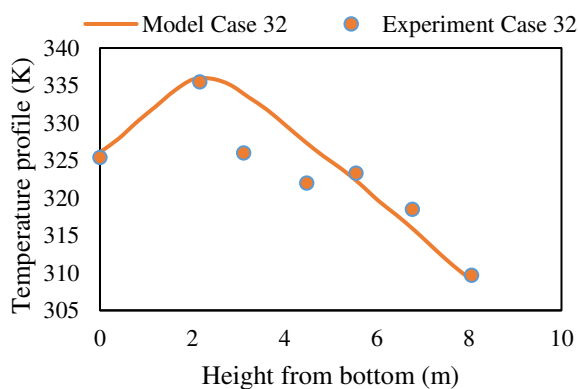


(b)

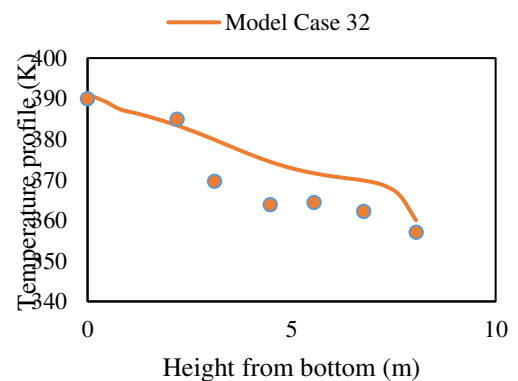


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(c)

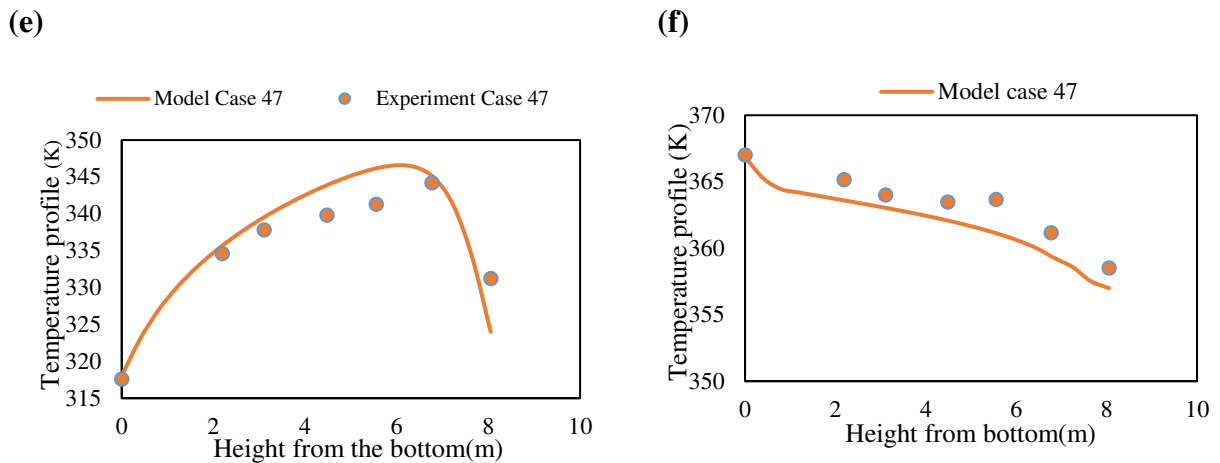


(d)



277

278



279 **Fig. 1.** Model predictions against experimental data for temperature profiles in the absorber (a,
 280 c, e) and in the stripper (b, d, and f) of the SRP pilot plant for the three cases.

281

282 3.2 Model validation using the Brindisi pilot plant data

283 The rate-based capture model was also validated using experimental data collected at the
 284 Brindisi pilot plant (Enaasen, 2015). It is a relatively large plant compared to the SRP pilot
 285 plant described in section 3.1. The pilot plant uses a flue gas produced from one of the four
 286 units (each with capacity of 660 MW_e) of a coal-fired power plant. The absorber and stripper
 287 have diameters of 1.5m and 1.3m, and packing heights of 22m and 11m respectively. It can
 288 capture up to 2500 kg of CO₂/h from a flue gas slipstream and has a maximum capacity of 9212
 289 m³/h which corresponds to about 0.45 % of the total flue gas produced from the unit four of
 290 the power plant (Lemaire et al., 2014). The solvent flow rate can be varied between 20-80 m³/h.
 291 The main process conditions, dimensions of the absorbers and the strippers, and the type of
 292 packings used in the pilot plants for the selected cases are summarized in 6.

293 **Table 6**

294 Pilot plant data from the Brindisi CO₂ capture plant (Enaasen, 2015).

Cases	2	3	4	5	7
Flue gas flow rate (Nm ³ /h)	9876	9929	9893	9949	9921
Flue gas CO ₂ concentration (dry vol%)	11.00	12.50	12.00	10.40	11.00
Flue gas temperature (°C)	46.20	44.80	45.60	44.70	46.90
Lean solvent flow rate (m ³ /h)	30	30	30	35	35
Lean solvent temperature (°C)	46.90	47.00	47.00	47.00	47.10
Lean solvent MEA concentration (wt%)	29.60	29.80	29.80	29.80	29.70
	Absorber		Stripper		

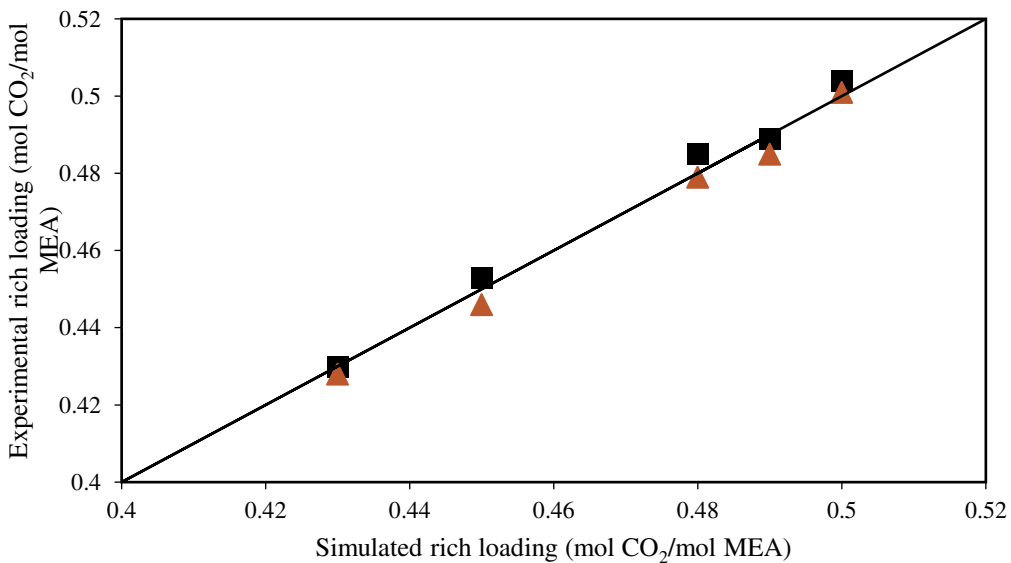
Diameter (m)	1.50	1.30
Packing height (m)	22	11
Pressure (bar)	1.00	1.84

295

296 Five experimental cases (Table 6) with the least relative deviations in steady-state CO₂ mass
 297 balance were selected for the model validation among the 12 experimental cases reported. Two
 298 sets of packings were used in the absorber and stripper during the model validation. The first
 299 set of packing (Mellapak 250X and IMTP 50) is the original packings used in the Brindisi pilot
 300 plant during the experiments while the second set of packing (IMTP 40 and Flexipac 1Y) is the
 301 packing used in the SRP pilot plant. This was done to:

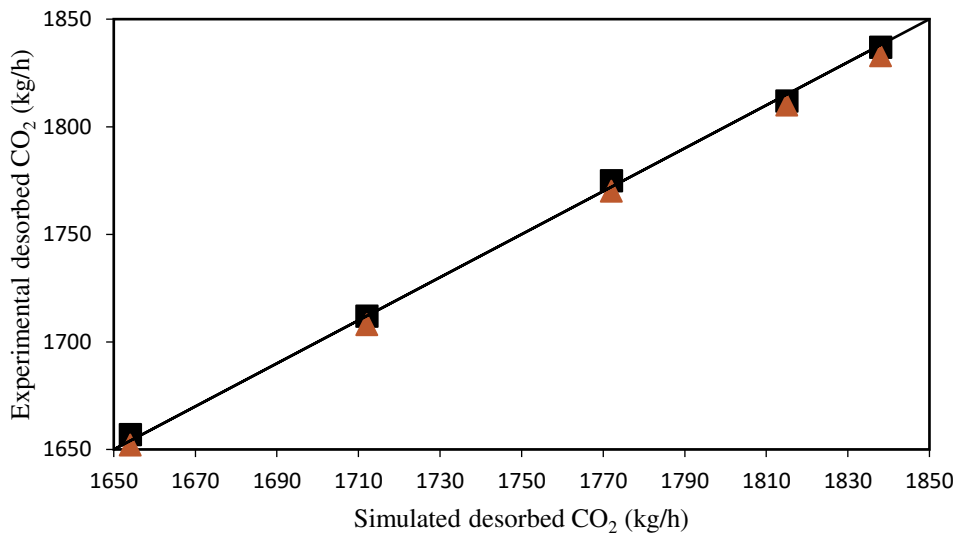
- 302 1. Enable the scale-up from the SRP pilot plant to the Brindisi pilot plant in order to
 303 validate the proposed scale-up approach presented later in section 4.
- 304 2. Enable the scale-up of the Brindisi pilot plant model (using the second set of packing)
 305 to a commercial CO₂ capture plant (Section 5.2) using the proposed scale-up method.
 306 The results obtained from the scale-up will be compared to those obtained from a
 307 commercial CO₂ capture plant designed by Canepa et al. (2013) using the same set of
 308 packings and the GPDC method.

309 The parity plot of the rich solvent CO₂ loading, desorbed CO₂ and specific duty predicted by
 310 the model using the two sets of packings against experimental data are shown in Figs. 2-4. The
 311 validation results show good agreement between the model predictions and experimental data.
 312 The results further demonstrate that the sets of packing used in the columns have identical
 313 performance in terms of rich CO₂ loading, amount of CO₂ desorbed and specific duty.



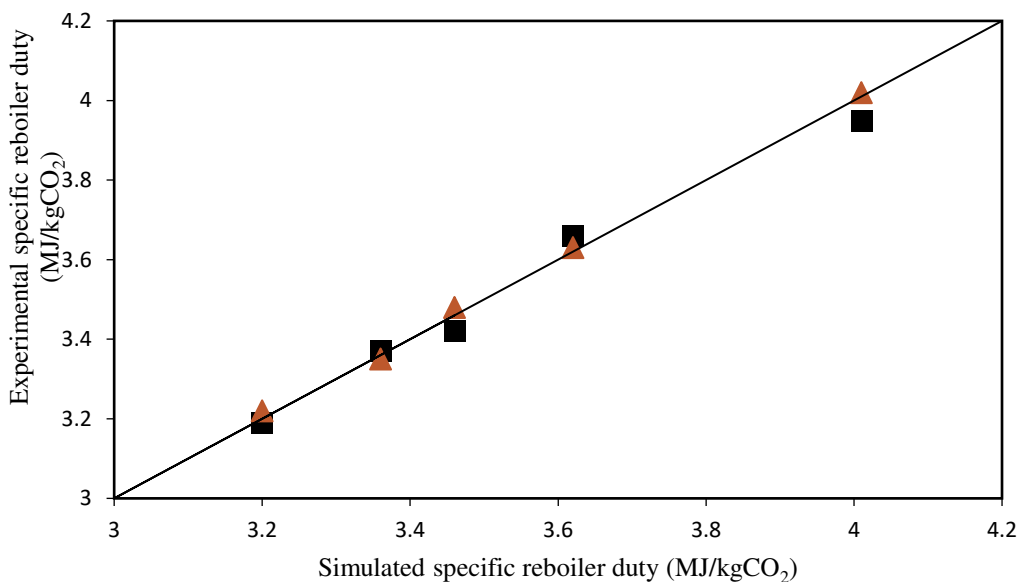
314

315 **Fig. 2.** Experimental values of rich loading (Enaasen, 2015) compared to simulated values
 316 obtained with set 1 packing (■) and set 2 packing (▲) and the dark line represents equal
 317 experimental and simulated rich loadings.



318
 319 **Fig. 3.** Experimental values of desorbed CO₂ (Enaasen, 2015) compared to simulated values
 320 obtained with set 1 packing (■) and set 2 packing (▲). The dark line represents equal
 321 experimental and simulated desorbed CO₂.

322

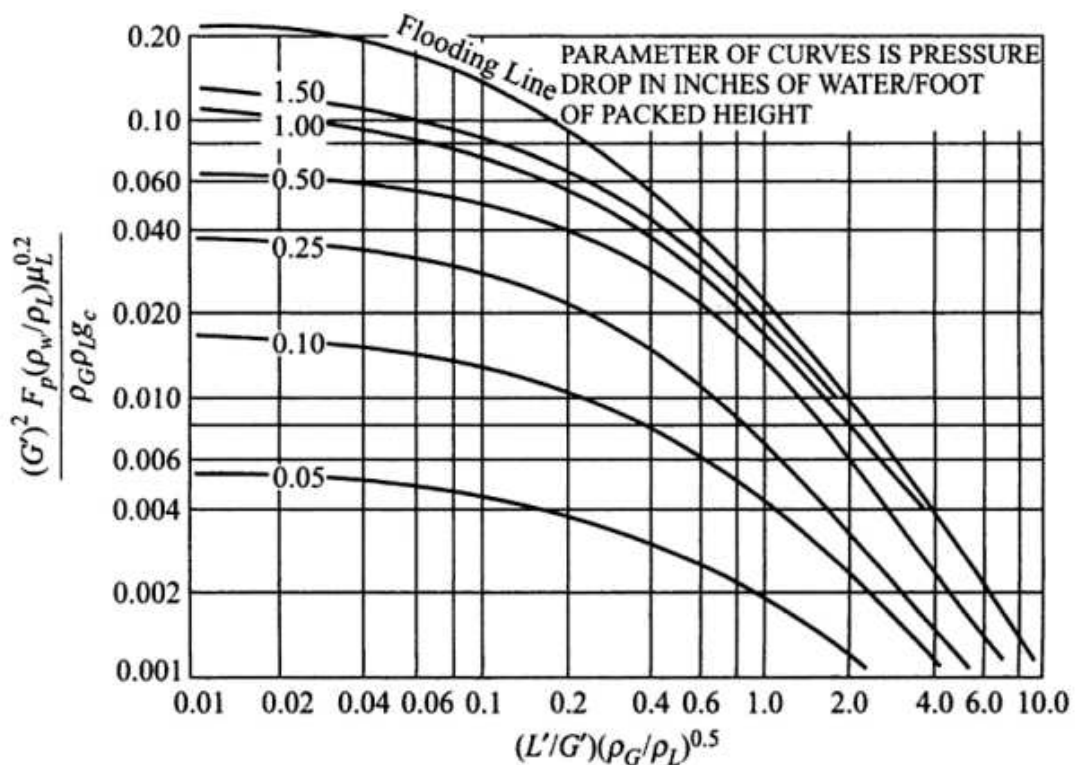


323
 324 **Fig. 4.** Experimental values of specific duty (Enaasen, 2015) compared to simulated values
 325 obtained with set 1 packing (■) and set 2 packing (▲), the dark line represents equal
 326 experimental and simulated specific duty.

327

328 **4. A newly proposed method for estimating the diameter of absorber and stripper**

329 The diameter of a packed column is a key parameter that must be determined in the design of
 330 a packed bed absorber or stripper. The columns are designed in such a way to avoid flooding
 331 because flooding reduces their efficiency and sometimes causes the column to breakdown (Liu
 332 et al., 2019). Since the flooding point establishes the upper limit of the hydrodynamic capacity
 333 at which the packed column can operate, the velocity of the gas at flooding condition is
 334 particularly important and is a vital design parameter for the packed column (Brunazzi et al.,
 335 2008). Sherwood et al. (1938) developed the first generalized correlation chart for predicting
 336 flooding points in random dumped packings using experimental data from an air-water system.
 337 The chart which contained only one curve was later modified by Lobo et al. (1945). The
 338 ordinate of the chart includes the ratio a/ε^3 for characterising the packing size and shape. Leva
 339 (1954) added several isobaric curves to determine the pressure drop in the packed beds. In
 340 addition, Leva (1954) determined that the ratio a/ε^3 did not adequately predict the packing
 341 hydraulic performance and proposed the use of packing factor to characterise packing size and
 342 shape. Eckert (1970) further modified the chart and calculated the packing factor from
 343 experimentally determined pressure drops. The modified Eckert version (Fig. 5), known as the
 344 Sherwood-Leva-Eckert (SLE) GPDC chart has been the standard for pressure drop and
 345 flooding points prediction in a column packed with random packings for many decades.



346

347 **Fig. 5.** Generalized pressure drop correlation for packings (Eckert, 1970)

348 In later versions of the GPDC chart developed by Strigle (1994) for random packings and Kister
349 et al. (2007) for structured packings, only the pressure drop curves were retained while the
350 flooding curve was omitted (Wolf-Zöllner et al., 2019a). Kister and Gill (1991) developed a
351 correlation for predicting the flooding point and pressure drop in packed columns and an
352 expression for the flooding curve was written in equation form particularly for the relationship
353 between the abscissa and the ordinate as follows (Piché et al., 2001).

354

$$CP = A \log^2(F_{LV}) + B \log(F_{LV}) + C \quad (5)$$

355 Where F_{LV} is the flow parameter. It is the ratio of the kinetic energy of the liquid to the kinetic
356 energy of the gas entering the packed column (Kister et al., 2007), and it is represented as
357 follows:

$$F_{LV} = \frac{L}{G} \sqrt{\frac{\rho_L}{\rho_G}} \quad (6)$$

358 The value of the flow parameter is low for vacuum operation but high for operations involving
359 high pressures or high liquid/vapour loading such as gas absorption operation. The CP in Eq.
360 5 is the capacity parameter and it is given by:

361

$$CP = \sqrt{V_{G,fl}^2 \left(\frac{\rho_G}{\rho_L - \rho_G} \right)^{0.1} F_P} \quad (7)$$

362

363 The pressure drop at which incipient flooding occurs in columns packed with modern random
364 packings has been correlated and expressed as a function of the packing factor F_P by Kister and
365 Gill (1991) as shown by Eq. 8.

$$\Delta P_{fl} = 0.115 F_P^{0.7} \quad (8)$$

366 Eq.8 also applies to structured packings and has been found to predict very well the pressure
367 drop at flooding point for structured packings (Geankoplis, 2014; Kister and Gill, 1992). The
368 equation is particularly applicable to packings with F_P between 10 and 60 ft^{-1} , thus, it is capable
369 of predicting the pressure drop at flooding in packed columns from as low as 0.57 in. $\text{H}_2\text{O}/\text{ft}$

370 for packings with F_p of 10 ft^{-1} to as high as approximately 2 in. $\text{H}_2\text{O}/\text{ft}$ for packings with F_p of
 371 60 ft^{-1} . However, the equation only gives an optimistic prediction of the flooding point pressure
 372 drop at F_p beyond 60 ft^{-1} and should therefore not be used with F_p above this value (Geankoplis,
 373 2014).

374 The expressions for determining the parameters A, B and C in Eq.5 together with their range
 375 of application are summarised in Table 7. These parameters are determined using the flooding
 376 point pressure drop calculated from Eq. 8.

377 **Table 7**

378 Expressions for parameters in Eq.5 (Piché et al., 2001)

Parameters	Expression	Range of application
A	$0.07 \ln(\Delta P_{fl}) - 0.11$	$0.5 \leq \Delta P_{fl} \leq 5.0 \text{ inH}_2\text{O}/\text{ft}$
B	$-0.25 \ln(\Delta P_{fl}) - 0.89$	$0.5 \leq \Delta P_{fl} \leq 1.0 \text{ inH}_2\text{O}/\text{ft}$
B	-0.89	$1.0 \leq \Delta P_{fl} \leq 5.0 \text{ inH}_2\text{O}/\text{ft}$
C	$0.12 \ln(\Delta P_{fl}) + 0.71$	$0.5 \leq \Delta P_{fl} \leq 5.0 \text{ inH}_2\text{O}/\text{ft}$

379

380 Eq. 5 can be re-written in the form shown below;

$$CP = A(\log F_{LV})^2 + B \log(F_{LV}) + C \quad (9)$$

381 By equating Eqs. 7 and 9 and substituting for F_{LV} in the resulting equation. An expression of
 382 the form in Eq 10 can be written for the flooding velocity ($V_{G,fl}$).

$$383 \quad V_{G,fl} = 0.3048 \left[\left(\frac{\rho_G}{\rho_L - \rho_G} \right)^{-0.5} v^{-0.05} F_p^{-0.5} \left\{ A \left(\log \left(\frac{L}{G} \sqrt{\frac{\rho_G}{\rho_L}} \right) \right)^2 + B \left(\log \left(\frac{L}{G} \sqrt{\frac{\rho_G}{\rho_L}} \right) \right) + C \right\} \right]$$

384

(10)

385 The flooding velocity (upper limit of the rate of gas flow) in a packed column can be calculated
 386 from Eq.10 once process information such as the flow rate, density and kinematic viscosity of
 387 the individual phase are known. Also important is the packing factor. The values of the density
 388 and kinematic viscosity can be obtained from open literature, experiments or chemical process
 389 simulation software such as Aspen Plus[®] and ProMax. Eq. 10 has limitation over the range of
 390 its application. This is because the correlation presented in Eq. 8 which formed the basis of the
 391 equation only gives a good prediction of the flooding point pressure drop between F_p range of

392 10 to 60 ft⁻¹. Therefore, it is recommended that it should only be used to estimate the flooding
393 velocity in a column packed with packings with F_P in the range specified above.

394 The diameter required by a given gas and liquid flow rate in a packed column is based on the
395 maximum allowable pressure drop and the maximum operational capacity (MOC). The values
396 for the MOC can range from 60 to 86 percent, thus, packed columns are usually designed to
397 operate at about 60-80 percent of the flooding velocity (Marx-Schubach and Schmitz, 2019). In
398 this work, it is assumed that the column operates at 70 percent of the flooding velocity, hence
399 the superficial gas velocity at operating condition was calculated as follows:

$$V_G = 0.7V_{G.fl} \quad (11)$$

400

401 The diameter of the column required to perform the absorption operation at 70% of flooding
402 velocity can be calculated from the expression in Eq. 12.

$$D = \sqrt{\frac{4G}{\pi V_G \rho_G}} \quad (12)$$

403

404 **5. Model Scale-up**

405 The design of a commercial PCC plant by the scale-up of PCC pilot plant model requires scale-
406 up calculations to be performed to determine the size of the absorber and the stripper. One
407 important parameter that determines the size of these columns is the amount of flue gas to be
408 treated by the commercial PCC plant. In most cases, the flue gas is thousands of times the
409 amount at pilot scale. For instance, the commercial PCC plant designed by Lawal et al. (2012)
410 and Canepa et al. (2013) process about 5000 and 2200 times the amount of flue gas at the
411 pilot scale. Considering that the CO₂ capture process generally involves many interacting
412 variables, accurate scale-up of the process to a commercial PCC plant that is capable of
413 processing flue gas that is thousands of times the amount at pilot scale is, therefore, a very
414 complicated exercise. To avoid this complication, a two-stage scale-up of the validated models
415 presented in section 3 is carried out as follows:

- 416 (1) The validated model of the SRP CO₂ capture pilot plant (flue gas flow rate 0.15 kg/s)
417 is scaled up to the size of the Brindisi CO₂ capture pilot plant (flue gas flow rate 3.22
418 kg/s) to validate the proposed scale-up method.

419 (2) The validated model of the Brindisi pilot plant is scaled up to a commercial CO₂ capture
420 plant capable of serving a 250 MW_e CCGT power plant producing 356 kg/s of flue gas.
421
422

423 **5.1 Scale-up of the SRP CO₂ capture pilot plant to the Brindisi CO₂ capture pilot plant**

424 In order to validate the approach proposed in section 4, the SRP pilot plant is scaled up to the
425 size of the Brindisi pilot plant. Based on the amount of flue gas processed, the Brindisi pilot
426 plant has the capacity that is about 22 times of the SRP pilot plant. It is a relatively large PCC
427 pilot plant that is attached to a full-scale coal-fired power plant and operated on flue gas from
428 the power plant. The steps involved in the scale-up calculations are provided in the following
429 subsections:

430 **5.1.1 Estimation of lean solvent flow rate**

431 The lean solvent flow rate required to capture 90% of the CO₂ in the flue gas entering the
432 absorber of the Brindisi pilot plant is estimated based on the absorption capacity of 0.2 mol
433 CO₂/mol MEA, lean solvent MEA concentration of 30 wt%, CO₂ mass fraction of 0.1608 and
434 flue gas mass flow rate of 3.22 kg/s. The estimation is carried out by assuming a constant flow
435 rate for the gas and the solvent throughout the absorber column. The lean solvent flow rate
436 required for the absorption operation is estimated using the approach of Agbonghae et al.
437 (2014) presented in Eq. 13.
438

$$L_{Lean} = \frac{Gx_{CO_2}\varphi_{CO_2}}{100z(\alpha_{Rich} - \alpha_{Lean})} \left[\frac{M_{MEA}}{44.009} \left(1 + \frac{1 - \omega_{MEA}}{\omega_{MEA}} \right) + z\alpha_{Lean} \right] \quad 13$$

439 With regard to the stripper, the total solvent flow is equivalent to the sum of the mass flow rate
440 of the rich solvent and reflux rate while the gas flow rate is equivalent to the boil-up rate needed
441 to maintain the CO₂ loading in the lean solvent at 0.23 CO₂/per mole MEA. Based on these
442 calculations, the solvent flow rate to the absorber and stripper was estimated to be 10.88 kg/s
443 and 11.5 kg/s respectively. The gas (vapour) phase required for the desorption of CO₂ was
444 estimated to be 1.62 kg/s.
445

446 **5.1.2 Estimation of columns diameter**

447 The diameter of the absorber and the stripper is estimated using Eqs. 10 to 12 presented in
 448 section 4. Information regarding the density and kinematic viscosity of the MEA solvent was
 449 obtained from the SRP pilot plant model. Column packings used in the SRP pilot plant were
 450 adopted in the Brindisi plant. The absorber was packed with IMTP 40 with F_P of 78.7 m^{-1} (24
 451 ft^{-1}) and the stripper was packed with Flexipac 1Y with F_P of 168.3 m^{-1} (51.3 ft^{-1}). Using the
 452 given flue gas and the estimated solvent flow rates together with the values of parameters
 453 provided in Table 8, the superficial gas velocities in the absorber and the stripper were
 454 estimated from the flooding gas velocity as 1.83 and 1.20 m/s respectively.

455 **Table 8**

456 Parameters used to estimate the flooding velocity in the absorber and the stripper

Parameter	Absorber	Stripper
$\rho_L \text{ (kg/m}^3\text{)}$	1017.06	1019.88
$\rho_G \text{ (kg/m}^3\text{)}$	1.03	1.02
A	-0.11	-0.07
B	-0.91	-0.89
C	0.72	0.79

457
 458 Based on the gas velocities, the diameter of the absorber and the stripper were calculated from
 459 Eq. 12 to be 1.46 m and 1.28 m respectively. The values obtained for the absorber and the
 460 stripper diameter are similar to the values of 1.5 m and 1.3 m reported for the absorber and the
 461 stripper of the Brindisi pilot plant. The percentage deviations of the estimated column
 462 diameters from those of the Brindisi pilot plant are 2.6 % and 1.54 %, which are within an
 463 acceptable range. The fact that the method proposed herein is able to estimate the diameter of
 464 the absorber and the stripper of an existing plant validates the approach and demonstrates that
 465 it can confidently be used to estimate the diameter of a column required for an absorption
 466 process.

467 **5.1.3 Estimation of packing height**

468 The height of packing (Z_T) require for a given separation in a packed column is most often
 469 expressed in terms of the overall gas-phase mass transfer coefficient and the gas composition.
 470 Based on this, the packing height of the column can be calculated with the expression (Seader
 471 et al., 2006).

$$Z_T = \frac{G_i}{K_G a P} \int_{y_{CO_2,in}}^{y_{CO_2,out}} \frac{dy}{y - y^*} \quad (14)$$

472 The right-hand side of Eq.14 can be written more conveniently as a product of two terms
 473 involving the height and number of transfer units.

$$Z_T = H_{OG} \cdot N_{OG} \quad (15)$$

474 The N_{OG} is the number of (gas) transfer units and can be expressed as:

$$N_{OG} = \int_{y_{CO_2,in}}^{y_{CO_2,out}} \frac{dy}{y - y^*} = \ln \left(\frac{y_{CO_2,in}}{y_{CO_2,out}} \right) \quad (16)$$

475 The larger the value of N_{OG} , the higher the height of the packed column needed to achieve the
 476 required separation. Eq. 16 assumes that the term y^* which is the concentration of CO_2 in
 477 equilibrium with the bulk concentration is negligible because of the fast reaction between CO_2
 478 and the MEA solution and because of the negligible equilibrium partial pressure of CO_2
 479 (Aroonwilas and Veawab, 2004; Fu et al., 2014; Khan et al., 2011).

480 The H_{OG} is the height of a transfer unit, it shows the efficiency of the packing i.e. the smaller
 481 the value of H_{OG} the more efficient the contacting (Coulson and Richardson, 2002). The value
 482 of H_{OG} was computed by:

$$H_{OG} = \frac{G_i}{K_G a P} \quad (17)$$

483 Dugas (2006) in a series of experiments performed at the SRP pilot plant reported the mass
 484 transfer performance of the packing (IMTP 40) in terms of $K_G a$ in the absorber used to absorb
 485 CO_2 from the flue gas using MEA. The concentrations of the MEA in the solvent and CO_2 in
 486 the flue gas, as well as the operating conditions i.e. temperature and pressure used to obtain the
 487 $K_G a$ values are similar to that of the Brindisi pilot plant. Considering that the same packing is
 488 used in the Brindisi pilot plant, the $K_G a$ values are not expected to change markedly. In view
 489 of this, the $K_G a$ values reported by Dugas (2006) were used to estimate the H_{OG} of the absorber.

490 Based on the above, the packing height (Z_T) of the absorber was estimated to be 22.55 m using
 491 the values of the parameters summarised in Table 9.

492

493

494 **Table 9**

495 Calculated values of parameters used to estimate the absorber packed bed

Parameters	Value
N_{OG}	4.1
K_{Ga} (kmol/m ³ s. bar)	1.22×10^{-2}
G_i (kmol/s m ²)	0.06
P (bar)	1.00
H_{OG} (m)	5.50

496

497 The packing height of the stripper could not be determined using the same approach for the
 498 absorber because the K_{Ga} values for the runs with the Flexipac 1Y in the stripper were not
 499 reported (Dugas, 2006). During the experiments, negative CO₂ driving force was encountered
 500 at the top of the stripper, that made it impossible to calculate the log mean driving force and
 501 the mass transfer coefficient of the Flexipac 1Y packing. Therefore, the packing height of the
 502 stripper was determined using a different approach that involves the summation of the HETPs
 503 of stages in the stripper. This is the same approach used in Agbonghae et al. (2014) to estimate
 504 the stripper packing height. The packing height of a stripper with N number of stages can be
 505 estimated as follows; (Agbonghae et al., 2014).

$$506 \quad Z_{T,Stripper} = \sum_{i=2}^{N-1} HETP_i \quad (18)$$

507 The approach was implemented using the calculator block in Aspen Plus[®] to automatically
 508 adjust the ending stage number of the packed section to the number of stages while fixing the
 509 starting stage of the packed section. The starting stage for the stripper was fixed at 2, also a
 510 design specification for the lean loading was set at 0.23 mol CO₂/mol MEA. Starting with a
 511 generic total stage number of 5, the number of stages in the stripper was continuously increased
 512 by 1, till a certain point where a further increase had a negligible effect on the reboiler duty.
 513 Using this approach, the packing height of the stripper was determined to be 11.4 m. Details of
 514 this approach can be obtained is available in Agbonghae et al. (2014). Table 10 shows how the

515 scale-up result for the diameter and packing height of the absorber and stripper compare with
 516 the pilot plant.

517 **Table 10**

518 Comparison of results between the scale-up and the pilot plant measurement

	Pilot plant		Scale-up	
	Absorber	Stripper	Absorber	Stripper
Diameter (m)	1.50	1.30	1.46	1.28
Height (m)	22	11	22.55	11.40

519

520 **5.2. Scale-up of the Brindisi CO₂ capture plant to commercial CO₂ capture plant**

521 Having applied the approach to scale-up between existing capture plants in section 5.1, the
 522 Brindisi CO₂ capture plant was scaled up to deal with the flue gas equivalent to that discharge
 523 by a 250 MW_e CCGT power plant described in Canepa et al. (2013). Based on the amount of
 524 flue gas processed, the capacity of the commercial CO₂ capture plant is about 110 times that of
 525 the Brindisi CO₂ capture plant. The operating conditions of the columns and the input
 526 conditions of the flue gas after treatment to remove acid gases, oxygen and particulates matter
 527 are given in Table 11. These conditions were chosen to be the same as those reported for the
 528 case without exhaust gas recirculation in Canepa et al. (2013). This was done to enable the
 529 comparison of results obtained from this study with those obtained from that study Canepa et
 530 al. (2013) who had previously scaled up from the SRP pilot plant based on these same
 531 conditions using the GPDC method and assumed pressure drop of 412 Pa/m of packing.

532 **Table 11**

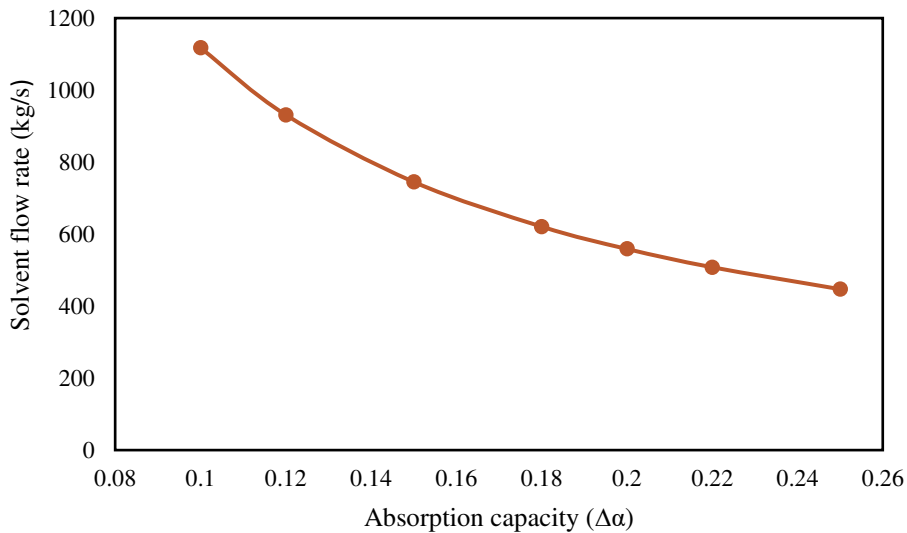
533 Inlet conditions of the PCC capture plant (Canepa et al., 2013)

Compositions (Mass fraction)	Value
CO ₂	0.076
H ₂ O	0.047
N ₂	0.862
Argon	0.015
Flue gas temperature (K)	313
Lean solvent temperature	313
Lean MEA concentration (wt%)	30
CO ₂ Capture level (%)	90
Mass flow rate (kg/s)	356
Absorber pressure (kPa)	101
Absorber packing type	IMTP no. 40

Stripper pressure (kPa)	162
Stripper packing type	Flexipac 1Y

534

535 Going by the CO₂ mass fraction, flue gas mass flow rate, lean MEA concentration and the CO₂
536 capture level in Table 11, the solvent flow rate required by the commercial CO₂ capture plant
537 to treat the flue gas was estimated with Eq. 13. The amount of solvent flow required is
538 dependent on its absorption capacity. The impact of the absorption capacity ($\Delta\alpha$) on the solvent
539 flow rate required by the commercial CO₂ capture plant is shown in Fig. 6. The solvent flow
540 rate to the absorber was estimated to be 669 kg/s. Likewise, the solvent flow rate to the stripper
541 was obtained to be 583 kg/s while the vapour flow rate (boiled-up rate) was obtained to be 58
542 kg/s.



543

Fig. 6 Solvent flow rate at different absorption capacity

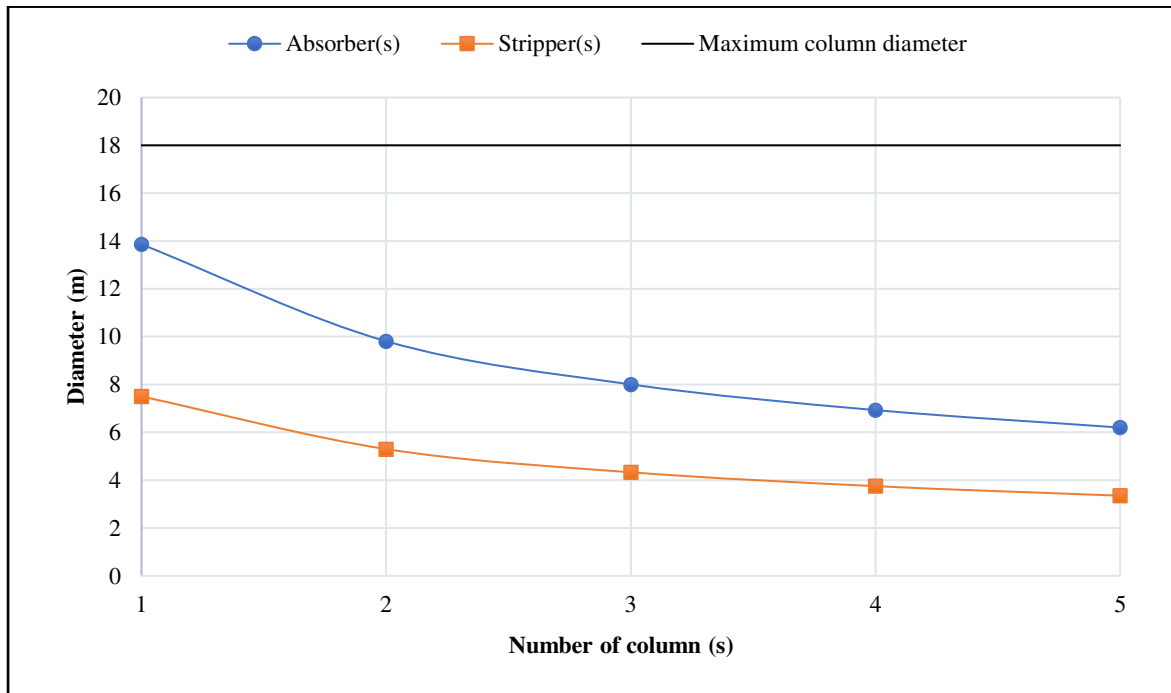
544

545 The diameter of the absorber and the stripper required by the commercial CO₂ capture plant
546 were determined as earlier illustrated in section 5.1.2 using Eqs. 10 – 12. Physical properties
547 such as density and kinematic viscosity useful for calculation were obtained from the Brindisi
548 CO₂ capture pilot plant model simulation. The flooding velocity ($V_{G,f}$) and the operating
549 superficial gas velocity V_G were determined to be 3.25 m/s and 2.27 m/s in the absorber, and
550 1.83 m/s and 1.28 m/s in the stripper respectively. The vapour flow rate was far lower in the
551 stripper than in the absorber thereby making the rich amine flow rate the deciding factor in the
552 design, hence a smaller diameter than the absorber.

553 Based on the superficial gas velocities in the columns, the diameter of the absorber and the
554 stripper required by the commercial CO₂ capture plant was determined to be 13.86m and 7.50
555 m respectively. The relationship between the diameter and the number of columns required by

556 the capture plant is presented in Fig. 7. This was based on what can be delivered by the state-
 557 of-the-art technology and maximum column diameter of 18 m for a commercial CO₂ capture
 558 plant (IEA-GHG, 2006; Reddy et al., 2013, 2008; Scherffius et al., 2013). Moreover, absorbers
 559 of similar diameter have been designed and built by Fluor just as strippers of similar diameter
 560 have been constructed and used for SO₂ stripping in power plants (Dutta et al., 2017; Reddy et
 561 al., 2008).

562



563

564

565 **Fig. 7** Relationship of the diameter of columns and the number of columns

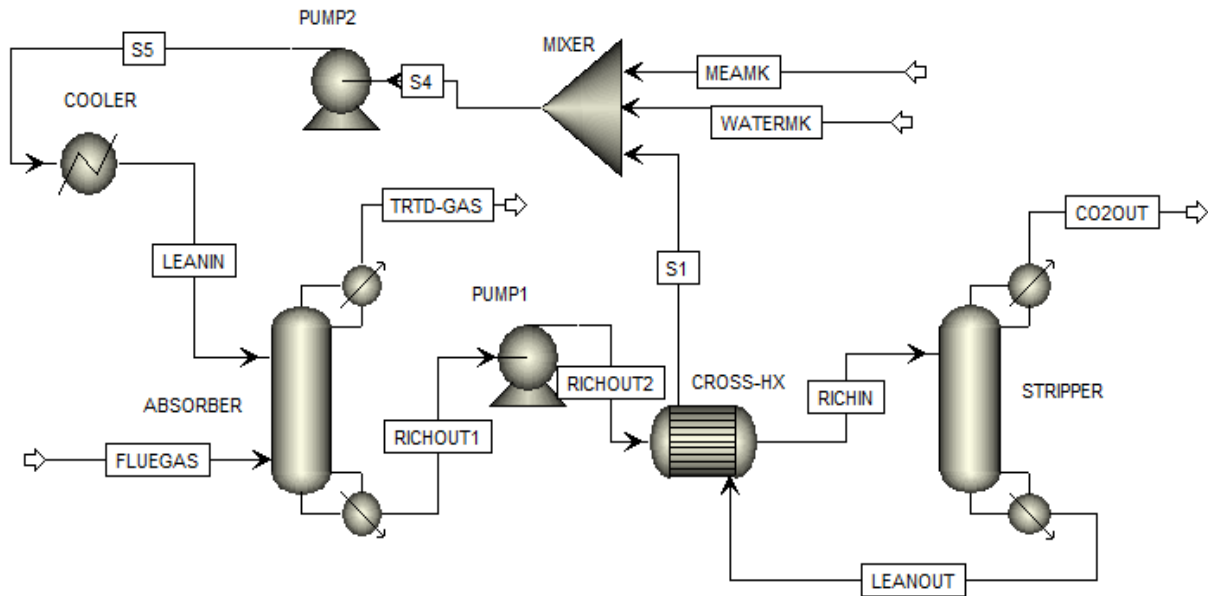
566

567 The diameter calculated for the absorber and the stripper in this study was found to be below
 568 this value. As a result, a single absorber and a single stripper were selected for the commercial-
 569 scale CO₂ capture plant in order to minimize the number of absorption trains and reduce the
 570 complexity of the plant. As a consequence, the plant footprint and capital cost are reduced. The
 571 packing height of 28.5 m was arrived at for the absorber using the approach presented in section
 572 5.1.3. This same value of packing height was used for the stripper.

573

574 **6. Simulation of the commercial CO₂ capture plant**

575 The commercial-scale CO₂ capture plant was simulated in Aspen Plus® V8.4 and the flowsheet
 576 that was developed for the plant in this work is shown in Fig. 8. The comparison of results
 577 obtained from the final simulation with Canepa et al. (2013) is presented in Table 12.



578
 579 **Fig. 8** Process flowsheet of the commercial CO₂ in Aspen Plus®

580 It can be observed from Table 12 that this study achieves lesser equipment size than those
 581 reported in Canepa et al. (2013) for the absorber and the stripper columns. In Canepa et al.
 582 (2013), two absorbers each of diameter 9.5 m and a stripper of diameter 8.2 m were required
 583 by the CO₂ capture plant to treat flue gas from a 250 CCGT power plant. In this study, a single
 584 absorber and stripper of diameters 13.86 m and 7.5 m respectively were designed for the same
 585 capture plant. The higher column diameter reported by Canepa et al. (2013) might be due to
 586 the pressure drop of 412 Pa/m packing assumed in columns sizing is higher than the actual
 587 pressure drop in the absorber and the stripper of a CO₂ capture plant for a 250 CCGT power
 588 plant.

589
 590
 591
 592
 593

594 **Table 12**

595 Comparison of key results obtained from this work with those obtained by Canepa et al. (2013)

	Canepa et al. (2013)	Results from this work
Lean solvent flow rate (kg/s)	720.46	705.23
L/G ratio (mol/mol)	2.29	2.23
Lean solvent loading (mol CO ₂ /mol MEA)	0.30	0.30
Rich solvent loading (mol CO ₂ /mol MEA)	0.45	0.47
CO ₂ capture level (%)	90	90
Flooding ratio (%)	Not reported	70
Absorber		
Number of absorbers	2 ^a	1
Absorber packing	IMTP no. 40	IMTP no. 40
Absorber diameter (m)	9.50	13.86
Absorber packing height (m)	30	28.50
Absorber pressure drop (Pa/m)	412	241
Stripper		
Number of strippers	1	1
Stripper diameter (m)	8.20	7.50
Stripper packing height (m)	30	28.50
Stripper pressure drop (Pa/m)	412	57
Reboiler temperature (°C)	117	115.70
Reboiler duty (MW)	121	115.30
Specific duty (GJ/tonCO ₂)	4.97	4.69
Condenser temperature (°C)	25	25

596 ^a A single absorber will result in diameter of 14.10 m.

597 The commercial-scale CO₂ capture developed in this study achieved a pressure drop of 241
598 Pa/m and 57 Pa/m of packing in the absorber and in the stripper respectively. The pressure drop
599 in the stripper is much lower than the absorber because the vapour flow in the stripper is much
600 lower and the structured packing used in the stripper is expected to provide lower gas-phase
601 pressure drop than the random packing used in the absorber. The pressure drop in the absorber
602 and the stripper are less for this study indicating a reduction in power loss due to pumping
603 which would translate to a reduction in the operational costs of the capture process.

604 Meanwhile, the absorber and the stripper packing height in this study are also smaller. Canepa
605 et al. (2013) reported the same packing height of 30 m for the absorber and the stripper in their
606 study without giving the details of how the values were arrived at. This was despite using a
607 structured packing (Flexipac 1Y) with higher mass transfer efficiency and lower HETP (which
608 should reduce packing height) in the stripper. The results in this study will support more

609 accurate estimation of the capital cost of the process since according to Abu-Zahra et al. (2007)
610 the absorber and the stripper account for about 55% and 17% of the total equipment purchase
611 cost for the whole CO₂ capture process. The specific duty of the CO₂ capture plant attained a
612 value of 4.69 GJ/ton CO₂ representing a 5.63% reduction in the value reported by Canepa et
613 al. (2013). The solvent flow rate is also less for this study because more CO₂ is absorbed as
614 reflected in the rich loading which is slightly higher in this study. The lower solvent flow rate
615 would reduce the energy consumption for pumping and regeneration thereby reducing the
616 operating cost of the process.

617 **7. Conclusions**

618 A steady-state model for the solvent-based post-combustion CO₂ capture plant using MEA has
619 been developed and validated at pilot scale in Aspen Plus[®]. The validation results showed good
620 agreement between the model predictions and the pilot plants measurements. A new scale-up
621 method for estimating packed column diameter based on the use of flooding gas velocity is
622 proposed in this paper. The scale-up method was validated by applying it to the scale-up
623 between two existing pilot plant sizes. The method was able to estimate the diameter of the
624 absorber and the stripper with deviations of 2.6 % and 2.54 % respectively. The validation
625 showed that the method could be used to estimate the diameter of the packed column used in
626 the CO₂ capture process. Furthermore, it was used to scale up the validated model from pilot
627 scale to commercial scale to process flue gas from a 250 MW_e CCGT power plant. The results
628 obtained show that estimates using the GPDC method in literature may be significantly higher
629 than required. In addition, with our approach, it was found that for commercial-scale cases, the
630 solvent flow rate and energy consumption were less by about 2.12% and 5.63% compared to
631 the GPDC approach. Therefore, the capital and operating costs for the process using the newly
632 proposed scale-up method could be lower based on our estimates of the column dimensions,
633 solvent flow rate and specific reboiler duty.

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