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Modelling and Simulation of an Industrial Riser in Fluid Catalytic Cracking Process

Yakubu M. John, Raj Patel, Iqbal M. Mujtaba*

Chemical Engineering Division, School of Engineering, Faculty of Engineering and Informatics, University of Bradford, Bradford, BD7 1DP, UK

*Corresponding author I.M.Mujtaba@bradford.ac.uk

Abstract

Fluid Catalytic Cracking (FCC) unit is an important unit of modern refineries and any improvement in the unit's operations and design to increase yield and meet the ever increasing demand for fuel brings about the overall profitability of the FCC. In this work, simulation of an FCC riser of varied diameter was carried out to improve the unit's operations and design, and the results are compared with risers of different diameters. The riser with varied diameter produces 53.4 wt%, a 3.18% increased yield of gasoline at low catalyst to oil (C/O) ratio of 1.27 compared to 51.7 wt% from a 1 m diameter riser. At increased C/O ratio, more gases and coke are produced in the varied diameter riser. Larger diameter demands more catalyst but yields more gases. Process variables can be directly correlated with yield of gasoline, which can aid process design.

Keywords: FCC Unit, Riser, Variable Diameter, Simulation, Modelling.

1. Introduction

The FCC unit is the workhorse of modern refineries (Bollas et al., 2007), which converts gas oil into lighter hydrocarbons used as valuable transportation fuels such as gasoline and diesel. A typical barrel of crude produces approximately 20% straight run gasoline. However, demand is nearly 50% per barrel and hence there is the need for an efficient process to increase the gasoline production. In the FCC unit, gasoline is produced in the riser and therefore it must be given considerable attention for improvement in gasoline yields.

To meet the demand for gasoline, many researchers have considered the simulation of the riser as a major strategy to improve the yield of gasoline. To do this, some important success factors like the riser design and operations must be improved. Two important factors to consider in trying to achieve optimum yield of gasoline in the riser, is to have uniform

33 catalyst density and very effective hydrodynamics. In situations where the catalyst activity is
34 excellent but the yield poor, the cause would be attributed primarily to the riser
35 hydrodynamics (Kalota and Rahmim, 2003), which is a function of riser design. Therefore,
36 riser diameter is an important factor to consider because of its effect on the riser
37 hydrodynamics.

38 Although a lot of work has been carried out on the modelling of the riser, it is done by
39 considering the riser to be of a uniform cross section (Fernandes et al., 2007, Duduku
40 Krishnaiah, 2007, Gupta and Subba Rao, 2003, Elshishini and Elnashaie, 1990). For some,
41 the riser comprises of a number of equal sized compartments (or volume elements) of circular
42 cross section, but not varied diameters (Gupta et al., 2007), and for others it comprises of a
43 cylindrical vertical vessel where cracking of gas oil is carried out using a catalyst in a
44 vaporised upward fashion (Han and Chung, 2001a). Even when a comprehensive three-
45 dimensional (3-D) heterogeneous riser model was applied to simulate the turbulent gas–solid
46 flow and reaction in a polydisperse FCC riser, the entire zones of the riser were considered as
47 a uniform cross sectional tubes (Li et al., 2013).

48 The riser unit has many sections; feed preheater, the vaporization section and the riser, which
49 are sometimes modelled differently. Although an attempt to simulate the riser unit with
50 varied diameter (between 1 m at the bottom to 1.4 m at the top) was made (Novia et al.,
51 2007), only a quarter of the riser was considered because they modelled the riser unit in two
52 sections; the vaporization section (found to have no chemical reactions) as 1 m diameter and
53 the riser section as 1.4 m, a uniform cross section. They also included the vaporization
54 section in the riser unit model. In some cases, the model of the vaporization section was
55 included in the riser unit simulation but the length of the riser (uniform cross section)
56 considered did not include the vaporization section (Han and Chung, 2001a, Han and Chung,
57 2001b). It is also clear that the vaporization section of the riser unit has unique
58 hydrodynamics and can be treated differently, because it takes about 3% of the riser residence
59 time (Ali and Rohani, 1997). For this reason, the riser has been modelled differently from the
60 vaporization section with the assumption that the gas oil vaporizes instantaneously (Ahari et
61 al., 2008, Al-Sabawi et al., 2006, Araujo-Monroy and López-Isunza, 2006). Therefore,
62 modelling the riser unit by having different diameters for the vaporization and riser sections,
63 is different from modelling the system where the diameter of the riser is varied. This is what
64 this work sets to achieve; to model the riser section as a varied diameter with three different
65 cross sections.

66 The riser unit of the FCC unit of Kaduna Refinery and Petrochemicals Company (KRPC) in
67 Nigeria is a vertical cylinder but with varied diameters. This design is such that the reaction
68 proceeds as the catalyst and vapour mixture flows up through the riser. The lower part of the
69 riser is sized to provide sufficient pick up velocity and as cracking proceeds, the riser
70 diameter is increased to handle the increasing volume and provide the desired reaction time.
71 The mixture then flows through the remainder of the vertical riser.

72 This work modelled the riser according to geometric differences of the riser and validated
73 against industrial data. gPROMS software is used for the simulation with C/O ratio, catalyst
74 temperature and gas phase temperature are used as manipulating variables. The various
75 effects of the riser geometry on the conversion of gas oil and yield of gasoline were
76 determined.

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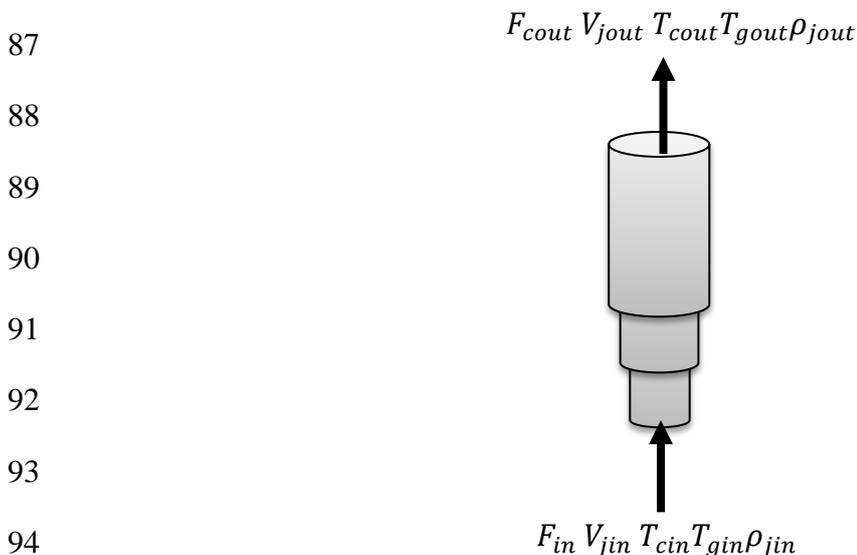
78 **Process Modelling**

79 This section presents the description of the riser and its model assumptions, the model
80 equations, degree of freedom analysis, the parameters used and method of solution of the
81 model.

82 **1.1 The Riser**

83 The riser has always been modelled as a single vertical tube or cylinder but risers can have
84 varied diameters. The riser unit of the FCC unit of KRPC is a type with varied diameters as
85 shown in Figure 1.

86



95

Figure 1: The Riser

96 It is a vertical cylinder with three different compartments, each of different diameter and
97 height. For simplicity, the connection between each compartment is made flat as shown in
98 Figure 1. The first compartment at the bottom has a diameter of 1.0 m and 3.965 m height.
99 The middle compartment has a diameter of 1.35 m and 3.753 m height and the third
100 compartment at the top has a diameter of 1.6 m and 17.6 m height. The entire height of the
101 riser is 25.36 m. The C/O ratio of the unit varied from 2.0 to 6.5, as set by the production
102 unit. The riser is modelled as a one-dimensional plug flow reactor without axial and radial
103 dispersion, and mass and energy balance equations for the catalyst and gaseous phases are
104 obtained under the following assumptions:

- 105 • the hydrocarbon feed instantly vaporizes as it comes into contact with the hot catalyst
106 from the regenerator, then moves upwards in thermal equilibrium with the catalyst
107 and there is no loss of heat from the riser (Ali et al., 1997).
- 108 • The cracking reactions only take place in the riser, on catalyst surface and fast enough
109 to justify steady state model.
- 110 • The vaporization section of the riser was not considered in the simulation.
- 111 • The momentum equations of the system were not included in the simulation.
- 112 • The rate of dispersion and adsorption inside the catalyst particles are negligible.
- 113 • The coke deposited on the catalyst does not affect the velocity of the fluid.

114 At the entrance of the riser, the feed vaporizes immediately when it comes in contact with the
115 regenerated catalyst and flows pneumatically upward in the riser as cracking reactions goes
116 on the surface of the catalyst to form products. The products in this case are gasoline, gases
117 and coke based on the four lumped model. The four lumped model to represent the kinetic
118 model which determines the weight fractions of components in product stream as well as the
119 reactants involved in the riser was obtained from the literature (Lee et al., 1989) and
120 presented in Figure 2. The relevant information related to Figure 2 are presented in Table 1.

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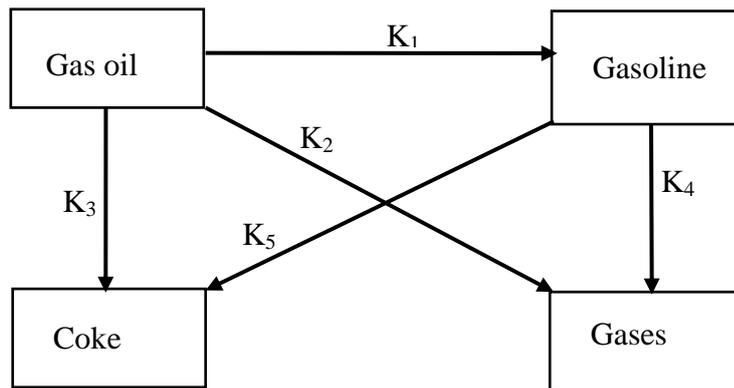


Figure 2: Four-lumped model of gas oil cracking reactions (Lee et al., 1989).

The formulation of a kinetic model that includes all chemical reactions responsible for the catalytic cracking of gas oil is extremely difficult and therefore, most researchers group the components into lumps to make it easier to account for the various valuable petroleum fractions. The kinetic model shown in Figure 2 is the breaking down of gas oil into gases, coke and gasoline. It is the most acceptable and widely used for its accuracy in consolidating the very important refinery fractions. The cracking reaction is endothermic and the heat required for endothermic gas oil cracking is supplied from the regenerator by burning coke formed during catalyst deactivation in the riser. Thus, accurate prediction of the coke formed due to catalyst deactivation is crucial. The coke formed aids heat integration and reactor temperature control which is one of the advantages of the four-lump model (Han and Chung, 2001a).

In Figure 2, K_1 , K_2 , K_3 , K_4 and K_5 are the overall rate constants for the cracking reactions while their kinetic parameters are shown in Table 1. The cracking of gas oil to form gasoline, gases and coke is considered to be a second order reaction, while the cracking of gasoline to form gases and coke is considered a first order reaction.

155 Table 1: Kinetic parameters for gas oil cracking (Han and Chung, 2001b)

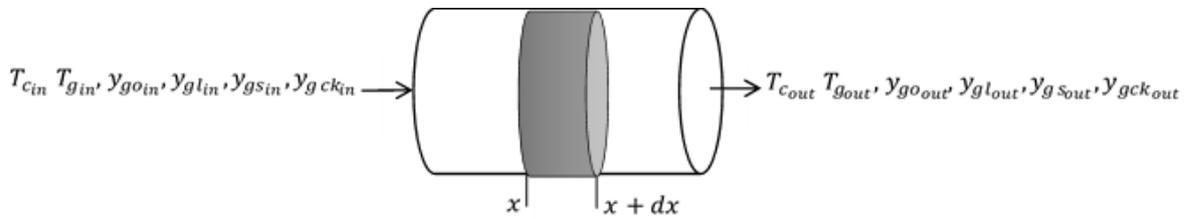
4-lump cracking reactions	Frequency factor (s ⁻¹)	Activation Energy (kJ/kg mol)	Heat of reaction (kJ/kg)	Reaction Path	Order of reaction
Gas oil – Gasoline	1457.50	57,359	195	1	2
Gas oil –C ₁ -C ₄ gases	127.59	52,754	670	2	2
Gas oil- Coke	1.98	31,820	745	3	2
Gasoline–C ₁ -C ₄ gases	256.81	65,733	530	4	1
Gasoline- Coke	6.29x10 ⁻⁴	66,570	690	5	1
Catalyst deactivation	1.1x10 ⁻⁵	49,000			

156

157 1.2 The Model Equations

158 The riser shown in Figure 1 is modelled as a one-dimensional tubular reactor using mass and
 159 energy balance equations. The riser composition varies along its length, and because there is
 160 reaction going on in the riser, the dependent variables were deduced from the energy and
 161 material balance carried out on a differential element of volume as shown in Figure 3.
 162 Equations 1 and 2 are deduced temperatures of catalyst and gases respectively. Figure 3
 163 shows the inlet and outlet compositions of the control volume.

164



165

166 Figure 3: A control volume of the riser

167

$$168 \frac{dT_c}{dx} = \frac{\Omega h_p A_p}{F_c C_{pc}} (T_g - T_c) \quad (1)$$

$$169 \frac{dT_g}{dx} = \frac{\Omega}{F_g C_{pg}} [h_p A_p (T_c - T_g) + \rho_c \epsilon_c Q_{react}] \quad (2)$$

170 Equations 3 – 6 are deduced mass fractions of gas oil, gasoline, gases and coke respectively
 171 from the mass balance carried out on the control volume of the riser:

$$172 \frac{dy_{go}}{dx} = \frac{\rho_c \epsilon_c \Omega \theta_c}{F_g} R_{go} \quad (3)$$

173
$$\frac{dy_{gl}}{dx} = \frac{\rho_c \varepsilon_c \Omega \phi_c}{F_g} R_{gl} \quad (4)$$

174
$$\frac{dy_{gs}}{dx} = \frac{\rho_c \varepsilon_c \Omega \phi_c}{F_g} R_{gs} \quad (5)$$

175
$$\frac{dy_{ck}}{dx} = \frac{\rho_c \varepsilon_c \Omega}{F_g} R_{ck} \quad (6)$$

176 The rates of reaction for gas oil R_{go} , gasoline R_{gl} , light gas R_{gs} , and coke R_{ck} , are given as

177
$$R_{go} = -(K_1 + K_2 + K_3)y_{go}^2 \quad (7)$$

178
$$R_{gl} = (K_1 y_{go}^2 - K_4 y_{gl} - K_5 y_{gl}) \quad (8)$$

179
$$R_{gs} = (K_2 y_{go}^2 - K_4 y_{gl}) \quad (9)$$

180
$$R_{ck} = (K_3 y_{go}^2 - K_5 y_{gl}) \quad (10)$$

181 The rate constants K_i , of reaction path $i = 1, \dots, 5$ and their corresponding frequency factors
182 k_{i0} are given as:

183
$$K_1 = k_{10} \exp\left(\frac{-E_1}{RT_g}\right) \quad (11)$$

184
$$K_2 = k_{20} \exp\left(\frac{-E_2}{RT_g}\right) \quad (12)$$

185
$$K_3 = k_{30} \exp\left(\frac{-E_3}{RT_g}\right) \quad (13)$$

186
$$K_4 = k_{40} \exp\left(\frac{-E_4}{RT_g}\right) \quad (14)$$

187
$$k_5 = K_{50} \exp\left(\frac{-E_5}{RT_g}\right) \quad (15)$$

188

189 Q_{react} is the rate of heat generation or heat removal by reaction and can be written as

190
$$Q_{react} = -(\Delta H_1 K_1 y_{go}^2 + \Delta H_2 K_2 y_{go}^2 + \Delta H_3 K_3 y_{go}^2 + \Delta H_4 K_4 y_{gl} + \Delta H_5 K_5 y_{gl}) \phi_c \quad (16)$$

191

192 Where the gas volume fraction, ε_g , and catalyst volume fraction, ε_c , can be obtained from:

193
$$\varepsilon_g = 1 - \varepsilon_c \quad (17)$$

194 The catalyst volume fraction, ε_c , can be written as

195 $\varepsilon_c = \frac{F_c}{v_c \rho_c \Omega}$ (18)

196 Cross sectional area of the riser, Ω , is given as

197 $\Omega = \frac{\pi D^2}{4}$ (19)

198 Effective interface heat transfer area per unit volume between the catalyst and gas phases,
199 A_{ptc} is derived as:

200 $A_{ptc} = \frac{6}{0.72 d_c} * (1 - \varepsilon_g)$ (20)

201 The catalyst deactivation is given by:

202 $\phi_c = \exp(-\alpha_c C_{ck})$ (21)

203 Where;

204 $\alpha_c = \alpha_{c0} \exp\left(\frac{-E_{1c}}{RT_g}\right) (R_{AN})^{\alpha_{c*}}$ (22)

205 and

206 $C_{ck} = C_{ckCL1} + \frac{F_g y_{ck}}{F_c}$ (23)

207 The density of the gas phase is given by:

208 $\rho_g = \frac{F_g}{\varepsilon_g v_g \Omega}$ (24)

209 The riser pressure is given by:

210 $P = \rho_g \frac{RT_g}{M_{wg}}$ (25)

211 The ratio of the mass flowrate of catalyst to the mass flowrate of gas oil is the C/O ratio and it
212 is given by:

213 $C/O \text{ ratio} = \frac{F_c}{F_g}$ (26)

214 **1.3 Degree of Freedom Analysis**

215 The model equations are made up of six (6) ordinary differential equations (ODEs) and
216 twenty (20) algebraic equations (AEs), making a total of twenty six (26) equations. The riser

217 model contains thirty four (34) unknown variables as shown in Table 2. Therefore, the model
 218 is found to have 8 degrees of freedom which are specified in Table 3.

219

220 Table 2: Unknown variables in the riser model equations

Variable type	Symbol	No. of Unknown variables
Temperature	T	4
Pressure and Flowrate	P	4
Weight fraction and density	y_i, ρ	5
Heat rate	Q, A_{ptc}	2
Area and volume fraction	Ω, ε	3
Reaction coefficient	$K_i, R_i, \phi_c, \alpha_c, \Delta H$	16
Total of unknown variables		34

221

222 There is eight degree of freedom for the model equations and they are presented in Table 3.

223 The first six variables in Table 3 are boundary conditions at $x = 0$, the entrance of the riser.

224

225 Table 3: Variables to satisfy degree of freedom

Variable	Value
y_{go} (Weight fraction of gas-oil)	1
y_{gl} (Weight fraction of gasoline)	0
y_{gs} (Weight fraction of gases)	0
y_{ck} (Weight fraction of gas-oil)	0
T_g (Temperature of gas oil, K)	513
T_c (Temperature of catalyst, K)	933
F_c (Catalyst mass flowrate, kg/s)	44.91
F_g (Gas oil mass flowrate, kg/s)	35.36

226

227 Table 4 summarizes the parameters used in this simulation and were obtained from industry
 228 and literature.

229

230

231

232 Table 4: Riser inputs parameters (Han and Chung, 2001b, Nuhu et al., 2012, Ahari et al.,
 233 2008)

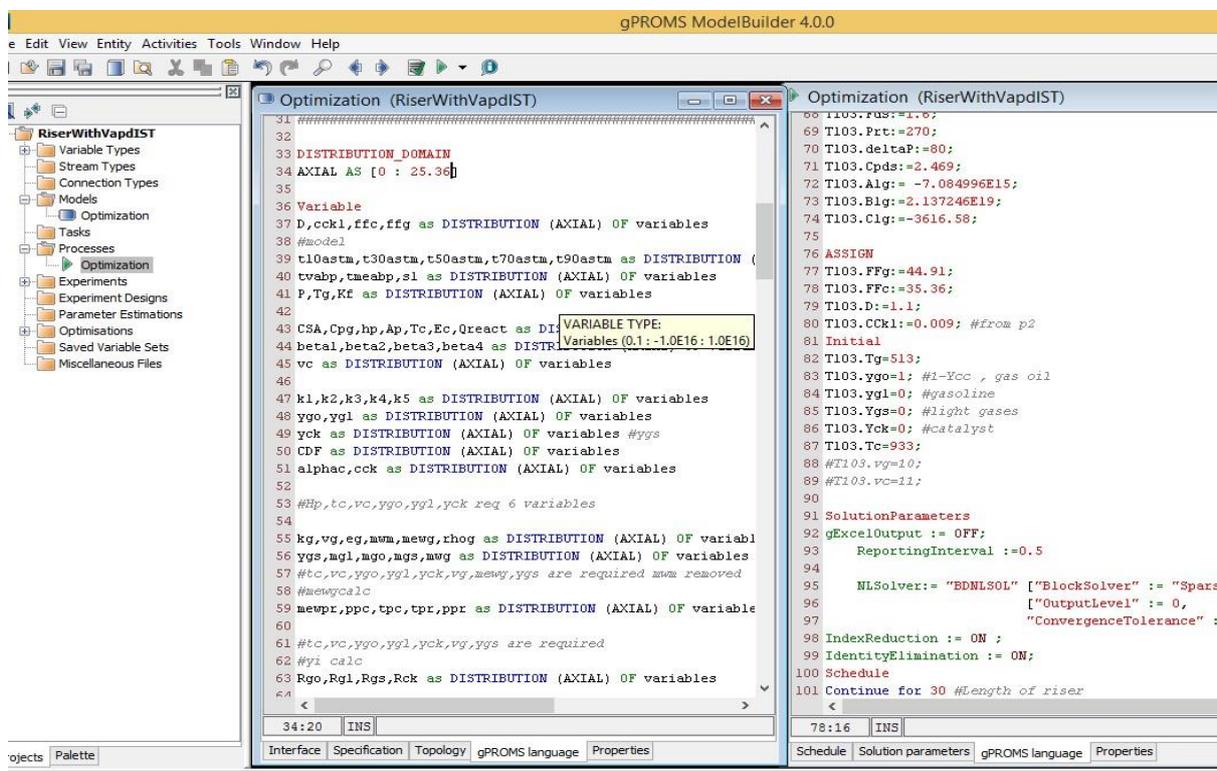
Variable	Value
D (Diameter, m)	1.0, 1.35, 1.6
Riser Length (m)	25.368
T10, T30, T50, T70, T90 (TBP distillation temp at distilled vol%, °C)	351, 380, 409, 445, 490
M _{wgo} (Molecular weight gas oil)	371
M _{wgl} (Molecular weight gasoline)	106.7
M _{wgs} (Molecular weight light gases)	40
M _{wck} (Molecular weight coke)	14.4
D _c (Average particle diameter, m)	0.00007
S _c (Average sphericity of catalyst particles)	0.72
S _g (Specific gravity)	0.897
R _{AN} (Aromatics to Naphthenes ratio in liquid feedstock)	2.1
C _{ckc} (Coke on catalyst, kg coke/kg catalyst)	0.001
α_{c0} (pre-exponential factor of α_c)	1.1×10^{-5}
α_{cs} (Catalyst deactivation coefficient)	0.1177
C _{pg} (Heat capacity of Gasoline, kJ/kg K)	3.33
C _{pc} (Heat capacity of catalyst kJ/kg K)	1.15
ρ_c (Density of catalyst, kg/m ³)	1410
P, Pressure (kPa)	250

234

235 1.4 Model Solution

236 The equations generated for the riser are a set of Differential Algebraic Equations (DAEs) of
 237 Index 1 and gPROMS is used to solve them. gPROMS is a general process modelling system
 238 for simulation, optimisation and control (both steady state and dynamic) of highly complex
 239 processes such as the FCC unit. It is one of the available equation oriented software suitable
 240 for the type of equations developed for the riser of FCC unit. All solvers have been designed
 241 specifically for large-scale systems and there are no limits regarding problem size other than
 242 those imposed by available machine memory (Mujtaba, 2012). In spite of the robustness of
 243 gPROMS, there is no known literature of the use of the software to solve the models of the
 244 FCC unit. This is the first attempt and gPROMS proves to be a reliable software. The riser

245 model is constructed in the model section and the parameters are specified in the process
 246 section of the gPROMS software 4.0.1. as shown in Figure 4. The distributed domain is
 247 defined in the ‘MODEL’ section as ‘DISTRIBUTION DOMAIN’ under which the
 248 distributed variables are defined as DISTRIBUTION (AXIAL) as seen in Figure 4. The
 249 boundary conditions are specified in the ‘PROCESS’ section under the ‘ASSIGN’ subsection,
 250 while the other parameters are specified in the ‘SET’ section. The gPROMS software is
 251 capable of analysing the set of equations to determine the stiffness of the system and calls on
 252 the appropriate solvers, in this case a differential-algebraic solver (DASolver) capable of
 253 solving the system of DAE of the riser model.
 254



255
 256 Figure 4: gPROMS platform for the riser model

257

258 2. Results and Discussions

259 The manipulated variables for the simulation are catalyst-to-oil ratio (C/O), gas oil inlet
 260 temperature and catalyst inlet temperature. The results obtained are presented in Figures (5-
 261 12) and Tables 5 - 9. In Table 5, the results for two different configurations were considered
 262 in the simulation; a 1 m diameter riser and a varied diameter riser. This is to enable
 263 comparison of the two configurations and to study the effect of the diameter variation on the
 264 riser column. In the first simulation run, C/O ratio of 1.27 and 2.4 were used for both 1 m

265 diameter riser and the varied diameter riser at catalyst inlet temperature of 933 K and gas oil
 266 inlet temperature of 513 K. The results are presented in Table 5 along with Plant data for
 267 validation of the model.

268

269 Table 5: Riser outlet weight fractions and temperatures at input C/O ratio of 1.27 and 2.4

Parameter	1 m Diameter		Varied Diameter		Plant data
	1.27	2.4	1.27	2.4	
C/O ratio	1.27	2.4	1.27	2.4	
Temperature of Gas, T_g (K)	579.44	688.54	563.62	685.68	796
Temperature of Catalyst, T_s (K)	584.03	691.97	564.88	686.20	
Gas oil Fraction	0.197	0.065	0.122	0.028	
Gasoline fraction	0.517	0.466	0.534	0.265	0.53
Gases	0.158	0.352	0.200	0.589	0.25
Coke	0.127	0.116	0.14	0.119	0.11

270

271 In the 1 m diameter riser and at C/O ratio of 1.27, the gas oil conversion is 80.3% (0.197 wt
 272 %) producing 51.7% (0.517 wt %) yield of gasoline, 15.8% (0.158 wt %) gases and 12.7%
 273 (0.127 wt %) coke. This indicates that gasoline yield deviated from the plant value (0.53 wt
 274 %) by -2.5%, gases yield deviated from the plant value (0.25 wt %) by -58.22% and coke
 275 yield deviated from the plant value (0.11 wt %) by 13.38%. There is a decrease in the yield of
 276 gases compared with the plant value, which is better for a case where gasoline is the desired
 277 product and needs to be improved to meet market demand. However, in this case it did not
 278 result in higher gasoline yield but produced more coke when compared with plant data. The
 279 gas phase exit temperature is not expected to be more than 800 K, beyond which most of the
 280 gasoline will be converted in a secondary reaction to gases. In this case, the temperature is
 281 579.44 K, which is much lower than the plant exit temperature and it is the reason for the low
 282 yield of gases.

283 At C/O ratio of 2.4 for the same 1 m diameter riser, the gas oil conversion is 93.5% (0.065 wt
 284 %) producing 46.6% (0.466 wt %) yield of gasoline, 35.2% (0.352 wt %) gases and 11.6%
 285 (0.116 wt %) coke. This indicates that gasoline yield deviated from the plant value (0.53 wt
 286 %) by -13.73%, gases yield deviated from the plant value (0.25 wt %) by 28.98% and coke
 287 yield deviated from the plant value (0.11 wt %) by 5.17%. It shows an increase in the yield of
 288 gases compared with the plant value, and resulted in lower gasoline yield compared to plant

289 value due to secondary reactions of gasoline to form gases and coke. The gas phase exit
290 temperature is 688.54 K, which is lower than the plant exit temperature.

291 Comparing the results obtained at C/O ratio of 1.27 and C/O ratio of 2.4 for 1 m diameter
292 riser, it is clearly seen that the gas phase temperature of 688.54 K at C/O ratio of 2.4 is higher
293 than 579.44 K at C/O ratio of 1.27 which explains why the gasoline yield at C/O ratio of 1.27
294 is higher due to less heat available for gasoline secondary reaction to form gases. For the
295 same reason, the yield of gases is higher for C/O ratio of 2.4 than for C/O ratio of 1.27. In
296 the 1 m diameter riser, the higher the C/O ratio, the lower the yield of gasoline and coke, but
297 higher gas oil conversion and yield of gases.

298 Data in Table 5 also shows that in the varied diameter riser and at C/O ratio of 1.27, the gas
299 oil conversion is 87.8% (0.122 wt %) producing 53.4% (0.534 wt %) yield of gasoline, 20.0%
300 (0.200 wt %) gases and 14.0% (0.140 wt %) coke. Gasoline yield deviated from the plant
301 value (0.53 wt %) by 0.749%, gases yield deviated from the plant value (0.25 wt %) by -
302 25.0% and coke yield deviated from the plant value (0.11 wt %) by 21.43%. There is a
303 decrease in the yield of gases compared with the plant value which, in this case gives higher
304 gasoline yield though produced more coke when compared with plant data. The gas phase
305 exit temperature is 563.62 K, which is lower than the plant exit temperature and the reason
306 for the low yield of gases.

307 At C/O ratio of 2.4 for the same varied diameter riser, the gas oil conversion is 97.2% (0.028
308 wt %) producing 26.5% (0.265 wt %) yield of gasoline, 58.9% (0.589 wt %) gases and 11.9%
309 (0.119 wt %) coke. This indicates a 100% deviation of gasoline yield from the plant value
310 (0.53 wt %), gases yield deviated from the plant value (0.25 wt %) by 57.55% and coke yield
311 deviated from the plant value (0.11 wt %) by 7.56%. There is an increase in the yield of gases
312 compared with the plant value and decrease in gasoline yield compared to plant value which
313 is due to secondary conversion of gasoline to gases and more coke. With the increase in
314 diameter of the riser at the top, more residence time for catalyst is created, thereby increasing
315 the secondary reaction of gasoline. The gas phase exit temperature is 685.68 K, which is
316 lower than the plant exit temperature.

317 Comparing the results for C/O ratio of 1.27 and C/O ratio of 2.4 for varied diameter riser, the
318 gas phase temperature of 685.68 K at C/O ratio of 2.4 is higher than 563.62 K at C/O ratio of
319 1.27 which explains why the gasoline yield at the lower C/O ratio is higher due to gasoline
320 secondary reaction to form gases. Also, the yield of gases is higher for C/O ratio of 2.4 than
321 for ratio of 1.27. In the varied diameter riser, it can be concluded that the higher the C/O

322 ratio, the lower the yield of gasoline and coke, but higher gas oil conversion and yield of
 323 gases.

324 For both varied and 1 m diameter risers, the yield of gasoline and coke is higher at 1.27 C/O
 325 ratio than the 2.4 C/O ratio, and more yield of gases and higher gas oil conversion for C/O
 326 ratio of 2.4 than for ratio of 1.27. In conclusion, it is better to use the varied riser at C/O ratio
 327 of 1.27 and 933 K catalyst temperature, because it gives a difference of 0.749% increase of
 328 gasoline with less yield gases and coke.

329 Varying the inlet temperature of catalyst can affect the cracking temperature in the riser and
 330 eventually impact on the yields of the product. The catalyst temperature was increased by 20
 331 °C, from 933 K to 953 K for both 1 m diameter and varied diameter risers. The resulting
 332 yields of the lumps are presented in Tables 6 – 8. Table 6 shows the yield of cracking lumps
 333 at 953 K and C/O ratio of 1.27 for both 1 m diameter riser and varied diameter riser along
 334 with plant data for validation.

335 In the 1 m diameter riser, the gas oil conversion is 82.0% (0.18 wt %) producing 52.7%
 336 (0.527 wt %) yield of gasoline, 16.9% (0.169 wt %) gases and 12.5% (0.125 wt %) coke. The
 337 yield of gasoline deviated from the plant value (0.53 wt %) by -0.57%, gases yield deviated
 338 from the plant value (0.25 wt %) by -47.92% and coke yield deviated from the plant value
 339 (0.11 wt %) by 12.00%. A decrease in the yield of gases compared with the plant value is
 340 observed, but produced more coke when compared with plant data.

341

342 Table 6: Riser outlet weight fractions and temperatures at input C/O ratio of 1.27

Parameter	1 m Diameter	Varied diameter	Plant data
C/O ratio	1.27	1.27	
Temperature of Gas, T_g (K)	587.8	572.24	796
Temperature of Catalyst, T_s (K)	591.40	573.40	
Gas oil fraction	0.180	0.109	
Gasoline fraction	0.527	0.534	0.53
Gases	0.169	0.218	0.25
Coke	0.125	0.138	0.11

343

344 For the varied diameter riser, the gas oil conversion is 89.10% (0.109 wt %) producing 53.4%
 345 (0.534 wt %) yield of gasoline, 21.8% (0.218 wt %) gases and 13.8% (0.138 wt %) coke. The
 346 yield of gasoline deviated from the plant value (0.53 wt %) by -7.49%, gases yield deviated

347 from the plant value (0.25 wt %) by -14.67% and coke yield deviated from the plant value
348 (0.11 wt %) by 20.29%. There is a decrease in the yield of gases compared with the plant
349 value, but an increase in the amount of coke produced when compared with plant data. At
350 catalyst inlet temperature of 953 K, the gas phase exit temperature for the 1 m diameter riser
351 is 587.8 K, which is much lower than the plant exit temperature (796 K) of the gas phase but
352 higher than the exit temperature (572.24 K) of the gas phase for the varied diameter riser.
353 These exit temperatures at catalyst inlet temperature of 953 K are higher compared with the
354 exit temperatures at catalyst inlet temperature of 933 K (see Table 5), meaning that increasing
355 the inlet catalyst temperature has a direct influence on the riser exit temperatures and it
356 results in higher yield of gases at higher catalyst inlet temperature. The yield of gasoline at
357 catalyst inlet temperature of 933 K (Table 5) and 953 K (Table 6) at C/O ratio of 1.27 for the
358 varied diameter riser remained the same (0.534 wt %), which is not the case with the 1 m
359 diameter riser where at 953 K gasoline mass fraction is 0.527 wt % and at 933 K it is 0.517
360 wt %, showing an increase of gasoline yield. This shows that the increase in temperature did
361 not affect the yield of gasoline in the varied diameter riser, but reduced the yield of coke
362 (from 0.14 wt % at 933 K to 0.138 wt % at 953 K). For both risers (1 m diameter and varied
363 diameter), there is more gas produced at higher temperature with lower coke yield.
364 Table 7 shows the yield of cracking lumps at catalyst inlet temperature of 953 K and C/O
365 ratio of 1.84 for both 1 m diameter riser and varied diameter riser along with plant data for
366 validation. Results at C/O ratio of 1.84 at catalyst inlet temperature of 953 K in Table 7 are
367 compared with results at C/O ratio of 1.27 in Table 6 at the same catalyst temperature. In the
368 1 m diameter riser, the gas oil conversion is 90.8% (0.092 wt %) producing 51.9% (0.519 wt
369 %) yield of gasoline, 27.1% (0.271 wt %) gases and 11.9% (0.119 wt %) coke. The yield of
370 gasoline deviates from the plant value (0.53 wt %) by -2.11%, gases yield deviates from the
371 plant value (0.25 wt %) by 7.75% and coke yield deviates from the plant value (0.11 wt %)
372 by 7.56%. This shows an increase in the yield of gases compared with the plant value, but
373 produces more coke when compared with plant data.

374
375
376
377
378
379
380

381 Table 7: Riser outlet weight fractions and temperatures at input C/O ratio of 1.84

Parameter	1 m Diameter	Varied diameter	Plant data
C/O ratio	1.84	1.84	
Temperature of Gas, T_g (K)	653.64	647.64	796
Temperature of Catalyst, T_s (K)	657.37	648.33	
Gas oil fraction	0.092	0.043	
Gasoline fraction	0.519	0.400	0.53
Gases	0.271	0.432	0.25
Coke	0.119	0.124	0.11

382

383 For the varied diameter riser, the gas oil conversion is 95.7% (0.043 wt %) producing 40.0%
 384 (0.40 wt %) yield of gasoline, 43.2% (0.432 wt %) gases and 12.4% (0.124 wt %) coke. The
 385 yield of gasoline deviates from the plant value (0.53 wt %) by -32.5%, gases yield deviates
 386 from the plant value (0.25 wt %) by -42.12% and coke yield deviates from the plant value
 387 (0.11 wt %) by 11.29%. here, there is decrease in the yield of gasoline, but produced more
 388 gases and coke when compared with plant data. The 1 m diameter riser has the better yield of
 389 gasoline than the varied diameter riser. Also, the varied diameter riser produced more gases
 390 and coke than the 1 m diameter even though the 1 m diameter riser has the higher gas phase
 391 exit temperature. Comparing the yields at C/O of 1.27 and 1.84 at catalyst inlet temperature
 392 of 953 K from Tables 6 and 7 respectively, it can be concluded that operating the varied
 393 diameter riser at C/O ratio of 1.27 gives the better yield of gasoline which is the desired
 394 product. However, less coke is produced at C/O ratio of 1.84.

395

396 Table 8 shows the yield of cracking lumps at catalyst inlet temperature of 953 K and C/O
 397 ratio of 2.4 for both 1 m diameter riser and varied diameter riser along with plant data for
 398 validation. In the 1 m diameter riser, the gas oil conversion is 94.5% (0.055 wt %) producing
 399 43.0% (0.43 wt %) yield of gasoline, 40.0% (0.400 wt %) gases and 11.3% (0.113 wt %)
 400 coke. The yield of gasoline deviated from the plant value (0.53 wt %) by -52.49%, yield of
 401 gases deviated from the plant value (0.25 wt %) by 37.5% and coke yield deviated from the
 402 plant value (0.11 wt %) by 2.65%. There is an increase in the yield of gases compared with
 403 the plant value, giving rise to decrease in gasoline yield and produced more coke when
 404 compared with plant data.

405

406 Table 8: Riser outlet weight fractions and temperatures at input C/O ratio of 2.4

Parameter	1 m Diameter	Varied diameter	Plant data
C/O ratio	2.4	2.4	
Temperature of Gas, T_g (K)	703.18	700.82	796
Temperature of Catalyst, T_s (K)	706.15	701.26	
Gas oil fraction	0.055	0.023	
Gasoline fraction	0.430	0.210	0.53
Gases	0.400	0.657	0.25
Coke	0.113	0.116	0.11

407

408 For the varied diameter riser, the gas oil conversion is 97.3% (0.023 wt %) producing 21.0%
409 (0.40 wt %) yield of gasoline, 65.7% (0.657 wt %) gases and 11.6% (0.116 wt %) coke. The
410 yield of gasoline deviates from the plant value (0.53 wt %) by -152.38%, gases yield deviates
411 from the plant value (0.25 wt %) by 61.95% and coke yield deviates from the plant value
412 (0.11 wt %) by 5.17%. This gives a very high decrease in the yield of gasoline and high yield
413 of gases but produced more coke when compared with plant data. The 1 m diameter riser has
414 the better yield of gasoline than the varied diameter riser. Also, the varied riser produced
415 more gases and coke than the 1 m diameter even though the 1 m diameter riser has the higher
416 gas phase exit temperature. Comparing the yields at C/O of 1.27, 1.84 and 2.4 at 953 K from
417 Tables 6, 7 and 8 respectively, it can be concluded that operating the varied diameter riser at
418 C/O ratio of 1.27 gives the better yield of gasoline which is the desire product, however, less
419 coke is produced at C/O ratio of 1.84 and lesser coke at the C/O ratio of 2.4. The varied riser
420 is the better choice and C/O ratio 1.27 appears to be the best condition to operate at 953 K.

421

422 Table 9 shows the simulation results when considering four different risers; 1 m diameter
423 riser, 1.35 m diameter riser, 1.6 m diameter riser and the varied diameter riser. These were
424 simulated at 933 K and C/O ratio of 1.84 and results obtained from the products at the exit of
425 the risers were compared with plant data. In the 1 m diameter riser, the gas oil conversion is
426 89.4% (0.106 wt %) producing 53.0% (0.53 wt %) yield of gasoline, 24.40% (0.244 wt %)
427 gases and 12.1% (0.121 wt %) coke. The yield of gasoline did not deviate from the plant
428 value (0.53 wt %), it is the same (0.0% deviation). The yield of gases deviates from the plant
429 value (0.25 wt %) by -2.45% and coke yield deviates from the plant value (0.11 wt %) by
430 9.09%.

431

432 Table 9: Riser outlet weight fractions and temperatures at input $T_s = 933$ K

Parameter	1 m diameter	1.35 m diameter	1.6 m diameter	Varied diameter	Plant data
C/O ratio	1.84	1.84	1.84	1.84	
Temperature of Gas T_g (K)	641.73	636.28	633.90	634.85	796
Temperature of Cat. T_s (K)	645.88	637.42	634.50	635.66	
Gas oil fraction	0.106	0.061	0.045	0.051	
Gasoline fraction	0.530	0.470	0.410	0.440	0.53
Gases	0.244	0.342	0.412	0.380	0.25
Coke	0.121	0.127	0.129	0.128	0.11

433

434 In the 1.35 m diameter riser, the gas oil conversion is 93.9% (0.061 wt %) producing 47.0%
435 (0.47 wt %) yield of gasoline, 34.20% (0.342 wt %) gases and 12.7% (0.127 wt %) coke. The
436 yield of gasoline deviates from the plant value (0.53 wt %) by -12.77%, yield of gases
437 deviates from the plant value (0.25 wt %) by 22.9% and coke yield deviates from the plant
438 value (0.11 wt %) by 13.38%. For the 1.6 m diameter riser, the gas oil conversion is 95.5%
439 (0.045 wt %) producing 41.0% (0.41 wt %) yield of gasoline, 41.0% (0.41 wt %) gases and
440 12.9% (0.129 wt %) coke. The yield of gasoline deviates from the plant value (0.53 wt %) by
441 -29.27%, yield of gases deviates from the plant value (0.25 wt %) by 39.32% and coke yield
442 deviates from the plant value (0.11 wt %) by 14.72%. In the varied diameter riser, the gas oil
443 conversion is 94.9% (0.051 wt %) producing 44.0% (0.44 wt %) yield of gasoline, 38.0%
444 (0.38 wt %) gases and 12.8% (0.128 wt %) coke. The yield of gasoline deviates from the
445 plant value (0.53 wt %) by -20.46%, yield of gases deviates from the plant value (0.25 wt %)
446 by 34.21% and coke yield deviates from the plant value (0.11 wt %) by 14.06%.

447 The gas phase exit temperature decreases with increase in diameter for risers of 1 m, 1.35 m
448 and 1.6 m diameters. Likewise, the yield of gasoline decreases with decrease in diameter for
449 the same risers, but yield of gases increases with increase in diameter. This is because as
450 diameter increases, the residence time for catalyst increases causing secondary reaction for
451 gasoline being converted into gases and coke, hence the decrease in the gas phase
452 temperature. The trend in these risers (1 m diameter, 1.35 m diameter and 1.6 m diameter)
453 correlates in a polynomial fashion with coefficient of determination, $R^2 = 1$ as follows:

454

$$455 \text{ Yield of gasoline} = -0.1143(\text{riser diameter})^2 + 0.0971(\text{riser diameter}) + 0.5471 \quad (27)$$

456 Gas phase temperature = $10.086(\text{riser diameter})^2 - 39.273(\text{riser diameter}) + 670.92$
457 (28)

458 Yield of coke = $-0.0152(\text{riser diameter})^2 + 0.053(\text{riser diameter}) + 0.0833$ (29)

459 Yield of gases = $0.028(\text{riser diameter}) - 0.036$ (30)

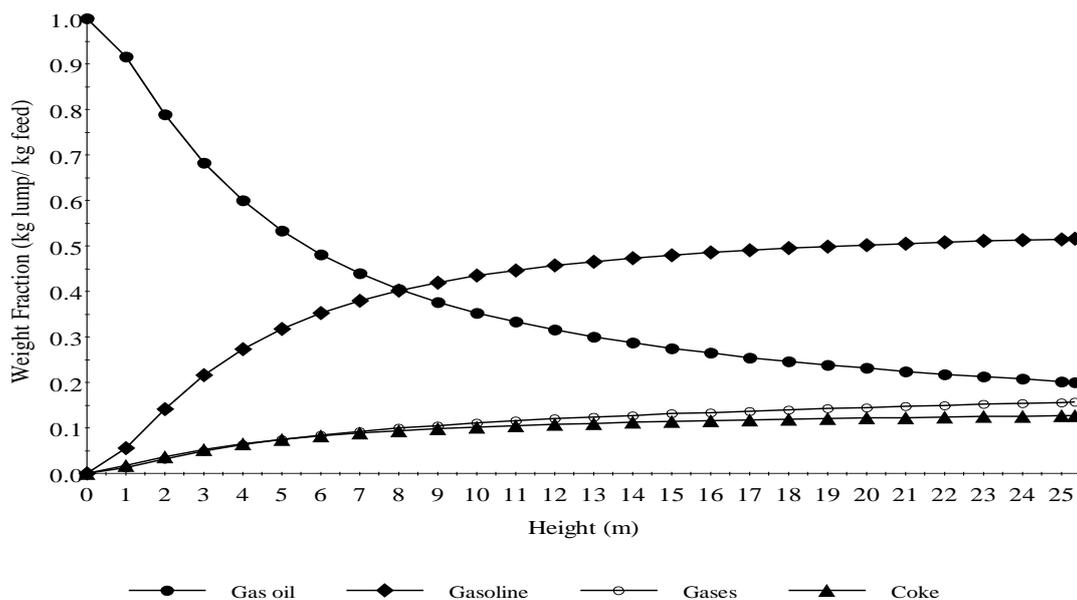
460

461 These Equations (27 – 30) are only viable for the conditions they were obtained, but can be
462 used for the typical range of riser diameters (0.61 m to 2.13 m) (Sadeghbeigi, 2012).

463 The results in Table 9 show that the varied diameter riser behaves like a reactor in-between
464 risers of diameter 1.35 m and 1.6 m. This gives rise to a gasoline yield higher than in the 1.6
465 m diameter riser but lower than in the 1.35 m diameter riser. In conclusion, the 1 m diameter
466 riser catalyst inlet temperature of 933 K has the best yields for gasoline and the lowest values
467 for gases and coke.

468 When gas oil comes in contact with the catalyst, it begins to crack to form cracked lumps;
469 gasoline, gases and coke. The profiles of the products of this gas oil cracking are presented in
470 Figure 5. The gas oil inlet temperature is 513 K, the C/O ratio is 1.27 and the inlet
471 temperature of catalyst is 933 K. Gas oil is cracked to produce three lumps; gasoline, gases
472 and coke.

473 The conversion of gas oil reaches 90 wt% at the exit of the riser and 70% of that conversion
474 is attained at 13.3 m of the riser. The coke concentration increases logarithmically from 0
475 wt% at the inlet to 13.0% at the exit of the riser. The gasoline increases logarithmically from
476 0 wt% at the inlet of the riser to its maximum yield of 51.7 wt% in the first 14 m of the riser
477 and then essentially levels out. At the exit of the riser, the yield is 50.0 wt%. The yield of the
478 gases increases logarithmically from 0 wt% at the inlet of the riser to a maximum of 15.8
479 wt% at the exit. Being an intermediate in a series reaction of consecutive reactions, the
480 gasoline is expected to rise to a maximum and then fall. It has reached its maximum because
481 of the consistent yield as seen in Figure 5, and about to fall if there is any secondary reaction.
482



483

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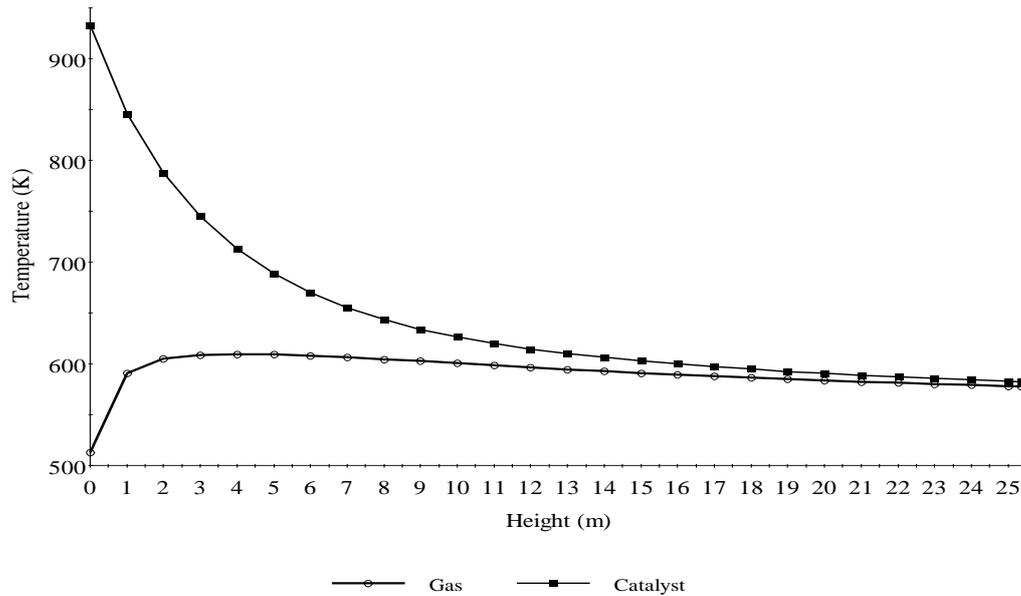
Figure 5: Weight fraction of components with 1 m diameter riser

485

486 The gasoline yield rises slightly throughout the riser and reaches to about 50 wt% at the exit
 487 of the riser which compares favorably with the value of 50 wt% obtained by Han and Chung
 488 (2001a) and 53 wt% in the plant as shown in Table 4.

489 The gases formed in this model logarithmically increased from 0 wt% at the inlet of the riser
 490 to 15.8 wt% at the exit of the riser as seen in Figure 3. This is expected, as the gases being a
 491 product of a multiple series-parallel reactions, should rise from a minimum to a maximum
 492 and then later levels out. The gases profile in this work compares well with that of Han and
 493 Chung (2001a). The coke composition also follows a similar logarithmic trend. However, in
 494 Figure 3, the coke is 12.7 wt% at the riser outlet for this model and it is much higher
 495 compared the plant value of 11.0 wt% shown in Table 5.

496 The temperature profile of both catalyst and gas phases presented in Figure 4 was obtained at
 497 the maximum C/O ratio of 2.4 which appears to have produced the lowest amount of coke
 498 deposited on catalyst at catalyst inlet temperature of 933 K and 513 K gas oil inlet
 499 temperature. The temperature of the catalyst-phase starts from about 933 K and decreases
 500 for the first 8 m and then essentially levels out. The temperature profiles of the gas phase
 501 starts from about 513 K and rises to a peak in the first 3 m of the riser and essentially levels
 502 out for the remaining portion of the riser. Both profiles approach the same value with
 503 temperature difference of about 1 K which is necessary for the completion of the reaction.
 504 The temperature profiles obtained in this work are similar to those obtained in many
 505 literatures (Han and Chung, 2001b, Souza et al., 2006, Ali et al., 1997).



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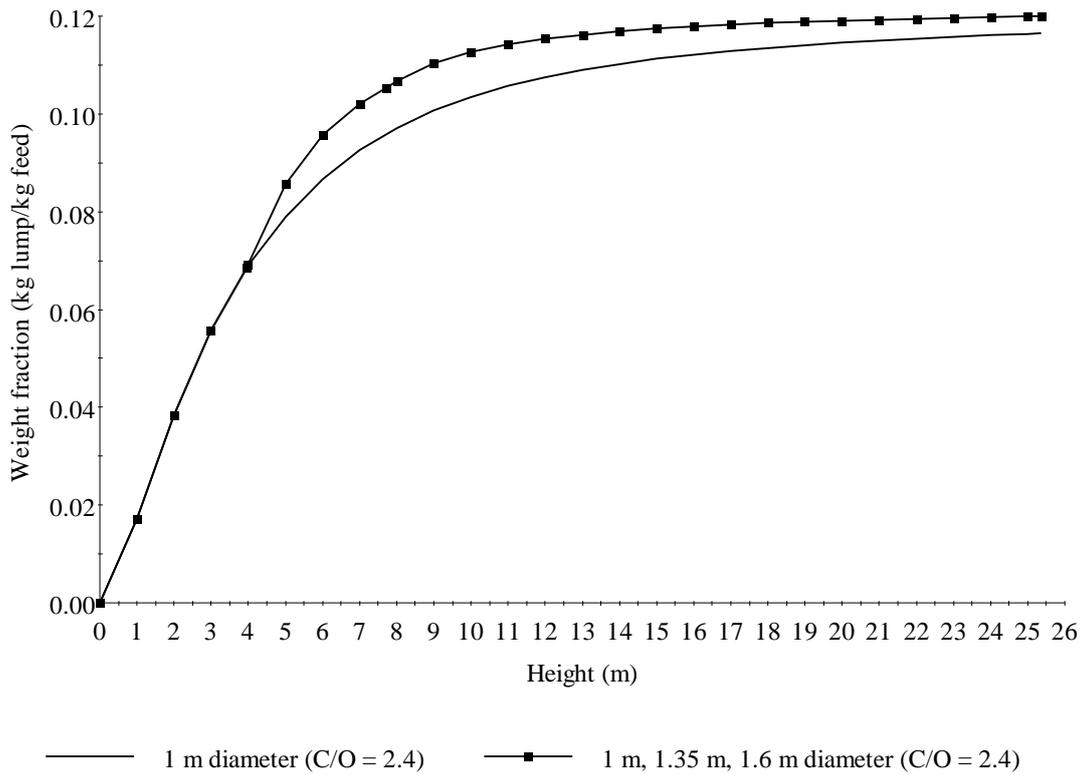
Figure 6. Temperature profile of 1 m diameter riser

508

509 The yield of coke as a lump is significant in FCC operation because of heat integration. The
 510 deactivated catalyst is regenerated by burning off the coke deposited on it and the resulting is
 511 used for the cracking of gas oil. Figure 7 compares the profiles of coke in two different risers
 512 (1 m diameter and varied diameter). The coke weight fraction profiles for both risers follows
 513 the same logarithmic trend from 0wt % at the riser entrance to 0.08 wt % at first 4 m of the
 514 riser height, then the profile for the 1 m diameter riser begin to levels out while the profile for
 515 the varied diameter riser continue to rise and eventually levels out. The exit concentrations of
 516 coke differ with more coke deposited on the catalyst for the varied diameter riser. This is
 517 possibly because of increased residence time of the catalyst in the varied diameter riser which
 518 increases the catalyst deactivation.

519

520



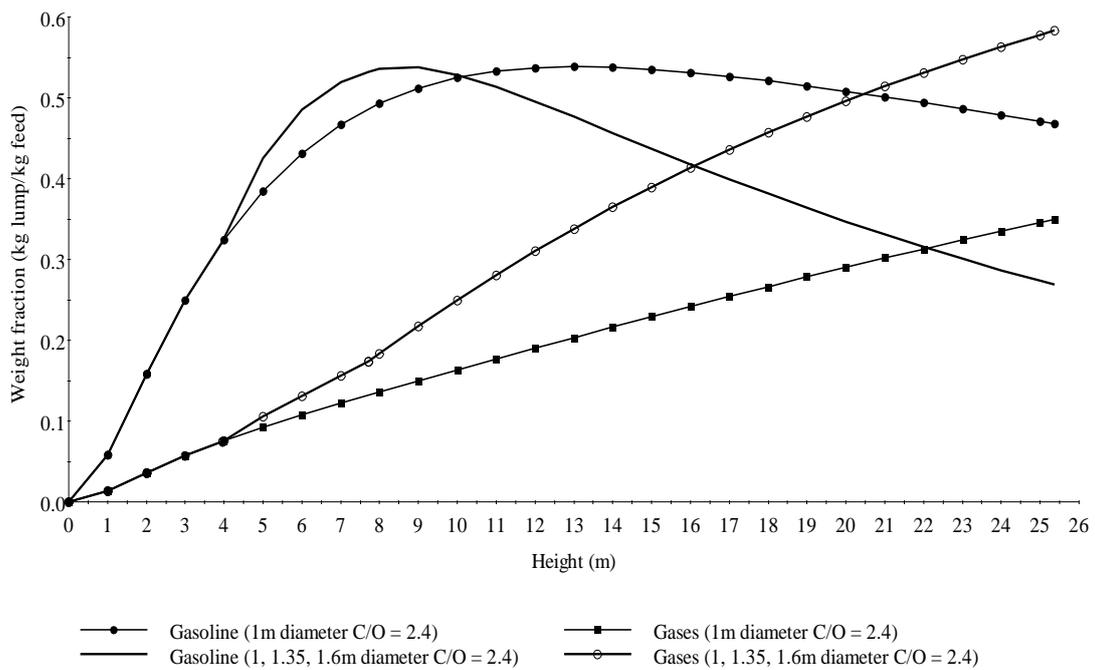
521

522

Figure 7. Weight fraction of coke in the riser at C/O ratio of 2.4 at 933 K.

523

524 Similarly, the profiles of gasoline and gases at C/O ratio of 2.4 and catalyst temperature of
 525 933 K for 1 m diameter and varied diameter risers are presented in Figure 8.



526

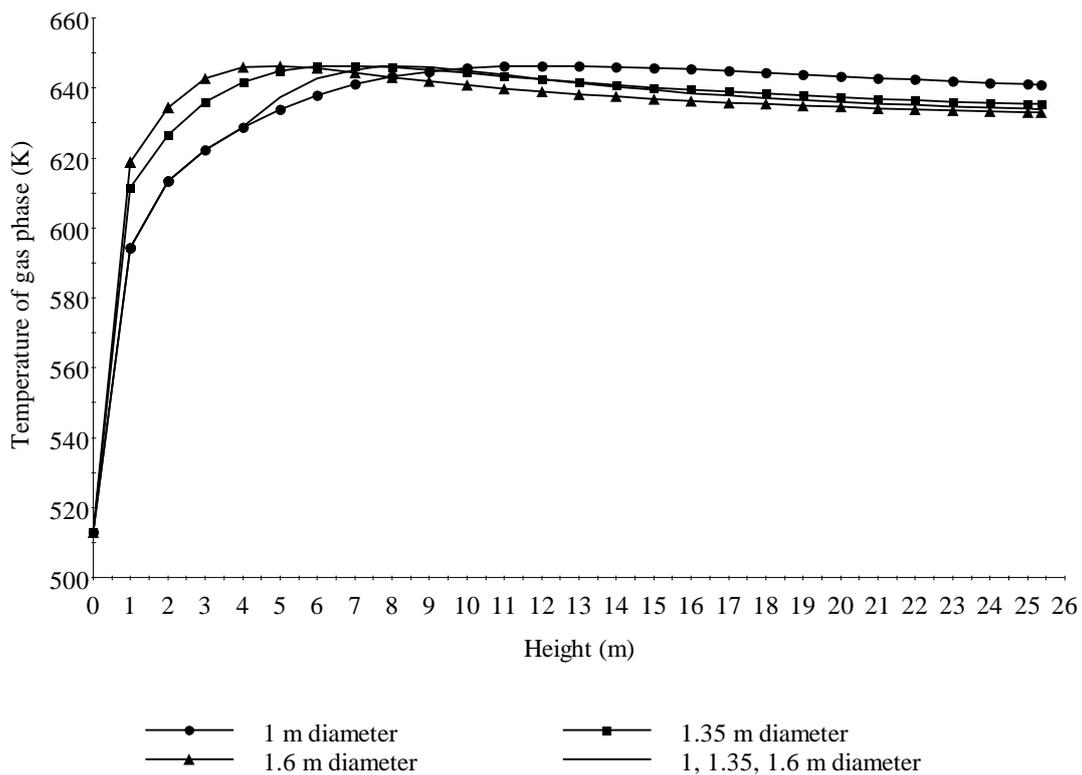
527

Figure 8. Weight fraction of Gasoline and Gases at C/O ratio = 2.4, $T_g = 933$ K.

528 The weight fraction of gasoline rise from 0 wt % at the entrance of the 1 m diameter and the
 529 varied diameter risers and then peaking to over 50% of the yields at 10 m height of the riser
 530 then level out at the exit of the riser. However, for the varied diameter riser, the outlet weight
 531 fraction of gasoline dropped drastically compared to that of the 1 m diameter riser. This is
 532 due to increased volume of the riser as the diameter increased and consequently the residence
 533 time for catalyst increased, causing a secondary conversion of gasoline to gases, which
 534 explain why there is more gas in the varied diameter riser. This shows a trend that the C/O
 535 ratio of 2.4 favours the 1 m diameter riser because of higher gasoline yield, though it has
 536 higher coke yield too.

537 Four different risers (1 m diameter, 1.35 m diameter, 1.6 m diameter and varied diameter)
 538 were simulated at C/O ratio of 1.8 and catalyst temperature 933 K. The gas phase temperature
 539 profiles of the four risers are shown in Figure 9.

540



541

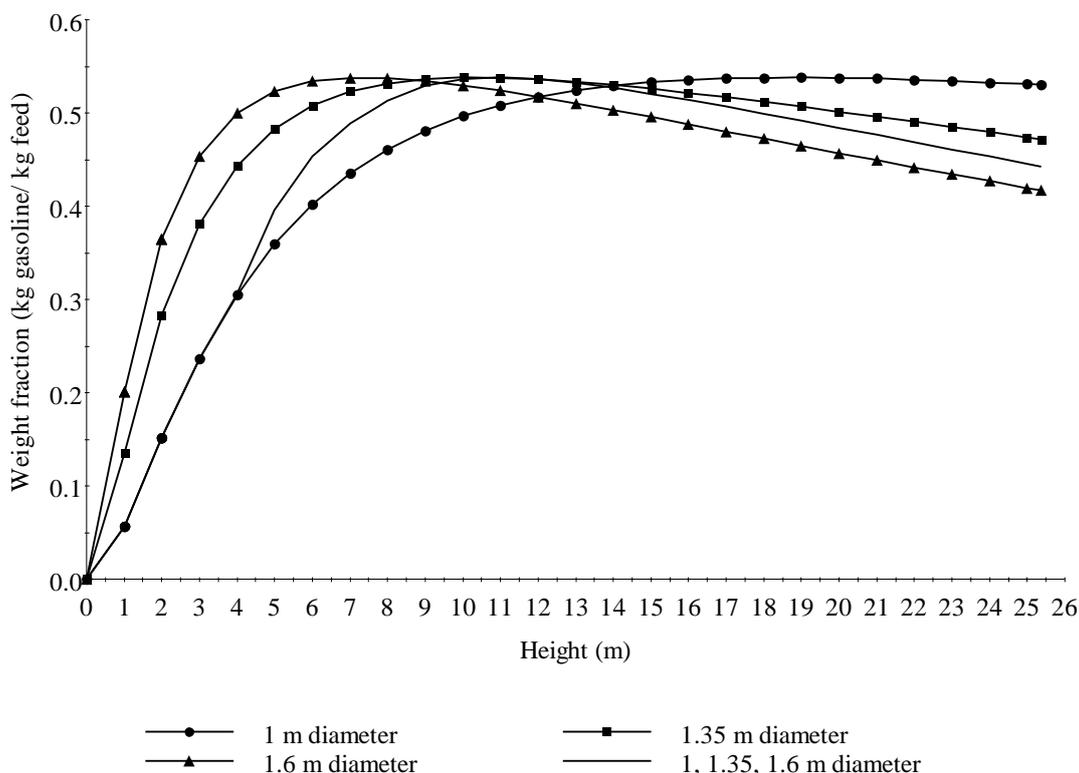
542 Figure 9. Temperature profiles of the four risers

543

544 The gas phase temperature profiles show great heat interactions at the inlet (first 1 m) of all
 545 the risers irrespective of the geometries. However, at the middle of the riser and towards the
 546 exit, the profile of the 1 m diameter riser shows higher temperature output due to less catalyst

547 residence time than the catalyst residence time in the varied diameter riser. This means that
 548 the energy interactions in the riser are greatly influenced by the riser geometry. The larger the
 549 diameter of the riser, the lower is the gas phase exit temperature.

550 Similarly, for the four different risers the profiles of the gasoline yields are presented in
 551 Figure 10. The 1 m diameter riser produces more gasoline than the other risers, with the
 552 varied diameter riser gasoline yield slotting in-between those of risers with diameter 1.35 m
 553 and 1.6 m. Also, the 1.6 m diameter riser produced the poorest gasoline yield at this
 554 condition.



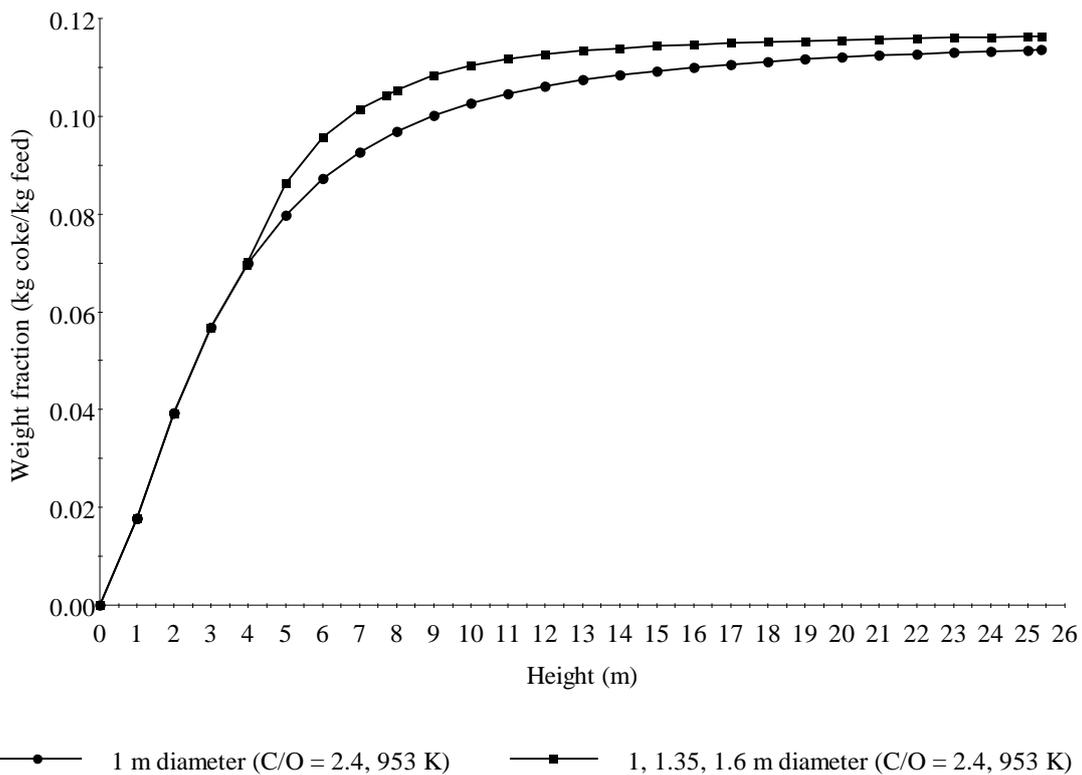
555

556 Figure 10. Gasoline yield (C/O ratio = 1.84, T_g = 513K, T_s = 933K)

557 The yield of gasoline at C/O ratio of 2.4 for 1 m diameter and varied diameter shown in
 558 Figure 8 and the yield of gasoline at C/O ratio of 1.84 for 1 m diameter and varied diameter
 559 shown in Figure 10 are qualitatively similar, however, quantitatively, the profiles of gasoline
 560 yields for both risers at C/O ratio 2.4 shows drastic decrease towards the exit of the risers.
 561 This is because increased C/O ratio means more catalyst is made available in the riser which
 562 favours secondary reaction of gasoline.

563 The temperature of catalyst was increased from 933 K to 953 K at C/O ratio of 2.4 and the
 564 profile of coke yield for 1 m diameter riser and varied diameter riser are presented in Figure

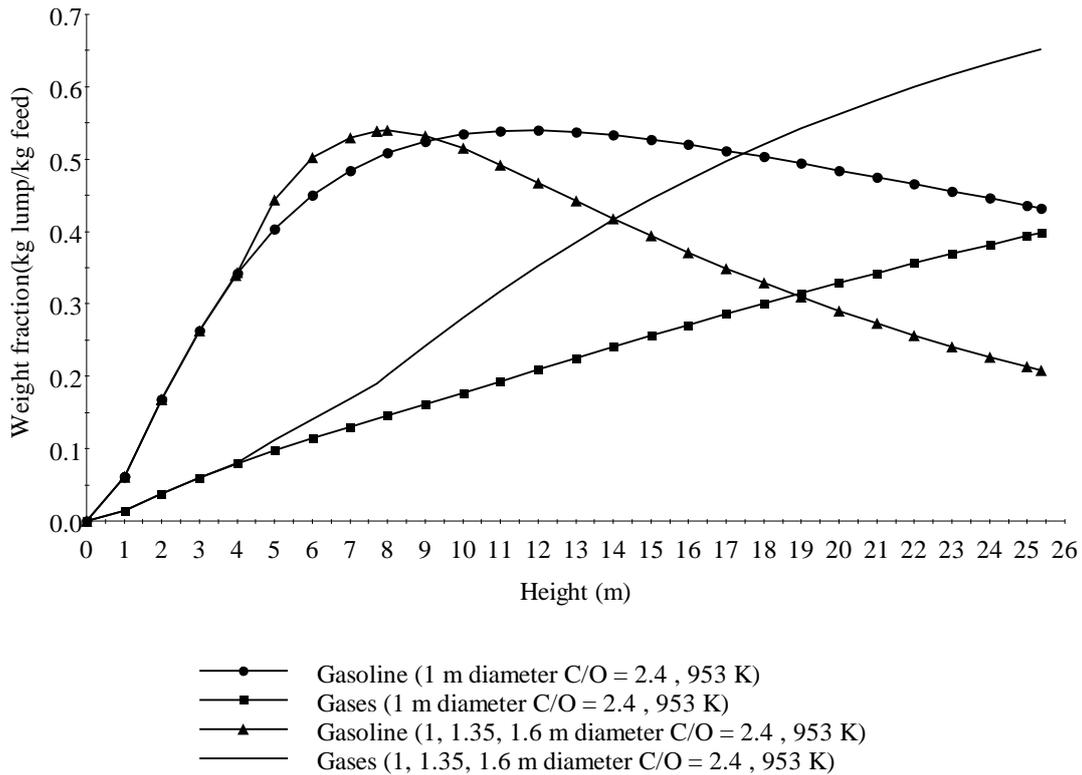
565 11. The coke deposited on catalyst for both risers at 953 K followed a similar qualitative
 566 trend as in coke deposited on catalyst for both risers at 933 K shown in Figure 7. The coke
 567 yield for the 1 m riser diameter is lower than in the varied diameter riser even when there is
 568 an increase of 20 °C on the catalyst temperature. In general, the higher the catalyst
 569 temperature the lower the amount of coke deposited on the catalyst.



570

571 Figure 11. Weight fraction of coke at catalyst temp (953 K).

572 Similarly, the profiles of the weight fractions of gasoline and gases for 1 m diameter riser and
 573 varied diameter riser are presented in Figure 12 for the same conditions as those in Figure 11.
 574 The yield of gasoline and gases for both risers at 953 K followed a similar qualitative trend as
 575 in the yield of gasoline and gases for both risers at 933 K shown in Figure 8. The coke yield
 576 for the 1 m riser diameter is lower than in the varied diameter riser even when the catalyst
 577 temperature increases by 20 °C. In general, the higher the catalyst temperature the higher the
 578 yield of gases, which is an undesired product, and lower the yield of gasoline. This happens
 579 because being an endothermic reaction and more heat is injected, most of the gasoline is
 580 converted to gases.



581

582

Figure 12. Weight fraction at 20 °C increase in catalyst temperature.

583

584 3. Conclusions

585 A varied diameter riser along with risers of uniform diameters was simulated and the
 586 following conclusions were made:

- 587 • The riser with varied diameter produces better yield of gasoline (53.4 wt %) at low
 588 catalyst to oil ratio with much closer values to plant data than that of 1 m diameter
 589 riser. At increased C/O ratio of 2.4, more gases are produced in the varied diameter
 590 riser which deviated well from the plant data and produced lower gasoline yield. This
 591 could be as a result of increased catalyst flux (Bollas et al., 2007) making more
 592 catalyst surface area available for further conversion of gasoline to gases and even
 593 coke.
- 594 • Increasing diameter of the riser, results in greater catalyst density and more heat for
 595 further cracking.
- 596 • The higher the C/O ratio and temperature, the lower the gasoline yield in all cases and
 597 the higher the yield of gases showing an inverse relationship between C/O ratio and
 598 temperature and between C/O ratio and riser diameter.

- 599 • With an increase of 20 °C catalyst temperature, gases yields increased more with
600 increasing C/O ratio, while gasoline yield is best at the lowest C/O ratio (1.27).
601 • Increased diameter of the risers results in increased yield of gases.
602 • The riser with varied diameter behaves like the combination of all risers of different
603 diameters. Its responses fall between the risers of diameters of 1.35 m and 1.6 m.
604 • With the relationship between C/O ratio, diameter and yields, refiners can easily
605 choose the plant yield at given diameter of riser or C/O ratio.
606 • Further work is required which will include detail hydrodynamics of the various units
607 of the FCC unit in relation to using a varied diameter riser and develop correlations
608 that are applicable to all FCC models.
609

610 **Notation**

A	Surface area, m ²
A_{ptc}	Effective interface heat transfer area per unit volume, m ² /m ³
C	Mole concentration, kg mole/m ³
C_{pg}	Gas heat capacity, kJ/kg K
C_{ps}	Solid heat capacity, kJ/kg K
D	Diameter, m
d_c	Catalyst average diameter, m
E	Activation energy, kJ/kg mole
F	Mass flow rate, kg/s
H	Specific enthalpy, kJ/kg
ΔH	Heat of reaction kJ/kg
h	Enthalpy of reaction kJ/kg
h_p	Interface heat transfer coefficient between the catalyst and gas phases
h_T	Interface heat transfer coefficient, kJ/m ² s K
k_{i0}	Frequency factor in the Arrhenius expression, 1/s
K_i	Rate coefficient of the four-lump cracking reaction, 1/s
K_g	Thermal conductivity of hydrocarbons
L	Length, m
M_w	Molecular weight
P	Pressure , kPa

Q_{react}	Rate of heat generation or heat removal by reaction, kJ/s
R	Ideal gas constant, 8.3143 kPa m ³ /-kg mole K or kJ/kg mole K
RAN	Aromatics-to-naphthenes ratio in liquid feedstock
S_c	Average sphericity of catalyst particles
S_g	Total mass interchange rate between the emulsion and bubble phases, 1/s
T	Temperature, K
u	superficial velocity, m/s
V	Volume, m ³
y	Weight fraction
Z_g	Gas compressibility factor

Greek

Ω	Cross-sectional area
ρ	Density, kg/m ³
\emptyset	Catalyst deactivation function
ε	Voidage
α	Catalyst deactivation coefficient
α_c^*	exponent for representing α
μ_g	viscosity

Subscript

cc	Coke on catalyst
ck	coke
g	Acceleration m/s ²
gl	gasoline
go	Gas oil
gs	gases
MABP	Molal average boiling temperature, K
MeABP	Mean average boiling temperature, K
pc	pseudo-critical
pr	pseudo-reduced
Rs	Riser

611 **Appendix A**

612 Table A.1 and Equations A1 – A24 are correlations of physical and transport parameters
 613 adopted from the literature (Han and Chung, 2001a, Han and Chung, 2001b).

614 Table A.1: Distillation Coefficients

Volume % distilled	a	b
10	0.5277	1.0900
30	0.7429	1.0425
50	0.8920	1.0176
70	0.8705	1.0226
90	0.9490	1.0110

615

616 Heat capacity of gas, C_{pg} , is

617
$$C_{pg} = \beta_1 + \beta_2 T_g + \beta_3 T_g^2 \quad (A.1)$$

618 Where β_1 , β_2 , β_3 and β_4 catalyst decay constant given as

619
$$\beta_1 = -1.492343 + 0.124432K_f + \beta_4 \left(1.23519 - \frac{1.04025}{S_g} \right)$$

620 (A.2)
$$\beta_2 = (-7.53624 \times 10^{-4}) \left[2.9247 - (1.5524 - 0.05543K_f)K_f + \beta_4 \left(6.0283 - \right. \right.$$

 621
$$\left. \left. \frac{5.0694}{S_g} \right) \right]$$

622 (A.3)

623
$$\beta_3 = (1.356523 \times 10^{-6})(1.6946 + 0.0884\beta_4) \quad (A.4)$$

624
$$\beta_4 = \left[\left(\frac{12.8}{K_f} - 1 \right) \left(1 - \frac{10}{K_f} \right) (S_g - 0.885)(S_g - 0.7)(10^4) \right]^2 \text{ For } 10 < K_f < 12.8 \quad (A.5)$$

625 Else $\beta_4 = 0$ for all other cases

626 K_f is the Watson characterization factor written as

627
$$K_f = \frac{(1.8T_{MeABP})^{\frac{1}{3}}}{S_g} \quad (A.6)$$

628 Where M_{wg} is the molecular weight of the gas and can be calculated using

$$M_{wg} = 42.965 \left[\exp(2.097 \times 10^{-4} T_{MeABP} - 7.787 S_g + 2.085 \times 10^{-3} T_{MeABP} S_g) \right] (T_{MeABP}^{1.26007} S_g^{4.98308}) \quad (A.7)$$

$$T_{MeABP} = T_{VABP} - 0.5556 \exp[-0.9440 - 0.0087(1.8 T_{VABP} - 491.67)^{0.6667} + 2.9972(SI)^{0.3333}] \quad (A.8)$$

Where T_{VABP} , the volume average boiling temperature and (SI) is slope given as

$$(SI) = 0.0125(T_{90ASTM} - T_{10ASTM}) \quad (A.9)$$

$$T_{VABP} = 0.2(T_{10ASTM} + T_{30ASTM} + T_{50ASTM} + T_{70ASTM} + T_{90ASTM}) \quad (A.10)$$

The ASTM D86 distillation temperatures are calculated using

$$T_{10ASTM} = a_{10}^{-\frac{1}{b_{10}}} (T_{10TBP})^{\frac{1}{b_{10}}} \quad (A.11)$$

$$T_{30ASTM} = a_{30}^{-\frac{1}{b_{30}}} (T_{30TBP})^{\frac{1}{b_{30}}} \quad (A.12)$$

$$T_{50ASTM} = a_{50}^{-\frac{1}{b_{50}}} (T_{50TBP})^{\frac{1}{b_{50}}} \quad (A.13)$$

$$T_{70ASTM} = a_{70}^{-\frac{1}{b_{70}}} (T_{70TBP})^{\frac{1}{b_{70}}} \quad (A.14)$$

$$T_{90ASTM} = a_{90}^{-\frac{1}{b_{90}}} (T_{90TBP})^{\frac{1}{b_{90}}} \quad (A.15)$$

Where a_i and b_i are distillation coefficients (Table A.1) and T_{iTBP} is the TBP distillation temperature.

Interface heat transfer coefficient between the catalyst and gas phases, h_p ,

$$h_p = 0.03 \frac{K_g}{d_c^{\frac{2}{3}}} \left[\frac{|(v_g - v_c)| \rho_g \epsilon_g}{\mu_g} \right]^{\frac{1}{3}} \quad (A.16)$$

Thermal conductivity of hydrocarbons

$$K_g = 1 \times 10^{-6} (1.9469 - 0.374 M_{wm} + 1.4815 \times 10^{-3} M_{wm}^2 + 0.1028 T_g) \quad (A.17)$$

M_{WM} is the mean molecular weight of the combined catalyst and gas

$$648 \quad M_{WM} = \frac{1}{\left(\frac{y_{go}}{M_{wgo}} + \frac{y_{gl}}{M_{wgl}} + \frac{y_{gs}}{M_{wgs}} + \frac{y_{ck}}{M_{ck}}\right)} \quad (A.18)$$

$$649 \quad M_{wgo} = M_{wg} \quad (A.19)$$

$$650 \quad M_{wgs} = 0.002M_{wH_2} + 0.057M_{wC_1} + 0.078M_{wC_2} + 0.297M_{wC_3} + 0.566M_{wC_4} \quad (A.20)$$

651 The viscosity of the gas

$$652 \quad \mu_g = 3.515 \times 10^{-8} \mu_{pr} \frac{\sqrt{M_{WM} P_{pc}^{\frac{2}{3}}}}{T_{pc}^{\frac{1}{6}}} \quad (A.21)$$

$$653 \quad \mu_{pr} = 0.435 \exp\left[\left(1.3316 - T_{pr}^{0.6921}\right)P_{pr}\right] T_{pr} + 0.0155 \quad (A.22)$$

$$654 \quad T_{pc} = 17.1419 \left[\exp\left(-9.3145 \times 10^{-4} T_{MeABP} - 0.5444 S_g + 6.4791 \times 10^{-4} T_{MeABP} S_g\right) \right. \\ \left. \times T_{MeAB}^{-0.4844} S_g^{4.0846} \right] \quad (A.23)$$

$$655 \quad P_{pc} = 4.6352 \times 10^6 \left[\exp\left(-8.505 \times 10^{-3} T_{MeABP} - 4.8014 S_g + 5.749 \times 10^{-3} T_{MeABP} S_g\right) \right. \\ \left. \times T_{MeAB}^{-0.4844} S_g^{4.0846} \right] \quad (A.24)$$

$$657 \quad T_{pr} = \frac{T_g}{T_{pc}} \quad (A.25)$$

$$658 \quad P_{pr} = \frac{P}{P_{pc}} \quad (A.26)$$

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