

1 **Improved syngas processing for enhanced Bio-SNG production: A**
2 **techno-economic assessment**

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11
12 **Abstract**

13 This study analyzes the modification of an existing process configuration train for
14 Bio-SNG production using an indirectly-heated circulating fluidized bed gasifier.
15 Taking the process design of the Gothenburg Biomass Gasification (GoBiGas)
16 project, we investigate four modifications to the process design in order to
17 analyze what the potential effect from implementation of the results from state-
18 of-the-art research activities on bio-syngas conversion is. Firstly, aromatic
19 compounds are converted into Bio-SNG. Secondly, olefin hydration and
20 hydrodesulfuration units are combined in a high-temperature hydrodesulfuration
21 unit. Thirdly, the methanation section is modified and the pre-reformer unit in the
22 syngas conditioning section is eliminated. Finally, H₂S and CO₂ removal are
23 combined in the same unit. In order to provide a comprehensive comparison of
24 current GoBiGas process and the configuration investigated in this work, process
25 flowcharts and energy and material balances are provided. The study reveals that
26 the investigated configuration has the potential to reduce capital investment and
27 operating costs. Considering Phase II of GoBiGas project, a potential reduction
28 of 29% of the capital investment and 7 €/MWh of produced Bio-SNG could be
29 achieved comparing planned and investigated configurations. The results prove
30 that investigated modifications can have a large impact in the future
31 commercialization of Bio-SNG.

32 **Keywords:** Bio-SNG; biomass; process optimization; techno-economic
33 assessment

34 1. Introduction

35 To substitute natural gas with gas from lignocellulosic biomass (Bio-SNG) is an
36 important contribution to the reduction of greenhouse gas (GHG) emissions,
37 providing that the biomass can be regarded as a renewable carbon source. Bio-
38 SNG can be used as a substitute of diesel provided car-engines are adopted for
39 natural gas as a fuel. The main strengths of Bio-SNG compared to other biofuels
40 (diesel-FT and ethanol) are its versatility (domestic, heating and transportation
41 fuel) and it can be directly injected to the European natural gas grid. Considering
42 the high prices of natural gas in Europe (30-40 €/MWh) [1], the substitution of
43 natural gas by Bio-SNG seems to be the most favored option.

44

45 The development of plants for Bio-SNG production is in progress mainly in
46 Europe and there are currently three active demonstration projects: the Milena-
47 SNG plant in The Netherlands, and the Gothenburg Biomass Gasification
48 (GoBiGas) project and the Biomass-to-Gas (Bio2G) project in Sweden. Figure 1
49 shows the main difference between the projects with respect to Bio-SNG
50 generation, prior to the gas upgrading steps. The Milena-SNG project has been
51 proposed by the Energy Research Center of The Netherlands (ECN) for the
52 production of 10-12 MW_{Bio-SNG}. The plant will use the Milena technology for
53 biomass gasification and the OLGA technology for gas cleaning [2]. The GoBiGas
54 project is a two-step project initiated by the local utility *Göteborg Energi AB* for
55 the production of Bio-SNG. The first phase (Phase I), which is in operation, is a
56 20 MW_{Bio-SNG} plant and a second phase (Phase II) of 100 MW_{Bio-SNG} is under
57 planning. The Phase I plant is located in Gothenburg (Sweden) and it uses a dual
58 bed indirectly-heated circulating fluidized bed (i-CFB) gasifier developed by
59 REPOTEC [3]. The Bio2G project is under planning by E.ON and it targets the
60 production of 200 MW_{Bio-SNG}. The plant will be located in Malmö (Sweden) and
61 will use a pressurized oxygen blown direct fluidized bed (d-FB) gasifier developed
62 by the Gas Technology Institute (GTI) connected to a tar-reformer. Both projects
63 use wood pellets (for start-up) and forest residues as biomass feedstock. A
64 previous project was coordinated by the German Biomass Research Centre
65 (DBFZ): a demonstration plant of 1 MW_{Bio-SNG} in Güssing (Austria) using
66 REPOTEC gasifier [4,5].

67

68

69 Figure 1. General overview of active demonstration projects for Bio-SNG
70 production (Europe) [2,3,6,7].

71

72 Regarding literature on Bio-SNG, Kopyscinski has reviewed SNG production
73 from coal and biomass gasification [8]. Van der Maijden et al. provide a
74 comparison of three different biomass gasification technologies (entrained flow,
75 d-FB and i-CFB) on Bio-SNG production [9]. Gassner and Maréchal, Heyne and
76 Harvey and ECN have recently published different techno-economic
77 assessments for Bio-SNG production [10-13]. However, these works cannot be
78 compared with any of the demonstration Bio-SNG projects from Figure 1. Hence,
79 some technical aspects of these projects on bio-syngas cleaning and conditioning
80 were not included.

81

82 The main bottlenecks for the commercialization of Bio-SNG are the operating
83 cost and capital investment since these are higher than today's fossil fuel
84 alternatives for which there is basically no capital cost. For the operating cost, the
85 biomass price contributes most and, thus, any increment of Bio-SNG production
86 (i.e. energy efficiency) is important in order to reduce the impact of biomass price.
87 The energy efficiency can be increased either by enhancing the conversion of
88 char in the gasifier or reducing the energy losses during syngas conditioning. The
89 capital investment can be reduced if some process units can be eliminated or if
90 their size and complexity can be reduced.

91

92 In this study, we compare the production of Bio-SNG from lignocellulosic biomass
93 taking the Phase II of the GoBiGas project and a modified configuration
94 investigated in this study as a basis for the design. The comparison is carried out
95 in terms of a techno-economic assessment. The investigated modifications have
96 the potential to enhance Bio-SNG production and reduce capital and operating
97 costs. The biomass pretreatment and gasification process are excluded from the

98 study, being the focus on the syngas treatment steps (cleaning and conditioning).
99 We describe the fundamentals of each modification and the implications in the
100 process design and economics.

101

102 **2. Methodology**

103 **2.1. Description of GoBiGas process**

104 Phase II of GoBiGas project considers the construction of a 100 MW_{Bio-SNG} plant.
105 Figure 2 shows an overview of the process, where both wood pellets and forest
106 residues can be used as feedstock. The gasification technology is by means of
107 an i-CFB gasifier operating at atmospheric pressure and at a temperature of 800-
108 900 °C. The produced gas is de-dusted and cooled and heavy tars removed in
109 an oil scrubber using rapeseed-oil methyl ester (RME). Light tar (aromatic
110 compounds) along with chlorine and nitrogen compounds are removed
111 downstream using an adsorption bed with activated carbon. The tar-free bio-
112 syngas is compressed up to approximately 35 bar. The bio-syngas enters the
113 olefin hydrogenator, where olefins are converted into alkenes, and the
114 hydrodesulfuration (HDS) reactor, where organic sulfur is converted into H₂S.
115 Downstream, H₂S is removed and the H₂/CO molar ratio adjusted in a water-gas-
116 shift (WGS) reactor to meet the requirements of the methanation section. The
117 pre-reformer (pre-methanator) unit converts all alkenes into methane, but also
118 produces a CO₂-rich gas. The CO₂ content of the cleaned bio-syngas is reduced
119 before methanation. The methanation is carried out using in-series adiabatic
120 reactors. Finally, the raw Bio-SNG meets natural gas standards after water
121 removal.

122

123

124 Figure 2. Planned Phase II GoBiGas process (100 MW_{Bio-SNG}) overview.

125

126 **2.2. Process modeling**

127 Material and energy balances for planned GoBiGas and investigated
128 configurations are obtained using Aspen Plus® process simulator (version 8.4).
129 Biomass pretreatment, gasification, de-dusting and oil scrubber sections are not
130 included in the study since they are not affected by the process modifications in
131 the investigated configuration. Both the investigated and the planned GoBiGas
132 configurations are designed to have the same Bio-SNG production (100 MW_{Bio-}
133 SNG). The Soave-Redlich-Kwong (SRK) thermodynamic model is applied for the
134 simulation of both configurations. Details of the modeling of process units for both
135 configurations are given in Annex A.

136

137 **2.3. Description of process modifications**

138 The modifications to the GoBiGas syngas cleaning and conditioning investigated
139 in this work have the aim to:

- 140 • Enhance the energy efficiency of the plant, reducing the amount of
141 biomass feedstock (reducing operating cost).
- 142 • Eliminate several operating units in the downstream processing of bio-
143 syngas (reducing capital investment).

144 This is carried out by four process modifications to the planned GoBiGas
145 configuration:

- 146 1. Replacement of the adsorption beds with activated carbon by a zeolite
147 bed. The modification allows light-tar (aromatics) to remain in the syngas
148 stream and be converted into Bio-SNG in the methanation section.
- 149 2. Combination of olefin hydrogenation and HDS units in a new high
150 temperature HDS unit. The modification allows the reduction of one
151 processing unit.
- 152 3. The pre-methanation unit is eliminated in order to avoid the formation of
153 CO₂ and, for methanation, the TREMP™ technology from Haldor Topsoe
154 A/S is used, where the excess heat from methanation is balanced with the
155 partial recycle of converted bio-syngas. The modification allows the
156 reduction of one processing unit.

157 4. Combined removal of CO₂ and H₂S using an amine-absorption process.
158 The modification allows the reduction of one processing unit. This
159 modification is possible since the pre-reformer unit is eliminated.

160

161 As indicated in section 2.2, biomass pretreatment, gasification, de-dusting and oil
162 scrubber are not modified in the investigated configuration. Hence, Figures 3 and
163 4 show the flowcharts of downstream processing of the bio-syngas after the
164 removal of heavy tars (oil scrubbing). Figure 3 shows the flowchart of the planned
165 GoBiGas configuration (Phase II) and Figure 4 the flowchart of the investigated
166 configuration incorporating the four modifications described above. Table 1 gives
167 the technical differences between the configurations. There are some process
168 units that have been eliminated in the configuration investigated in this work
169 (adsorption bed, olefin hydrogenation, H₂S removal and pre-reformer). There is
170 a new gas compressor for the recycle of bio-syngas in the first methanation
171 reactor.

172

173

174 Figure 3. Process flowchart for the planned GoBiGas configuration. It excludes
175 biomass pretreatment, gasification, de-dusting and oil scrubber.

176

177

178 Figure 4. Process flowchart for the configuration investigated in this study. It
179 excludes biomass pretreatment, gasification, de-dusting and oil scrubber.

180

181 Table 1. Technical aspects and comparison of process units for the planned
182 GoBiGas plant and the configuration investigated in this study. In italics the
183 units that have not been modeled in this study.

Unit	GoBiGas	Investigated in this work
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<i>Grinding</i>	<i>Only for forest residues</i>	<i>Unchanged</i>
<i>Drying</i>	<i>Only for forest residues</i>	<i>Unchanged</i>
<i>Gasification section</i>	<i>i-CFB gasifier operating with steam addition at atmospheric pressure</i>	<i>Unchanged</i>
<i>Filtering</i>	<i>Ceramic filter for de-dusting</i>	<i>Unchanged</i>
<i>Oil scrubber</i>	<i>Removal of heavy tars (C₆₊) using RME.</i>	<i>Unchanged</i>
Adsorption beds (activated carbon)	Removal of light tars (aromatics), Cl and N-compounds.	Not present in this configuration
Zeolite bed	Not present in this configuration	Removal of Cl and N-compounds
Compression of raw syngas	Multi-step (6) compression from atmospheric pressure to 35 bar	Unchanged
Olefin hydrogenation unit	Hydrogenation of olefins It operates at 450 °C	Not present in this configuration
HDS unit	Conversion of organic sulfur compounds into H ₂ S It operates at 350 °C	Conversion of organic sulfur compounds into H ₂ S and hydrogenation of olefins to alkenes It operates at 550 °C
H ₂ S and CO ₂ removal	It is done in two different units: The first unit removes H ₂ S before the WGS unit	A single unit for both H ₂ S and CO ₂ removal before methanation

	The second unit removes CO ₂ before methanation section	
ZnO guard-bed	Removal of H ₂ S and Cl to ppb level	Unchanged (only location)
WGS unit	Correction of H ₂ /CO molar ratio for methanation Unaffected by methane or alkenes presence	It is located prior to the zeolite-bed In order to operate under sulfur-atmosphere the catalyst should be a Mo-based catalyst [24]
Pre-reformer (Pre-methanation)	Outlet methane content is approximately 40%	Not present in this configuration
Methanation section	There are four adiabatic reactors in-series Recirculation of syngas is not necessary since the pre-reformer has increased the methane content in the stream	There are three adiabatic reactors There is a partial recycling after the first reactor to reduce the temperature rise in the first reactor to 700 °C

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2.3.1. Conversion of aromatics into Bio-SNG

186

Table 2 shows the composition of a raw bio-syngas after removal of heavy tar in an oil scrubber. About 18% of the energy in the raw syngas corresponds to olefins (mainly ethylene) and 10% to aromatic compounds (mainly benzene). The presence of olefins and aromatics has been avoided in the planned GoBiGas configuration. However, while olefins are converted into alkenes in the olefin hydration reactor, aromatics (light tar) are removed in the adsorption beds. This removal of aromatic compounds from the bio-syngas implies an important energy loss.

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195 Some authors have studied the impact of higher hydrocarbons in Bio-SNG
196 production. They concluded that whereas olefins are very reactive and will form
197 coke at the methanation reactor inlet, the presence of alkenes and aromatics has
198 no effect on the methanation unit [13-15]. Moreover, literature only recognizes
199 the effect of olefins on methanation catalysts [8,16]. However, the behavior of
200 aromatics compounds on methanation catalyst (at biomass-derive syngas levels)
201 should be confirmed for long-operating times.

202

203 In this study, we investigate to leave aromatic compounds in the syngas, since
204 as it is described above their removal represents an important energy loss and
205 they have no harmful effect on the methanation unit. In order to avoid the removal
206 of aromatics in the syngas cleaning, the adsorption beds with activated carbon
207 are replaced by a zeolite bed, which only removes chlorine and nitrogen
208 compounds.

209

210 Table 2. Molar composition (%) of the raw bio-syngas after the oil scrubber
211 (based on [17]).

H₂	24.6
CO	36.3
CO₂	15.2
H₂O	10 ppm
CH₄	13.6
C₂H₂	0.4
C₂H₄	4.6
C₂H₆	0.4
C₃H₆	0.2
C₃H₈	0.0
Benzene	1.0
Other light tar (aromatics)	0.1
H₂S	250 ppm

COS	10 ppm
Other S (thiophene)	500 ppm
HCl	10 ppm
HCN	10 ppm
NH₃	20 ppm
N₂	3.6

212

213

2.3.2. Combination of olefin hydration and HDS

214 The removal of sulfur-containing species is crucial in the synthesis of biofuels
 215 from bio-syngas. In Bio-SNG production, H₂S and organic sulfur deactivate
 216 methanation catalyst [18,19]. The HDS unit operates at 350-400 °C [20]. In the
 217 HDS unit, organic sulfur is converted into H₂S and organic Cl into HCl [20].
 218 Furthermore, the WGS reaction is at equilibrium [20]. The HDS catalysts are
 219 CoMo- or NiMo-based catalysts, which are hardly deactivated by coke formation
 220 [20,21]. In a conventional HDS unit processing natural gas, hydrogen must be
 221 provided at a large excess with respect to the sulfur (and Cl) species [20]. In the
 222 case of biomass, the hydrogen excess is already provided in the syngas.

223

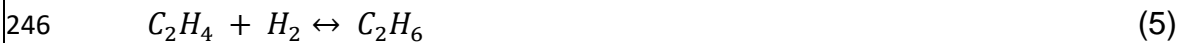
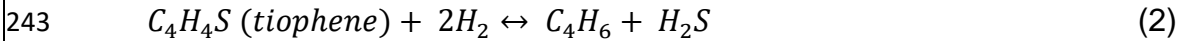
224 It is possible to hydrogenate olefins into alkenes in a HDS unit processing bio-
 225 syngas. The HDS catalyst is also active for olefins hydrogenation and the
 226 presence of sulfur species in the stream does not deactivate the hydrogenation
 227 of olefins [20,21]. The HDS unit can withstand aromatic compounds in the bio-
 228 syngas as demonstrated by Zuber et al. [13]. However, the operating temperature
 229 of the HDS unit should be 500-550 °C (higher than in conventional HDS units)
 230 [13,14]. Available information on the operation of HDS at high temperature is
 231 scarce, but important efforts are being made to have a complete characterization
 232 of HDS behavior at high temperatures [13,14,22]. The impact of high
 233 temperatures on catalyst performance at long operating times is unknown.

234

235 In this study, the HDS unit is operated at 550 °C (i.e. a high temperature HDS
 236 unit) allowing the hydrogenation of olefins in the bio-syngas. The olefin

237 hydrogenation unit is, therefore, eliminated. The high temperature HDS unit is
 238 modeled using the following set of reactions (all reactions at equilibrium). The
 239 HDS unit from the planned GoBiGas configuration is modeled using reactions 1-
 240 3.

241



248

249 Figure 5 gives the outlet composition (equilibrium) from the high temperature
 250 HDS unit considering bio-syngas composition from Table 1. The gray zone
 251 corresponds to the feasible operating window (500-550 °C). Since the required
 252 temperature is high, olefin conversion decreases with the operating temperature
 253 (approximately 95% conversion for 550 °C). However, in this study the higher
 254 value for temperature is selected in order to reduce the formation of COS.

255

256

257 Figure 5. Effect of temperature on the investigated high temperature HDS unit
 258 including olefin hydrogenation (outlet reactor conditions). Dotted lines refer to
 259 conversion and solid lines to molar concentration. Notes: C_{2+} content is
 260 approximately molar 20%. Operating pressure 35 bar.

261

262 **2.3.3. Modification of the methanation section and removal of the**
 263 **pre-reformer**

264 The pre-reformer or pre-methanation unit (both terms indistinctly used for Bio-
265 SNG production) is an adiabatic reactor for the conversion of higher
266 hydrocarbons (alkenes) into methane [16]. Along with alkenes, a fraction of H₂
267 and CO is converted into methane (equilibrium approach). The required steam-
268 to-carbon molar ratio required to prevent coke formation is around 3-4 [16]. As
269 commented in sections 2.3.1. and 2.3.2., olefins and sulfur species must be
270 avoided to prevent catalyst deactivation. The main drawback of using a pre-
271 reformer is the generation of a CO₂-rich stream, which needs to be conditioned
272 (CO₂ removal) before entering the methanation section. For the methanation
273 section, syngas composition must have a stoichiometric ratio (defined as $S =$
274 $\frac{H_2 - CO_2}{CO + CO_2}$, using molar fractions) of 3.00 [16].

275

276 In order to avoid the formation of CO₂, the pre-reformer unit is eliminated in the
277 configuration investigated in this work. Furthermore, we investigate the use of the
278 TREMP™ technology from Haldor Topsoe A/S, where the excess heat from
279 methanation is balanced with the partial recycling of converted bio-syngas [23].

280

281 **2.3.4. Combined CO₂ and H₂S removal**

282 The removal of the pre-reforming unit implies that sulfur removal can be placed
283 downstream (before the methanation section). This is possible since the WGS
284 unit is sulfur-tolerant [24]. Therefore, the removal of CO₂ and H₂S can be done in
285 the same unit. In this study, we investigate the combined removal of CO₂ and
286 H₂S using an amine-absorption process.

287

288 **2.4. Process economics**

289 Results from the process simulations for both planned and investigated
290 configurations are used to estimate capital and operating costs. Purchase
291 equipment cost (PEC) are taken from literature on techno-economic assessment
292 of thermochemical biorefineries. Firstly, the purchase costs are scaled and the
293 effect of inflation is corrected using the CEPCI (Chemical Engineering Plant Cost

294 Index) index. Then, the installed equipment cost is calculated by multiplying the
295 purchase cost by an installation factor. Thus, the total installed cost (TIC) is
296 calculated by adding up the cost of the individual equipment. The indirect costs
297 are estimated as percentages of TIC. The expected accuracy of this estimate is
298 $\pm 30\%$, typical of a study estimate [25]. Finally, the operating costs are calculated
299 according to Table B.2. Fixed operating costs are calculated as a percentage of
300 TIC while variable operating costs are calculated based on the cost of
301 consumables. Once the capital and operating costs are calculated, the minimum
302 Bio-SNG selling price is calculated as the Bio-SNG price which makes the net
303 present value of the project zero with a 10% internal rate of return (IRR). The
304 discounted cash flow analysis considers Bio-SNG as the single product of the
305 plant. The economic parameters used for the discounted cash flow analysis along
306 with the data for the capital and operating costs calculation are presented in
307 Annex B.

308

309 **3. Results**

310 **3.1. Energy and material balances**

311 Table 3 gives the resulting energy efficiencies and biomass feedstock for planned
312 GoBiGas and investigated configurations. This is due to the conversion of light
313 aromatics and the simplification of the process layout, as explained in section 2.
314 Whereas the planned GoBiGas configuration achieves a syngas-to-SNG energy
315 efficiency of 87.6%, the investigated configuration can achieve up to 96.1%. The
316 energy efficiencies are calculated using the higher heating value basis to SNG
317 production as described in [26]. In both configurations the value for the gasifier
318 energy efficiency is taken from literature [17]. The overall energy efficiency
319 (biomass-to-SNG) is, therefore, higher for the investigated configuration (71.1%).
320 The higher energy efficiency of the investigated configuration reduces the
321 required biomass feedstock to the plant in an 8.8% for a constant Bio-SNG
322 production. A remarkable difference is the amount of captured CO₂ available in
323 the plant. Since the pre-reformer unit is eliminated in the investigated
324 configuration, the captured CO₂ in the plant is also reduced (4.3%).

325

326 The heat demands and sources of both configurations are presented in Figures
 327 6 and 7 using the so-called composite curves (i.e using Pinch analysis). In these
 328 figures, biomass pretreatment, gasification, de-dusting and oil scrubber are not
 329 included since they remain equal in both configurations. The composite curves
 330 reveal the heat demand for the configuration investigated is 28% lower, in spite
 331 of its higher thermal level (550 °C vs. 250 °C). The heat demand for both
 332 configurations is assumed to be fulfilled by the heat release in the combustor
 333 gasifier, including the high-temperature demand of the HDS unit at 550 °C
 334 (approximately 4.5 MW_{th}). Considering the cooling demand, the investigated
 335 configuration achieves a reduction of 58% because of the reduction of processing
 336 units and corresponding intercooling.

337

338 Table 3. Energy efficiencies and fuel input for the planned GoBiGas and
 339 investigated configurations (production of 100 MW_{Bio-SNG}).

	Planned GoBiGas configuration	Investigated configuration
<i>Gasifier efficiency (% HHV) [17] (not modeled in this study)</i>	74.0	
Syngas-to-SNG efficiency (% HHV)	87.6	96.1
Biomass-to-SNG efficiency (% HHV)	64.8	71.1
Biomass input (dry tonne/day)	660	602
CO ₂ captured (t/h)	18.6	17.8

340

341

342 Figure 6. Composite curve for the planned GoBiGas configuration (blue: heat
 343 sources; orange: heat demands). Biomass pretreatment, gasification, de-
 344 dusting and oil scrubber are not included.

345

346

347 Figure 7. Composite curve for the configuration investigated in this work (blue:
 348 heat sources; orange: heat demands). Biomass pretreatment, gasification, de-
 349 dusting and oil scrubber are not included.

350

351 **3.2. Process economics**

352 Table 4 shows the reduction in the purchase equipment costs of the configuration
 353 investigated compared to the planned GoBiGas configuration (Phase II). There
 354 is a significant reduction thanks to the lower equipment-size and equipment
 355 reduction, except for the WGS reactor. The total capital investment of the
 356 configuration investigated is 29.1% lower than in the planned GoBiGas
 357 configuration for the production of 100 MW_{Bio-SNG}. Total capital investments are
 358 91.2 and 64.7 M€₂₀₁₀, respectively.

359

360 In order to analyze the impact of biomass feedstock on both configurations,
 361 Figure 8 presents the sensitivity of biomass price on the minimum Bio-SNG
 362 selling price. The minimum selling price Bio-SNG strongly depends on biomass
 363 price. However, the impact is slightly lower in case of the investigated
 364 configuration thanks to the reduction of required biomass feedstock for a constant
 365 Bio-SNG production. Considering the comparison between both configurations,
 366 the minimum selling price of Bio-SNG would drop 7 €/MWh for a price of 100 €
 367 per dry tonne of biomass for the investigated configuration.

368

369 Table 4. Comparison of purchase equipment cost (PEC) of the configuration
 370 investigated in this study.

Process unit	% PEC reduction compared to planned GoBiGas configuration
Grinding	8.9
Drying	6.3
Gasification section	6.5

Filtering	6.3
Oil scrubber	5.9
Adsorption beds	Eliminated
Compression of raw syngas	6.3
Olefin hydrogenator	Eliminated
HDS reactor	6.0
Zeolite bed	New equipment
H ₂ S removal	Eliminated
ZnO guard-bed	13.1
WGS reactor	-9.1
Pre-reformer	Eliminated
CO ₂ removal	9.6 (CO ₂ and H ₂ S removal)
Methanation section	49.8
Compression (recycling in methanation section)	New equipment

371

372

373 Figure 8. Sensitivity of Bio-SNG minimum selling price with biomass price for
374 planned GoBiGas and investigated configurations (100 MW_{Bio-SNG} assumed in
375 both cases).

376

377 4. Discussion and conclusions

378 This study demonstrates that there are potential benefits from incorporating
379 process modifications in the downstream processing of bio-syngas for the
380 production of Bio-SNG. Taking the planned 100 MW_{Bio-SNG} GoBiGas plant (Phase
381 II) as an example, investigated process modifications could improve energy
382 efficiency while reducing capital and operating costs. The conversion of light tars
383 (aromatics) into Bio-SNG largely increases the process efficiency in the
384 conversion of bio-syngas. Besides, the reduction of process equipment and the
385 combination of some units (olefin hydration/HDS and H₂S/CO₂ removal) allows

386 an important reduction of capital investment (29%). The operating cost, mainly
387 biomass feedstock, is reduced thanks to the higher energy efficiency of the
388 investigated configuration. However, the results presented are based on previous
389 lab or pilot scale investigations that need to be further demonstrated before being
390 applied to commercial Bio-SNG production. The positive results of this study
391 prove the large positive impact that these investigations can have in the future
392 commercialization of Bio-SNG.

393

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398

399

400 **Annex A. Modeling of process units and Bio-SNG specifications for both** 401 **configurations**

402 The process units have been modeled using Aspen Plus®. The name in brackets
403 is the corresponding name in the process simulator.

404

- 405 • Heat exchanger [Heater]: pressure drop 3 psi
- 406 • Pump [Pump]: $\eta = 90\%$
- 407 • Knock-out drum [Flash2]: adiabatic, no pressure drop
- 408 • Compressor: isentropic stage efficiency, $\eta = 72\%$
- 409 • Adsorption beds (activated carbon): 100% HCl recovery, 100% HCN
410 recovery, 10 ppm benzene slip
- 411 • Zeolite-bed [Sep]: 100% HCl recovery, 100% HCN recovery, 10% N₂ recovery
- 412 • Amines [Sep]: 95% CO₂ recovery or 95% H₂S recovery, 1% H₂ slip, 2% CO
413 slip, 2% CH₄ slip, 2% C₂ slip, 20% C₃₊ slip. Power consumption 4.5 MJ/kg

- 414 CO₂ or H₂S equivalent. There is no CO₂ compression or conditioning for
415 sequestration.
- 416 • Olefin hydrator [RStoic]: T= 450 °C (isothermal), pressure drop 5 psi, reactions
417 (3-6) at equilibrium
 - 418 • HDS reactor [RStoic]: pressure drop 5 psi, isothermal
 - 419 ○ Planned GoBiGas configuration (T=550 °C): Equilibrium of reactions
420 (1-6)
 - 421 ○ Investigated configuration (T= 350 °C): Equilibrium of reactions (1-3)
 - 422 • WGS reactor [RStoic]: T= 250 °C (isothermal), pressure drop 5 psi, reactions
423 (1,3) at equilibrium
 - 424 • Pre-reformer [RGibbs]: (adiabatic), pressure drop 5 psi, steam/C (able-to-be
425 reformed) ratio of 3, minimization of Gibbs free energy
 - 426 • Methanation [RGibbs]:
 - 427 ○ Planned GoBiGas configuration: 4 in-series adiabatic reactors with
428 inter-cooling. Minimization of Gibbs free energy.
 - 429 ○ Investigated configuration: 3 in-series adiabatic reactors with inter-
430 cooling. There is a partial recirculation of the outlet stream from the first
431 reactor (15.7%). Minimization of Gibbs free energy.
 - 432 • ZnO guard-bed [Ideal Sep]: 99.99% recovery of H₂S, 95% recovery of COS
433 and other organic sulfur. T = 45 °C
 - 434
 - 435 • Composition of the SNG (specifications) for both configurations:
 - 436 ○ T = 45 °C
 - 437 ○ CH₄ content (molar) = 99.46% (GoBiGas), 98.41% (investigated)
 - 438 ○ CO₂ content (molar) = 0.08% (GoBiGas), 1.16% (investigated)
 - 439 ○ H₂ content (molar) = 0.46% (GoBiGas), 0.42% (investigated)
 - 440 ○ Upper Wobbe index (MJ/Nm³)= 52.8 (GoBiGas), 51.6 (investigated)

441

442 **Annex B. Calculation of total capital investment, operating cost, and**
 443 **discount cash flow analysis**

444

445 The data for the calculation of the purchase and installation of equipment is
 446 shown in Table B.1. The TCI is calculated from the costs of installed equipment
 447 and indirect costs (21.9% of installed equipment) [27]. The data for the calculation
 448 of the operating cost is shown in Table B.2. The economic assumptions for the
 449 calculation of the IRR are shown in Table B.3.

450

451 Table B.1. Data for capital cost calculations of the most important equipment.^a

Unit	Base purchase cost	Reference year	Scale factor ^b	Units	Base scale	Installation factor ^c	Reference
Grinding	0.13 MUS\$	2003	1	dry tonne/day	2140	2.47	[27]
Drying	0.39 MUS\$	2008	0.70	dry tonne/day	1100	2.47	[27]
Gasification section	10.7 MUS€	2010	0.72	MW _{th} of biomass (LHV)	100	2.47	[11]
Filtering	2.7 MUS\$	2008	0.70	MW _{th} of biomass (HHV)	500	2.47	[27]
Oil scrubber	18.5 M€	2010	0.65	Nm ³ /h	135497	1	[27]
Water scrubber	3.4 MUS\$	2010	0.70	m ³ /s	9	2.47	[11]
Compression of raw syngas	5.85 MUS\$	2009	0.70	MW _e	5.44	1.32	[27]

Olefin hydrator	3.45 kUS\$	2007	0.67	t/h	65.77	2.47	[28]
HDS unit	3.45 kUS\$	2007	0.67	t/h	65.77	2.47	[28]
Zeolite bed	22.6 kUS\$	2002	1	Nm ³ /s	8	3	[27]
CO ₂ and H ₂ S removal	15.4 MUS\$	2001	0.65	lb/s of eq. CO ₂	27.83	2.47	[27]
ZnO guard-bed	22.6 kUS\$	2002	1	Nm ³ /s	8	3	[27]
Pre-reformer	41 MUS\$	2002	0.60	kmol/h reformed	1277	1	[27]
WGS unit	0.32 MUS\$	2002	0.56	lb/s	98.45	1	[29]
Methanation	43.8 kUS\$	2007	0.67	kg/s 1 st reactor	149.6 9	2.47	[28]

452 ^a A conversion rate of 1.35 US\$/€ has been used in this study.

453 ^b Scaling equation: $\text{Cost}/\text{Cost}_{\text{base}} = (\text{Scale}/\text{Scale}_{\text{base}})^n$.

454 ^c The installation factor is 1 if the base cost already includes the indirect costs.

455

456 Table B.2. Data for calculating operating costs (assuming 7500 operating hours
457 per year).

Fixed operating costs (% TIC)	
Labor	1.56
Maintenance	1.50
General expenses	3.07
Management and operation services	0.44
Cost of goods sold, marketing, logistics and others	1.32
Insurance	0.50
Total	8.39
Variable operating costs	

Biomass	50-150 €/dry tonne
Chemicals and absorbents (CO ₂ and H ₂ S removal)	2.4 €/t CO ₂ eq.
Waste water	0.57% variable costs
Gasifier bed material	0.81% variable costs
Diesel (oil scrubber)	5.00% variable costs
Boiler chemicals	0.14% variable costs
Water demineralization	2.43% variable costs
Pre-reformer and methanation catalyst	3.00% variable costs
Reactor catalyst (olefin hydration, HDS, WGS, pre-methanation and methanation)	1.5% variable costs
Ash disposal cost	0.36% variable costs

458

459 Table B.3. Economic assumptions for discounted cash flow analysis [27].

Parameter	Value
Debt/Equity	0/100%
Plant life	20 years
Depreciation (linear)	10 years
Salvage value	0 M€
Construction period	1 year
Income tax	30%
Working capital	1-month operating costs
Land	6% TCI

Working capital and cost of land are recovered at the end of plant life.

460

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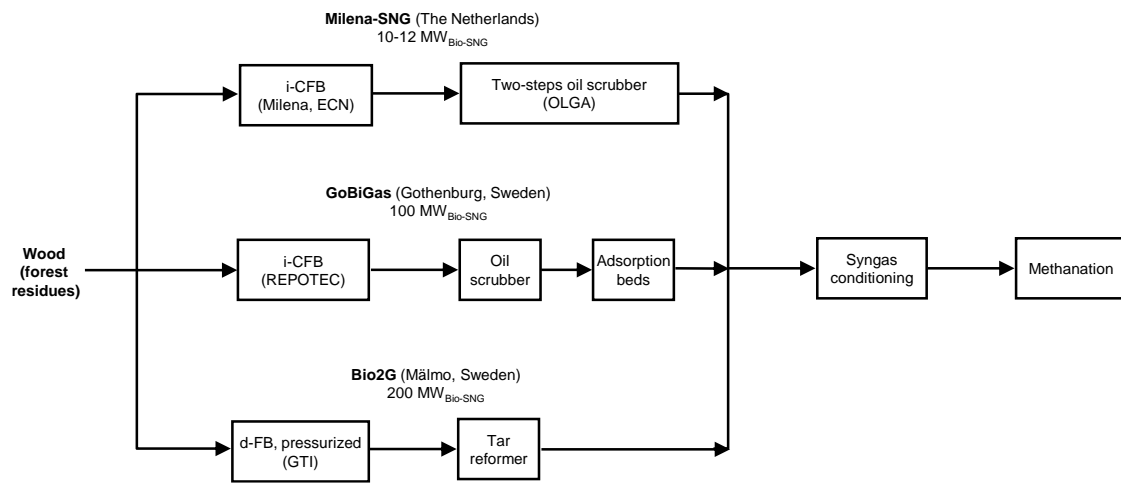
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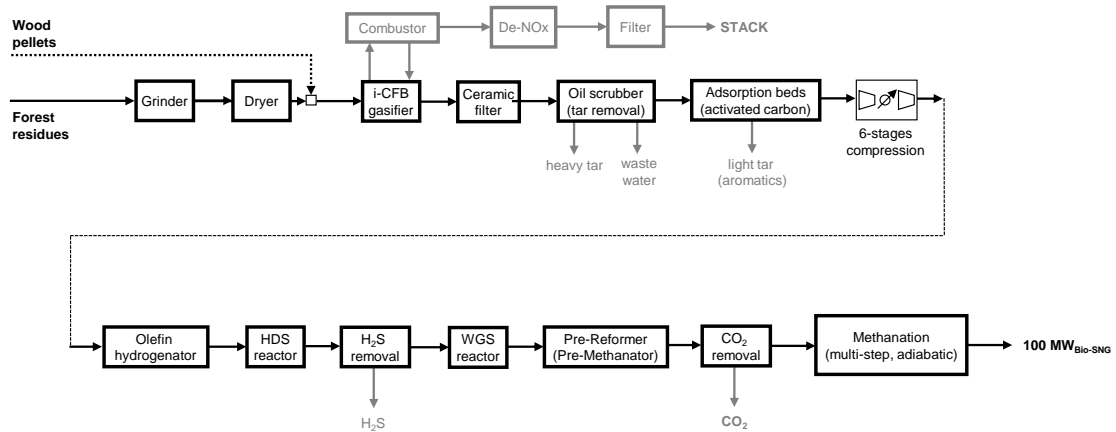
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568 Figure 2.

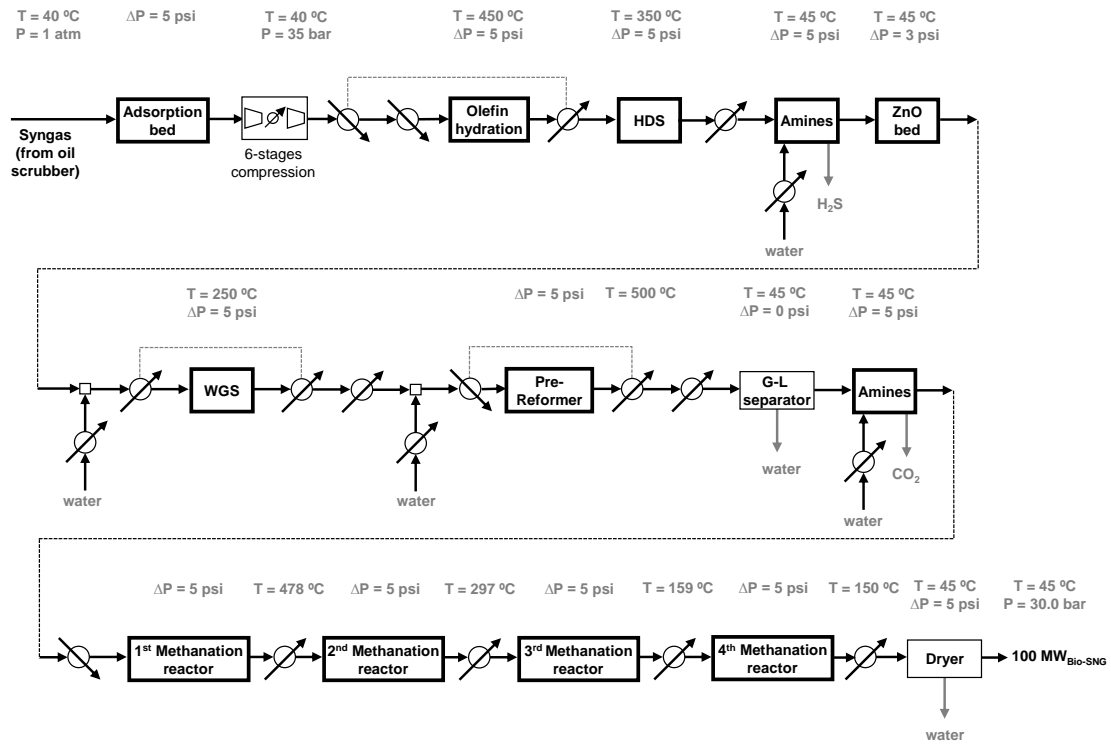


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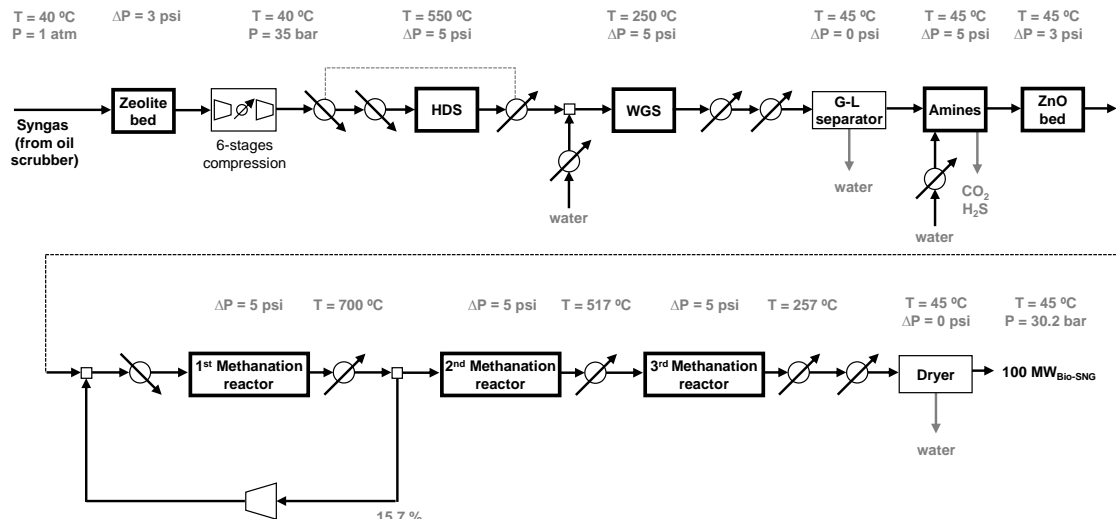
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575 Figure 4.

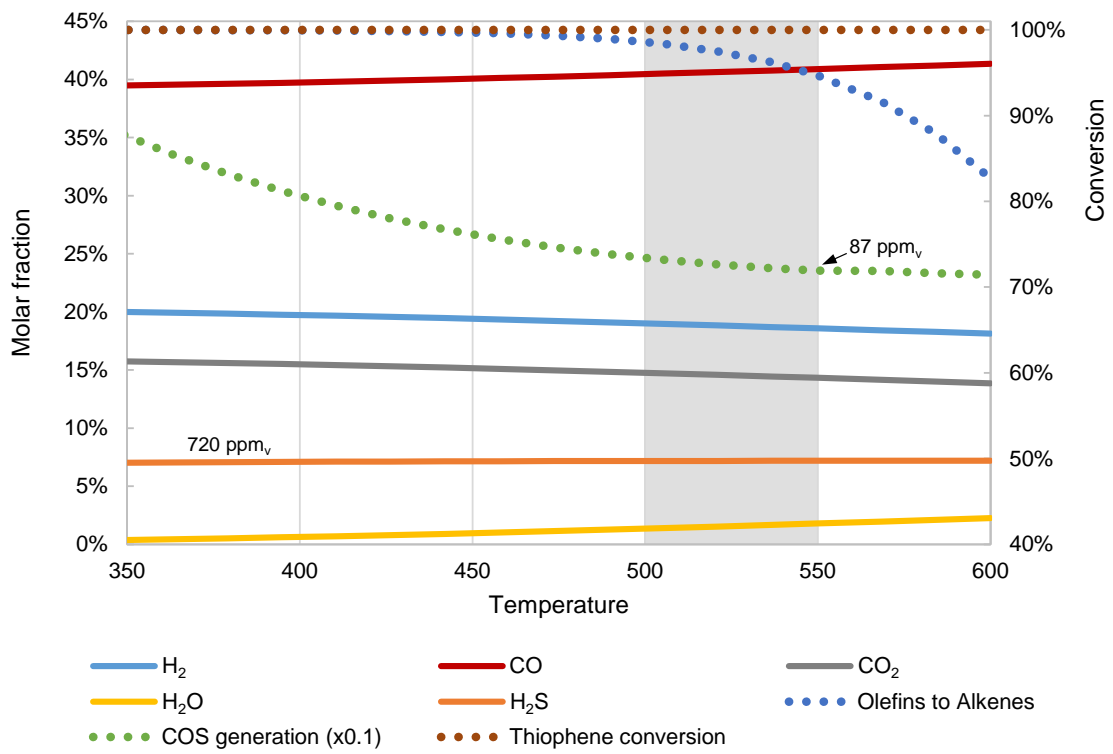


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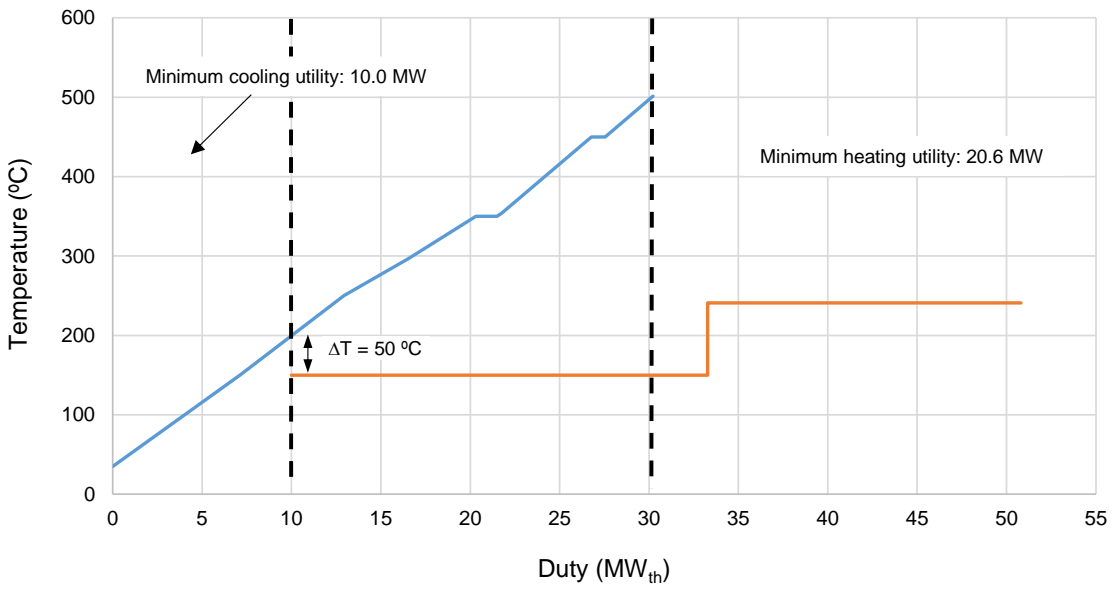
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582 Figure 6.

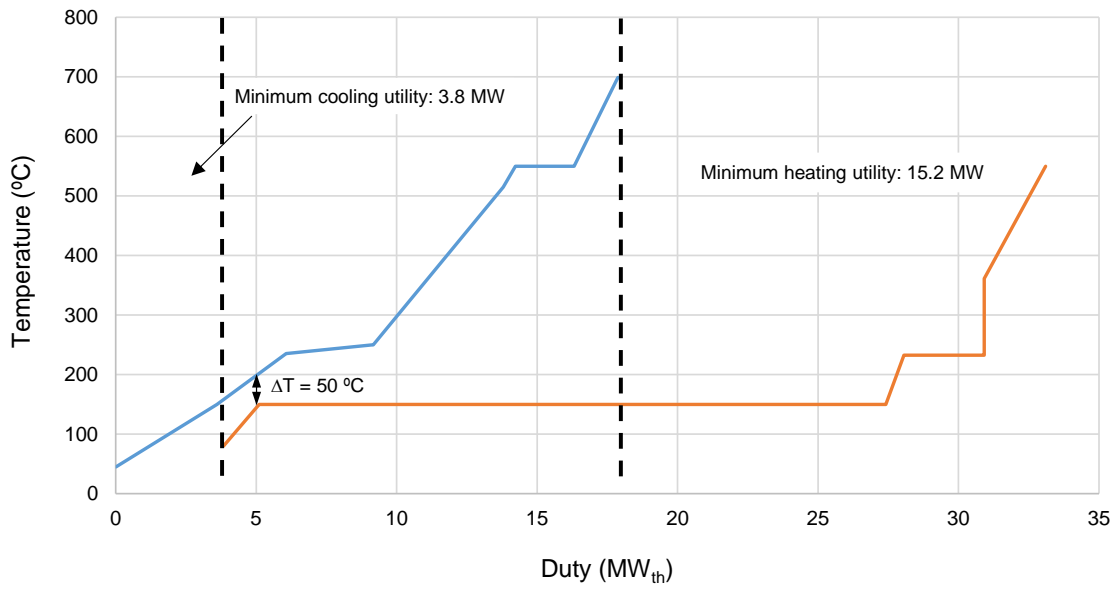


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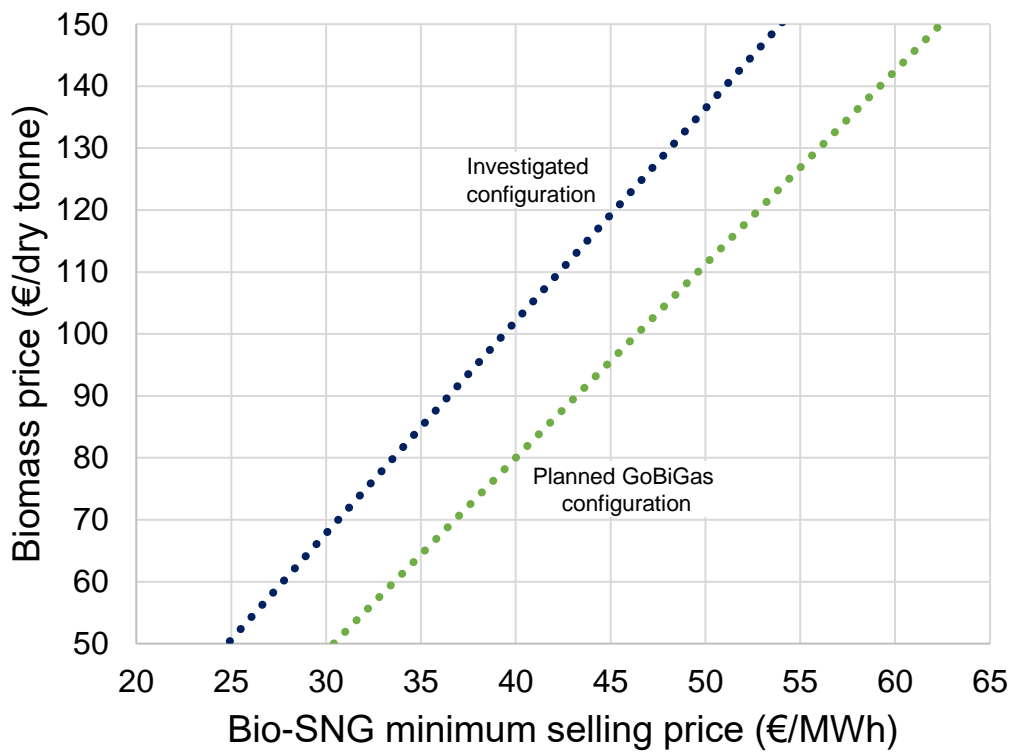
586 Figure 7.



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