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CO₂-mitigation options for the offshore oil and gas sector

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Abstract

The offshore extraction of oil and gas is an energy-intensive process leading to the production of CO₂ and methane, discharged into the atmosphere, and of chemicals, rejected into the sea. The taxation of these emissions, in Norway, has encouraged the development of more energy-efficient and environmental-friendly solutions, of which three are assessed in this paper: (i) the implementation of waste heat recovery, (ii) the installation of a CO₂-capture unit and (iii) the platform electrification. A North Sea platform is taken as case study, and these three options are modelled, analysed and compared, using thermodynamic, economic and environmental indicators. The results indicate the benefits of all these options, as the total CO₂-emissions can be reduced by more than 15 % in all cases, while the avoidance costs vary widely and are highly sensitive to the natural gas price and CO₂-tax.

1. Introduction

1 The extraction and processing of oil and gas on offshore petroleum fields was responsible for up to 26 %
2 of the total CO₂-emissions of Norway in 2011, and this share will likely stay significant in the coming years.
3 These emissions are mainly caused by the combustion of natural gas on-site for power generation purposes,
4 and they are subject to a hydrocarbon fuel tax, which has increased from 210 (\$35) [1] to 410 NOK (\$67) [2]
5 per ton of CO₂. A reduction of greenhouse gas emissions can therefore result in lower operating costs and
6 possibly higher energy savings, and this can be achieved by (i) decreasing the energy demand of the processes
7 implemented on the platform, (ii) increasing the efficiency of the power generation plant, (iii) implementing
8 a carbon capture unit, or by (iv) electrifying the facility.

9 (i) Regarding the first possibility, many studies assess the performance of oil and gas separation plants
10 by analysing thermodynamic inefficiencies and estimating the potentials for reducing the energy and
11 exergy losses.

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12 de Oliveira Jr. and Van Hombreeck [3] conducted the first exergy analysis of a Brazilian oil and gas
13 platform, and they showed that most inefficiencies are related to the petroleum heating and separation
14 steps. Voldsund et al. [4] assessed the performance of a Norwegian offshore facility and pinpointed
15 the gas compression process as the most exergy-destroying process. Nguyen et al. [5] conducted a
16 similar analysis on another plant, located in the Norwegian Continental Shelf, and the same conclu-
17 sions were drawn. These authors extended later their analysis to two other actual cases [6]. Their
18 findings highlighted the impacts of different field properties and export specifications on the power,
19 heating and cooling demands. The more recent work of Nguyen et al. [7] suggested that promoting
20 process integration on oil and gas facilities would be an effective measure for improving their energy
21 performance, and this was illustrated by comparing different facility configurations.

- 22 (ii) The second possibility has been discussed in several recent works, with the analysis of SRC (steam
23 Rankine cycle) and ORC (organic Rankine cycle).

24 Kloster et al. [8] argued that the implementation of steam cycles is technically feasible and would result
25 in a strong decrease of the CO₂-emissions, as proven with three case studies located in the North Sea [9].
26 In addition, Nord et al. [10] pinpointed many practical challenges, since there are stringent constraints
27 with regards to the maximum weight and space allowable on-site. Pierobon et al. [11] proposed a
28 methodology for designing ORCs for offshore applications. Barrera et al. [12] evaluated the influence
29 of different field conditions, such as the gas-to-oil ratio, on the design of an optimum ORC. These
30 studies lack a systematic approach: they investigate, for example, the implementation of a waste heat
31 recovery cycle offshore and present an optimisation of these systems with regards to weight, efficiency
32 and economic aspects. However, they do not take into account system integration issues – very little
33 attention is given on the interactions between the power and processing plants, and on the ways to
34 actually improve the performance of the overall system. This may result in sub-optimum solutions, as
35 illustrated with the case of SRCs [13].

- 36 (iii) The third possibility, which is to integrate a CCS (carbon capture and storage) process has been only
37 discussed in a couple of works.

38 For example, the study of Sánchez and de Oliveira [14] proved that the integration of CO₂-capture,
39 offshore, with monoethanolamine, is technically feasible, at the expense of a significant energy penalty.

- 40 (iv) Last but not least, the fourth possibility consists of electrifying the platform.

41 This can be achieved by connecting it to the onshore electric grid [15] or to offshore wind power
42 facilities [16]. Offshore electrification is claimed to present several technical and operational benefits,
43 such as higher facility availability, lower maintenance costs, and higher system efficiency, at the expense

44 of high investment costs. At present, a few offshore platforms located in the North and Norwegian
45 Seas are electrified, and new projects on the electrification of other ones are ongoing.

46 At the knowledge of the authors, all these studies focus on a single mean to reduce CO₂-emissions
47 (i.e. waste heat recovery, CO₂-capture, electrification) in the oil and gas offshore sector, but none actu-
48 ally compares them, and none evaluates each option considering process, thermodynamic, economic and
49 environmental aspects.

50 The present work aims at addressing these gaps. The novelty of the present work is to compare all
51 the above-mentioned means of reducing CO₂-emissions, by modelling at first each option, then evaluating
52 the associated costs and environmental impacts, and finally investigating which process parameters and
53 economic factors play an importance. The objectives are therefore to (i) evaluate the prospects for inte-
54 grating CO₂-mitigation processes (waste heat recovery, CO₂-capture, and electrification) on an existing oil
55 and gas platform, (ii) assess their thermo-environmental (i.e. thermodynamic, economic and environmental)
56 performance, (iii) and identify the possible trade-offs, by performing multi-objective optimisations.

57 2. Methodology

58 The methodology followed in this work has been extensively described in Bolliger et al. [17]: it builds
59 on a combination of flowsheeting techniques, energy integration models, economic evaluations using the
60 correlations of Turton et al. [18] and a multi-objective optimisation routine [19], based on a genetic algorithm
61 (Figure 1). The aims of the present methodology are to:

- 62 • extract, analyse and optimise simultaneously various process designs considering process, energetic,
63 economic and environmental indicators;
- 64 • develop flow-sheets that represent appropriately the physical and chemical processes taking place in
65 oil and gas processes, as well as in CO₂-capture systems;
- 66 • pinpoint possible systems improvements by process and energy integration;
- 67 • perform consistent economic evaluations based on cost correlations;
- 68 • assess the environmental impacts of integrating CO₂-capture processes in offshore conditions;
- 69 • identify the most optimal configurations by multi-objective optimisations and comparing systematically
70 CO₂-mitigation processes.

71 This framework may be divided into three steps: modelling, analysis and optimisation. The models are
72 of two types: mathematical models, which describe the physical behaviour of the system under interest, and
73 thermo-economic models, which evaluate the thermodynamic, economic and environmental performance of

74 that system. The mathematical models are built on data collection provided by the project partners and
 75 on information from the literature. They are embedded in a system superstructure in a way to include all
 76 possible options and connections between different sub-systems.

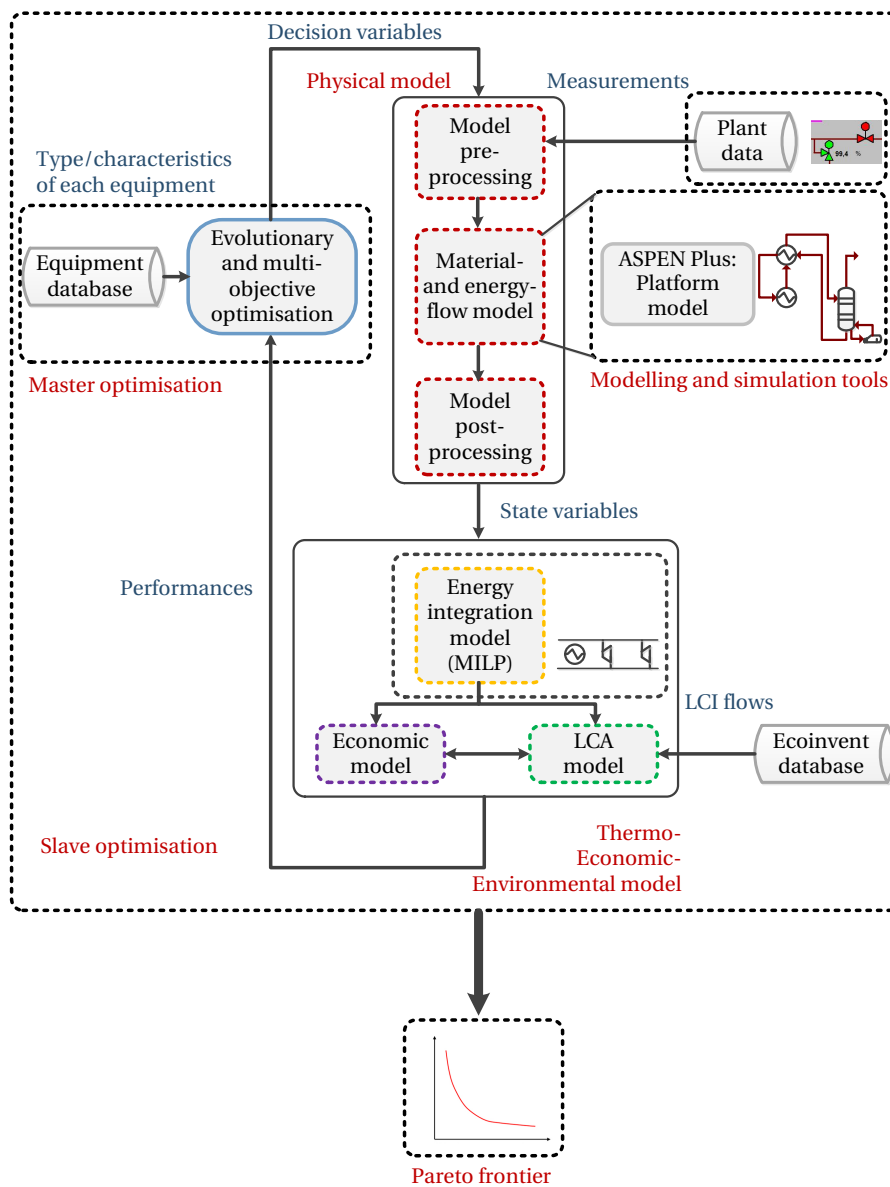


Figure 1: Illustration of the methodology and computational framework used for the modelling, simulation and optimisation of offshore platforms with CO₂-mitigation.

77 Each mathematical model follows a calling sequence in three phases: pre-processing, simulation and
78 post-processing. The pre-processing phase selects the process model of interest, collects and transfers the
79 decision variables. The simulation phase runs the mathematical model, computing the material and energy
80 flows using a flowsheeting software such as Aspen Plus [20]. In the case of an offshore platform, the main
81 thermodynamic relations describing the energy and exergy balances are presented in Appendix A. The
82 post-processing phase extracts the information of interest from the process models, such as temperatures,
83 pressures and state variables, and sends it to the thermo-economic models. The information transfer from
84 one phase or model to another is handled by using a Matlab-based platform named OSMOSE.

85 The thermodynamic performance of the system of interest can now be evaluated in the thermoeconomic-
86 environmental analysis models, since all the material flows and energy requirements have been computed.
87 Energy, Pinch, Total Site and exergy analyses are performed. The reader is referred to the textbooks and
88 articles of Linnhoff [21] for an overview of the basics of the Pinch analysis technique, of Smith [22] for an
89 overview of the Total Site Analysis method, and of Kotas [23] for the energy and exergy assessment methods.

90 The thermoeconomicenvironmental analysis models include an energy-integration model, which builds
91 on the use of the Pinch analysis technique [24]. The aim is to minimise the external energy demands and
92 associated operating costs, and the model is formulated as a MILP (Mixed Integer Linear Programming)
93 problem, whose formulation is presented in Appendix B.

94 The economic performance is then evaluated by performing cost estimations using, among others, the
95 correlations of Turton et al. [18]. The economic model is discussed in Appendix C. The environmental
96 impacts are estimated by conducting a life cycle assessment, following the approach of Gerber et al. [25].
97 The impacts are normalised with regards to the FU (functional unit) of this system, which is here taken as
98 1 Sm^3 of oil equivalent exported to the shore. The LCI (life cycle inventory) flows, e.g. the elementary flows
99 associated with, for instance, the pollutant emissions during operation, are calculated and their impacts are
100 computed from the Ecoinvent[®] database [26].

101 In the next step, multi-objective optimisations are conducted to define the system configurations that,
102 for example, simultaneously minimise the economic costs or environmental impacts, while maximising the
103 internal heat recovery and the thermodynamic performance [27]. The optimisation problem is based on the
104 *decision variables* used in the pre-processing phase of the mathematical models, and on *performance indica-*
105 *tors* calculated in the thermoeconomicenvironmental analysis models. For each evaluation, the performance
106 of the simulated system is re-assessed, and an evolutionary algorithm is used to emulate the values of the
107 decision variables. The latter are then re-sent through the pre-processing phase, closing the evaluation loop.

108 This mathematical problem includes discrete and continuous variables, as well as linear and non-linear
109 relationships among them. It is therefore a MINLP (Mixed Integer Non-Linear Programming) problem,
110 decomposed in this work into two sub-problems, namely a *master* and a *slave* problem. As this optimisation
111 includes possibly conflicting objectives, the results do not consist of a single solution but as a set of Pareto-

112 optimal ones.

113 **3. System description**

114 *3.1. Oil and gas platforms*

115 Oil and gas platforms present similar structural designs, but they process petroleum with different
116 characteristics (e.g. viscosity) and operate on fields with different properties (e.g. temperature, pressure,
117 water- and gas-to-oil ratios). An oil and gas facility can be divided into two main sub-systems: a processing
118 plant, where oil, gas and water are separated and treated, before being exported to the shore or rejected
119 to the environment; and an utility plant, where the power and heat required for driving the separation and
120 compression processes are produced.

121 The platform taken as case study exploits an oil field in the Norwegian Continental Shelf region, ap-
122 proaching its *end-life*. In the last decade, the oil and gas production has declined, while the water extraction
123 has increased, resulting in a feed water cut over 85–90 mol% for most wells. The process flow diagram,
124 together with additional data, is given in [Appendix D](#). The facility is characterised by a power demand of
125 about 19 MW, a heating demand smaller than 5 MW and a cooling demand greater than 30 MW, which is
126 met by processing more than 2000 m³/h of seawater at about 8 °C and lifted on-site. The Pinch point of the
127 overall facility is located at about 150 °C, illustrating that most heat discharged to the environment is at
128 low (under 100 °C) temperatures. The heating demand takes place between 150 and 230 °C and is satisfied
129 by recovering heat from the exhaust gases at about 330 °C. The thermal efficiency of the gas turbines varies
130 between 23 % (current operating point) to 34 % (nominal design point), and the energy intensity, on the
131 higher heating value basis, amounts to 4.6 %. The total daily CO₂-emissions produced locally reach about
132 450–460 tons, of which more than 90 % are correspond to the operation of the gas turbines.

133 *3.2. CO₂-mitigation*

134 *3.2.1. Superstructure*

135 The different technological options that can be implemented to reduce the CO₂-emissions of the oil
136 and gas plant (waste heat recovery and CO₂-capture) are investigated and included in a general system
137 superstructure (Figure 2). The integration of CO₂-capture on oil and gas platforms is not common, as it
138 faces several technical and economical challenges. CO₂-separation with acid gas (i.e. CO₂ and H₂S) removal
139 is a mature technology and is widely applied in hydrocarbon processing industries such as refineries. It is also
140 a common process for purifying hydrogen after steam reforming processes, such as in integrated gasification
141 combined cycle plants. Most applications are nevertheless onshore, and CO₂-separation offshore is limited
142 to natural gas platforms if the initial CO₂-content is considered too high for export.

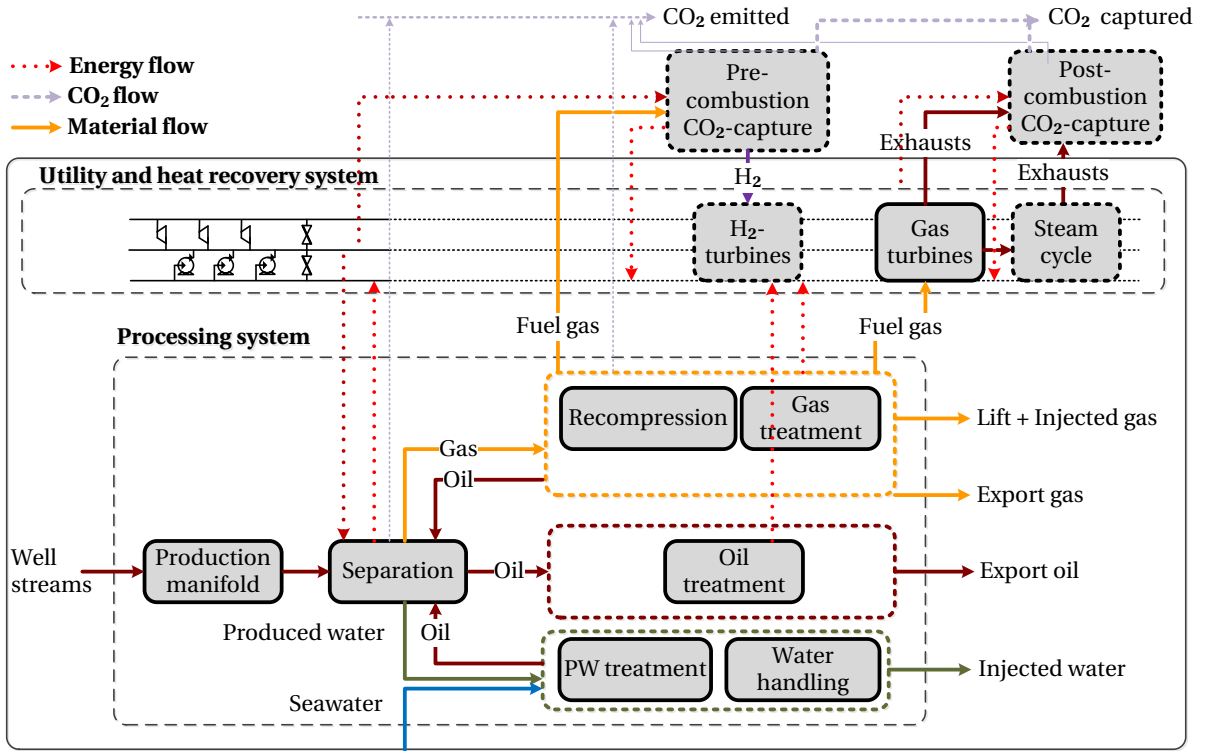


Figure 2: A generalised superstructure for integration of CO₂-capture on offshore platforms.

143 3.2.2. Waste heat recovery

144 Heat-to-power technologies are necessary to convert waste heat into power, of which the steam Rankine
 145 cycles are the predominant ones. Such technologies are mature for onshore applications, but the integration
 146 of these technologies offshore was performed only on three facilities [8] and has faced several practical
 147 challenges [28]. Organic Rankine and Kalina cycles are not considered in this study, as they have never
 148 been demonstrated in offshore applications and/or are not widely implemented. Different scenarios for the
 149 integration of steam networks can be considered, which differ (i) by the selection of the cooling utility (water
 150 and air), (ii) by the recovery of waste heat from either one or two gas turbines, (iii) the possible use of an
 151 intermediary heating loop. The details are presented in a previous work of the authors [13].

152 3.2.3. CO₂-capture

153 There exists a large variety of CO₂-separation technologies (absorption, adsorption, cryogenic distillation,
 154 membrane separation), and the present work focuses on absorption technologies. Acid gases such as CO₂
 155 are bound to an organic solvent, either chemically (chemical absorption), based on the CO₂-dissociation into
 156 hydrogen carbonates (HCO₃⁻), or physically (physical absorption), based on the solubility differences of CO₂
 157 in the feed gas and the liquid solvent [29]. Chemical absorption with alkanolamines such as monoethanolamine

158 (MEA) is preferred for treatment of acid gases with low CO₂ partial pressures [30]. Chemical absorption with
159 alkanolamines such as triethanolamine (TEA), or physical absorption with methanol (MeOH - Rectisol[®])
160 and mixtures of polyethylene glycol esters (DEPG - Selexol[®]) are preferred for high CO₂ partial pressures
161 (≥ 7 bar) [29]. The CO₂-separation process takes place in two main columns: an absorption column, in
162 which the solvent circulates at counter-current of the feed gas, removing CO₂, and a regeneration one, in
163 which CO₂ is recovered at high purity, generally by heating up or depressurising the solvent-CO₂-mixture.

164 There exist different pathways of CO₂-capture [31] based on these separation technologies:

- 165 • pre-combustion;

166 The term *pre-combustion* CO₂-capture refers to the CO₂-removal from carbonaceous fuels before com-
167 bustion, by, for instance, converting the primary fuel into hydrogen and CO₂ [32]. Natural gas is first
168 converted into a synthesis gas with carbon monoxide, hydrogen, water and CO₂, by partial oxidation,
169 steam reforming, auto-thermal reforming, which is a combination of the two previous paths, or crack-
170 ing [33]. CO₂ at high partial pressure is then separated from the synthesis gas by physical (methanol
171 or DEPG) or chemical (TEA) absorption [34].

172 The major challenges encountered with these technologies are namely their high investment costs, their
173 applicability to new plants only, and the difficulties associated with hydrogen combustion (e.g. higher
174 radiation share in the heat transfer to the cooled components and different fluid properties). The
175 first hydrogen-fired gas turbines will likely run on hydrogen combustion with air, as the production of
176 pure oxygen is costly and hydrogen premixed combustion is challenging. With some exceptions, the
177 current turbine materials cannot withstand temperatures above 1500 °C [35]. Other issues that may
178 be encountered are the flame stability and the production of nitrogen oxides, and research on these
179 topics is currently on-going [36].

- 180 • oxy-combustion;

- 181 • post-combustion.

182 The term *post-combustion* CO₂-capture refers to the removal of CO₂ from the flue gases of a power
183 plant (e.g. coal- or gas-fired), i.e. after the combustion process. At the difference of the exhausts
184 from coal-fired plants, which have a high CO₂-concentration, the exhausts from a gas turbine have
185 a CO₂-concentration of only 3 to 5%, as well as a relatively low pressure (near atmospheric), and
186 chemical absorption with MEA is preferred.

187 The rest of this work focuses on the pre- and post-combustion paths, as the oxy-combustion one is the
188 least mature and requires, by definition, production of pure oxygen, which is generally done by cryogenic
189 air separation at very low temperatures.

190 3.2.4. Electrification

191 The term *electrification* refers to the supply of electricity from *shore* to offshore oil and gas installations.
 192 A platform can be *partly* or *fully* electrified, depending on whether the heating demand, if any, is satisfied by
 193 electric heaters or by gas-fired burners, and on part of or all the electrical consumption is supplied by land-
 194 based electricity (Figure 3). The distance from the shore has an impact on the economic cost of electrifying
 195 a platform. However, no numbers could be found in the literature, and the analysis of current studies shows
 196 that electrification may be viable for short and medium distances.

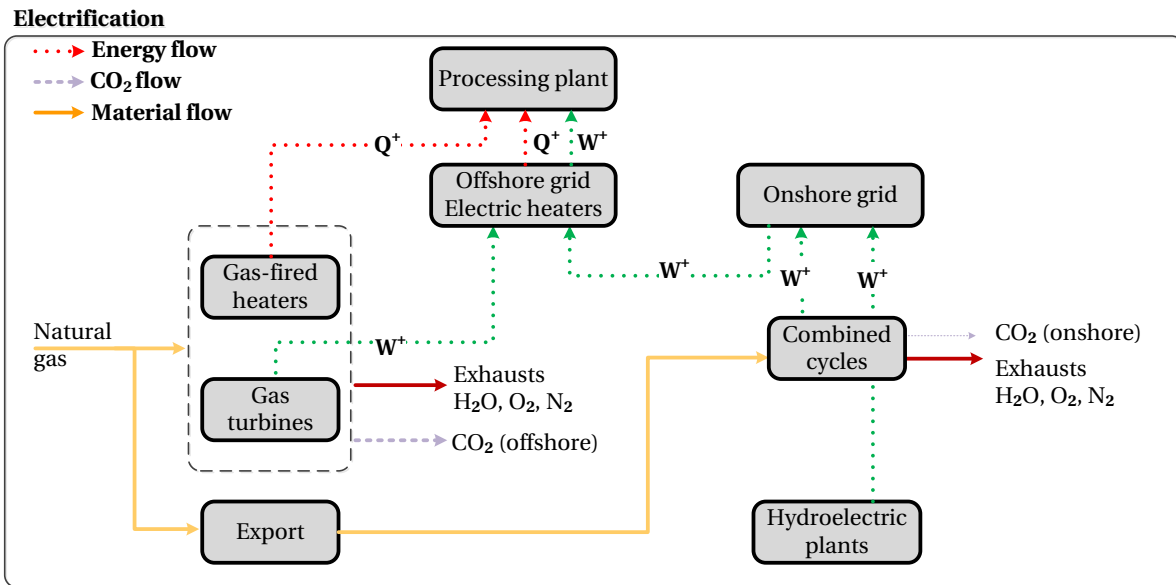


Figure 3: A generalised superstructure for integration of electrification on offshore platforms.

197 The system set-up of an electrified platform can differ from one facility to another, but, for simplicity,
 198 the transmission losses are assumed constant and equal to 8%. In practice, they vary with the power duty of
 199 the platform, on the type and geometry of the power transmission cable. Different electrification scenarios
 200 can be proposed, depending on how the power is supplied offshore:

- 201 • *no electrification*: all the power demand is satisfied by on-site utilities such as gas turbines;
- 202 • *partial electrification*: part of the power duty is met with power produced on onshore facilities;
- 203 • *total electrification*: all the required power is imported from the shore.

204 and on how the heating needs are met:

- 205 • *internal heat recovery* between process streams;

- 206 • *electric heaters* fuelled with on- or offshore power;
- 207 • *heat pumps* driven by on- or offshore power;
- 208 • *burners and furnaces* fuelled by natural gas.

209 Two electrification scenarios are considered in the rest of this study, referred as Scenario 1, where all the
210 power demand is satisfied with power from the mainland, but the heating demand is ensured by natural
211 gas combustion in heaters and furnaces. This scenario is likely for offshore platforms where recovering heat
212 from the gas turbine exhausts is not sufficient, if the oil is too viscous or heavy, or if the feed petroleum
213 has a low temperature. In Scenario 2, it is assumed that the heating demand is satisfied by electric heating,
214 which is likely if the heating demand is minor, and all power is supplied from shore.

215 Different electrification sources can be considered, depending on the case study and country of interest.
216 However, only the hydro- and combined cycle power plants options are assessed in this work, as these
217 possibilities are the ones that are mostly discussed in the literature. Those are the most relevant one in
218 the case of Norway, where the average electrical supply is mainly ensured by hydro-electric dams, and the
219 marginal one builds on the import of electricity from e.g. Denmark and Germany, where state-of-the-art
220 gas-fired power plants are implemented. The type of electrical production is relevant when performing an
221 environmental assessment of the different electrification options. A given solution may correspond to smaller
222 local emissions (site-scale) but greater global ones (country-scale).

223 3.2.5. *Practical limitations*

224 There are stringent limitations on oil and gas platforms with regards to the maximum space or weight
225 tolerable on-site, and system layouts with low equipment inventory may therefore be preferred to more
226 complex configurations. This limits the maximum efficiency of a waste heat recovery system and the CO₂-
227 separation potential of a CO₂-capture plant. For these reasons, only ‘simple’ waste heat recovery cycles,
228 built downstream the gas turbines, without extraction, and without supercritical fluids, are considered in
229 this work. The implementation of organic cycles based on CO₂ or alkanes (e.g. ethane and propane) for
230 recovering heat from the gas compression process is not analysed.

231 As processes with low volume may be favoured, the integration of pre-combustion CO₂-capture systems
232 may be relevant for future offshore platforms. The volume flow rate of the fuel gases is smaller than in
233 post-combustion plants, implying that the power generation system would be more compact. Only the
234 design set-ups with low equipment inventory are investigated, i.e. with one absorber and a scrubber, with
235 solely two pressure levels. In addition, there may be a need for a back-up power system or storage on-site
236 to compensate for possible failures of the electrical grid if the platform is directly connected to it.

237 3.3. Process modelling

238 The models of the oil and gas platform and of the CO₂-capture by monoethanolamine are developed
239 using the commercial flowsheeting software Aspen Plus[®] version 7.2 [20]. The modelling details of the steam
240 network, with the corresponding decision variables, are presented in Nguyen et al. [5] (production levels,
241 cooling and heating sources), and in Tock et al. [30] for the chemical absorption process (CO₂-loading,
242 absorption and regeneration temperatures and pressures).

243 The gas turbines are simulated based on the work of Pierobon et al. [37]: the part-load behaviour is
244 considered to predict the CO₂-emissions along a year of operation, taking the SGT-500 gas turbine as case
245 study. The model of the CO₂-capture units with monoethanolamine is developed using the commercial
246 flowsheeting software Aspen Plus[®] version 7.2 [20], based on the electrolyte NRTL method [38] for the
247 liquid phase and the Redlich-Kwong [39] EOS (equation of state) for the vapour phase. The models of
248 the physical absorption units are developed using the Perturbed-Chain Statistical Associating Fluid Theory
249 EOS [40].

250 4. System performance

251 4.1. Performance indicators

252 The system performance can be evaluated by a large variety of thermodynamic and economic parameters,
253 and this study focuses on:

- 254 • the energy efficiency of the cogeneration plant η , on a higher heating value (Δh^0) basis;

$$\eta_{CC} = \frac{\dot{Q}^- + \dot{W}^-}{\Delta h_{FG}^0 \cdot \dot{m}_{FG}^+} \quad (1)$$

255 where \dot{Q} and \dot{W} represent the energy transfers with heat and power. The superscripts $+$ and $-$ denote
256 the input and output flows. This definition considers that the heat output is useful, as it is the case
257 for a combined cycle with heat extraction, i.e. for a combined heat and power utility plant.

- 258 • the exergy efficiency of this plant ε , based on a dead state of 8 °C and 1 atm, as these are the average
259 air conditions in the North Sea;

$$\varepsilon_{CC} = \frac{\dot{E}_Q^- + \dot{E}_W^-}{\Delta h_{FG}^0 \cdot \dot{m}_{FG}^+} \quad (2)$$

260 where \dot{E}_Q and \dot{E}_W represent the exergy transfers with heat and power, and Δh^0 the exergy content
261 of the fuel gas.

- the grassroot costs C_{gr} of the CO₂-capture or waste heat recovery unit;

A precise analysis of the grassroot costs of a CO₂-capture or a waste heat recovery plant on an offshore platform is difficult because of (i) the very few, if none, studies on this topic, as well as the absence of case studies, (ii) the lack of knowledge on the necessary space and associated cost for installing the equipment items, (iii) the different approaches and cost correlations for evaluating the economics of a chemical or physical absorption plant.

Discussions with process and design engineers of offshore platforms stressed these three challenges. In particular, the cost of the space required for implementing these processes is expected to vary greatly from one platform to another, depending on the actual configuration of the platform. On the one hand, a well-design steam Rankine cycle may have a power capacity as high as each gas turbine installed on-site. In this case, this waste recovery unit could replace one of the gas turbines, and no additional space would be required. On the other hand, it is necessary to have redundant equipment to prevent system failures and avoid the resulting economic losses associated with a shut-down of the processing plant.

- the relative variation of natural gas δ_{NG} exported to the shore (increase or decrease);
- the reduction of CO₂-emissions δ_{CO_2} caused by the decrease of fuel gas consumption and/or the possible integration of a CO₂-mitigation plant;
- the changes in operating costs C_{op} ;

The changes in operating costs are related to (i) the additional cooling water and pumping demands on-site for the steam condensation, (ii) the maintenance costs of the additional equipment, (iii) the money savings with the reductions of CO₂-taxes, (iv) the money earnings with the increases of gas sales, and (v) the make-up of monoethanolamine, methanol and DEPG because of degradation issues and losses.

An accurate calculation of the system operating costs would require to know the economic value of the exported gas. However, the latter cannot be estimated accurately, because the gas is sent through the Åsgard pipeline system, where it is mixed with natural gas from the other petroleum fields located in the northern part of the North Sea. These hydrocarbon flows have very different chemical compositions and physical properties, and they are treated in onshore plants where they are refined into a large variety of hydrocarbon products (e.g. natural gas, liquid petroleum gases, diesel, naphtha, etc.).

Calculating a precise economic value of a single natural gas stream is therefore infeasible, as this would require (i) the measurements of the exact flow rates and compositions of the gas streams from each offshore facility, and these flows vary significantly with time, and (ii) a forecast of the variations of the

294 natural gas prices on the market, for different countries. On the contrary, the relative variation of the
 295 export gas flow δ_{NG} is a clearer performance indicator, as it depends solely on the facility under study.
 296 Finally, it is assumed that the integration of these additional processes does neither result in an
 297 increase of the number of operators, nor in a higher operator's salary. The economic performance
 298 of these processes can be assessed with regards to the potential fuel gas savings and reductions in
 299 CO₂-taxes.

- 300 • the cost of CO₂-avoidance CAC, defined as the ratio of the increase in investment costs over the
 301 decrease of CO₂-emissions;

$$CAC = \frac{C_{gr,CCS} - C_{gr,ref}}{\dot{m}_{CO_2,ref} - \dot{m}_{CO_2,CCS}} \quad (3)$$

- 302 • the cost of electricity COE, defined here as the levelised cost of electricity;

$$COE = \frac{\sum_{t=1}^n (C_{gr} + C_{op} + C_{mt} + C_{tax})(1+r)^{-t}}{\sum_{t=1}^n \dot{W}_{el}(1+r)^{-t}} \quad (4)$$

303 Economic analyses of carbon capture units may be evaluated using either the electricity (COE) or
 304 the CO₂-avoidance (CAC) cost. The first one illustrates the cost of using the produced natural gas
 305 in the on-site gas turbines, while the second one reflects the costs of reducing the CO₂ emissions.
 306 They depend on factors such as the production cost of the fuel gas, the capital costs with the well
 307 construction and the CO₂-tax on the hydrocarbon production.

- 308 • the power capacity or consumption of the additional systems \dot{W} .

309 In the following economic analysis, the base value assumed for the natural gas is taken as 8.08\$/GJ,
 310 which corresponds to the average sales price of natural gas in Norway for the year 2010, in the case of the
 311 Statoil company [41]. The base value assumed for the CO₂ tax is taken as 65 \$/t_{CO₂}, which corresponds to
 312 the current value in Norway for the offshore sector [2]. These estimates are used to perform the sensitivity
 313 analyses and to give some indications on the expected electricity and avoidance costs.

314 Other performance indicators such as the net present value and payback period provide valuable infor-
 315 mation on the economic profitability of a given investment. However, calculating these values is not relevant,
 316 as these indicators would present high inaccuracies due to the uncertainties of the capital (as reasonable
 317 information on the space cost is missing) and operating costs (since an exact estimate of the value of the
 318 exported natural gas cannot be given).

319 4.2. Multi-objective optimisation

320 4.2.1. Objective functions

321 The variety of process indicators that can be derived when evaluating the performance of an oil and gas
322 platform with CO₂-mitigation shows a competition between several factors. For example, the integration of
323 a post-combustion CO₂-capture unit results in a lower energy efficiency, higher investment costs but lower
324 CO₂-emissions. The trade-off between these competing factors is assessed by performing multi-objective
325 optimisations, applying an evolutionary algorithm. The objectives considered in the optimisation procedure
326 are (i) the maximisation of the power capacity \dot{W} or of the exergetic efficiency ε , (ii) the minimisation of
327 the investment costs C_{inv} , and (iii) the minimisation of the CO₂-emissions δ_{CO_2} .

328 4.2.2. Decision variables.

329 The optimal system configurations are displayed under the form of a Pareto optimal frontier, which
330 separates the research domain into the feasible but sub-optimal solutions, the feasible and Pareto-optimum
331 solutions, and the infeasible ones. A solution is said Pareto-optimum if a better-off with respect to one
332 objective results in a worse-off with respect to another one. The master decision variables amount to 48, of
333 which 18 are related to the operation of the steam cycle (e.g. pressures, temperatures, vapour fraction) and
334 5 to the selection of the cooling utility (e.g. process water and temperatures) and are shown in Nguyen et
335 al. [13]. The decision variables related to the CO₂-capture processes are presented in [Appendix E](#).

336 5. Performance comparison

337 5.1. Process optimisation

338 5.1.1. Pre-combustion

339 The impact of pre-combustion CO₂-capture technologies on the performance of the oil and gas platform
340 and on its utility plant can be assessed by performing a multi-objective optimisation, analysing the trade-off
341 between the exergy efficiency and the CO₂-capture rate (Figure 4). For simplicity, the possibility of purifying
342 the H₂-rich fuel obtained after the CO₂-capture process, by integrating pressure swing adsorption, is not
343 investigated.

344 The Pareto frontiers for the different pre-combustion processes show that the integration of CO₂-capture
345 results in a lower efficiency of the power plant, because of the energy used for heating (solvent regeneration),
346 cooling (solvent refrigeration, with physical absorption), and power (CO₂-compression and solvent pumping).
347 Three types of configurations with pre-combustion CO₂-capture are investigated: (i) a set-up without waste
348 heat recovery [n]; (ii) a set-up with steam Rankine cycle, where the recovered heat only comes from the
349 exhaust gases [s]; and (iii) a more advanced set-up, where heat can be recovered from the turbine exhausts
350 and different sections of the offshore system [a].

351 The impact of integrating a simple steam Rankine cycle, which utilises the heat from the gas turbine
 352 exhausts only, is investigated at first. The advantage of such configuration can be visualised by the horizontal
 353 shift of the Pareto frontiers, which illustrate that the gain in exergy efficiency is about 15 to 20 %-points,
 354 whether physical or chemical absorption is implemented. The third set-up, which is the most complicated,
 355 leads to an increase in efficiency for all physical and chemical solvents, at the expense of a higher number of
 356 equipment items. This configuration may not be chosen in practice for offshore implementation, since the
 357 gains in efficiency and fuel savings may be judged marginal with regards to the added complexity. However,
 358 it sets a limit on the maximum performance that could be expected.

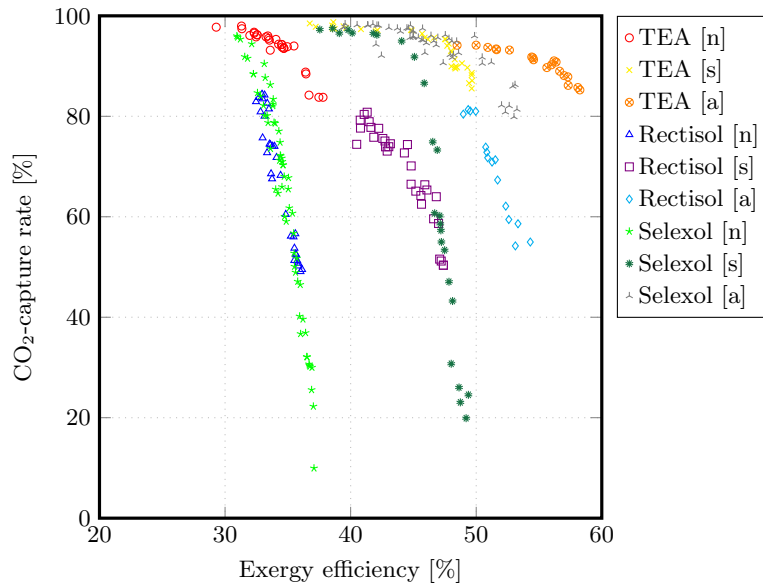


Figure 4: Pareto-optimal solutions for the site-scale integration of pre-combustion CO₂-capture processes on offshore platforms: trade-off between the CO₂-emissions and exergy efficiency, without steam cycle [n], with simple [s] and advanced [a] steam cycle.

359 5.1.2. Post-combustion

360 Similarly, the impact of post-combustion CO₂-capture technologies on the performance of the oil and gas
 361 platform and on its utility plant can be evaluated by performing a multi-objective optimisation, analysing
 362 the trade-off between the increase in power generation due to the waste heat recovery cycle, and the CO₂-
 363 capture rate (Figure 5). Two types of configurations are investigated: (i) a set-up with only waste heat
 364 recovery, without chemical absorption, and (ii) a set-up with post-combustion CO₂-capture and a steam
 365 Rankine cycle, to visualise the energy penalty induced by the capture unit. All the configurations with
 366 post-combustion CO₂-capture include a waste heat recovery cycle, because this would result in a higher
 367 thermal efficiency, and the chemical absorption of CO₂ is favoured at relatively low temperatures (about

368 40 °C).

369 The maximisation of the net power capacity and the minimisation of the CO₂-emissions are clearly
370 conflicting objectives, since CO₂-capture is favoured with large flows of solvent and high regeneration tem-
371 perature. Significant amounts of heat from the exhaust gases are required and cannot be used in the steam
372 network for electricity generation purposes.

373 The Pareto frontiers illustrate the trade-off between the increase in net power capacity and the reduction
374 in CO₂-emissions. In general, the integration of a steam network allows for a greater power production,
375 ranging from 3 to 8 MW at design conditions. The efficiency increases to about 35–40 % when the steam
376 cycle is run at full-load conditions, and between 28 and 33 % when run for the normal operating conditions.
377 All the optimal steam cycles consist of a single production level at a pressure between 8 and 13 bar and
378 a condensation level below 0.5 bar. The cooling utility used in the steam condenser is the process water,
379 i.e. the cooling water from the processing plant at about 16.5 °C. The export of natural gas to the shore
380 increases by up to 14 %, and this goes along with a reduction of the CO₂-emissions of up to 16–20 %.

381 If post-combustion CO₂-capture is integrated, the total CO₂-emissions of the platform can be reduced by
382 up to 70 %, of which \simeq 10–20 %-points are related to the steam network, and \simeq 50–60 %-points are associated
383 with the CO₂-capture unit. The remaining CO₂-emissions are caused by the flaring and secondary gas
384 turbines on-site. The export of natural gas also increases, although the savings in fuel gas are not as
385 significant as if only a steam cycle was integrated, because of the power demand of the CO₂-sequestration
386 unit. The CO₂-capture rate varies between 83 and 93 %, which corresponds to a reduction of the total
387 CO₂-emissions of the platform between 55 and 70 %. Two configurations, which have a CO₂-avoidance rate
388 of 55 and 70 %, respectively, are discussed further in the energy integration and economic analyses.

389 5.2. Energy integration

390 5.2.1. Pre-combustion

391 The introduction of CO₂-capture processes by physical absorption (Figure 6) results in different conclu-
392 sions whether the Rectisol or Selexol technology is integrated. The first one systematically implies a need
393 for cooling down the solvent (methanol) and the feed gas below ambient temperatures, increasing the elec-
394 tricity consumption as a refrigeration cycle should be implemented. The second one creates a refrigeration
395 demand only if high CO₂-capture rates are of interest, since the CO₂-solubility in DEPG is higher at low
396 temperatures.

397 In terms of power consumption, the differences between the Selexol and Rectisol processes are minor,
398 because the synthesis gas should preferably be compressed at pressures higher than 40–50 bar in both cases
399 to ensure high CO₂-partial pressure, and the solvent pumping process has a negligible power demand in
400 comparison to the CO₂-compression.

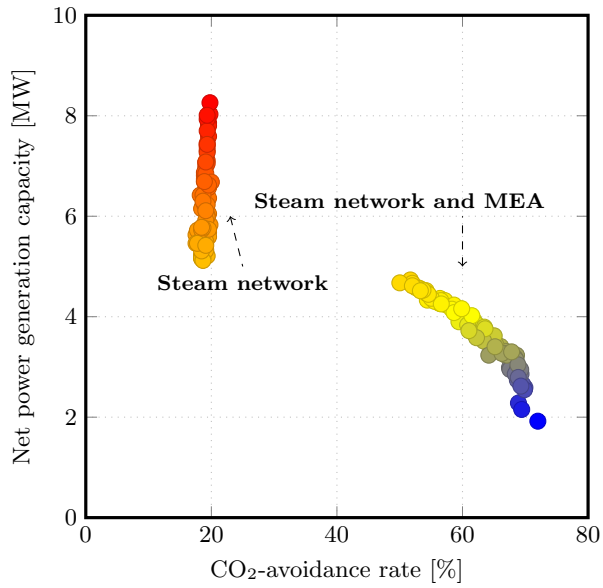


Figure 5: Pareto-optimal solutions for the site-scale integration of post-combustion CO₂-capture processes on offshore platforms: trade-off between the CO₂-emissions and net power capacity.

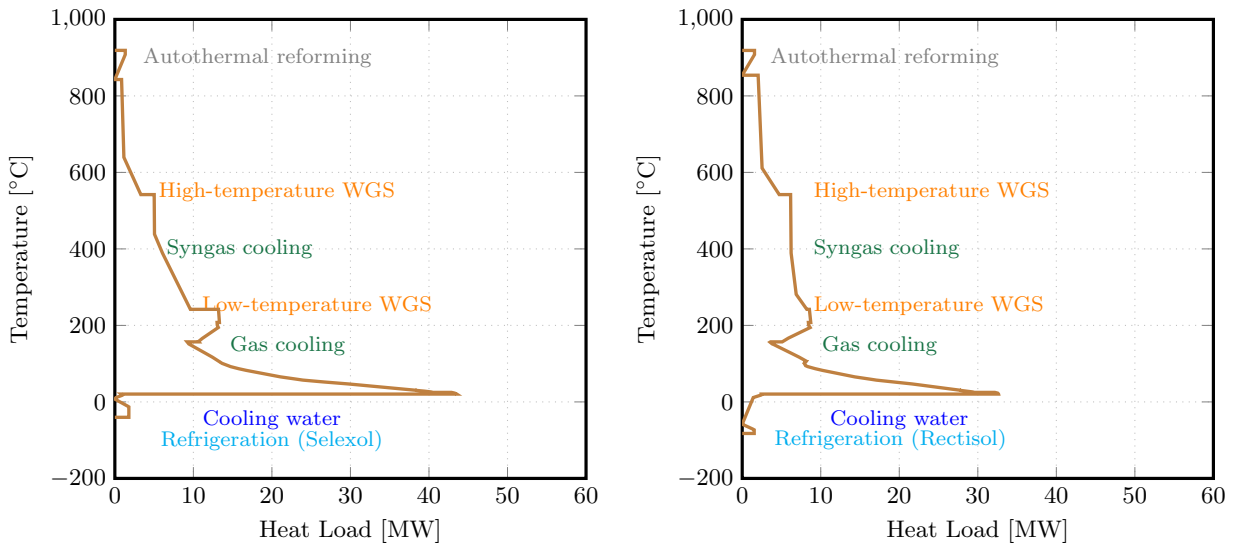


Figure 6: Balanced Composite Curves for an oil and gas platform with pre-combustion CO₂-capture based on autothermal reforming and using Selexol (left) and Rectisol (right).

401 There is a large variety of sets of operating conditions for which a CO₂-capture rate can exceed 80 %. It
 402 is assumed in the following examples that the H₂-fuelled gas turbines replace the current SGT-500, which
 403 satisfy the baseline power demand of 16,500 kW and the additional power consumption, which can be split
 404 (Figure 7 and Figure 8) into the power demands of the synthesis gas preparation process, the DEPG pumping
 405 in the Selexol process, the CO₂-compression and the refrigeration cycle. The same trends are found with

406 system configurations based on Rectisol, with a higher share of the refrigeration cycle because of the lower
 407 temperatures of the methanol solvent ($\simeq -70^{\circ}\text{C}$) compared to the DEPG ($\simeq -10^{\circ}\text{C}$).

Syngas composition (mol %)	After preparation	After Selexol	
H ₂	54.4	64.4	
CO	1.74	-	DEPG purity (solvent, before absorber): 97.1 mol %
CO ₂	17.9	3.6	
N ₂	25.7	30.0	
H ₂ O	0.2	-	

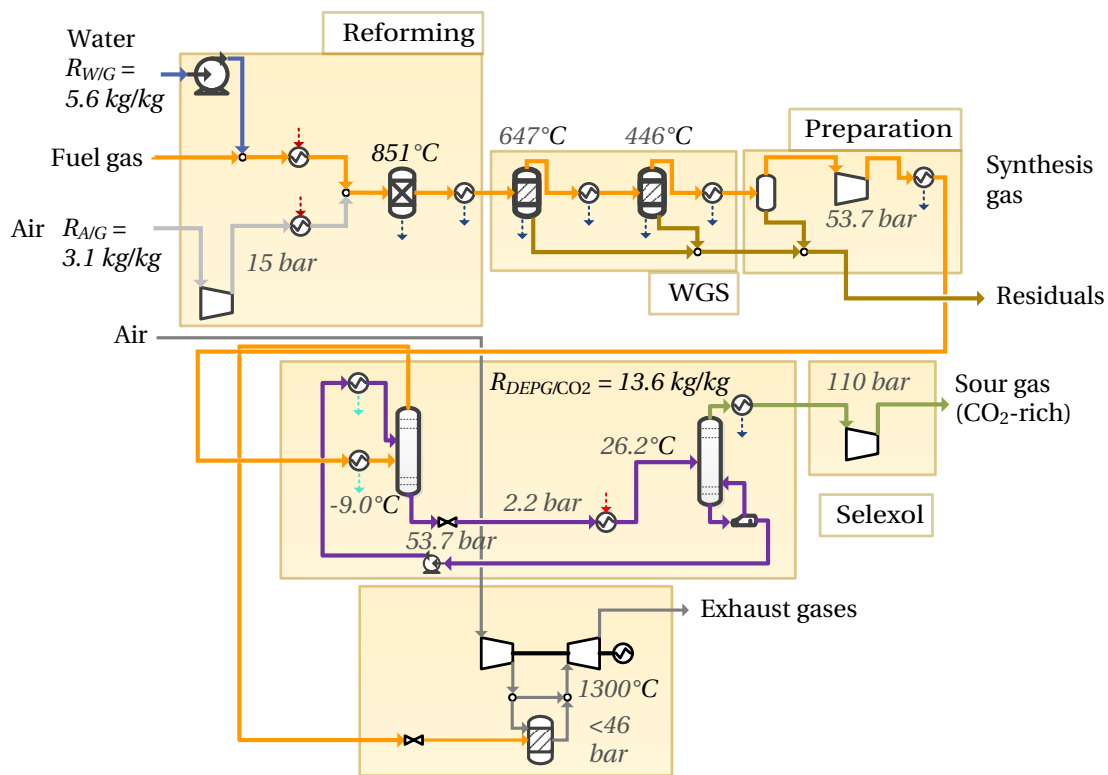


Figure 7: Example of configuration of pre-combustion CO₂ capture with autothermal reforming and Selexol.

408 Chemical absorption processes with TEA (Figure 9) may also compete with physical absorption ones
 409 since the CO₂-content may be as high as 25% for a H₂-purity of 75%. The heating requirements at low
 410 temperatures ($\leq 150^{\circ}\text{C}$) increase in such cases, as a chemical absorption unit is always characterised by a
 411 heating demand (Figure 10) for regenerating the amine solvent. The highest H₂-purity is reached for chemical
 412 absorption with TEA along with steam methane reforming: it can exceed 90% because the produced syngas
 413 is not diluted with nitrogen, while it is limited to 65–70% if the reforming process is autothermal. The use

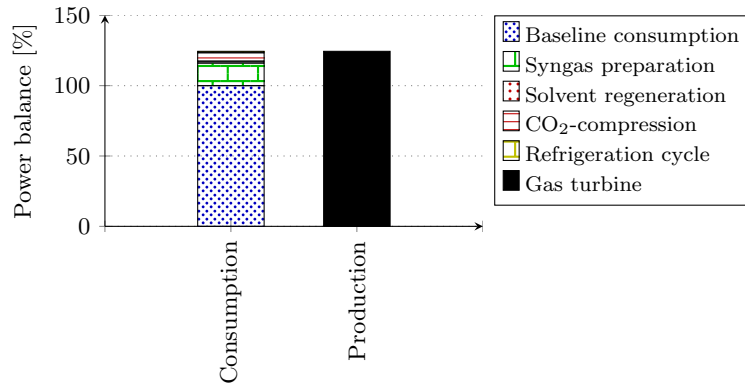


Figure 8: Power balance for an example of layout with ATR and Selexol.

414 of chemical absorption avoids large pressure drops in the CO₂-capture unit and the need for refrigeration
 415 (Figure 11), which explains the slightly higher electrical efficiency of pre-combustion CO₂-capture processes
 416 with amines in the later optimisations.

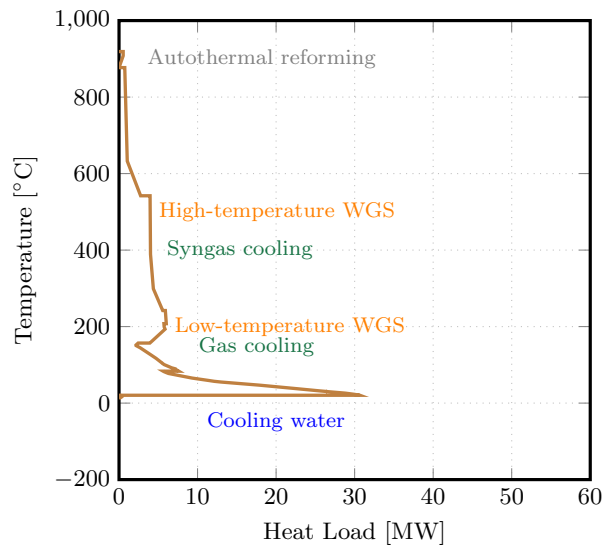


Figure 9: Balanced Composite Curve for an oil and gas platform with chemical absorption based on TEA.

417 Moreover, the energy required to regenerate the chemical solvent can be covered by utilising the heat
 418 from the water-gas shift reactors and syngas coolers, and that results in a smaller demand for cooling
 419 water compared to the process layouts with physical absorption. The introduction of a cogeneration utility
 420 together with a chemical absorption plant with TEA can be beneficial since it would result in a better match
 421 between the temperatures of the regeneration process and the hot utilities. However, the maximum amount
 422 of electricity that can be generated is smaller than if a physical absorption unit is integrated, because less
 423 heat is available in the temperature range of 300 to 600 °C.

424 The losses of CO₂ with the knock-out water are negligible in all cases, representing less than 0.5 % of
 425 the total carbon entering the capture unit. However, there are losses of hydrogen with the CO₂ sent to
 426 sequestration, which limit the CO₂-purity to an upper bound of 96–97 %. The implementation of a 2-stage
 427 regeneration plant may be beneficial, but this configuration is not further studied in this work.

Syngas composition (mol %)	After preparation	After TEA	TEA purity (concentration, before absorber): 24.4 wt %
H ₂	53.4	63.9	
CO	0.4	-	
CO ₂	18.8	3.4	
N ₂	27.2	32.7	
H ₂ O	-	-	

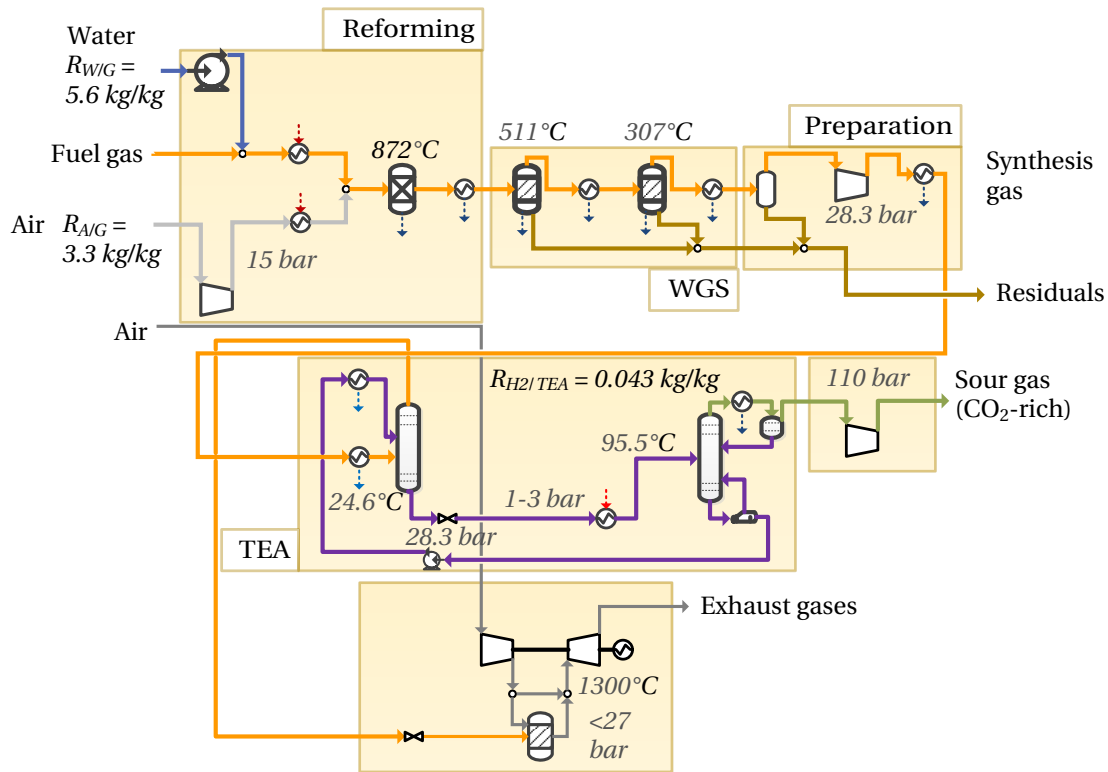


Figure 10: Example of configuration of pre-combustion CO₂ capture with autothermal reforming and TEA.

428 5.2.2. Post-combustion

429 The installation of CO₂-capture plant together with a steam cycle (Figure 12 and Figure 13) presents
 430 by far the greatest potential for CO₂-reduction, but only reduces the exergy losses with the exhaust gases
 431 taking place at temperatures of 100 to 330 °C. There is a potential for recovering exergy at low temperatures,

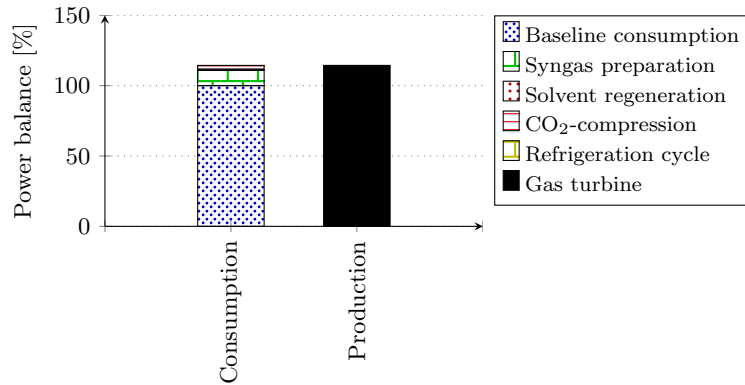


Figure 11: Power balance for an example of layout with autothermal reforming and TEA.

432 which is, in the proposed configurations, dissipated with cooling water, but such option may be particularly
 433 challenging. Smaller quantities of heat are dissipated into the environment with cooling water for the process
 434 designs including CO₂-capture, as a fraction of the heat contained in the exhaust gases is used to regenerate
 435 MEA instead and dissipated into the environment.

436 The comparison of the two configurations with post-combustion capture, and a CO₂-avoidance rate of
 437 55 and 70 %, show that the net power capacity and CO₂-capture potential are clearly conflicting objectives
 438 (Figure 14). Integrating a CO₂-capture plant seems more appropriate when it is performed in conjunction
 439 with a steam cycle. The heat required in the desorption column is provided by the cooling of the exhaust
 440 gases after the gas turbines and possibly prior to the steam cycle, illustrating the synergy between the
 441 bottoming cycle and the CO₂-capture plant.

442 The power demand at normal operating conditions increases by about 12 %, as a consequence of the
 443 power demand for compressing the CO₂ and pumping the MEA. High CO₂-capture rates result in a higher
 444 heating demand in the amine regeneration process, which should be satisfied by recovering larger amounts of
 445 waste heat from the turbine exhausts, and in a greater power consumption in the CO₂-compression process.
 446 The purity of the CO₂-rich stream exceeds 97 % on a molar basis.

447 5.2.3. Electrification

448 The higher gas export results in larger cooling duty in the gas treatment section and greater power de-
 449 mand (Figure 15). However, platform electrification does not change significantly the temperature-enthalpy,
 450 or temperature-exergy profiles of an oil and gas platform. The exergy destruction and losses in the gas
 451 turbines are eliminated, but they are replaced with the ones related to the onshore plants (combined cycles
 452 or hydroelectric facilities), to the gas-fired heaters (if any), and to the transmission cables (power losses).
 453 The exact values of these irreversibilities are not calculated in this work, but they are expected to be smaller
 454 because of the greater efficiency of the onshore power plants and the smaller fuel gas consumption.

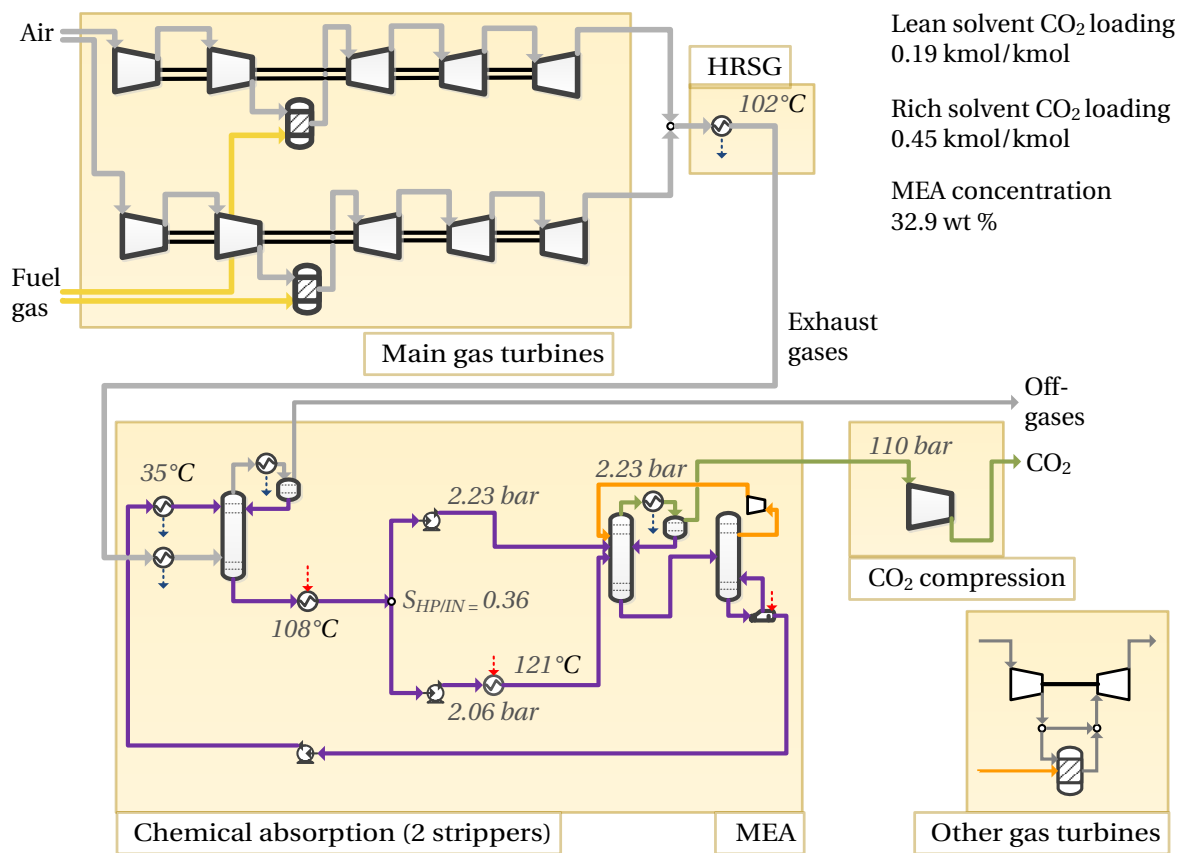


Figure 12: Example of configuration of post-combustion CO₂ capture with MEA.

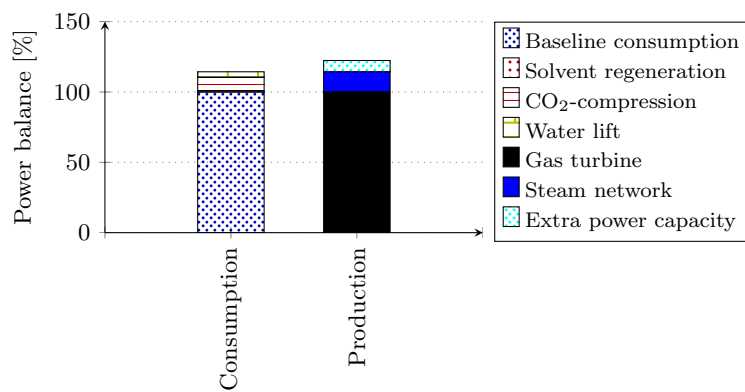


Figure 13: Power balance for an example of layout with MEA.

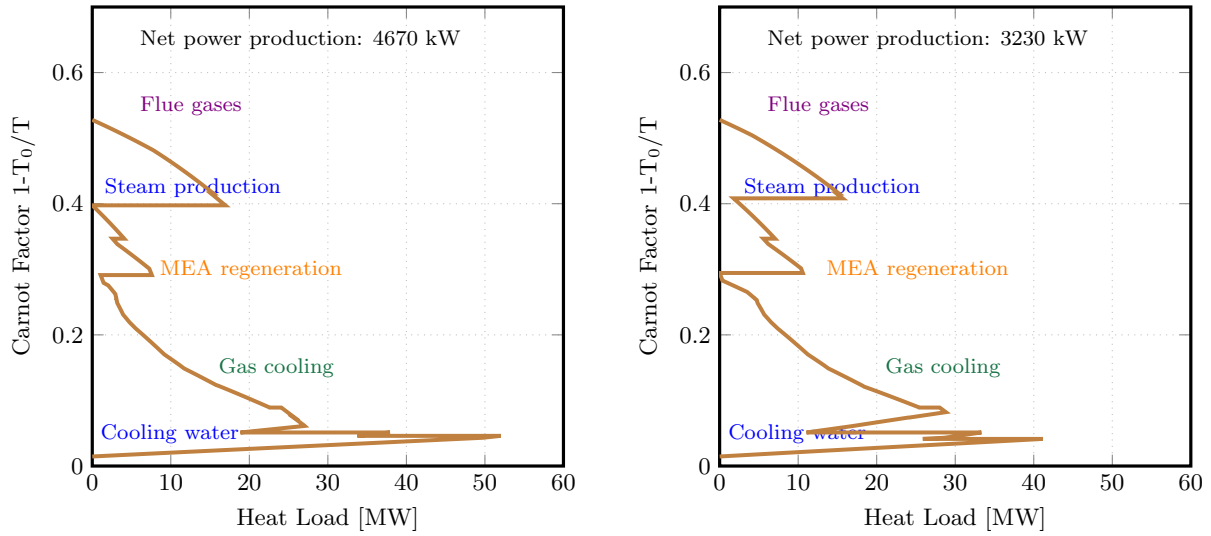


Figure 14: Balanced Grand Composite Curves, on an exergy basis, of the oil and gas platform system, including a steam Rankine and a post-combustion CO₂-capture unit, with low (left) and high (right) CO₂-capture rate.

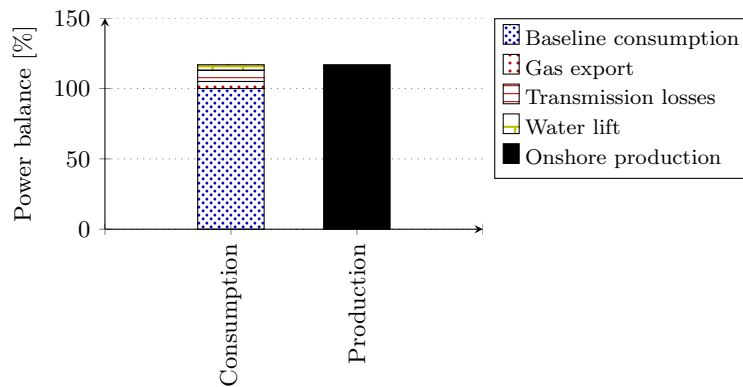


Figure 15: Power balance for an example of layout with electrification (Scenario 1).

455 5.3. Economic performance

456 5.3.1. Economic build-up

457 A preliminary assessment of the economic build-up of the capital costs of such plants (Figure 16) suggests
 458 that pre-combustion CO₂-capture processes are more costly than post-combustion ones, which is consistent
 459 with the conclusions drawn in the literature. The additional costs, related to the initial costs of the processes
 460 installed on the Draugen platform, vary between 10 to 40%: the application of the economic correlations
 461 suggests that a configuration based on the Selexol process is the most costly, but the differences with the
 462 other system layouts are within the range of uncertainty given in Turton et al. [18].

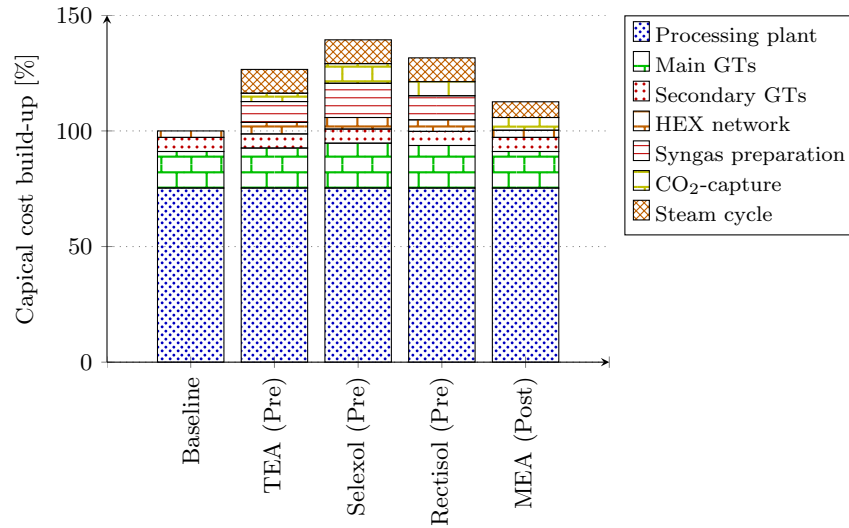


Figure 16: Capital cost build-up for offshore platforms with and without CO₂-capture, based on the process equipments of the Draugen platform.

463 5.3.2. Sensitivity analyses

464 The economic profitability of CO₂-mitigation options is impacted by several parameters and possible
 465 future policies in the offshore sector [42], and evaluating the sensitivity of the electricity and avoidance costs
 466 is thus of interest [43]. The ones that are considered the most important are the following:

- 467 • natural gas price;
- 468 • CO₂ tax;
- 469 • capital costs of CO₂-injection wells.

470 Their effects are assessed by performing sensitivity analyses on each of these three parameters, based
 471 on estimations found and deduced from studies available in the literature. Two post-combustion cases,
 472 derived from the multi-objective optimisation routine, are used as references for analysing the sensitivity of
 473 the electricity and CO₂-avoidance costs for process configurations with medium- and high CO₂-reduction
 474 potential.

475 The baseline values assumed for conducting the sensitivity analyses are a natural gas price of 8.08 \$/GJ,
 476 a lifetime of 30 years, site-specific costs of M\$15, and a carbon tax of 65 \$/t_{CO₂}. The choice of these
 477 economic data is based on the current situation in Norway. Care should therefore be exercised when
 478 drawing conclusions about the economic profitability of CO₂-mitigation options, as the conclusions drawn
 479 in the present study are based on a specific European case. The sensitivity of the COE and CAC is then
 480 investigated by keeping these parameters but one constant, and vary each single parameter individually,
 481 assuming it will not influence the value of the others.

482 For readability, only the sensitivity analysis on the natural gas price is presented in the main matter,
483 the other ones being shown in [Appendix F](#).

484 *Natural gas price.* This evolution is difficult to predict, as it is highly different between European and
485 American countries. In the past, the oil and gas prices have followed similar movements, but these trends
486 have changed, at least in North America, because of geopolitical factors and the exploitation of shale gas.
487 Changes in oil price were accounting for more than 40–65 % of the changes in gas price until 2009, but
488 only for less than 25 % since [44]. An increase of the gas prices, as described by most economic analyses,
489 would push towards the implementation of waste heat recovery cycles because of the possible energy and gas
490 savings. European gas prices are currently about twice as high as American ones, and they are projected to
491 be around 10, 12 and 16 \$/GJ in 2020, 2030 and 2050, respectively [45]. The boundary values used for the
492 sensitivity analysis on the natural gas price are derived from this resource, and validated with other works
493 performed by the European Commission [42] and the International Energy Agency [46]. Further works [47]
494 also suggest that this resource price will increase over years, as a result of the depletion of the natural gas
495 resources.

496 The electricity and CO₂-avoidance costs clearly follow a linear dependence on the natural gas price
497 (Figure 17). At high natural gas prices, an oil and gas platform without CO₂-capture appears to be the
498 most competitive option if no carbon tax is set, but the implementation of waste heat recovery seems
499 promising, as it results in fuel savings by up to 15 % points. On the contrary, the interest for CO₂-capture
500 plants is small, since these processes reduce the energy efficiency of the power generation unit, resulting in
501 greater natural gas consumption.

502 The introduction of a carbon tax, in this case of 65 \$/t_{CO₂}, increases the profitability of oil and gas
503 platforms with CO₂-capture. The first CCS process configuration appears to be competitive over a large
504 range of natural gas prices, while the second one is only competitive for a resource price below 4 \$/GJ. The
505 same trends can be visualised by analysing the variations of the CO₂-avoidance cost with the natural gas
506 price, which increase more sharply in the second case.

507 5.3.3. *Economic scenarios*

508 There is a clear trade-off between the economic performance and the degree of CO₂-abatement of oil and
509 gas platforms, and the process configurations that are optimal will obviously differ depending on the field
510 and the future economic scenarios (Table 1). The impacts of variations of parameters such as the natural gas
511 price, CO₂ tax, economic lifetime and capital costs are investigated by comparing three different scenarios,
512 a *low* scenario, which corresponds to a situation where no CO₂-tax is imposed on the offshore sector, a
513 *baseline* scenario, which represents the current economic context in Norway, and a *high* scenario, which is
514 characterised by a high CO₂-tax.

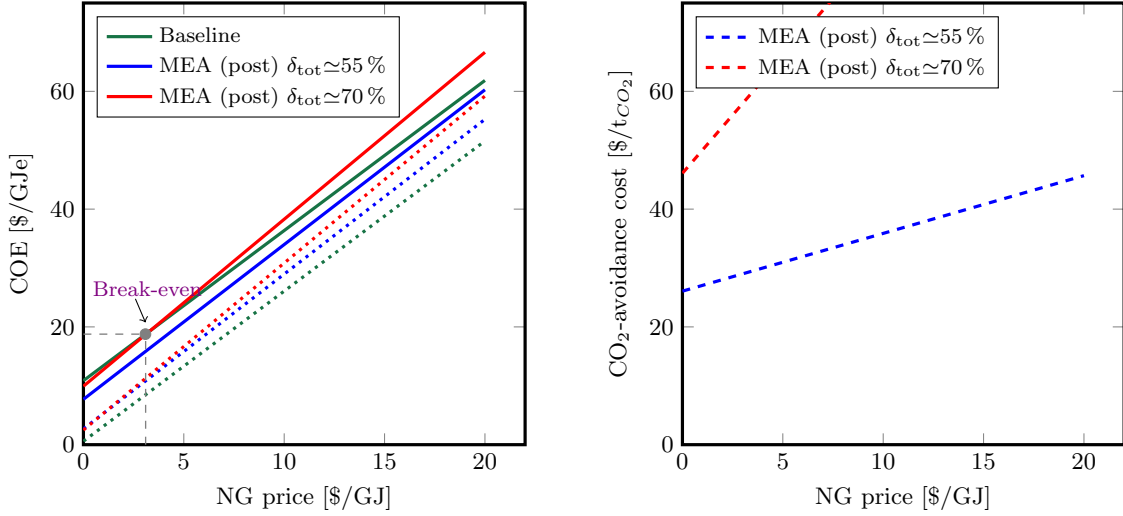


Figure 17: Sensitivity of the electricity production (COE) and avoidance (CAC) costs of an offshore platform with and without integration of post-combustion CO₂-capture to the natural gas price, with (solid) and without (dotted) CO₂-tax.

515 The low/high scenarios represent therefore extreme case scenarios, presenting the highest/lowest natural
 516 gas price and lowest/highest CO₂-tax that may be expected in the future, based on the current estimates
 517 and trends. The baseline scenario presents strong similarities with the ones discussed in the works of the
 518 European Commission [42] and the International Energy Agency [46,47]. This approach of considering three
 519 values (low, middle and high) for the fuel costs was applied in other works, such as the techno-economic
 520 analysis conducted by the Zero Emission Platform [48]. The following results should not be viewed as precise
 521 technoeconomic results, but shall be seen as indications on the trends that can be expected, together with
 522 the results of the sensitivity analyses.

Table 1: Tested economic scenarios on the post-combustion CO₂-capture unit.

Scenario	Base	Low	High
Natural gas price [\$/GJ]	8.08	16.16	4.04
CO ₂ -tax [\$/tCO ₂]	65.6	0.00	131.2
Expected lifetime [years]	30	20	40
Capital costs (CO ₂ -wells) [M\$]	≥ 15 and ≤ 30	≥ 30	≤ 15

523 High CO₂-capture rates are favoured with high CO₂-tax, small well capital costs, and low gas costs,
 524 because the large economic penalties on the CO₂-emissions compensate the additional investment costs of a
 525 CO₂-capture unit and the possible benefits with a greater gas export. Medium CO₂-capture rates, i.e. with
 526 a steam cycle only or with a small capture unit capacity, are preferable with high fuel gas production costs,
 527 since the integration of a waste heat recovery cycle allows for a smaller gas consumption on-site (Figure 18).

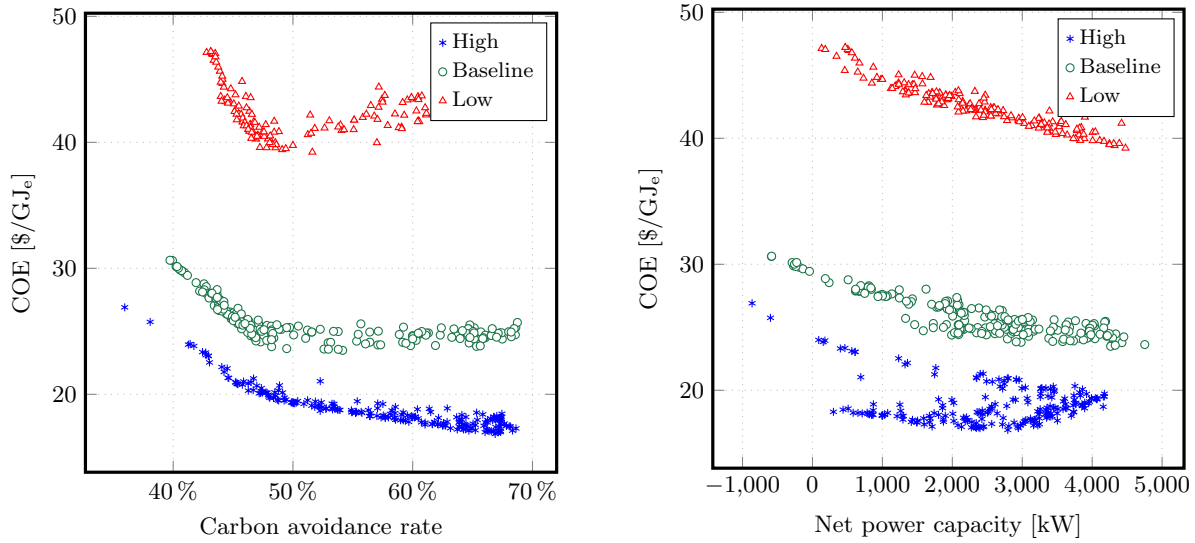


Figure 18: Pareto-optimal solutions for the site-scale integration of CO₂-capture with steam networks: trade-off between the electricity cost (with CO₂-tax), CO₂-emissions and net power capacity.

528 5.4. Environmental impact

529 5.4.1. Local and global emissions

530 There is likewise a lack of knowledge on the environmental impact of CO₂-capture plants on offshore
 531 platforms. The integration of such processes obviously results in a reduction of the local CO₂-emissions, but
 532 the installation of additional equipment items and the discharge of amines to the environment may have other
 533 harmful impacts. In the case of electrification (Figure 19), the local fuel gas CO₂-emissions are completely
 534 eliminated in Scenario 2 and decreased by about 90–95 % in Scenario 1. The remaining emissions consist of
 535 the release of methane, CO₂ and other greenhouse gases by flaring and venting. The global emissions are
 536 reduced by up to 45 % in Scenario 2 and by more than 50 % in Scenario 1, if the supplied power comes from
 537 gas-fired combined cycle power plants with a thermal efficiency of 55 %. This bigger decrease in Scenario
 538 1 can be explained by the lower transmission and conversion losses. There are, a priori, no CO₂-emissions
 539 associated with fuel consumption if hydraulic power is used instead of natural gas.

540 5.4.2. Life cycle assessment

541 The environmental performance of the whole process chain, i.e. from the resource extraction to the
 542 decommissioning of the offshore platform, can be evaluated based on a life cycle assessment, considering
 543 several impacts (e.g. GWP) and various analyses methods (e.g. CML 2001 [49]). Only the results of the
 544 IPCC07 method are presented in the main matter, the results of the other assessments are given in [Appendix](#)
 545 [G](#).

546 All configurations combining a steam network and a CO₂-capture unit have, overall, a beneficial effect

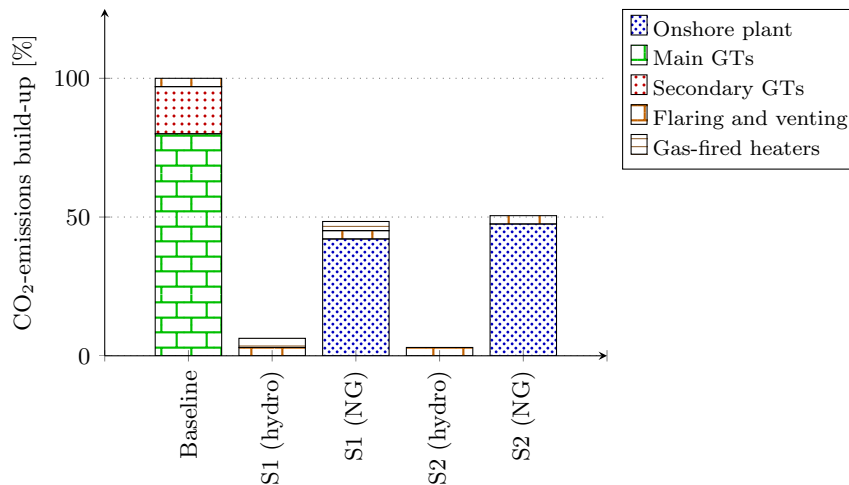


Figure 19: CO₂-emissions build-up for offshore platforms with and without electrification.

547 with respect to the global warming potential impacts (Figure 20), because of the reductions in greenhouse
 548 gas emissions over the facility lifetime. The major contribution to the remaining GWP impact corresponds
 549 to the fossil CO₂-emissions that are not sequestered. The global warming potential effects associated with
 550 the construction and manufacturing of the process components are negligible in comparison.

551 Similarly, all configurations including electrification have a beneficial effect with regards the global warm-
 552 ing potential impacts, because of the lower CO₂-emissions over the lifetime of the offshore facility. The main
 553 contributions to the remaining impacts correspond to (i) the fuel emissions associated with the combustion
 554 of natural gas in power plants, (ii) the process components, including the voltage cables, and (iii) the flaring
 555 and venting discharges. The contributions from the voltage cables are much smaller than the contributions
 556 of the other process components, and are negligible compared to the emissions of the non-sequestered CO₂.
 557 However, these results build on the assumption that there is enough power on the electrical grid to meet the
 558 power demands offshore. The picture would likely be different if additional power plants have to be built as
 559 it would require additional resources and materials.

560 6. Conclusion

561 Different CO₂-mitigation options for the offshore oil and gas sector were presented and compared,
 562 based on thermodynamic, economic and environmental performance indicators. The integration of pro-
 563 cesses such as steam bottoming cycles, pre- and post-combustion CO₂-capture has been analysed, using
 564 a multi-disciplinary approach that combines thermodynamic analysis tools with optimisation routines and
 565 process integration methods.

566 In all cases, the integration of a steam network is revealed to be profitable, with an increase of the power
 567 generation capacity of up to 8 MW and a greater gas export of up to 16%. The integration of pre- and post-

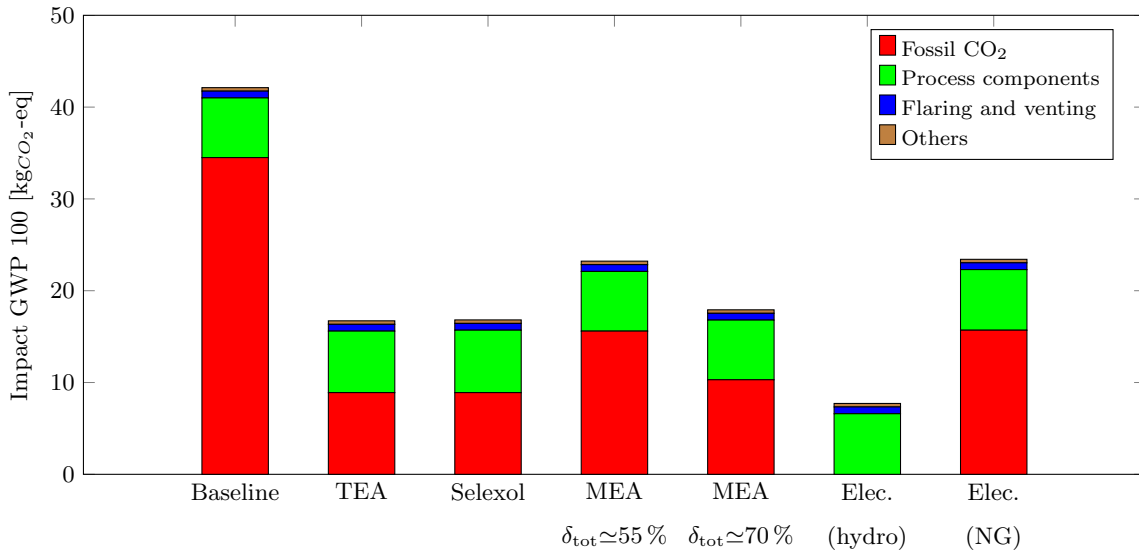


Figure 20: Comparison of the scenarios with and without CO₂-capture: IPCC 07 (impact method) – GWP 100 (impact category)

568 combustion CO₂-capture processes seems to be technically feasible, resulting in an energy penalty, but in
 569 lower CO₂-emissions and therefore greater operating profits. The multi-objective optimisations help assess
 570 the trade-off between the gain in power production, reduction in CO₂-emissions, and additional investment
 571 costs. The sensitivity analyses demonstrate that the economic context, i.e. the value of the CO₂-tax, the
 572 economic field lifetime, and the natural gas market price, has a critical importance on the profitability of
 573 all these options.

574 Electrifying an oil and gas platform is shown to be beneficial both from a thermodynamic and environ-
 575 mental perspective, because onshore power plants have a higher efficiency than the gas turbines installed on
 576 offshore platforms. The reductions in CO₂-emissions strongly depend on whether the electricity available
 577 on the grid is generated from gas-fired combined cycle power plants or hydroelectric facilities.

578 Finally, it can be concluded that the combination of process integration tools and life cycle assessments
 579 illustrates the benefits of mitigating CO₂-emissions in the offshore oil and gas sector. Each option may
 580 be promising or competitive in the future, if economic incentives are taken to push the offshore industries
 581 towards the development of more energy-efficient and environmental-friendly solutions.

582 Acknowledgments

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 584 2034/E30 led by Teknova is acknowledged.

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694 Appendix A. Thermodynamic relations

In the case of an offshore platform, energy enters and exits this system with material streams (e.g. petroleum feed, imported gas, fuel air, as well as oil, gas and produced water), with power (e.g. imported or exported electricity from the mainland or to other platforms) and with heat (e.g. heat losses by component radiation). Without considering the special cases with imported gas (e.g. lift purposes) or power import (e.g. electrification), the energy balance for the processing and utility plants of the oil and gas facility can then be expressed as:

$$\dot{H}_{\text{feed}} + \dot{W}_{\text{UT}} + \dot{Q}_{\text{UT,heat}} = \sum_k \dot{H}_k + \dot{Q}_{\text{PP,cool}} \quad (\text{A.1})$$

$$\dot{H}_{\text{k,fuel}} + \dot{H}_{\text{air}} = \dot{Q}_{\text{UT,cool}} + \dot{Q}_{\text{UT,heat}} + \dot{W}_{\text{UT}} \quad (\text{A.2})$$

695 where:

- 696 • \dot{H} stands for the energy rate carried with the ingoing material flows (feed denoting the feed streams
697 from the wells);

- 698 • or for the outgoing streams;
- 699 • \dot{W} for the energy transported with power, imported or exported to the mainland or other platforms.
- 700 • \dot{W}_{UT} is the power consumed within the separation and treatment modules, as well as in electric heaters,
- 701 which is produced in the utility plant;
- 702 • $\dot{Q}_{PP,heat}$ is the heat entering the processing plant, generally by direct heat exchange with the exhausts
- 703 of a gas turbine, or by indirect heat exchange, by using a heating medium (e.g. hot water or hot
- 704 glycol);
- 705 • $\dot{Q}_{PP,cool}$ is the heat entering the processing plant, generally by direct heat exchange with the exhausts
- 706 of a gas turbine, or by indirect heat exchange, by using a heating medium (e.g. hot water or hot
- 707 glycol);
- 708 • $\dot{Q}_{UT,cool}$ is the energy transferred from the power plant to the cooling medium (e.g. cooling air,
- 709 seawater or glycol-water mixtures) in, for instance, a steam condenser.

In details, the exergy balances can thus be expressed as:

$$\dot{E}_{feed} + \dot{E}_{imp} + \dot{E}_{air} + \dot{E}_{cw} + \dot{E}_{imp}^W = \sum_k \dot{E}_k + \dot{E}_{exh} + \dot{E}_{rw} + \dot{E}_{exp}^W + \dot{E}_{d,OP} \quad (A.3)$$

$$\dot{E}_{feed} + \dot{E}_{UT}^W + \dot{E}_{UT,heat}^Q = \sum_k \dot{E}_k + \dot{E}_{PP,cool}^Q + \dot{E}_{d,PP} \quad (A.4)$$

$$\dot{E}_{k,fuel} + \dot{E}_{air} = \dot{E}_{UT,cool}^Q + \dot{E}_{UT,heat}^Q + \dot{E}_{UT}^W + \dot{E}_{exh} \quad (A.5)$$

710 where:

- 711 • \dot{E} denotes the exergy flow associated with a given stream of matter;
- 712 • \dot{E}^W denotes the exergy transferred with power, and has the same value than its energy;
- 713 • \dot{E}^Q denotes the exergy transferred with heat, and has a smaller value than its energy, as it depends
- 714 on the temperatures of the environment and at which the heat transfer takes place;
- 715 • \dot{E}_d is the exergy destroyed in the overall (OP), processing (PP) and utility plants (UP);
- 716 • imp denotes the imported gas for injection or power generation, air for the air processed through the
- 717 gas turbines, cw stands for the seawater used for cooling needs, exh for the exhaust gases, rw for the
- 718 treated and rejected cooling water, and k for the several oil and gas streams.

719 **Appendix B. Energy integration**

720 The heat cascade problem is formulated as a MILP (Mixed Integer Linear Programming) problem,
 721 assessing precisely the heating and cooling requirements of the system under study. It follows the work of
 722 Maréchal et al. [50], with the aim of minimising the operating costs associated with the selection and use of
 723 the utilities (e.g. process water and seawater). The heating demand is satisfied by waste heat recovery from
 724 the turbine exhausts, while the cooling demand is met by processing seawater or recovering cooling water
 725 from the processing plant.

726 At first, the stream temperatures are corrected (T^*) considering individual temperature differences
 727 ($\Delta T_{\min}/2$) of 2 K, 4 K and 8 K for phase-changing, liquid and gas streams.

$$T_{\text{hot}}^* = \left(T_{\text{hot}} - \frac{\Delta T_{\min}}{2} \right) \quad (\text{B.1})$$

$$T_{\text{cold}}^* = \left(T_{\text{cold}} + \frac{\Delta T_{\min}}{2} \right) \quad (\text{B.2})$$

728 The resulting mass and energy balances are established for the process subsections based on the process
 729 models. The aim is to evaluate how their energy requirements can be satisfied, by promoting internal heat
 730 recovery and minimising the operating costs associated with the external utilities. The objective function
 731 of this optimisation problem can be written as:

$$\min \sum_{s=1}^{N_s} f_s \cdot \dot{C}_{O,s} + c_{e+} \dot{W}_g^+ - c_{e-} \dot{W}_g^- \quad (\text{B.3})$$

732 with:

733 N_s , the number of subsystems, processes and utilities s ;

734 f_s , the level of utilisation of the process or sub-system s ;

735 $\dot{C}_{O,s}$, the operating costs associated with the sub-system s ;

736 c_{e+} and c_{e-} , the purchase and sales prices for electricity;

737 \dot{W}_g^+ and \dot{W}_g^- , the imported and exported electricity from and to the grid.

738 The multiplication factor f_s represents the usage rate of the sub-system s . It is equal to 1 if the sub-
 739 system of interest is a process unit, i.e. a unit with a determined flow-rate and energy demand, and is
 740 variable in a certain range if it is a utility unit, i.e. a unit with variable flow-rate (e.g. seawater, with a flow
 741 rate that can be varied by the operator to match the system cooling demand).

742 Each utility unit is therefore defined by a minimum ($f_{\min,s}$) and maximum usage rate ($f_{\max,s}$), which
 743 correspond to the lower and upper usage bounds. For example, for seawater, the maximum usage rate on

744 an offshore platform is determined by the maximum amount of water that can be processed in the seawater
 745 lift pumps. Moreover, when comparing and selecting different utilities, such as air and seawater for cooling
 746 purposes, an optimal solution may be to use only seawater, and an integer variable y_s is added in the MILP
 747 problem to account for these on/off aspects. In this example, the activation variable would be equal to 1
 748 for seawater and 0 for air.

$$\forall s \in [1, n_s], y_s \in \{0, 1\}, f_{\min, s} y_s \leq f_s \leq f_{\max, s} y_s \quad (\text{B.4})$$

749 This objective function is subject to process and thermodynamic limitations, such as the mass and energy
 750 balances, as well as the heat cascade constraints:

- 751 • heat balance for each temperature interval k :

$$\forall k \in [1, N_k], \sum_{s=1}^{N_s} f_s \cdot \left(\sum_{h_k=1}^{N_{s_{h,k}}} \dot{Q}_{N_{s_{h,k}}} - \sum_{c_k=1}^{N_{s_{c,k}}} \dot{Q}_{N_{s_{c,k}}} \right) + \dot{R}_{k+1} - \dot{R}_k = 0 \quad (\text{B.5})$$

- overall heat balance:

$$\dot{R}_1 = 0, \dot{R}_{N_k+1} = 0, \forall k \in [2, N_k] \dot{R}_k \geq 0 \quad (\text{B.6})$$

- electricity consumption:

$$\sum_{s=1}^{N_s} f_s \dot{W}_s^- - \dot{W}_{\text{aux}}^+ + \varepsilon_{g,s} \dot{W}_g^+ \leq 0 \quad (\text{B.7})$$

- electricity export:

$$\sum_{s=1}^{N_s} f_s \dot{W}_s^- - \dot{W}_{\text{aux}}^+ + \varepsilon_{g,s} \dot{W}_g^+ - \frac{\dot{W}_g^+}{\varepsilon_{s,g}} = 0 \quad (\text{B.8})$$

- overall electricity balance:

$$\dot{W}^+ \leq 0, \dot{W}^- \leq 0 \quad (\text{B.9})$$

752 with:

753 N_k , the number of temperature intervals k ;

754 \dot{W}_s^- , the net power production of the sub-system s in reference conditions;

755 \dot{W}_{aux}^+ , the auxiliary power consumption on-site, not assigned to any sub-system s ;

756 $\varepsilon_{g,s}$ and $\varepsilon_{s,g}$ the conversion efficiency from the grid to the sub-system s , and from the sub-system s to
 757 the grid;

758 $\dot{Q}_{N_{s_{h,k}}}$ and $\dot{Q}_{N_{s_{c,k}}}$, the thermal loads associated with the hot stream h and cold stream c in the
 759 temperature interval k and sub-system s ;

760 \dot{R}_k , the residual heat load from the temperature interval k , cascaded to the lower one $k + 1$.

761 These constraints ensure that the problem is sound from a thermodynamic point of view, and that the
 762 utilities that are the most interesting from an economic perspective are selected.

763 **Appendix C. Economic evaluation**

764 The retrieval of the mass and energy balances from the process and energy integration models can then be
 765 used for performing preliminary cost estimations, following the method of Turton et al. [18]. The grassroot
 766 costs C_{gr} represent the total investment costs, deduced from the bare module costs C_{bm} and purchased
 767 equipment costs C_{pc} . The total grassroot costs are calculated following these four steps:

1. the purchased-equipment costs of each item C_{pc} are estimated by cost correlations, such as the ones of Turton et al. [18], which have an uncertainty of $\pm 30\%$, or by estimation charts, assuming atmospheric pressure conditions and carbon steel construction:

$$\log_{10} C_{\text{pc}} = k_1 + k_2 \log_{10} A + k_3 (\log_{10} A)^2 \quad (\text{C.1})$$

768 where k_1 , k_2 and k_3 are constants and A is the capacity or size parameter specific to the component
 769 under study (e.g. heat transfer area for heat exchangers).

2. the bare module costs C_{bm}^0 are obtained, adjusting the purchased-equipment costs with pressure (f_p) and material (f_m) factors:

$$C_{\text{bm}}^0 = C_{\text{pc}} (b_1 + b_2 f_m f_p) \quad (\text{C.2})$$

770 where b_1 and b_2 are constants. In some cases, these correlations should be adapted to include design-
 771 type and temperature factors to correct these base costs.

3. the actualised bare module costs C_{bm} are computed, considering the inflation between the reference year of the cost data and the date of the estimate with the Marshall Swift Indexes:

$$C_{\text{bm}} = C_{\text{bm}}^0 \left(\frac{\text{MSI}}{\text{MSI}^0} \right) \quad (\text{C.3})$$

4. the grassroot costs C_{gr} , i.e. the total investment costs when installing the equipment items on a new production site, are deduced from:

$$C_{\text{gr}} = (1 + \alpha_1) \sum_i C_{\text{bm},i} + \alpha_2 \sum_i C_{\text{bm},i}^0 \quad (\text{C.4})$$

772 where the factor α_1 ($\simeq 0.18$), which depends on the process conditions, accounts for the contingencies
 773 ($\simeq 0.15$) and fees ($\simeq 0.03$), and the factor α_2 ($\simeq 0.35$), which is independent of the process operation,
 774 accounts for the auxiliary facilities and site development.

775 The capacity factor and values of the coefficients k used in this work to calculate the purchased equipment
776 costs for the most important equipments are the following:

- 777 • for centrifugal pumps with electric drives, the capacity factor is the shaft power in kW, and k_1 , k_2 and
778 k_3 are 3.5793, 0.3208 and 0.02850;
- 779 • for centrifugal compressors, the capacity factor is the fluid power in kW, and k_1 , k_2 and k_3 are 2.9945,
780 0.9542 and 0;
- 781 • for axial gas turbines, the capacity factor is the shaft power in kW, and k_1 , k_2 and k_3 are 3.5137,
782 0.5888 and 0;
- 783 • for vertical vessels such as scrubbers, the capacity factor is the height in m, and k_1 , k_2 and k_3 are
784 3.6237, 0.5262 and 0.2146;
- 785 • for horizontal vessels such as 2- and 3-phase separators, the capacity factor is the length in m, and k_1 ,
786 k_2 and k_3 are 3.3592, 0.5905 and 0.1106.

787 The vessel costs are directly related to their dimensions, which are themselves deduced from the volume
788 flowrate of the feed flow. A maximum diameter of 3 m, height of 4 m and height to diameter ratio of 4/1 is
789 considered for vertical vessels.

790 The column costs are calculated with the correlations presented in Ulrich et al. [51]

791 **Appendix D. Platform information**

792 The offshore platform investigated in this work (Figure D.21) has been in production for more than 20
793 years and processes oil, gas and water. The initial hydrostatic pressure and reservoir temperature were about
794 165 bar and 71 °C [52]. Petroleum and subsurface water enter the platform system through the *production*
795 *manifolds*, in which the streams from the several wells are depressurised and mixed. The pressures at the
796 inlet of the production manifold range between 8 and 46 bar, and the average temperatures are about 63
797 °C.

798 Oil, gas and water are then dissociated by gravity in two *separation* stages, the first ones operating at
799 the production manifold conditions, the second one at about 65–75 °C and 1.6–1.8 bar. The recovered gas is
800 first re-compressed to the initial feed pressure, and compressed further to 179–189 bar. Most high-pressure
801 dry gas leaving the final compression stage is used for gas lift. The remaining gas is partly exported to the
802 coast and partly consumed in gas turbines for local power generation. The crude oil is stored and exported
803 to the shore, while the produced water is treated and discharged into the sea. The gas turbines amount to
804 five, of which three provide the power required in the processing plant, and the remaining two are run in
805 case of water injection. Additional process data is provided in Nguyen et al. [5].

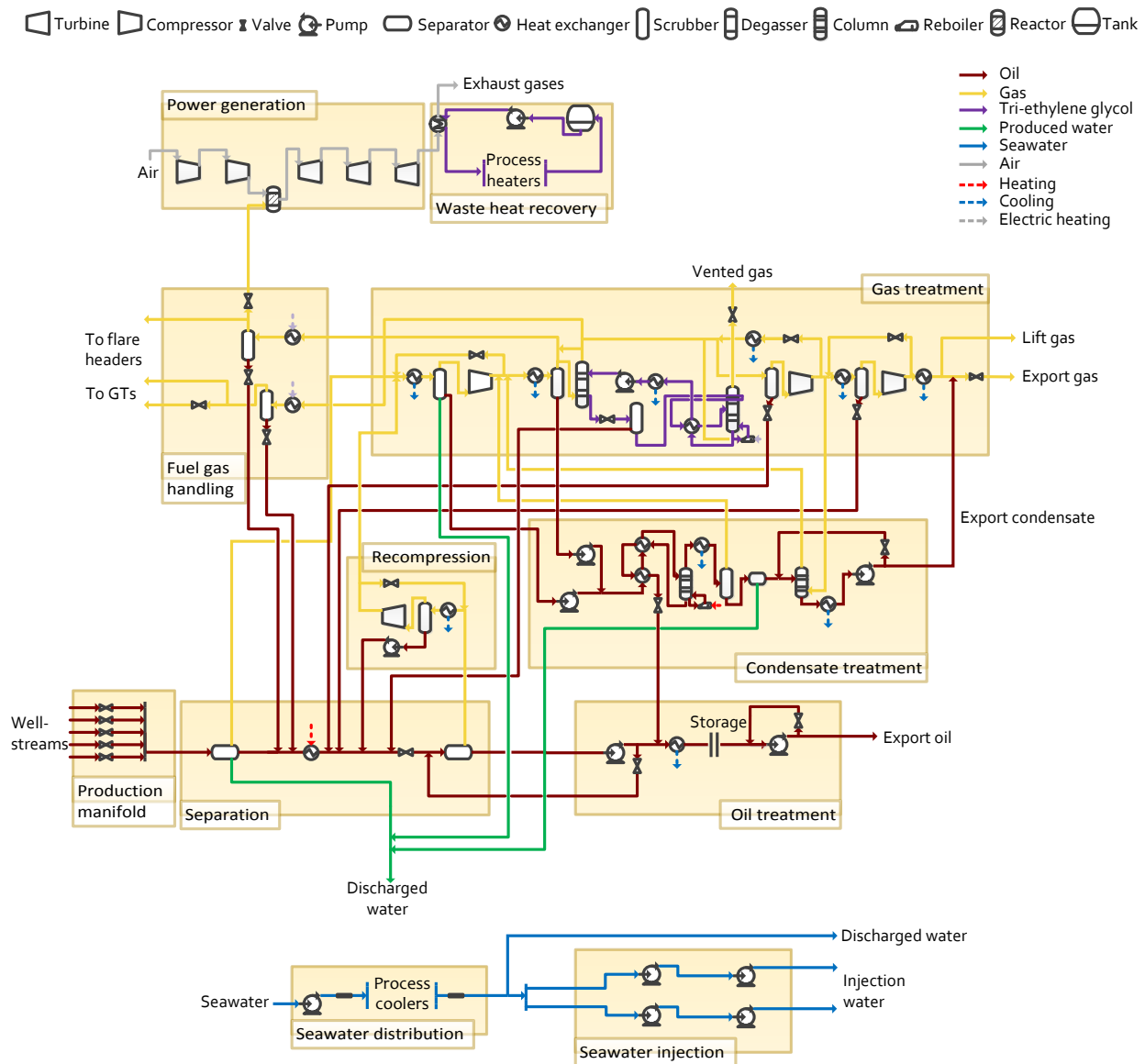


Figure D.21: Process flow diagram of the Norwegian Sea offshore platform investigated in this work, based on input by Norske Shell A/S and data collected from the literature. For ease of reading, only the most significant recycling loops are drawn, and only one gas turbine is shown. The addition of chemicals such as glycol and methanol to prevent hydrate formation is not indicated, and redundant components that are run in parallel (e.g. pumps and separators) are merged into one.

806 **Appendix E. Multi-objective optimisation**

807 The decision variables of the multi-objective optimisation problem are presented, in the case of the CO₂-
808 capture processes, as follows. They correspond to the selection and configuration of the CO₂-capture unit
809 (e.g. equipment sizes) and amount to 14 in the case of a chemical absorption unit with MEA (Table E.2), 5
810 in the case of a physical absorption module with MeOH (Table E.3), 5 with DEPG (Table E.4), 6 with TEA
811 (Table E.5). 13 other decision variables (Table E.6) are related to the design of the natural gas pre-processing
812 and of the associated utilities in the CO₂ pre-combustion path.

813 **Appendix F. Sensitivity analyses**

814 The sensitivity analyses of the electricity production (COE) and avoidance (CAC) costs of an offshore
815 platform to different factors such as the CO₂-tax are presented as follows.

816 *Oil price.* These variations of the oil prices may have an impact on the profitability of waste heat recovery
817 and CO₂-capture plants, as these could result in variations of the gas prices. In the case that Norwegian oil
818 and gas producers sign supply agreements based on oil indexation, instead of spot prices, the installation
819 of new equipments will be directly impacted by the oil price fluctuations on the market. Such contracts are
820 agreed on a 15-20 year basis, implying that the profitability of these additional processes will be influenced
821 for more than a half of their lifetime. In the last decade, increasing oil prices pushed towards extended
822 exploitation of oil fields, and the use of CO₂ for enhanced oil recovery could therefore be favoured. On the
823 contrary, decreasing oil prices would discourage the use of costly oil recovery techniques, and the integration
824 of CCS processes would be unfavourable without a rise of the carbon taxes.

825 *CO₂-tax.* Similarly, the taxation on CO₂ depends on the industrial sector and country of application: it is at
826 the moment about \$ 65 per tonne of CO₂ in the Norwegian petroleum sector, and it will most likely rank as
827 one of the highest CO₂-taxes in Europe. In Europe, the foreseen values of the CO₂-taxes range between \$ 20
828 and \$ 40 in the near-future and between \$ 65 and \$ 75 in the long-term. More precisely, the cost projections
829 suggest that carbon tax prices will rise only moderately in the coming decade, reaching about \$ 35 in 2030,
830 as a consequence of the financial crises and surplus of allowances and international credits. These costs
831 may then increase significantly, up to \$ 55 in 2050, to support low-carbon technologies and energy efficiency
832 measures [45,53].

833 As suggested by the first sensitivity analysis, the CO₂-tax also has a strong impact on the electricity
834 and CO₂-avoidance costs (Figure F.22). For a natural gas price of 8.08 \$/GJ, which is in the range of
835 the production costs estimated by the oil companies operating petroleum fields in Norway, the break-even
836 values are about 35 and 100 \$/tCO₂ for the first and second configurations, respectively. This large difference

837 between the break-even values illustrates that the implementation of CO₂-capture processes may be feasible
 838 or economically profitable only over a certain range of CO₂-capture potentials.

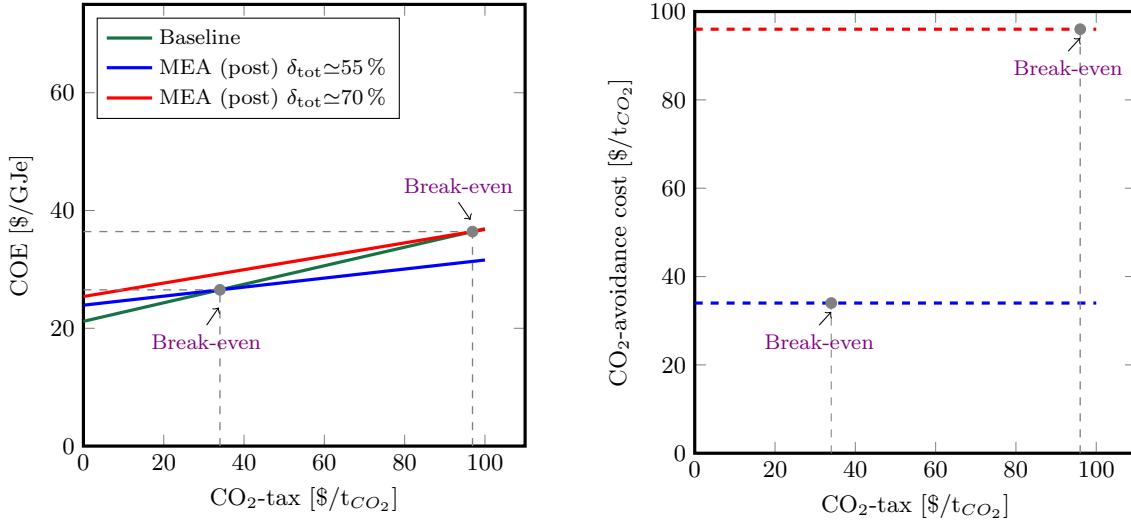


Figure F.22: Sensitivity of the electricity production (COE) and avoidance (CAC) costs of an offshore platform with and without integration of post-combustion CO₂-capture to the carbon taxation.

839 *CO₂-injection wells.* The present work assumes that CO₂ could and would be injected into the reservoir for
 840 possible enhanced oil recovery or long-term storage. Depending on the field, existing wells may be used if not
 841 producing any longer, or new ones may have to be built. In that case, these expenses are site-specific and the
 842 cost estimates vary widely from one study to another (Figure F.23). They consist of the costs for building
 843 the injection wells, cementing the wells, installing corrosion resistant casing, drilling and constructing the
 844 pipelines, and have been estimated to about 15 M\$ in the case of the Sleipner platform. The electricity
 845 and CO₂-avoidance costs of the process configurations with CCS are highly sensitive to the site-specific
 846 costs, and those sensitivity analyses suggest that an offshore platform with a high degree of CO₂-reduction
 847 may only be economically viable, in the future, with a further increase of the CO₂-tax, and unlikely for
 848 all petroleum fields. The economic profitability of CO₂-capture and storage on offshore fields is expected
 849 to decrease sharply if CO₂ cannot be injected on-site and has to be transported to the shore or to other
 850 platforms, as suggested by feasibility studies on that topic.

851 Appendix G. Environmental impacts

852 The benefits of CO₂-capture processes combined with steam Rankine cycles can also be drawn with
 853 regards to the acidification and eutrophication potentials, as the on-site NO_x emissions are decreased by
 854 about 25%. Although the chemical absorption process induces an energy penalty, the overall natural gas
 855 consumption for the platforms on which carbon capture is implemented is decreased, because this results in

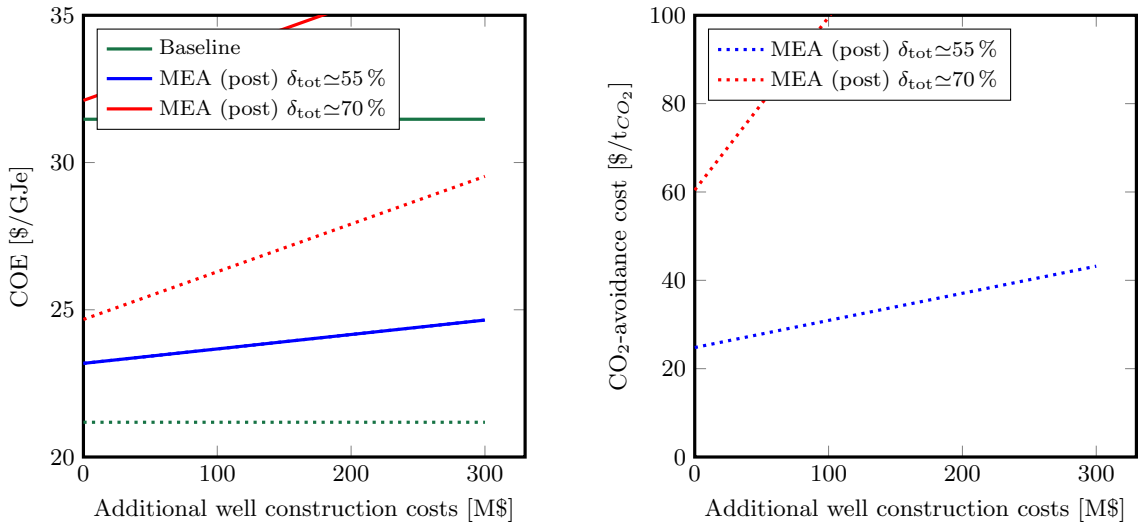


Figure F.23: Sensitivity of the electricity production (COE) and avoidance (CAC) costs of an offshore platform with and without integration of CO₂-capture to the additional construction and capital costs.

856 a smaller depletion of the gas resources. Similarly, the impact on human health is reduced because of the
 857 smaller emissions of CO₂, nitrous oxides and pollutants, and this is illustrated with both the Ecoindicator
 858 99 [54] and Impact 2002+ methods (Figure G.24 and Figure G.25).

859 For the climate change impact category, the main contributions are caused by the emissions of fossil CO₂
 860 from the gas turbines ($\approx 85\%$). The emissions associated with the manufacturing and installation phases of
 861 the system components play the major role ($\approx 85\%$) for the ecosystem impact, and the greatest impact on
 862 human health derives from the NO_x-emissions ($\approx 60\%$). The same conclusions can be drawn when applying
 863 the Ecoindicator 99 approach. One of the main differences is that the climate change impacts are considered
 864 within the human health category, and the impact decrease is more marked as it is affected by the reductions
 865 of both CO₂ and nitrous oxide emissions.

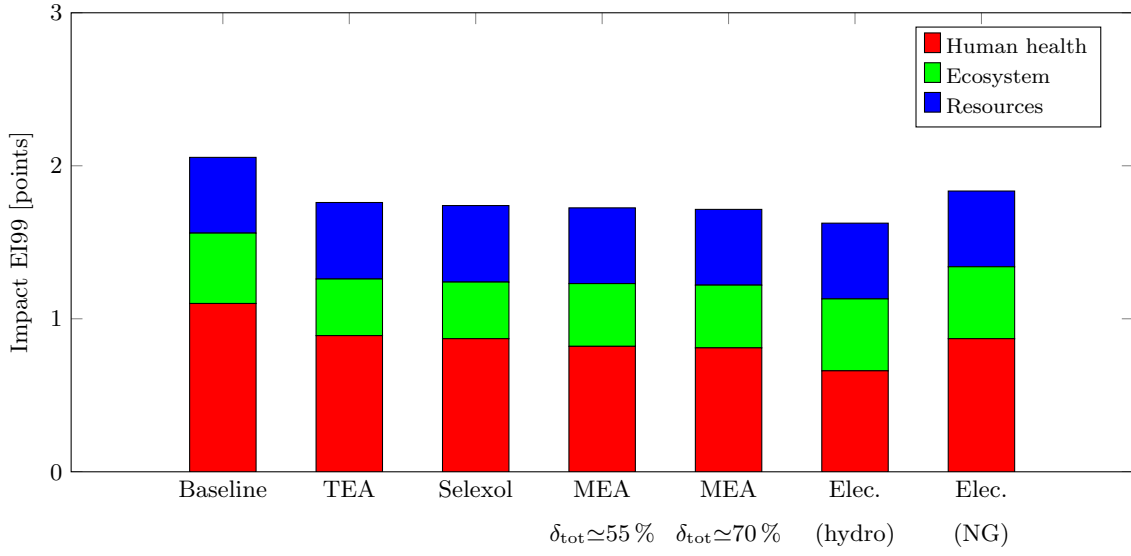


Figure G.24: Comparison of the scenarios with and without CO₂-capture: EI99 (hierarchical approach).

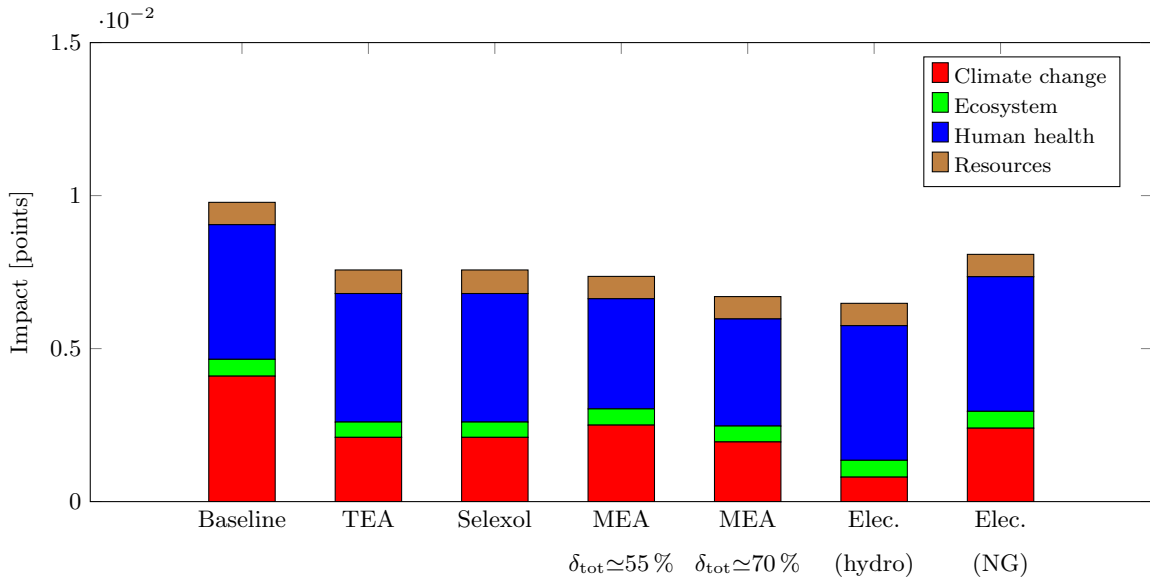


Figure G.25: Comparison of the scenarios with and without CO₂-capture: Impact 2002+.

Nomenclature

Abbreviations

+ Material-/Energy-flow entering the system
 - Material-/Energy-flow leaving the system

CAC Carbon avoidance cost
 CC Combined cycle
 CCS Carbon capture and storage
 COE Cost of electricity
 DNA Dynamic Network Analysis
 EOS Equation of State
 eq equivalent
 FG Fuel gas
 FU Functional Unit
 GE Exported gas
 GWP Global Warming Potential
 IPCC Intergovernmental Panel on Climate Change
 LCA Life Cycle Assessment
 LCI Life Cycle Inventory
 MILP Mixed Integer Linear Programming
 MINLP Mixed Integer Non-Linear Programming
 MOO Multi-Objective Optimisation
 MSI Marshall Swift Index
 NG Natural gas
 OE Exported oil
 ORC Organic Rankine cycle
 PR Peng-Robinson
 ref reference
 SRC Steam Rankine cycle

Greek letters

α_1 Contingencies factor (grassroot costs)
 α_2 Auxiliary factor (grassroot costs)

Δh^0 Heating value, kJ/kg
 Δk^0 Specific exergy, kJ/kg
 δ Relative variation, %
 η Energy efficiency, %
 σ Energy intensity, %
 I Environmental impact, kg/FU

Roman letters

\dot{m} Mass flow, kg/s or t/h
 \dot{Q} Heat, kW
 \dot{R}_k Residual heat load, kW
 \dot{W} Work, kW
 \dot{W}_{aux}^+ Auxiliary power consumption on-site, kW
 \dot{W}_g^+ Imported electricity from the grid, kW
 \dot{W}_s^- Net power production (reference conditions), kW
 \dot{W}_g^- Exported electricity to the grid, kW
 A Capacity or size parameter
 C_{bm}^0 Bare module costs, \$
 C_{gr}^0 Grassroot costs, \$
 C_{op} Operating costs, \$
 C_{pe} Purchased-equipment costs, \$
 c_{e+} Purchase price of electricity, \$/kW
 c_{e-} Sales price of electricity, \$/kW
 f_m Material factor (bare module costs)
 f_p Pressure factor (bare module costs)
 f_s Level of utilisation of a process or sub-system
 N_k Number of temperature intervals
 N_s Number of subsystems, processes and utilities
 p Pressure, bar
 T Temperature, °C or K
 T^* Corrected temperature, °C or K

Table E.2: Set of the master decision variables used in the multi-objective optimisation of the CO₂-capture unit based on chemical absorption with an aqueous solution of monoethanolamine.

Variable	Type	Unit	Range
Lean solvent CO ₂ loading	continuous	kmol/kmol	[0.18–0.25]
Rich solvent CO ₂ loading	continuous	kmol/kmol	[0.4–0.5]
Split fraction	continuous	-	[0;1]
Rich solvent preheat temperature	continuous	°C	[95–105]
Rich solvent reheat temperature	continuous	°C	[115–125]
LP stripper pressure	continuous	bar	[1.7–2.1]
HP/LP pressure ratio	continuous	-	[1–1.5]
Number stages absorber	continuous	-	[10–17]
Number stages HP stripper	continuous	-	[8–15]
Number stages LP stripper	continuous	-	[6–10]
Absorber diameter	continuous	m	[6–12]
LP stripper diameter	continuous	m	[2–5]
HP stripper diameter	continuous	m	[3–6]
MEA concentration (solvent)	continuous	wt %	[30–40]

Table E.3: Set of the master decision variables used in the multi-objective optimisation of the CO₂-capture unit based on physical absorption with methanol.

Variable	Type	Unit	Range
MeOH/CO ₂ ratio	continuous	kmol/kmol	[10–15]
Absorber temperature	continuous	°C	[-70–0]
Absorber pressure	continuous	bar	[15–60]
Regenerator pressure	continuous	bar	[1–10]
Regenerator temperature	continuous	°C	[20–100]

Table E.4: Set of the master decision variables used in the multi-objective optimisation of the CO₂-capture unit based on physical absorption with DEPG.

Variable	Type	Unit	Range
DEPG/CO ₂ ratio	continuous	kg/kg	[8–14]
Absorber temperature	continuous	°C	[-18–173]
Absorber pressure	continuous	bar	[10–60]
Regenerator pressure	continuous	bar	[1–10]
Regenerator temperature	continuous	°C	[25–100]

Table E.5: Set of the master decision variables used in the multi-objective optimisation of the CO₂-capture unit based on chemical absorption with an aqueous solution of TEA.

Variable	Type	Unit	Range
TEA concentration	continuous	kg/kg	[0.25–0.40]
H ₂ -TEA ratio	continuous	kg/kg	[0.035–0.055]
Absorber temperature	continuous	°C	[20–45]
Absorber pressure	continuous	bar	[15–30]
Regeneration pressure	continuous	bar	[1–130]
Regeneration temperature	continuous	°C	[25–120]

Table E.6: Set of the master decision variables used in the multi-objective optimisation of the pre-combustion CO₂-capture path.

Variable	Type	Unit	Range
SMR temperature	continuous	°C	[450–950]
ATR temperature	continuous	°C	[500–950]
Reforming pressure	continuous	bar	[1–30]
Air-to-carbon ratio (ATR)	continuous	kg/kg	[3–4.5]
Steam-to-carbon ratio (ATR)	continuous	kg/kg	[1.5–6]
High-temperature water-gas-shift	continuous	°C	[250–420]
Low-temperature water-gas-shift	continuous	°C	[150–250]
Water-gas-shift pressure	continuous	bar	[1–30]
CO ₂ -capture unit	discrete	-	{0 – 3}
Oxygen-to-hydrogen ratio	continuous	kmol/kmol	[0.4–0.7]
H ₂ -turbine combustion pressure	continuous	bar	[5–50]
Exhaust gas temperature	continuous	°C	[100–200]
Low-pressure level (refrigeration cycle)	continuous	bar	[0.1–5]