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# HYBRID TANDEM CATALYTIC CONVERSION PROCESS TOWARDS HIGHER OXYGENATE EFUELS

E-TANDEM

## E-TANDEM - Deliverable report

## D5.5 – Techno-economic assessment of e-fuel production concept

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Project summary

The E-TANDEM Horizon Europe project aims to unlock a green and efficient production route for a new higher oxygenate diesel-like fuel for the marine and heavy-duty transport sectors, from $CO_2$ as the only carbon source, and renewable electricity as the only energy source. The process consists of a solid oxide co-electrolysis step to convert $CO_2$ and $H_2O$ into syngas, followed by a tandem reactor in which the e-syngas is converted to higher oxygenates. The E-TANDEM project aims to demonstrate the new e-fuel production route at bench scale and assesses its commercialization perspective through techno-economic and life cycle analyses.

# Publishable summary

The E-TANDEM project aims to develop a direct and effective conversion pathway to produce higher oxygenate fuels (alcohols and ethers) for the maritime and heavy-duty transport sectors from $CO_2$ as the only carbon source and renewable electricity as the only energy source. The process design combines syngas generation via co-electrolysis of $CO_2$ and $H_2O$, e-syngas conversion through tandem Fischer-Tropsch and reductive hydroformylation, downstream product purification, and higher alcohol dehydration to yield the final higher ether product. In this deliverable, the techno-economic potential of the proposed e-fuel production route is assessed for a 1 MW electrolyzer demonstration facility, as simulated by project partner AVL in WP4. In addition, detailed geographical and technological sensitivity analyses are performed to select suitable e-fuel production locations and to identify the main cost drivers for the E-TANDEM e-fuel production concept for the 1 MW demonstration scale, as well as for potential future larger scale plant sizes (100-1,500 MW).

The capital expenditures (CAPEX) and operating expenditures (OPEX) of the 1 MW E-TANDEM demonstration facility are estimated based on (i) mass- and energy balances from project partner AVL in WP4 and (ii) detailed designs of the required process equipment. The total direct inside battery limit (ISBL) investment of the 1 MW facility is around 10 M€, with reactors (co-electrolysis, tandem reactor), syngas compressors, and heat exchangers as the main cost drivers of the proposed process. Besides, the total production costs are around 6 $€\cdot L_{fuel}^{-1}$ for the 1 MW plant, mainly due to high personnel, maintenance, and depreciation charges at this small production scale. The E-TANDEM e-fuel production costs can be reduced to less than 2 $€\cdot L_{fuel}^{-1}$ by increasing the electrolyzer capacity from 1 MW to ≥100 MW. For the 1 MW facility, the selection of suitable production locations is governed by the emission intensity of the local electricity grid, which should be below 100 $g_{CO2}\cdot kWh^{-1}$. On the contrary, for larger electrolyzer capacities (≥100 MW) suitable production locations should possess a combination of green and low-cost electricity (<0.05 $€\cdot kWh^{-1}$), as fuel production costs are dominated by local electricity costs at those larger scales. Under current market conditions, the E-TANDEM e-fuel production costs (1.5-2.0 $€\cdot L_{fuel}^{-1}$) are forecasted to be higher than biodiesel and fossil diesel prices (<1.0 $€\cdot L_{fuel}^{-1}$). Consequently, regulations on emission taxations and/or e-fuel subsidies or quotas are needed to commercialize E-TANDEM e-fuels in the current stage. As a first strategy, the ether fuel products can be blended with conventional diesel-type fuels, e.g., marine gas oil (MGO), to mitigate fuel costs and accelerate market uptake.

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# Abbreviations & Definitions

| Abbreviation | Explanation |
|---|---|
| BoP | Balance of plant |
| CAPEX | Capital expenditures |
| CEPCI | Chemical engineering plant cost index |
| Co-SOE | Solid Oxide co-Electrolysis |
| DAC | Direct air capture |
| FT | Fischer-Tropsch |
| GHG | Greenhouse gas |
| HOEF | Higher oxygenate e-fuel |
| HVO | Hydrotreated vegetable oil |
| ISBL | Inside battery limits |
| MGO | Marine gas oil |
| NG | Natural gas |
| OPEX | Operating expenditures |
| OSBL | Outside battery limits |
| OSN | Organic solvent nanofiltration |
| PSA | Pressure swing adsorption |
| RHF | Reductive hydroformylation |
| S/F | Solvent-to-feed ratio |
| SPD | Slurry phase distillate |
| TSA | Temperature swing adsorption |

# 1 Introduction

The European Green Deal aims to ensure climate neutrality in Europe by 2050 [1]. One of the main objectives of this Green Deal is to drastically reduce the greenhouse gas (GHG) emissions of the transportation sector, as this sector accounts for 30% of the total EU GHG emissions [2]. In part, this can be accomplished through electrification of road transport. However, for other sectors, like the maritime transport, alternative production routes to produce energy dense liquid fuels need to be developed to achieve net-zero emissions. E-TANDEM's ambition is therefore to develop a direct and effective conversion pathway to produce higher oxygenate fuels (higher alcohols and ethers) from renewable carbon resources. Specifically, E-TANDEM combines solid oxide co-electrolysis of $CO_2$ and $H_2O$ (co-SOE) with Fischer-Tropsch synthesis and alcohol dehydration. Co-electrolysis of $CO_2$ and $H_2O$ using renewable energy represents an attractive route towards e-syngas [3,4]. This e-syngas can, in turn, be applied to generate higher alcohols via a tandem reaction consisting of (i) Fischer-Tropsch synthesis and (ii) reductive hydroformylation [5]. After alcohol product purification, the higher alcohols are further converted into higher ethers via a dehydration step.

This report belongs to work package 5: *Socio-economic and LCA assessment, dissemination, communication & exploitation strategies*. Within this work package, deliverable 5.5 corresponds to Task 5.1: *Techno-economics of e-fuel production concept*. The main objective of this task is to assess the new e-fuel production concept from a techno-economic perspective and evaluate its conversion efficiency and competitiveness against alternative fuels, e.g., fossil-based diesel, biodiesel, and alternative e-fuels. Additionally, a sensitivity analysis is applied to examine the overall economic feasibility of the e-fuel production concept. The production scale of the demonstration plant of which the techno-economic aspects are assessed is based on a 1 MW co-electrolysis capacity.

The major inputs for the techno-economic assessment are the mass- and energy balances obtained from the process simulations performed by project partner AVL in Task 4.3. The capital expenditures (CAPEX) of the e-fuel production concept are estimated via a bottom-up approach in which the total capital investment is estimated as a summation of the investment per piece of equipment needed. To this end, each individual piece of equipment is designed using information from the mass- and energy balances (WP4) and using experimental performance information of catalysts, membranes, etc., from WP2 and WP4. In addition, the calculated total e-fuel production costs comprise variable production costs (e.g., utilities, raw materials, consumables) and fixed production costs (e.g., personnel, maintenance). A technological sensitivity analysis is carried out using the main cost drivers to assess the financial vulnerability of the E-TANDEM demonstration plant. Additionally, a geographical sensitivity analysis is performed to identify suitable production locations for the 1 MW demonstration facility, as well as for potential future larger scale (100-1,500 MW) e-fuel production facilities.

# 2 Methods

## 2.1 Background

The methods applied to assess the techno-economics of the e-fuel production concept proposed in E-TANDEM can be subdivided into four parts:

- **Equipment design and purchased equipment cost estimation**. Based on the mass- and energy balances retrieved from the process simulations performed in WP 4 and the experimental performance information from WP2 and WP4, the individual pieces of equipment are designed. Directly coupled with the equipment design, the purchased equipment costs of each individual piece of equipment are estimated using methods described in detail in **Section 2.2.1**.
- **Capital expenditures – factorial method**. The total capital expenditures of the e-fuel production concept are determined from the individual equipment costs using a factorial method to translate purchased equipment costs to installed equipment costs, and to normalize all costs to Western Europe, 2025.
- **Operating expenditures**. The operating expenditures are estimated based on the mass- and energy balances and the equipment design. They comprise variable production costs (e.g., raw materials and consumables) and fixed production costs (e.g., personnel and process maintenance).
- **Sensitivity analysis.** A technological sensitivity analysis is conducted with the main cost drivers to examine the financial vulnerability of the E-TANDEM production concept. Besides, a geographical sensitivity analysis is carried out to identify suitable e-fuel production locations.

## 2.2 Procedures

### 2.2.1 Capital expenditures (CAPEX)

**Co-SOE system**

The targeted capacity of the co-SOE system in E-TANDEM is 1 MW. In the base case scenario, the electrolyzer is operated with a current density of 0.31 $A\cdot cm^{-2}$, corresponding to the current density experimentally employed by DTU, at single-cell level, in WP2 (see D2.3). The required electrolyzer stack area for this current density is around 116 $m^2$. The electrolyzer stack design is based on a report on industrial guidelines for solid electrolyzer technologies by the Institute for Sustainable Process Technology [6]. The available area per stack layer is estimated at 800 $cm^2$, and the number of layers per stack at 60 (i.e., 4.8 $m^2$ per stack) [6]. Consequently, the required number of stacks is 25 and the capacity per stack is about 40 kW.

The total electrolyzer costs are subdivided into electrolyzer stack costs and balance of plant (BoP) costs. The electrolyzer stack costs comprise costs related to layer materials, interconnects, gas manifolds, weld connectors, sealing rings, pressure vessel, and stack assembly. Here, the layer materials are based on the metal-supported cell design experimentally used by project partner DTU in WP2. The BoP costs are calculated based on the heat-integrated design of the co-SOE system, as developed by project partner AVL in WP4. Herein, the BoP consists of an air filter, an air blower, five heat exchangers, two oxidation catalyst units, a recirculation ejector, and an AC/DC converter (rectifier). Note that the water evaporator costs are calculated separately in this study and are, hence, not included in the BoP investment.

**Tandem reactor and etherification reactor**

The conversion of syngas into higher alcohols via a combination of Fischer-Tropsch (FT) and reductive hydroformylation (RHF) occurs in the tandem reactor, which is operated at 180 bar(a) and 200 °C. The tandem reactor is designed based on the industrially applied Sasol slurry phase distillate (SPD) reactor. This Sasol SPD reactor is a bubble column for FT synthesis, in which syngas is bubbled through a slurry phase [7], which thereby strongly resembles the experimental tandem reactor applied in WP2. The tandem reactor is dimensioned by downscaling the industrial Sasol SPD reactor, considering equal syngas linear velocity and slurry volume fraction, in which it is implicitly assumed that hydrodynamics and reaction phenomena are translatable to smaller reactor capacities. The etherification reactor is dimensioned based on the corresponding reactor performance measured by project partner MPG in WP2, considering a constant ether production rate over the required reaction time of 72 h to achieve near complete alcohol conversion. The resulting tandem reactor diameter and height are 0.7 m and 4.0 m, while the etherification reactor diameter and height are 0.9 m and 3.6 m. The capital investment of the tandem reactor is estimated from the required steel costs, by regarding this reactor as a vertical pressure vessel, according to the design guidelines in ref. [8], as it is operated at high pressures. Additional costs related to, e.g., reactor lining material and cooling coils are included by means of a high installation factor. On the contrary, the etherification reactor investment is calculated by considering it as a jacketed, agitated, and glass-lined reactor [8] instead of a pressure vessel, as the etherification step is carried out at atmospheric pressure.

**Adsorption columns**

The syngas stream from the co-SOE system contains $H_2O$ and $CO_2$ impurities that must be removed before feeding the syngas into the tandem reactor. This is accomplished by a combination of a temperature swing adsorption (TSA) and a pressure swing adsorption (PSA) step. In the TSA step, a molecular sieve 3A zeolite is applied as desiccant to dry the syngas stream [9]. Subsequently, an activated carbon adsorbent is used to remove the $CO_2$ impurities in the PSA step [10,11]. The TSA $H_2O$ removal step is employed first to avoid interference of $H_2O$ adsorption with the adsorption of $CO_2$ on the activated carbon adsorbent. The TSA step is carried out in a two-column system with complete $H_2O$ removal. The temperature swings are between 40 °C (adsorption) and 180 °C (desorption). The adsorption takes place at 20 bar(a), while a syngas purge is used in the desorption step to facilitate $H_2O$ desorption. Similar to the TSA step, a two-column PSA system is used to achieve the desired $CO_2$ recovery (>90%) [10]. Pressure swings from an adsorption pressure of 20 bar(a) to a desorption pressure of 3 bar(a) are applied in the adsorption/desorption cycle at 40 °C. The required adsorbent amounts are calculated from the respective mass flows of $H_2O$ and $CO_2$ to be removed from the syngas feed, and the adsorption capacities of the concerning adsorbents [9,11]. The dimensions of the adsorption columns are estimated from the required adsorbent amounts and corresponding densities, considering 80% of the column volume occupied by adsorbent, and a column aspect ratio of 4. Similar to the tandem reactor, the column investment is calculated from the required steel mass, considering the adsorption columns as vertical pressure vessels [8].

**Fractional distillation tower**

The fractional distillation tower has two main functions: (i) separate the waxes, naphtha and light alcohols ($C_{4-}$) from the higher alcohol products, and (ii) fractionate the higher alcohol products ($C_{5+}$)

into three product fractions based on carbon chain length. The column has two feed streams, originating from the liquefied gaseous outlet of the tandem reactor and the unretained liquid outlet of the tandem reactor, respectively. This fractional distillation tower comprises five column sections that are individually dimensioned. The Sulzer BXPlus Gauze packing is selected as a suitable column internal for this column, as this packing type is recommended to separate fatty alcohols [12]. The fractional distillation tower investment consists of (i) the vessel steel investment, and (ii) the investment in column internals.

### Extraction columns

The paraffinic byproducts are separated from the targeted higher alcohol product fractions using liquid-liquid extraction. Here, mixtures of acetonitrile, water, and methanol are selected as extractant to isolate the higher alcohol products from the paraffins [13]. The process requires three extraction columns, as the composition of the extractant mixture differs per higher alcohol product fraction. The columns are operated at atmospheric pressure and ambient temperature, and the Sulzer extraction packing is considered suitable for this application [14]. Similar to the distillation column investment, the capital investment of the extraction columns consists of the vessel investment plus the expenditures on column internals.

### Azeotropic distillation

The higher alcohol products are separated from the acetonitrile/water/methanol extractant mixtures via azeotropic distillation to recover the extractants and to obtain purified alcohol products. Three azeotropic distillation columns are required to individually recover the extractants. The Sulzer AYPlus DC packing is recommended for separating mixtures of higher alcohols and water [12], and is therefore considered to be suitable for the azeotropic distillation steps targeted here. The column investment is again calculated as a combination of the vessel investment and the column packing investment.

### Vessels

In addition to the distillation and extraction columns designed in detail previously, the demonstration plant contains several vessels for vapor-liquid separation, utility storage (extractants, molecular catalyst), and $CO_2$ buffering. The capital investment in these vessels is estimated based on the required steel costs.

### Retention of molecular catalyst

A combination of a solid and a molecular catalyst is used to convert syngas into higher alcohols in the tandem reactor. While the solid catalyst is mechanically retained inside the tandem reactor through the use of a gauze in the reactor outlet, an organic solvent nanofiltration (OSN) membrane is employed in the liquid reactor outlet to retain and recycle the molecular catalyst. The required area of the OSN membrane is estimated based on the optimal experimental retention performance measured by MPG in WP4, and the anticipated molecular catalyst flow in the E-TANDEM demonstration plant. This resulted in a required membrane area of ca. 12 $m^2$. Industrial OSN modules are available in membrane area sizes up to 32 $m^2$ [15], hence one OSN membrane module is needed for the demonstration plant. The purchased equipment costs of this OSN filtration step are estimated at 32 k€ based on a commercial quotation for such an OSN membrane module.

### Pressure changers

The capital expenditures of the compressors are estimated based on the corresponding driver power [16]. Two reciprocating compression stages pressurize the unconditioned syngas from the co-SOE pressure of 1 bar(a) to the TSA and PSA adsorption pressure of 20 bar(a). Besides, the conditioned syngas stream is pressurized further in a three-stage reciprocating compressor from 20 bar(a) to the tandem reactor pressure of 180 bar(a). Furthermore, a centrifugal tail gas recycle compressor is used to recycle the unconverted syngas. For pumps the capital expenditures consist of the pump itself and the pump driver investment, where all pumps are considered to be single stage centrifugal pumps, and in which the pump drivers are taken as explosion proof motors, according to the design guidelines in ref. [8].

### Heat exchangers

The process contains 43 heat exchangers that are individually dimensioned by means of the required heat exchange area based on the total amount of energy to be exchanged, the driving force for heat exchange, and an estimated overall heat transfer coefficient. The capital investment of the five heat exchangers that are part of the co-SOE process section are included in the BoP investment of the co-SOE system. The CAPEX of 35 of the other 38 heat exchangers is estimated by considering a double pipe heat exchanger configuration. This holds for the reboilers of the fractional distillation and ether distillation towers, and for all coolers, heaters, and condensers [8]. By contrast, electrical reboilers are employed for the three azeotropic distillation columns, because of the high heating demand of these reboilers (>10 kW each). The purchased equipment costs of these electrical reboilers are estimated based on furnace equipment costs correlations by Loh et al. [17]. All heat exchangers are considered to be constructed from stainless steel type 304, because of the risk from corrosive compounds, except for the steam conditioning heat exchangers.

### Influence of the solvent for the molecular catalyst

An organic high-boiling solvent is needed to dissolve and transport the molecular catalyst. To this end, squalane (boiling point: 350 °C at 1 atm) is experimentally used by E-TANDEM partners CSIC and MPG in WP2 and WP4. However, due to its structural complexity, the molecular catalyst and the corresponding squalane solvent were not included in the process simulations in Task 4.3. As a consequence, the stream sizes around the tandem reactor as obtained from the process simulations are underestimated, as they exclude the solvent flow. This specifically holds for the heater of the tandem reactor feed and for the three heat exchangers that are part of the liquid outlet of the tandem reactor. These heat exchangers are not designed including the solvent flow, as this would change the detailed heat exchanger design of the entire process. However, a solvent recycle pump is included in the CAPEX assessment, and the size of the flash drum in the liquid outlet of the tandem reactor is designed including the solvent flow, to incorporate the solvent recirculation requirement in the financial assessment. The size of the tandem reactor itself is calculated based on the syngas flow and is, thereby, independent of the solvent flow.

### Factorial method

The purchased equipment costs per unit operation are multiplied by an installation factor to account for the installation of the concerning equipment into the demonstration plant. This installation factor covers costs related to, e.g., construction material, piping, instrumentation, process control, and electrical [8], and differs per type of equipment. In addition, cost multiplication factors related to the costing reference year (Chemical Engineering Plant Cost Index, CEPCI [8,18]) and the plant location are applied. Here, the plant location factor is estimated to be ca. 1.1 for Western Europe relative to the U.S. Gulf Coast [19].

### 2.2.2 Operating Expenditures (OPEX)

The OPEX assessment is based on a plant operation of 8,000 hours per year. Besides, the applied USD-to-euro exchange rate is: 1 USD = €0.87. Similar to the CAPEX, the OPEX are evaluated for Western Europe, 2025.

**Raw materials**

The carbon dioxide feedstock is considered to originate from direct air capture (DAC). A lower end DAC price of 100 $€\cdot t_{CO2}^{-1}$ is included in this case, corresponding to a future projected DAC price [20]. The fresh water supply is obtained from reverse osmosis desalination, which costs about 2 $USD\cdot m^{-3}$ (i.e., ca. 1.75 $€\cdot t_{H2O}^{-1}$) [21].

**Consumables**

The consumables of the E-TANDEM demonstration plant (1 MW) comprise various catalysts, solvents, membranes, adsorbents, and extractants.

*Tandem reactor*

The required amounts of solid and molecular catalysts are estimated using the maximum higher alcohol productivity experimentally obtained by project partner CSIC in WP2 (D2.2), and the higher alcohol product mass flow in the tandem reactor outlet from the process simulations in WP4. This results in a total Co amount of 99 kg. The experimental cobalt ratio between the solid and the molecular catalyst is about three, implying that ca. 74 kg of Co is needed in the solid catalyst and ca. 25 kg in the molecular catalyst. A Co loading of 25% is taken for the solid catalyst [22], resulting in a total required solid catalyst mass of 296 kg. The market price of the solid catalyst is estimated at 27.44 $€\cdot kg_{cat}^{-1}$ based on Ma et al. [22], normalized to 2025 using the CEPCI. This results in a total solid catalyst value of about 8 k€ in the tandem reactor cycle. The lifetime of the Co-based solid catalyst is estimated at two years [23], yielding fresh solid catalyst costs of about 4 $k€\cdot yr^{-1}$.

The molecular catalyst costs contain costs for the cobalt carbonyl complex ($Co_2(CO)_8$) and the ligand. The cobalt carbonyl precursor is only available in experimental amounts (<100 g) and costs around 10 $€\cdot g^{-1}$ on that scale [24]. It is estimated that the price of this cobalt carbonyl precursor can be reduced to 2 $€\cdot g^{-1}$ when produced at an industrial scale. The commercially available trioctylphosphine ($POct_3$) is selected as a ligand and costs around 15 $kUSD\cdot t^{-1}$ (i.e., 13 $k€\cdot t^{-1}$) in Western Europe [25]. For the molecular catalyst composition experimentally employed in WP2 and WP4, a total amount of about 150 kg of the molecular catalyst is needed. With the listed cobalt complex and ligand prices, the total precursor costs (i.e., cobalt carbonyl + ligand) are about 1 $€\cdot g_{cat}^{-1}$, resulting in a total molecular catalyst value of about 150 k€. The lifetime of the molecular catalyst is estimated to be about 1-2 years based on its amount lost through permeation upon retention, as experimentally measured by project partner

MPG in WP4. The molecular catalyst consumable costs per year are then about 91 k€·yr$^{-1}$. Note that it is implicitly assumed here that the required fresh molecular catalyst flow rate depends on the amount lost via the OSN membrane permeate and not on possible intrinsic catalyst activity losses over time.

The solvent of the molecular catalyst, squalane, is only commercially available in liter scale for a price in the range 40-50 €·L$^{-1}$ (i.e., 50-60 €·kg$^{-1}$) [26,27]. The required squalane solvent mass flow rate is approximated based on the solvent-to-molecular catalyst mass ratio experimentally employed by CSIC in WP2. This results in a total solvent mass flow of about 58 kg·h$^{-1}$. When considering a squalane retention by the OSN membrane of 99.7%, the fresh squalane costs are about 75 k€·yr$^{-1}$.

*Retention of molecular catalyst*

The replacement costs of the OSN membrane module are calculated to be 6.4 k€·yr$^{-1}$ based on the membrane module capital investment of 32 k€ (**Section 2.2.1**) and an estimated membrane lifetime of five years for OSN membranes [28]. Herein, it is assumed that the entire membrane module needs to be replaced at the end of lifetime.

*Adsorbents*

A molecular sieve 3A adsorbent is used in the TSA step for $H_2O$ removal and an activated carbon adsorbent is employed in the PSA step to remove $CO_2$ from the syngas feed of the tandem reactor. The required amount of molecular sieve 3A is about 40 kg and its market price is estimated at 8.70 €·kg$^{-1}$ [29]. When considering a lifetime of three years [30], this yields total molecular sieve costs of only 115 €·yr$^{-1}$. The required amount of activated carbon adsorbent is about 1,200 kg and the corresponding price for the European market is around 2.25 €·kg$^{-1}$ [31]. When taking into account a lifetime of two years [32], this results in fresh activated carbon adsorbent costs of around 1.4 k€·yr$^{-1}$.

*Extractants*

The extractants used in the three extraction columns of the E-TANDEM demonstration plant are based on Li et al. [13]. An acetonitrile/water mixture is used as extractant in extraction columns 1 and 2 for extraction of the relatively short chain higher alcohol fractions, whereas an acetonitrile/methanol mixture is applied as extractant for the longest chain higher alcohol fraction in column 3. The total mass flow of the acetonitrile/water extractant (i.e., column 1 + 2) is around 95 kg·h$^{-1}$, corresponding to ca. 38 kg·h$^{-1}$ of acetonitrile and ca. 57 kg·h$^{-1}$ of water for an acetonitrile/water mass ratio of 1:1.5 [13]. Besides, the required mass flow of the acetonitrile/methanol extractant is about 85 kg·h$^{-1}$, of which 68 kg·h$^{-1}$ is acetonitrile and 17 kg·h$^{-1}$ is methanol for an acetonitrile-to-methanol mass ratio of four [13]. The European market prices of the deionized water, acetonitrile, and methanol constituents are estimated at 1.75 €·$t_{H2O}^{-1}$ [21], 1.07 €·$kg_{acetonitrile}^{-1}$ [33], and 0.65 €·$kg_{methanol}^{-1}$ [34], respectively. For an extraction-azeotropic distillation cycle time of 10 minutes, and an extractant loss of ca. 3-4% per cycle, around 20% of extractant is lost per hour. Consequently, the respective fresh acetonitrile/water and fresh acetonitrile/methanol extractant costs are about 65 k€·yr$^{-1}$ and 134 k€·yr$^{-1}$.

*Dehydration (etherification) catalyst*

An AmberLyst™ 36 ion exchange resin is selected as a suitable dehydration catalyst in the etherification reactor, as recommended in deliverable report D2.1. Around 38 kg of this resin is needed. The

AmberLyst™ 36 catalyst costs 539 €·kg$^{-1}$ [35] and has an estimated operating service lifetime of five years [36]. Consequently, the annual resin costs are around 4 k€·yr$^{-1}$.

**Utilities**

The utilities required to operate the E-TANDEM demonstration plant are electricity, cooling water, and thermal oil. The electricity users comprise the co-SOE system (stack and balance of plant) and the compressors, blowers, pumps, and electrical reboilers (azeotropic distillation columns). In addition, electrical power consumptions of the required ancillary instrumentation and agitation equipment are included in the total electricity assessment. The total process electricity demand is around 1.5 MW. In the base case scenario, an electricity price of 0.05 €·kWh$^{-1}$ [37,38] is utilized, resulting in total electricity costs of about 600 k€·yr$^{-1}$.

A coolant is needed in the coolers, water separators, and condensers. Cooling water is selected as a cooling medium, because of its low costs and its applicability to the temperature ranges of the cooling and condensation steps within the process. In total, an energy duty of about 0.5 MW needs to be covered with cooling water, corresponding to ca. 39 m$^3$·h$^{-1}$ of cooling water. The Western European cooling water costs in 2025 are estimated around 0.08 €·m$^{-3}$ [39], leading to total cooling water costs around 26 k€·yr$^{-1}$.

The process contains two thermal oil cycles, combining the water feed heater with the Tandem reactor heat, and integrating syngas cooling with the azeotropic distillation column feed heaters. These thermal oil cycles cover temperature ranges between 20 and 240 °C. DOWTHERM™ A is a commonly used heat transfer fluid in the chemical industry [40]. It is applicable in the temperature range from 12 to 400 °C [41], and is therefore selected as a suitable heat transfer fluid here. The required DOWTHERM™ flow rates are calculated using its density and heat capacity, which are taken as 990.7 kg·m$^{-3}$ and 1.814 kJ·kg$^{-1}$·K$^{-1}$, respectively, across the applied temperature range [42]. The total required amount of DOWTHERM™ A is then calculated considering a cycle time of five minutes for both thermal oil cycles, leading to a total demand of about 2.9 m$^3$. The price of DOWTHERM™ A is estimated around 100 €·L$^{-1}$ [43], resulting in a total thermal oil value of about 290 k€. The service lifetime of the fluid is around ten years [44], yielding annual DOWTHERM™ costs of about 29 k€·yr$^{-1}$.

**Carbon tax**

Increased taxes are anticipated to be levied in the near future to chemical companies for emitting greenhouse gases from their production facilities with the aim of making the chemical industry more sustainable. Here, the focus is on carbon dioxide emissions, and the influence of carbon taxation on the viability of the E-TANDEM production facility is included in the financial assessment. A carbon tax of 57 €·$t_{CO2}^{-1}$ is considered, which was the average across 23 European countries in April 2025 [45]. The total carbon dioxide emissions contain emissions from the generation of regional grid electricity and emissions related to the incineration of the tail gas stream for generating internal process heat. Note that the carbon tax is, therefore, only imposed to $CO_2$ emissions related to the e-fuel production phase, and not to emissions related to the $CO_2$ resource and/or other life cycle aspects.

### Fixed production costs

The fixed production costs are independent of the production capacity and comprise costs related to personnel, process maintenance, equipment replacement, rent of land, property taxes, plant insurances, corporate overhead charges, and depreciation charges. It is expected that two operators per shift are needed to operate the 1 MW demonstration plant. Four shifts are needed to allow for weekends and holidays, resulting in eight operators in total. An operator salary of 60,000 $€\cdot yr^{-1}$ is taken, valid for Germany in 2025 [46], yielding total operator salary costs of 480 $k€\cdot yr^{-1}$. The management labor costs are estimated at 25% of the operator costs (120 $k€\cdot yr^{-1}$) [47], while the salary overhead costs are 40% of the total labor costs (240 $k€\cdot yr^{-1}$) [47]. The total personnel costs are thereby 840 $k€\cdot yr^{-1}$. The process maintenance costs are traditionally estimated at 3% of the ISBL capital investment [47], i.e., ca. 330 $k€\cdot yr^{-1}$. It is anticipated that the electrolyzer stacks need to be replaced periodically, as their expected lifetime, estimated at five years at a current density of 0.31 $A\cdot cm^{-2}$ [48], is shorter than the estimated plant lifetime of 25 years. Consequently, the stack replacement costs are around 64 $k€\cdot yr^{-1}$. The costs related to rent of land, property taxes, and plant insurances are all estimated at 1% of the (ISBL + OSBL) investment, i.e., all ca. 146 $k€\cdot yr^{-1}$ [47]. Besides, the corporate overhead charges are taken as 20% of the total labor costs (168 $k€\cdot yr^{-1}$). Lastly, the annual depreciation charge is calculated based on the total depreciable capital and an anticipated plant lifetime of 25 years, yielding ca. 416 $k€\cdot yr^{-1}$. The total fixed production costs are about 2.2 $M€\cdot yr^{-1}$, corresponding to ca. 3.7 $€\cdot L_{fuel}^{-1}$.

### Normalization per liter of product

The total synthetic ether production capacity of the E-TANDEM demonstration plant is about 13 $kg\cdot h^{-1}$ (i.e, 16 $L\cdot h^{-1}$). However, the paraffinic byproducts recovered from the extraction step and the wax and naphtha byproducts from the fractional distillation tower are also valuable fuel constituents. Following a hydrotreating upgrade, the latter would provide a renewable, non-oxygenated hydrocarbon backbone to the HOEF, or alternatively be directed to other e-fuel applications, such as jetfuel or gasoline. This upgrading step has not been considered in the techno-economic assessment presented in this report. When including the paraffins, waxes, and naphtha, as valuable products, the total fuel equivalent production capacity amounts to ca. 55 $kg\cdot h^{-1}$ (i.e., 75 $L\cdot h^{-1}$). The total production costs are normalized per liter of fuel equivalent to allow for a fair comparison with conventional fossil and alternative sustainable fuel production routes.

## 2.3 Sensitivity Analysis

The sensitivity analysis is divided into a geographical and a technological sensitivity analysis. The multifactor geographical sensitivity analysis focuses on the selection of the most suitable production location for the e-fuel demonstration plant. On the contrary, the technological sensitivity analysis identifies the influence of variations in several crucial process parameters on process economics.

### Geographical sensitivity analysis

The three key requirements of e-fuel production facilities are: (i) access to low carbon electricity (wind, solar, hydro, nuclear), (ii) access to $CO_2$ resources (e.g., industrial or biogenic $CO_2$), and (iii) access to open sea for fuel product transportation. Based on these three prerequisites, 16 potential plant locations are shortlisted, as presented in **Figure 2.1**. The focus here is on regions that have the potential to satisfy the listed prerequisites, even if, e.g., the existing electricity grids are currently still carbon intensive. In short, the considered plant locations range from the windy regions of Patagonia and Northeast China to sunny Spain, Morocco and Saudi Arabia, to hydropowered sites in Norway and Tasmania, to industrial hubs in Germany, France, and on the U.S. Gulf Coast (**Figure 2.1**).

***Figure 2.1.*** *Selected possible plant locations for the e-fuel production facility.*

The following local characteristics are incorporated in the multifactor geographical sensitivity analysis: electricity price, emission intensity of the current electricity grid, carbon taxation measures, $CO_2$ feedstock price, plant operator salary, and local capital investment. Moreover, the evaluation of the fuel production costs and production-related $CO_2$ emissions is carried out considering Western Europe and 2025 as reference location and year. For this reason, transportation costs and transportation-related $CO_2$ emissions are included, considering the Port of Rotterdam as the representative harbor for Western Europe. For a complete list of the local production characteristics, see **Appendix B**.

In this geographical sensitivity analysis, the electricity price for large consumers in energy-intensive industries in the period 2024-2025 is taken per country [49,50]. Besides, the electricity emission intensity is based on the grid electricity mix in 2025 [49]. Grid electricity is considered, as the e-fuel plant needs to be connected to the regional electricity grid, instead of, e.g., decentralized solar or wind farms, to be able to operate continuously. The imposed carbon tax corresponds to carbon taxation measures acting in the concerning countries in the period 2023-2025 [51–54]. The local $CO_2$ source is defined based on announced e-fuel production projects in the concerning countries and regions [55–60], or based on the available natural resources and/or industrial facilities. The corresponding $CO_2$ source costs are estimated based on currently valid price ranges for the concerning $CO_2$ capture technologies [61,62]. All fuels are considered to be transported from the production location to the Port of Rotterdam via maritime cargo transport, using MGO as transportation fuel. The emission intensity of fuel combustion in cargo ships is estimated at 40 $g_{CO2}\cdot t_{cargo}^{-1}\cdot km^{-1}$ [63]. Besides, the transportation distance is approximated by means of the shortest possible sea distance [64]. The fuel transportation costs are calculated based on the average of the container rates on three popular routes, i.e., Shanghai-Los Angeles, Shanghai-New York, and Shanghai-Dallas [65], yielding transportation costs of ca. 0.05 €·$t_{cargo}^{-1}$·$km^{-1}$. The local operator salaries are estimated for chemical plant operators [66,67], and evaluated and currency corrected at 25 February 2026. Lastly, the local capital investment is calculated from the base case CAPEX, evaluated for Germany, using plant location factors for industrial plants valid in 2021 [19].

The geographical sensitivity analysis is not only performed for a demonstration plant with an electrolyzer capacity of 1 MW, i.e., the capacity targeted in the E-TANDEM project, but also for plants with an electrolyzer capacity of 100 MW and 1,500 MW, respectively. While the 1 MW base case corresponds to a production of ca. 0.4 ktpa of fuel constituents, the 100 MW intermediate scale and 1,500 full industrial scale plants would correspond to around 44 ktpa and 660 ktpa of fuel, respectively. The total direct investments for a 100 MW and 1,500 MW plant capacity are estimated from the base case 1 MW total direct investment using the six-tenths rule [8]. For the higher production capacities, the required amounts of raw materials, consumables, utilities, and related $CO_2$ emissions, are multiplied by a factor of 100 and 1,500, respectively, assuming that these amounts scale linearly with production capacity. Besides, for the base case 1 MW plant, a total number of 8 operators is considered (4 shifts, 2 operators per shift). On the contrary, for the two larger plants, a total of 20 operators is considered (4 shifts, 5 operators per shift), as estimated based on Towler and Sinnott [47].

# 3 Results & Discussion

## 3.1 Results

### 3.1.1 Base case analysis

The total inside battery limit (ISBL) investment is around 10 M€ and comprises expenditures on reactors, columns, filters, vessels, pumps, compressors, and heat exchangers (**Figure 3.1a**). After inclusion of outside battery limit (OSBL) costs (40% of ISBL [8]), i.e., investments related to site modifications, and detailed engineering and contingency charges (25% of (ISBL+OSBL) [8]), the total process investment is around 18 M€ (**Figure 3.1a**). Main contributors to the plant CAPEX are the reactors, pressure changers, and heat exchangers. The total reactor investment is substantial, particularly because of the electrolyzer costs. The total electrolyzer investment, i.e., stack + balance-of-plant (BoP), is about 1.5 M€, corresponding to ca. 1,500 $€\cdot kW^{-1}$ for a 1 MW electrolyzer capacity. The CAPEX of SOEs, including BoP, typically ranges between 1,000-5,000 $€\cdot kW^{-1}$ [68]. The E-TANDEM electrolyzer investment is on the lower end of this range, because of the cost-effective metal-supported cell design applied. Besides, compressor and heat exchanger investment are considerable, because of transitions from high temperature, low pressure equipment, like the co-SOE system, to high pressure, relatively low temperature units, like the Fischer-Tropsch-based tandem reactor, within the proposed process.

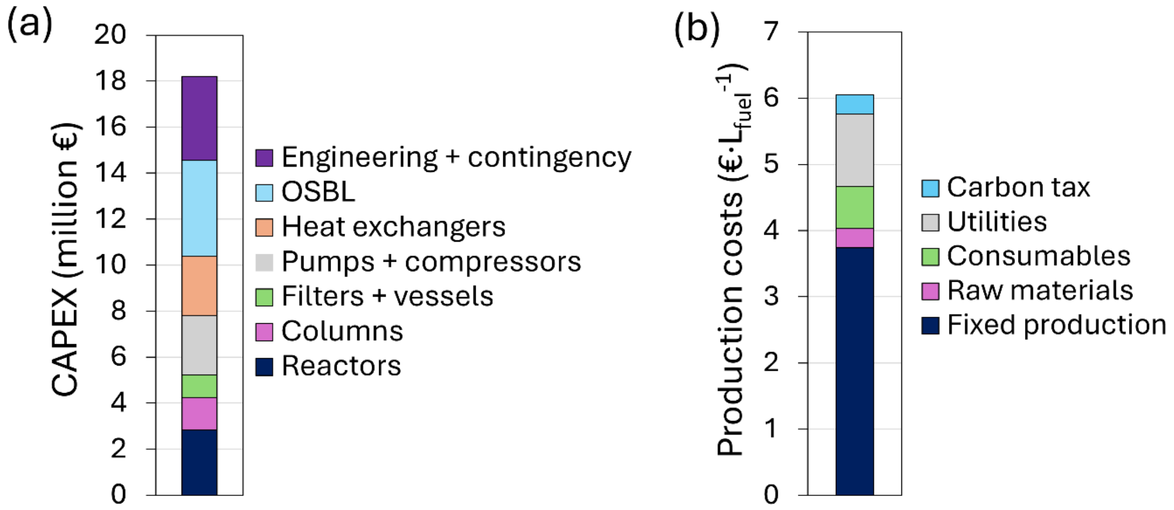

***Figure 3.1.*** *(a) Capital investment of the E-TANDEM demonstration plant (1 MW), and (b) the corresponding e-fuel production costs. Data is valid for Western Europe, 2025, OSBL indicates the outside battery limit investment.*

For the limited 1 MW electrolyzer capacity of the E-TANDEM demonstration plant, the e-fuel production costs are about 6 $€\cdot L_{fuel}^{-1}$ and are dominated by the fixed production costs (**Figure 3.1b**). At this production scale, the personnel costs, process maintenance costs, and depreciation charges, are high per liter of fuel produced. In addition, consumables (e.g., catalysts, extractants) and utilities (primarily electricity) are significant contributors to the e-fuel production costs (**Figure 3.1b**). A detailed breakdown of the CAPEX and OPEX results is available in **Appendix B**.

### 3.1.2 Sensitivity analysis

#### *3.1.2.1 Geographical sensitivity analysis*

The geographical sensitivity analysis of the 1 MW e-fuel production facility divides the 16 potential site locations into three groups (**Figure 3.2**). At this production scale, the e-fuel production costs are high (4-7 €·$L_{fuel}^{-1}$), irrespective of the local process and electricity market characteristics. This implies that the main criterion for selecting suitable plant locations at this scale is the emission intensity of the electricity (<100 $g_{CO2}$·$kWh^{-1}$) to minimize the process carbon footprint. Suitable countries for the 1 MW plant, i.e., group I, are Norway, Spain, France, Brazil, Uruguay, and Canada (**Figure 3.2**). Group II in **Figure 3.2** comprises countries governed by natural gas-based electricity grids (United States, Egypt), and countries that are currently transitioning from fossil-based to renewable electricity grids (Argentina, Chile, Germany). The latter countries are anticipated to become suitable e-fuel production locations, as their grids are forecasted to become even more renewable in the coming decades. For the countries in group III, drastic changes of the electricity resources are needed to become suitable production locations. It should be noted that this analysis concerns national average electricity grids. Locally or regionally decentralized electricity grids can differ substantially from the national average. As an example, the electricity grid in Tasmania is mainly based on renewable hydropower, making Tasmania specifically a much more suitable e-fuel production location than the Australian average. Lastly, it should be noted that the e-fuel production costs for a 1 MW scale (4-7 €·$L^{-1}$) are considerably higher than the large scale production costs of fossil diesel and biodiesel (**Figure 3.2**), irrespective of the production location.

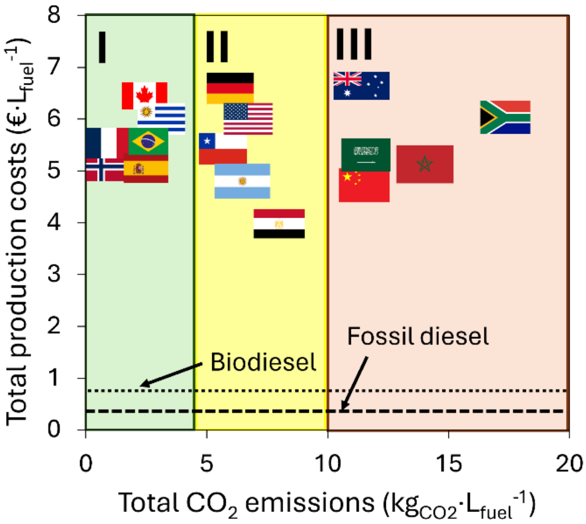

*Figure 3.2. Geographical sensitivity analysis of the base case 1 MW e-fuel production facility. The total production costs cover variable as well as fixed production costs, while the $CO_2$ emissions cover the production-related $CO_2$ emissions including electricity generation, tail gas incineration, and fuel product transportation. Note that negative $CO_2$ emissions related to the usage of $CO_2$ as fresh process feedstock (ca. -2.83 $kg_{CO2}$·$L_{fuel}^{-1}$) are not included in this graph. Roman numerals indicate grouping by similar e-fuel production suitability.*

For the intermediate (100 MW) and future industrial (1,500 MW) production scale, the geographical distribution of the production costs (**Figure 3.3**) significantly differs from the distribution of the 1 MW facility in **Figure 3.2**. Total production costs can be lowered from ca. 6 €·$L_{fuel}^{-1}$ to about 1.5-2.5 €·$L_{fuel}^{-1}$ by increasing the production capacity from 1 MW to 1,500 MW. For the larger production capacities, Norway, Spain and France (group II in **Figure 3.3**) are preferred to Uruguay, Brazil, and Canada (group

I in **Figure 3.3**) as e-fuel production location. Despite expensive labor and equipment investment, the production costs are lower in the three European countries, because of relatively low industrial electricity prices as compared to Uruguay, Brazil, and Canada. At these larger capacities, the total production costs are dominated by the variable production costs, i.e., mainly the electricity costs, instead of the fixed production costs. Electricity costs are currently relatively high in Canada, Uruguay, and Brazil, because of high infrastructural federal charges. Groups III and IV in **Figure 3.3** comprise countries that still have to transition from fossil to renewable electricity grids. The production costs of the group IV countries are low, as carbon taxation measures are absent, hence carbon emissions related to their fossil electricity generation are not penalized. Lastly, it should be noted that for a 1,500 MW facility, a hydropower station or a complete wind mill or solar panel farm needs to be connected to satisfy the enormous electricity demand.

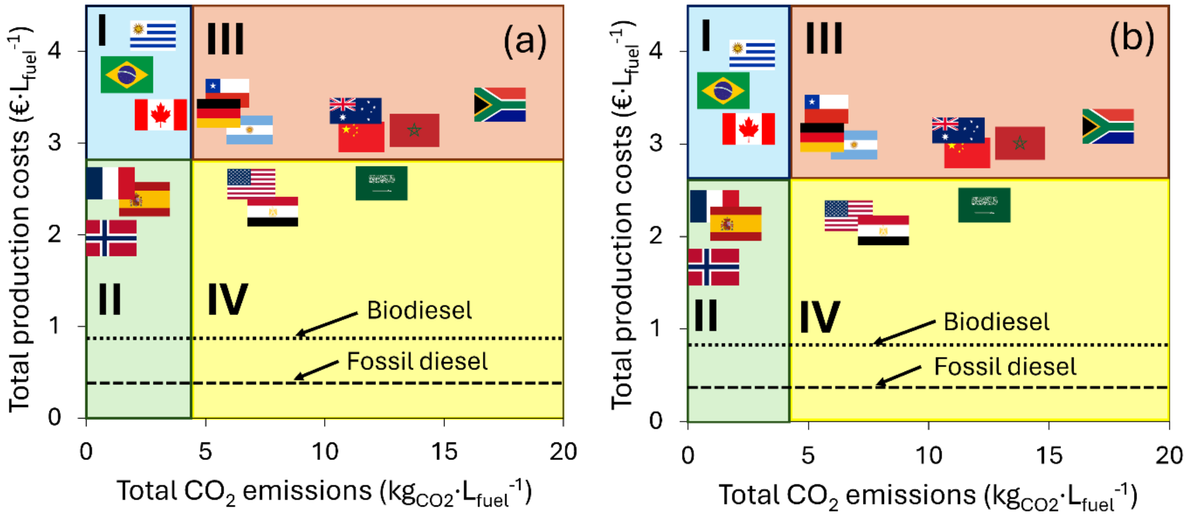

*Figure 3.3. Geographical sensitivity analysis of the e-fuel production facilities with electrolyzer capacities of (a) 100 MW, and (b) 1,500 MW. The total production costs cover variable as well as fixed production costs, while the $CO_2$ emissions cover the production-related $CO_2$ emissions including electricity generation, tail gas incineration, and fuel product transportation. Note that negative $CO_2$ emissions related to the usage of $CO_2$ as fresh process feedstock (ca. -2.83 $kg_{CO2}\cdot L_{fuel}^{-1}$) are not included in these graphs. Roman numerals indicate grouping by similar e-fuel production suitability.*

In short, for the larger production scales (100 and 1,500 MW) suitable production locations should possess a combination of renewable (<100 $g_{CO2}\cdot kWh^{-1}$) and low-cost (<0.05 €·$kWh^{-1}$) electricity, as in those cases the production costs are governed by the electricity contribution. Even for the high production capacities, the E-TANDEM e-fuel production costs are higher than fossil diesel and biodiesel production costs (**Figure 3.3**). For this reason, regulatory frameworks providing taxes on fossil fuels or subsidies and quotas on the production and utilization of e-fuels will be needed to stimulate widespread implementation of these e-fuels. Also note that the carbon footprint of the industrial production of fossil diesel and biodiesel are around 0.6 $kg_{CO2}\cdot L_{fuel}^{-1}$ and 0.4 $kg_{CO2}\cdot L_{fuel}^{-1}$ [69], respectively, which emphasizes the requirement to use fully renewable electricity in e-fuel production to environmentally outcompete alternative diesel production routes. Nevertheless, it is noted that negative $CO_2$ emissions related to the usage of $CO_2$ as fresh process feedstock (ca. -2.83 $kg_{CO2}\cdot L_{fuel}^{-1}$) in the E-TANDEM e-fuel demonstration facility are not included in **Figure 3.2** and **Figure 3.3**. Lastly, it is a promising observation that European countries can be suitable e-fuel production locations, as this could reduce the European dependency on other continents in satisfying the fuel demand in hard-to-abate transport sectors in the upcoming decades.

### *3.1.2.2 Technological sensitivity analysis*

**Current density of co-SOE system**

The electrolyzer stack costs strongly depend on the required electrolyzer area and, thereby, on the applied current density. Next to the base case current density of 0.31 A·cm$^{-2}$, an additional case is considered in which the electrolyzer is operated with an industrially recommendable current density of 1.0 A·cm$^{-2}$ [6]. Operation at the higher current density reduces the total stack investment from about 318 k€ to about 170 k€. In contrast to the stack investment, the BoP investment is considered independent of the electrolyzer current density. Because of the relatively cheap electrolyzer cell design, an increment in current density only leads to a marginal reduction in plant CAPEX (**Table 3.1**).

Operation at a higher current density is anticipated to accelerate cell degradation, which influences the operating costs related to stack maintenance and replacement. Here, it is considered that only the stack materials need to be replaced periodically. The stack lifetime is estimated to be five years for 0.31 A·cm$^{-2}$ and two years for 1.0 A·cm$^{-2}$, respectively [48]. As a consequence, stack replacement costs are similar for the high and low current density cases, 397 k€·yr$^{-1}$ and 376 k€·yr$^{-1}$, respectively, despite the lower stack capital investment at higher current densities.

**Operating pressure of the tandem reactor**

In the base case scenario, the tandem reactor is operated at 180 bar(a), while the connected membrane for retaining the molecular catalyst has a maximum operating pressure of 60 bar(a). It would be beneficial to operate the entire molecular catalyst cycle at 60 bar(a). When considering tandem reactor operation at 60 bar(a) instead of 180 bar(a), the corresponding reactor investment is about halved (**Table 3.1**), as less construction materials are needed, due to reduced wall thicknesses at lower pressures. Note that it is implicitly assumed here that the same higher alcohol productivity can be attained at 60 bar(a) as experimentally obtained at 180 bar(a) in D2.2. Besides, the number of required syngas compression stages is reduced from five to three, lowering the total investment in pressure changing equipment from ca. 2.6 M€ to ca. 2.2 M€ (**Table 3.1**).

Lower pressure operation of the tandem reactor reduces the boiling point of its content. Consequently, the size of the gaseous outlet is increased (from 330 kg·h$^{-1}$ to 458 kg·h$^{-1}$), whereas the size of the liquid outlet is reduced (from 52 kg·h$^{-1}$ to 15 kg·h$^{-1}$) when lowering the tandem reactor pressure. This influences the dimensions of all process equipment that are part of the liquid and gaseous recycles around the reactor. Despite a reduction in compressor cooling requirements, the total heat exchanger investment is slightly higher at lower tandem reactor pressures (**Table 3.1**), as larger heat exchangers are needed in the gaseous reactor outlet. Overall, the capital investment is 5-10% lower when operating the tandem reactor at 60 bar(a) instead of 180 bar(a) (**Table 3.1**). As a result of the lower reactor pressure, the electricity demand for compression is slightly lowered, which reduces the total production costs from 6.05 €·$L_{fuel}^{-1}$ to 5.90 €·$L_{fuel}^{-1}$.

**Alternative syngas conditioning method**

The base case process contains a TSA-PSA combination to remove $H_2O$ and $CO_2$ impurities from the syngas feed. Because of the significant investment in adsorption columns, it is suggested to replace the $CO_2$ removal step by membranes. Polaris$^{TM}$ membranes from MTR Industrial Separations are selected as suitable membrane type to selectively separate $CO_2$ from the syngas feed [70]. The Polaris$^{TM}$ membrane also isolates the $CO_2$ from the $H_2O$ [71], the latter can subsequently be removed from the syngas feed through stage-wise condensation. Moreover, the Polaris$^{TM}$ membrane is known to be unaffected by the presence of $H_2O$ [71], implying that an additional $H_2O$ removal step is omitted from the process.

The ambient $CO_2$ permeance through Polaris$^{TM}$ membranes from gas mixtures is estimated at 750 gpu for a 20 bar(a) feed pressure [72]. This corresponds to a $CO_2$ membrane flux of about $3.35\cdot10^{-5}$ $mol_{CO2}\cdot m^{-2}\cdot s^{-1}$ for a trans-membrane pressure of 1 bar. The $CO_2$ recovery of a single stage Polaris$^{TM}$ membrane is up to 80% [70], implying that a 2-stage membrane system is needed to attain a targeted total $CO_2$ recovery of 90+%. The process simulations by AVL in WP4 revealed that the $CO_2$ flows to be removed per membrane stage are around 154 $kg_{CO2}\cdot h^{-1}$ (i.e., ca. 0.97 $mol_{CO2}\cdot s^{-1}$) and 31 $kg_{CO2}\cdot h^{-1}$ (i.e., ca. 0.19 $mol_{CO2}\cdot s^{-1}$), respectively. Consequently, the respective required Polaris$^{TM}$ membrane areas per stage are about 29,100 $m^2$ and 5,800 $m^2$. The area-normalized price of Polaris$^{TM}$ membranes, including module and skid investment, is in the range 50-100 $USD\cdot m^{-2}$ [71]. When considering the lower end of this range for future industrial applications, the total Polaris$^{TM}$ membrane costs per stage are around 1.26 M€ and 0.25 M€, respectively. **Table 3.1** shows that the $CO_2$ filter investment is higher than the reduction in adsorption column investment when replacing the TSA-PSA combination by a $CO_2$ removing membrane. Besides, investments in pressure changers and heat exchangers slightly increase, because of the additional need for stage-wise $H_2O$ condensation in the syngas compression section. In total, the process is around 10% more expensive when using membranes instead of adsorption columns for syngas conditioning.

## Solvent-to-feed ratio of the extraction step

The paraffinic byproducts are separated from the higher alcohol product fractions through the use of extraction, based on the experimental performance from Li et al. [13]. Mixtures of acetonitrile/water and acetonitrile/methanol are used as extractants in mass-based solvent-to-feed ratios (*S/F*) of 4:1 [13]. The acetonitrile and methanol are relatively expensive components, hence it would be financially beneficial to minimize extractant usage. In case the *S/F* ratio could be reduced from 4:1 to 2:1, this would lead to a total direct investment reduction of about 4% (**Table 3.1**), due to a reduction in the size of the extraction columns, azeotropic distillation columns, and connected reboilers and condensers. Besides, this *S/F* reduction results in a 6% decline in fuel production costs from 6.05 $€\cdot L_{fuel}^{-1}$ to 5.71 $€\cdot L_{fuel}^{-1}$, due to a lower fresh extractant demand and a lower electricity demand for the reboilers of the azeotropic distillation columns.

***Table 3.1.** Technological sensitivity analysis, covering cases in which (i) the co-SOE current density ($I_A$) is increased from 0.31 A·cm$^{-2}$ to 1.0 A·cm$^{-2}$, (ii) the tandem reactor pressure ($P_{reactor}$) is decreased from 180 bar(a) to 60 bar(a), (iii) the TSA-PSA combination for syngas conditioning is replaced by a $CO_2$ removing filter, and (iv) the solvent-to-feed ratio (S/F) of the extraction is reduced from 4:1 to 2:1.*

| Equipment | Base case | $I_A$ → 1.0 A·cm$^{-2}$ | $P_{reactor}$ → 60 bar(a) | TSA-PSA → $CO_2$ removing membrane | *S/F* → 2:1 |
|---|---|---|---|---|---|
| Co-SOE (stacks +BoP) | 1,487,800 | 1,339,700 | 1,487,800 | 1,560,900 | 1,487,800 |
| Tandem reactor | 747,800 | 747,800 | 359,100 | 747,800 | 747,800 |
| Etherification reactor | 594,200 | 594,200 | 594,200 | 594,200 | 594,200 |
| Columns | 1,416,000 | 1,416,000 | 1,416,000 | 883,300 | 1,328,000 |
| Pressure changers | 2,570,100 | 2,570,100 | 2,207,900 | 2,643,200 | 2,567,800 |
| Heat exchangers | 2,606,500 | 2,606,500 | 2,683,900 | 2,625,700 | 2,301,600 |
| Vessels/filters | 975,400 | 975,400 | 964,100 | 2,428,300 | 927,300 |
| ***Total direct investment:*** | ***10,397,900*** | ***10,249,800*** | ***9,713,100*** | ***11,483,500*** | ***9,954,700*** |
| ***Relative difference:*** | **-** | ***-1.5%*** | ***-6.6%*** | ***+10.4%*** | ***-4.3%*** |

**Tornado diagrams**

An additional sensitivity analysis is carried out in which the influence of relative fluctuations (+10% and -10%) in the values of several main contributors to the total fuel production costs is investigated (**Figure 3.4**). This allows for identification of the parameters governing the e-fuel production costs. At the 1 MW scale targeted in E-TANDEM, the salaries of the plant operators are an important factor (**Figure 3.4a**), as fixed production costs play a dominant role at this small scale. On the contrary, for larger scale industrial plants (1,500 MW), the e-fuel production costs are governed by variable production costs, making electricity price and consumable prices the most crucial factors (**Figure 3.4b**).

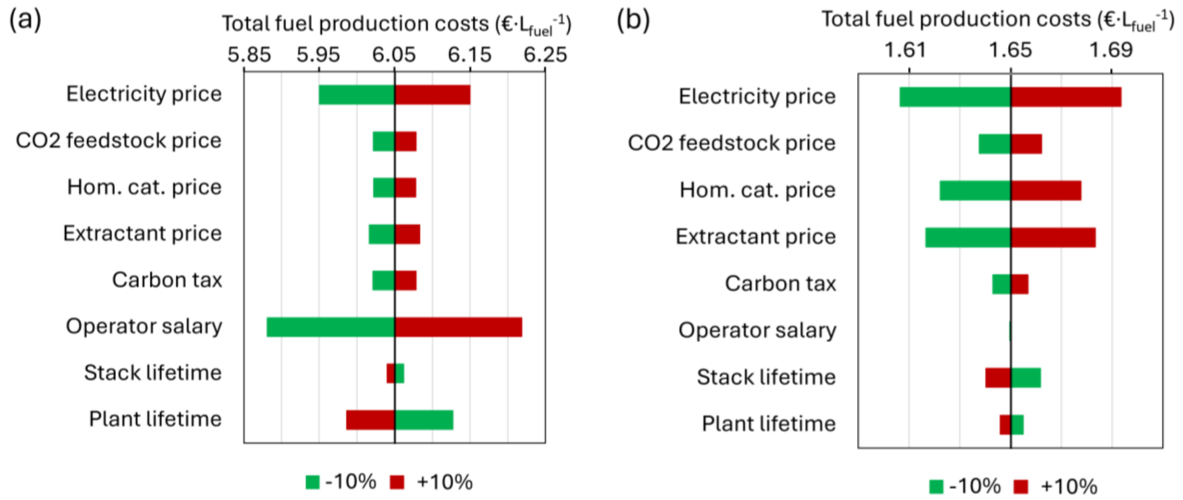

***Figure 3.4.*** *Tornado diagrams presenting the influence of main contributors to the total e-fuel production costs for (a) the 1 MW base case scenario, and (b) the 1,500 MW Norwegian scenario, in case parameters increase or decrease by 10% with respect to the base case value.*

**Dynamic operation of co-SOE system**

In the base case scenario, the E-TANDEM demonstration plant is considered to operate 8,000 h·yr$^{-1}$. However, on neither of the shortlisted geographical regions the wind will be permanently blowing, nor will the sun be permanently shining. Ideally, one would operate the electrolyzer only when there is low-cost, renewable electricity at hand. Specifically, a low industrial electricity price of 0.02 €·kWh$^{-1}$ is considered here at full load operation. While the electrolyzer operates dynamically, continuous e-fuel production is still desired. For this reason, an additional syngas storage vessel and gas blower are required to temporarily store and subsequently transport the syngas. Herein, the syngas is considered to be stored at 20 bar(a), right after the syngas conditioning steps. The total direct investment increases from ca. 10.4 M€ to ca. 11.7 M€ when operating the electrolyzer dynamically, due to the additional syngas storage costs. However, electricity costs are reduced from ca. 600 k€·yr$^{-1}$ to ca. 330 k€·yr$^{-1}$, because of the usage of only low-cost electricity to operate the co-SOE system.

### *3.1.2.3 Comparison with fossil and alternative sustainable fuel production routes*

The E-TANDEM e-fuel production costs are compared with (i) conventional marine gas oil (MGO), (ii) hydrotreated vegetable oil (HVO) biodiesel, (iii) biomethanol, and (iv) e-methanol. The volumetric energy density of methanol is about half that of MGO, biodiesel, and E-TANDEM e-fuel (**Figure 3.5**), making biomethanol and e-methanol less compatible with existing fleet infrastructures. The biodiesel production costs are currently only slightly higher than MGO production costs. However, biodiesel costs are expected to increase in the coming decades, due to a growing demand for vegetable oils and waste-based feedstocks, in combination with a looming supply crunch. The E-TANDEM e-fuels have a volumetric energy density that is similar to biodiesel and MGO, but its production costs are higher, even for large scale plants (>100 MW, **Figure 3.5**). **Figure 3.5** shows that the gaps in production costs to be closed through regulatory frameworks, like emission taxations and e-fuel subsidies, are about 0.60 €·L$_{fuel}$$^{-1}$ between HOEF and biodiesel (2050), and about 1.20 €·L$_{fuel}$$^{-1}$ between HOEF and MGO.

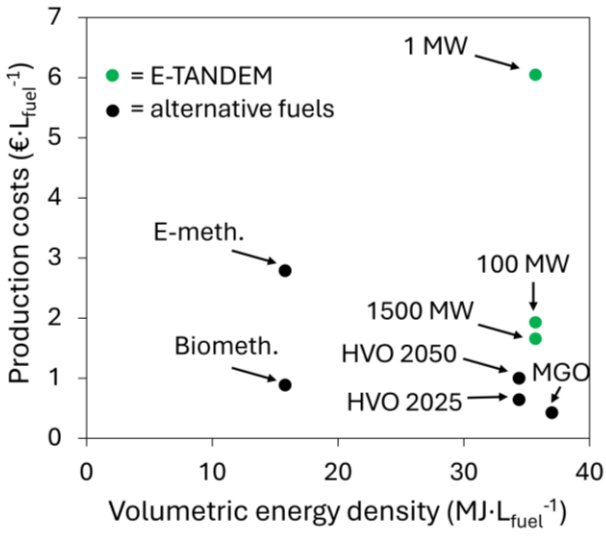

***Figure 3.5.*** *Comparison between the production costs of the E-TANDEM e-fuels, conventional fossil fuels and alternative sustainable fuels. MGO = marine gas oil, HVO = hydrotreated vegetable oil. Data on alternative fuels is based on Solakivi et al. [73], e-methanol production based on 5 MW plant and an electricity price of 0.08* $€·kWh^{-1}$*. E-TANDEM production data based on the base case scenario (1 MW, 0.05* $€·kWh^{-1}$*) and Norwegian large scale scenarios (100 and 1,500 MW, 0.025* $€·kWh^{-1}$*). Volumetric energy density is based on lower heating values.*

**Blending with MGO**

As initial commercialization strategy, the E-TANDEM HOEF products can be blended with conventional marine gas oil (MGO) to optimally meet the fleet infrastructure requirements and to mitigate production cost per liter of fuel. Blends containing 10%, 20%, 30%, and 50% of the produced HOEF with MGO are considered here, based on E-TANDEM deliverable report 3.1, which has shown those fuel blends to be stable. The fuel production costs of these blends are calculated using MGO production costs of about 0.43 $€·L_{fuel}^{-1}$ (507 $€·t_{fuel}^{-1}$) [74] and the Norwegian 1,500 MW E-TANDEM e-fuel production costs of 1.65 $€·L_{fuel}^{-1}$, i.e., 7.48 $€·L_{HOEF}^{-1}$ (**Table 3.2**). **Table 3.2** shows that fuel costs can be below 1 $€·L_{fuel}^{-1}$ even when blending substantial amounts of e-fuel (up to 50%) from the Norwegian 1,500 MW E-TANDEM facility with MGO. The compatibility of 10% and 20% HOEF ether blends with waterborne and automotive fuel norms was confirmed in D3.1. Note that those blends concerned ether mixtures, i.e., only HOEF constituents, excluding the paraffinic byproducts generated in E-TANDEM. Since the paraffinic byproducts are chemically even closer to traditional diesel fuel constituents, it can be safely assumed that inclusion of the electrically produced paraffinic byproducts from E-TANDEM in fuel blends will not harm the fuel norm conformation.

***Table 3.2.*** *Fuel production costs of blends of E-TANDEM e-fuels (1,500 MW, Norway) and marine gas oil (MGO). Note that a distinction is made here between cases in which only the HOEF ether products are utilized in the fuel, and in which all relevant fuel constituents (HOEF + paraffinic byproducts) are utilized.*

| E-TANDEM fuel content [%] | MGO content [%] | Fuel production costs, only HOEF [$€·L_{fuel}^{-1}$] | Fuel production costs, HOEF + paraffinic byproducts [$€·L_{fuel}^{-1}$] |
|---|---|---|---|
| 10 | 90 | 1.13 | 0.55 |
| 20 | 80 | 1.84 | 0.67 |
| 30 | 70 | 2.54 | 0.79 |
| 50 | 50 | 3.95 | 1.04 |
| 100 | 0 | 7.48 | 1.65 |

## 3.2 Contribution to project (linked) Objectives

This deliverable contributes to project objectives 5.1 and 5.2 of the E-TANDEM project:

- To assess the new e-fuel production concept from a techno-economic perspective and evaluate its conversion efficiency and competitiveness against alternative fuels, e.g., fossil and second generation BtL and HVO biofuels (5.1)
- To perform a sensitivity analysis for the overall economic feasibility of the fuel production concept (5.2)

The techno-economic characteristics of the E-TANDEM e-fuel production concept are presented and explained in **Section 3.1.1**, and compared with conventional fossil and alternative sustainable fuels in **Figure 3.5**. Besides, a detailed sensitivity analysis is applied in **Section 3.1.2** to assess the overall economic feasibility of the E-TANDEM e-fuel production concept, to identify suitable production locations of the e-fuel production facility at different production scales, and to identify the main cost drivers influencing the viability of the process at various production capacities.

## 3.3 Contribution to major project exploitable result

This deliverable supports KER 1: “Simulative investigation, optimization tools, and processes for E-fuel production plants”. The design of all process equipment needed to operate the E-TANDEM e-fuel production facility, as discussed in **Section 2.2.1**, represents the basis for detailed design, engineering, and construction of the 1 MW demo-scale e-fuel production facility. Besides, the corresponding techno-economic assessment presents information on the required capital investment and on the concerning e-fuel production costs (**Section 3.1.1**). Moreover, the detailed sensitivity analysis in **Section 3.1.2** identifies the process parameters that are most strongly influencing the process performance and economics, and thereby provides guidelines for further optimization of the e-fuel production plant and simulation model.

# 4 Conclusion and Recommendation

This deliverable (i) provides an overview of the capital expenditures and production costs of the E-TANDEM e-fuel production concept, (ii) identifies suitable production locations at various production scales, and (iii) identifies the main cost drivers of the concept at various production capacities. The total direct investment of the E-TANDEM demonstration plant (1 MW) is around 10 M€. Besides, the e-fuel production costs are high (ca. 6 $€\cdot L_{fuel}^{-1}$) at this small scale, but can be reduced to less than 2 $€\cdot L_{fuel}^{-1}$ by increasing the electrolyzer capacity from 1 MW to more than 100 MW. Also, for the 1 MW scale, the production costs are high (4-7 $€\cdot L_{fuel}^{-1}$), irrespective of the production location, as fixed production costs are dominating the fuel costs at this scale. For this reason, the main criterion for selecting suitable production locations of the 1 MW plant is the emission intensity of the local electricity grid, which should be below 100 $g_{CO2}\cdot kWh^{-1}$. On the contrary, for larger electrolyzer capacities (>100 MW), suitable production locations should possess a combination of green and low-cost electricity (<0.05 $€\cdot kWh^{-1}$), as fuel production costs are dominated by the electricity contribution for these larger production capacities. Lastly, the e-fuel production costs are forecasted to be higher than biodiesel and fossil diesel (<1.0 $€\cdot L_{fuel}^{-1}$), even for industrial e-fuel production sites (ca. 1.6 $€\cdot L_{fuel}^{-1}$). Consequently, regulatory frameworks are needed, covering emission taxations and/or e-fuel subsidies or quotas, to increase the incentive to commercially implement e-fuels. As a first strategy, E-TANDEM e-fuels can be blended with conventional MGO to mitigate fuel costs.

# 5 Interconnections with other deliverables

The individual pieces of process equipment are designed in this deliverable based on inputs from the mass- and energy balances retrieved from the process simulations reported by project partner AVL in **Deliverable 4.3**. In addition, experimental performance information of, e.g., the tandem reactor (**Deliverable 2.2**), the co-SOE system (**Deliverable 2.3**), molecular catalyst retention (**Deliverable 4.1**) and the etherification reactor, is used to dimension the required process equipment. Also, the expenditures on consumables, utilities, and raw materials are estimated using the mass- and energy balances resulting from the process simulations in **Deliverable 4.3**. Lastly, the production costs of fuel blends of E-TANDEM HOEF with MGO, as shown in **Table 3.2** of this deliverable report, are in line with the stable fuel blend compositions confirmed by project partner OWI in **Deliverables 3.1** and **3.2**.

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**Project partners:**

| # | Partner short name | Partner Full Name |
|---|---|---|
| **1** | CSIC | AGENCIA ESTATAL CONSEJO SUPERIOR DE INVESTIGACIONES CIENTIFICAS |
| **2** | MPG | MAX-PLANCK-GESELLSCHAFT ZUR FORDERUNG DER WISSENSCHAFTEN EV |
| **3** | DTU | DANMARKS TEKNISKE UNIVERSITET |
| **4** | OWI | OWI SCIENCE FOR FUELS GMBH |
| **5** | UNR | UNIRESEARCH BV |
| **6** | T4F | TEC4FUELS GmbH |
| **7** | AVL | AVL LIST GMBH |
| **8** | ZU | SVEUCILISTE U ZAGREBU, FAKULTET STROJARSTVA I BRODOGRADNJE |
| **9** | UCT | UNIVERSITY OF CAPE TOWN |
| **10** | KAUST | KING ABDULLAH UNIVERSITY OF SCIENCE AND TECHNOLOGY |

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# 8 Appendix A - Quality Assurance Review Form

The following questions should be answered by all reviewers (WP Leader, reviewer, Project Coordinator) as part of the Quality Assurance procedure. Questions answered with NO should be motivated. The deliverable author will update the draft based on the comments. When all reviewers have answered all questions with YES, only then can the Deliverable be submitted to the EC.

NOTE: This Quality Assurance form will be removed from Deliverables with dissemination level "Public" before publication.

| ***Question*** | **WP Leader** | **Reviewer** | **Project Coordinator** |
|---|---|---|---|
| | NAME (Organisation) | Markus Goll (AVL) | Gonzalo Prieto (CSIC) |
| ***1.*** *Do you accept this Deliverable as it is?* | Yes / No (elaborate) | Yes | Yes |
| ***2.*** *Is the Deliverable complete?*<br>- *All required chapters?*<br>- *Use of relevant templates?* | Yes / No (elaborate) | Yes | Yes |
| ***3.*** *Does the Deliverable correspond to the DoA?*<br>- *All relevant actions preformed and reported?* | Yes / No (elaborate) | Yes | Yes |
| ***4.*** *Is the Deliverable in line with the E-TANDEM objectives?*<br>- *WP objectives*<br>- *Task Objectives* | Yes / No (elaborate) | Yes | Yes |
| ***5.*** *Is the technical quality sufficient?*<br>- *Inputs and assumptions correct/clear?*<br>- *Data, calculations, and motivations correct/clear?*<br>- *Outputs and conclusions correct/clear?* | Yes / No (elaborate) | Yes | Yes |
| ***6.*** *Is created and potential IP identified and are protection measures in place?* | Yes / No (elaborate) | Yes | Yes |
| ***7.*** *Is the Risk Procedure followed and reported?* | Yes / No (elaborate) | Yes | Yes |
| ***8.*** *Is the reporting quality sufficient?*<br>- *Clear language*<br>- *Clear argumentation*<br>- *Consistency*<br>- *Structure* | Yes / No (elaborate) | Yes | Yes |

# 9 Appendix B – Technical Details

## 9.1 Geographical characteristics

***Table 9.1.*** *Geographical characteristics of the shortlisted production site locations for the e-fuel production concept. Fuel transport emissions and costs are calculated considering the port of Rotterdam as destination. Data is obtained across the timespan 2023-2025. References: electricity characteristics [49,50], carbon taxation [51–54], $CO_2$ source identification [55–60], $CO_2$ source pricing [61,62], fuel transportation characteristics [63–65], local operator salary [66,67], and local capital investment [19].*

| Country (Region) | Electricity source(s) | Carbon feedstock | Electricity price [USD·$MWh^{-1}$] | El. emission intensity [$g_{CO2}$·$kWh^{-1}$] | Carbon tax [USD·$t_{CO2}^{-1}$] | $CO_2$ feedstock price [€·$t_{CO2}^{-1}$] | Fuel transport emissions [$kg_{CO2}$·$L_{cargo}^{-1}$] | Fuel transport costs [€·$L_{cargo}^{-1}$] | Local operator salary [€·$yr^{-1}$] | Local CAPEX 1 MW plant [M€] |
|---|---|---|---|---|---|---|---|---|---|---|
| Argentina (Bahía Blanca) | Wind, solar | NG processing | 89 | 300 | 5 | 20 | 0.36 | 0.43 | 10,900 | 8.9 |
| Australia | Hydro, wind | Plantation forestry | 70 | 550 | 0 | 60 | 0.61 | 0.72 | 59,900 | 10.9 |
| Brazil (Port of Açu) | Wind, solar | Biogenic and industrial $CO_2$ | 122 | 60 | 0 | 60 | 0.29 | 0.34 | 13,900 | 8.9 |
| Canada (Ontario) | Nuclear | Industrial $CO_2$ | 102 | 120 | 10 | 30 | 0.20 | 0.23 | 55,300 | 9.0 |
| Chile (Patagonia) | Wind | Biogenic | 100 | 259 | 5 | 60 | 0.41 | 0.48 | 15,800 | 8.9 |
| China (Heilongjiang) | Wind | Biogenic | 65 | 565 | 0 | 60 | 0.62 | 0.73 | 21,700 | 8.2 |
| Egypt (Suez Canal) | Wind, solar | Biogenic | 40 | 380 | 0 | 60 | 0.18 | 0.22 | 3,900 | 10.0 |

| | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| France (Dunkerque) | Nuclear | Steel plants | 65 | 30 | 50 | 44 | 0.01 | 0.01 | 49,500 | 10.4 |
| Germany (Hamburg) | Wind, solar | Biogenic | 90 | 300 | 64 | 60 | 0.02 | 0.02 | 61,700 | 10.4 |
| Morocco (Dakhla) | Wind, solar | Industrial $CO_2$ | 100 | 660 | 0 | 30 | 0.09 | 0.10 | 13,700 | 10.0 |
| Norway (North) | Hydro | Biogenic, steel plants | 25 | 18 | 140 | 60 | 0.05 | 0.05 | 49,800 | 12.5 |
| Saudi Arabia (Port of Neom) | Wind, solar | DAC | 55 | 590 | 0 | 100 | 0.21 | 0.25 | 32,500 | 10.0 |
| South Africa (Humansdorp) | Wind, solar | Biogenic, industrial $CO_2$ | 82 | 850 | 13 | 60 | 0.38 | 0.45 | 26,800 | 11.9 |
| Spain (Bilbao) | Solar | Refinery | 55 | 90 | 17 | 60 | 0.04 | 0.05 | 36,400 | 10.4 |
| United States (Gulf Coast) | Wind, solar | Biogenic | 45 | 320 | 0 | 60 | 0.28 | 0.33 | 72,200 | 9.5 |
| Uruguay (Río Uruguay) | Wind, solar | Bioethanol | 125 | 115 | 159 | 30 | 0.36 | 0.42 | 8,000 | 8.9 |

## 9.2 Breakdown of capital expenditures (CAPEX)

The total base case capital investment (Western Europe, 2025) of the E-TANDEM demonstration plant is summarized in **Table 9.2**. The Chemical Engineering Plant Cost Index for 2025 is around 800.0 [75]. Besides, the location factor for Western Europe relative to the U.S. Gulf Coast is estimated at 1.1 [8].

***Table 9.2.*** *Capital investment of the E-TANDEM demonstration plant (1 MW), Western Europe, 2025.*

| Equipment | Cost | Note |
|---|---|---|
| ***Reactors*** | | |
| *Co-SOE (stack)* | €317,900 | |
| *Co-SOE (balance of plant)* | €1,169,900 | Balance of plant (BoP) investment |
| *Tandem reactor* | €747,800 | |
| *Etherification reactor* | €594,200 | |
| ***Subtotal:*** | ***€2,829,900*** | |
| Columns | €1,416,000 | |
| Pressure changers | €2,570,100 | |
| Heat exchangers | €2,606,500 | |
| Vessels and filters | €975,400 | |
| ***Total direct investment (TDI)*** | ***€10,397,900*** | ***Inside battery limit (ISBL)*** |
| *Total allocated investment (TAI)* | *€4,159,200* | *Outside battery limit (OSBL), 40% of ISBL [8]* |
| *Engineering cost + contingency* | *€3,639,300* | *25% of (TDI+TAI)* |
| **Total process investment** | **€18,196,400** | |

## 9.3 Breakdown of operating expenditures (OPEX)

***Table 9.3.*** *Operating expenditures of the E-TANDEM demonstration plant (1 MW), Western Europe, 2025.*

| Contributor | Cost [$€\cdot yr^{-1}$] | Note |
|---|---|---|
| ***Variable production costs*** | | |
| Raw materials | 172,200 | Considering $CO_2$ from direct air capture for a future target price of 100 $€\cdot t_{CO2}$ [76] |
| Consumables | 382,200 | |
| Utilities | 652,100 | |
| Carbon taxation | 170,400 | Based on an EU average carbon tax of 57 $€\cdot t_{CO2}$ [45] |
| ***Subtotal:*** | ***1,377,000*** | ***Equals 2.30 $€\cdot L_{fuel}^{-1}$*** |
| ***Fixed production costs*** | | |
| Personnel | 840,000 | |
| Maintenance | 375,500 | Include co-SOE stack replacement costs |
| Rent of land | 145,600 | |
| Property taxes | 145,600 | |
| Plant insurances | 145,600 | |
| Corporate overhead | 168,000 | |
| Depreciation | 415,900 | |
| ***Subtotal:*** | ***2,236,200*** | ***Equals 3.74 $€\cdot L_{fuel}^{-1}$*** |
| **Total production costs:** | **3,613,200** | **Equals 6.05 $€\cdot L_{fuel}^{-1}$** |