Chair of Energy Systems
Technical University of Munich

CESTEA:
The TUM Chair of Energy Systems Techno-Economic Analysis Method

Advancing Techno-Economic Analysis for Evaluating Power-to-X and Biomass-to-X Processes

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2024/05

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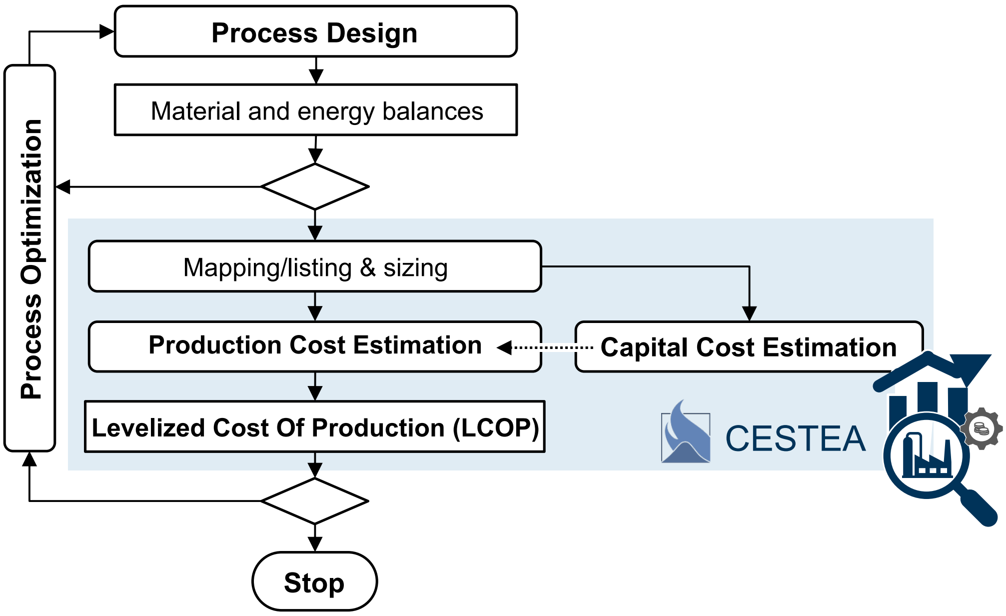
May, 2024

Abstract — This study presents the TUM CESTEA (TUM Chair of Energy Systems Techno-Economic Analysis) method, developed for economically analyzing Power-to-X (PtX), Biomass-to-X (BtX), and Power-and-Biomass-to-X (PBtX) processes. This method extends on standard factor-based cost estimation methods by incorporating specific adaptations and assumptions crucial for accurately evaluating these energy transformation processes. It specifically addresses the unique cost elements associated with key process components such as biomass treatment and electrolysis, thereby offering detailed insights into capital and production cost estimation. The primary objective is to establish a clear and systematic approach to cost estimation, enhancing the economic comparability of PtX, BtX, and PBtX technologies. The study emphasizes the importance of the Levelized Cost of Production (LCOP) as a critical economic metric for assessing these processes. Furthermore, the respective composition of the LCOP, consisiting of fixed, variable and depreciation costs, must be taken into account in the economic evaluation.



The calculation of LCOP follows a series of well-defined steps, aimed at ensuring a comprehensive and reliable economic analysis:

- Capital cost estimation of the main process equipment and its depreciation using the annualized cost method.
- Determination of variable production costs based on material and energy balances from process simulations.
- Fixed Costs of Production based on literature estimations.

Contents

List of Abbreviations

AEL	Alkaline Elektrolysis
AGR	Acid Gas Removal
$APEA$	Aspen Process Economic Analyzer
ASU	Air Separation Unit
BtX	Biomass-to-X
$CapEx$	Capital Expenditures
$CCOP$	Cash Costs of Production
$CEPCI$	Chemical Engineering Plant Cost Index
DAC	Direct Air Capture
DAC	Direct Air Capture
$eBtX$	electrified Biomass-to-X
FCI	Fixed Capital Investment
$FCOP$	Fixed Costs of Production
FT	Fischer-Tropsch
$ISBL$	Inside Battery Limits
$LCOP$	Levelized Cost of Production
$MACRS$	Modified Accelerated Cost Recovery System
$MOGD$	Mobil's Olefins to Gasoline and Distillates
MTO	Methanol to Olefins
$OpEx$	Operating Expenditures
$OSBL$	Outside Battery Limits
$P\&ID$	Piping and Instrumentation Diagram
$PBtX$	Power-and-Biomass-to-X
$PEMEL$	Proton-Exchange-Membrane Electrolysis
PFD	Process Flow Diagram
PSA	Pressure Swing Adsorption
PtX	Power-to-X
TCI	Total Capital Investment
$TCOP$	Total Cost of Production
TEA	Techno-Economic Analysis
$TPEC$	Total Purchased Equipment Cost
$VCOP$	Variable Costs of Production
WC	Working Capital
WGS	Water-Gas Shift

1 Background

In the course of anthropogenic climate change due to massive greenhouse gas emissions, it is necessary to transform industrial sectors which mainly rely on fossil based raw materials. Power-to-X (PtX), Biomass-to-X (BtX), and hybrid Power-and-Biomass-to-X (PBtX) or directly electrified BtX (eBtX) processes can produce chemicals and fuels from sustainable raw materials as renewable electricity, biomass residues and CO_2 from ambient air [1]. Techno-economic analyses (TEA) are essential to evaluate the innovative processes, identify the best process options and to give guidelines for required technology improvements and policies.

This publication introduces a comprehensive TEA method, known as Chair of Energy Systems Techno-Economic Analysis (CESTEA), designed specifically for BtX and PtX cost estimates. The basic approach for TEA is established in existing literature [2, 3]. However, these methods focuses on a wide range of conventional production processes in the chemical industry. For renewable and sustainable chemicals and fuel production processes, the established approach cannot be applied directly. Instead the specific differences between the renewable and fossil manufacturing paths must be taken into account and thus implicit assumptions must be adjusted. Notably, the presented method holds applicability for PtX and BtX processes. This includes guidelines for the dealing with biomass and electricity resources, as the largest shares of the variable operating costs, as well as clear allocation of the most important investment cost drivers, namely gasifier, synthesis and electrolyzer.

The CESTEA method, aims at suggesting a general framework for cost estimation including key steps, considerations, and parameters that are applicable across different processes within the PtX, BtX, and hybrid domains. The method is then applied to PtX, BtX and combinations of the two processes (PBtX, eBtX). Base case assumptions are outlined, providing a foundation for the cost estimation. These assumptions cover specific technologies along the process chain, and key economic factors specific to the processes under consideration. The aim of the method presented is to balance the level of detail required for a sufficiently accurate cost estimate with the ability to compare different methods.

2 Method details

TEA is a methodology based on a process design, combining technical and economic performance indicators, to assess the technical and economic viability of a production process. The final goal of a TEA is to concentrate all process data into one economic indicator, typically the Total Cost of Production (TCOP) [3] or the Levelized Cost of Production (LCOP). The TCOP/LCOP combines all costs of plant operations, product sales, and contribution to corporate functions such as management and research and development and allows for techno-economic process optimization [2, 3].

Usually, a scale-up approach can be applied to estimate costs based on similar existing projects, especially in large blue chip chemical plants as they have been designed and built for decades. However, when such specific information or data from existing reference processes is missing, a more detailed TEA methodology becomes necessary. This is especially important for novel processes such as PtX, BtX, PBtX and eBtX plants that have not yet been established. The general methodology presented here combines elements of process design and modeling, with plant dimensioning and capital cost as well as operating cost estimation, as shown in Figure 1. It is based on established cost estimation methods and correlations proposed by Peters et al. [2] and Towler and Sinnott [3].

As shown in Figure 1, the TEA generally starts based on the process design, typically a concept sketch, block schematic, process flow diagram (PFD) or detailed piping and instrumentation diagram (P&ID) allowing for material an energy balances [3]. The process design results in a more or less detailed overview of necessary equipment. The mapping or listing of the entire equipment is followed by the sizing of the equipment to allow for capital cost estimation (see Section 2.1.1).

The level of detail for any estimation of any process costs may vary from an order of magnitude estimate for concept screening purposes to a detailed estimate prepared from detailed engineering to be used by subcontractors for bids, or by owners for check estimates [4]. General flowsheets, PFDs or more detailed P&IDs are critical as primary scope-defining documents to estimate production costs [3]. They are also useful in determining the level of project definition, and thus the extent and maturity of estimate input data [4].

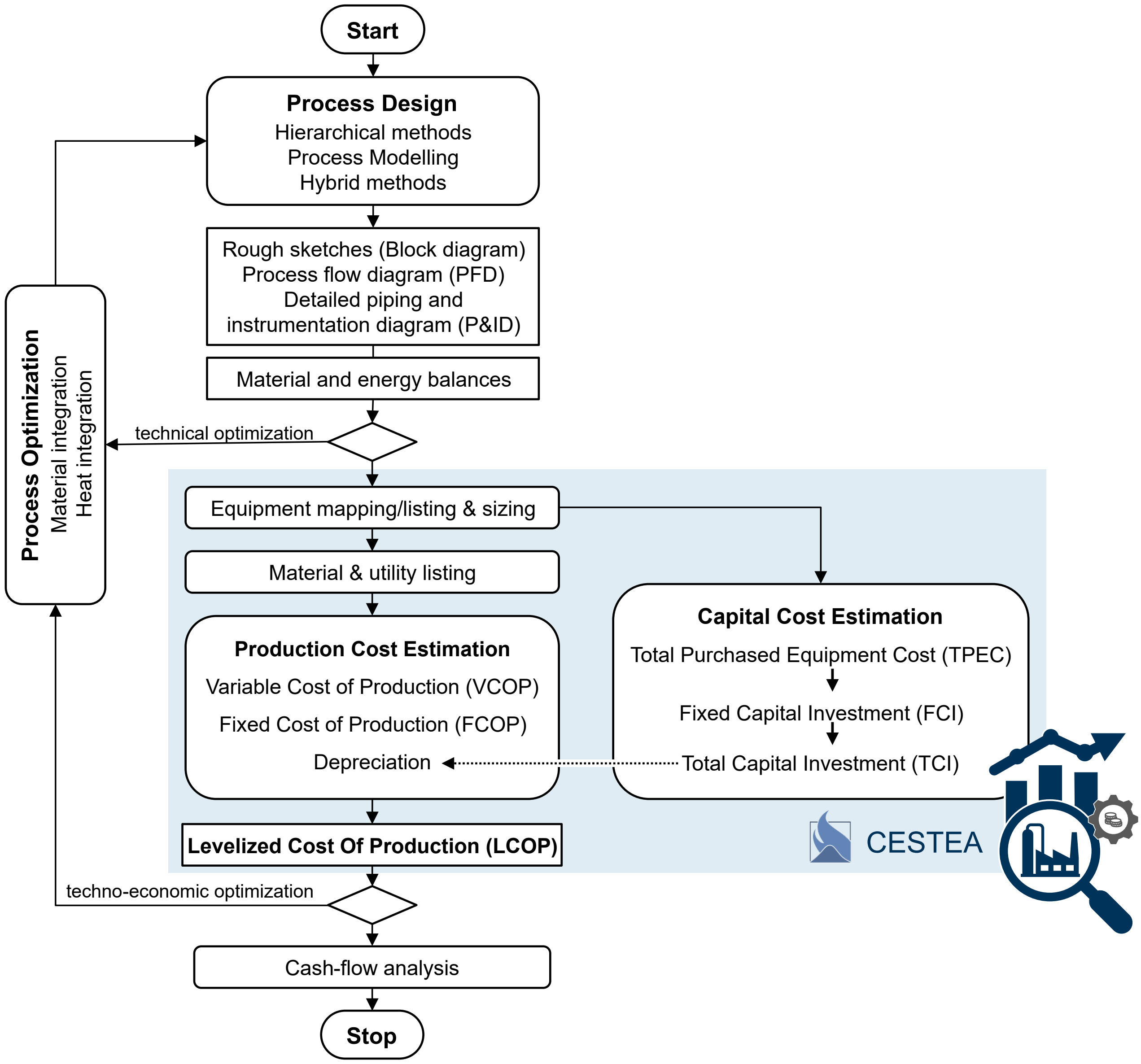


Figure 1 Simplified overview of comprehensive methodology of TUM Chair of Energy Systems Techno-Economic Analysis (CESTEA) as introduced in this paper.

The estimation levels are referred to by a number of terms, but the five classes shown in Figure 2 represent the level of detail and accuracy of estimation defined by the Cost Estimate Classification System for the process industries by the Association for the Advancement of Cost Engineering [4]. The system includes a mapping of the various phases and stages involved in project cost estimating, accompanied by a versatile maturity and quality matrix. It is designed to be adaptable across a diverse range of industries, encompassing sectors such as chemical manufacturing, petrochemicals, and hydrocarbon processing [4], making it highly applicable to PtX, BtX, PBtX and eBtX processes.

Commercial software can also be used for cost estimates, as they follow a similar workflow of process modelling, followed by mapping and sizing of equipment and the estimation of capital and operating plant cost. Such software may also allow sensitivity analyses to be carried out to understand how changes in key parameters affect the overall efficiency of the process. The Aspen Process Economic Analyzer (APEA), for examples, uses the ICARUS™ evaluation engine licensed by Aspen Technology Inc., and is one of the most widely used software solutions for estimating chemical plant costs [3]. There are, however, limitations to the use of software in TEA, as described in detail in Section 3.3.

Increasing project definition & accuracy

Class V — Order of Magnitude Estimate*
Concept Screening
based on very limited information using stochastic estimating methods

Process Flowsheet	Equipment Mapping	Utility Requirements	Accuracy*
None	None	None	-30% to +50%

Class IV — Study Estimate*
Feasibility Estimate
based on limited information typically using stochastic estimating methods

Process Flowsheet	Equipment Mapping	Utility Requirements	Accuracy*
Process flow diagram (PFD) for main process systems	Preliminary sizing & material specifications based on energy and material balances	Rough quantities (steam, water, electricity, etc.)	-15% to +30%

Class III — Preliminary Estimate*
Budget, Authorization or Control Estimate
based on sufficient data to permit the estimate to be budgeted using deterministic estimating methods

Process Flowsheet	Equipment Mapping	Utility Requirements	Accuracy*
PFD, utility flow diagrams, preliminary piping & instrumentation diagram (P&ID)	Engineered sizing & material specifications	Preliminary heat balances based on process simulation	-15% to +30%

Class II — Definitive Estimate*
Project Control or Bid/ Tender
based on almost complete data using deterministic estimating methods

Process Flowsheet	Equipment Mapping	Utility Requirements	Accuracy*
Engineered PFD, piping & instrumentation diagram (P&ID)	Engineered specifications & vessel sheets	Engineered heat balance based on engineered PFD	-5% to +15%

Class I — Detailed Estimate*
Contractor's or Check Estimate or Bid/Tender
based on complete data, typically for discrete parts or sections of the total project using deterministic estimating methods

Process Flowsheet	Equipment Mapping	Utility Requirements	Accuracy
Detailed piping & instrumentation diagram (P&ID)	Detailed engineered specifications	Engineered heat balance based on detailed P&ID	-5% to +15%

*Name and accuracy of estimate according to ANSI Standard Reference Z94.2-1989

Figure 2 Classes of capital cost estimates and selected required information for respective types, based on AACE International [4], including typical purpose of estimate, short description, required level of detail in process definition (process flowsheet, equipment listing and utilities requirements), as well as expected accuracy of estimate.

The level of detail of the CESTEA methodology presented in this paper is between Class III and IV: cost estimation is based on a PFD or a process simulation, which allows equipment mapping, sizing and material specifications, as well as a determination of utilities. The actual cost estimation involves deterministic estimating methods for major equipment cost at an assembly level rather than individual components. Factoring and other stochastic methods are used to estimate less significant equipment. Fixed Costs of Production (FCOP) are also calculated using stochastic methods (Class IV). The Variable Costs of Production (VCOP) are estimated more precisely (Class II/III) because they allow a deterministic method to be employed (e.g. the real costs for raw materials, electricity, steam etc.).

2.1 Total Cost of Production (TCOP) Estimation

Figure 3 shows the general structure of the TCOP which is the sum of variable costs, fixed costs and depreciation of invested capital. VCOP are directly proportional to the plant capacity utilization and FCOP accrue independently of the plant operation rate [3]. Based on the TCOP estimation the LCOP as a value for annual, product specific TCOP for full capacity can be calculated:

$$LCOP = \frac{TCOP}{production\ rate} \tag{1}$$

The term capital expenditure (CapEx) is used as a counterpart for the overall investments to be made. This term can be equated with fixed capital investment (FCI), i.e. total capital investment (TCI) without working capital (WC) as defined in Section 2.1.1. The CapEx in the form of a depreciation is not necessarily to be considered in the LCOP calculations. This is due to the fact that the investment, although it must naturally be amortized in the first set of years of a plant's lifetime (amortization period, see also Section 2.1.5), would no longer have to be allocated to the TCOP at the end of this period. That also means that the comparison with existing, already depreciated plants would be unfairly distorted. In fact, investment costs represent foremost an investment barrier. That is especially the case for very high acquisition costs in the billions, as in the chemical industry and also in the case of large-scale PtX, BtX, PBtX, eBtX systems.

In literature, the term operating expenditures (OpEx) is frequently used. Since the term comes more from business and accounting, this terminology is not used consistently and rather confusing in the TEA context. When OpEX is used, it should refer to Cash Costs of Production (CCOP) as defined above with the amortization of capital investment in the form of a depreciation not yet considered.

The differentiation and cost allocation of VCOP and FCOP differs in the literature. Peters et al. claim that the TCOP (denoted as TPC) consists of manufacturing costs (VCOP and FCOP) and general expenses. All expenses directly associated with the manufacturing operation are included in the VCOP estimate. They include costs for direct operating labor, supervisory labor directly applied to the manufacturing operation, raw materials, its transportation, unloading, etc., utilities, catalysts, and solvents, plant maintenance and repairs, operating supplies, laboratory supplies, as well as royalties. The fixed charges are composed of operation independent costs as property taxes, property insurance and depreciation. Plant overhead costs are independent of production rate and include expenses related to the employees or facilities indispensable and yet not directly related to production. The costs related to the recreational facilities, medical equipment, restaurants storage facilities, etc. and the salaries paid to their staff are included in the overhead costs. General expenses accrue for administration, distribution and marketing and research and development. [2]

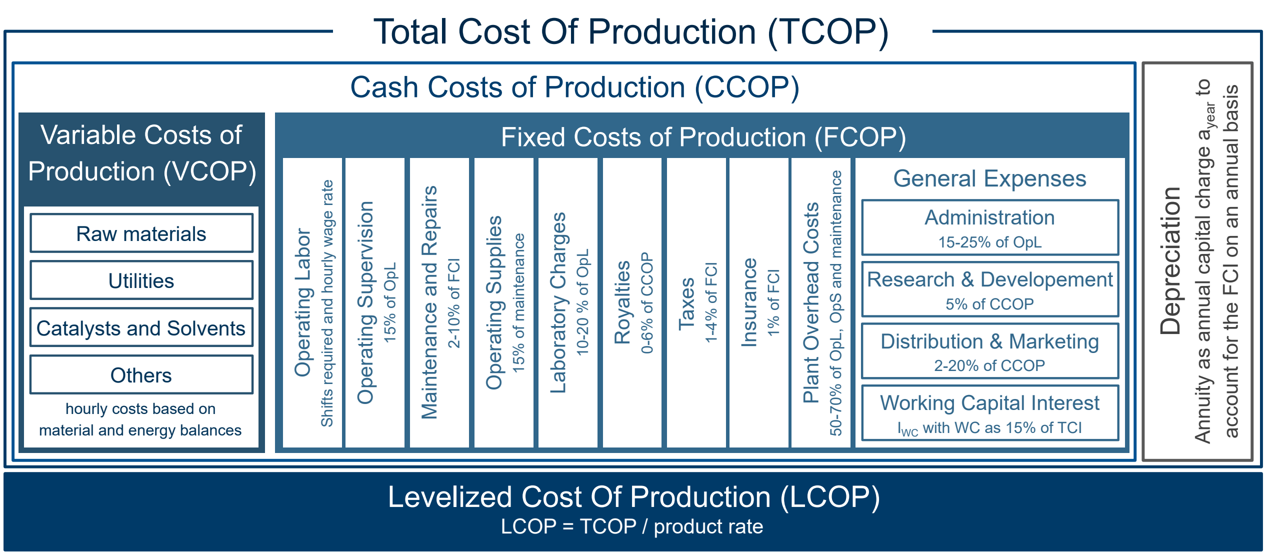


Figure 3 Overview of Total Cost of Production (TCOP) structure in CESTEA adapted from Peters et al. [2], Towler and Sinnott [3], including cost factor ranges and estimates from [2] for Fixed Costs of Production (FCOP).

Towler and Sinnott [3] also distinguish between variable costs and fixed charges but the exact delimitation is different as in Peters et al. [2]. The TCOP are calculated as the sum of VCOP, FCOP and the amortization of the capital investment. In contrast to Peters et al. [2], labor, royalty and maintenance costs are included in the fixed charges and are not seen as variable costs. Towler and Sinnott [3] justify this for maintenance costs by claiming that the costs are often regardless from the plant production rate, because equipment wear is mainly caused by changes in the production rate and not stationary operation [3]. Plant overhead costs and general expenses Towler and Sinnott also include in the Fixed Costs of Production [3]. Another difference in the methodology is that Towler and Sinnott consider depreciation of the investment costs as a single cost factor denoted as annual capital charges and do not include the depreciation in the fixed charges [3].

2.1.1 Capital Cost Estimation

In preparation for any industrial plant's operations, substantial financial resources are required to procure and erect the necessary equipment. The underlying cost estimate is based on the cost of the equipment required [2]. The sum of all fixed capital investment (FCI) combined with the working capital (WC) to be retained is referred to as the total capital investment (TCI).

FCI is the cost for designing, acquisition, and construction the necessary production and operating equipment. According to Peters et al., this cost can be further divided into manufacturing FCI, also known as direct cost, and non-manufacturing FCI, also known as indirect cost [2]. Towler and Sinnott further differentiate the direct costs into the inside battery limits (ISBL) investment, encompassing the cost of the plant itself, and the offsite or outside battery limit (OSBL) investment, accounting for modifications and improvements that must be made to the site's infrastructure [3].

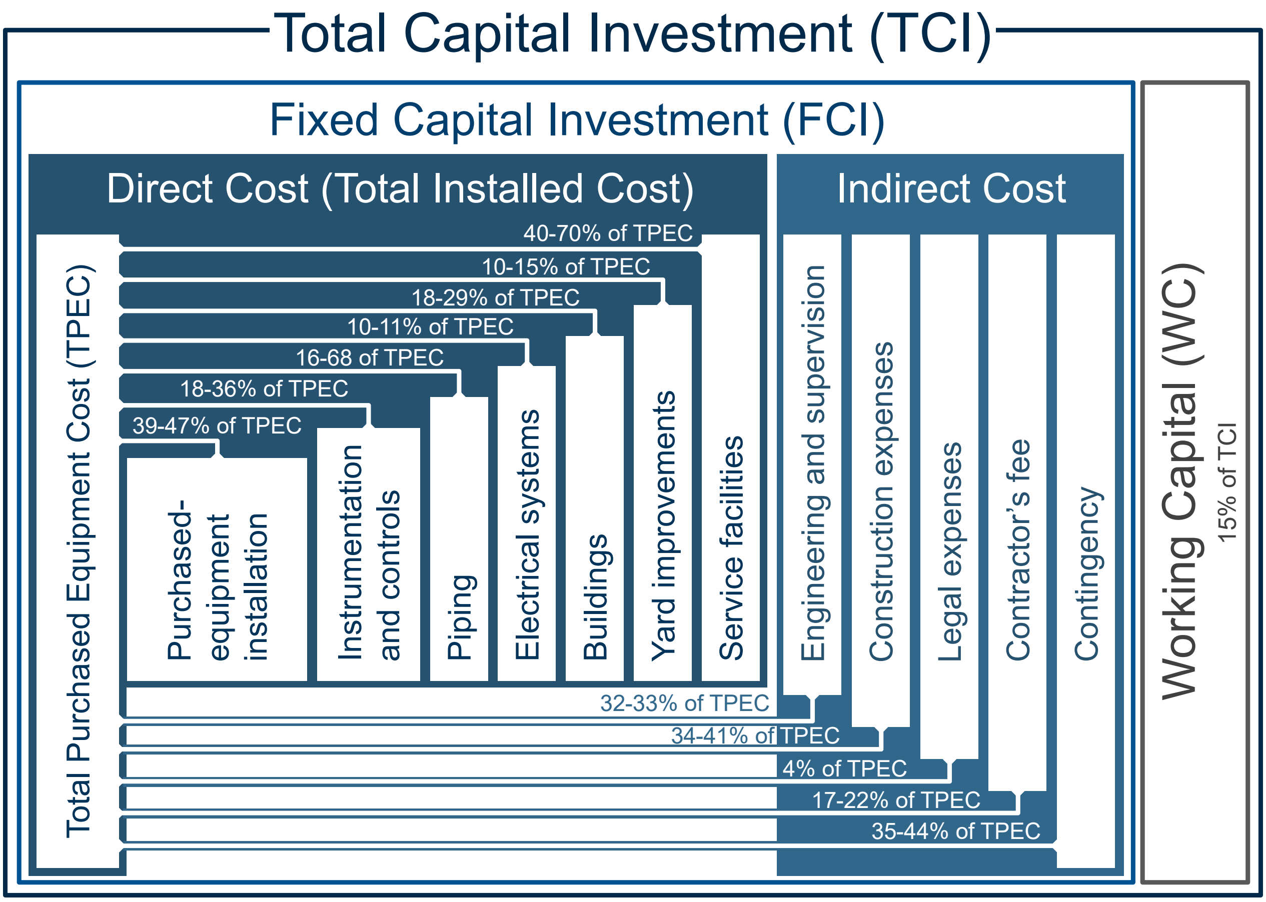


Figure 4 Overview of CESTEA Total Capital Investment (TCI) structure using the factorial method. Ratio factors based on Total Purchased Equipment Costs (TPEC) according to Peters et al. [2].

WC is the amount of additional financing required beyond the cost of building the plant to put the plant into operation and which is tied up in maintaining plant operation [3]. An industrial company's WC consists of the total amount of raw materials, consumables and supplies held in stock, as well as finished products and semi-finished products in the manufacturing process. In addition, there are accounts receivable, cash on hand for the monthly payment of operating costs such as wages and salaries, purchases of raw materials as well as accounts payable for goods, services and taxes. As customers are generally granted a payment term of 30 days, the WC required for trade receivables generally corresponds to the production costs for one month of operations. [2]

While Towler and Sinnott recommend between 5 % and 30 % of FCI for estimating WCs, with 15 % being a typical value for general chemical and petrochemical plants [3], Peters et al. suggest 10 to 20% of TCI [2]. However, it should be noted that for a higher level of detail, WC would be best estimated using production costs rather than capital investment [3].

There are several methods to estimate TCI with different accuracy levels including power factors applied to the plant capacity ratio, investment cost per unit of capacity and turnover ratio [2]. In this paper, the factorial method is presented to derive the FCI while WC is estimated as 15 % of TCI. This method of cost estimation based on Lang [5] uses the total installed costs and indirect costs as a function of the Total Purchased Equipment Cost (TPEC) [2, 3]. Figure 4 shows the respective cost structure of the TCI, including FCI as percentage of TPEC based on Peters et al. [2].

In this method, TPEC is used as sole basis to calculate all direct cost such as the provision of service facilities and the complete construction and installation of the plant, including piping, controls, services, etc., (these total investment costs are summarized as total installed cost) and indirect cost for engineering, construction, legal, contractors and contingency. The factors used are best determined based on the type of process, complexity of construction, materials required, location of the plant, past experience, and other factors that depend on the particular plant. The values used here are based on averages for various percentages for typical chemical plants according to Peters et al. and are shown in Table 1. The expected uncertainty is realistically in the range of ±30 %. [2]

Thus, to estimate FCI, TPEC must be estimated, which is the sum of all major equipment costs EC_i without installation, piping etc. according to Equation 2.

$$TPEC = \sum EC_i \tag{2}$$

Towler and Sinnott first calculate the ISBL costs via Equation 3 in their FCI estimation [3].

$$C_{ISBL} = \sum_{i=1}^{i=M} EC_i((1 + f_P)f_M + \sum f_j) \tag{3}$$

Other ratio factors (denoted as "Lang factors") f_j are surcharge factors on the major equipment costs EC_i to estimate the further costs for piping f_P, erection and installation, electrical, instrumentation and control as well as civil, buildings and structures [3]. If equipment is manufactured from stainless steel and the equipment costs EC_i are deployed for carbon steel, a material factor f_M of 1.3 is added [3]. In Table 2, the respective ratio factors are described and typical values for solid, solid-fluid and fluid processing plants are entered similar to the ratio factors from Peters et al. in Table 1. The fixed capital investment C_{FCI} is calculated with the following formula [3]:

$$C_{FCI} = C_{ISBL}\,(1 + \mathrm{OS})\,(1 + \mathrm{DE} + \mathrm{X}) \tag{4}$$

Table 2 also provides factors to calculate the offsite OS, design and engineering DE and contingency X costs [3].

The FCI of a fluid processing plant is 5.82 times higher as the purchased equipment costs, which is about 15 % more as the factor of 5.04 which is suggested in table Table 1. The main difference in the cost factors is that Towler and Sinnott apply factors for design and engineering and contingency on the summed ISBL and OSBL costs, which results in high design and engineering costs of 1.25 times EC (Table 1: 0.33 EC). The values of the other cost factors lay close to each other and the factors only slightly deviate in their denotation. Overall, the methodology of the FCI determination in Towler and Sinnott and Peters et al. is similar with Towler and Sinnott being slightly more conservative leading to higher FCI. The CESTEA method uses cost factors proposed by Peters et al. as shown in Table 1 [2].

2.1.2 TPEC Estimation Methods and Application in CESTEA

In general, there are four main options to determine EC based on the equipment specifications as shown in Figure 5. Each method has its advantages and limitations, and the choice depends on the specific project requirements and constraints (see Section 3.2). The choice of TPEC estimation method should consider the level of detail required, the availability of data, and the desired accuracy of the cost estimate.

Recent data on actual quotations from manufacturers is the most accurate source of purchased equipment costs and thus recommended. If no such data is available, the fabrication method provides an unbiased estimate. The overall equipment cost is calculated by summing component costs, machine costs, labor costs, and adding suitable amounts for supervision, overhead, and manufacturer's profit [3].

Utilizing cost estimates derived from previous orders offers a satisfactory level of accuracy while maintaining reasonable effort. When used to price new equipment, adjustments must be made in terms of sizing by employing a so-called scaling factor SF and in terms of price development by considering the relevant cost index ratio. For this so-called Cost Escalation Method Equation 5 is used, where EC_{ref} refers to the actual reference component costs, S_i is the scale of the employed equip-

Table 1 Ratio factors for estimating Fixed Capital Investment (FCI) from Total Purchased Equipment Cost (TPEC) as used in preliminary estimates and studies for solid, solid-fluid and fluid processing plant from Peters et al. [2].

		All	**Solid**	**Solid-fluid**	**Fluid**
		% of FCI		% of TPEC	
Direct Investments	Investment for purchased equipment	15-40	100	100	100
	Purchased-equipment installation	6-14	45	39	47
	Instrumentation and controls (installed)	2-12	18	26	36
	Piping (installed)	4-17	16	31	68
	Electrical systems (installed)	2-10	10	10	11
	Buildings (including services)	2-18	25	29	18
	Yard improvements	2-5	15	12	10
	Service facilities (installed)	8-30	40	55	70
	Total Direct Investment (TDI)		269	302	360
	Total Installed Cost (TIC)			TDI×TPEC	
Indirect Investments	Engineering and supervision	4-20	33	32	33
	Construction expenses	4-17	39	34	41
	Legal expenses	1-3	4	4	4
	Contractor's fee	2-6	17	19	22
	Contingency	5-15	35	37	44
	Total Indirect Investment (TII)		128	126	144
	Indirect cost (IC)			TII×TPEC	
Fixed Capital Investment (FCI)		TIC + IC	397	428	504
Working Capital (WC)		15% of TCI	70	75	89
Total Capital Investment (TCI)		FCI + WC	467	503	593

ment and S_{ref} the scale of the reference equipment. Here, costs are updated to the the current payment year $year_i$ using the Chemical Engineering Plant Cost Index (CEPCI). A respective currency exchange rate can be used to convert the costs to the desired currency.

$$EC_i = EC_{ref} \cdot \left(\frac{S_i}{S_{ref}} \right)^{SF} \cdot \frac{CEPCI_{year_i}}{CEPCI_{ref}} \quad (5)$$

The employed CEPCI consists of a composite index assembled from a set of four sub-indexes as shown in Figure 6. Each index and subindex is the weighted sum of several components mostly corresponding to Producer Price Indexes, updated and published monthly by the U.S. Department of Labor's Bureau of Labor Statistics. The index is employed primary as a process plant construction index, was established using a base period of 1957-1959 as 100. It was last revised in 2002, to account for changes in labor productivity. [6] The CEPCI is updated monthly and published in each issue of Chemical Engineering [7]. If reliable cost data is lacking, cost curves and correlations by Peters et al. [2], Towler and Sinnott [3] or others, or software solutions can be used for preliminary estimates.

To account for different scenarios with varying unit sizes, the cost escalation method (Equation 5) is recommended to determine equipment cost EC_i in the BtX/PtX framework. Depending on the technology, single sources (e.g. standard equipment, reactors etc.) or several cost references (EC_{ref}) from literature, manufactures or former quotations for the same unit operation (e.g. biomass gasifier, entire gas cleaning section) are employed (see Section 2.2). However, regarding consistency throughout the analysis, it is best to opt for a single reference for equipment costs. References with sizes (S_{ref}) more than ten times smaller or larger than the plant equipment size (S_i) are to be excluded to narrow down the selection of relevant references for more accurate estimates.

Scaling factors SF in equipment cost estimation are crucial as they represent the equipment's economy of scale behavior. It is important to understand the technology-specific scaling to ensure that these factors accurately reflect the cost dynamics of each component within BtX, PtX, e-/PBtX processes. These factors depend on various factors such as size, capacity, complexity, and technology maturity. For equipment showing significant cost reduction with size, the scaling factor may be lower. Other technology might, however, not follow the same principle. For example, in the case of electrolysis, stacking multiple cells may not result in significant cost reduction, while chemical reactors show massive cost reduction with size (Section 2.2.4).

Table 2 Ratio factors for estimating Fixed Capital Investment (FCI) from equipment cost (EC) as used for solid, solid-fluid and fluid processing plant from Towler and Sinnott [3].

	Solid	**Solid-fluid**	**Fluid**
Purchased equipment costs	EC	EC	EC
f_{er} Equipment installation	0.6	0.5	0.3
f_P Piping	0.2	0.6	0.8
f_I Instrumentation and control	0.2	0.3	0.3
f_{el} Additional electrics (e.g. lighting)	0.15	0.2	0.2
f_C Civil engineering work	0.2	0.3	0.3
f_X Buildings and structures	0.1	0.2	0.2
f_L Lagging/insulation and paint	0.05	0.1	0.1
ISBL costs C_{ISBL}	2.5 EC	3.2 EC	3.2 EC
OS Offsites	0.4	0.4	0.3
DE Design and Engineering	0.2	0.25	0.3
X Contingency	0.1	0.1	0.1
OSBL costs C_{OSBL}	0.4 C_{ISBL}	0.4 C_{ISBL}	0.3 C_{ISBL}
Fixed capital investment C_{FCI}	4.55 EC	6.05 EC	5.82 EC

Furthermore, recognizing the limitations on the size of specific equipment is essential. While some equipment, like chemical reactors, may follow economies of scale principles, there are practical limits to their size. Larger sizes might require the use of multiple units or a train of reactors. Different types of equipment may have different maximum sizes. For instance, in gasification, fluidized bed gasifiers may have a practical limit up to 50 MW_{th}, while entrained flow gasifiers can be built to sizes exceeding 500 MW_{th} [8]. Since understanding these limitations is crucial for accurate scaling, detailed research into the capacity limitations of specific equipment is necessary.

In CESTEA, standard equipment costs, i.e. costs of compressors, heat exchangers and pumps, are estimated based on correlations in Towler and Sinnott [3]. Other equipment costs are mainly calculated from literature values by utilizing the cost escalation method. Adapted from Towler and Sinnott [3], a material factor of 1.3 is applied on the TPEC including costs for piping, if the equipment is to be manufactured from stainless steel and the correlation is originally for carbon steel [3]. Fixed capital investment costs of innovative technologies as Direct Air Capture (DAC) and large-scale water electrolysis are estimated with specific investment costs from the literature as described in Section 2.2. To estimate the equipment costs EC_i, precise material and energy balances and equipment operating characteristics such as dimensions, specifications, and operating parameters of the equipment are required which can best be determined from process simulations.

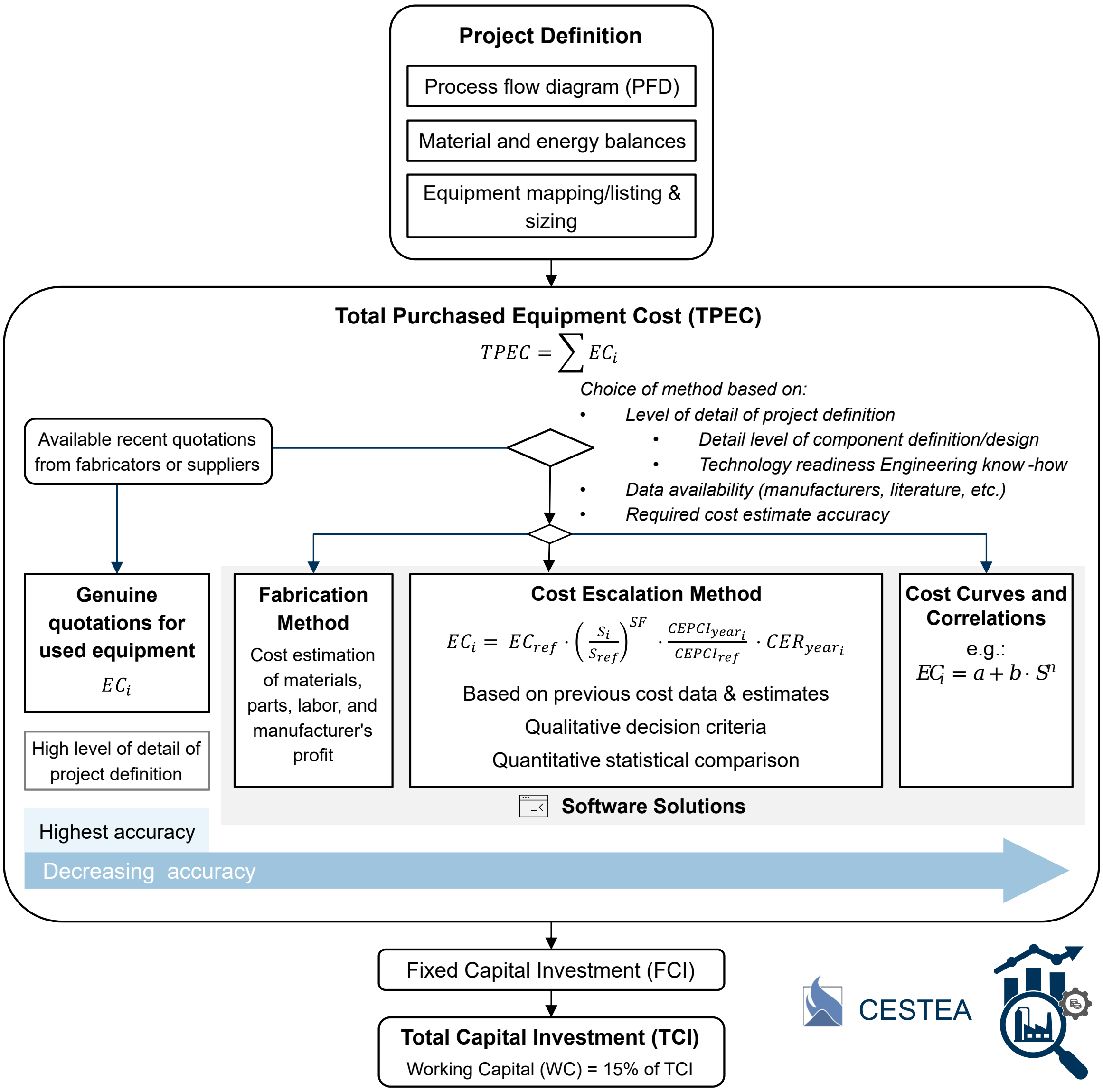


Figure 5 Overview of CESTEA methodology to estimate capital costs.

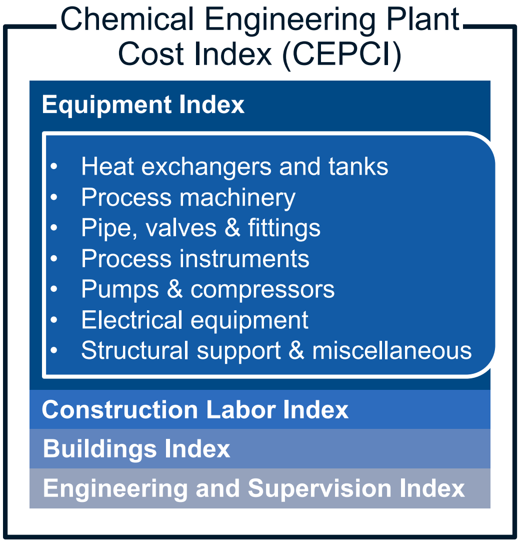


Figure 6 Overview of general Chemical Engineering Plant Cost Index (CEPCI) structural including sub-indexes according to [6].

Leveraging data from entire facility cost estimates can be used to check for plausibilty or even validate the cost estimate. This methodology involves deriving your own equipment costs for individual units, summing them up to obtain the TPEC for the entire pretreatment section, and then comparing this cost estimate to one where someone has estimated the entire section as a whole.

To derive the TPEC, equipment costs (EC_i) are summarized after ratio factors (see Figure 4) are applied to determine the FCI and TCI as demonstrated in Figure 5. The proposed methodology in this work focuses on the investment and production costs of the considered renewable chemicals. Regarding the investment decision these are the important factors to consider because the investment costs can be seen as the capital at risk while the LCOP are the minimal selling price of the product. Revenue estimation is not considered because revenues highly depend on fluctuating market prices and individual purchase agreements resulting in high uncertainty. Similarly, a cash flow analysis is excluded as it implies assumptions about revenue streams.

2.1.3 Variable Costs of Production

According to Peters et al., variable production costs include all the expenses related to the production of the product, encompassing operating labor and supervision costs, expenses for raw materials, utilities, catalysts and solvents, maintenance and repairs, as well as other costs for operating supplies, laboratory charges and royalties [2]. Towler and Sinnott include operating labor and supervision costs, as well as maintenance and repair expenses in the fixed-charges, nevertheless the general approach for determining variable costs is similar [3]. In standard economics, any labor costs are often not seen as variable costs at all. Consequently, this method follows the methodology approach by Towler and Sinnott for the determination of the VCOP as presented in Figure 3 [3].

Raw materials, Utilities, Catalysts and Solvents
In the chemical industry, the procurement of raw materials represents a significant cost factor in production operations. The term ”raw materials” generally encompasses materials that are directly consumed in the manufacturing process to produce the final products. [2] As mentioned above, the amounts of raw materials, utilities and consumables are determined based on process energy and material balances typically from the process simulations. Accurate material balances of the process are essential to establishing the process raw materials requirements [2, 3].

Obtaining direct price quotations from potential suppliers is preferable when estimating raw material costs. In cases where such quotations are not accessible, published prices are utilized. During preliminary cost analyses, market prices can also serve as a basis for estimating raw material costs. In instances where transportation charges are applicable, they should be incorporated into the raw material costs. Typically, these charges are estimated at around 10 % of the raw material cost, although they can vary significantly. The proportion of raw material costs to the total product cost differs greatly across different types of plants. In chemical plants, raw material costs generally fall within the range of 10-60 % of the TCOP. [2].

The cost for utilities and other consumables, such as steam, electricity, process and cooling water, compressed air, fuels, refrigeration, as well as waste treatment and disposal, depends on the required amount, plant location and source. Direct price rates from potential utility providers and suppliers are desirable for cost estimating purposes. Utilities and other consumable costs amount to about 10-20 % of the TCOP for typical chemical processes [2].

Consumable process materials as catalysts and solvents degrade in longer operation and have to be replaced with a specific periods [3]. The cost of catalysts and solvents can be significant and should be estimated based on the requirements and prices for the process in question [2]. The choice of catalyst is specific to the reaction pathways involved in BtX, PtX and PBtX processes. Different catalysts are used for various chemical transformations, such as Methanol or Fischer-Tropsch (FT) synthesis, and should be evaluated at an early stage of process development. Conducting a thorough cost-benefit analysis is essential to evaluate the economic viability of using specific catalysts and solvents. This analysis should consider both upfront costs and long-term performance. While the conversion efficiency of a catalyst in facilitating the desired reactions is a critical factor, the stability and lifetime associated with any catalysts is crucial. Different reactor technologies, such as microstructured catalyst coated reactors and slurry reactors, have unique effects on catalyst stability and lifetime. Mechanical stress, abrasion, and the nature of the reaction environment vary significantly. The reactor setup also massively influences the method of cyclic catalyst replacement. A microstructure, catalyst coated reactor, for example, would require different replacement options than a slurry reactor allowing continuous catalyst removal and replacement with makeup catalyst. Consideration should be given to the end-of-life implications of catalysts. Some catalysts may be suitable for recycling or regeneration, while others may require proper disposal methods. The same applies to solvents, for which the possibility of recycling and reuse should be examined.

In addition, recycling can lead to cost savings by reducing the need for purchasing new raw materials. Reusing materials within the production process can result in operational efficiencies and economic benefits. Other materials produced during the process may have market value as by-products. These could be materials with applications in other industries or products that can be sold for additional revenue. Materials generated in the process that cannot be recycled or sold as by-products are classified as waste and must be disposed of. In some cases, the proper management of waste involves treatment to meet environmental standards, concentration to minimize volume, and disposal methods that align with regulatory requirements and sustainability principles. [3]

2.1.4 Fixed Costs of Production

Fixed Costs of Production (FCOP) refer to the costs that are paid irrespective of the production rate. According to Peters et al., these charges include depreciation, tax, and insurance. General expenses are seen as independent cost factor contributing to the TCOP [2]. Towler and Sinnott exclude depreciation and additionally include labor costs, maintenance, royalties and plant overhead charges in the Fixed Costs of Production [3].

In this method, the FCOP composition is adapted from Towler and Sinnott [3] and the cost factors are taken from Peters et al. [2]. Thus, the FCOP include operating labor and supervision, maintenance, operating supplies, laboratory, laboratory charges, royalties, taxes, insurance, plant overhead costs and general expenses.

Operating Labor and Supervision In estimating operating labor costs, hourly wage rates are typically considered, varying across skill levels, industries, and locations. For chemical processes, operating labor generally constitutes around 1-20 % of the TCOP. Preliminary cost estimates (Class III) use either company experience with similar processes or published information on comparable processes to determine operating labor quantities. Scaling plant capacities up or down often involves a non-linear relationship between labor requirements and production rate, with a commonly used scaling factor of 0.2 to 0.25 for the capacity ratio [2]. Peters et al. also provide a methodology for typical labor requirements if a plant flowsheet is available based on either individual process equipment or principal processing steps on the flowsheet. Alternatively they propose rules of thumb either based on the production capacity of the plant or based on experience with similar processes. [2] However, with advancements in technology and automation within the process industry, relying solely on relating employee-hour requirements to equipment, processing steps, or production quantities can lead to inaccurate results. It is crucial to use up-to-date data to ensure accuracy in estimating employee-hour requirements.

This method uses a diagram from Peters et al., where the operating labor hours per day and process are plotted over the production capacity of a plant, to estimate the required labor hours [2]. Furthermore, the labor costs are assumed to be independent of the operating hours of the plant.

Direct supervisory labor is an essential component of manufacturing operations, with its quantity closely linked to the total operating labor, complexity of the operation, and product quality standards. Its cost can be estimated to approximately 15 % of the operating labor cost. When operating at reduced capacities, supervision costs generally remain fixed at the 100 % capacity rate. [2]

Maintenance and Repairs In the process industries, the annual total plant cost for maintenance and repairs typically ranges from 2-10 % of the fixed capital investment [2], with 7 % considered a reasonable value [2]. For operating rates below the plant capacity, the maintenance and repair costs are estimated at 85 % of the cost at 100 % capacity for a 75 % operating rate, and 75 % of the cost at 100 % capacity for a 50 % operating rate [2]. Towler and Sinnott [3] estimate the maintenance costs to be lower with 3-5 % of the ISBL costs. Since the FCI of PtX, PBtX and BtX processes is very high and the electrolyzer stack replacement is to be considered independently, the maintenance and repair costs are estimated to 3 % of the FCI.

Operating Supplies Consumable items, including lubricants, test chemicals, custodial supplies, and similar provisions, fall under the category of operating supplies. The annual cost associated with these supplies typically amounts to about 15 % of the total cost for maintenance and repairs. [2]

Laboratory Costs The expenses related to laboratory tests for operational and product quality control are typically estimated by considering the employee-hours involved and multiplying them by the applicable rate. As a rough estimate, these expenses may account for about 15 % of the operating labor costs. [2]

Royalties Since patents are necessary to protect manufacturing processes and products, the purchase of patent rights or license fees based on the amount of material produced is required. An approximate range of patent and royalty costs for patented processes is 0-6 % of the total production costs without considering depreciation (CCOP). Because of many involved process steps in PtX, PBtX and BtX processes, in CESTEA, the royalties are set to 4 % of the CCOP. [2]

Plant Overhead Costs Additional costs emerge for medical care at the plant, safety measures as property protection, employee salary overhead, facilities as restaurants and other minor interventions. All these additional costs are included in the plant overhead costs, which are assumed to be 60 % of the sum of operating labor, supervision and maintenance costs. In this consideration, replacement of electrolyser stacks is excluded from the maintenance costs. [2]

Taxes and Insurance The assessment of local property taxes is contingent upon the specific location of the plant and the governing regulations within the region. Typically, plants situated in densely populated areas incur annual property taxes of approximately 2-4 % of FCI while property taxes ranging from 1-2 % of FCI can be estimated for sparsely populated areas, which leads to 2 % of FCI being a reasonable value regardless of the location. The determination of property insurance rates relies on factors such as the nature of the manufacturing process and the level of available protective facilities. Here, these rates are estimated to around 1 % of the FCI. [2]

General Expenses While executive and administrative expenses cannot be directly allocated to manufacturing costs, it is crucial to include these costs for a comprehensive TEA. Administrative expenses encompass salaries and wages of administrators, secretaries, accountants, computer support staff, engineering and legal personnel, along with costs associated with office supplies, equipment, external communications, administrative buildings, and other overhead items related to administrative activities. The magnitude of these costs can vary significantly between plants and may depend on whether the plant is a new establishment or an expansion of an existing one. In the absence of precise cost data from company records or for preliminary estimates, administrative costs can be approximated as 20 % of operating labor. [2]

Constant advancements in the chemical industries lead to the continual development of novel methods and products through research and development activities. To maintain a competitive position, forward-thinking companies allocate resources to research and development expenses. These expenses encompass various aspects, including salaries and wages for personnel directly engaged in research and development, fixed and operating expenses for machinery and equipment, material and supply costs, and fees for consultants. In certain sectors such as pharmaceuticals, research expenses can dominate the overall product cost. Within the chemical industry, research and development costs typically represent approximately 5 % of CCOP. [2]

Distribution and marketing expenses encompass salaries, wages, supplies, and other costs related to sales offices, sales representatives' salaries, commissions, and traveling expenses, shipping expenses, container costs, advertising expenses, and technical sales services. The magnitude of distribution and marketing costs varies significantly across different plant types, depending on factors such as the specific material being produced, other products sold by the company, plant location, and company policies. Typically, these costs range from 2-20 % of CCOP for most chemical plants. The higher end of the range typically applies to new or low-volume products targeting a large customer base, while the lower end is observed for high-volume products like bulk chemicals. For liquid hydrocarbon fuels 3 % of CCOP seems to be suitable. [2]

Working capital interest costs As highlighted in Section 2.1.1, WC refers to the extra funds needed to start and maintain plant operations. From a cash flow perspective, WC is recovered at the end of the plant's life because it doesn't depreciate. This leads to WC often being omitted from TCOP or LCOP calculations in many TEA methodologies, overlooking the fact that these funds are bound through the entire plant lifetime. Assuming similar costs of capital as for other investment costs, the interest on the WC throughout the plant's lifespan must be considered. Consequently, we include the interest costs of WC as part of the general expenses according to Equation 6, using the interest rate i and plant lifetime PL.

$$I_{WC} = WC \cdot ((1+i)^{PL} - 1) \tag{6}$$

2.1.5 Depreciation

The establishment of a manufacturing plant requires an initial investment in equipment, buildings, and other tangible assets, which is recovered through the allocation of depreciation as a part of the TCOP. A fixed percentage of FCI can be charged as depreciation consistently over a predetermined number of years. This constant annual depreciation over a fixed duration allows for easy integration into the TCOP since it does not require cash flow analysis over a number of years. It should be noted that this method is only appropriate when the time value of money is not a factor to be considered. [2]

Alternatively, the Modified Accelerated Cost Recovery System (MACRS) is a commonly employed approach for calculating depreciation, resulting in varying annual depreciation amounts. MACRS allows for greater accelerated depreciation allowing businesses to deduct greater amounts during the first few years and relatively less later. [9]

MACRS is designed to provide businesses with a systematic way to recover the cost of their investments over time as a critical component of tax planning and financial management for businesses that own and use depreciable assets. It is important to note that cash-flow analysis, on the other hand, involves a broader examination of a company's operational and financial activities. Other factors, such as operating expenses, financing activities, and changes in working capital, also play significant roles in cash flow analysis. While both, MACRS and cash flow analysis, are integral components of financial management, the application of any of the two in TEA should be guided by a clear understanding of the project's context, assumptions, and the specific goals of the analysis. It is crucial to maintain a balance between the level of detail required for informed decision making and the resources available for analysis.

In design estimates, an interest rate is established as a cost if external sources are required for borrowing the necessary funds. Thus, interest is considered a definite cost when funding for plant investments relies on borrowed capital. Annual interest rates typically range from 5-10 % of the total value of borrowed capital. [2]

Here, we employ a linear depreciation using the annualized cost method. In the annualized cost method, a fixed interest rate i is used to calculate the so-called annual capital charge ratio or annuity factor [2, 3]. Applying the annuity factor on the FCI, the net present value of the investment is linearly distributed over the desired time of amortization n_A ($\leq$ plant lifetime PL), resulting in annual payments the so-called annuity or annual capital charge. Plant life time is not equivalent to the considered amortization period n_A, since generally its preferable to pay the investment back as fast as possible to reduce the total loan interest. The annual depreciation a_{year} can be calculated according to Equation 7. The working capital is recovered after the last year of operation, which is why the working capital is not considered in the depreciation but the payed interest contribute to the FCOP.

$$a_{year} = FCI \cdot \frac{i(1+i)^{n_A}}{(1+i)^{n_A} - 1} \tag{7}$$

2.2 Process Specific Equipment Cost Estimation

Understanding the BtX, PtX, and e-/PBtX processes is crucial for accurate equipment cost estimation. Figure 7 presents a streamlined block flow diagram illustrating the main stages involved in the processes.

Typically, BtX operations necessitate an initial pretreatment phase for biomass, preparing it for subsequent conversion into syngas via gasification. The composition and quality of the syngas, heavily influenced by the biomass feedstock and the chosen conversion technology, contain various impurities that significantly impact the catalysts used in the synthesis phase. This necessitates rigorous gas cleaning procedures, especially for the removal of sour gases, to ensure the syngas is suitable for further conversion into high-value products. An additional step often involves adjusting the H_2/CO ratio in the syngas through water-gas shift (WGS) reaction, as it is typically too low to directly utilize. This adjustment is necessary for the production of liquid fuel fractions via pathways such as Methanol-to-Hydrocarbons, Fischer-Tropsch synthesis or methanation. [10]

Conversely, PtX processes have less complex gas purification requirements. Hydrogen production is achieved through electrolysis, and CO_2 capture is necessary, either from concentrated sources like power plant flue gases or the cement industry, or from the atmosphere via direct air capture (DAC). Depending on the targeted synthesis, hydrogen and CO_2 can be directly converted, or CO_2 may first need transformation into CO through reverse WGS (rWGS). [11]

Hybrid PBtX or eBtX processes address key challenges inherent to the above. Notably, syngas derived from BtX processes typically exhibits a hydrogen deficiency, which leads to carbon losses during the syngas conditioning phase due to the WGS reaction. On the other hand, PtX processes are constrained by the high energy demands of carbon capture technologies and large hydrogen demands. The electrification of the BtX process, for example through integration of hydrogen produced via electrolysis, effectively mitigates these issues. [1]

2.2.1 Solid Fuel Conversion

Solid fuel handling for use in BtX processes is a complex topic highly depending on the employed biomass feedstock and its characteristics as well the employed gasification technologies and its requirements. The production of chemicals and fuels through biomass gasification has been extensively researched and economically evaluated. Dieterich et al. provide an overview of current technical and economic studies [12].

The capital cost estimation for each step in this process involves a careful consideration of various factors. Different biomass feedstocks (e.g., wood, agricultural residues, energy crops) may require different pretreatment approaches. Various pretreatment technologies exist, including physical, chemical, and biological methods. Each method has its own capital cost implications. For example, physical methods may involve drying, thermal treatment and size reduction, while chemical methods may include acid or alkali treatments. The choice of gasification technology (e.g., fluidized bed, fixed bed, entrained flow), as well as the integration of pretreatment and gasification process, will impact the choice and requirements of the pretreatment technology. Capital costs in the pretreatment section may also include equipment for biomass collection, transportation, and storage.

After technology selection, and process design, e.g. in the form of a process model, exist, estimating the costs for solid fuel conversion, especially in the context of pretreatment and gasification, is challenging, particularly when using software tools or standard cost correlations. Unlike for fluid materials, solid fuels are hardly applicable to standardized methodology or generic cost correlations, due to their specific feedstock characteristics. Cost correlations may be based on technologies that were developed for completely different processes (e.g. drying for paper industry, milling and gasification of coal instead of torrefied biomass etc.), while BtX technologies typically involve advanced or novel approaches, that may not be well-

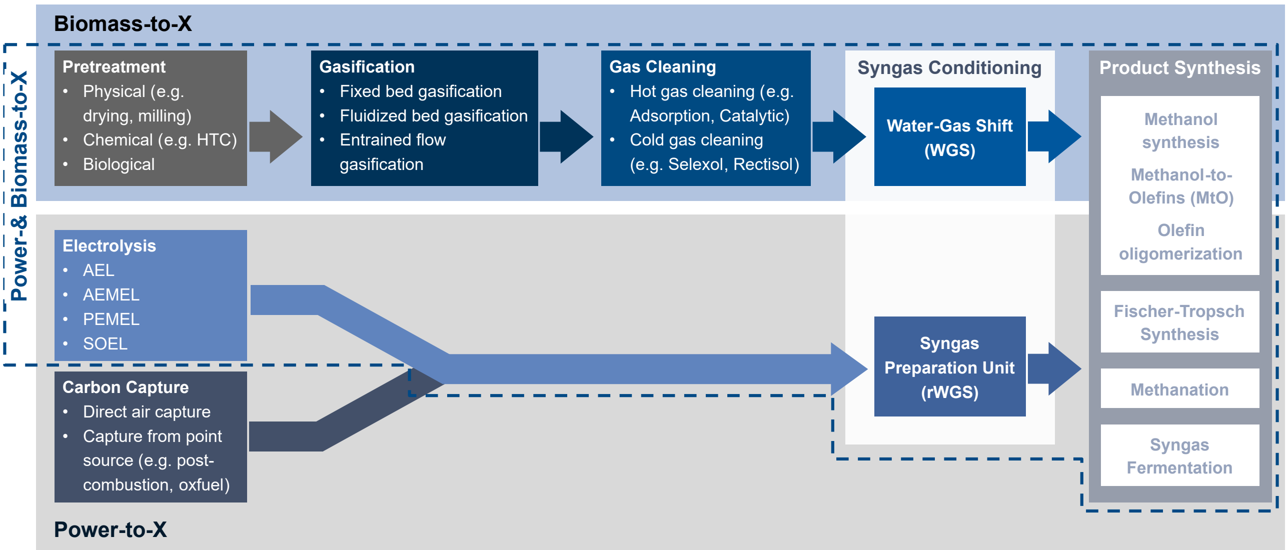


Figure 7 Simplified block diagram of BtX, PtX and e-/PBtX process scheme.

represented in existing cost databases or correlations. Furthermore, scaling up solid fuel conversion processes can be challenging due to factors such as heat and mass transfer limitations, potential reactor design changes, and variations in feedstock properties at larger scales. These challenges may not be adequately addressed by standard cost correlations.

Given these challenges, a customized approach to cost estimation is required for solid fuel conversion processes. The proposed best practice for cost estimates tailored to the specific characteristics of the feedstock and process is to use literature data from detailed cost estimating studies using the cost escalation method and general methodology introduced above. In the following, we present what we consider to be the most reliable cost references from the literature. Cost estimates using cost correlations from literature [2, 3] or simulation software such as APEA are best for standard equipment such as heat exchangers, pumps, compressors and commercial separators. Based on this literature review, a default scaling factor of 0.7 for all major solid fuel equipment (dryer, torrefaction reactor, mill, gasifier) seems reasonable, if not stated otherwise by manufacaturers or literature. While for all major pretreatment equipment (dryer, torrefaction reactor, mill), FCI and TCI are best derived using the "solid plant type" ratio factors presented Table 1, "solid-fluid plant type" ratio factors are best used for the gasifier itself.

Biomass Preparation, Handling, and Pretreatment Given the wide range of options for biomass handling and pretreatment, not all options can be discussed in detail here. The typical process chain for entrained-flow gasification will therefore be considered as an example.

Tipping bunkers and fuel processing are required as a general unit for a BtX pretreatment plant. Svanberg et al. provide a good cost estimate for that entire section [13]. The following process consists of a dryer with subsequent gas-solid separation, a torrefaction reactor with subsequent gas(-liquid)-solid separation, a solid cooler and a mill for particle size reduction. Swanson et al. [14] estimate the cost of a biomass rotary dryer based on Couper [15]. Mobini et al. estimate the total capital investment for a pellet mill [16] as well as for a torrefaction process including torrefaction reactor, separation steps, torgas burner and torcoal cooler [17]. Hannula provides capital cost estimates based on a self-consistent set of component-level capital cost data including feedstock handling and belt drying [18]. Cost data is based on literature sources, vendor quotes and discussions with industry experts, as well as estimates based on similar equipment and engineering judgment [18]. Seider et al. provide good equipment cost data for pneumatic conveyors, and dust collectors (electrostatic precipitator, bag filters, cyclones, etc.), fired heater for pyrolysis or torrefaction and milling (hammer mills, ball mills, etc) [19].

Hamelinck and Faaij [20] and Tijmensen et al. [21] reported cost estimates for biomass conveyers, grinding, storage, dryer, feeding systems . Batidzirai et al. collected cost data on biomass pretreatment including a chipper, rotary drum type dryers, a moving bed torrefaction reactor, a hammer mill, pelletizing, and a pellet mill [22]. Their cost data is combined with detailed capacity limits, primary references and unit specific scaling factors [22].

Bergman et al. also presented a cost estimate for an stand alone torrefaction plant including a rotary drum dryer, torrefaction, torgas fired heater, and an indirectly water cooled rotary drum product cooler. Their process design is based on both experimental and simulation work. For the torrefaction reactor, an indirectly heated screw reactor, a rotating drum and a moving-bed reactor are compared. [23]

Uslu et al. provide more cost estimates for torrefaction, fast pyrolysis and pelletization [24]. Uslu et al. and Andersson et al. provide an estimate of an entire "black-box" pretreatment section using torrefaction (Uslu et al. additionally investigate cost data for fast pyrolysis and pelletization) [24, 25]. Hannula also reports capital cost estimates for feedstock handling and drying [18]. Dieterich et al. include a good overview of cost estimates for the pretreatment section before entrained flow gasification [12].

Gasification One equipment typically not modeled explicitly in process simulation is the feeding system. a lockhopper system is the best setup used for pressurized feeding of dried biomass to the gasifier. Swanson et al. provides a good estimate using a fabrication method for lockhopper feeding systems based on coal processing for a previous IGCC project in 1993 [14]. Seider et al. provide an equation to calculate the cost of hydrocyclone used for slag separation [19]. Slag handling systems for handling slag is also listed in other studies [26–28].

The main costs in the gasification section are attributable to the gasifier itself. While there are numerous studies providing relatively good data for coal gasifier, little is known of the potential cost of large-scale biomass gasifiers, especially when it comes to entrained flow gasification. Most studies use a scaling factor of 0.7 which is recommended here. As the following brief literature review shows, even recent TEA publications are generally based on cost estimates dating back up to 25 years.

For biomass gasification, Williams et al. and Faaij et al. provided the first fluidized bed gasifier cost estimates in 1995 and 1998 [29, 30]. Tijmensen et al. provide cost estimates for biomass feeding systems for gasification purposes, as well as for the Batelle Columbus and IGT fluidized bed gasifier [21]. Larson et al. assume several cost reduction opportunities when biomass gasification moves from first-of-its kind to commercially available technology. The learning effects considered include eliminating specialized engineering services, making technological improvements, and reducing contingency costs [31]. Nexant Inc. evaluate costs of a low-pressure, indirect system using the BCL gasifier, and a high-pressure, direct system using GTI gasification technology [32]. Larson et al., also estimate cost for a pressurized oxygen-blown fluidized bed reactor based on the GTI gasifier design [33]. Hannula provides capital cost estimates based on a self-consistent set of component-level capital cost data for a pressurized O_2 CFB as well as an atmospheric steam fluidized bed gasifier including air separation unit (ASU) [18].

Most author estimate the cost for a biomass entrained flow gasifiers based on cost estimates for coal entrained flow gasifiers [14, 26, 34–38]. Williams et al. assumes capital costs of a biomass entrained flow gasifier based on the Shell coal gasifier design assuming that costs are the same as for the IGT fluidized bed gasifier as both are pressurized and the lower cost associated with the higher throughput of an entrained-bed design is assumed to be offset by the higher cast associated with higher temperature operation [29]. The US National Energy Technology Lab (NETL) conducted multiple studies on the economical performance of coal fuel IGCC plants since 1998 incorporating entrained flow gasifiers. Most of which include all balance of plant and oxygen supply via ASU cost estimates as well. Previous studies includes a series of capital cost estimates for entrained flow, oxygen-blown, slagging gasifiers (PED-IGCC-98-001 for GE/Texaco; PED-IGCC-98-002 for Shell, and PED-IGCC-98-003 for ConocoPhillips/Destec). Capital cost are estimated using APEA and designs include a comparison of raw syngas cooling and full water quenching. [34]

In 2001, Frey and Akunuri used the APEA to estimate the cost of coal IGCC using the Texaco design comparing syngas cooling and full quench [35]. Further IGCC studies on bituminous coal gasification provide cost estimates for Shell, ConocoPhillips and GE coal entrained flow gasifier capital cost [39].

The NETL Coal To Liquids Study performed in 2007 includes a cost estimate based on Illinois No. 6 coal and ConocoPhillips gasification technology. The employed gasifier consists of a two-stage, oxygenated, slagging entrained-flow gasifier. The syngas exiting the gasifier is cooled in a fire tube syngas cooler to produce high pressure steam and then passed into a water scrubber to remove particulates and trace components. [26]

A further TEA was performed in 2015 using coal and biomass gasification via oxygen-blown entrained flow gasification to produce FT fuels [37]. NETL supplied a cost update in 2022 based on internal references [38]. Kreutz et al. estimate the cost of an oxygen-blown entrained flow coal gasification system based on the ratio of the investment costs to those for a fluidized bed GTI coal gasifier at 2.1 [27]. Kerdoncuff estimate the total investment for a slurry,Andersson et al. that of a dry-fed pressurized entrained flow biomass gasifier [25, 40]. Other estimates including that of [28] and [41] highly underestimate entrained flow gasifier equipment cost. Dieterich et al. provide a comprehensive comparison for entrained flow gasifier cost estimates from literature [12].

2.2.2 Gas Cleaning and Acid Gas Removal

Gas cleaning in BtX, eBtX and PBtX processes is a crucial step to remove impurities and contaminants from the power-derived syngas or gasified biomass before further conversion. The main objectives of gas cleaning are to ensure the syngas meets the quality requirements for downstream processes and to protect the downstream equipment from fouling or damage. Gas cleaning typically involves particulate removal to ensure solid particles, ash, and dust are separated from the syngas stream. Methods such as cyclones, filters, or electrostatic precipitators are commonly used. Depending on the used gasifier technology, tar compounds, which can be harmful to downstream catalysts and equipment, are removed through processes like tar cracking, scrubbing, or catalytic conversion. The most elaborate cleaning step is acid-gas removal (AGR) where compounds such as hydrogen sulfide (H_2S) and carbon dioxide (CO_2) are removed to improve the syngas quality. Techniques such as amine scrubbing or pressure swing adsorption (PSA) are often employed. Additional steps may be included to further purify the syngas, depending on the specific requirements of the downstream processes.

Processes consisting of standard equipment such as compressors, pumps, flash drums, absorbers, and columns, are generally well-known and characterized by mature technologies with good data availability. Thus, a combination of fabrication method, cost escalation and cost correlations, as well as cost estimation software can be used to estimate the cost of purchasing equipment for gas cleaning processes. While software such as APEA allows for detailed input of equipment specifications and operating conditions to generate accurate cost estimates, the basic engineering and process flow diagram must sustain a high level of detail to account for major gas cleaning equipment. Furthermore, there may be some equipment or components within the gas cleaning section where APEA or similar methods might not be as suitable. For example, if the plant requires specialized or custom-designed equipment for unique process requirements, the cost estimation for such equipment may require more detailed engineering analysis or vendor quotes. Additionally, if there are significant variations in equipment specifications or operating conditions that are not well-captured by standard cost correlations, a more detailed cost estimation approach may be necessary. Depending on the employed technology, solvent or adsorbent costs must also be considered in the VCOP. Not only the ability of the solvent or adsorbent, but also the energy and capital costs required for its regeneration must be carefully considered [42]. The same generally applies to any type of cleaning and purification cost estimate.

Especially warm or hot gas cleaning technologies present specific challenges in cost estimation due to their specialized nature and the need for high-temperature operation. In hot gas filtration, for example, particles are removed from high-temperature gas streams using ceramic filters or other high-temperature materials, which necessitates suitable filter materials to be selected, design considerations for high-temperature operation and the need for special maintenance procedures. The same applies to ceramic heat exchangers, which are used to recover heat from high-temperature gas flows to increase energy efficiency. Cost estimation challenges include the selection of suitable ceramic materials, design for thermal expansion and contraction, and considerations for high-temperature corrosion. Cost estimation for catalytic hot gas cleaning employing catalysts to convert gaseous contaminants into less harmful or inert substances at high temperatures require the selection of appropriate catalyst materials, reactor design for high-temperature operation, and considerations for catalyst regeneration and replacement.

Pressure Swing Adsorption (PSA) is a gas separation technology used in gas cleaning processes and is commonly used for selectively removing impurities such as CO_2, H_2, or N_2 from gas streams. Since PSA systems can be complex, with multiple adsorption beds, valves, and control systems ,estimating the cost of each component and the overall system requires detailed knowledge of the design specifications. Furthermore, the cost of the adsorbent material used in the PSA system can vary depending on the type and quality of the material. Estimating the cost of adsorbent material requires knowledge of market prices and consumption rates. Couling et al. estimate cost for CO_2 removal based on a PSA model [42] Kreutz et al. estimate costs for a PSA system to remove H_2 from the syngas based on PSA bed size to be scaled according to the purge gas flow and not the H_2 flow [36]. Swanson et al. also provide a sizing and fabrication method for a PSA unit to remove H_2 from syngas. [14]. Rosner et al. compared syngas cleaning using Selexol wash and ZnO sorbent, CO_2 and H_2 PSA [43]. Tijmensen et al. also provide cost estimates for a ZNO guard bed for FT application based on the fabrication method [21].

2.2.3 Direct Air Capture

For PtX processes CO_2, can be provided directly from ambient air by utilizing Direct Air Capture (DAC) technologies, leading to climate neutral carbon supply. DAC is a technology in an early development stage and economic data is limited [44]. There are several DAC technologies, whereby the most developed processes are the liquid solvent approach and the solid sorbent processes [45, 46].

The solid sorbent technology has the potential of low carbon capture costs, because the working principle is quite simple and the regeneration energy can be provided at low temperatures in the form of steam, which enables simple heat integration [44]. Fasihi et al. estimate the captured CO_2 specific fixed capital investment costs of a solid sorbent DAC process to $730\,€/\mathrm{t_{CO_2}}$ but they claim that economic data for solid sorbent DAC is limited and this estimation is only based on one specific source [44]. Especially the sorbent costs are uncertain but a crucial cost factor [46]. Due to the modular structure of the solid sorbent DAC units, strong specific cost reduction with scaling are unlikely [46]. Nevertheless, Fasihi et al. predict a large cost reduction for the future due to increased manufacturing scales and connected learning effects [44].

The feasibility of the high temperature liquid solvent DAC process from the Canadian company Carbon Engineering has already been proven by a small-scale pilot plant with a capacity of $600\,\mathrm{kg_{CO_2}/d}$ [45]. Furthermore, liquid solvent DAC is suitable for large scale operations, because the components can benefit from effects of economy-of-scale, all process parts are well known from other applications and the process can be operated continuously [45]. Nevertheless, cost data on liquid solvent DAC is also limited [44]. Based on data from Keith et al. [45], Fasihi et al. assume the specific investment costs of the electrified DAC process to be $815\,€/\mathrm{t_{CO_2}}$ (year 2020) [44]. In economic terms, only scales above $100\,\mathrm{kt_{CO_2}/a}$ are reasonable [45]. The use of specific costs neglects economies of scale, but the unclear cost data also complicates determining scaling factors.

The specific costs of the DAC process include all capital costs except for the working capital. That means that the specific costs multiplied with the captured mass flow of CO_2 result in the fixed capital investment costs of the DAC process.

2.2.4 Electrolysis

Since for PtX and e-/PBtX processes, electrolysis technologies have a significant impact on investment costs, detailed and up-to-date cost estimates, especially for novel technology options, are essential [1]. The equipment costs of electrolyzers consist of stack costs and balance of plant costs [47]. Balance of plants costs include costs of additional equipment and auxiliary systems as pumps for power supply, hydrogen processing, cell cooling, water supply and storage tanks [47]. The balance of plant components are standard equipment and thus have lower specific costs at larger scales [48]. However, stacks do not show potential for economy of scale effects due to the modular cell design and furthermore, they are the main cost factors [49]. Consequently, electrolyzer equipment costs EC, including balance of plant equipment, are typically given specifically in $€/\mathrm{m^2}$ or $€/\mathrm{kW}$

[48–56]. The prices proposed by different works cover a wide range. In the research work carried out by Holst et al. [48] for Proton-Exchange-Membrane Electrolysis (PEMEL) and Alkaline electrolysis (AEL) systems with capacities as high as 100 MW, today's costs per kilowatt excluding the installation costs are in a range of $650 - 1000\,€/\mathrm{kW}$. Saba et al. [57] conducted a literature review of the investment costs and learning curves of PEMEL and AEL between 1990 and 2017. Anghilante et al. [54] demonstrated a bottom-up cost evaluation of SOEL, reaching installed costs of $309 - 618\,€/\mathrm{kW}$ for an integrated, and $380 - 727\,€/\mathrm{kW}$ for a stand-alone system. Böhm et al. estimate today's specific costs for PEMEL and SOEL to $1200\,€_{2017}/\mathrm{kW}$ and $2250\,€_{2017}/\mathrm{kW}$, respectively [49]. Furthermore, they project much lower costs for 2050 (about $350-650\,€_{2017}/\mathrm{kW}$ for PEMEL, AEL and SOEL) based on an earlier published methodology [58] by proposing technology learning ("economies of manufacturing scale") and scaling effects ("economies of unit scale") [49]. Böhm et al. claim that only for scales smaller than 5 MW economies of unit scale are relevant [49]. To prevent overestimation of scaling effects, it seems to be reasonable to utilize specific cost as proposed in Böhm et al. [49] for the calculation of the electrolyzer equipment costs EC.

As for other process equipment, the fixed capital investment costs of electrolyzers are the sum of the direct and the indirect investment costs. However, for electrolyzers the ratio factors on the EC should be considered to be smaller as for standard equipment (see Table 2), since the investment costs for the electrochemical cells are disproportionately large Holst et al. [48], Herz et al. [50]. Table 3 presents factors to determine the FCI of the electrolyzer from the estimated purchased equipment costs. The total factor of 1.52 on the EC of the electrolyzer is taken from [48], where the capital costs of a $100\,\mathrm{MW}$ PEM electrolyzer are determined. The installation costs are assumed to be $12\,\%$ of the electrolyzer purchase costs EC James et al. [59] and thus other costs as engineering, housing, instrumentation and piping are estimated to be $40\,\%$ of EC.

With today's electrolyzer stack lifetimes, stacks have to be replaced before the end of plant operation, leading to additional costs. The costs for replacing stacks are included in the FCOP and discussed in Section 2.3.2.

Table 3 Estimation of the fixed capital investment cost of an electrolyzer from Holst et al. [48], James et al. [59].

Description	Value
Equipment costs	EC
Installation	0.12
Other costs	0.40
Fixed capital investment FCI	1.52 EC

2.2.5 Chemical Reactors

Auxiliary equipment in reactor sections, such as compressors, heaters, coolers, and pumps, is typically standard and can be estimated using commercial software or cost correlations. However, certain reactor setups present unique challenges. For example, FT slurry bed reactors are challenging due to the complex nature of the reaction and the need for careful control of temperature, pressure, and reactor design. Factors such as the size and capacity of the equipment, the materials of construction, and the complexity of the design must be carefully considered. Specialized cost estimation methods, such as detailed engineering analysis or cost escalation, may be necessary to accurately estimate the cost of equipment in these reactor setups.

Water-Gas Shift and Syngas Preparation A WGS reactor is applied in BtX processes to shift the synthesis gas produced in a gasifier to higher H_2/CO ratios (e. g. for FT) by converting CO and water to CO_2 and H_2. The cost estimation of WGS reactors is relatively straight-forward using the cost escalation method. Tijmensen et al. provide cost estimates for WGS systems for BtX purposes based on the fabrication method [21]. Kreutz et al. also tabulated the overnight capital costs of a WGS reactor and its heat exchangers [36].

In PtL and PBtX processes, rWGS reactors are used to convert CO_2 in synthesis gas to CO for CO requiring downstream synthesis steps as FT synthesis [60]. In conventional setups, high temperatures required for the rWGS reaction are typically achieved through natural gas combustion. However, when aiming to avoid fossil fuels, alternative heating methods such as electrical heating may be used [61]. Using electrical heating for rWGS reactors can introduce additional challenges and costs. Electrical heating systems must be designed to provide the high temperatures required for the reaction efficiently and reliably, adding complexity to the cost estimation process. For rWGS reactors equipped with a fired heater, Rezaei and Dzuryk provide a good cost estimation basis [62]. Using electric heating, the same cost levels as that of a gas-fired reactor without a furnace can be assumed. However, the estimation of Rezaei and Dzuryk is provided for a very large scale reactor system with an external heat duty of 312 MW. Baltrusaitis and Luyben provide an economic comparison of various processes to produce syngas from methane including a separate cost analysis of the reactor and the furnace, which can also be utilized to estimate rWGS reactor costs [63]. Many literature studies [56, 60, 64, 65] estimate the rWGS reactor costs based on costs of a WGS reactor provided by Kreutz et al. [36]. Hannula gives an estimation of rWGS unit (catalytic reformer) costs but does not provide further details on the cost determination procedure [18].

Fischer-Tropsch Synthesis The FT synthesis is a commercialized process to convert synthesis gas consisting of carbon monoxide and hydrogen catalytically to a mixture of short and long chain hydrocarbons [11]. FT synthesis with cobalt catalyst is carried out in fixed bed or slurry bed reactors [11].

As mentioned above, FT slurry bed reactors are challenging due to the complex reactor design. Most of the techno-economic process studies in literature [14, 28, 41, 64, 66, 67] use the comprehensive work performed by Bechtel modeling a biomass-based gasification, Fischer-Tropsch liquefaction and combined-cycle power plant in the 1990s [68–72] to estimate FT slurry reactor costs. Tijmensen et al. [21] and Kreutz et al. [27] provide cost correlations for the capital investment costs of a fixed bed FT reactor which are applied in literature studies on PBtX FT processes [56, 60, 73].

Methanol Synthesis Methanol synthesis is a commercialized process, converting synthesis gas to methanol and water and pure methanol can be obtained by distillation [73]. Typically water cooled tubular fixed bed reactors are industrially applied [11]. Dieterich et al. collected literature data for the cost estimation of the methanol reactor, methanol distillation and whole methanol reaction loops including all components [12]. Hennig and Haase calculate the methanol reactor cost from base investment costs of [75] scaled to the produced methanol mass stream [74]. Petersen et al. use the same approach but different cost data for the reactor and also provide a cost correlation for the methanol distillation [73]. Hannula estimate costs of the methanol reactor loop including compressors, and costs for distillation to fuel or chem-grade methanol based on literature data scaled with the produced energy stream of methanol [18].

The problem of scaling with the produced amount of methanol is that the correlations do not account for changing reactor dimensions to keep space velocity constant resulting from changes of the feed gas composition (e. g. higher CO_2 fractions lead to higher recycle streams). Therefore, Rivera-Tinoco et al. and Lacerda de Oliveira Campos et al. claim that the costs of a tube-bundle methanol reactor are equivalent to the costs of a shell and tube heat exchanger filled with catalyst [52, 76]. This simplifies determining reactor costs if the heat transfer area of the reactor tubes are known, since heat exchanger cost correlations as from Towler and Sinnott can be applied.

Methanol to Olefins The Methanol to Olefins process (MTO) utilizes acidic catalysts to produce olefins from methanol [77]. Olefins can be used as base chemicals for plastic production and also to produce fuels in oligomerization. The general process mainly consists of standard equipment units as heat exchanger, distillation columns, compressors and gas-liquid separators. However, the costs of the MTO reactor are more uncertain. Ruokonen et al. calculate the MTO reactor costs with the APEA but size their reactors manually with space velocities from literature [78]. They provide stream tables and costs of the MTO process which enables the application of the cost escalation method by assuming a proper scaling factor. Trippe gives capital costs and a scaling factor of a reactor to synthesize olefins from dimethyl ether [79], which can be assumed to be similar to the costs of a MTO reactor since dimethyl ether is an intermediate product of MTO [77]. Onel et al. provide data for the cost escalation of MTO and methanol to propylene units [80]. However, the unit costs appear to be very high in comparison to the costs also stated for methanol synthesis [80]. Xiang et al. also show MTO unit cost data for applying the cost escalation method, leading to comparably high MTO investment costs [81].

Olefin oligomerization Olefins can be oligomerized via Mobil's Olefins to Gasoline and Distillates (MOGD) process to produce fuel range hydrocarbons as for instance jet fuel components [77]. Similar to MTO, Ruokonen et al. calculate the MOGD reactor costs with the APEA but size the reactor with a space velocity from literature [78]. With cost escalation, the resulting costs are similar as calculated with values for olefin oligomerization to gasoline, given in [79]. Hennig and Haase economically assess a methanol to gasoline process with dimethyl ether as intermediate [74]. The process design is very similar to the MOGD process which is why their cost parameters (adapted from [75]) can be potentially applied for calculating the MOGD reactor costs. It should be mentioned that the resulting costs are clearly lower as with cost escalation from [78, 79], which is questionable.

2.3 Process Specific Production Cost Assumptions

While the capital costs estimated above are included in the Total Cost of Production (TCOP) in the form of depreciation or cash flow analysis, the Cash Costs of Production (CCOP) are based on the Variable and Fixed Cost of Production (VCOP see Section 2.3.1 and FCOP see Section 2.3.2). The applied structure of the TCOP in the CESTEA methodology, as displayed in Figure 3, relies on both Towler and Sinnott [3] and Peters et al. [2].

For BtX, PtX and e-/PBtX processes, the distinction of VCOP and FCOP is based on Towler and Sinnott, since raw materials, utilities, catalysts and solvents are the only costs which are directly proportional to the plant production rate [3]. VCOP are calculated based on mass and energy balances, while FCOP are estimated based on cost factors as demonstrated in Peters et al. [2]. The annual depreciation is calculated from the capital cost estimation which was introduced in the previous sections including the used annuity method to account for depreciation in TCOP calculations.

2.3.1 Variable Costs of Production

In BtX, PtX, and e-/PBtX processes, the VCOP, including all expenses directly associated with the manufacturing operation, typically have the largest share in the LCOP. Here, feedstock cost and plant capacity scale linearly, while investment costs show economy of scale for larger plants [1]. The cost structure, however, varies depending on the process and scenario. For example, in PtX, CO_2 serves as carbon source, while in BtX biomass supplies both carbon and hydrogen to the process. Consequently, feedstock costs which are typically very location dependent, represent the primary cost component [1]. In BtX, feedstock costs and utilities account for about 25 % of the LCOP, while in PtX processes, the contribution of electricity or hydrogen costs can be significantly higher, sometimes exceeding 50 % of the LCOP [12]. In e-/PBtX processes, on the other hand, biomass still serves as carbon and hydrogen hydrogen source while additional H_2 is supplied via electrolysis resulting in higher electricity demand and respective cost structures.

Comparing production cost assumptions across different studies for BtX, PtX, and e-/PBtX processes is challenging due to varying assumptions and boundary conditions. The location of the plant is crucial, as raw material and electricity costs are a major cost driver (see Section 3.1). Site-specific economic analyses are necessary to accurately estimate costs for electricity and biomass supply. Dynamic operation, especially in response to fluctuating electricity prices, adds complexity to economic assessments. [1]

Biomass Cost The cost of biomass is a decisive factor in the detailed TEA, often representing a significant portion of up to 30 % of the VCOP in BtX processes [12]. Biomass residues are typically the only carbon-neutral source considered in these processes, classified as second-generation biofuels.

A wide variety of biomass residues can be used as feedstock in BtX processes, including wood, forestry and agricultural residues, organic wastes, and sewage sludge. When selecting a biomass residue source for a BtX plant, logistical considerations in the entire value chain must be evaluated as biomass availability varies by region, with low energy density and challenges in seasonal storage to compensate for fluctuations in supply [1]. Thus, estimating biomass cost is generally challenging.

Transportation infrastructure at the plant site plays a critical role, with road, rail, or ship transport necessary for biomass delivery. While long-distance transportation of woody biomass is common for large biomass-fired power plants or pulp mills, locally-sourced biomass offers advantages in transport economics and sustainability, particularly for bulky biomass types like agricultural waste and forest residues. To optimize transport needs and reduce greenhouse gas emissions, for BtX plants it is advised to limit transport distances to less than 100 km, typically via road transport.

The RC-EU-TIMES model for bioenergy potentials in the EU and neighbouring countries reports a detailed GIS-based estimation for the physical availability and costs of biomass residues for every location and time combined with the technological options. The derived ENSPRESO database includes cost estimates at national and regional levels for the 2010-2050 period [82, 83].

Potential agricultural residues for BtX applications include "primary resi-dues" such as dry and wet manure coming from cattle, "secondary residues" like olive pits, and "solid agricultural" including waste from the cutting of permanent crops as well as straw and stubble residues. In forestry residues, "primary residues" include logging residues and other pre-commercial thinnings, while "secondary residues" cover woodchips and pellets, sawdust and black liquor. Other waste biomass sources are classified as "primary residues" covering residues from landscape care management, roadside verges and abandoned lands, and "tertiary residues" consisting of biomass residues from different industries and municipal solid waste. [84]

For biomass types already traded in the market, market prices are used as a proxy for cost levels. For other biomass categories, national-specific labor and machinery costs for production, harvesting, and collection up to the roadside are considered including logistics costs, estimated using country-specific transportation costs in different supply chains. The calculations are performed at the NUTS2 regional level, but the input required for the model is at the country level. To derive supply costs at the required aggregation level, the weighted average of the supply cost for each NUTS2 region are used. The costs are converted from Euros per ton to Euros per GJ using crop and feedstock-specific conversion factors. The prices of some feedstock varied depending on the scenario considered, with assumptions made about mobilization, market demand, and technological learning. In the High availability scenario, prices were assumed to be 10% lower than the reference scenario due to more efficient mobilization measures and lower competition from non-energy sectors. Conversely, in the Low scenario, a 10% higher price was assumed than in the reference scenario. [82]

Carbon Dioxide Cost In the context of PtX processes, CO_2 is considered a raw material. CO_2 can either be purchase from a large scale carbon capture facilit or be captured on-site. Its costs, particularly those related to capture, depend on various factors including the plant location, the CO_2 source and employed capture technology, as well as the transport infrastructure if CO_2 is not directly provided at the site. If purchased, the levelized costs of CO_2 capture depending on location and CO_2 source range from $25\,\$/\mathrm{t}$ for chemical, cement, and steel industry to more than $300\,\$/\mathrm{t}$ for DAC [85]. Many studies assume constant CO_2 purchase prices, e. g. $37.75\,€/\mathrm{MWh}$ in Albrecht et al. [56]. Brynolf et al. provide an overview of economic assumptions in literature studies, including the assumed CO_2 purchase costs [86]. However, for detailed economic analysis of PtX processes, the CO_2 capture process should be either directly included in the process modeling, or the economic evaluation or cost data for one specific CO_2 capture technology should be applied.

Electricity Cost In PtX and e-/PBtX processes, the cost of renewable electricity or H_2 has the most significant impact on the VCOP making the supply of renewable electricity or H_2 crucial and highly dependent on the location of the plant and electricity generation. For PtX and e-/PBtX processes, the supply of electricity can be centralized, meaning it is generated at the facility's location, or decentralized, where it is transported to the plant via transmission lines. Similarly, H_2 supply can also be centralized or decentralized, with H_2 either produced on-site or transported to the site after production elsewhere. However, the latter option is generally only feasible for PtX and PBtX processes and is highly dependent on the process design as steady-state operation in the syngas train is a desirable option. In addition, H_2, whether produced on-site or not, may need to be stored for peak shaving, such as during day-night shifts in PV electricity generation, as load shifting is often not a viable option. [1]

When estimating the cost of electricity, it is important to distinguish between electricity purchased from the grid and electricity generated in-house. Purchased electricity typically involves a fixed cost per kWh or MWh based on the prevailing electricity rates. On the other hand, the cost of electricity generated in-house includes various factors requiring a more detailed assessment.

In the CESTEA method for PtX and e-/PBtX processes, the cost of electricity purchased from the grid can be assumed based on current electricity rates for industrial consumers in the relevant geographic region. It is important to consider any incentives or discounts available for industrial consumers since many companies receive tax exemptions and/or other levies and purchasing strategies differ including long term, short term or intercapacity access [87]. Another important challenge of this type of cost estimation is the exclusive use of electricity from renewable energy sources and its volatility. Since renewable electricity prices fluctuate with on-peak and off-peak electricity rates during the day, as well as seasonal variation, considering such a variable electricity price complicates the cost calculation procedure without adding a significant degree of accuracy. Therefore, cost calculations with a fixed electricity price including taxes and levies are essential. Additionally, future trends in electricity prices and potential regulatory changes should be taken into account to provide a realistic estimate of the cost of purchased electricity over the project's lifetime. Assumed renewable electricity costs in literature studies vary widely. While Albrecht et al. assume an electricity price of $100\,€/\mathrm{MWh}$ [56], Marchese et al. provide a range of generation costs of $41 - 210\,€/\mathrm{MWh}$ for renewable energies [65]. Herz et al. use $77\,€/\mathrm{MWh}$ based on the German wind electricity costs for energy intensive industries in 2020 [50] and Hennig and Haase optimistically utilize electricity cost of $44\,€/\mathrm{MWh}$ [74].

Given the challenges related to estimating cost of electricity purchased from the grid and the resulting broad range of cost estimates above, it might be beneficial to estimate the cost to produce the electricity required for the PtX or e-/PBtX process in-house. If only renewable electricity is to be used, the cost estimation would need to consider the specific cost of renewable electricity sources available in the region. This can include sources such as wind, photovolatic (PV), hydroelectric, or biomass-based power generation. Such a cost estimate requires the consideration of various factors such as the initial investment in generating equipment, maintenance costs, electricity generation per year in the form of capacity factors for renewable, as well as any additional costs associated with transmission, storage and taxes. One useful criterion for a comprehensive cost estimate for all technologies, capacity factors, and countries is the expected levelized cost of electricity (LCOE). Ruiz et al. provide detailed GIS-based estimations for the availability and costs of renewable electricity for every location and time combined with the technological options using the RC-EU-TIMES model. The derived ENSPRESO database includes capacity factor distributions for Wind and PV as a result of a technology matrix combining possible technologies and resource scenarios. [84] Dalla Longa et al. and Nijs estimate LCOE based on all combinations of technology and site characteristics using a yearly capacity factor for each technology and country combination including CAPEX and OPEX over the plants lifetime [88, 89].

Oxygen Cost and Revenue In BtX processes with autothermal gasification, pure O_2 is essential. Similar to CO_2, O_2 can be obtained either by purchasing it or by producing it on-site via ASU. The choice between purchasing and on-site production depends on factors such as availability, cost, and logistics. Purchasing O_2 may be more convenient in most brown field cases, while on-site production can offer more control over the supply and potentially lower costs in the long run especially in green field processes. For on-site production, O_2 costs of $25\,€/\mathrm{t}$ can be assumed [12, 56]. If O_2 is purchased from an ASU operator, O_2 costs are about $100\,€/\mathrm{t}$ [90].

In e-/PBtX and PtX processes, if H_2 is produced on-site by electrolysis, so is O_2. Consequently, several studies consider the revenue from selling surplus O_2 [52, 56, 65]. de Saint Jean et al. include revenues from O_2 sales only in optimistic scenarios, with prices ranging from 20 to $70\,€/\mathrm{t}$ [51]. Fasihi et al. conclude that an O_2 sale price above $20\,€/\mathrm{t}$ is unrealistic, despite some studies reporting sales prices of about $80\,€/\mathrm{t}$ [91]. This discrepancy may be due to the potential future surplus of concentrated O_2 from widespread electrolyzer applications. In a future hydrogen economy with widespread electrolysis-based hydrogen production, the price of O_2 could be significantly impacted. The large-scale production of H_2 from electrolysis would likely lead to a surplus of O_2 as a byproduct, which could drive down the market price of O_2. As the demand for H_2 generally increases and electrolysis technologies become more efficient and widespread, the surplus of O_2 could become substantial, further reducing its market value. However, the exact impact on O_2 sales prices for e-/PBtX and PtX processes depends on various factors, including the rate of adoption of electrolysis technologies, the growth of the hydrogen market, and the development of alternative uses for O_2. To avoid unrealistic estimates, the sale price of O_2 should be neglected or only be considered in very optimistic scenarios.

Solid Slag Disposal Cost Solid slag from biomass gasification in BtX processes can meet regulatory standards and thus be used as a construction material in road construction, cement production, and other applications if properly processed. However, the feasibility of generating revenue from solid slag sales depends on factors such as the quality and quantity of the slag produced, local market demand for construction materials, transportation costs, and regulatory requirements for using slag in construction. Detailed feasibility studies are required to assess the potential revenue from solid slag sales and to ensure that the process for producing slag meets the necessary standards and regulations for its use in construction.

On the other hand, if the slag does not meet quality standards or if there is limited demand, it may be more appropriate to treat it as a disposal cost. Thus, solid slag disposal costs are typically considered as operating costs depending on the plant's operation and can include expenses related to handling, transportation, and disposal of the slag [14, 92–94]. Solid disposal costs of 40 $\$_{2016}$/t are proposed by Del Alamo et al. [93].

Water Costs and Steam Revenue Water costs are an important consideration in BtX, PtX, and e-/PBtX processes, as they can impact both the operational efficiency and the overall economics of the processes. Fresh process water is typically used in various stages of the process, such as for syngas quenching after gasification and AGR. The cost of fresh process water can vary depending on the source and treatment required. Demineralized water is often used as feed for electrolyzers in PtX and PBtX processes with on-site H_2 production. The cost of demineralized water includes the cost of treatment to remove minerals and impurities to meet the specifications required for electrolysis. Cooling water is essential for maintaining optimal operating temperatures in various process units. The cost of cooling water includes the cost of water itself as well as the energy required for cooling. However, most cooling water can be recycled which is why the cooling is the major cost driver. Finally, wastewater treatment is necessary to comply with environmental regulations and ensure responsible water management. The cost of wastewater treatment includes the cost of treatment technologies and disposal.

The prices of these different types of water can vary significantly depending on factors such as location, availability of water sources, treatment requirements, and local regulations. There are several sources providing industry water and utility cost data, including governmental databases and information from industry associations and consultancies. Table 4 summarizes water cost assumptions.

Table 4 Cost assumptions for fresh, demineralized, and cooling water supply, as well as waste water disposal.

Type	Price	Source
Fresh water	$2.05\,€_{2018}/\mathrm{m}^3$	[95]
Demineralized water	$4.10\,€_{2018}/\mathrm{m}^3$	[3][a]
Cooling water	$0.10\,€_{2019}/\mathrm{m}^3$	[96, 97]
Waste water	$2.97\,€_{2018}/\mathrm{m}^3$	[95]

[a] Assumption that demineralized water is twice as expensive as fresh water adapted

As process water prices vary in different countries, the work of Tetzner and Bittner is used exemplaryly for the average fresh water and wastewater prices in 2018 in one German federal state [95]. Since the price of the demineralized water highly depends on the input and output water qualities, its costs can vary. Therefore, as suggested by Towler and Sinnott, demineralized water is assumed to be twice as expensive as process water [3]. Cooling water prices and costs are provided by Intratec including costs related to clarified water make-up, chemicals and electricity required to drive cooling tower and pumps motors [97].

As the use of electrolyzers for hydrogen production is expected to increase especially for applications such as PtX processes, concerns about water availability and sustainability, particularly in regions facing water scarcity or competing demands for freshwater resources, might be justified. Furthermore, there could be challenges related to the availability of cooling water due to changes in water availability, temperature, and environmental regulations due to global warming. This could impact the design and operation of BtX, PtX, and hybrid processes, potentially leading to increased water consumption or the need for alternative cooling technologies. Addressing these issues in a TEA could help in identifying potential risks and opportunities associated with water use in BtX, PtX, and hybrid processes. Strategies such as water recycling, or use of alternative cooling technologies, must be evaluated to enhance the sustainability and resilience of these processes in the face of future challenges.

In BtX, PtX, and PBtX processes, steam is a by-product that can be generated from waste heat produced during various unit operations. This steam can be utilized within the plant for heating purposes or can be sold to other facilities, providing additional revenue streams. The availability and pricing of steam depend on the process design, location of the plant, and local market conditions. Utilizing waste heat to generate steam not only improves the overall efficiency of the plant but also contributes to the sustainability of the process by reducing the reliance on fossil fuels for heating purposes.

In BtX, PtX and hybrid processes, the majority of the heat demand can be matched using waste heat from chemical reactions and surplus steam is produced [56, 65]. Albrecht et al. assume that produced steam is sold for the price of steam produced by a natural gas boiler resulting in low pressure (4 bar) steam sale prices of $25.7\,€_{2014}/\mathrm{t}$ and $26.3\,€_{2014}/\mathrm{t}$ for medium pressure (25 bar) steam [56]. Lower temperature heat can be used in district heating, also generating revenues [56].

However, the applicability of steam networks and district heating heavily depend on the plant location. Location and existing infrastructure are closely linked to the decision to build a greenfield or brownfield plant. In addition to existing transport infrastructure, the latter offers the advantage of being integrated into existing industrial processes including the option to profitably supply waste heat [1]. The decision to build a greenfield or brownfield plant depends on various factors, including the availability of raw materials and utilities such as biomass, water and electricity, as well as general process design. BtX and e-/PBtX plants are more likely to be built brownfield due to the complexities of biomass supply chains and the cost structures associated with BtX processes. For PtX plants supplied with CO_2 from DAC, on the other hand, the requirement for only water and electricity could indeed enable greenfield plants in locations where these resources are readily available, hindering the assumption of additional revenue from steam sales.

Catalyst costs Estimating makeup costs for catalysts can be challenging due to the variability in catalyst prices and the lack of standardization across different processes and equipment. One approach is to use cost data from similar processes or applications where catalysts are used. Another approach is to use cost correlations or cost indices that relate catalyst prices to other variables such as reactor size, process conditions, or production capacity. These correlations can provide a rough estimate of makeup costs based on known factors that influence catalyst prices. It is also important to consider the specific requirements of the synthesis process and the catalyst used. Factors such as catalyst lifespan, regeneration requirements, and the availability of alternative catalysts can all impact the makeup costs and should be taken into account in the estimation process.

In CESTEA, replacement costs of catalysts are calculated using gas hourly space velocity GHSV, catalyst bed density, replacement rate and specific catalayst costs. Dividing the syngas flow rate entering the reactor by the GHSV, the total amount of catalyst required can be dertermined. To consider the catalyst makeup due to deactivation and attrition, the total catalyst mass is multipled with the catalyst replacement rate.

GHSV, catalyst bed density, and makeup rate proposed by Swanson et al. are used to calculate the catalyst needed for the (sour) WGS reactor [14].

In the rWGS reactor, a Ni catalyst on alumina is employed [65]. The catalyst space time, used to determine the catalyst mass, is derived from experiments on a commercial Ni catalyst [98]. It is assumed that $0.01\,\%$ of the catalyst have to be replaced every day and the catalyst costs are adapted from the from the costs of a tar reformer catalyst [99].

Table 5 WGS, rWGS and FT reactor design assumptions and catalyst costs.

Parameter	**WGS**^{a}	**rWGS**	**FT**
GHSV in $1/\mathrm{h}$	1000	172000^{b}	595^{d}
Bed density in $\mathrm{kg_{cat}}/\mathrm{m^3}$	897	1200^{b}	250^{d}
Cat. makeup in $\%/\mathrm{d}$	0.091	0.01^{c}	0.5
Cat. cost in $\$_{2007}/\mathrm{kg_{cat}}$	17.64	13.29^{c}	33.07^{a}

^{a}Swanson et al. (FT: Co-based catalyst), ^{b}Jess et al., ^{c}Phillips et al., ^{d}Fox et al.

To calculate the amount of catalyst used in the FT reactor, data for the space velocity and bed density provided by Fox et al. is utilized [70]. Table 5 presents the respective values including an assumed catalyst makeup rate and a Cobalt catalyst price from [14].

In the conversion of synthesis gas to fuels with methanol as intermediate, catalyst costs emerge in the methanol synthesis, MTO and MOGD processes. The catalyst costs and lifetime in the methanol synthesis are based on [99]. The catalyst mass can be calculated from kinetic simulation models or by applying commercial GHSV as given in [11]. In the MTO and MOGD reactors, a ZSM-5 catalyst is used [77]. The costs and lifetimes are taken from Ruokonen et al. [78]. If kinetics are applied, the catalyst mass in the MTO reactor can be calculated from the simulation model. The weight hourly space velocity WHSV in the MOGD reactor is taken from Harandi [100]. Table 6 provides all relevant data.

Table 6 Catalyst costs and additional assumptions for the methanol, MTO and MOGD process.

Parameter	Value	Source
Methanol GHSV	$\backsim 10000\,\mathrm{h}^{-1}$	[11]
Catalyst (MeOH) cost	$21.36\,\$_{2007}/\mathrm{kg}$	[99]
MeOH catalyst makeup	$0.075\,\%/\mathrm{d}$	[99]
ZSM-5 catalyst price	$8.4\,€_{2020}/\mathrm{kg}$	[78]
ZSM-5 catalyst makeup	$0.15\,\%/\mathrm{d}$	[78]
MOGD WHSV	$0.5\,\mathrm{kg}/\mathrm{kg_{cat}h}$	[100]

2.3.2 Fixed Costs of Production

Operating labor and supervision The hourly salary paid to all the operating personnel is assumed to be constantly 40 €/h. To calculate the number of personnel for the plant the chart proposed by Peters et al., in which the number of employee-hours/(per day)/(per process) can be obtained by knowing the plant production capacity, is utilized [2]. After obtaining the number of worker-hours/day/process, it is multiplied by the number of processes in the plant and $8000\,\mathrm{h}$ as the amount of yearly hours where personnel is needed. The capacity of units in different cases is leading to different values for each case. The operating supervision costs are considered as 15 % of the labor costs. In considerations of the plant operating hours, the labor costs are assumed to stay constant.

General maintenance Costs of general maintenance are assumed to annually account for 3 % of the FCI based on Peters et al. [2]. In this methodology, maintenance costs are included in the Fixed Costs of Production and consequently, part load operation doesn't effect the maintenance costs.

Electrolyzer maintenance For electrolyzers, additional to the general maintenance costs, costs evolve for replacing stacks at the end of their lifetime [50, 51]. To calculate the arising annual stack replacement cost, the degradation rate of the cell stacks and the cost breakdown of the electrolysis system are to be known. Data from Böhm et al. [49] and Herz et al. [50] regarding the share of the stacks in the equipment costs EC and the stack lifetime is exemplary presented in Table 7.

Table 7 Electrolysis stack cost share and lifetime of PEMEL and SOEL today (2020) and projected (2050).

Technology	Stack EC share [49]	Lifetime in h [50]	
		2020	2050
PEMEL	60 %	66790	89509
SOEL	30 %	45473	88700

The annual costs for stack replacement can be calculated by multiplying the electrolyzer purchase costs EC, the stack share in the EC and the yearly operational hours divided by the electrolyzer lifetime. Also costs for the installation of the stacks must be considered in the purchase costs [51]. Consequently, the installation factor of 12 % [59] is additionally applied on the EC for calculating the annual stack replacement costs. Since the stack replacement costs depend on the operating hours of the plant, strictly speaking, the stack replacement is a variable cost factor. Nevertheless, this methodology includes it further on in the fixed charges.

Other fixed charges As already mentioned in Section 2.1.4, the other Fixed Costs of Production are estimated based on Peters et al. [2] and the working capital interest costs are calculated based on the working capital, the interest rate and the plant lifetime. Peters et al. quantify the range of typical interest rates to 5-10 %, which is why 7 % is utilized as standard value in this methodology (also used in [101]). However, lower interest rates could be realistic for projects with government funding. The plant lifetime is assumed to be 25 years [65].

The working capital interest costs are very high for PtX/PBtX processes, since the working capital is assumed to be directly dependent on the investment costs. In operation high cash reserves have to be available to bridge time differences between the production and sale of products, because high value products are produced.

3 Limitations

Estimating costs for BtX, PtX, and e-/PBtX processes involves understanding the specific equipment, methods, and data sources used. The CESTEA method presented in this paper uses cost estimation based on a process flow diagrams or process simulation. Since each process has its unique requirements and challenges, influencing cost estimation, a detailed understanding of these processes and their associated TEA methodologies is crucial for accurate equipment cost estimation.

Based on deterministic estimating methods for major equipment and variable cost and factoring and other stochastic methods to estimate less-significant areas of the process as well as Fixed Costs of Production, the level of detail of the CESTEA method is limited to Class III and IV (see Figure 2). Acknowledging that a too-detailed methodology might limit the exploration of various process options, the approach presented in the following seeks to strike a balance. It recognizes that an overly detailed approach could be resource-intensive and restrictive in terms of the number of processes that can be investigated. On the other hand, the methodology should be sufficiently detailed, as too much imprecision could affect the reliability of the cost estimates and limit the ability to draw meaningful conclusions or make serious statements about the processes. The ultimate objective is to facilitate process comparison. This suggests that the methodology is designed not just for estimating costs in isolation but also to allow for a meaningful evaluation and comparison of different PtX, BtX, and hybrid processes. Furthermore, the accuary of cost estimation is heavily influence by the plant's boundary conditions such as process design, location, assumed full-load hours etc. (see Section 3.1), as well as inherent accuracy limits when estimating equipment cost (see Section 3.2).

CESTEA is designed to allow for the investigation of various process options aligning with the dynamic and evolving nature of technologies and processes in the PtX and BtX domains. Commercial software can help estimate costs by considering equipment costs, utility costs, other variable costs, and associated expenses. Most of the methods used in CESTEA are incorporated in such cost-estimating software. It often employs industry-standard costing methodologies and correlations. APEA, for examples, determines equipment costs, bulk costs, and installation costs from the costs of materials and labor, it can, in theory, give reasonably good estimates [3]. As discussed in Section 3.3, the lack of documentation or the use of outdated data and methodologies in commercial cost estimation software can, however, introduce significant risks to the accuracy of cost estimates, potentially leading to sub-optimal decision-making and project outcomes. Users should take proactive steps to increase the reliability of the automated cost estimation.

3.1 Boundary Conditions

Understanding and accurately estimating the cost of production is crucial for evaluating the economic feasibility of novel BtX, PtX, and e-/PBtX processes. The key factors that influence the accuracy of any cost estimate are mainly dependant on the facility's location. For example, most plant and equipment cost data are given for plant location in North America or western Europe as the main centers of the chemical industry. Thus, for equipment cost estimation, a location factor can be used if the plant location differs. The Cash Costs of Production (CCOP), on the other hand, are massively affected by factors such as raw material availability, utility an labor cost, regulatory requirements, and market conditions. The cost of biomass residues, for example, not only depends on the type of biomass used, but also the regional and seasonal availability, as well as competing market conditions. Also, the cost of biomass production and of harvesting, and transport, pretreatment cost up-to the conversion gate including the cost made after harvesting for preprocessing, and forwarding and transport to the place of collection must be considered where relevant. [82] Biomass can be sourced locally, regionally, or transported from a distance, with both decentralized and centralized pretreatment options available. Furthermore, while biomass residues are sometimes treated as waste and assumed to be cost-free, or even profitable to dispose of, there is often an already existing market for biomass residues, including competing use in biogas plants, thermal use in heat or steam generation, or second use in construction materials or packaging material. Seasonal fluctuations in biomass availability, along with the need for diverse transportation modes, present other significant challenges in estimating the economic viability of biomass supply. [1] Thus, high accuracy of cost estimates is only expected for site-specific economic analyses.

Deciding between a greenfield or brownfield plant is a critical early decision that can significantly impact the overall project feasibility and economics. Greenfield plants offer the advantage of starting with a clean slate, allowing for optimal design and integration of new technologies. However, they often require more infrastructure development and may face challenges in obtaining permits and approvals. On the other hand, brownfield plants are typically easier to permit and may have existing infrastructure that allows integration into other processes and shared utilities. The choice between greenfield and brownfield should consider factors such as site availability, proximity to feedstock and markets, infrastructure, regulatory environment, and overall project goals and constraints. Furthermore, competing market conditions, including the availability and pricing of utilities, by-products or materials, also play a significant role in cost estimation. Additionally, considering whether the estimate is for the current situation or a future scenario is important, as future conditions may vary due to factors such as technological advancements, changes in regulations, and shifts in market dynamics.

Another crucial parameter affecting accuracy is the assumed full-load hours and interest rate. As full-load hours, other than for example feedstock costs, show a strongly non-linear behaviour towards Levelized Cost of Production (LCOP), stationary operation with the least possible downtime is a key factor for the lowest possible cost [12]. However, assuming high full-load hours without considering the technical challenges and potential supply chain issues in BtX, PtX, and e-/PBtX processes could lead to overly optimistic cost estimates. It is important to be realistic and factor in the complexities of the process, including the reliability of feedstock supply, potential downtime for maintenance, and the need for operational flexibility. The main challenges for achieving high full-load hours in BtX, PtX, and e-/PBtX processes include feedstock availability and quality, supply chain challenges and infrastructure, and technology readiness and reliability. For gasification-based processes, the novel biomass gasifier is the most likely bottleneck. The limitations due to technical challenges, such as maintaining stable operation and addressing potential issues with feedstock variability or gas quality, limit the expected full-load hours for BtX and e-/PBtX to around 8000, with 7500 being a good base case assumption. In PtX processes, the limiting factor varies depending on the specific process and design. Some potential bottlenecks include the availability and cost of renewable electricity or hydrogen, the efficiency and reliability of the electrolysis process, and the overall integration of different process steps. A more conservative approach to estimating full-load hours can help ensure that cost estimates are grounded in reality and account for the uncertainties inherent in such processes.

3.2 Accuracy Limits in Equipment Cost Estimation

To determine the Total Purchased Equipment Cost (TPEC) the four methods used in CESTEA (see Section 2.1.2 are generally limited in terms of accuracy and availability of data. Recent genuine data from fabricators or suppliers provide the most accurate cost estimate but requires access to up-to-date quotations from suppliers, which may not always be available. The Fabrication Method estimates the cost of equipment based on the actual component design, taking into account factors such as materials, labor, and fabrication processes. However, this method typically requires detailed engineering, understanding of fabrication steps, knowledge of machinery costs, and labor requirements. The Cost Escalation Method, on the other hand, uses existing cost data and scales it to the current or future project needs. This method relies on accurate scaling factors and existing cost data to estimate future costs, making it useful for predicting cost trends but potentially less accurate if the scaling factors, the CEPCI, or other data are not up to date. Cost Curves and Correlations use empirical relationships to estimate equipment costs. They, though easily calculated, suffer from low accuracy and are only applicable to standard equipment and may lack accuracy for complex or unique projects.

When choosing a method for calculating TPEC, a well-defined project with clear equipment specifications may benefit from more detailed methods, while projects with less defined requirements may require more general estimates. Similarly, the level of detail in the design of individual components affects the accuracy of the cost estimate. Methods that account for specific component designs may be more accurate but require detailed design information. The readiness of the technology being used can also impact the choice of method. Established technologies benefit from well-know estimation methods, while novel or complex technologies may require more detailed approaches that require a high level of engineering knowledge may be more accurate but may also be more time-consuming and costly to implement. Combining multiple methods or using sensitivity analysis can help improve the reliability of the cost estimate.

Table 8 Exemplary EC_i and TPEC calculation procedure for the water-gas shift (WGS) step in BtX in M€.

WGS unit	Min	Max	Mean
WGS Reactor	**3.10**	**4.30**	**3.70**
First Source	3.10	3.10	3.10
Sec. Source	4.30	4.30	4.30
Cooler	**0.11**	**0.11**	**0.11**
First Source	0.11	0.11	0.11
G-L Sep.	**0.06**	**0.06**	**0.06**
First Source	0.06	0.06	0.06
Heater	**0.02**	**0.02**	**0.02**
First Source	0.02	0.02	0.02
TPEC	**3.29**	**4.49**	**3.89**

In CESTEA, the cost escalation method is recommended to determine equipment cost (see Section 2.1.2). Where possible, several cost references from literature, manufactures or former quotations for the same unit operation are used to decrease uncertainty in the cost estimate. Derived cost ranges should always be systematic compared to other source of the same equipment, thoroughly discussed and checked for plausibility. This methodology seeks to provide a more robust and reliable estimate for the equipment cost. Instead of providing a single final value, the approach aims to calculate a cost range for the equipment including maximum, minimum, and average cost values. The average derived equipment cost can be used if the spread is in a reasonable range. This acknowledges the inherent variability and uncertainty in estimating costs and provides stakeholders with a range of potential values.

Table 8 exemplary shows the generic TPEC calculation procedure for a WGS unit. Using average values smooth out outliers and provides a more representative cost estimate. Here, the deviation from the average is in a range of ±15 %. With no single high-quality cost estimate, such as recent manufacturer quotation or detailed fabrication method, available, this average TPEC of 3.89 M€ is the cost estimate used.

3.3 Process Simulation and Commercial Software Solutions

TEA software can be valuable tools for conducting cost evaluations and feasibility studies. However, any inaccuracies or uncertainties in the process simulation model can affect the reliability of cost estimates. The issue of potentially unqualified users providing input data in the context of process simulation and economic analysis software like APEA is a common challenge in engineering and industrial settings, as accurate simulation and economic analysis depend on the availability of precise input data provided by the user. The user's active input of technical data and specifications can significantly impact the accuracy of the simulation. Misunderstandings or errors in data entry, such as unit conversions or process parameters, can propagate through the simulation model, affecting the reliability of the economic analysis. Users need to ensure that process simulation models are well-defined and represent the actual process accurately.

The accuracy and reliability of any simulation results also depend heavily on the quality and the methodologies used by the software. The software itself may make certain assumptions and use industry-standard correlations, which may not perfectly capture the specific nuances of every process. If documentation is lacking, or if the data and methodologies employed by the software are outdated, it can lead to incorrect results without the user being aware of it. It can become challenging for users to validate the accuracy of the results independently. This lack of transparency can erode confidence in the software's predictions. To mitigate these risks, it's crucial for users to:

- Ensure correct technical data and specification inputs to the simulation.

- Seek updated documentation: regularly check for updated documentation from the software developer to ensure that the methodologies and algorithms used are current and aligned with industry best practices.
- Validate results: independently validate the results obtained from commercial software where possible. Cross-checking estimates with other reliable sources or conducting sensitivity analyses can provide additional assurance.
- Engage with software support: if there are uncertainties or lack of clarity, users should reach out to the software support for clarification. Developers often provide assistance in understanding the software's capabilities and limitations.

Overall, TEA software can be safe to use when applied appropriately and with careful consideration of its limitations. Users should approach TEA as a tool to inform decision-making rather than as a definitive answer, recognizing that it provides valuable insights but may not capture all nuances of a real-world scenario.

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Acknowledgments

The investigations have been supported by the German Federal Ministry of Education and Research (BMBF) within the project REDEFINE H2E (01DD21005) and H2 Living Laboratory Burghausen (03SF0705B) which is gratefully acknowledged. In addition, the authors acknowledge the cooperation within the Network TUM.Hydrogen and PtX.

Declaration of interests

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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