

## HEURISTICS FOR CAPITAL COST ESTIMATION: A CASE STUDY ON BIOREFINERY PROCESSES

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### ABSTRACT

Capital cost estimation is required during all stages of a process development to decide on project viability, screen process alternatives and monitor expenses. General cost engineering practices have been established, while a variety of costing techniques are available according to the process design stage, starting from rapid order-of-magnitude techniques to detailed cost methods. This study investigates the ability of several quick methods to produce accurate early-stage estimates for pioneer biorefinery plants. For the purposes of this work two biorefinery processes were simulated to serve as validation models. The results show discrepancies among the methods, while their reliability is dubious as they were developed for the needs of the petrochemical industry. The development of a new method or the modification of existing cost techniques is imperative in order to accurately estimate the biorefineries' capital costs.

### INTRODUCTION

Optimal plant design should define a process that can create profit under certain operating parameters. Chemical Engineers must be familiar with all cost elements that are associated with the production process because net profit is defined as the plant income minus the operating and capital expenses [1]. Due to limited capital budgets and stringent financial resources, companies are reluctant to allocate funding to new projects while they require continuous monitoring of expected costs and expenses at all stages of the process development. This laborious task is assigned to the researchers, who must be comfortable in providing accurate estimates justifying their proposals [2]. Cost engineering falls within the boundaries of practical engineering and deserves to be studied as any other empirical field [3].

### CAPITAL COST ESTIMATION

The development of a new chemical process and the design and construction of the plant is a long process which requires large investment funds. The initial conception is followed by experimental trials in the laboratory and simple flowsheets and if it is approved, it gradually moves on to pilot-scale and then commercial scale with more detailed designs. An accurate capital cost estimate must be determined for a proposed plant in order to decide on project size and viability [4].

### CAPITAL INVESTMENT

The capital needed to supply the necessary manufacturing and plant facilities is called the **fixed-capital investment**, while that necessary for the operation of the plant is termed the **working capital**. The sum of the fixed-capital investment and the working capital is known as the total capital investment. The fixed-capital portion may be further subdivided into manufacturing fixed-capital investment and non-manufacturing fixed-capital investment [1].

### FIXED CAPITAL INVESTMENT

**Manufacturing fixed-capital investment** (or direct costs or battery limits capital costs) represents the capital necessary for the installed process equipment with all auxiliaries that are needed for complete process operation (i.e. piping, instruments, insulation, foundations, and site preparation). Fixed capital required for construction overhead and for all plant components that are not directly related to the process operation is designated as the **non-manufacturing fixed-capital investment** (or indirect costs or offsites). The construction overhead cost consists of field-office and supervision expenses, engineering and construction costs, contractor's fees, and contingencies [1].

### WORKING CAPITAL

The working capital for an industrial plant consists of the total amount of money invested in (1) raw materials and supplies carried in stock, (2) finished products in stock and semifinished products in the process of being manufactured, (3) accounts receivable, (4) cash kept on hand for monthly payment of operating expenses, such as salaries, wages, and raw-material purchases, (5) accounts payable, and (6) taxes payable. The raw-materials inventory included in working capital usually amounts to a 1-month supply of the raw materials valued at delivered prices. Finished products in stock and semifinished products have a value approximately equal to the total manufacturing cost for 1 month's production. Because credit terms extended to customers are usually based on an allowable 30-day payment period, the working capital required for accounts receivable ordinarily amounts to the

production cost for 1 month of operation. The ratio of working capital to total capital investment varies with different companies, but most chemical plants use an initial working capital amounting to 10 to 20 percent of the total capital investment. This percentage may increase to as much as 50 percent or more for companies producing products of seasonal demand because of the large inventories which must be maintained for appreciable periods of time [1] [5].

#### DEFINITION OF INSIDE/OUTSIDE BATTERY LIMITS

According to the American Association of Cost Engineers, for process plants, fixed capital investment can be divided into outside battery limits or off sites (OBL) and inside battery limits (IBL). "Inside Battery limits" (IBL) comprises one or more geographic boundaries, to specify the area where production takes place, enclosing all associated equipment and production facilities. "Outside Battery limits" (OBL) includes facilities such as storage, utilities, administration buildings, or auxiliary facilities [5] [6].

#### TYPES AND CLASSIFICATION OF COST ESTIMATES

Capital cost estimates are made since the very start of project development until final commissioning and start-up of the plant. Initial cost estimates are essential to decide whether to continue allocating funds or discard the project and opt for more promising alternatives. For early estimates, quick methods are employed that usually do not cost much to apply. Since not adequate information is available, some substantial error is inevitable. As more process details are defined during project development, capital estimates become more accurate and the error ranges narrow significantly.

The classification of capital cost estimates is still not universally standardised despite efforts that have been made to overcome this problem [3]. The Cost Estimate Classification System maps the phases and stages of asset cost estimating together with a generic maturity and quality matrix that can be applied across a wide variety of industries. There are numerous ways of classifying the different types of capital cost estimates for chemical plants (Table 1). Quantified predictions can only be done for a limited amount of time ahead, since uncertain market conditions and rapid technology development result in sharp price inflation.

Not all estimation stages are to be followed during a project development. Management often decides to proceed in commissioning a new plant very soon after its conception because of favourable market conditions or business agreements. Moreover, if the new project resembles a plant already built by the company, detailed estimates can be produced very early during the project development [7].

**Table 1.** Classification of cost estimates [8]

AACE Classification Standard	ANSI Standard Z94.0	AACE Pre-1972	Association of Cost Engineers (UK) ACostE
Class 5 L: -20% to -50% H: +30% to +100%	Order of Magnitude Estimate -30/+50	Order of Magnitude Estimate +40% to -20%	Order of Magnitude Estimate Class IV -30/+30
Class 4 L: -15% to -30% H: +20% to +50%	Budget Estimate -15/+30	Study Estimate +30% to -20%	Study Estimate Class III -20/+20
Class 3 L: -10% to -20% H: +10% to +30%		Preliminary Estimate +25% to -15%	Budget Estimate -10/+10
Class 2 L: -5% to -15% H: +5% to +20%	Definitive Estimate -5/+15	Definitive +15% to -7%	Definitive Estimate Class I -5/+5
Class 1 L: -3% to -10% H: +3% to +15%		Detailed estimate +6% to -4%	

**EXISTING METHODS FOR CAPITAL COST ESTIMATION****A. POWER LAW OR EXPONENTIAL METHODS**

The capital cost of a plant can be derived from historical data by extrapolating on the total cost or parts of the cost of a similar plant (if it exists). The following equation proposed by Williams (1947) [9] is used:

$$\frac{C_1}{C_2} = \left(\frac{Q_1}{Q_2}\right)^n \times t \quad (1) \text{ where,}$$

$C_1$  = Cost of the item at size or scale  $Q_1$

$C_2$  = Cost of the reference item at the size or scale  $Q_2$

$n$  = Scale exponent or cost capacity factor

$t$  = correction factor for relocation, inflation, currency exchange, temperature, pressure, materials of construction

The  $n$  value of 0.6 is typically used for chemical plants, and for this reason, the relationship is often dubbed the '6/10 rule'. Information on the value of the  $n$  exponent can be found in common chemical engineering handbooks [1] [7].

**B. FACTORIAL ESTIMATES**

Factorial methods can be used to produce study estimates and preliminary estimates. This estimating method requires a detailed list of all equipment and their relevant cost necessary to run the production process. It is based on the postulation that all parts of the capital cost are correlated to the purchased equipment cost. Lang (1948) [10] was the first to introduce this idea by proposing that the capital cost of a plant is related to its equipment cost by a factor  $f$  related to the type of process (fluids, solids, solids-fluids).

$$\text{Capital Cost} = f_{\text{INST}} \times (\text{Equipment Cost})_{\text{PURCH}} \quad (2)$$

Values for Lang's factors can be found in common engineering handbooks. The Lang factors have been improved and modified by various authors throughout the years. Hand, Wroth, Brown, Holland, Harper, Guthrie, Chilton, Hirsch and Glazier, Miller and Garrett are some of the authors who have worked and proposed such types of estimations [5] [7] [11]. The last report referring to factorial estimating, published by Woods [12] in 2000, proposes a list of 500 pieces of equipment each with its installation factor. Considering the amount of information and details required, factorial estimating is most probably not suitable for conceptual estimating but rather Class 4 or 3 appraisals.

**C. STEP COUNTING METHODS-FUNCTIONAL UNIT BASIS**

Step counting methods attempt to correlate the capital cost of a process with the values of its fundamental process parameters. These values might be the number of significant process steps, plant capacity, temperature, pressure, materials of construction, number of significant process steps etc. The term significant process step (or functional unit) defines all equipment and auxiliaries necessary to complete an operation in the production stream, as defined by the American Association of Cost Engineers.

Gore (1969), Stallworthy (1970), Wilson (1971), Bridgwater (1976,1981), Taylor (1977), Viola (1981), Klumpar, Brown and Fromme (1983) are some of the authors who collected historical plant data and attempted to find a correlation between the capital cost and process parameters. Each of the authors presented their own methodology for specific types of chemical processes providing their definition of the functional unit and proposing basic parameters which affect the costs.

**D. HEAT LOSSES METHOD**

With a view to provide simple concepts and tools for quick economic estimation, J.P. Lange produced a simple correlation that indicates that the total project investment is related to the amount of energy loss of the process. The assumption is derived by the significant impact of the heat of reaction on the total cost. For accommodating the heat of reaction, dedicated equipment and therefore dedicated costs are required. In both, exothermic and endothermic, reaction types, handling the heat of reaction demands bigger surface areas for heat exchanging that contributes a great amount on the total investment. The total energy loss is defined as the difference between the Low Heating Values of the plant intake (including feed and fuel streams) and that of the product stream leaving the plant. Accordingly, the energy loss accounts for the losses through heat of reaction, heat of fuel combustion and heating value of purge streams.

$$\text{Investment [M\$ 1993]} = 3.0 \times (\text{energy losses [MW]})^{0.84} \quad (3)$$

Equation 3 is derived by correlating a collection of fuel manufacturing plants, where the  $R^2$  amounts to 0.94. However, it is stated that the energy loss is not a good indicator for small-scale, heat neutral reactions or for batch processes used for manufacturing fine and specialty chemicals [13]. The total installed equipment cost could be considered to be approximately 60% of the total investment, as indicated by Peters & Timmerhaus [1].

**CASE STUDY: BIOREFINERY PLANTS**

It is only recently that society has recognized the increasing potential of biorenewable resources towards a more sustainable production of energy, chemicals and materials. Biomass (or biorenewable resource) is defined as organic material of recent biological origin, whose processing can act as a competitive alternative to the conventional fossil-based production processes in terms of economic potential and environmental impact.

Increasing energy demand and environmental awareness are the main driving forces directing society to allocate financial resources for research and development of biomass conversion processes to useful materials and fuels [14]. Furthermore, biorefineries offer an excellent initiative for promoting development and creating new employment opportunities in rural areas, while leading towards oil independent societies employing greener practices [14] [15]

The early conceptual stages of a bio renewable production process must be accompanied by order-of magnitude cost estimates and serve as a selection guide to the most economical route. Published information from commercial biorenewable projects can help in evaluating the economy of various biorefinery technology variants. As these projects are pioneer and often one-of-a-kind, sensitive economic information is usually kept secret and not available in the public literature. Therefore, a large number of scientific papers relies on the application of well-established rapid costing techniques, in order to approximate the worth of a biorefinery process given the information available at the time. Usually the estimated cost is near to the actual cost but might have some deviations from it and is expected to provide guidance for the planner or designer to justify his/her approach.

This study aims at determining this assumption and making recommendations regarding the use of these costing methods for estimating biorefinery processes. Five ‘functional unit’ techniques and J.P. Lange’s heat losses method were selected for the purposes of the study and were modified with suitable relocation [16] and inflation factors(CEPCI) [17] to estimate Capital Costs in M\$, 2011 for a greenfield plant built in the US. As the methods under study date as far back to the 1970’s, currency and location were selected as such as to avoid any additional calculation errors. Furthermore, the majority of the methods were developed at a time of extreme inflation in the UK and thus, the use of a complex US factor is suggested instead [18]. Details on the correlations used can be found in Table 2. Details on the methods and explanation on parameters used can be found in the original articles published by the corresponding authors.

**Table 2.** “Functional Unit” techniques

Author	Correlation	Reported Accuracy
Wilson [19]	$C = 6.38 \cdot 10^{-6} \cdot f \cdot N \cdot (AUC) \cdot F_m \cdot F_p \cdot F_T \quad (4)$ <p>C: Capital Cost in M\$ (US,2011), f:investment factor obtained from a graph of the factor versus (AUC), N:number of main plant items, AUC: average unit cost  <math>F_m</math>: factor for specific materials of construction, other than carbon steel, <math>F_p</math>, <math>F_T</math>:correction factors for design pressure and temperature, V:average throughput</p>	-30/+50%
Taylor [20]	$C = 0.17 \cdot \sum_1^N (1.3)^S \cdot Q^{0.39} \quad (5)$ <p>C: Capital Cost in M\$ (US,2011),Q:capacity in ktns/year, N:number of ‘significant process steps’ (e.g. filter, react, distil), S: ‘complexity score’ determined for each process step to take account of factors such as relative throughput, materials of construction, reaction time, storage time, temperature, pressure, multistreaming and special conditions.</p>	±30%
Bridgwater [21]	$C = 380 \cdot 10^{-6} \cdot N \cdot \left(\frac{Q}{s^{0.5}}\right)^{0.85} \cdot \left(\frac{T \cdot n}{N}\right)^{-0.17} \cdot \left(\frac{P \cdot n'}{N}\right)^{0.14} \quad (6)$ <p>C: Capital Cost in M\$ (US,2011), N: number of functional Units, Q: Capacity (ktns/year), s: process conversion</p>	±20%
Klumpar, Brown, Fromme [22]	$C = 3.54 \cdot 10^{-4} \cdot F \cdot N \cdot G^{0.57} \quad (7)$ <p>C: Capital Cost in M\$ (US,2011), <math>F = 2 \cdot (10^{T+P+M})</math>, N: number of process modules, G: capacity in kg/hr</p>	±30%
Petley [4]	$C = 0.093 \cdot Q^{0.44} \cdot N^{0.486} \cdot T_{max}^{0.038} \cdot P_{max}^{-0.02} \cdot F_m^{0.341} \quad (8)$ <p>C: Capital Cost in M\$ (US,2011), Q: Capacity in (kt/yr), N: number of Functional Units</p>	N.A.
Lange [13]	$C = 1.8 \cdot (LHV_{in} - LHV_{out})^{0.84} \quad (3)$ <p>C: Capital Cost in M\$ (US,2011), <math>LHV_{in}</math>: Lower Heating Value of inlet+fuel, <math>LHV_{out}</math>: Lower Heating Value of outlet</p>	N.A.

For assessing the reliability of these costing techniques, two bio-refinery processes were selected: “Biomass Indirectly Heated Gasification” and “Production of Ethanol by Fermentation of SynGas”. The first process has been analyzed and various reports refer to its thermochemical technology as it resembles the classic gasification of coal. However, little is known on the capital cost of such a commercial plant in operation. On the other hand, Syngas Fermentation is an innovative route to ethanol production. It involves state-of-the-art technology, while only one commercial plant is currently in operation.

#### **i)INDIRECTLY-HEATED BIOMASS GASIFICATION: PROCESS SIMULATION**

The process was modelled using Aspen Plus, the only commercially available software capable of handling sorbent solid components, to handle the rigorous material and energy balances included. The NREL’s design for the Batelle Columbus Indirectly Heated Gasifier is taken as a base case. The system thermodynamics are modelled using the RKS-BM (Redlich Kwong Soave with Boston Matthias alpha function) – as proposed by Spath et al. [23]- which is widely used in coal treatment processes. The Gasifier (RYield) operates at low pressure (1.6 bara) and is of a dual fluidized bed design. Heat for the endothermic gasification reactions is supplied by circulating hot olivine sand between the Gasifier vessel and the char combustor (RStoic). The gasification medium is steam. The temperature of the boiler is 982C and the temperature of the Gasifier results from the energy balance around the Gasifier and the boiler. The temperature of the Gasifier is 870C. Particulate removal is performed through cyclone separators. The majority of the olivine and char (99.9% of both) is separated in the primary Gasifier cyclone and sent to the char combustor. A secondary cyclone removes 90% of any residual fines. The char that is formed in the Gasifier is burned in the combustor to reheat the olivine. The primary combustor cyclone separates the olivine (99.9%) from the combustion gases and the olivine is sent back to the Gasifier. Ash and any sand particles that are carried over are removed in the secondary combustor cyclone (99.9% separation) followed by an electrostatic precipitator which removes the remaining residual amount of solid particles. The sand and ash mixture is diluted with water and discarded as waste. After immediate cooling of the produced gas at a temperature of 148C and at a point where the gas remains above its dew point, a filter is used to remove ash from the gas stream. The gas is then cooled by heating deaerator feed water. Any condensate is removed, and the gas is then scrubbed (RadFrac) with water to remove HCl, along with most of the residual ammonia content and any residual particulate material. The product gas still contains H<sub>2</sub>S.

The flowsheet is illustrated in Figure 1. According to the results obtained from AspenPlus, the gasification of 40tn/hr of dry pine chips produces 45tn/hr (356 kTe/yr) of clean Synthesis Gas. Estimation of cold gas efficiency, defined as the fraction of the higher heating value (HHV) of the produced SynGas to the HHV of the dry biomass feed, shows that it is approximately 73% [24].

#### **ii)PRODUCTION OF ETHANOL BY SYNGAS FERMENTATION:PROCESS SIMULATION**

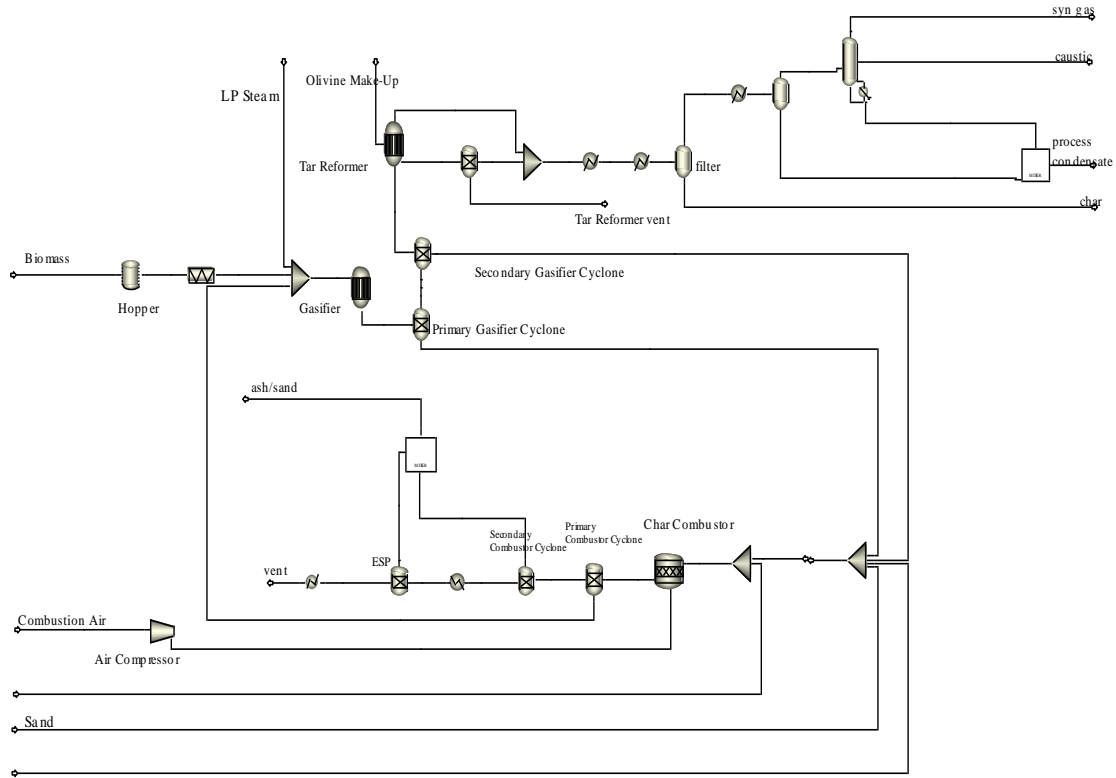
Feed gas containing predominantly CO and H<sub>2</sub> passes into the fermenter. The fermenter (RStoic) operates at 3 bar and 37 C, and contains bacteria (*C. Ljundahlia*) which convert part of the CO and H<sub>2</sub> into ethanol, producing CO<sub>2</sub> and consuming water. Unconverted gas which passes out of the fermenter is cooled and chilled to maximize recovery of ethanol. The gas is then scrubbed by a counter-current flow of water, recycled from the distillation area. Liquor from the fermenter, which contains 3% w/w ethanol and 10,000 mg/l of biomass, is pumped through the membrane unit. The membrane permeate passes to the distillation area, and the thickened retentate returns to the fermenter, first passing through a cooler to remove the heat generated by the ethanol formation and thereby control the fermenter operating temperature. The fermentation liquor from the membrane separator is heated to 100 C and passes into the first distillation column which operates at 1.9 bar (Stripping column-RadFrac). The column removes dissolved gases (mostly CO<sub>2</sub>) and removes about 92% of the water as a bottoms product. The gases are vented to the atmosphere from the top of the column. The ethanol is withdrawn from the column as a vapour side stream of 28% w/w ethanol. The column bottoms water contains 0.2% w/w ethanol and is recycled to the fermenter via an interchanger where it preheats the column feed. The nutrients present in the column feed are retained in the column bottoms. The vapour stream passes to the rectifier (RadFrac) to produce 92.5% w/w ethanol which is withdrawn from the column top as vapour at 116 C. The concentration is limited by the existence of an azeotrope at 95.6% w/w ethanol. The rectifier bottoms stream contains 0.05% ethanol and is cooled and passes to the gas scrubber, before returning to the fermenter. There is a small make-up of water to the top of the scrubber. The ethanol vapour stream from the rectifier passes to the ethanol dehydration unit. Molecular sieve dryers are used to remove the remaining water to produce anhydrous ethanol (>99.7% w/w). The dryer system consists of a pair of vessels which contain 3A molecular sieve. The ethanol/water vapour stream passes through the dryer bed and the water vapour is absorbed by the molecular sieve. The dried ethanol vapour is condensed and pumped off to storage. A portion of the dried vapour is returned to the second bed, which operates under vacuum. The water is desorbed, and the water/ethanol vapour stream is condensed in a vacuum set. The condensed ethanol/water mixture is reheated to 95 C by interchange with the dry product stream and recycled to the rectification column.

SynGas fermentation is also simulated using AspenPlus. The input data are collected from personal communication with A.C. Kokossis and A.Yang [25]. The base property method used is BWR-LS (BWR- Lee Starling), which is the only method, that is able to satisfy the convergence criteria. BWR-LS is suited for gas and

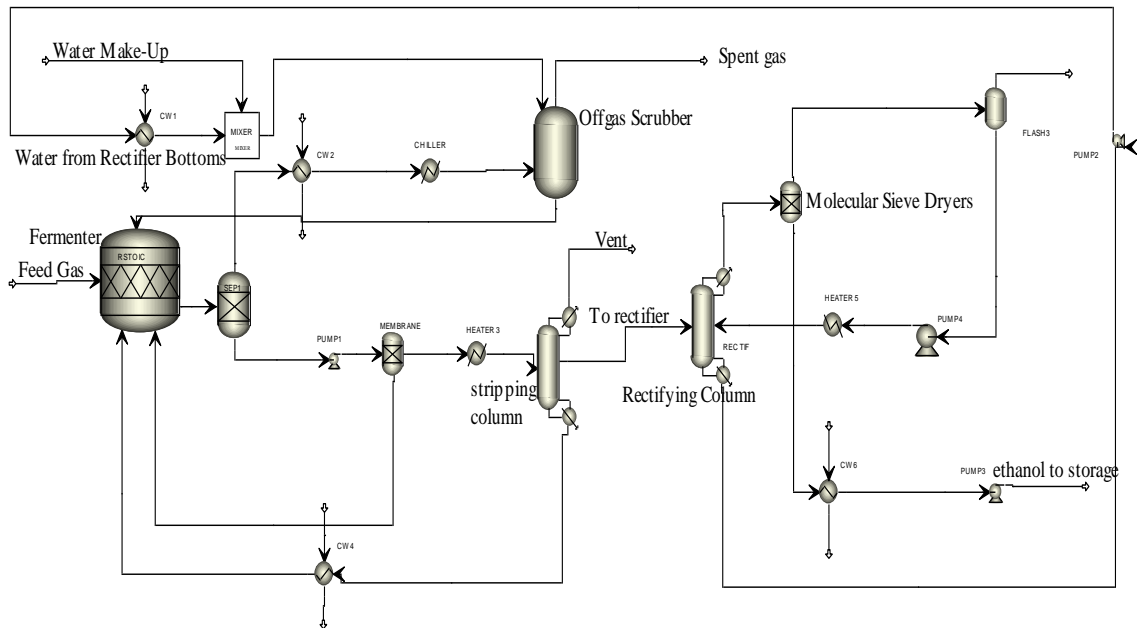
refinery processing, as well as for hydrogen- containing systems and therefore, it is assumed that the achieved results are feasible. Simulation of the process provides a capacity of approximately 4tn/hr (33kt/yr) of dry Ethanol from 15tn/hr of feed gas. The flowsheet of the simulated process is shown in Figure 2 [24].

Details on both processes as well as stream results can be found in *Tsagkari M.V.*

**Figure 1.** Aspen Plus process flow diagram detailing the indirectly heated biomass gasification [24]



**Figure 2.** Aspen Plus process flow diagram detailing the Ethanol production by SynGas Fermentation [24]



## RESULTS & CONCLUSIONS

Careful application of the costing methods for both processes produced the results in Table 4.

**Table 3.** Comparison of costing techniques

Process	Biomass Gasification (356kt/y)	Syngas Fermentation to Ethanol (33kt/yr)
<b>Capital Cost in M\$,2011,US</b>		
Wilson	110	42.5
Taylor	41.75	36.31
Bridgwater	147.7	36.1
Brown, Fromme, Klumpar	151	61.27
Petley	122	31.25
Lange	86.1	16.3
<b>Average</b>	109.76	37.29
<b>Reported Cost (M\$,2011,US)</b>	<b>67.9</b>	<b>22.36</b>

Wilson's method overestimates the gasification cost by 62%, which is mostly due to the temperature correction factor ( $F_T$ ), which grows exponentially with increasing temperatures and thus, the contribution of temperature into cost growth is misinterpreted. Syngas fermentation, on the other hand, is overestimated by 90%. We would expect invalidation of the model for this process because of negative economies of scale (multistreaming). Moreover, Wilson derived his method on a collection of 16 processes, 4 of which were solids-fluids type and 12 fluids only- the method inevitably is more suitable for fluid processes. Taylor's method (-38%) correctly predicts the gasification, while it misses syngas fermentation capital costs by 62%. Taylor accurately correlated temperature impacts on process cost growth. However, it should be noted that Taylor derived his method based on the regression analysis of 45 real plants during the 1970's. Unfortunately, this period was a time of sharp currency fluctuations in the UK, which cannot be appropriately represented by the CEPCI used in this study. Taylor proposes recorreling his equation, but collection of a significant amount of real process data for such a work is quite a toilsome task. Bridgwater overestimates gasification by 117%, while overestimating fermentation by only 62%. Brown et al.'s method shows the most disappointing results despite the huge effort and detailed methodology presented by the authors with overestimation greater than 100% for both processes. Petley's correlation overestimated by 80% the gasification and by 40% the fermentation. As expected, Lange's Heat losses method predicted quite accurately (26%) the strongly exothermic gasification process, as well as the syngas fermentation, which is quite surprising as the author states that his method is not appropriate for heat neutral reactions.

All in all, Lange's method provides the best prediction for both processes, showing the least deviation from the reported costs. However, the author claims that his correlation should not be trusted for heat neutral reaction processes and thus, its applicability shall be viewed with caution. From the rest of the methods, it is Taylor's method which shows the best results (-38%, +62%). His method is quite thorough and includes a lot of parameters which affect the cost. Some of them (such as storage time, multistreaming, materials of construction) are not certainly known during the R&D stages of a project and thus, the method belongs mostly to Class 4 estimates and not Class 5. However, as mentioned earlier, its "good" results should be viewed with distrust. All techniques were derived from conventional chemical or petrochemical plants, which might be similar but certainly do not reflect the specifics of modern biorefineries and therefore, their credibility is dubious.

## RECOMMENDATIONS

It's beyond doubt that no significant progress has been made in the field of early cost estimation (Class 5). Existing costing techniques are outdated and their results should be used with extreme caution even if they seem to correctly approach the cost. New or modified costing techniques should be created to meet the specifics of the pioneer biorefinery processes in order to accurately estimate their capital cost. This requires the collection of extensive process and cost data from existing biorefinery plants and extensive study of their inherent correlations to offer a quick tool for the R&D engineers to select the optimal processing route.

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## REFERENCES

- [1] M. Peters, K. Timmerhaus and R. West, *Plant Design and Economics for Chemical Engineers*, McGraw-Hill, 2004.
- [2] A. K. Coker, *Ludwig's Applied Process Design for Chemical and Petrochemical Plants*, Gulf Professional Publishing, 2007.
- [3] B. Amigun, "Processing Cost Analysis of the African Biofuels Industry with special reference to capital cost estimation techniques," University of Cape Town, Cape Town, 2008.
- [4] G. J. Petley, "A Method for Estimating the Capital Cost of Chemical Process Plants-Fuzzy Matching," Loughborough University, Loughborough, 1997.
- [5] J. R. Cooper, *Process Engineering Economics*, Marcel Dekker, Inc., 2003.
- [6] S. Ereev and M. Patel, "Recommended methodology and tool for cost estimates at micro level for new technologies," PROSUITE Deliverable 2.2, 2011.
- [7] A. Gerrard, *Guide to Capital Cost Estimating*, IChemE, 2000.
- [8] P. Christensen, L. R. Dysert, J. Bates, D. J. Burton, R. C. Creese and J. K. Hollmann, "Cost Estimate Classification system-as applied in engineering, procurement, and construction for the process industries," AACE, Inc., 2005, 2011.
- [9] R. Williams, "'Six-tenths factor' aids in approximating costs," *Chemical Engineering*, vol. 54, pp. 124-125, 1947.
- [10] H. J. Lang, "Simplified approach to preliminary cost estimates," *Chemical Engineering*, vol. 55, pp. 112-113, 1948.
- [11] C. Karaboiki, "Techno-economic analysis of solid recovered fuel (SRF) energy systems via gasification with emphasis on fixed capital investment," NTUA, Athens, 2013.
- [12] D. R. Woods, *Rules of Thumb in Engineering Practice*, John Wiley & Sons, 2007.
- [13] J.-P. Lange, "Fuels and chemicals manufacturing: guidelines for understanding and minimizing the productions costs," *CATECH*, vol. 5, pp. 82-95, 2001.
- [14] F. Cherubini, "The biorefinery concept: Using biomass instead of oil for producing energy," *Energy Conversion and Management*, vol. 51, pp. 1412-1421, 2010.
- [15] E. d. Jong, A. Higson, P. Walsh and M. Wellisch, "Bio-based Chemicals: Value Added Products from Biorefineries," IEA Bioenergy –Task42 Biorefinery, 2012.
- [16] Compass International, *Global Construction Cost and Reference Yearbook*, Compass International, Inc, 2012.
- [17] "<http://www.chemengonline.com/pci-home>," [Online].
- [18] R. Sinnott, Coulson and Richardson's *Chemical Engineering*, Volume 6, Elsevier, 2005.
- [19] G. Wilson, "Capital Investment for chemical plant," *British Chemical Engineering and Process Technology*, vol. 16, pp. 931-934, 1971.
- [20] J. Taylor, "The 'process step scoring' method for making quick capital estimates," *Engineering and Process Economics*, vol. 2, pp. 259-267, 1977.
- [21] A. Bridgwater, "Step counting methods for preliminary capital cost estimation," *Cost Engineering*, vol. 23, pp. 293-303, 1981.
- [22] I. V. Klumpar, R. F. Brown and J. W. Fromme, "Rapid Capital Estimation based on process modules," in *AACE Transactions*, 1983.
- [23] P. Spath, A. Aden, T. Eggeman, R. M., W. B. and J. J., "Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier," National Renewable Energy Laboratory, 2005.
- [24] M. V. Tsagkari, "Cost models for biorefinery processes: a comparative study of heuristic and computational methods," National Technical University of Athens, Athens, 2011.
- [25] A. Kokossis and A. Yang, *Personal Communication*, 2011.