Pathways to Decarbonize the PVC Value Chain in 2050

Appendix 4

Cost Estimating Methodology and Assumptions

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General Considerations

Estimating the fixed capital requirements and cash operating costs for the various processes involved in the current and potential future operations in the PVC chain is difficult and carries significant uncertainties because of the nature of the information available and the uses to which it must be put. The first difficulty is that, in many instances, the bases for the cost information available are not clearly defined: it may be unclear whether they represent engineering estimates of study or preliminary estimating quality, averages of other reported values, or extrapolations from other analogous examples of unknown quality. In most cases the values are reported for a plant or process at standard capacity, often simply in terms of dollars per unit of capacity, with no indication of the range of validity of such a single point estimate. Finally, many estimates do not provide breakdowns of the components of the estimated costs. Consequently, it is not possible to determine how they would change under circumstances that differed from those under which the estimates were developed.

In searching the literature and engaging in conversations with industry experts to develop cost estimates for this research, sources have been sought that provided information at the greatest level of detail possible. For fixed capital requirements the desirable level of detail would include estimates of delivered equipment costs for major items, total direct plant costs, indirect costs for engineering and construction, fees, owners' expenses, and the contingency allowance applied to the estimate. For cash operating costs the desired level of detail would include estimates of unit costs of the materials and supplies consumed, the utilities and fuels consumed, the labor required, and the fixed costs related to the capital invested such as maintenance materials and supplies, taxes and insurance.

If information is presented with this level of detail and is accompanied with text that describes the process that the estimate represents, the analyst can review it critically and form judgments about its quality and how it may be used in the analysis. This level of detail would also allow estimates of how the costs would change when the process or plant size is larger or smaller than the one on which the estimate is based through the use of scaling exponents.[1] If

[1] Scaling exponents are used to relate costs at different capacities by relationships of the form $(C_2/C_1) = (S_2/S_1)^n$, where *n* is the scaling exponent. A value of 0 implies no change in cost with capacity; a value of 1 implies a linear relationship between cost and capacity. Many process plants have values of n between 0.6 and 0.9 for capital requirements. Scaling exponents should also be applied to the components of operating costs, although this is less commonly done.

this level of detail is unavailable, or if the required capacity or projected operating conditions are significantly different from those that applied to the original estimate, considerable engineering judgment is required to develop appropriate estimates applicable to a new situation.

If the original estimates were developed at the preliminary level of estimating quality, the uncertainty in the derived estimates of capital requirements would be in the range of ±20% at best. If the original estimate were developed at the study level, the uncertainty in the derived estimate of capital requirements would be in the range of ±30% at best. For well-known processes with established cost histories, order-of-magnitude estimates of capital requirements may be made based on a known location, capacity, and utilities and site services availability, but the uncertainties could be as high as 50%. In general, uncertainties in estimated cash operating costs are lower, typically of the order of 5–15% unless there are significant uncertainties in the unit costs of materials, energy, or labor.

Investment decisions are not made on the basis of fixed capital requirements and cash operating costs alone. Working capital requirements must be included, but they are small for most of the technologies under consideration here. However, the investors expect a return on the capital committed that is commensurate with the project's perceived risk, the financing options available, the time projected for engineering, design, construction and start-up, and other potential uses for the funds. This is often estimated by applying a capital charge factor to the amount of capital at risk. For this study, the capital charge factor has been assigned a value of 10% of the capital requirement per year, and it is added to the estimated cash operating cost to give an estimated total operating cost.[2]

In estimating the costs of decarbonizing the PVC chain with different technologies, it was necessary to obtain cost information on CO2 removal from combustion gases, on the cost of production of blue hydrogen, and on the cost of capturing emissions from the production of ethanol by fermentation. The following sections describe the results of analyzing the cost information contained in sources relevant to the technologies used in the PVC chain

Capturing CO2 Emissions from Combustion and Process Gases

A report from NETL (2014) containing estimates of the costs of capturing CO2 from industrial sources contains brief process descriptions and summary cost estimates for decarbonization systems applied to nine industrial processes. Six of the processes generate CO2 at purities exceeding 99%, while three of the processes—production of refinery hydrogen, steel and cement—generate CO2 at concentrations between 23 and 45%. The decarbonization processes for capturing the emissions from low-purity gases are based on treating CO2-containing gases with aqueous solutions containing MDEA, which is circulated in a closed loop to remove the

[2] For the production of blue hydrogen by autothermal reforming of natural gas, a capital charge factor of 5.8% has been used so as to be consistent with the methodology used by the source of the estimates, NETL (2022)

CO_2 from inert gases and concentrate and compress it for sequestration.[3] These three processes generate much higher concentrations of CO_2 than would be found in combustion gases. Since these gases may contain on the order of 5% CO_2 by volume, the costs estimated for them cannot be applied directly to decarbonization of PVC-chain emissions. The process descriptions and cost information on these applications is presented at a rather low level of detail, which limits the extent to which they can be used to interpolate or extrapolate the costs to PVC chain applications. Specifically, the capital requirements breakdown for the base erected costs for treating cement kiln gases includes only estimates for the following accounts: Ductwork and piping; MDEA unit; pre-cooler–compressor–after-cooler unit, cooling-water chiller, boiler and balance of plant (I&C, site, buildings, etc.). The MDEA unit constitutes over 60% of the total erected cost.

The text describing the estimating methodology notes that some elements of the costs presented, as for the compressors and boilers, were estimated by scaling values obtained from vendor quotes, while others were obtained by scaling values from other studies. The costs presented for the MDEA unit were scaled from an earlier NETL (2011) study of the costs of blue hydrogen production, but no description of the scaling methodology used was given. The estimating methodology described in the two NETL reports, and the presentation of the results were similar, but the MDEA units were quite different. The earlier report on blue hydrogen production operated at about 26.5 atmospheres and was sized to treat about 1.4 million tonnes/year of CO_2, at a concentration of about 12.7%, while the later report operated at atmospheric pressure. The one treating the combustion gases from cement kilns was sized to treat about 1.1 million tonnes/year CO_2 at 22.4% concentration.

The estimated capital cost of the MDEA system for treating atmospheric pressure cement kiln gases was only about 75% as high on a per tonne CO_2 basis as that for blue-hydrogen production, but costs for treating combustion gases would be higher. This is because almost 4.5 times the volume of gases would have to be treated at the same CO_2 production rate, which would increase the size of the vessels required to treat them. Furthermore, the lower CO_2 concentration would require a larger circulation rate of MDEA solution, thereby increasing utility requirements per unit of CO_2 treated, and likely would limit the efficiency of the scrubbing process as well. Only those sections of the process that treated pure CO_2, such as the main CO_2 compressor, would have the same costs.

In order to estimate the increase in costs attending the treatment of combustion gases containing 5% CO_2 compared with gases containing 22.4%, preliminary material and energy balances were developed for both cases for plants sized to treat gases containing 1.14 MMt/yr. CO_2. The balances were then used to develop estimates of costs for major items of equipment and services including vessels, drums and tankage, exchangers, rotating equipment, boilers and cooling and refrigeration systems, as well as the requirements for steam, power, and water

[3] Methyldiethanolamine, or MDEA, is a solvent that removes CO_2 and other acid gases from gas streams by sorption at low temperatures and is regenerated by driving off the CO_2 at higher temperatures.

consumption. The estimated number of stages in the scrubbing vessel was increased by 30% to account for the treatment of a more dilute CO2 stream. The difference in bare erected costs were estimated using typical factors for installation, and the total cost difference was then estimated using the same allowances for engineering and fees, contingencies, and owner's costs that were applied in the estimation of costs to treat combustion gases from kilns.

Using this methodology, the capital requirements to treat the same amount of CO2 at a concentration of 5% are estimated to be 80% higher than those to treat gases containing 22.4% CO2, with costs for the MDEA system essentially doubling and costs for utilities increasing by 50%. The breakdown of operating costs was reviewed to determine if the estimates presented remained appropriate to the treatment of more dilute gases, and the changes developed from the energy balances were applied to the utility accounts. This resulted in an increase in estimated total operating costs of 45%, with utilities costs increasing by 60% and labor and other variable operating costs increasing by 20%.

The adjusted costs were applied to decarbonizing the processes in the PVC chain by updating them to 2020 using appropriate indices, applying fuel and power costs established for the Case and year of the estimate, and scaling the costs to the amount of CO2 being treated. Because the CO2 concentration was lower, the estimate of carbon-capture efficiency was decreased from 95% to 90%, even though the number of contact stages in the scrubbing vessel was increased. Given that the original bases used to develop the cost estimates upon which this estimate is based are not defined clearly, the uncertainty in estimated capital requirements cannot be less than -10 to +30% in capital requirements and -5 to +15% on operating costs.

Amine-based scrubbing processes have been used for many years in natural gas processing, refinery operations and the chemical process industries to remove acid gases from process streams, but they have not been used widely to remove CO2 from combustion gases as there has been no incentive to do so. Some demonstration plants for CO2 removal from combustion gases based on amine scrubbing have been built and operated with varying degrees of success, including one currently producing pharmaceutical grade CO2. It is to be expected that process improvements and cost reductions will attend the more widespread use of this technology to decarbonize combustion gases from industrial applications.

Estimating the Costs of Blue Hydrogen Production

The NETL (2022) study that compares hydrogen production technologies presents data and information on the processes used and the breakdown of estimated costs in far greater detail than the references used to estimate the costs of CO2 removal from gas streams using MDEA scrubbing technology. This study presents technical and economic analyses of grey hydrogen production by methane steam reforming and of blue hydrogen production by methane steam reforming and by autothermal reforming. The latter has 21% lower capital requirements and 3% lower total operating costs than methane steam reforming with CCS and is the technology of interest for providing blue hydrogen for decarbonizing the PVC chain.

The estimating methodology used in the study is described clearly and is based on a combination of a bottoms-up approach using vendor cost estimates as well as a scaling approach based on data from previous studies carried out by the contractor. Contingencies were applied at both the process and project level, but no process contingency was applied to the MDEA CO2 removal system because of its history of commercial-scale use in refinery hydrogen-production plants. However, the estimators recognize that CCS systems in this application are not fully mature, and its early deployment may result in higher costs. On the other hand, costs should decrease in the future as the learning curve develops. Construction costs were based on a USGC model but used Midwest labor rates. The basis for the cost estimates is December 2018, reasonably contemporaneous with the base case year of 2020 used in the study of PVC decarbonization.

The report presents both a simplified process schematic and a process flow diagram showing major equipment items including vessels, exchangers, rotating equipment, etc. as well as the mass flows, temperatures and pressures throughout the process.[4] A major equipment list is provided showing equipment descriptions and types, design conditions, the numbers installed and spares. The cost estimate is built up in detail showing the costs of equipment and materials, direct and indirect labor required for installation (summing to the bare equipment costs), costs for engineering and fees, and process and project contingencies (summing to total plant costs) for all major items. A breakdown of estimated owner's costs is provided as well as the costs incurred during construction and commissioning, and the sum is the estimate of the total as spent capital. For a plant with the capacity to produce 660 t/day of hydrogen, the estimated capital requirement is $968 million, with an estimated uncertainty of -15% to +25%. This approach is consistent with a Class 4 estimate.

A well-developed breakdown of operating costs is also presented. It is based on a staffing profile for operating labor with an allowance of 25% for administration and support. Maintenance materials and labor costs are taken as percentages of the estimated capital with values that are typical of those used in the CPI. Consumable materials and energy costs are estimated from the material and energy balances developed for the process and their unit costs. The costs of CO2 transportation and sequestration are developed separately and estimated based on expected costs for sequestration at a number of locations. The total estimated cash costs amount to $1.33/kg H2. Applying a return to capital of 5.8% gives a total cost of $1.59/kg H2. Fuel costs amount to 48% of the total with natural gas at $4.42/MMBTU. The estimated uncertainty due to capital is about -2% to +4%. No uncertainty was applied to the other components of the operating costs, but the total uncertainty would be in the range of -5% to +10%.

The amount of hydrogen required for decarbonization of a site depends upon the amount of unabated CO2 emitted there. This varies by a factor of 50 from the largest CHP plant to the smaller self-standing PVC plants.[5] We assume that all these plants would purchase hydrogen from merchant sources, whose capacities would increase as the demand grows. Merchant hydrogen costs were estimated based on the price of natural gas for the case and year under study, assuming that typical hydrogen plant sizes would double between 2030 and 2040 and redouble by 2050, with the attendant economies of scale.

[4] Also called an Energy and Mass Balance Diagram.

[5] Excluding the small specialty plants and the smallest two PVC plants.

For decarbonization of crackers, either by substitution of hydrogen for natural gas fuel or by electrification of the cracker, the issue of the disposition of cracker fuel gases must be addressed.[6] The fuel gases are mostly methane and for this study we have assumed that they would be reformed in on-site ATR plants with CCS systems rather than being flared or burned for fuel elsewhere. The hydrogen produced would either be burned along with merchant material or sold at the estimated price for merchant material. Because the cost of an on-site air separation plant in the 660 t/day ATR/CCS complex amounts to almost 40% of the capital requirements, we assume that the smaller plants reforming fuel gases would purchase oxygen from merchant sources rather than generate it on-site. Diseconomies of scale will make this material more costly to produce than merchant hydrogen.

Capturing CO2 Emissions from Corn-Based Ethanol Production

Cost estimates for the capture of emissions from the fermentation step of corn-based ethanol production have been developed by the National Energy Technology Laboratory (NETL 2014) and are presented in Chapter 7, Sections 7.1 through 7.1.10 of its report. The information in these sections include a brief discussion of (1) the size distribution of plants in the industry and the two major process options used, (2) the impact of plant size on the style of compressor required to handle the CO2 recovered, and (3) the assumed CO2-to-ethanol ratio and operating conditions in the fermentation step. The sections also include a simplified block-flow diagram of the assumed CO2 recovery process, tables summarizing the results of the cost estimate, and an estimate of the impacts of scale on the required break-even price for recovered CO2.

The description of the industry is reasonable given the amount of capacity additions and deletions and improvements in plant operating practices that have occurred since the report was issued. The plant size chosen, 50 million gallons per year, is at the high end of the mid-range of plants, which is a reasonable base case although there are a number of plants that are two to four times as large. The specific emission factor chosen is quite close to the stoichiometric amount of CO2 produced in a simplified fermentation reaction and is generally representative of reported industry practice.

However, the authors describe the fermentation reactions as taking place at 140–180° C and the CO2 leaving the fermentation step being 100% pure. These are both inaccurate assumptions. Fermentation typically takes place at temperatures in the range of 30–32° C, although newer plants practicing simultaneous sachorification and fermentation (SSF) may carry out fermentation at 32–35° C.[7] At these temperatures, the average composition of the gases leaving the fermentation reactor is about 1.8% ethanol by weight, 2.2% water and only 96% ethanol. While a gas cooler is shown in the schematic in Exhibit 7-2, the cooling duty shown in the energy balance in Exhibit 7-3 does not reflect the amount of heat that must be removed to condense most of the ethanol and water before the CO2 stream enters the

[6] Fuel gases are generated as a byproduct of the reactions that produce ethylene from the feedstock used. The amount generated depends on the feedstock composition. See Appendix 2, Industry Structure, for a more-complete description of cracker operation.
[7] In the SSF process both steps are carried out sequentially in the same vessel, reducing both capital requirements and process cycle time.

compressor. The estimated power requirements for the compressor and the duty of the aftercooler are essentially correct even though the inlet gas is not 100% CO2. The cooled, compressed CO2 at 2200 psig would be about 99.94% pure and of pipeline quality.

The estimated capital requirements in Exhibit 7-5 contain equipment cost estimates for the compressor, its pre-cooler, and a chilling unit, as well as estimates of the costs for ductwork and piping and the balance of the plant. These presumably sum to the total of direct plant costs, but the ratio of the total equipment costs to the total direct costs is some 2.5 times higher than for a typical plant, so direct plant costs for equipment installation, services, etc. may have been underestimated. Furthermore, the estimate apparently does not recognize the fact that the fermentation reactors that generate the CO2 operate batchwise whereas the compressor would operate most efficiently with a steady gas flow at its suction. It is not clear that a gas accumulator or a specialized control system for the compressor has been included in the estimate.

The allowance of 8.5% of direct costs for engineering, contract management, home office expense, and fees is also considerably lower than is typical for such projects and may have been underestimated. On the other hand, the allowance for owner's costs of more than 24% of the estimated total plant costs is far higher than normal for a typical project. An estimating contingency of 20% was applied, which would be appropriate for a study to preliminary level estimate but would have been more appropriately applied to the total project cost, which includes the owner's costs. If the direct plant costs are correct, however, the net result of substituting more typical values for the ones used and applying the 20% contingency properly would be a relatively small change in the estimated total.[8]

The cash operating cost breakdown presented in Exhibit 7-5 presents estimates of total labor costs, the costs of taxes and insurance, variable costs for maintenance materials, and other consumables and power costs. No contingency allowance was applied to the operating cost estimate. The staffing level derived from the labor cost estimate appears reasonable. The purchased power requirement apparently reflects only the amount used by the compressor, with no allowance for balance-of-plant power. The estimate of property taxes and insurance of less than 1.6% of the total capital requirement is lower than typically assumed, but the estimated variable O&M costs are significantly higher than typically assumed, and no breakdown of its components is reported. Substituting more-typical values for the ones used and applying a 10% contingency allowance does not result in a significant change to the estimated total operating costs.

The assumption that such a project could be completed within a year is reasonable, subject to the ability of a compressor manufacture to provide the equipment within this time frame and requires the application of a total-as-spent multiplier of 1.022 to the overnight costs used in the

[8] The American Association of Cost Engineers recommends the application of contingency allowances of 15–30% for budget type estimates of Class 4 or 5, which describe the type of estimate presented in this case.

estimate. An additional multiplier of 1.01 was applied in recognition that this would be a retrofit project and so more costly than a greenfield one, but this is a very low allowance for a project that would have to be implemented while the plant remains in operation. Finally, the cost base chosen was mid-2011, so they must be adjusted to 2020 with a multiplier of 1.018.[9]

Given the errors in the assumed process description and the ambiguities in the estimated capital requirements and operating costs described above, this estimate should be considered no better than of study level quality and the uncertainty in the estimated values is likely in the range of -10 to +30%. The estimated dependence of total operating cost, or breakeven CO2 selling price, on plant size is presented in Exhibit 7-9 and shows the expected relationship, with unit costs decreasing as size increases. No information is provided on the methodology used to scale the costs with size, but the scaling exponent implied from the curve is about 0.71, which is not unreasonable.

Works Cited

NETL. 2022. "Comparison of Commercial State of the Art Fossil-Based Hydrogen Production Technologies" DOE/NETL-2022/3241. April 12, 2022.

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[9] The ratio of the CEPCI indices is 596.2/585.7