

Techno-economic analysis of a novel small-scale blue H₂ and N₂ production system

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ABSTRACT

This study presents an economic analysis of a blue H₂-N₂ production system, using a novel intensified reformer system with a hydrogen production efficiency of 80%. The system's ability to produce both high-purity H₂ and N₂ creates opportunities for small-scale blue H₂ and distributed ammonia production. The system consists of three identical, optimized fixed-bed reforming reactors, a heat recovery system, and shift reactors. A dynamic model was developed to simulate three small-scale H₂ production systems: 2.8 tpd, 7.1 tpd, and 17.1 tpd, enabling an evaluation of their economic viability. The results indicate that the cost of H₂ production ranges from 2.7 to 3.1 USD/kgH₂. Sensitivity analysis reveals that natural gas and CO₂ transportation costs have a significant impact on the variability of H₂ price. This research provides valuable insights into the economic feasibility of small-scale blue hydrogen production, offering a pathway to support the broader adoption of hydrogen technologies in advancing a cleaner energy future.

Keywords: Dynamic Modelling, Process Intensification, Process Design, Technoeconomic Analysis, Hydrogen, Nitrogen.

INTRODUCTION

The use of hydrogen is critical for decarbonization efforts, serving as a replacement for fossil fuels. Approximately 80% of the projected future hydrogen demand stems from emerging applications, with around 27% attributed to the transportation sector [1]. Hydrogen use in medium- and heavy-duty vehicles, maritime shipping, aviation, and rail is among top priorities of the US Clean Hydrogen Strategy and Roadmap [2]. Manufacturers are already transitioning towards hydrogen-powered heavy-duty trucks, utilizing either fuel cells or internal combustion engines [3]. To support the widespread adoption of hydrogen, ensuring its availability and the development of infrastructure at points of use is essential.

However, centralized hydrogen production requires transporting and storing hydrogen, which has proven to be challenging and a significant bottleneck in hydrogen implementation. One proposed solution is blending hydrogen with the current natural gas infrastructure [4]. However, this practice has faced criticism due to safety

concerns. Another approach under consideration involves storing hydrogen in the form of ammonia or methanol. This method is considered more economical for larger scale applications, such as continental export and import, given the long supply chain to the point of use [5].

An alternative strategy is smaller, distributed hydrogen tailored to the specific local needs. This approach aligns with hydrogen implementation in the transportation sector, which requires numerous hydrogen refuelling stations along vehicle routes. On-site hydrogen production can be achieved through either green hydrogen or blue hydrogen, depending on the location characteristics and the availability of energy sources and materials. Green hydrogen, while more environmentally friendly, is less cost-effective compared to blue hydrogen [5]. Locations with access to natural gas, biogas or limited renewable energy resources, should consider blue hydrogen technology as a more cost-effective solution for hydrogen production.

The discussion regarding blue hydrogen technology has predominantly focused on large-scale industrial

applications. Steam methane reforming with carbon capture storage (SMR-CCS), the most mature blue H₂ technology, continues to face challenges in achieving CO₂ capture rates above 80% [6], primarily due to CO₂ dilution in the flue gas stream. However, emerging technologies like chemical looping (CL) hydrogen production, show promising potential for blue hydrogen production, as they can achieve higher CO₂ concentrations in the downstream process, thereby facilitating more efficient carbon capture [7].

Our previous research demonstrated an optimized fixed-bed reactor for chemical looping hydrogen production that generates high concentrations of hydrogen while simultaneously producing concentrated streams of CO₂ and N₂ in three separate streams [7]. Our following work presented an optimized control strategy for an integrated multi-reactor system, incorporating heat recovery system and shift reactor components, specifically designed for small-scale hydrogen and nitrogen production [8].

This study presents a techno-economic analysis of a novel blue hydrogen and nitrogen production system designed to meet local demands. Prior studies primarily focused on large-scale hydrogen production, whereas this work provides cost estimates for small-scale distributed H₂ production. Thus, it provides valuable insights for stakeholders assessing the feasibility and economic viability of small-scale hydrogen deployment, enabling more informed decision-making on investment and implementation.

METHODOLOGY

Small-scale Definition

First, the scale of H₂ production must be defined. Most studies on CL-based H₂ production plants focus on capacities above 60 tons/day [9-13]. Consequently, we define H₂ production of less than 20 tons/day as small-scale. This scale is intended for local applications, such as supporting hydrogen refuelling stations [14].

The Blue H₂-N₂ Production System

Figure 1 illustrates the novel blue H₂-N₂ production system discussed in this work. The system consists of three major subsystems: (i) a train of reforming reactors, (ii) a heat recovery system, and (iii) a gas purification system. The reactor system uses the intensified reforming reactor discussed in [7], operating on the chemical-looping reforming (CLR) principle, with successive oxidation (OX), reduction (RED), and reforming (REF) stages. A minimum of three reactors operates simultaneously at different CLR stages to ensure a continuous product stream. In each stage the reactors produce three different streams: a N₂-rich stream from OX, a CO₂-rich stream from RED, and a H₂-rich stream from REF.

The heat recovery system includes preheaters and steam generators installed along each downstream path to preheat the reformer feed. The hot streams used in the heat recovery system originate from the product streams of the multi-reformer system.

The gas purification system is designed to purify the primary product of each stream. In the N₂-rich stream, a methanator converts remaining CO and CO₂ into methane. In the CO₂-rich stream, a condenser removes steam to ensure a high CO₂ concentration output suitable for capture. For the H₂-rich stream, a water gas shift reactor converts remaining CO into H₂, and a pressure swing adsorption (PSA) separates H₂ from the other gases.

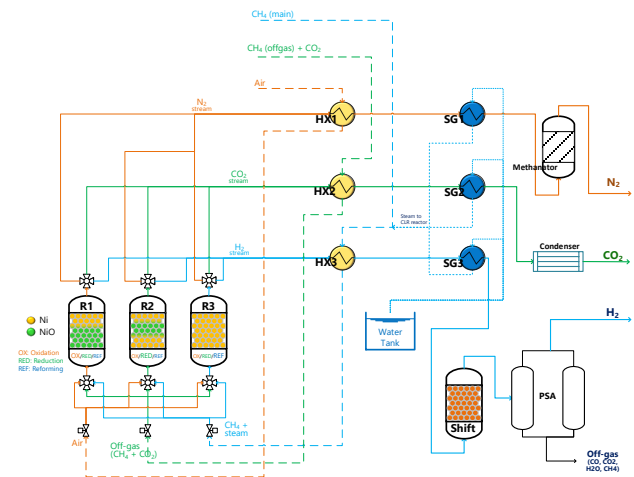


Figure 1. Simplified flowchart of the blue H₂-N₂ production system discussed in this work.

Modeling and Simulation

The fixed bed reactors for the reforming and shift reactors were modeled using a 1D heterogeneous model, that incorporates mass, energy, and momentum balances, reaction kinetics, and mass-heat transfer correlations. The kinetics of the CL reformer is documented in [7], while that for the shift reactor is provided in [8]. The heat recovery system was modeled as a counter-current heat exchanger, incorporating mass and energy balance equations. The system's compressor was modeled as a two-stage compressor with intercoolers, operating with a mechanical and polytropic efficiency of 90%.

Details on the integrated blue H₂-N₂ production model, along with its validation and dynamic optimization can be found in [8]. The system is comprised of a set of Differential-Algebraic Equations (DAEs), which were solved using the equation-based process modeling platform gPROMS Process 2023.2.0 [15], with the DAEBDF solver.

To simulate systems of different scales, the reactor's length-to-diameter (L/D) ratio was maintained, while the inlet molar flow rate and reactor volume were

adjusted. Key system parameters, such as the steam-to-methane ratio, the CO₂-to-fuel ratio during the reduction stage, and the steam distribution across the steam generators, were kept constant.

Economic Analysis

This techno-economic analysis covers from the purchase of natural gas to the downstream treatment of the CO₂ by product. The analysis was conducted following the framework outlined by DOE/NETL [16], as presented in Table 1.

Table 1: Methodology for techno-economic analysis used to estimate H₂ production cost based on the framework by DOE/NETL [16].

Components	Formula
<u>Capital Cost</u>	
- Bare Erected Cost of j-components (<i>BEC_j</i>)	$\sum BEC_j$
Total <i>BEC</i>	
Engineering Procurement and Construction (<i>EPC</i>)	10% (<i>BEC</i>)
Contingency (<i>CON</i>)	20%(<i>BEC</i> + <i>EPC</i>)
Total Plant Cost (<i>TPC</i>)	<i>BEC</i> + <i>EPC</i> + <i>CON</i>
Owner Cost & Other Financial Cost (<i>O&OFC</i>)	20% (<i>BEC</i>)
Total Overnight Cost (<i>TOC</i>)	<i>TPC</i> + <i>O&OFC</i>
Total as spent Cost (<i>TASC</i>)	<i>TOC</i> + 18% <i>TOC</i>
Capital Cost per Year (<i>TASC_{pa}</i>)	<i>TASC</i> × <i>CCF</i>
<u>Annual Operation and Maintenance Cost (<i>O&M</i>)</u>	<i>O&M</i>
Total Annual Cost (<i>TAC</i>)	<i>TASC_{pa}</i> + <i>O&M</i>
H ₂ cost (USD/kg)	$\frac{TAC}{H_{2prod}}$

The capital cost was determined by summing the costs of the individual reactor components, incorporating additional costs such as Engineering, Procurement, and Construction (EPC), contingency, and overnight costs, resulting in the Total as Spent Cost (TASC). The annual capital cost was then calculated by multiplying the TASC by a capital charge factor (CCF) of 0.159, which assumes an interest rate of 10% and a plant lifetime of 20 years [17]. Component costs for the blue H₂ and N₂ production system were estimated using reference costs provided by DOE/NETL [16], as summarized in [11], and scaled using the exponential scaling formula shown in Eq. (1). The required capacity of each equipment for this cost estimation was obtained from simulation. Additionally, a balance of plant cost – piping, instruments, controls, and facilities – was included and assumed to be 0.8 of the total equipment cost [17], except for the CL reactor component.

$$Cost_j = Cost_{ref} \left(\frac{Capacity_j}{Capacity_{ref}} \right)^f \quad (1)$$

The annual operating and maintenance (O&M) costs were estimated by combining fixed and variable costs based on assumptions from [11]. Labor costs were assumed to be 90,000 USD/year/person. The O&M costs also accounted for the transportation of CO₂ to local storage sites or for utilization. The transportation cost was based on estimates from [18] for 1 million tons CO₂ per year transported within a 100-mile distance. Table 2 summarizes the key assumptions for estimating operating expenses.

Table 2: Assumptions for the calculation of the operation and maintenance (O&M) costs.

Components	Formula
<u>Fixed Costs</u>	
Labor [USD/person/year]	90,000
Maintenance	0.7% <i>TASC</i>
Insurance	1.5% <i>TASC</i>
<u>Variable Costs</u>	
Oxygen carriers [USD/kg]	20
Shift catalyst [USD/kg]	50
Desulphurization/methanator catalyst [USD/kg]	15
Lifetime oxygen carriers [years]	2
Lifetime catalyst [years]	5
Natural gas cost [USD/GJ _{LHV}]	6.5
Steam cost [USD/ton _{steam}]	0.13
CO ₂ transportation for 1 Mtpa, <100 miles distance, with low monitoring [USD/tonCO ₂]	16.5
Electricity cost [USD/MWh]	55

RESULT AND DISCUSSIONS

We selected the proposed blue H₂-N₂ production system with three H₂ production scales: 2.8, 7.1, and 17.1 tons of H₂ per day. These production scales represent the potential demand for medium-sized hydrogen refuelling stations [14]. Simulation results for each scale were used to extract key parameters for the main equipment in the integrated system, as shown in Table 3. These parameters were subsequently used to estimate equipment costs using Eq. (1) and the points discussed in the previous section.

The techno-economic analysis was conducted using the methodology previously described, with the results summarized in Table 4. The analysis shows that the cost of blue H₂ for the proposed small-scale system ranges between 2.73 – 3.04 USD/kgH₂, with smaller scales incurring higher costs.

In terms of component costs, the CL reforming reactor accounts for the largest cost, comprising for 45–60% of the total equipment cost. Other significant components include the PSA unit, the heat recovery system, and CO₂ compressors.

Operational costs are dominated by the expense of natural gas followed by labor, electricity, and CO₂ transportation. However, natural gas costs are highly dependent on global market prices. Similarly, CO₂ transportation costs may vary based on the availability of local CO₂ utilization options, storage facilities, or reservoirs. Limited storage availability or longer transportation distances can increase costs. Although the cost of the oxygen carrier is relatively minor, it remains subject to variability depending on operational practices. The expense of oxygen carrier replacement may rise if the material demonstrates low durability, especially since CL technology still needs extensive field validation.

Figure 2 presents a sensitivity analysis of blue H₂ pricing with respect to potential changes in the natural gas prices, oxygen carrier price/lifetime, and CO₂ transportation costs. The analysis reveals that natural gas and CO₂ transport costs have the most significant impact on overall H₂ price. While natural gas prices are dictated by global market fluctuations, CO₂ transport costs depend on the availability and readiness of CO₂ storage and transportation infrastructure. Notably, the upper bound of \$88.9/tonCO₂ applies to scenarios involving small-scale CO₂ transport over long distances (500 miles) with extra monitoring for storage – conditions that are unlikely to be developed [18]. Previous studies suggest a practical cost range of \$4 to \$45/tonCO₂ for the transport and storage of CO₂ in Tier-1 countries such as the U.S [18]. In contrast, the cost of the oxygen carrier has a relatively minor impact on H₂ price unless its lifetime is short, necessitating more frequent replacement.

Figure 3 compares the price of the proposed blue H₂-N₂ production system with that of other CL-based H₂ production systems studied in the literature. Our proposed small-scale system is generally more expensive than the other technologies with larger H₂ production capacities. This discrepancy can be attributed to the much smaller production scale of our system, operating at two to three orders of magnitude below those reported in other studies. Additionally, this techno-economic analysis includes the cost of CO₂ transportation, which is often excluded in other studies. Despite these differences, Figure 3 reveals a general trend, offering valuable insights into alternatives for blue H₂ production at smaller scales.

CONCLUSIONS

This paper presents a techno-economic analysis based on the dynamic model of a novel blue H₂-N₂ production system designed for small-scale applications. Utilizing the innovative CL principle, the estimated H₂ price is around 3.0 USD/kgH₂, which is more expensive than that of larger-scale plants. However, this small-scale approach reduces the need for long-distance hydrogen transportation, as required in centralized

scenarios. It is important to emphasize that this scenario depends heavily on the natural gas infrastructure, which has been under scrutiny due to its fugitive methane emissions. Furthermore, ensuring access to local CO₂ utilization options and storage reservoirs is critical to minimize CO₂ treatment costs, enhancing the system's economic viability.

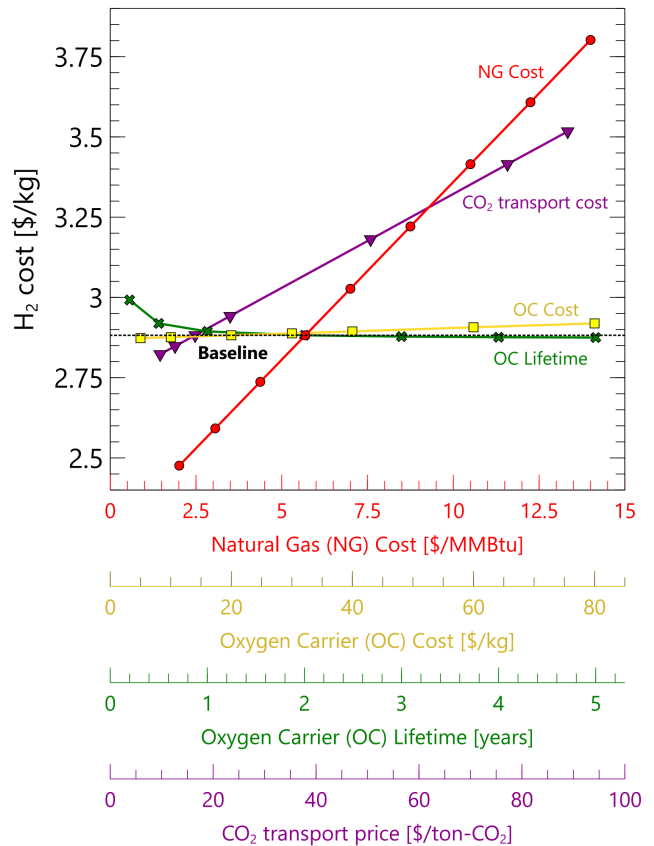


Figure 2. Sensitivity analysis of blue H₂ price relative to variations in natural gas prices, oxygen carrier (OC) price/lifetime, and CO₂ transport costs for the proposed system with 7.1 tonH₂ per day production rates.

Table 3: The key parameters for equipment price estimation. These values were obtained from the simulation of the proposed system at three different H₂ production rates: 2.8-, 7.1-, and 17.1-ton H₂ per day.

Equipment	Scaling parameter	H ₂ production rate		
		2.8 tpd	7.1 tpd	17.1 tpd
CL Reformer Reactor	Volume [m ³]	5.39	15.71	47.99
Shift Reactor	Shift gas flow [kg/s]	0.22	0.59	1.42
Heat recovery system	Heat duty per temp. [kW/K]	1.346	3.428	7.675
PSA	H ₂ stream flow [m ³ /s]	0.08	0.27	0.64
Desulphurizer	Methane feed [kg/s]	5.88	14.71	35.31
Methanator	CO ₂ in N ₂ -rich stream [g/s]	1.615	3.533	10.164
Air Blowers	Total air flow [kg/s]	0.47	1.18	2.83
CO ₂ compressors	Power [kW]	130	290	720
Heat rejection unit	Amount of heat [kW]	165.4	412.9	907

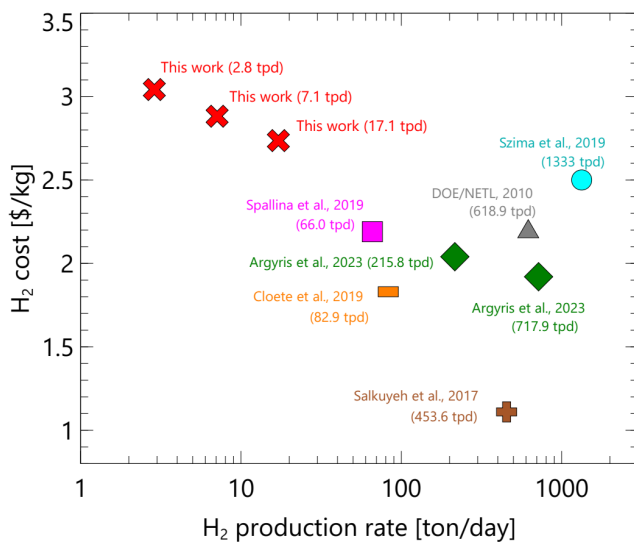


Figure 3. Comparison of H₂ prices for the proposed blue H₂-N₂ production system (red cross) with other blue H₂ technologies based on CL reforming principles from the literature [9–13], except DOE/NETL[16], which uses SMR-CCS.

REFERENCES

1. International Energy Agency (IEA), "Global hydrogen demand by sector in the Sustainable Development Scenario, 2019–2070," Paris, 2020.
2. "US National Clean Hydrogen Strategy and Roadmap," 2023.
3. Cummins, "Daimler Truck North America and Cummins collaborate to drive hydrogen fuel cell trucks forward in North America," May 2022.
4. A. Islam, T. Alam, N. Sheibley, K. Edmonson, D. Burns, and M. Hernandez, "Hydrogen blending in natural gas pipelines: A comprehensive review of material compatibility and safety considerations," *Int. J. Hydrogen Energy*, vol. 93, no. November, pp. 1429–1461, 2024, doi: 10.1016/j.ijhydene.2024.10.384.
5. Bloomberg NEF, "Hydrogen Economy Outlook," *Bloom. New Energy Financ.*, p. 12, 2020.
6. D. Schlissel and A. Juhn, "Blue Hydrogen: Not Clean, Not Low Carbon, Not a Solution," 2023. Accessed: Jun. 06, 2024.
7. A. R. Irhamna and G. M. Bolas, "Intensified reforming reactor for blue hydrogen and nitrogen production," *Int. J. Hydrogen Energy*, 2024, doi: 10.1016/j.ijhydene.2024.05.428.
8. Irhamna, A. R., & Bolas, G. M. Enhancing Hydrogen Infrastructure for Domestic Use: A Small-scale Blue Hydrogen and Nitrogen Production System (in preparation)
9. Y. Khojasteh Salkuyeh, B. A. Saville, and H. L. MacLean, "Techno-economic analysis and life cycle assessment of hydrogen production from natural gas using current and emerging technologies," *Int. J. Hydrogen Energy*, vol. 42, no. 30, pp. 18894–18909, 2017, doi: 10.1016/j.ijhydene.2017.05.219.
10. S. Cloete, M. N. Khan, and S. Amini, "Economic assessment of membrane-assisted autothermal reforming for cost effective hydrogen production with CO₂ capture," *Int. J. Hydrogen Energy*, vol. 44, no. 7, pp. 3492–3510, Feb. 2019, doi: 10.1016/j.ijhydene.2018.12.110.
11. V. Spallina, G. Motamedi, F. Gallucci, and M. van Sint Annaland, "Techno-economic assessment of an integrated high pressure chemical-looping process with packed-bed reactors in large scale hydrogen and methanol production," *Int. J. Greenh. Gas Control*, vol. 88, pp. 71–84, Sep. 2019, doi: 10.1016/j.ijggc.2019.05.026.
12. S. Szima et al., "Gas switching reforming for flexible power and hydrogen production to balance variable renewables," *Renew. Sustain. Energy Rev.*,

Table 4: The techno-economic analysis results for the proposed small-scale blue H₂ and N₂ production system at three different H₂ production rates: 2.8-, 7.1-, and 17.1-ton H₂ per day (tpd).

Cost Items	Price (in thousands USD)		
	2.8 tpd	7.1 tpd	17.1 tpd
<u>Capital Cost</u>			
Main equipment			
Reformer system	1993.2	5811.0	17752.8
Shift reactor	96.9	222.7	455.3
Heat recovery system	587.6	721.8	861.8
Pressure swing adsorption (PSA)	869.6	2912.5	6952.4
Desulphurizer	102.5	217.2	445.3
Methanator	18.4	34.4	80.1
Air Blowers	14.2	27.6	51.8
CO ₂ compressor	530.1	942.7	1799.5
Cooling plant/ heat rejection unit	185.3	342.0	579.7
Balance of plant: piping/instrument/control/facilities	1923.7	4336.7	8980.7
Total Plant Cost (TPC)	7328.3	18048.3	44005.4
Owner cost & other cost (O&OFC)	1465.7	3609.7	8801.1
Total Overnight Cost (TOC)	8794.0	21658.0	52806.4
Total as spent Cost (TASC)	10376.9	25556.4	62311.6
Capital charge rate factor (CCF)	0.159	0.159	0.159
TASC per year = CCF x TASC	1649.9	4063.5	9907.5
<u>Operating Cost</u>			
Fixed			
Labor	450	810	1350
Maintenance	11.5	28.4	69.4
Insurance	24.7	61.0	148.6
Variables			
Oxygen carrier replacement	15.6	31.9	67.1
Catalyst replacement	13.3	38.9	118.8
Electricity	218.8	486.8	862.9
Natural gas feedstock	660.0	1650.0	3959.9
CO ₂ transport	153.0	379.5	574.8
Total operating cost per year	1550.2	3495.3	7178.1
<u>Total annual cost</u>	3200.1	7558.8	17085.7
<u>H₂ production rate [tonH₂ per year]</u>	1052.0	2622.5	6244.1
<u>Cost of H₂ [USD/kgH₂]</u>	3.042	2.882	2.736
<u>Cost of CO₂ avoidance [USD/tonCO₂]</u>	149.8	137.3	125.9

- vol. 110, pp. 207–219, Aug. 2019, doi: 10.1016/j.rser.2019.03.061.
13. P. A. Argyris, J. Wong, A. Wright, L. M. C. Pereira, and V. Spallina, "Reducing the cost of low-carbon hydrogen production via emerging chemical looping process," *Energy Convers. Manag.*, vol. 277, Feb. 2023, doi: 10.1016/j.enconman.2022.116581.
 14. B. Heid, C. Martens, and M. Wilthaner, "Unlocking hydrogen's power for long-haul freight transport," 2022.
 15. Process Systems Enterprise, "gPROMS." 2024.
 16. US-DOE & NETL, "Assessment of hydrogen production with CO₂ capture volume 1: baseline state-of- the-art plants," US-Department Energy/National Energy Tech. Lab., vol. 1, 2011.
 17. G. Towler and R. A. Y. Sinnott, *Chemical Engineering Design: Principles, Practice and Economics of Plant and Process Design*. 2013.
 18. E. Smith, J. Morris, H. Khesgji, G. Teletzke, H. Herzog, and S. Paltsev, "The cost of CO₂ transport and storage in global integrated assessment modeling," *Int. J. Greenh. Gas Control*, vol. 109, p. 103367, 2021, doi: 10.1016/j.ijggc.2021.103367.

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