

Techno-economic Analysis for a Low Cost and Recyclable Oxygen Carrier

Junior NASAH^{1#}, Johannes VAN DER WATT¹, Srivats SRINIVASACHAR², Aaron KOENIG², Teagan NELSON², Mark MUSICH³

¹ University of North Dakota – Institute for Energy Studies, 2844 Campus Road, Grand Forks, ND, USA

² Envergenx, LLC, 10 Podunk Road, Sturbridge, MA, USA

³ University of North Dakota – Energy and Environmental Research Center, 15 N 23rd St, Grand Forks, ND, USA

#Corresponding Author

Abstract – The novelty of a chemical looping combustion (CLC) process is its ability to oxidize solid or gaseous fuels in a nitrogen-free environment eliminating the need for post combustion carbon dioxide capture. The technology is based on the oxygen carrier (OC); a metal oxide that at high temperatures can undergo a reduction reaction with a fuel, followed by an oxidation reaction with air. Successful development of CLC processes is dependent on the OC having: low cost, high attrition resistance under CLC conditions, and high reactivity for coal. In this paper, we discuss the development of a novel iron-based OC formulation produced from low cost and abundant raw materials. The active OC components are recyclable, with reformulated material performance comparable to the virgin OC, even when reformulated with ash contents of up to 10%. Furthermore, we discuss a techno-economic analysis for a centrally located OC facility for the novel OC, with a production capacity of 1 million metric tonnes per year. The facility serves a hypothetical market comprising 25 chemical looping combustion facilities, averaging 5 tonnes per hour OC replacement per facility. Estimates for the cost of the OC are around \$150/tonne, and include an energy-to-power option to improve process economics.

1 Introduction

Chemical Looping Combustion (CLC) is a promising oxy-combustion technology for producing a near-pure carbon dioxide (CO₂) stream from coal-based power generation. The majority of CLC technologies worldwide use the configuration of two interconnected bed reactors with an oxygen carrier (OC) that strips oxygen (O₂) from the air in one reactor (oxidizer) and delivers it to a fuel reactor (reducer) (1; 2; 3).

CLC development can be classified by OC type and reducer vessel design. OC development has focused on two applications, in-situ gasification (ig-CLC) and oxygen "uncoupled" (CLOU) (4). In the former, steam gasifies the fuel-producing carbon monoxide (CO) and steam (H₂O) which are then oxidized by the oxygen carrier. For CLOU processes, metal oxides are selected that release oxygen as a vapor under the operating conditions of the reducer, facilitating fuel combustion. The biggest contributor to the cost of OC is the raw material, which usually accounts for 50% of the total cost (5; 6), with CLOU-type OC development more expensive than ig-CLC OC due to more expensive raw materials (e.g. Copper) (5) and more expensive manufacturing techniques (5).

Reducer vessel design has focused on fluidized bed and moving bed designs. The commercial

maturity of fluidized beds favored their adoption by multiple researchers (5). However, the poor contacting between OC and gaseous fuels such as coal volatiles, make them best suited for CLOU processes (4). The moving bed design, pioneered by Ohio State University, is the most advanced of all CLC designs, with a 10 MW_e pre-FEED completed (6). The design ensures optimal solid-gas contacting between OC and fuel, making it ideal for ig-CLC.

A successful CLC development program requires the selection of the appropriate OC and a suitable reactor design to ensure optimal performance. Factors that will influence this decision include:

- Fuel conversion in the reducer which impacts the carbon capture rate and overall plant efficiency. This is characterized by the oxygen demand, Ω_{OD} , which is the oxygen (O₂) required for unconverted fuel species leaving with the reducer flue gas; and the char slip, CS, which is the unconverted solid fuel leaving the reducer to the oxidizer, and
- Oxygen carrier replacement costs which impact operating expenses for the facility. During the operation of a CLC facility, replacement of attrited or de-activated sorbent is required and estimated to be less than 1% of the coal feed rate (5; 6)

This paper presents the results of a class V AACE (7) techno-economic assessment (TEA) for a low-cost OC production facility serving a hypothetical CLC industry requiring approximately 1 million tonnes of OC per year.

2 Oxygen Carrier

2.1 Composition: The formulation of the novel OC consists of three major components: a CLC-active ingredient, Taconite, obtained from US Steel's Minnesota Taconite facility in Northern Minnesota, steel wastes from a steel facility in Ohio, provided by Carbontec Energy Corporation, and an alumina-based cement support serving as a "cold-bonding" agent to give initial strength to the constituents. In addition, two attrition inhibitors were included in the formulation to improve the high-temperature attrition resistance of the OC. The cement and attrition inhibitors were sourced from commodity suppliers to ensure sufficient supply for a commercial process. Table 1 summarizes the cost of the raw materials used for the oxygen carrier.

2.2 Preparation: The OC was prepared by a four-step mechanical mixing method consisting of: conditioning of raw materials to desired particle size (<100 μm), homogenization of raw materials, micro-pelletization of "green" pellets to desired particle size (300 μm – 2000 μm, a function of the CLC reactor design), and curing of the OC.

Table 1 Cost Estimate of OC Raw Materials

Consumable	Price (\$/tonne)
Taconite	\$100
Waste Steel	\$30
Cement	\$300
Attrition inhibitor 1	\$150
Attrition inhibitor 2	\$130

Table 2 Particle Size Distribution of the Manufactured OC.

Size (μm)	Weight (%)
400 – 250	36
250 – 180	36
180 – 150	14
150 - 110	14

2.3 Attrition Performance: The attrition performance of the manufactured OC was evaluated against an ilmenite provided by GE Power (formerly Alstom). Both jet and cyclonic attrition tests

were conducted using the apparatus described in Nelson *et al.*, 2019 (6). The procedure was as follows: A 31 g sample of the cured OC was prepared to the size distribution in Table 2. The attrition unit was heated to 900 °C prior to the addition of the OC sample. Redox cycling began after a 10-minute inert period with the bed under nitrogen only. Cycle conditions are given in Table 4 and Table 3 . A

Table 4 Cyclic Test Conditions in Jet Attrition Test System

	Reduction	Purge	Oxidation	Purge
Time (min)	4	2	4	2
Total Flow (lpm@70°F)	3, 4, 5	3, 4, 5	3, 4, 5	3, 4, 5
Jet vel. (m/s)	270, 360, 450	270, 360, 450	270, 360, 450	270, 360, 450
CO ₂ (vol%)	5	0	0	0
CO (vol%)	5	0	0	0
H ₂ (vol%)	5	0	0	0
O ₂ (vol%)	0	0	10	0

Table 3 Cyclic Test Conditions for Cyclonic Attrition Test System

	Ilmenite	OC
Spout Vel. (m/s)	10, 15, 20	
Temp. (°C)	900	900
Duration (min)	90 (high)	30 (low), 60 (med), 90 (high)

complete test sequence consisted of 15 cycles at 270 m/s, six cycles at 360 m/s, and six cycles at 450 m/s in order of increasing jet velocity (bed velocities calculated). Attrited material was collected on filters, which were replaced every two to three cycles, with new filters installed at the start of every reduction cycle. Filter loading was used to determine the attrition rate. At the completion of the test, the reactor was cooled under N₂.

Figure 1 shows the cumulative jet attrition performance for each operating velocity of the OC compared to ilmenite. The jet attrition method is an accelerated attrition method ideal for comparing performance between materials. The novel OC performed better than the ilmenite sample under identical jet attrition methods. Figure 2 shows cumulative cyclonic attrition performance of the OC vs. ilmenite, with attrition rates increasing as cyclonic impact velocities increase. Under cyclonic attrition conditions, the OC showed reduced attrition resistance compared to ilmenite as the velocities increased. These results agree with previous work (8) that showed under cyclonic attrition forces, comparative attrition rates could differ with particle-to-wall collisions (cyclonic) showing higher attrition tendencies than particle-to-particle attrition (jet).

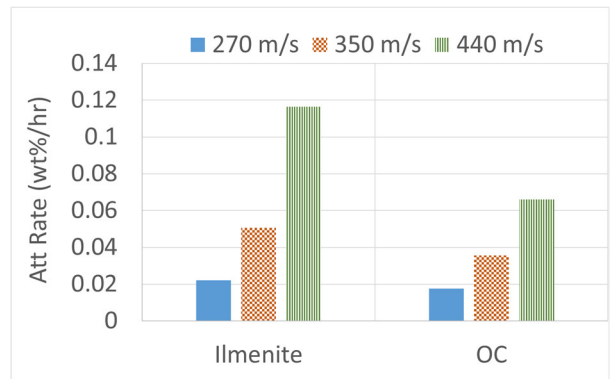


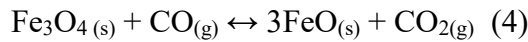
Figure 1 Jet attrition rate of manufactured OC vs ilmenite

2.3 Reactivity Performance: The OC reactivity was evaluated with a Thermogravimetric Analyzer - Differential Scanning Calorimetry (TGA-DSC) from TA Instruments. Reactivity evaluation was performed using CO / CO₂ as the reducing gas mixture with N₂ balance, see Table 5. Reactivity to H₂ / H₂O was not evaluated due to equipment limitations. Performance of the OC was determined by plotting the rate of oxygen (atomic) transfer (R_{O₂}) against the oxygen utilization (X) defined as:

$$R_{O_2} \left(\frac{mmol}{g \cdot min} \right) = \frac{1}{M_{t=0}} \cdot \frac{dN}{dt} \quad (1)$$

$$\frac{dN}{dt} \left(\frac{mmol}{min} \right) = \frac{(M_{t-1} - M_t) * 1000}{(t - t_{-1}) * 16} \quad (2)$$

$$X (\%) = \frac{M_{t=0} - M_t}{M_{t=0}} * 100 \quad (3)$$



Where M_t = mass of sample in TGA at time t (grams), N = Millimoles of oxygen (mmol), t = time (min).

Figure 3 summarizes the TGA reactivity results with the OC exhibiting 4 to 10 times faster reactivity than the ilmenite. The O₂ capacity of the OC was 2.4% and 4.8% for 4% and 10% CO respectively. At the 10% CO capacity, the rate of the reaction changes suggesting wüstite formation via reaction (4). This agrees with the composition analysis of the OC, which predicted a maximum O₂ capacity of 2.5% assuming hematite to magnetite reduction.

3 TEA for OC Production

3.1 Cost Estimate Scope: An AACE Class 5

estimate for process industries was adopted to develop the TEA for producing the OC (7). The design basis assumed an annual OC production rate at a capacity factor of 0.85, 990,000 tonne per year, or 133 tonne per hour. The chosen plant location was a 100-acre site near Gary/East Chicago,

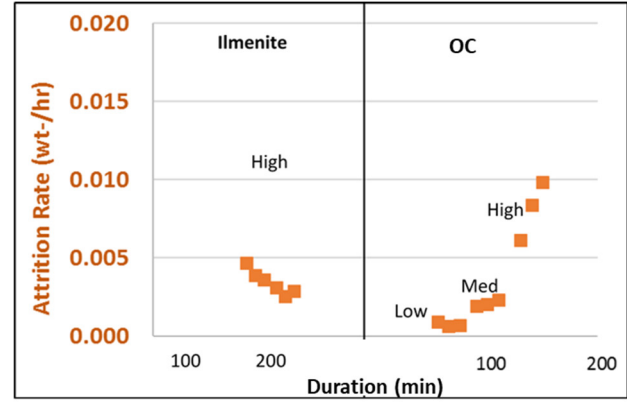


Figure 2 Cyclonic attrition of manufactured OC vs ilmenite

Table 5 Gas Composition during TGA-DSC of OC

Condition	Low	High
CO/CO ₂ ratio	0.33	0.50
% CO in Flow	4.3%	10.0%
(CO+CO ₂)/N ₂ ratio	20%	40%

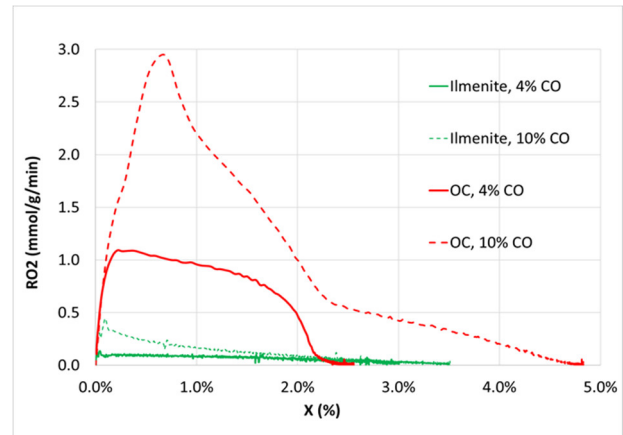


Figure 3 Reactivity of OC and Ilmenite during TGA testing

Indiana, a region of the Great Lakes. The site was chosen due to its proximity to suitable iron ore reserves.

3.2 Process Description: The facility consists of seven storage silos for the raw materials and OC products and six conveyors between storage and process units. Major process units were based on the manufacturing process; raw material homogenization using mixers and a primary mill, pelletization to desired particle size using an agglomerator, and screening of the agglomerated products using classifiers to produce green pellets later sent to the heat treatment (curing) step. Figure 4 is a block flow diagram of the process, and Table 6 is a heat and mass balance for streams entering each block in Figure 4. Oversize pellets are returned to the process via a secondary mill, while undersized pellets were returned to the mixer.

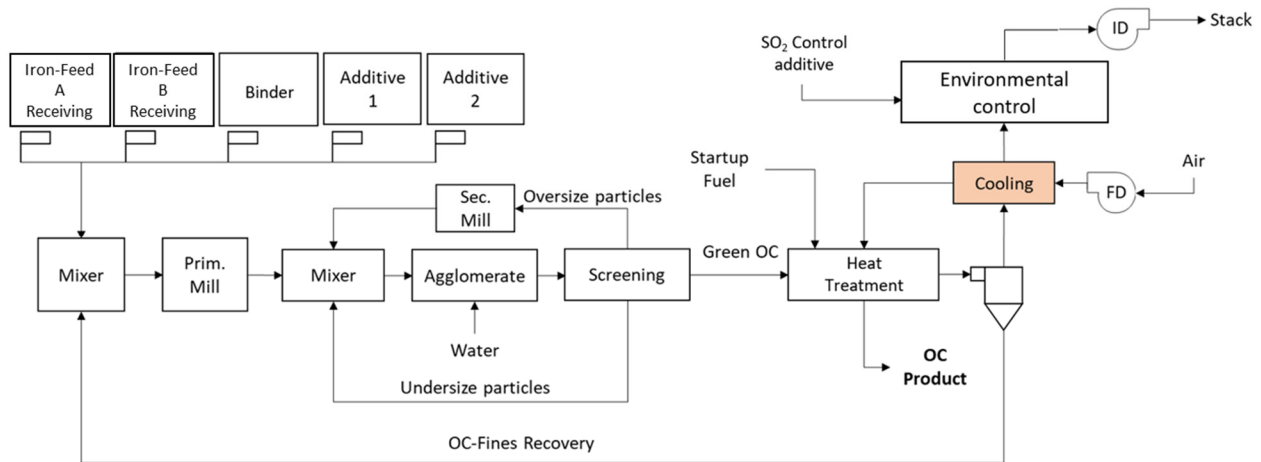


Figure 4 Block flow of OC manufacturing process

Table 6 Mass Balance of OC Facility

$T_o =$	Mixer	Mill 1	Mixer 2	Agglomerate	Mill 2	Heat Treatment	OC Pdt.	Cooling	Env. Ctrl
Mass Flow (kg/hr)	124,300	124,300	124,300	229,800	97,900	134,500	132,700	95,700	88,800
Temp. (°C)	20	89	89	77	67	67	900	900	156
Enthalpy Flow (Gcal/hr)*	-163	-163	-163	-325	-151	-202	-194	-27	-37
Fe _x O _y Content (wt.%)	0.40	0.40	0.40	0.39	0.37	0.37	0.76	0.06	<0.001
		Work (MW_e)		Water In (kg/hr)	Work (MW_e)	Q Out (MW_{th})			
		1.5		10,200	1.8	-35	-38	-10	

*Enthalpy flows from ASPEN Plus®

3.2.1 OC Curing: Curing is conducted in a low-velocity, bubbling fluid-bed to minimize elutriation and attrition of the OC. The calciner equipment adopted was the commercial scale Fluidized Bed Reactor (FBR) from SCHWING Technologies, which proposed a modular design based on their commercial-scale FBR array (9). An exothermic process, calcination proceeds

rapidly upon contact of the green OC with hot fluidizing air. The continuous stream of hot, calcined OC is discharged to a heat recovery system to reduce the OC temperature below 200°C prior to conveyor transport to silo storage. The calcined OC product consists primarily of Fe₂O₃. The slightly oxygen-depleted flue gas from the calciner contains OC dust, with the coarse fraction captured in a high-temperature (refractory-lined) cyclone. The cyclone underflow solid stream is returned to Mixer 1. The flue gas exits the cyclone and is sent to a heat recovery system.

3.2.2 Energy Recovery: During the calcination, the exothermic reaction produces an excess of 35 MW_{th} that needs to be removed to maintain operating temperatures. Two engineering approaches were considered to address this requirement. First, in bed cooling accomplished through internal water pipes. The excess heat would convert the water to steam that would then be sent to the Steam Turbine. Current commercial design of the FBR systems adds heat to the process, consequently, limited development work would be needed to modify the system to allow for heat extraction. The second option was in bed cooling accomplished through inert dilution. In this approach, cooled and cured OC from the heat recovery system is recirculated back to the calciner as an inert diluent for isothermal operation of the calciner. Heat recovery occurs post calcination. This approach will require that the capacity of the calciner and heat recovery system be increased to account for the recycled solids.

Similar to the calciner, two options were considered for cooling the product OC with energy recovery. In the first, hot OC from Calciner 1 is cooled in a solid-to-air heat exchanger, with the hot-air effluent discharged to a heat recovery steam generator (HRSG). In the second method, condensate in the turbine/condenser loop is fed directly to a solid-to-pressurized water heat exchanger, with the superheated water flashed in a steam drum. Steam from the HRSG or flash drum feeds a steam turbine. The resulting power is captively used by motors (conveyors, blowers, mixers, mills, pumps, and fans) to support electrical and mechanical needs. The first option was adopted due to the higher risk of tube failure for option two.

Finally, additional energy recovery is possible for the flue gas exiting the cyclone. An attemperation step drops the flue gas temperature sufficiently for use of a gas-to-gas heat exchanger where attemperated flue gas preheats fresh air for use in the calciner. This corresponds to an energy rejection of ~10 MW_{th}. Additional value engineering is required to minimize this heat loss. The flue gas leaving the heat exchanger is maintained at a temperature high enough to prevent moisture condensation in downstream processes.

A detailed engineering optimization was recommended for the energy recovery of the process, however, due to the conceptual nature of the TEA, this activity was considered beyond the scope of the current technology development profile. It was anticipated that if OC manufacturing were to proceed commercially, smaller facilities (~1000 – 10,000 tonnes/year) would be the primary focus eliminating the feasibility of significant energy recovery.

3.2.3 Environmental Controls: The flue gas leaving the gas-to-gas heat exchanger is routed to a conventional pulse-jet baghouse system exceeding 99.5% mass collection efficiency captures

particulate passing the High Temperature Cyclone. The baghouse solids stream is recycled to Mixer 1 in Green OC Preparation. The de-dusted flue gas is then processed in a packed-bed scrubber using slaked lime (Ca(OH)₂) slurry. The packed-bed scrubber discharge slurry is processed in a forced-air oxidizer that converts calcium sulfite to calcium sulfate. The slurry can be discharged to onsite ponds or dewatered for sale into the gypsum market. Cooling tower blowdown and scrubber liquid are sent to a zero-liquid discharge (ZLD) system.

3.3 Costing: Costing was performed by BARR Engineering Co.. The basis for the costing was a heat-mass balance of the OC facility developed using ASPEN Plus® and OEM quotes obtained for most pieces of equipment, combined with suitable scaling factors, prior experience, and Department of Energy (DOE) guidelines and studies (10; 11; 12). Process and project contingencies, Table 7, were applied following said guidelines (11) or based on the commercial maturity of the equipment for this application.

3.3.1 O&M Costs: Average burdened labor rate was assumed to be \$100/hr based on a 40 hr work week, with 30 personnel consisting of administrative, management, maintenance, and shift workers. Raw material costs (Table 1) were based on vendor quotes or USGS (13) estimates and included freight and delivery. Other costs were factored from (12) and are provided in Table 8. The cost of natural gas for process startup was estimated at \$3.3/MMBtu (14). Gypsum produced from the scrubber was a saleable by-product priced at \$35 per tonne. Table 9 is a summary of the annual O&M costs for the facility. With a heat recovery system that generates electricity, the facility saves \$3.6 million per year. However, the fuel usage was higher due to increased natural gas use to preheat the HRSG during startups.

Table 7 Process and Project Contingencies

Technology	Process	Project
Bulk Material Storage & Handling	10%	20%
Oxygen Carrier Processing	10%	20%
Heat Recovery System	10%	20%
Environmental Controls	20%	20%
Electrical Systems	10%	20%
Building & Facilities	10%	20%
Fluid Bed Calciner	20%	20%
High Temp Cyclone	20%	20%

Table 8 Cost of Consumables

Consumable	Price
Water (1000 gal)	\$5
Scrubber sorbent (tonne)	\$206
Scrubber solvent (gal)	\$0.004
Electricity (kWhr)	\$0.071
Scrubber Waste Disposal (gal)	\$0.006

Table 9 Operating and Maintenance Summary

	With Heat Recovery	Without Heat Recovery
Fixed Operating Costs, (\$)	17,786,000	15,736,000
Variable Operating Costs, (\$)	906,000	870,000
Consumables, (\$)	95,662,000	97,568,000
Waste Disposal, (\$)	705,000	705,000
Saleable By-Products, (\$)	-3,677,000	-76,000
Fuel Cost, (\$)	8,000	1,500
Total O&M, (\$)	111,390,000	114,805,000

3.3.2 Capital Costs: The costing was classified into six accounts with equipment costs based on OEM quotes or scaling factors (10). Table 10 summarizes the capital costs for the major process accounts. The difference between both options was the sum of the heat recovery system that consisted of the HRSG, steam turbine, and deaerator, which were quoted at \$2.9 million, \$4.2 million, and \$300,000, respectively.

Table 10 Total Capital Cost Summary of OC Manufacturing Facility

Category	BEC (\$)	EPC (\$)	Process Contingency (\$)	Project Contingency (\$)	TPC (\$)
Bulk Material Storage & Handling	2,984,000	448,000	344,000	755,000	4,531,000
Oxygen Carrier Processing	49,954,000	7,494,000	9,845,000	13,457,000	80,750,000
Heat Recovery System	35,892,000	5,383,000	4,129,000	9,082,000	54,486,000
Environmental Controls	27,944,000	4,192,000	6,427,000	7,712,000	46,275,000
Electrical Systems	21,628,000	3,245,000	2,487,000	5,472,000	32,832,000
Building & Facilities	19,914,000	2,988,000	2,290,000	5,038,000	30,230,000
Total w' Heat Recovery	158,316,000	23,750,000	25,522,000	41,516,000	249,104,000
Total w/o Heat Recovery	143,560,000	21,537,000	23,824,000	37,782,000	226,703,000

Table 11 Owners Cost

3.3.3 OC Sale Price: The OC sale price assumed a 30-year operation with an annual production of 989,000 tonne. The owner's cost (TOC) and total as-spent cost (TASC) were used to determine the OC sale price. The TOC was determined following guidelines from NETL (11), with final estimates shown in Table 12. The TASC was expressed in mixed-year, current dollars over the entire capital expenditure period. It was calculated from the total overnight cost (TOC), a finance

Description	Cost (with HR)	Cost (without HR)
Pre-Production Costs	\$21,592,000	\$20,225,000
Inventory Capital (0.5% TPC, 60-day consumables)	\$17,057,300	\$17,249,000
Land (\$1,000/acre)	\$300,000	\$300,000
Financing Cost (2.7% TPC)	\$7,244,000	\$6,592,000
Other Costs (15% TPC)	\$40,243,000	\$36,624,000
Total Overnight Costs (TOC)	\$354,721,000	\$294,789,000

Table 12 Annual Cost of Production and OC Sale Price

	With HR	Without HR
Total As Spent Cost (TASC), (\$)	440,563,000	366,128,000
First Year Capital Charge, (\$)	39,034,000	32,439,000
First Year Fixed Operating, (\$)	17,786,000	15,736,000
1 st Year Variable Operating, (\$)	93,596,000	99,067,000
Total Annual Cost, (\$)	150,424,000	147,244,000
OC Sale Price (\$/tonne)	152	149

structure assuming a BBB+ or higher company, to obtain a TASC/TOC nominal ratio of 1.242 for a three-year capital expenditure period, and a fixed rate charge of 0.0886 (11). Table 12 illustrates a final summary of the TASC and OC sale price of \$149 to \$152 per tonne without and with heat recovery, respectively.

4 Results Discussion

A sensitivity analysis of the process is summarized in Figure 5. The biggest impact on the cost of OC was the operating costs (fixed and variable). For the base case, annual operating costs are 27% of the TASC. Doubling or tripling these costs has the highest impact on OC sale price. Of these operating costs, the taconite (Iron-Feed A) and the binder are the biggest contributors as shown in Figure 5. Other factors that impact the price of OC is the fixed rate charge (FRC) and capital costs.

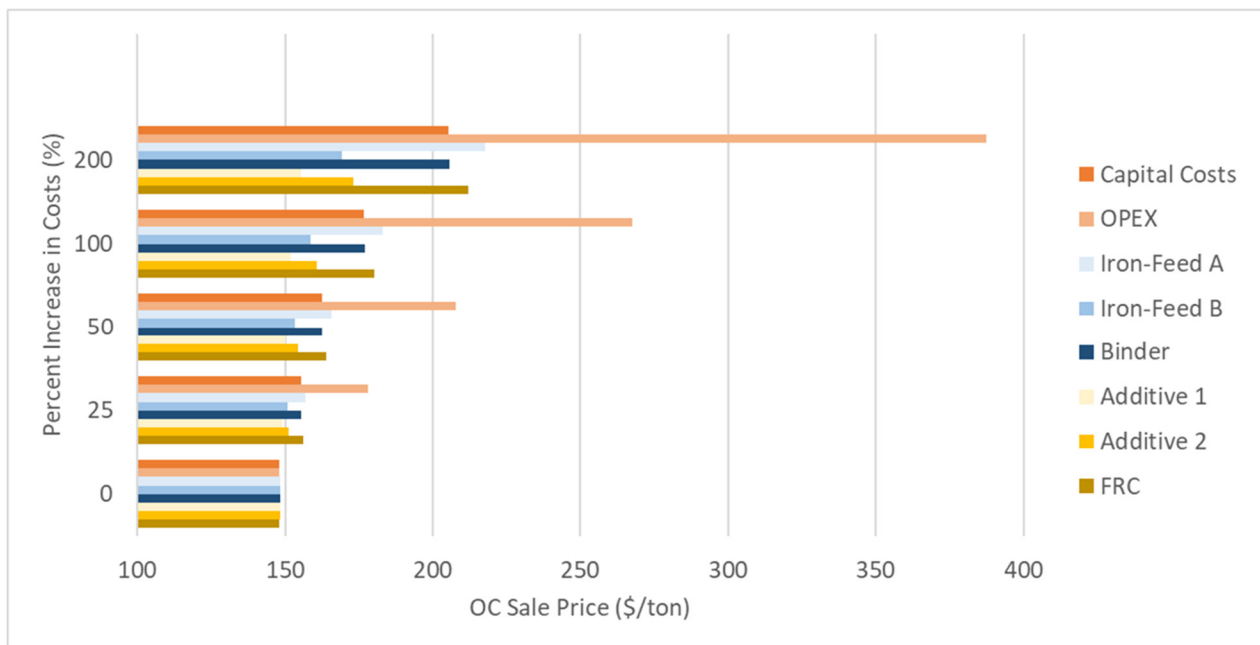


Figure 5 Sensitivity analysis of CAPEX, OPEX, raw materials, and financing on OC sale price

Stevens (5) performed a TEA on a similar iron-based OCs (taconite-alumina, mechanically mixed) and obtained costs of \$730 - \$970/tonne (adjusted from \$2011 to \$2021), Table 13. The biggest cost differences between our OC and the iron-based OC from Stevens (5) were the fixed costs (labor) and raw material costs. The costs for the main CLC ingredient (taconite and steel wastes) were similar, however, the final raw material cost was significantly higher than ours due to our use of low cost,

Table 13 TEA comparison with Stevens (5)

	Stevens (\$2021)	This Study (\$2021)
Taconite (\$/tonne)	73 - 245	65
Raw Material Cost (\$/tonne)	296 - 511	87
Fixed Costs (\$/tonne)	188 - 188	16
Variable Cost (\$/tonne)	81 - 81	12
Capital Recovery (\$/tonne)	161 - 188	33
OC Price (\$/tonne)	730 - 970	149

readily-available additives that kept the total raw material costs low. Fixed costs were an order of magnitude different, which was attributed to economies of scale; this study looked at 989,000 tonnes of OC a year versus 9,000 tonnes per year (factor of a 100) for Stevens (5), meanwhile, fixed costs are \$15 million vs. \$1.4 million (factor of 10) respectively. Economies of scale also explain the differences in other categories. If we scaled down our process to 9,000 tonnes/year by assuming similar costs with Stevens (5), for all factors affected by economies of scale (capital recovery, variable and fixed costs) except the raw materials, our OC sale price is approximately \$530 per tonne OC due to the lower costs for our raw materials.

5 Conclusion

A novel, low-cost method for producing an iron-based OC using a mechanical mixing method is presented. The method showed good attrition performance under the jet and cyclonic conditions, with reactivity up to eight times faster than the ilmenite benchmark. A Class V TEA was performed for a facility producing approximately 1 million tonnes of OC per year with an OC sale price of \$150 per tonne. Comparison with similar facilities suggest the proposed OC is cheaper due to the lower cost additives used in the manufacturing process.

6 Acronyms

BEC Bare Erected Cost	MW _{th} Megawatt thermal
CAPEX Capital Expenditures	N Millimoles of oxygen
CLC Chemical Looping Combustion	OC Oxygen Carrier
CLOU Chemical Looping Oxygen Uncoupled	OPEX Operating Expenditures
CS Char Slip	O ₂ Oxygen
CO Carbon monoxide	O&M Operating and Maintenance
CO ₂ Carbon dioxide	R _{O2} Rate of atomic oxygen transfer
EPC Engineering, Procurement and Construction	TASC Total As Spent Cost
FBR Fluidized Bed Reactor	TEA Techno-Economic Assessment
FEED Front-End Engineering Design	t time (min)
FRC Fixed Rate Charge	TPC Total Plant Cost
HRSG Heat Recovery Steam Generator	X Percent reduction of OC
ig Integrated Gasification	µm Micrometer
M _t mass of sample in TGA at time t	Ω _{OD} Oxygen demand
MW _e Megawatt electric	

7 Acknowledgements

This material is based upon work supported by the United States Department of Energy under award number DE-FE0031534.

Disclaimer: “This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.”

8 References

1. *Iron-based syngas chemical looping process and coal-direct chemical looping process development at Ohio State University*. Tong, Andrew, et al. 2014, *Applied Energy*, Vol. 113, pp. 1836-1845.
2. *Chemical-Looping Combustion of Solid Fuels - Status and recent progress*. Linderholm, Carl and Lyngfelt, Anders. 2017, *Energy Procedia*, Vol. 114, pp. 371-386.
3. *Parametric and dynamic studies on an iron-based 25-kWth coal direct chemical looping unit using sub-bituminous coal*. Bayham, Samuel, et al. May 1, 2015, *Applied Energy*, Vol. 145, pp. 354-363.
4. *Chemical looping combustion of solid fuels*. Adánez, J., et al. 65, s.l. : *Progress in Energy and Combustion Science*, 2018, pp. 6-66.
5. Stevens, Robert W, et al. *Oxygen Carrier Production Cost*. United States : s.n., 2019.
6. Velazquez-Vargas, Luis G., et al. *10 MWE CDCL Large Pilot Plant – pre-FEED Study*. United States : <https://doi.org/10.2172/163825>.
7. AACE International. 18R-97: Cost Estimate Classification System - As Applied in Engineering, Procurement, and Construction for the Process Industries. March 6, 2019.
8. *Reactive jet and cyclonic attrition analysis of ilmenite in chemical looping combustion systems*. Nelson, T., et al. 91, 2019, *International Journal of Greenhouse Gas Control*, p. 102837.
9. Fluidized Bed Powder Processing Systems. *SCHWING TECHNOLOGIES*. [Online] June 11, 2022. <https://schwing.tech/powder-processing>.

10. Zoelle, Alexander and Kuehn, Norma. *Quality Guidelines for Energy System Studies: Capital Cost Scaling Methodology: Revision 4 Report*. United States : s.n., 2019.
11. Theis, Joel. *Quality Guidelines for Energy Systems Studies: Cost Estimation Methodology for NETL Assessments of Power Plant Performance*. United States : s.n., 2019.
12. James, III, Robert E., et al. *Cost and Performance Baseline for Fossil Energy Plants Volume I: Bituminous Coal and Natural Gas to Electricity*. United States : s.n.
13. U.S. Geological Survey. *Mineral Commodities Summaries 2021: U.S. Geological Survey*. 2021.
14. U.S. Energy Information Administration. Henry Hub Natural Gas Spot Price (Dollars per Million BTU). *Henry Hub Natural Gas Spot Price (Dollars per Million BTU)*. [Online] October 10, 2021. eia.gov.