LOW-COST RECYCLABLE OXYGEN CARRIER AND NOVEL PROCESS FOR CHEMICAL LOOPING COMBUSTION

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University of North Dakota

Institute for Energy Studies

2844 Campus Rd Stop 8153

Grand Forks, Nd 58202

SUBCONTRACTORS

Envergex LLC

10 Podunk Road

Sturbridge, MA

Barr Engineering Company

4700 West 77th St., #200

Minneapolis, MN 55435

Microbeam Technologies Inc.

4200 James Ray Drive, Ste. 193

Grand Forks, ND 58203

PRINCIPAL INVESTIGATOR

Junior Nasah

Phone: 701-777-4307

nasah.domkam@und.edu

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Final Technical Report

Low-Cost Recyclable Oxygen Carrier and Novel Process for Chemical Looping Combustion

Author List:

University of North Dakota
Junior Nasah, Johannes Van der Watt, Michael Mann, Mark Musich

Envergex LLC
Srivats Srinivasachar, Aaron Koenig, Teagan Nelson

Microbeam Technologies, Inc.
Alex Benson, Matt Fuka, Steve Benson

Barr Engineering Co.

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Acronyms

AACE	Association for the Advancement of Cost Engineering
ATM	Atmosphere
CLC	Chemical Looping Combustion
CLOU	Chemical Looping Oxygen Uncoupled
CO	Carbon monoxide
CO_2	Carbon dioxide
$d_{p,avg}$	Average mean particle diameter
Fe_2O_3	Hematite (Iron III oxide)
Fe_3O_4	Magnetite (Iron II oxide)
FeO	Wustite (Iron I oxide)
Fe	Iron
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen sulfide
HCl	hydrochloric gas
Hg	Mercury
HHV	Higher Heating Value
HRSG	Heat Recovery Steam Generator
HX	Heat Exchanger
H_2O	Water
H_2S	hydrogen sulfide
ig-CLC	Integrated gasification Chemical Looping Combustion
kmol/hr	Kilomoles per hour
kW	Kilowatt
kWh	Kilowatt-Hour(s)
MMTPY	Million metric tonnes per year
MW	Megawatt
MW_e	Megawatt electric
MW_{th}	Megawatt thermal
m/s	Meters per second
N_2	Nitrogen
NaCl	Sodium chloride
NETL	National Energy Technology Laboratory
NO_X	Nitrogen Oxide
O_2	Oxygen
OC	Oxygen Carrier
PSD	Particle size distribution
QGESS	Quality Guidelines for Energy Systems Studies
SFB	Spout Fluid Bed
S	Sulfur
SO_2	Sulfur Dioxide
TEA	Techno-Economic Analysis
TGA	Thermogravimetric analyzer
μm	Micron
UND-IES	University of North Dakota Institute for Energy Studies
UNDEERC	University of North Dakota Energy and Environmental Research Center
Vol%	Volume percent
Wt%/hr.	weight percent per hour
%/hr	Percent per hour
°C	Degree celsius

1 Executive Summary

The novelty of a CLC process lies in the ability to oxidize a solid or gaseous fuel in a nitrogen-free environment to produce a near-pure carbon dioxide stream. This is made possible thanks to an OC, a metal oxide that at high temperatures can be oxidized in a fuel-free vessel (oxidizer). The oxidized metal oxide is then reduced by reacting with the fuel in a nitrogen-free vessel (reducer) to produce the near-pure stream of carbon dioxide (CO_2). The OC "loops" between the oxidation and reduction vessel.

In the reducer, the coal is converted primarily by steam gasification to carbon monoxide (CO) and hydrogen (H_2) followed by oxidation of the gasification products to CO_2 and water. The near-pure CO_2 stream from the reducer then goes through a series of cleanup steps prior to compression, eliminating the need for expensive post-combustion gas separation. Oxidation of the oxygen carrier leaving the reducer is exothermic, and provides the bulk of the steam evaporation duty required for the steam turbine. The steam superheat and reheat duties are provided by the hot exhaust gases leaving the oxidizer and reducer and the hot OC leaving the oxidizer to the reducer.

The University of North Dakota, through its Institute for Energy Studies and Energy & Environmental Research Center, partnered with Envergex LLC, Barr Engineering and Microbeam Technologies to develop a transformational enabling technology for advancement of Chemical Looping Combustion technology. Industrial support was provided by Carbontec Energy Corporation. The project targeted the two biggest challenges to chemical looping combustion:

- High costs of oxygen carrier replacement/loss due to expensive manufacturing and high replacement rates from physical attrition and/or decrease in reactivity.
- Inherently slow fuel char conversion which represents the rate-limiting step for chemical looping combustion and results in very large equipment sizes, and overall lower carbon dioxide capture efficiency.

The project activities were addressed in a series of seven tasks. Task 1 extended throughout the entire project and oversaw project management and execution. Task 2 and 3 focused on development and evaluation of the novel oxygen carrier. Task 4 to 7 focused on design of the novel reactor, testing with the novel oxygen carrier and a techno-economic assessment of the process. The list of tasks are:

Task 1 – Project management and planning

Task 2 – Laboratory scale oxygen carrier manufacturing and assessment

Task 3 – Modeling and laboratory-scale evaluation of oxygen carrier performance with coal

Task 4 – 10-kilowatt thermal integrated system installation

Task 5 – Scaled-up oxygen carrier manufacturing

Task 6 – 10-kilowatt testing

Task 7 – Process design and techno-economic analysis

Summary of Novel OC Production and Testing

The project team proposed a novel OC process that includes a low-cost OC and a reducer design based on a spout fluid bed. The OC was prepared by a mechanical mixing process using low-cost raw materials, and exhibits good attrition resistance under the CLC stresses. The proposed reducer design concept is a spout fluid bed (SFB) operated at low bed velocities to improve gas-solid contacting between OC and fuel. Key conclusions/achievements from activities focused on developing and testing a novel oxygen carrier include:

- A novel formulation method for producing high strength, low cost oxygen carriers was developed by sub-awardee Envergex LLC after evaluating over forty different formulations. Of these, two formulations of OC were produced – FEH31 and FEL3 and subjected to additional testing including jet and cyclonic attrition evaluation, fluidized bed testing, reactivity evaluation and ash/OC interaction behavior.

- The down-selected formulations were estimated to have comparable attrition performance when compared to an ilmenite baseline sample at low and medium kinetic power evaluations. Cyclonic attrition performance suggested attrition rates were less than 0.02wt.% at cyclonic impact velocities of 20 m/s.
- Fluid bed testing, thermogravimetric analysis, FactSage modelling and composition analysis were used to establish a maximum oxygen capacity for chemical looping reactions of 2 wt.% for both manufactured oxygen carriers. A recommended operating target of 1 wt.% was selected to ensure optimal heat balance between oxidizer and reducer.
- Cycled oxygen carrier tested in the fluidized bed was successfully recycled. The bed material containing coal ash was crushed and reformulated. Attrition testing confirmed similar performance with original material
- A minimum recommended OC-to-coal ratio was determined for both engineered oxygen carriers at desired bed hydrodynamics of a spout-fluid bed reactor
- Operating conditions for a 10 kW_{th} continuous system were determined from batch testing on a high temperature system and a cold flow system
- OC / ash interactions under accelerated conditions confirmed the onset of sintering and/or agglomeration can be expected after extended interaction under cyclic and non-cyclic conditions.

Summary of a TEA for a 1 MMTPY OC Facility

This report includes an assessment of the cost to construct a 1 million tonne per year oxygen carrier production facility based on a low-cost and novel process developed by Envergex LLC. A heat and mass balance of the process was developed by UND with input from Envergex LLC using ASPEN Plus®. The team developed a heat mass balance that was combined with a Mine Cost Estimate Software, and vendor quotations as references for developing the costs. The total project cost estimate, divided into 6 different code-of-accounts, corresponds to a Class 5 estimate class (AACE International Recommended Practice No. 18R-97) and the range of accuracy of -50- +100% accuracy.

Table 1-1 is a summary of the key findings for the facility including total project costs and projected sale price of the oxygen carrier.

Table 1-1 Cost Results Summary

Description for OC (Base Case)	With Heat Recovery	Without Heat Recovery
Total Project Cost ($)	268,285,000	244,159,000
Total Overnight Cost ($)	354,721,000	294,789,000
Total As Spent Cost ($)	440,563,000	366,128,000
Total Annual O&M	111,390,000	114,805,000
Total Annual Costs ($)	150,424,000	147,244,000
OC Sale Price ($/tonne)	152	149

Through sensitivity analysis, the sale price of OC was most greatly affected by the capital, O&M costs and the Fixed Rate Charge (FRC). The cost of iron-feed A (stream 1) and iron-feed B (stream 2) were the major O&M cost components. Figure 1-1 summarizes the sensitivity analysis. The oxygen carrier sale price was adopted for the second TEA – a 585 MW_e CLC power plant.

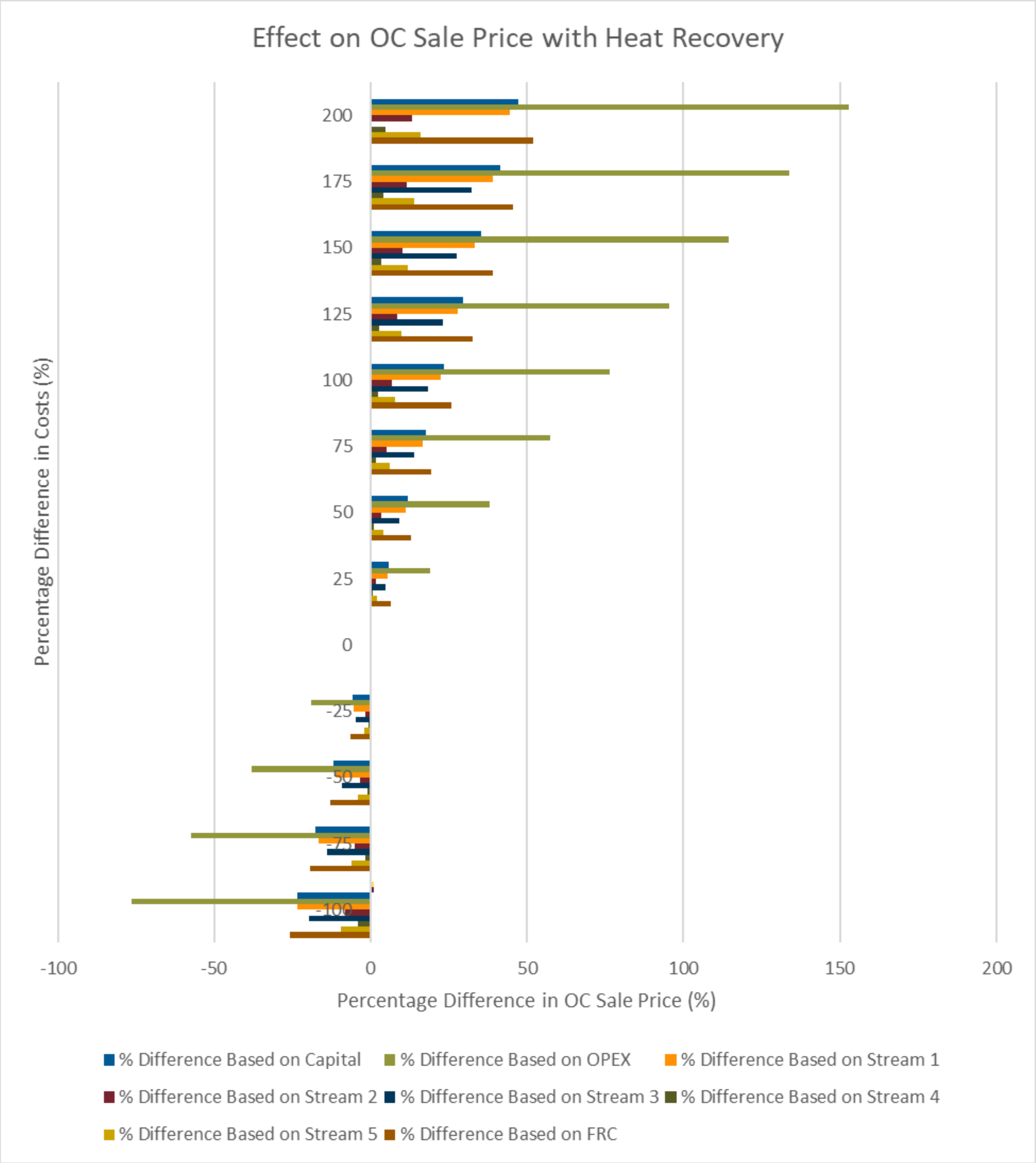

Figure 1-1 Sensitivity Analysis of 1 MMTPY OC Production Facility

Summary of a TEA for a 585 MW_e CLC Power Plant

This report includes an assessment of the cost to construct a 585 MW_e chemical looping combustion (CLC) power plant using a novel, low cost oxygen carrier (OC). A heat and mass balance of the facility was developed by UND using ASPEN Plus® with key process conditions determined from bench scale testing at UND's Institute for Energy Studies.

Table 1-2 summarizes key specifications, operating conditions and design configuration for the chemical looping process.

Table 1-2 Key Specifications and Operating Conditions of Proposed CLC Facility

CLC Reactor Design Specifications		
	Reducer	Oxidizer
Reactor Type	Spout Fluid Bed	Circulating Fluid Bed
Temperature (°C)	951	960
Pressure Drop (atm)	1.7	0.1
Solids Flow (1000 kg/hr)	24,500	25,000
Attrited OC (kg/hr)	1620	3390
Recycled OC (kg/hr)	2670	
OC Specifications		
Particle Size (µm)	600 - 800	
Cost ($/tonne)	150	

Table 1-3 Cost Results Summary

Description for CLC (Base Case)	
Plant Gross Capacity (MW_e)	685
Plant Net Capacity (MW_e)	585
Coal Flow Rate (kg/hr)	234,131
HHV Thermal Input	1,763,332
Plant Efficiency (%, HHV)	33.2
Plant Heat Rate (kJ/kWh)	10,851
CO_2 Capture Efficiency (%)	97.3%
CO_2 Purity	96.6%
Total Project Cost ($)	2,012,886,000
Total Overnight Cost ($)	2,459,765,000
Total As Spent Cost ($)	3,055,028,000
Total Annual O&M	182,964,000
Total Annual Costs ($)	453,639,000
Electric Sale Price ($/kWh)	0.104

The total project cost estimate, divided into 14 different code of accounts, corresponds to a Class 5 estimate class (AACE International Recommended Practice No. 18R-97) for the process industries and the range of accuracy of -50-+100% accuracy. Table 1-4 summarizes the total plant cost for different code of accounts. The Total Overnight Costs were found by approximating the capital costs, O&M costs, and owners costs. Based on a finance model from NETL (National Energy Technology Laboratory), the Total As-Spent Costs and Total Annual Costs were determined. The cost of electricity (based on annual production) was determined to be approximately $0.104/kWh.

Table 1-4 Total Plant Cost Breakdown by Code of Accounts

Item	Category	Total Project Cost ($)	$/kW
1	Coal Handling, Prep & Feed Systems	96,599,000	141
2	Coal Prep & Feed Systems	27,304,000	40
3	Feedwater & Miscellaneous	203,395,000	297
4	Chemical Looping Combustion	468,849,000	684
5	Gas Processing Unit	506,807,000	740
6	Green OC Production	1,994,000	3
7	HRSG, Ducting & Stack	24,225,000	35
8	Steam Turbine Generator	251,088,000	367
9	Cooling Water System	102,627,000	150
10	Ash & Carrier Waste Handling System	23,366,000	34
11	Accessory Electric Plant	87,272,000	127
12	Instrumentation & Control	30,922,000	45
13	Improvements to Site	90,294,000	132
14	Building & Facilities	98,144,000	143
	Total ($2021)	**2,012,886,000**	**2,939**

The key challenges facing CLC processes that this project sought to address include sorbent attrition resistance, sorbent costs, extent of coal conversion in reducer and purity of CO_2 produced from reducer. These create significant uncertainty in CLC process performance and impact the capital and operating expenses. By adopting a low cost and recyclable OC, the impact of OC price on electricity price is reduced significantly. The replacement costs of the OC are determined by the attrition rate, amount of attrited OC recycled and OC price. In our process, we assumed 53 percent recycle and an attrition rate of 0.02 percent of solid circulation. In this scenario, the impact of OC cost is negligible; a 100 percent increase in OC price results in less than one percent increase in the electric sale price of $0.104/kWh. The low cost and recyclable process significantly reduces the impact of OC price on the process.

A detailed and complete evaluation of the impact of the reducer design, which is a first-of-a-kind piece of equipment, was beyond the scope of the project. Estimated costs and performance were scaled based on much smaller commercially-available equipment and bench-scale data respectively. A high-level sensitivity of the impact of capital expenditures and adoption of an air separation unit were used to evaluate the impact of the reducer on the plant. The cost of the chemical looping unit including the reducer and oxidizer was estimated to be $685/kW; which is 23% of total project cost. If the capital cost of the CLC system doubles, the TPC increases by 25% and the cost of electricity is $0.119/kWh.

The results of this TEA confirm that CLC is a promising carbon dioxide removal technology if the technical challenges involved can be resolved. With key process features such as high OC circulation rates and the interconnected fluidized beds already considered significant technical challenges, it is important to reduce the impact of all other factors on the overall technology risk profile, specifically OC replacement costs and coal conversion performance.

2 Introduction

Chemical looping combustion (CLC) is a promising oxy-combustion technology at various stages of development (bench and pilot testing). The process adopts an oxygen carrier (OC) which consists of a metal oxide that provides inherent air separation by picking up oxygen from air (oxidation step) and then transferring the oxygen to a fuel (reduction step). The process occurs at high temperatures making it easier to minimize the energy not available to do work[1].

The oxidation-reduction (redox) reaction is summarized in Figure 2-1, which represents an ideal description of the process. The fuel oxidation step usually is endothermic and the hot OC acts as the heat source. The fuel is oxidized in the fuel reactor where carbon dioxide (CO_2) and steam (H_2O) are the primary products. A condenser is then used to obtain a pure CO_2 stream. An optional char separator is required if there are significant losses of unconverted char to the oxidizer. The reduced OC is then returned to the air reactor where it is oxidized by reacting with air in an exothermic step that provides the bulk of the heat duty for power production and for fuel conversion. Other heat sources consist of the hot gases leaving both reactors which are in thermal equilibrium with the OC.

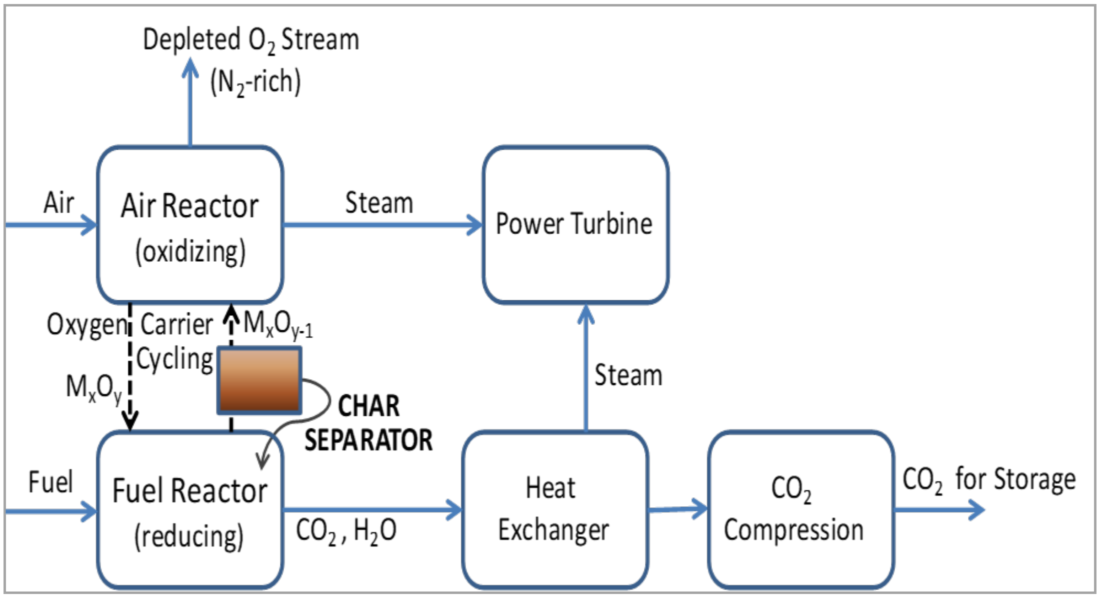

Figure 2-1 Simplified schematic of a chemical looping process

For coal conversion in a CLC process, two approaches are currently being considered – in-situ gasification (ig-CLC) and chemical looping oxygen uncoupled (CLOU)[2]. In ig-CLC, the solid fuel is gasified using steam and CO_2 to form hydrogen (H_2) and carbon monoxide (CO), which are then oxidized by the OC to water and CO_2. In CLOU, a special OC is selected that thermally decomposes under reducer conditions to release oxygen for direct combustion of the fuel. In CLOU processes, the reducer is exothermic.

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[1] McGlashan, N. R, 2008 Chemical-looping combustion – a thermodynamic study. *Proceedings of the Institution of Mechanical Engineers, Part C: Journal of Mechanical Engineering Science* 222 (6), pp. 1005-1019

[2] Adánez, J., Abad, A., Mendiara, T., Gayán, P., De Diego, L.F. and García-Labiano, F., 2018. Chemical looping combustion of solid fuels. *Progress in Energy and Combustion Science*, 65, pp.6-66.

3 Technical Project Summary

Task 1 – Project Management and Planning

A summary of major project management activities is presented below:

- An updated project management plan was provided on February 21, 2018.
- The project kick-off meeting was held on February 13, 2018
- Quarterly reports were provided on schedule during all budget periods.
- Monthly update meetings were scheduled with the federal project manager.
- All project deliverables were submitted, with two deliverables submitted after project end date:
  - 12/30/2019 – Report titled: “Equipment Design Package for 10 kW CLC Unit, Rev 0” submitted
  - 07/24/2020 – Report titled: “Oxygen Carrier Characterization and Testing Summary Report, Rev 0” submitted
  - 12/31/2021 – Report titled: “Techno-Economic analysis: Oxygen Carrier Production Plant” submitted
  - 01/14/2022 – Report titled: “Techno-Economic Analysis of a 585 MWe Chemical Looping Combustion Power Plant” submitted.
- The group presented its research findings virtually at the 2020 and 2021 NETL spring research meeting and in-person at the 2019 meeting.
- Technical presentations including final project presentation to NETL project managers was performed on 10/21/2019, 10/26/2020 and 11/30/2021.

Task 2 – Laboratory Scale Oxygen Carrier Manufacturing and Assessment

Activities in Task 2 were completed under three sub tasks. In subtask 2.1, over 40 oxygen carrier (OC) formulations were prepared and screened using a two-step process focused on initial stability. 19 of the down-selected OC formulations were then subjected to further screening. In subtask 2.2, the 19 down-selected formulations were then evaluated for attrition resistance and reactivity with two formulations down-selected for further testing. In sub-task 2.3 the down-selected formulations were subjected to comparative longer-term jet attrition testing.

Subtask 2.1 Laboratory-scale OC Manufacturing

The OC manufacturing focused on three main components for the composition: a chemical looping combustion active component, a support material and an attrition inhibitor. The sourcing of the components focused on commercial availability; only commodity materials were selected for manufacturing.

In addition to composition, the manufacturing process was sub-divided into three main groups – 1) High iron content (FEH) focused on naturally occurring iron ores, 2) Low iron content (FEL) focused on blends of naturally occurring ores and waste sources of iron, and 3) a non-iron OC group (FEC), focused on chemical looping oxygen uncoupled. Table 3-1 summarizes the number of samples down-selected for further testing in subtask 2-2.

Table 3-1 Number of samples down-selected for further testing

High Iron Content (FEH)	11
Low Iron Content (FEL)	3
Oxygen Uncoupled (FEC)	5

Prepared samples were homogenized using a ball mill and then micro-pelletized using a pan-mixer set-up shown in Figure 3-1. An initial screening consisting of a simple "thumb-crush" test was performed to see if the prepared material had acceptable strength for the next processing step. Next the samples were cured to activate the attrition inhibitor. The final screening step focused on evaluating attrition performance and reactivity. Figure 3-2 is an image showing the non-heat treated and heat-treated samples.

Figure 3-1 Image of Ball Mill Unit (left) and Pan Mixer (Right) for homogenizing and pelletizing sample

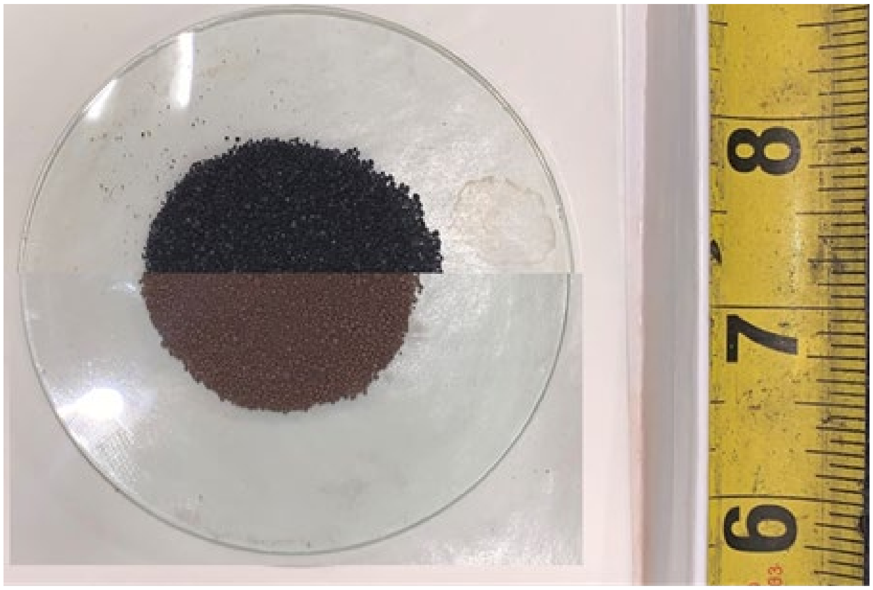

Figure 3-2 Image of pelletized sample showing heat treated (bottom) and non-treated sample

Subtask 2.2 OC Characterization & Laboratory-Scale Performance Testing

Nineteen samples passed the first two screening steps. These samples were further characterized using a jet attrition system for a further down-select to three formulations. In the jet attrition unit, the forces experienced by the OC are much greater than those expected in a commercial application due to the high jet velocities, implying "accelerated" attrition conditions. Therefore, the attrition rates determined are best used for a comparative assessment of different carriers and not for predicting actual attrition rates. Correlating jet attrition performance with real-world scenarios will require an appropriate scaling factor based on an actual CLC process. An alternate method, cyclonic attrition, was used to evaluate the down-selected formulations performance. Both attrition methods are further described below.

Methodology – Jet Attrition Evaluation

The 19 samples obtained from subtask 2.1 were sieved and crushed to obtain the required size distribution given in Table 3-2, as established by Nelson et al (2019)[3]. It is a fluidized bed reactor fitted with electric heaters and mass flow controllers supplying gases necessary to simulate the reduction and oxidation reactions during chemical looping. The distributor plate ensures high jet velocities are reached to create accelerated attrition conditions. Figure 3-3 is a schematic of the unit. Attrited material is collected on filters and a gas analyzer samples the outlet gas to determine attrition rate and reactivity data respectively. A settling chamber above the reactor is used to control the cut size of the attrited particles. Oversized particles, which escape the main attrition bed, are exposed to a lower superficial velocity and can drop back into the bed for additional attrition. nitrogen (N_2) is used to control the gas flow in the settling chamber and therefore the gas concentrations, as measured by the gas analyzer, are lower compared to the bed conditions. Each OC formulation underwent the same jet attrition test procedure summarized as follows:

- Agglomerates from the hot-bonding process were crushed to under 400 µm.
- A 31 g sample was prepared to the size distribution summarized in Table 3-2.

[3] Nelson, Teagan, et al. "Reactive jet and cyclonic attrition analysis of ilmenite in chemical looping combustion systems." *International Journal of Greenhouse Gas Control* 91 (2019): 102837.

- The attrition unit was heated to 900°C before addition of OC sample.
- Redox cycling began after a 10-minute inert period with the bed under nitrogen only. Cycle conditions are given in Table 3-3.
- A complete test sequence consisted of 15 cycles at low velocity, 6 cycles at medium velocity and 6 cycles at high velocity in order of increasing jet velocity (bed velocities calculated). The velocities correspond to a kinetic power input of 65, 150 and 300 W/kg respectively.
- Filters were replaced every two to three cycles, with new filters installed at the start of every reduction cycle. Filter loading was used to determine attrition rate.
- At completion of the test, the reactor was cooled under inert conditions (N_2) by shutting off heaters.

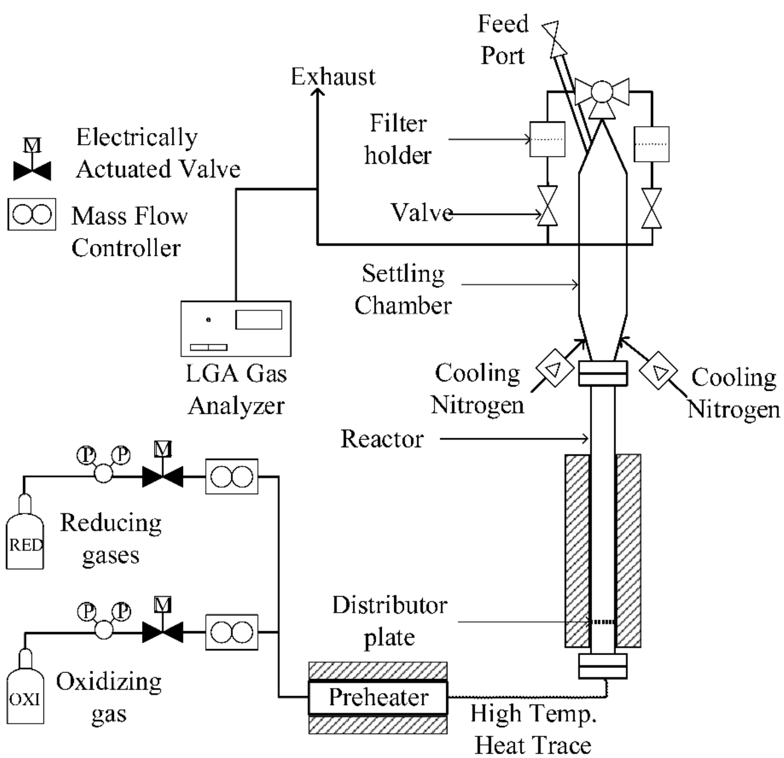

Figure 3-3 Attrition Evaluation Unit

Table 3-2 Particle Size Distribution of OC Evaluated in Attrition Unit

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

Table 3-3 Cycle conditions of jet attrition test.

	REDUCTION	PURGE	OXIDATION	PURGE
JET VELOCITY	Low / medium / high			
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
N_2 (VOL%)	80	100	90	100

Results - Jet Attrition Evaluation

Averaged attrition testing results for the 19 formulations are given in Figure 3-4 with the 9 best attrition performances benchmarked against an ilmenite sample (black box). A numerical index was used to express the attrition rate of materials. The attrition rate therefore represents the fraction of the bed material that is attrited and elutriated over a specific period (wt%/hr).

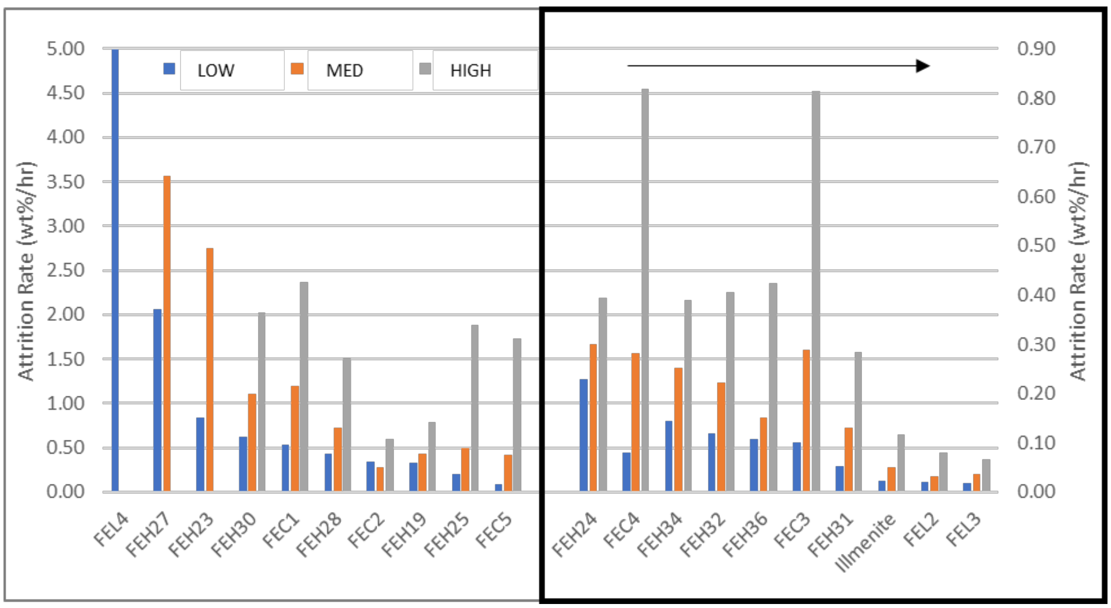

Figure 3-4 Jet Attrition Rates of 19 down-selected OC formulations.

Differences between formulations with and without attrition inhibitors were observed. The best performing formulation without inhibitors (FEH30) had an attrition rate of 0.62 %/hr at the 270 m/s condition or about five times that of formulations with inhibitors (FEH-19, 24-28, 31-36). Select attrition plots for different tests are compiled in Figure 3-5.

Two different inhibitors were used and attrition resistance was dependent on the type of attrition inhibitor used. FEH27 used a different inhibitor from FEH19, FEH24, FEH28, FEL2 and FEL3. FEH27 showed poorest attrition performance with an attrition rate of 2.1 %/hr at the 270 m/s condition. Meanwhile, for FEH25, FEH31 to FEH36, FEL4 and FEC1 to FEC5, the inhibitors were combined and a possible synergistic relationship was observed. Comparing the attrition rate of FEH28 and FEH31, FEH31 shows better attrition resistance.

The FEH formulation with the best performance was FEH31. The performance declined at higher attrition test conditions (360 and 450 m/s conditions). This formulation and others (FEH 32 – 36) had similar attrition inhibitor composition and was found to work best for the FEH group. The composition was adopted as the starting base for evaluating the low iron (FEL) and CLOU formulations (FEC).

The two low iron formulations FEL2 & FEL3, were the only ones tested that had attrition rates lower than the ilmenite benchmark. None of the CLOU-based formulations (FEC) showed performance better than FEH31 and so were not down-selected for further testing. Formulations down-selected for further evaluation were **FEH31, FEL2 and FEL3**.

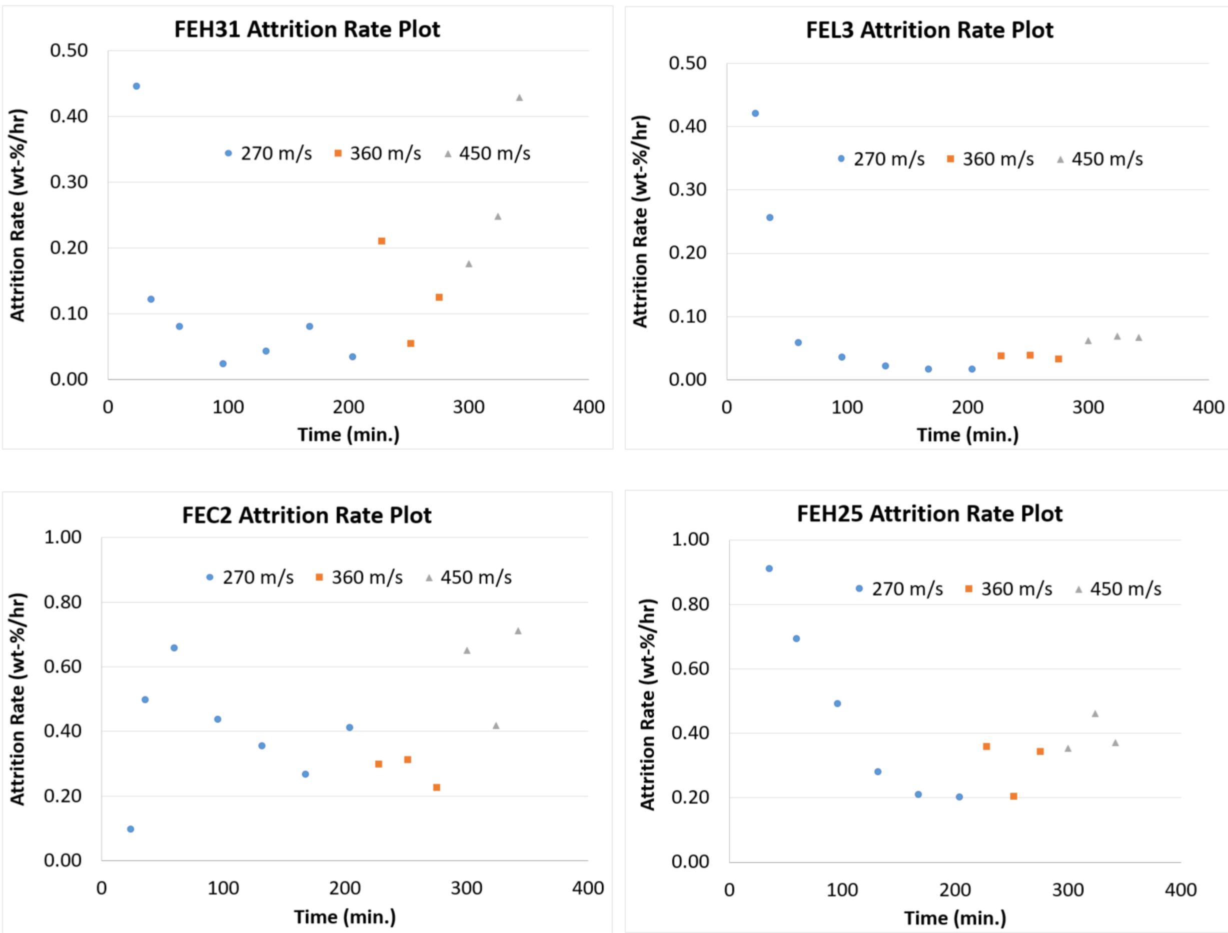

Figure 3-5 Select Attrition Plots – FEH31, FEL3, FEC2 and FEH25

Jet Attrition Reactivity Performance

CO conversion data of the formulations during attrition testing under reduction / oxidation (redox) cycling is provided in Figure 3-6. Achieving full conversion of the volume of reactants (CO and H_2) requires an O_2-equivalent weight change of 2.6, 3.4, and 4.3 wt.% of the OC for the 270, 360, and 450 m/s conditions respectively. As testing progressed, we observed that the reactor wall material (Inconel) was activated after multiple heating and cooling cycles, and began competing with the OC for the conversion of hydrogen. Therefore, hydrogen conversions during the jet attrition tests are not reported.

FEC3 (CLOU group) had the highest overall reported conversion and the highest redox potential, see Figure 3-6. Of the down-selected formulations, the FEL samples had the lowest redox potential (oxygen capacities of 2.1 and 2.3

wt% for FEL3 and FEL2 respectively) due to lower iron content. Reduction of Fe_3O_4 (magnetite) to FeO (wüstite) was also observed for the iron containing OC as the oxygen quantity required to achieve the measured gas conversion was greater than the OC capacity of hematite to magnetite alone. A representative example of a gas effluent profile during a 450 m/s condition cycle is given in Figure 3-7. Unconverted reactants are present only in the second half of the reduction phase or after all hematite is converted (2-minute mark on the graph); complete gas conversion occurs in the first half of the reduction phase, demonstrating the slower kinetics of Fe_3O_4 to FeO reduction versus Fe_2O_3 (hematite) to Fe_3O_4.

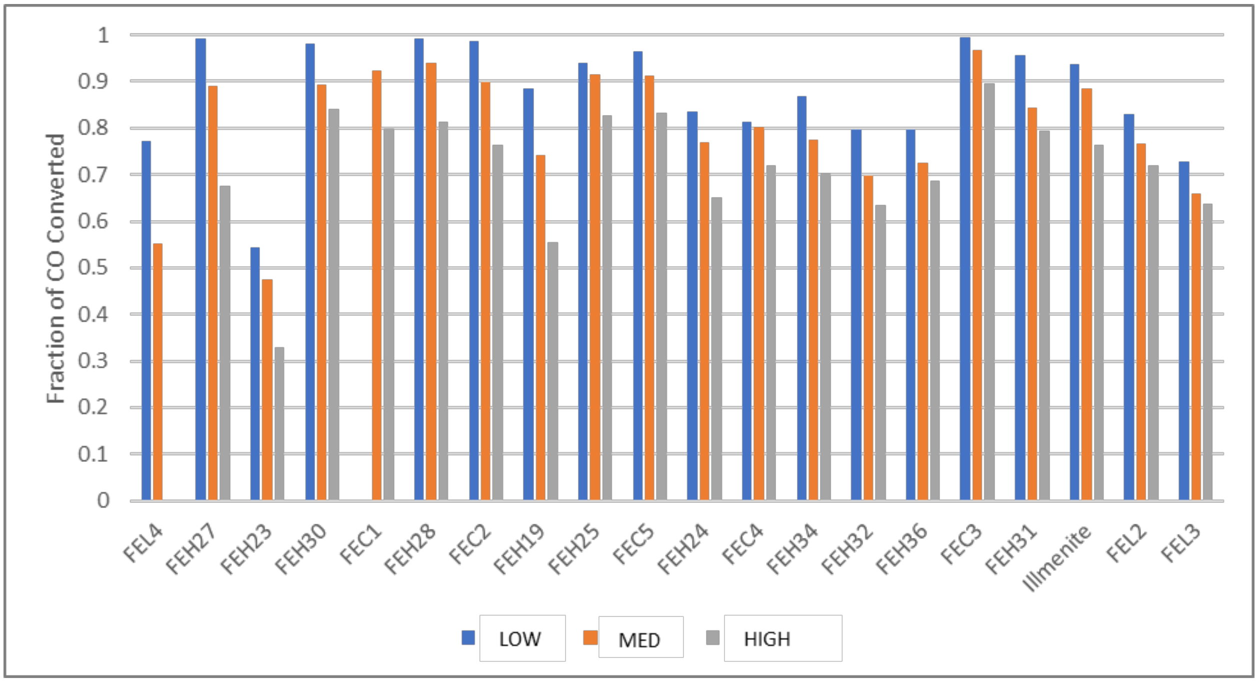

Figure 3-6 **Conversion of CO during Jet Attrition Screening.**

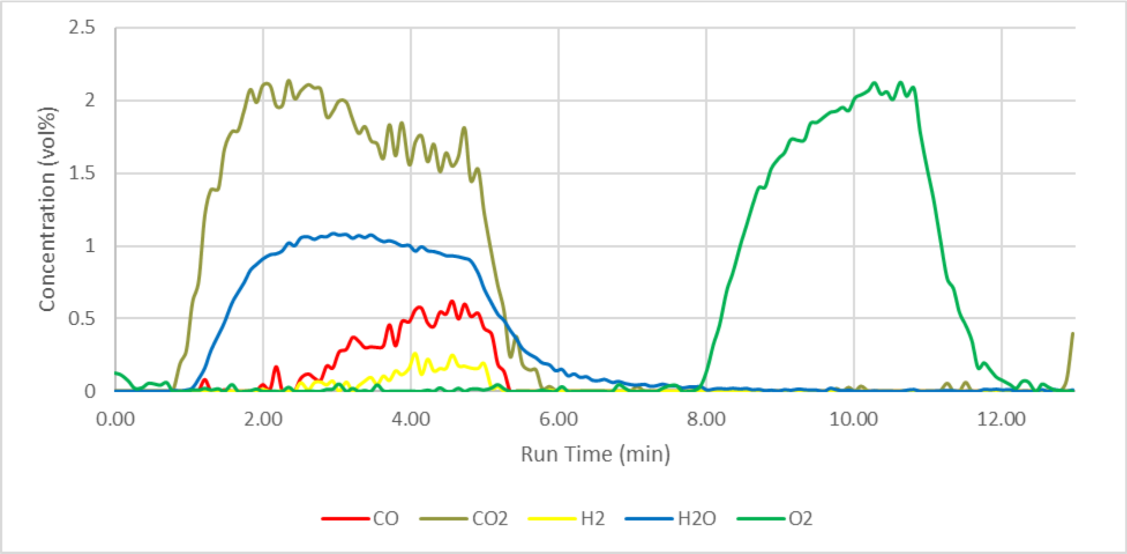

Figure 3-7 **Reactor Effluent during a High Jet Velocity Testing of FEH31**

Oxygen Carrier Activation during Jet Attrition Cycling

Improvement in reactivity performance was observed during the first 10 cycles of the fresh OC formulation. These improvements were in both rate of reaction and capacity, see Figure 3-8. These observations are best explained by an increase in particle porosity and migration of iron species to the particle surface. Nelson et al.[3] observed similar trends for ilmenite, where iron species migrated to the surface of the particle after numerous cycles.

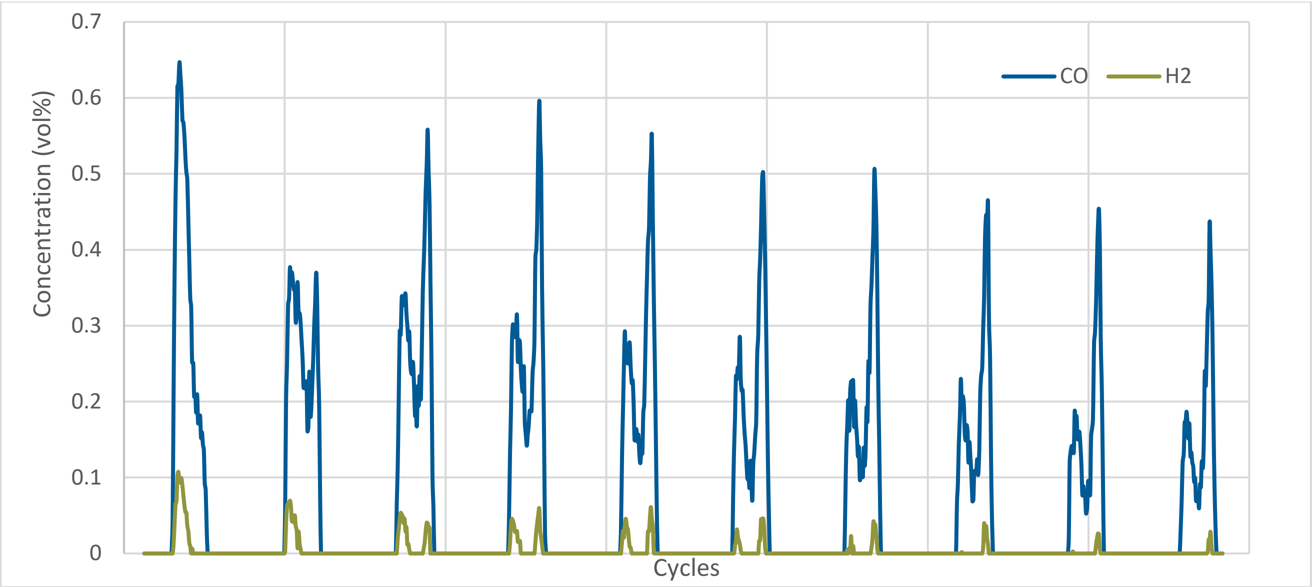

Figure 3-8 Effluent concentrations of CO and H_2 during the first 10 cycles of FEC3 showing conversion improvement.

Methodology -Cyclonic Attrition Evaluation of Down-Selected Formulations

With 3 formulations down-selected – FEH31, FEL2 and FEL3, the cyclonic attrition evaluation method was adopted to screen the formulations. Ilmenite was included in this evaluation as a benchmark. A cyclonic attrition system was adopted to assess the attrition behavior of ilmenite. The cyclonic attrition unit increases the frequency of collisions of particles, rather than operating under accelerated conditions associated with the jet attrition unit. The particles in the cyclonic attrition unit experience similar forces and operating conditions as what would be experienced in a full-scale CLC system - the frequency of impacts is the only difference. However, the cyclonic attrition process requires three times more material than the jet attrition system (100 g vs 33 g). Consequently, it was reserved for evaluating only down-selected materials. The cyclonic attrition test is performed on the same rig as the jet attrition tests with a larger reactor equipped with a spout and draft tube as shown in Figure 3-9. The OC is carried through the draft tube and then tangentially directed to the reactor walls. In effect, this setup simulates the forces expected in a gas-solid cyclone. The test matrix for cyclonic attrition is summarized in Table 3-4.

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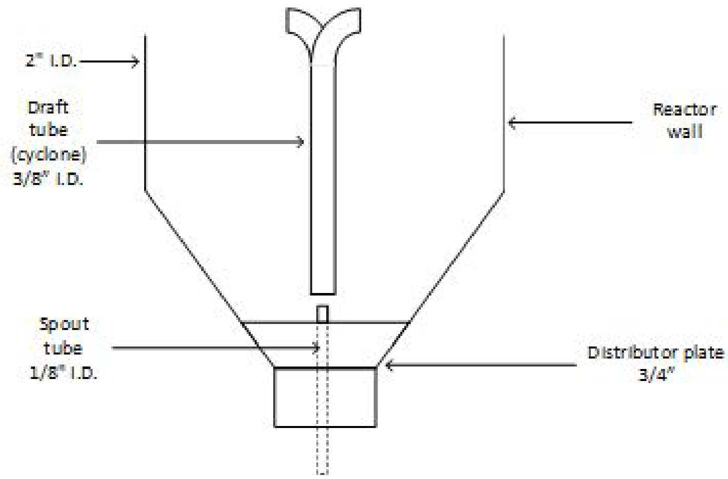


Figure 3-9 Schematic of Cyclonic Attrition Reactor

Table 3-4 Test Conditions for Cyclonic Attrition Testing

	Ilmenite	FEH31	FEL2	FEL3
Spout Velocity (m/s)	Low → Med → High			
Temperature (°C)	900	830	900	900
Run time (min)	90	30 → 60 → 90		
Bed Velocity	High	Low → Med → High		

Results – Cyclonic Attrition Testing

The results for the cyclonic attrition testing are summarized in Table 3-5 and Figure 3-10. The results suggest high cyclonic velocities could significantly impact attrition. Adoption of a low efficiency cyclone followed by a high efficiency cyclone could minimize losses via attrition. Key highlights from the results:

- The benchmark ilmenite had the lowest attrition averaging 0.0034 wt.% at 20 m/s.
- FEL3 had the lowest attrition of the down-selected formulations, averaging 0.0083 wt% at 20 m/s.
- Attrition rates increase with increase in cyclonic velocity, with FEH31 showing bigger increases

Table 3-5 Cyclonic attrition as a function of cyclone velocity (wt.% loss)

	LOW	MED	HIGH
FEH31	0.0015	0.0044	0.013
FEL2	0.0008	0.0025	0.012
FEL3	0.0011	0.0030	0.0083
ILMENITE			0.0034

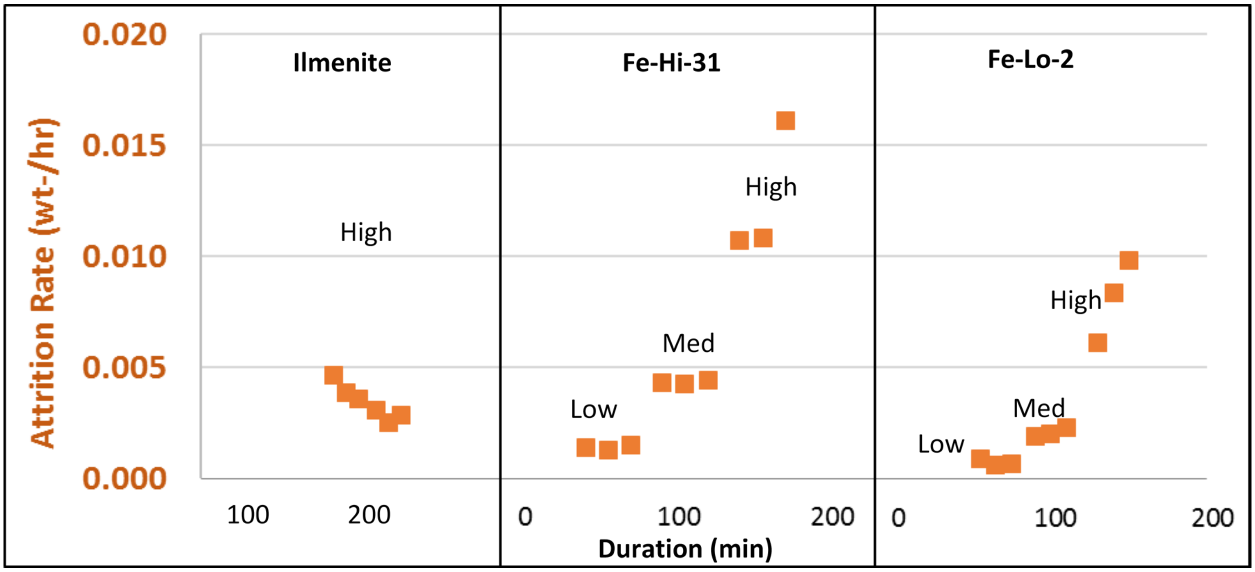


Figure 3-10 **Cyclonic attrition evaluation of ilmenite, FEH31 and FEL3 at increasing velocities**

Subtask 2.3 Longer Term Cyclic Operation and Recyclability Evaluation

This subtask focused on an extended jet attrition evaluation of the two down-selected formulations – FEH31 and FEL3. Approximately 140 cycles were evaluated. For the first 100 cycles the kinetic power input was reduced to 25 W/kg and a lower oxygen capacity of 1 wt.% to minimize accelerated conditions. For the next 40 cycles, 2 wt.% was adopted. Testing was performed over 5 sessions. Bulk density was measured between each session and samples analyzed with SEM.

Attrition rates and bulk densities of FEL3 and FEH31 over the extended tests are given in Figure 3-11. The bulk density stabilized by the second session (cycle 60) for FEL3 and FEH31. Upon increased oxygen utilization to 2 wt.% after cycle 97 the bulk density of FEH31 decreased from 1850 to 1750 kg/m^3 while FEL3 remained at its cycle-60 value of 1800 kg/m^3. SEM images appear to be similar at different cycles of FEH31 and FEL3, Figure 3-12 and Figure 3-13 respectively. Full conversion of carbon monoxide and hydrogen was achieved throughout the test for both formulations. Both oxygen carriers exhibited an increase in attrition rate followed by decline back to a baseline, but the start point and duration of this increase was different for each. This increase started earlier and was more severe for FEH31. Attrition rate of FEL3 was always at or below the rate of FEH31. Attrition rates at the conclusion of the test was 0.043 and 0.028 wt%/hr for FEH31 and FEL3 respectively.

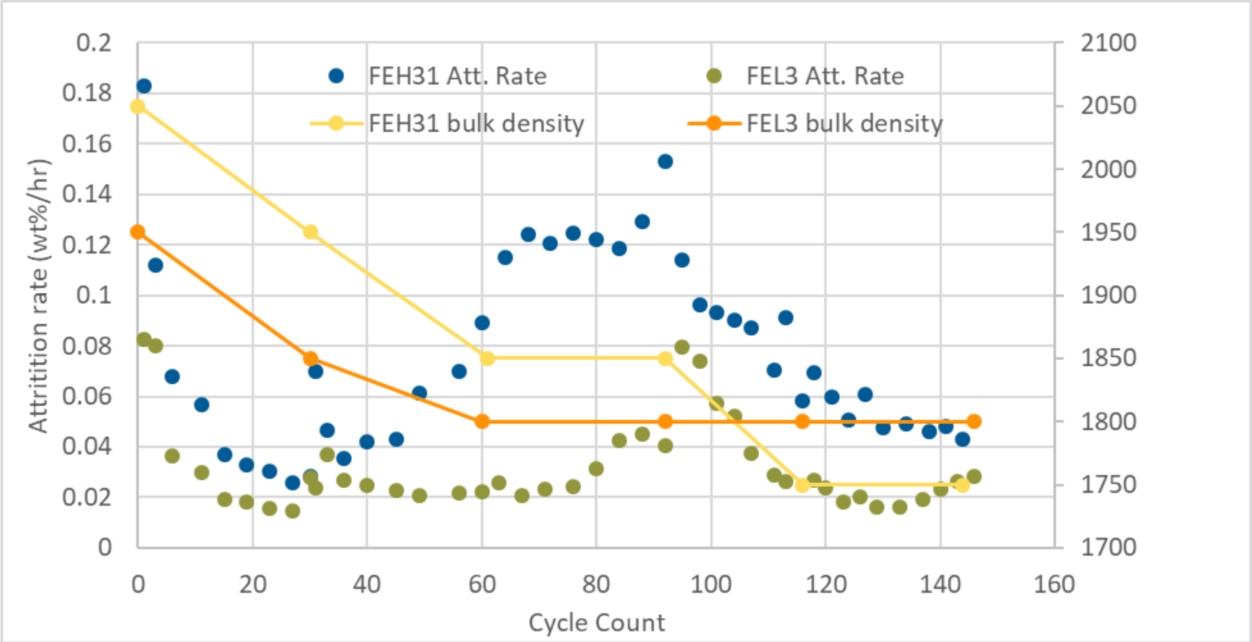


Figure 3-11 **Jet attrition and bulk density of FEH31 and FEL3 during extended jet attrition test.**

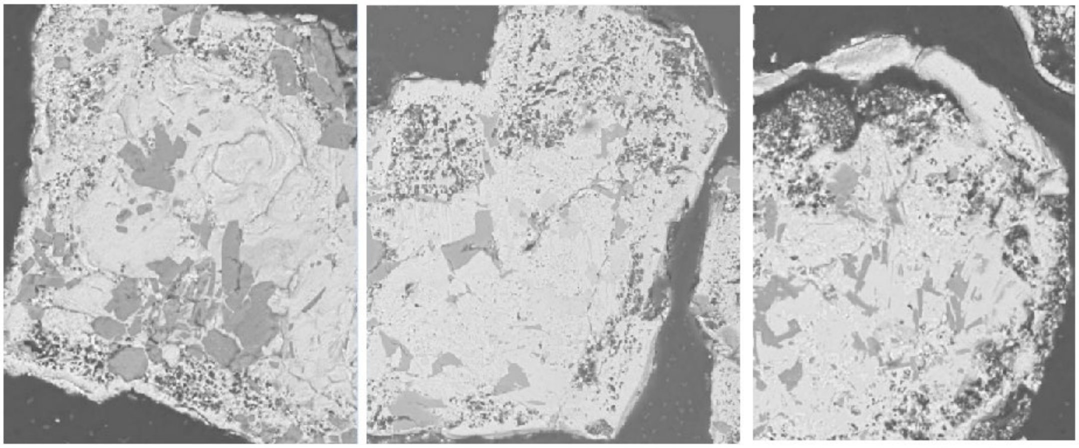


Figure 3-12 **SEM images at 500x magnification of FEH31 cross-sections after 30, 61, and 92 cycles (left to right) of the extended attrition test.**

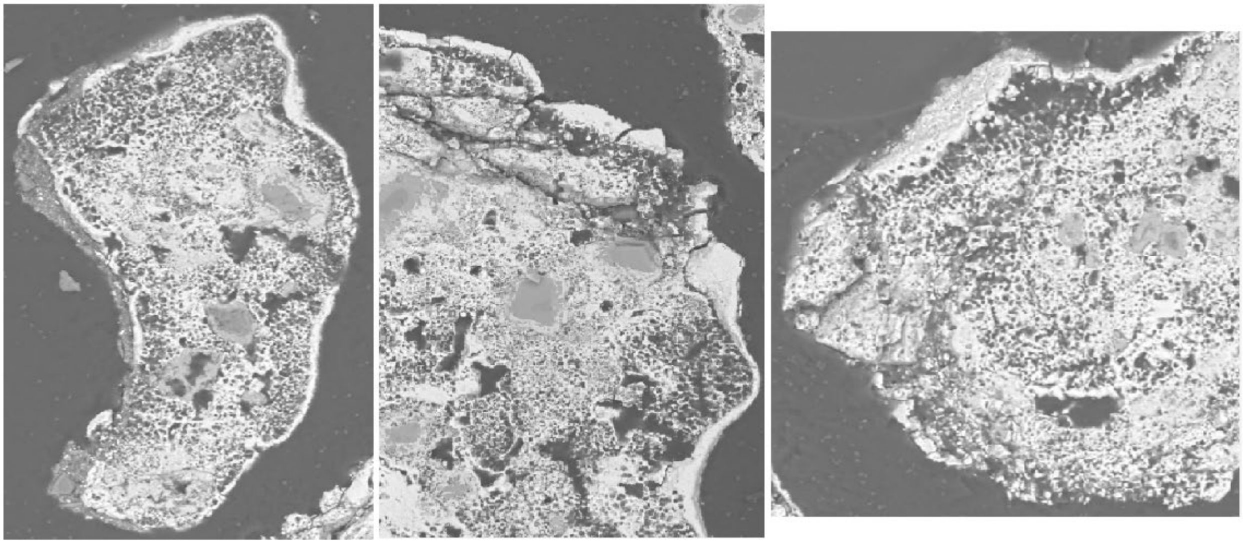


Figure 3-13 SEM images at 500x magnification of FEL3 cross sections after 60, 92, and 116 cycles (left to right) of the extended attrition test.

Task 2 Conclusion

Forty-six oxygen carrier (OC) formulations were evaluated via a multi-step screening process and one formulation, FEL3, was initially down-selected for evaluation with coal / char. However, due to challenges with procuring one of the ingredients, the alternate formulation FEH31 was selected for further evaluation.

Key achievements from this task was the development of an OC manufacturing process that includes a novel "attrition inhibitor" that showed very promising attrition performance. Additionally, the OC was developed from low cost raw materials minimizing the impact of the OC on operating performance of the CLC facility.

Task 3 – Modeling and Laboratory-Scale Evaluation of OC Performance with Coal

This task focused on evaluation of manufactured OC under varying conditions to determine kinetic parameters, reductant performance with coal, ash interactions to predict agglomeration/sintering tendencies using thermodynamic modelling, and finally, determine the recyclability of the OC.

Subtask 3.1 – Fluidized Bed Testing

The objective of this subtask was to determine the optimal operating conditions for converting coal char with ilmenite and the two down-selected OC – FEH31 and FEL3. The particle size distribution (PSD) tested is summarized in Table 3-6. The OC evaluation unit is a draft-tube assisted Spout-Fluidized Bed (SFB) with a single spout reactor described in Figure 3-14.

Table 3-6 PSD of Oxygen Carriers Evaluated in Batch Unit

µm	Fe-Hi-31 (wt.%)	Fe-Lo-3 (wt.%)	Ilmenite (wt.%)
250-420	0.2	18.4	67.7
177-250	41.0	26.9	
149-177	31.8	23.9	27.5
105-149	12.7	10.7	4.6
149-105	13.2	11.8	0.1
105-74	1.2	7.8	0.1
< 74	0	0.6	0.1
Mean Particle Size, dp,avg	214	211	230
Bulk Density, kg/m^3	~1800		2600

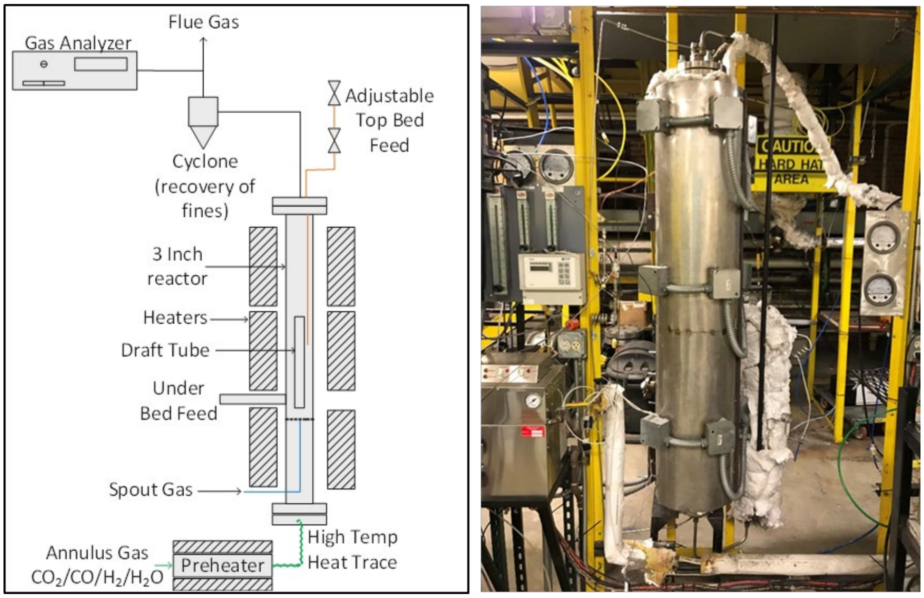


Figure 3-14 Schematic and Picture of 3 kW Laboratory Test Unit

SFBs have shown better solids mixing at bed velocities lower than for fluidized beds[4]. The geometry of the evaluation unit served as the base module for the 10 kW_{th} CLC reducer to be fabricated in Task 4. The SFB consists of a zone of high velocities known as the spouting region, and a zone of low velocities known as the annulus. Fluidizing gas is delivered separately to each zone.

For the fluidized bed testing, an 8 cm diameter (3-inch pipe), batch reactor was adopted. The unit includes a top bed feed port for solid fuels, a spout and draft tube to facilitate spouting and electric heaters. Testing with ilmenite focused on identifying CLC operating conditions for the 10 kW_{th} system. During testing, the spout of the unit was fluidized using Nitrogen and the annulus was fluidized using steam and CO_2. Gas composition was monitored using a Laser Gas Analyzer from Atmosphere Recovery Inc. Mass flow controllers were used to deliver CO_2, CO and H_2 to the bed plenum. A peristaltic pump and ceramic heaters are used to deliver steam to the system. Char was produced from a sub-bituminous coal sourced from the Absaloka mine, to minimize the effect of volatile release on the data reduction. Table 3-7 is a summary of operating conditions tested using the 3 kW_{th} system and Table 3-8 is a summary of the proximate and ultimate analyses of the char and coal.

Table 3-7 Operating Conditions for Laboratory Testing with 3 kW_{th} Batch System

Fuel	Char (from PRB coal sourced from Absaloka Mine)
Fuel Particle Size (µm)	225, **450**, 720
Fuel Feed Rate (g)	5 g batches
Temperature (°C)	~850
Inlet Gas Velocity (m/s)	Annular region: ~0.1; Spouting region: ~0.3 to 0.6
Settled Solids Height (m)	~0.25
Bed Inventory (kg)	Ilmenite ~2.2; FEH31 / FEL3 ~1.6 lg
Reduction Gas Composition (vol%)	20 H_2; 20 CO; N_2
Oxidation Gas Composition (vol%)	21 O_2; 79 N_2
Purge Gas Concentration (vol%)	100 N_2
Minimum Fluidization Velocity (cm/s)	Ilmenite ~3; FEH31 & FEL2 ~2

[4] van der Watt, J. G. *et al.* Development of a Spouted Bed Reactor for Chemical Looping Combustion. *J. Energy Resour. Technol.* **140,** 112002 (2018)

Table 3-8 Proximate and Ultimate Analysis of Coal and Char Tested in Batch Unit

	Char			Coal		
	As Det. %	Dry %	Dry/Ash Free %	As Det. %	Dry %	Dry/Ash Free %
	Proximate Analysis			Proximate Analysis		
Moisture	2.4	N/A	N/A	15.1	N/A	N/A
Volatile Matter	11.6	11.8	14.0	30.5	35.9	40.9
Fixed Carbon	71.2	72.9	86.0	44.1	52.0	59.1
Ash	14.8	15.2	N/A	10.3	12.1	N/A
	Ultimate Analysis			Ultimate Analysis		
Hydrogen	1.7	1.5	1.8	5.3	4.2	4.8
Carbon	77.1	78.9	93.1	54.3	63.9	72.8
Nitrogen	1.3	1.3	1.6	0.8	0.9	1.1
Sulfur	0.4	0.4	0.5	0.9	1.0	1.2
Oxygen	4.7	2.6	3.1	28.5	17.8	20.2
Ash	14.8	15.2	N/A	10.3	12.1	N/A

Ilmenite Testing Results

Effect of Fuel Particle Size and Steam/CO_2 ratio: Three char particle sizes ($d_{p,avg}$) were tested (Table 3-7) at a steam / CO_2 ratio of low and high. 5 g of char was injected in the annulus region at the top of the bed and the CO, H_2 and CO_2 gas concentrations monitored. Figure 3-15 summarizes the results obtained.

The smallest char particle size showed the fastest conversion to 90%. the carbon mass balance was 97% and 100% for low and high steam respectively, and 92% for all other tests. These suggest that there are minimal char losses from char elutriation out of the bed and very good solid-gas contact between the OC and the char gasification products CO and H_2. These positive observations are attributed to the lower operating velocity in the annulus of the spouted fluid bed design. Another key benefit was that operating at a lower steam flow rate did not adversely affect conversion. For the low steam condition, the total steam fed corresponds to 83% of the stoichiometric steam requirement (assuming only steam gasification).

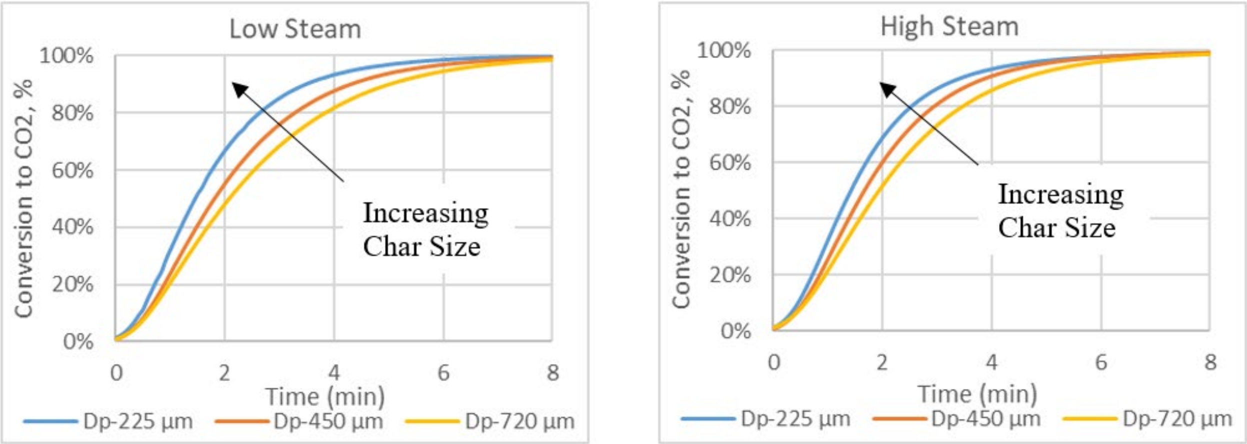


Figure 3-15 Effect of particle size and steam concentration on conversion of char

Optimum OC / Fuel Ratio: The minimum OC/Fuel ratio was determined by performing 5 g batch additions of char until H_2 break through was observed in the exit gas. Figure 3-16 summarizes the results of the char addition. After 6 injections (30 g), the conversion rate starts slowing down, see Figure 3-16. This suggests that 30 g char to 2 kg of ilmenite is optimal, for an OC/Fuel ratio of 67. A single injection of 30 g was performed to verify performance, and the conversion time was very similar to the 6 x 5 g batches.

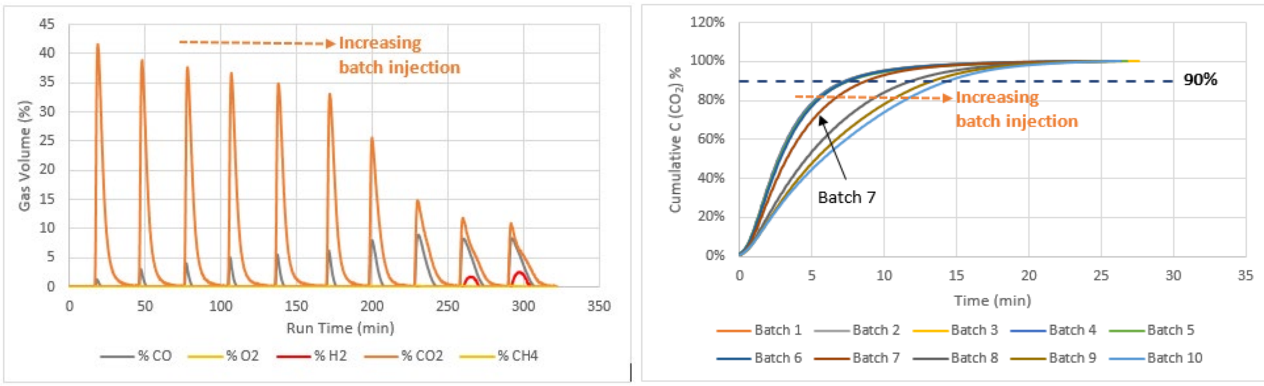


Figure 3-16 Effect of OC to Fuel ratio on solid fuel conversion

Effect of Spout Velocity: The spout velocity controls the internal solids circulation rate (mixing) in the SFB. A low and high circulation were tested to verify effect on performance. Figure 3-18 shows the effect of spout velocity on conversion of char fed at 5 g and 25g in the reactor. In both cases, the high spout velocity had faster CO_2 conversion.

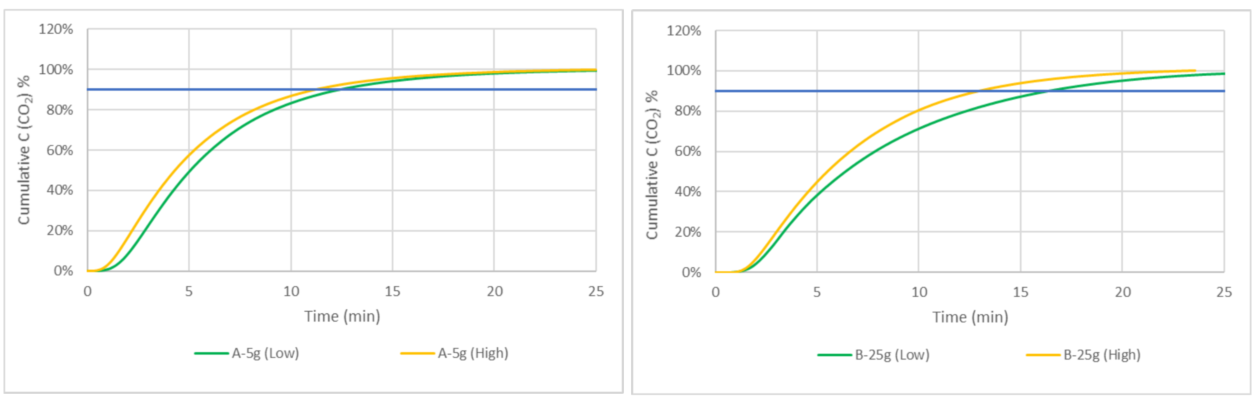


Figure 3-17 Effect of spouting velocity on conversion of solid fuel

FEL3 Testing Results

Effect of Bed Hydrodynamics and Steam/CO_2 ratio: Testing initially focused on identifying optimum operating conditions of the spouted fluid bed with FEL3 that minimize mass transfer limitations for conversion. Bed hydrodynamics refers to the velocity in the annulus and spouting region of the bed. Higher velocities will result in better solid mixing but shorter gas-to-solid residence time. Table 3-9 is the test matrix adopted and Figure 3-18 shows the results for the tests.

Table 3-9 Operating Conditions for FEL3 Batch Testing

Test	Spout Flow	Annulus Flow	Steam (Annulus)	Char Addition (g)	Temperature (°C)
1	**Low**	Low	High	5	850°C
2	**Med**	Med	High	5	850°C
3	**High**	High	High	5	850°C
4	**Low**	Low	High	5	850°C
5	**High**	Low	High	5	850°C
6	**Low**	Low	Low	5	850°C
7	**High**	Low	Low	5	850°C

To compare performance, only the time to 90% carbon conversion is considered. For tests 1 to 3, carbon conversion to 90% of all three samples occurs within a minute of each other (4.2 min to 4.9 min). It takes 12% and 17% longer for the low and high conditions respectively. At the low condition, the steam fed to the system is 84% of the stoichiometric requirement. This is desirable as steam is considered an operating expense. This observation suggests good gas-solid contacting is occurring in the annulus with some steam molecules involved in the char gasification reaction reacting more than once (similar to a moving bed).

Test 4 to 7 investigated the optimal spout flow and steam concentration in the annulus at two levels each. At high spout flows, it took less than 5 minutes to reach 90% conversion of carbon (solid plots in Figure 3-18). Reducing the steam composition of the annulus gas from 62% to 38%, resulted in a reduction in steam requirement per stoichiometric requirements of 95% to 59% and 72% to 46%. This implies savings in operating cost from reduced steam demand. With the above results, recommended operating conditions are high spout flow and low steam composition.

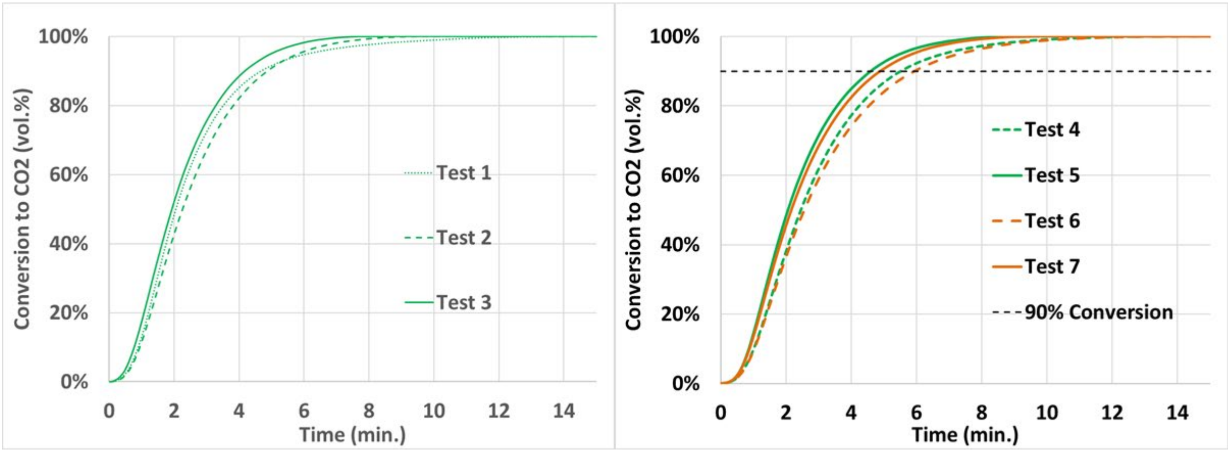


Figure 3-18 Effect of Spout Velocity on Conversion (left) and steam on conversion (right)

Char-to-Oxygen Carrier Ratio: Next, testing focused on identifying the optimal char-to-OC ratio by adding char in 5 g increments until a hydrogen breakthrough was observed. Figure 3-19 summarizes the results. After 4 injections (20 g), further addition of char to the system caused a process upset that resulted in a quick shutdown. This process upset was later attributed to deep reduction of the OC to FeO/Fe during FEH31 testing. However, due to the process upset, testing stopped and 83 was selected as the minimum OC-to-char ratio.

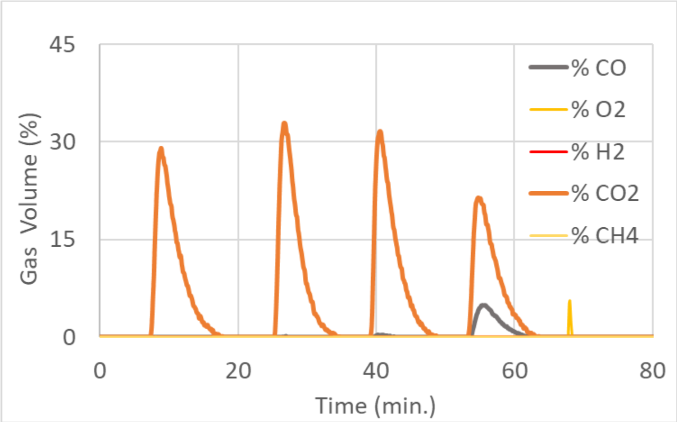
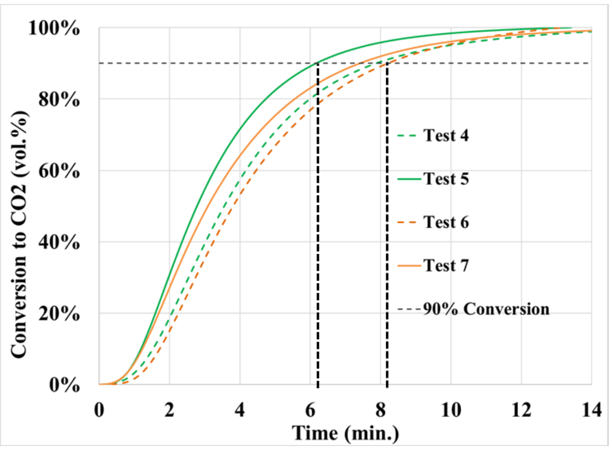


Figure 3-19 Effect of OC-to-fuel ratio on solid fuel conversion of FEL3

FEH31 Testing Results

Effect of Steam/CO_2 ratio: Testing focused on investigating the effects of spout / annulus flows and steam ratios on the conversion of carbon in the char to CO and CO_2, using a test procedure identical to Test 4 to 7 of the FEL3 OC (Table 3-9). Figure 3-20 compares the results of FEH31 and FEL3 under similar test conditions. Similar trends were observed for both FEH31 and FEL3, with the higher spout flows being quickest and the lower steam test conditions resulting in less steam used compared to the higher steam test. The observation again shows that the lower steam flow condition was not limiting to gasification compared to the higher steam flow. The results indicate that, similar to FEL3, the optimum conditions for operating the SFB unit is at high spout flow and low steam composition in the annulus fluidizing gas. This approach reduces the steam demand which is expected to be a major operating cost.

When comparing FEH31 to FEL3, the FEH31 results took approximately 25% longer to achieve 90% carbon conversion. This result was attributed to differences in test conditions – the test unit underwent maintenance between experiments, as it was unlikely that the OC is catalyzing the steam gasification process which is the rate limiting step.



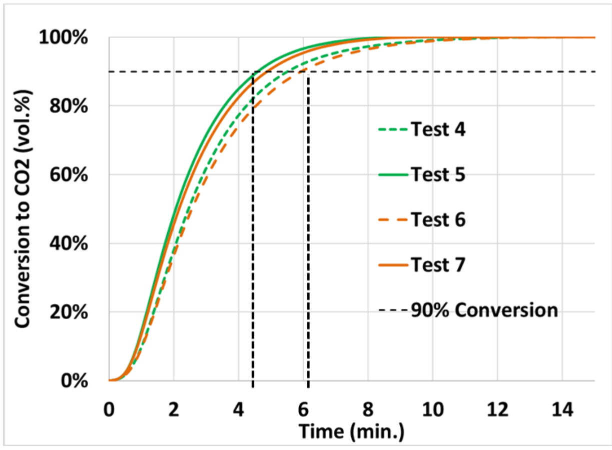


Figure 3-20 Comparison of Steam/CO_2 Ratio on conversion of FEH31 (left) and FEL3 (right)

Char-to-Oxygen Carrier Ratio: Similar to the testing with FEL3, this test campaign focused on injecting multiple batches of char into the OC bed until hydrogen breakthrough occurred. Figure 3-21 summarizes the results. The black line shows the CO_2 concentration and the peaks line up with each individual char injection. During testing, a very low concentration of Hydrogen was observed. This behavior was not observed during the FEL3 run and was initially attributed to char-steam gasification products - CO & H_2. By the third injection of char, the H_2 persisted after the CO_2 concentration dropped to 0, implying the H_2 source was not from char-steam gasification, and possibly from the steam-iron reaction of FeO and H_2O to form Fe_3O_4 and H_2. It was estimated that a negligible amount (<0.1%) of the fed steam was consumed to produce H_2. This amount though negligible, could increase as the OC is further reduced to FeO. After 4 injections, the oxygen carrier capacity of the OC was reduced by 2.7 wt%.

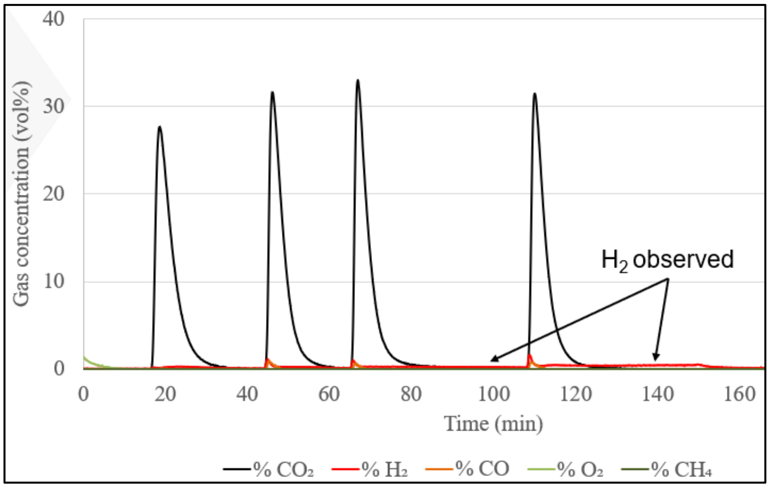


Figure 3-21 Effect of OC-to-fuel ratio on solid fuel conversion of FEH31 test day 1

The test was repeated on a separate day to determine the point where CO and H_2 release became prevalent, thus indicating depletion of the Fe_2O_3 content in the OC. The results are depicted in Figure 3-22.The material was oxidized completely before commencing with the batch char injections. H_2 and CO release was observed at the fourth char injection indicating that the oxygen capacity of the material was low, however, the persistent H_2 composition observed after the CO peak was zero suggest the reduced OC was being oxidized by the steam. At 3 injections, the oxygen carrier capacity of the OC is estimated at 2.1 wt%. Further testing discussed in Sub-task 3.2 confirmed that at reduction levels equivalent to 2wt.% oxygen capacity, Wustite formation (FeO) is occurring. In conclusion, the minimum OC-to-char ratio was averaged from both test days to give 83.

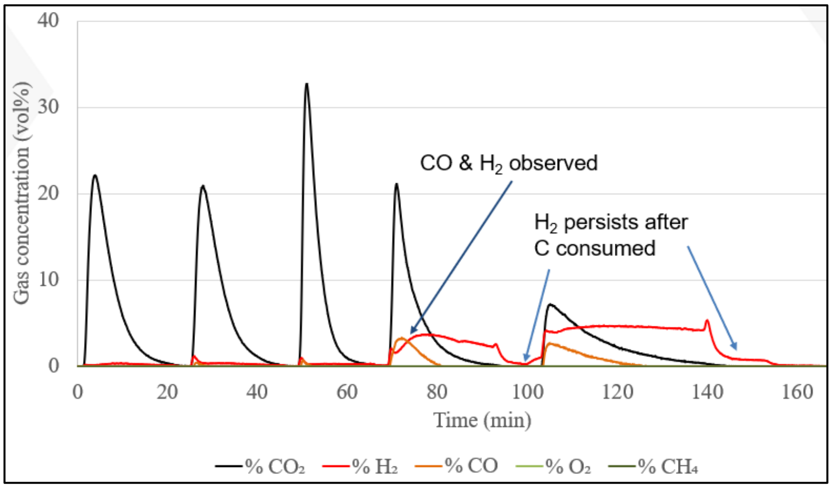


Figure 3-22 Effect of OC-to-fuel ratio on solid fuel conversion of FEH31 test day 2.

Subtask 3.2 – Experimental Evaluation of OC/Coal Ash Interactions

In this subtask, the OC was evaluated to determine the reactivity using a thermo-gravimetric analyzer followed by an evaluation of the OC/Coal ash interactions under oxidizing and reducing conditions.

Thermogravimetric Testing

Evaluation of all engineered oxygen carriers: FEH31 and FEL3 was performed with a Thermogravimetric Analyzer - Differential Scanning Calorimetry (TGA-DSC) from TA instruments. The objective of the TGA testing was to evaluate the active oxygen capacity of the materials and the reaction rates. Due to equipment limitations, the OCs were evaluated using carbon monoxide, carbon dioxide and nitrogen as the reducing gas mixture. Hydrogen and steam were not evaluated as reducing gases.

Performance of the OC is determined by plotting the rate of oxygen transfer (R_{O2}) against the oxygen utilization (X) defined as:

$$R_{O_2} = \frac{1}{M_{t=0}} \cdot \frac{dN}{dt} \quad (1)$$

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

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

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

Optimum gas flow conditions in the TGA-DSC were determined to be 210 milliliters per minute with a CO concentration of 4vol.%. Gas composition for testing the OC was then determined and are summarized in Table 3-10. Three temperatures: 800°C, 850° and 900°C were evaluated.

Table 3-10 Gas Conditions during TGA-DSC Evaluation 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%

Activation of the OC via Cycling

The engineered OCs were cycled 11 times, with the data obtained during the 11th cycle used for comparison. The FEH31 sample capacity stabilized after 4 cycles while the FEL3 took six cycles, see Figure 3-23.

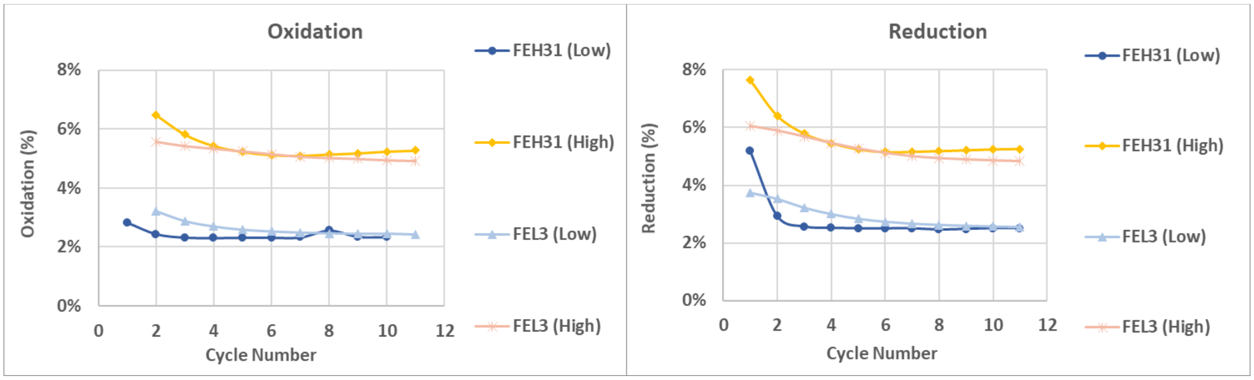


Figure 3-23 **Results of cycling engineered oxygen carriers**

Reactivity of OC

FEH31: The effect of concentration and temperature on the FEH31 sample is summarized in Figure 3-24 below, which shows the weight change during the reduction cycle tests in the TGA-DSC.

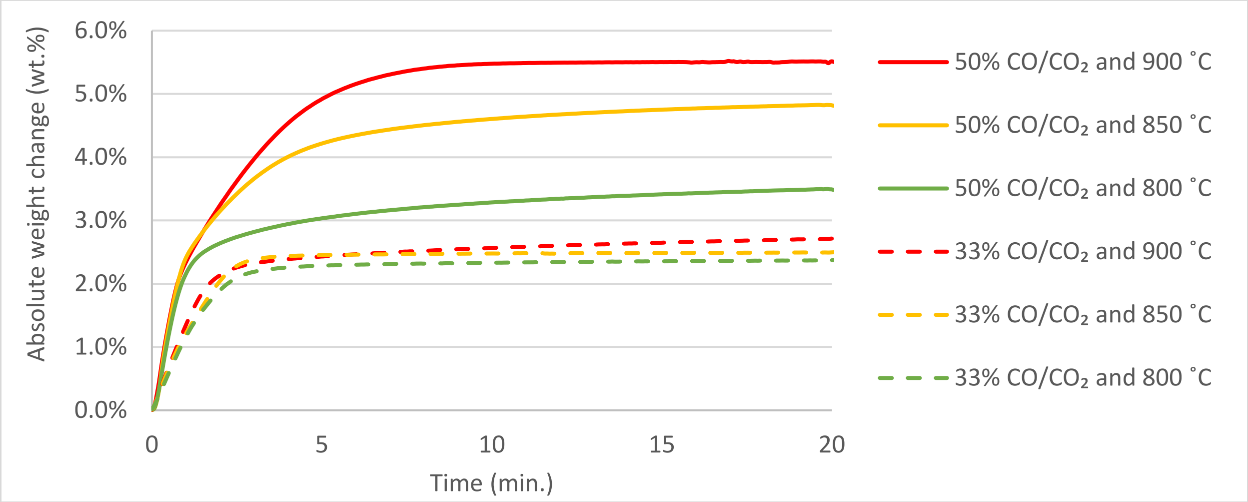


Figure 3-24 **Effect of temperature and gas composition on reduction of FEH31**

At the "Low" conditions, temperature dependency of the reduction step appears minimal (dashed lines). Meanwhile, at the "high" conditions, temperature dependency is significant after approximately one minute, which coincides with a change in the rate at of sample weight change. Assuming all the Iron present in FEH31 exists as Hematite, elemental analysis suggests that during reduction with CO, weight losses less than ~3% are most likely due to Hematite reduction to Magnetite (Fe_3O_4), meanwhile, reductions greater than 3% are due to Magnetite reduction to Wustite. Thermodynamically, magnetite reduction to wustite is temperature dependent[5], slower and controlled by pore diffusion kinetics[6], which would explain the profile observed during the "High" conditions. Plotting the TGA

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[5] Han et al. (2016) Reduction Behavior of Magnetite Pellets by CO-CO_2 Mixtures Using Direct Reduction Process. In: Hwang JY. Et al. (eds) 7th International Symposium on High-Temperature Metallurgical Processing. Springer, Cham

[6] Kuila et al. (2016) Kinetics of hydrogen reduction of magnetite ore fines. Int. J. of Hydrogen Energy, 41, 22, 9256-9266

results as the rate of oxygen transfer (R_{O2}) vs oxygen utilization (X) confirms there is a change in the order of reaction, synonymous to a switch from hematite reduction to magnetite reduction, see Figure 3-25.

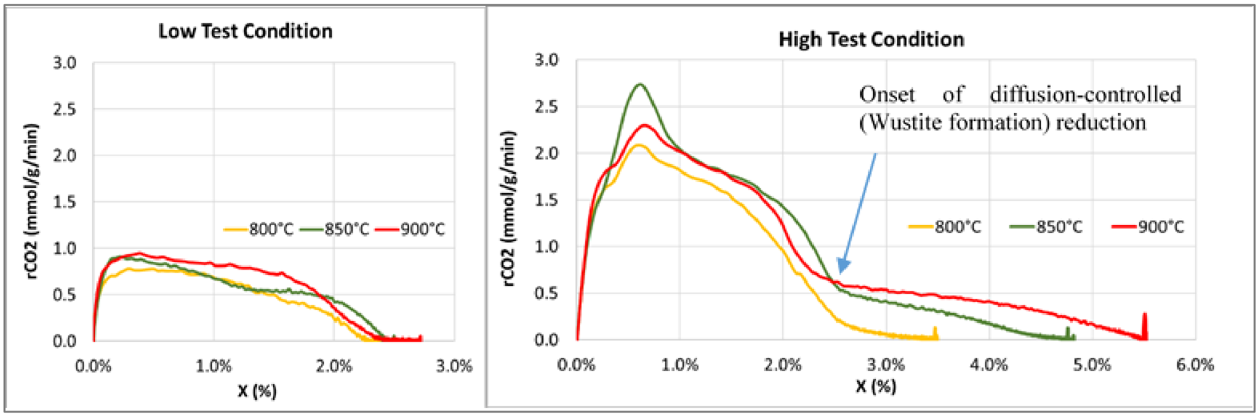

Figure 3-25 Rate of oxygen transfer during reduction of FEH31

FEL3: Evaluation of the effect of concentration and temperature on the FEL3 sample is summarized in Figure 3-26 below. At the low conditions (33% CO/CO2), temperature dependency of the reduction step appears insignificant for the 800°C and 850°C runs. At 900°C, the absolute weight change during the reduction step exceeds 3%. The composition of FEL3 suggests an oxygen content less than 2.5 wt% if all the iron gets reduced from Hematite to Magnetite. The weight change observed at 900°C suggests deeper reduction of the FEL3 with formation of Wustite.

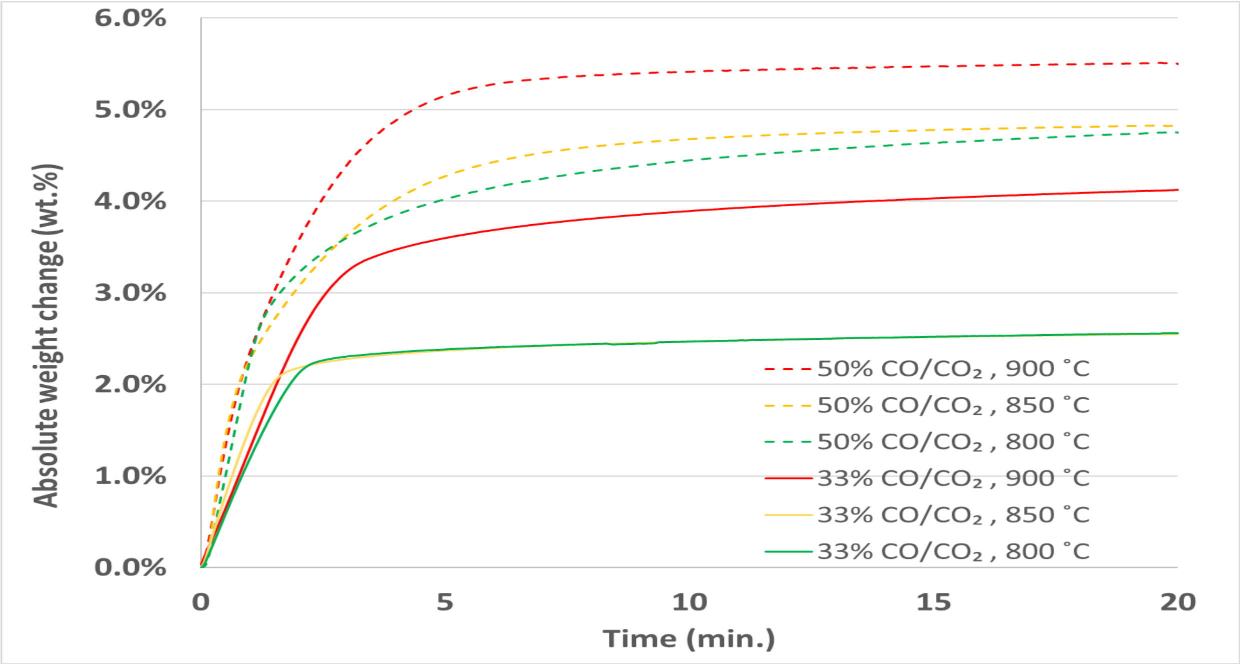

Figure 3-26 Effect of temperature and gas composition on reduction of FEL3

These results also provide possible answers to the process upset observed during the "Char-to-Oxygen Ratio" testing (subtask 3.1). The carbon content of the char at 20 g implied possible formation of Wustite. Attrition performance of FEL3 was evaluated under conditions with no formation of Wustite. The deeper reduction to Wustite implies operating conditions were beyond the desired operating range for the proposed CLC process.

At the high conditions (50% CO/CO2, Figure 3-27), temperature dependency is significant and suggests that Wustite formation predominates. This is expected due to the higher CO composition and the temperature dependence of the Wustite formation reaction[5,6]. Similar to FEH31, plots of the rate of oxygen transfer (R_{O2}) vs oxygen utilization (X) confirms there is a change in the order of reaction for the high conditions, synonymous to a switch from hematite reduction to magnetite reduction, Figure 3-27. The rate of Wustite formation for the Low condition at 900°C is low, as expected for the small CO concentration.

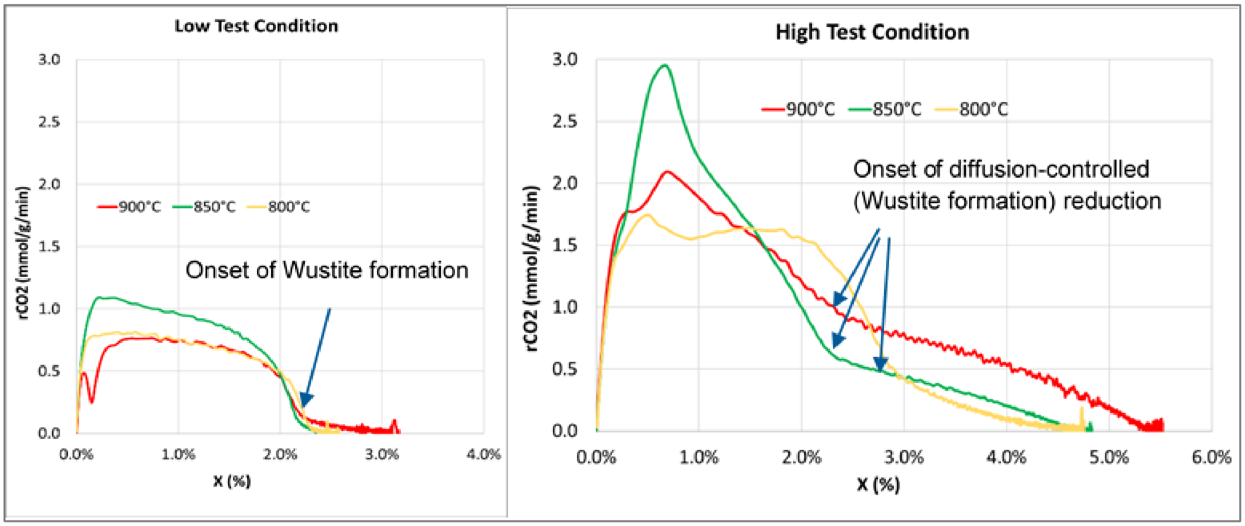

Figure 3-27 Rate of oxygen transfer during reduction of FEL3

Comparison of Engineered OC and Ilmenite

A comparison of the reactivity of the engineered OC and ilmenite shows the engineered OCs are 8 to 10 times higher than for ilmenite at CO concentrations of 4% and 10%, Figure 3-28. The oxygen capacity for ilmenite at 4vol% CO gas concentration was 3%, and just 2.5% for both engineered OCs, which was expected as the oxygen carrying capacity of ilmenite is higher than for the engineered OCs.

For a 10vol% CO concentration, the oxygen capacity of all the OCs increased – ilmenite increased to 3.5% while FEH31 and FEL3 increased from 2.5% to 5%. This increase was marked by a change in R_{O2} profile to second order suggesting that O_2 utilization is diffusion controlled. The higher CO concentration in the reducing gas resulted in deeper reduction of the engineered OCs through the occurrence of diffusion-controlled reduction.

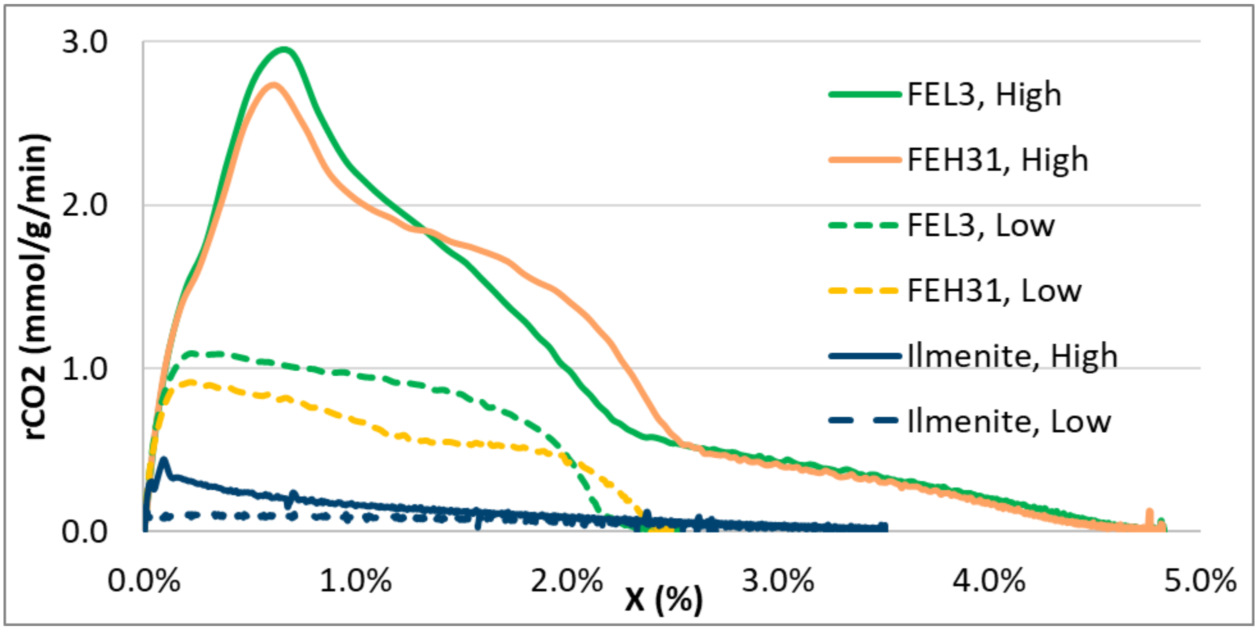

Figure 3-28 **Reactivity of ilmenite, FEH31 and FEL3 at 850°C**

Evaluation with Coal Ash

This task focused on evaluating the sintering potential of FEL3 blended with ashes produced from a sub-bituminous and lignite coal. The samples prepared were:

- Baseline ash
- 50/50 wt% blend of baseline ash and OC as-received
- 50/50 wt% blend of baseline ash and OC crushed.

A dextrin binder was added to produce small micro-pellets that were then tested in the different reactors. The prepared samplers were evaluated using two test methods:

- Non-cyclic testing in a Deltec high temperature furnace where the sample was exposed to the desired gas composition at different temperatures
- Cyclic testing of samples using the attrition test system (Figure 3-3) at a similar gas composition as the attrition testing (Table 3-3) and 900°C

Figure 3-29 is an image of the pellets after testing and Table 3-11 is a summary of the test matrix. Figure 3-30 to Figure 3-33 give comparisons between various pellets sintered at different temperatures and conditions. These images show the effect of time, temperature, coal ash type, and cycling on the deposits structure. In general, it was observed that higher temperatures, reducing conditions, and addition of OC resulted in stronger bonds between the particles and more molten material. These experimental results agree with the modeling efforts from FactSage.

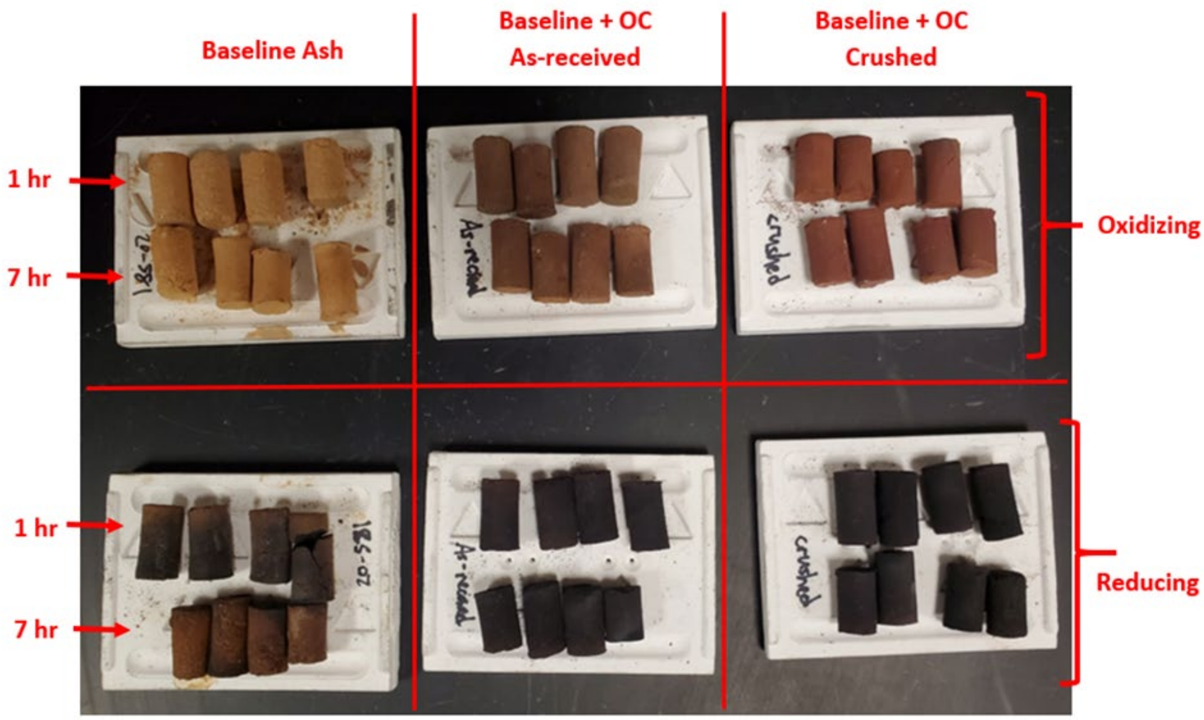

Figure 3-29 **Thermal treated ash and ash/OC blends under oxidizing and reducing conditions**

Table 3-11 **Test conditions for OC/Coal Ash Interactions**

	Sample Description	Temperature	Condition	
Sintered	Sub-bituminous 100%	900C, 1100C	Oxidizing: Air	Reducing: N_2: 80% CO_2: 8% CO: 12%
	Sub-bituminous/FEL3 50/50 wt.%	900C, 1100C		
	Lignite 100%	900C, 1100C		
	Lignite/FEL3 50/50wt.%	900C, 1100C		
Attrition Unit	Sub-bituminous 100%	900C	Table 3-3	
	Sub-bituminous /FEL3 50/50 wt.%	900C		
	Lignite 100%	900C		
	Lignite/FEL3 50/50wt.%	900C		

Effect of Time under oxidizing conditions: Figure 3-30 is an SEM image comparing the effect of time. Picture (a) in the figure shows Absaloka coal ash and the OC sample sintered for 7 hours and picture (b) shows sintering for sintering for 1 hour. The OC particles are still distinct from the finer ash dispersed around the particles suggesting no significant agglomerate formation or onset of sintering. The ash matrix (darker region) is less distinct for the 7 hr run making it harder to discern the OC particles.

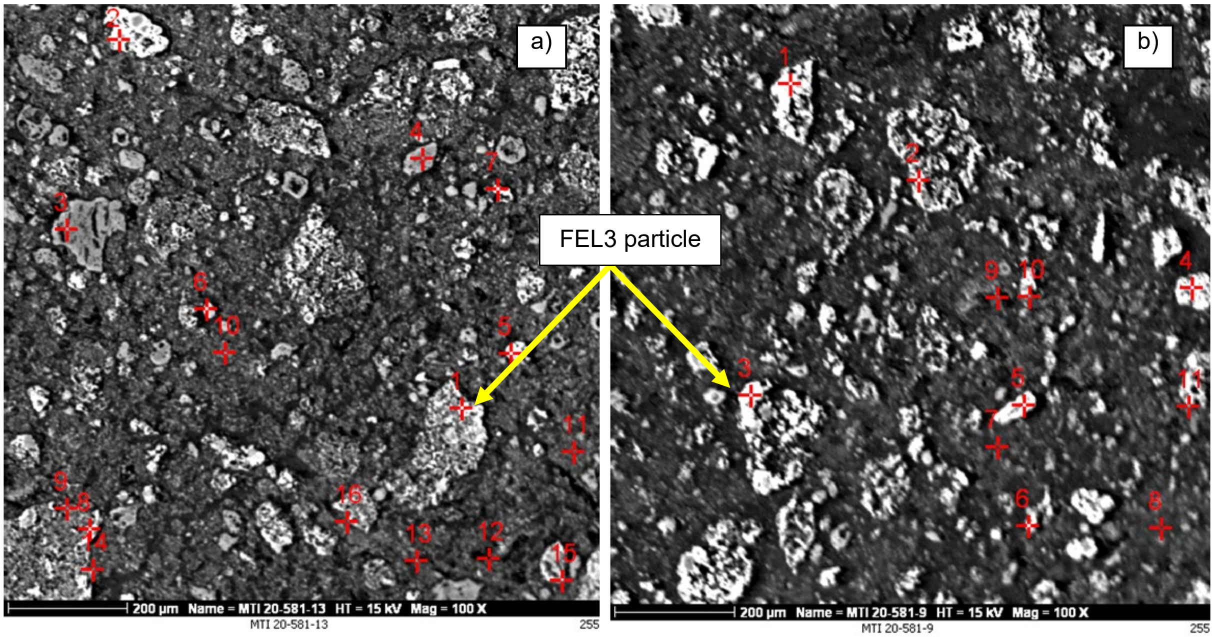

Figure 3-30 Morphology comparison of the effect of time on sintering under oxidizing conditions of absaloka coal ash and FEL3 at 900°C

Effect of Temperature under reducing conditions for 6 hours: Figure 3-31 shows the effect of 900°C (left) and 1100°C (right images) on the morphology. The top pictures (a and b) consist of blends of absaloka ash and FEL3. The bottom images (c and d) show only the absaloka ash with no FEL3. At 1100°C, the ash fuses with no distinct layers apparent between them, irrespective of the presence of OC or not. This suggests at high temperatures over long durations (dead zones in reactor) agglomerates are bound to form.

At 900°C, the onset of agglomeration is visible in picture (a) which contains the OC. In picture (c) with only ash present, mineral inclusions in the ash are visible with bright spots detected as calcium-rich alumino-silicates. The presence of the iron-rich OC with the ash in picture (a) apparently results in sintering at the accelerated conditions of 8% CO and 6 hours given the morphology is distinct from an oxidized sample at similar conditions (Figure 3-30). This implies sintering / agglomeration is more likely to occur in the reducer.

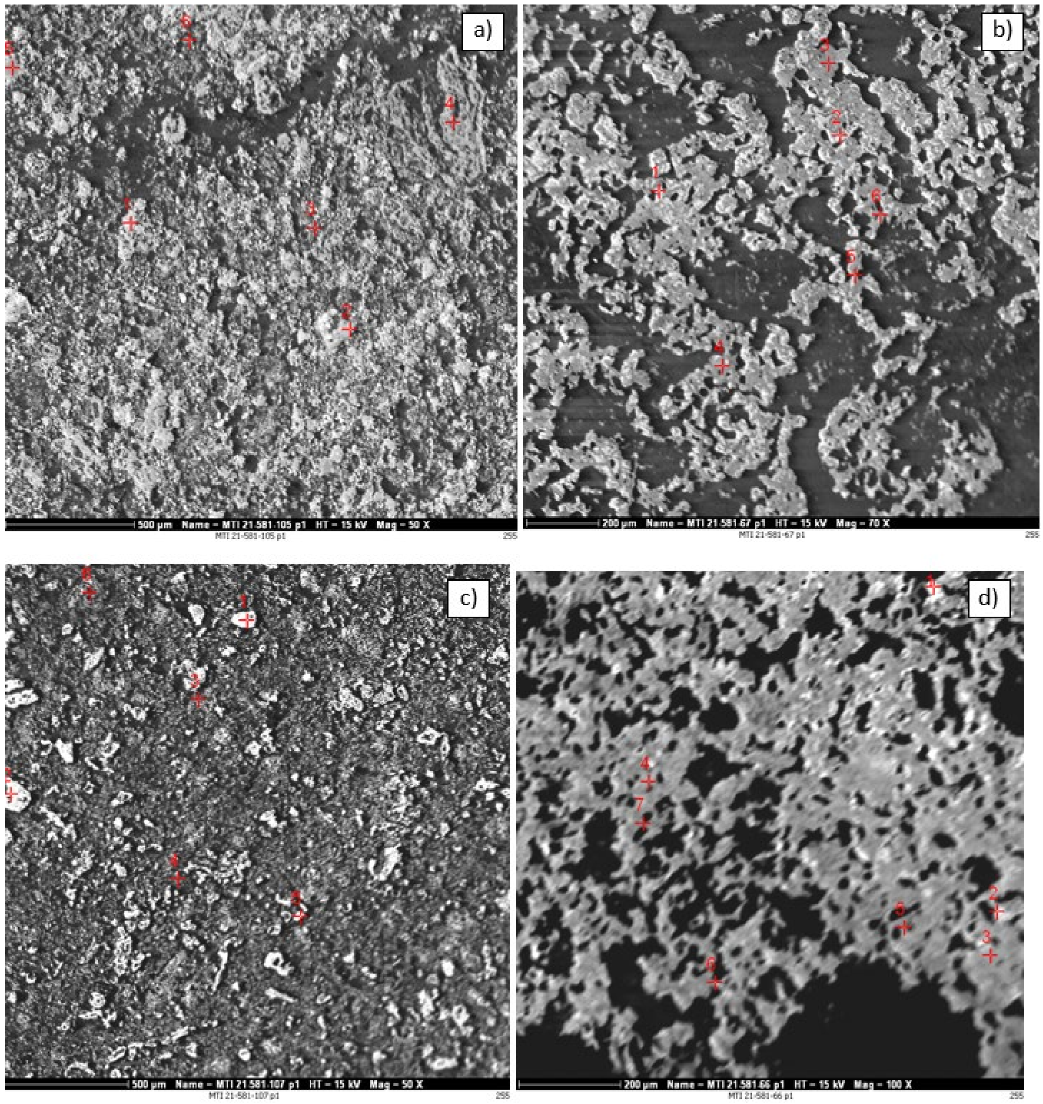

Figure 3-31 Effect of temperature and oxygen carrier presence on the morphology of ash

Effect of Ash Type: Figure 3-32 compares the sintering/agglomeration potential as a function of ash type. The sub-bituminous-based ash (left picture) shows particles are less distinct than for the lignite-based ash (right picture). Point analysis of the lignite-based ash showed that the distinct particles were included minerals rich in silica and the regions with high iron content were not distinct from the ash matrix. Both low rank coals are expected to show onset of sintering at 900°C.

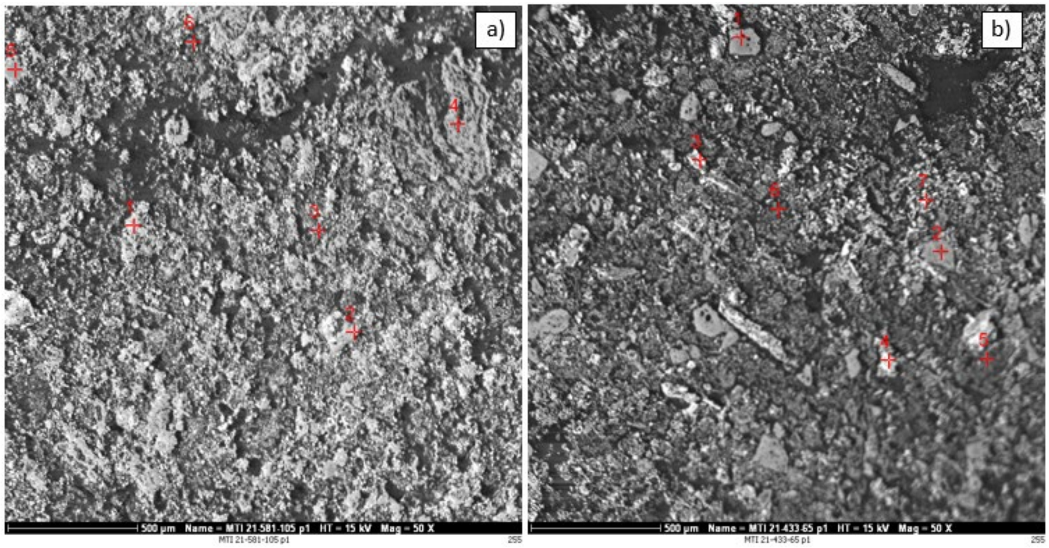

Figure 3-32 Effect of FEL3 blending with sub-bituminous ash (left) and lignite (right) at 900°C under reducing conditions

Effect of Cycling on Ash / FEL3 Interactions: The lignite coal ash was cycled at 900°C under reducing and oxidizing conditions (Table 3-11). The morphology post 6 hr cycling is shown in Figure 3-33 with image (a) lignite ash only, and (b) lignite ash blended with FEL3. The ash which was pulverized shows significant fusing during the cycling (image a). For the blended samples, it is not possible to clearly differentiate the ash from the OC suggesting interaction between the ash and the OC sample.

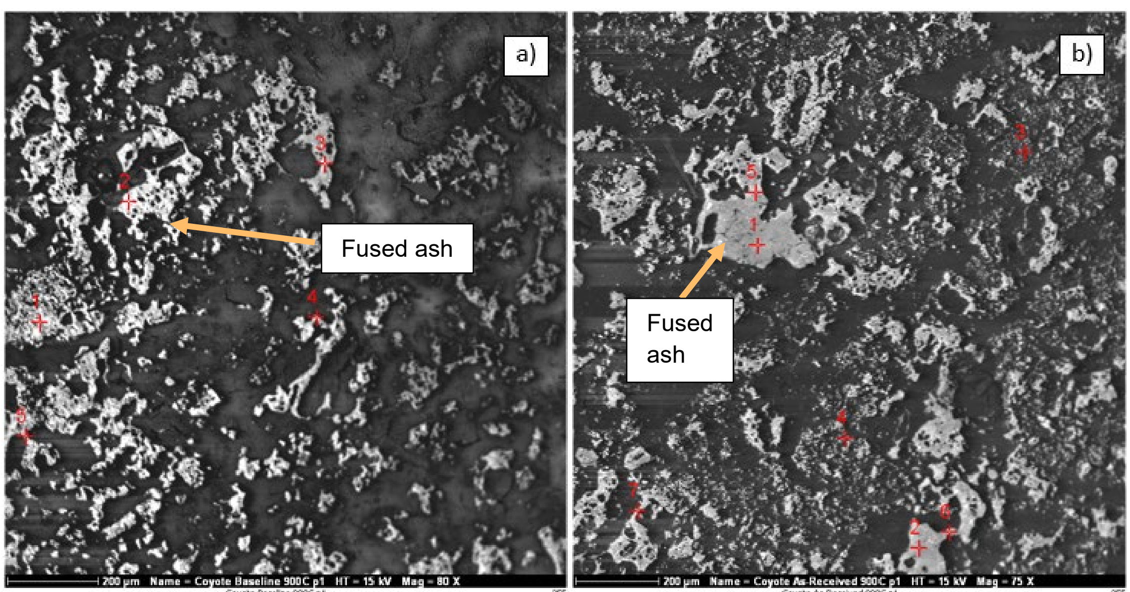

Figure 3-33 Effect of Cycling on ash and ash / FEL3 morphology at 900°C

Subtask 3.3 – Thermochemical Equilibrium Modelling

FactSage modeling was conducted by using the coal ash composition and OC compositions at various coal ash and OC ratios in air, to provide data for predicted OC conversion in both the reducing and oxidizing sections and to predict potential problematic deposit phases in the CLC. The temperature range in these modeling efforts was 600 °C – 1200 °C in both reducing and oxidizing conditions. A custom database was selected using the FactPS and FTOxid databases.

To begin modeling, the reducer section of the CLC was modeled by using the oxygen carrier, coal, and annular gas as inputs. After this modeling was completed, modeling of the oxidizing section of the CLC was completed by using the solids output predicted by FactSage from the reducing section and air as inputs. The solids input into the oxidizing section includes the reduced OC and coal ash, and removes any predicted FactSage liquids, or “sticky” material. The “sticky” material is removed from input into the oxidizing section as that material is predicted to agglomerate or “stick” to the walls of the reducing section.

Results of FactSage Modelling

Liquid Phase Formation

The appearance of liquids is considered an indication of the onset of sintering and/or agglomeration. The wt.% solids and liquids are given as a weight percent of the non-gaseous components of the materials modeled in the respective oxidizing and reducing sections of the CLC. Table 3-15 summarizes the temperature at which the onset of a liquid phase was predicted by FactSage modelling.

Figure 3-34 and Figure 3-35 depict predicted liquid formation for FEH31 and FEL3 respectively. Formation of the liquid phase is predicted to be more significant in the reducer section with predictions of up to 100% observed at higher temperatures. In the oxidizer, liquid phase predictions are under 20% at the higher temperatures modelled. This suggests that the reducing conditions in the reducer are key contributors to liquids forming.

In comparing the FactSage modeling to the sintering experiments, it was observed that FactSage predicted more liquids under reducing conditions and with the addition of OC. This is expected as thermochemical modelling does not factor reaction rates. Figure 3-30 seems to suggest this effect as the sample with the longer exposure appears to show a slight change in morphology, meaning these modeling results agree with the experimental results. The morphology results also show that with the increase in temperature, reducing conditions, and addition of OC, the liquid phase increases. The effect of melting and sintering can be seen in the images in the previous section.

Table 3-12 Temperatures at which onset of liquids is observed for OC / Ash blends

	Reducing		Oxidizing	
	FEH31	FEL3	FEH31	FEL3
25 Ratio	1000	850	-	900
75 Ratio	1000	850	-	900
100 Ratio	950	900	-	-
125 Ratio	950	850	-	1000
150 Ratio	950	850	-	1100
200 Ratio	950	850	-	-

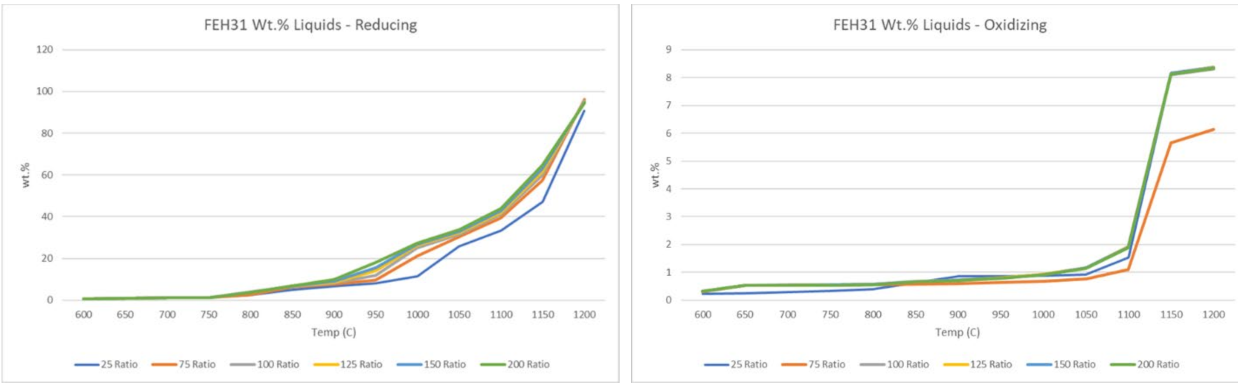

Figure 3-34 Liquid phase formation for FEH31 under oxidizing and reducing conditions

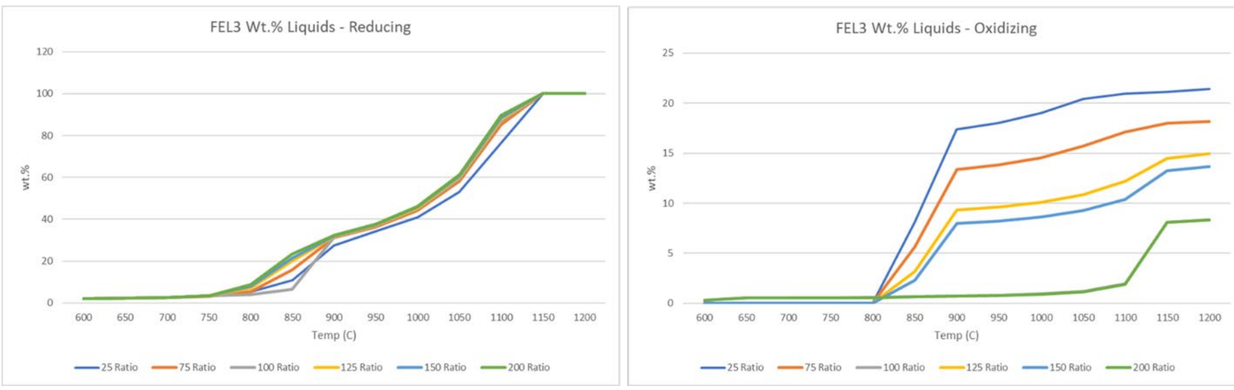

Figure 3-35 Liquid phase formation for FEL3 under oxidizing and reducing conditions

Subtask 3.4 – OC Fines Separation and Recyclability

The recyclability of the engineered OC is a key aspect of the proposed process that would significantly impact replacement costs for attrited oxygen carrier. For this task, FEL3 OC cycled in the 3 kW_{th} bench unit with the coal char was used. The total cycle numbers the sample had undergone were estimated at 20. The FEL3 was then ground down to initial particle size and reformulated using the formulation procedure minus any additives that are inactive and would be diluents. Only attrition inhibitors were added to the sample.

The sample was then subjected to the screening procedure outlined in task 2 and successfully passed all screening methods. The sample was then subjected to a jet attrition test similar to the original formulation. The results are presented in Figure 3-36 showing the performance of the recycled oxygen carrier is comparable with the original formulation for the low and medium conditions, with a slight increase at the high attrition condition. The recycled FEL3 condition performance is similar to FEH31 confirming the attrition inhibitor is a key aspect of the novel formulation method.

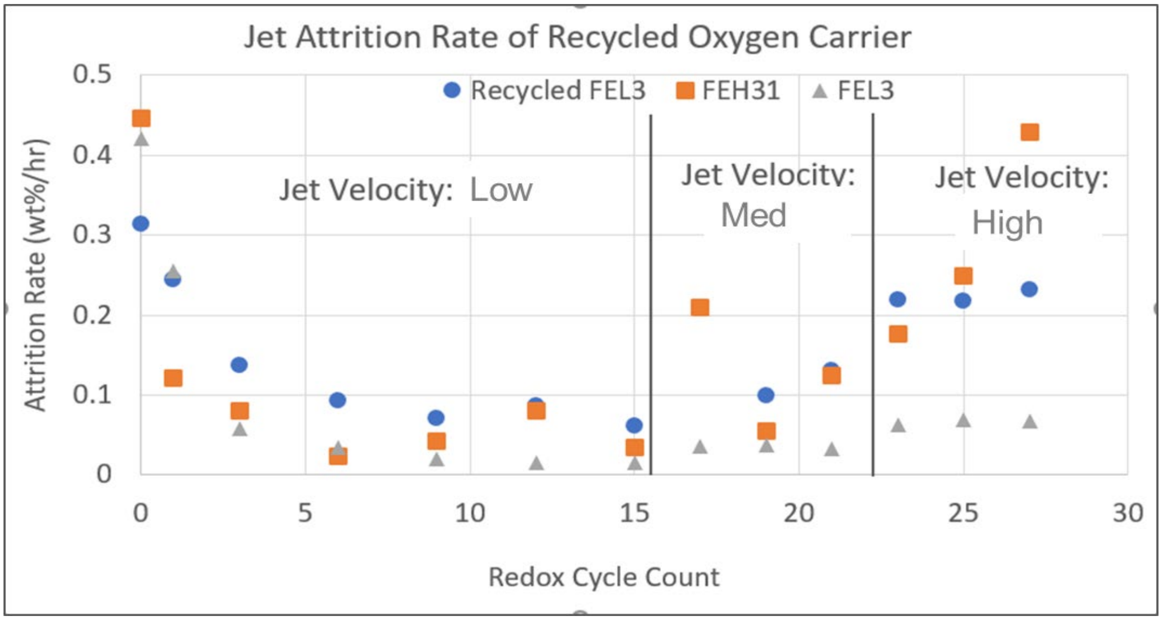

Figure 3-36 **Recycled FEL3 formulation compared with original formulation for FEL3 and FEH31**

Task 3 Conclusion

Three oxygen carriers – ilmenite, FEH31 and FEL3, were evaluated to determine key metrics in a CLC process as outlined below:

- Minimum OC-to-char ratio was 67 and 83 for ilmenite and engineered OC (FEL3 and FEH31) respectively. Given the ratio of carbon between the char and coal is 1.25, the recommended minimum OC-to-coal ratio is approximately 85 for ilmenite and 105 for FEL3 and FEH31. This minimum is based on reactivity only.

- During operation of the spout fluid bed, a high spouting flow was preferred due to the slightly faster 90% conversion of the carbon in the char.

- A low steam ratio is desirable during reduction reactions to reduce the operating cost associated with providing steam as a fluidizing agent

- For FEL3/FEH31, excessive reduction of the OC would result in the formation of FeO which might not be desirable. An oxygen carrying capacity of approximately 1wt.% was thus recommended for further testing.

- The reactivity of the FEL3/FEH31 OC are 4 to 8 times higher than for ilmenite.

- At 900°C, FEL3 reduction to wustite is likely to occur at CO concentrations greater than 4%.

- Possible onset of sintering / agglomeration was observed at operating conditions of ~900°C using both FactSage modelling and experimental testing. The kinetics of the liquid phase formation are key for determining the significance of the issue.

- FEL3 was successfully reformulated after multiple cycles including ash with performance that is comparable to initial formulation. This finding is significant regarding variable costs of operating a CLC facility.

Task 4 – 10 kW_{th} Integrated System Installation

Cold Flow – Design and construction

A cold flow unit was constructed to evaluate bed hydrodynamics and determine operating parameters for the 10 kW_{th} system design as shown in Figure 3-37. Ilmenite was the oxygen carrier (OC) selected as the design basis for the unit and was used for shakedown of the 10 kW_{th} system.

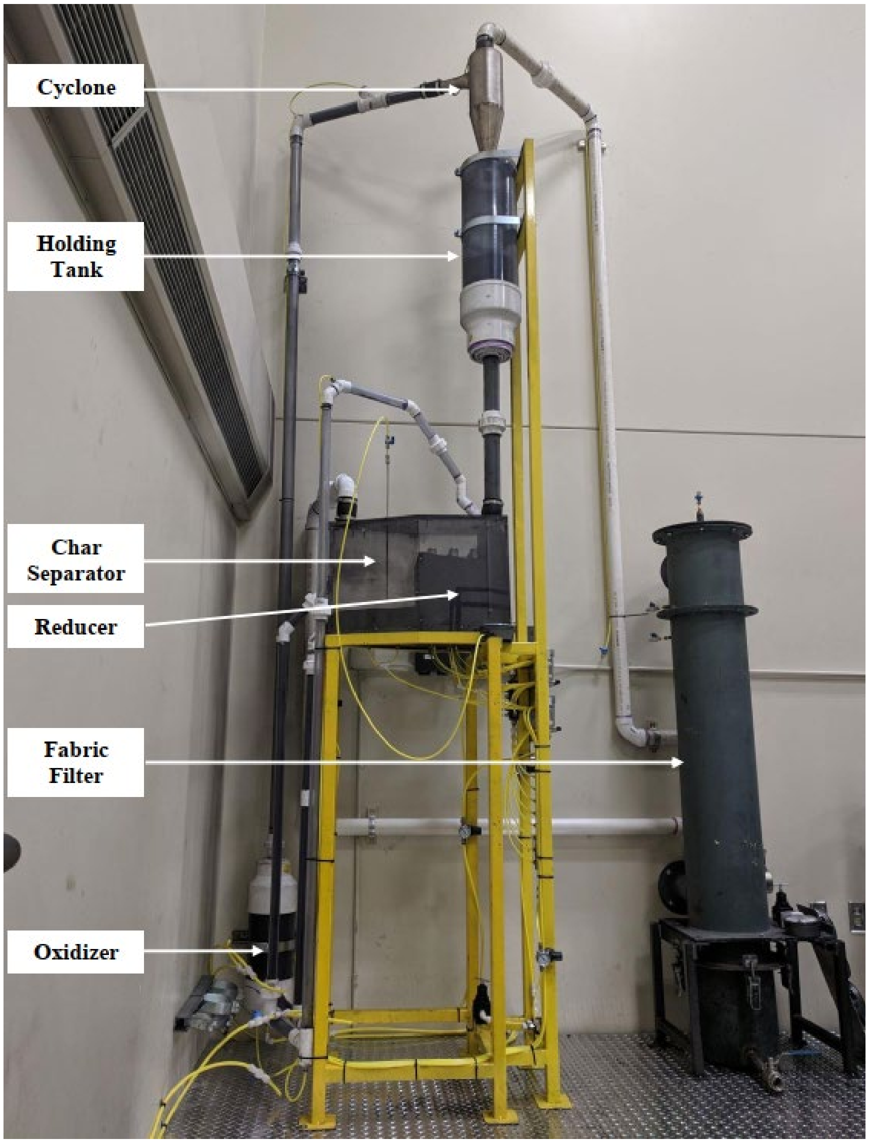

Figure 3-37 Constructed Cold Flow unit

$10kW_{th}$ system – Design and construction

The 10 kW_{th} system design is based on results from the cold flow unit testing and fluidized bed testing in subtask 3.1. It was designed for a solids throughput of approximately 200 kg/hr, a coal feed rate of 2 kg/hr, and a minimum oxygen carrier to coal ratio of 75. The total oxygen carrier inventory of the system was initially 100 kg, but reduced to ~ 50 kg with a re-design of the oxidizer. Below in Figure 3-38 is a 3-D representation of the system on the left, highlighting major sections of the equipment; (1) oxidizer, this design underwent modifications which are discussed below, (2) Riser Section, (3) Solids Downcomer, (4) Reducer section, and (5) the char separator. The picture to the right is the original system as it was constructed before shakedown testing.

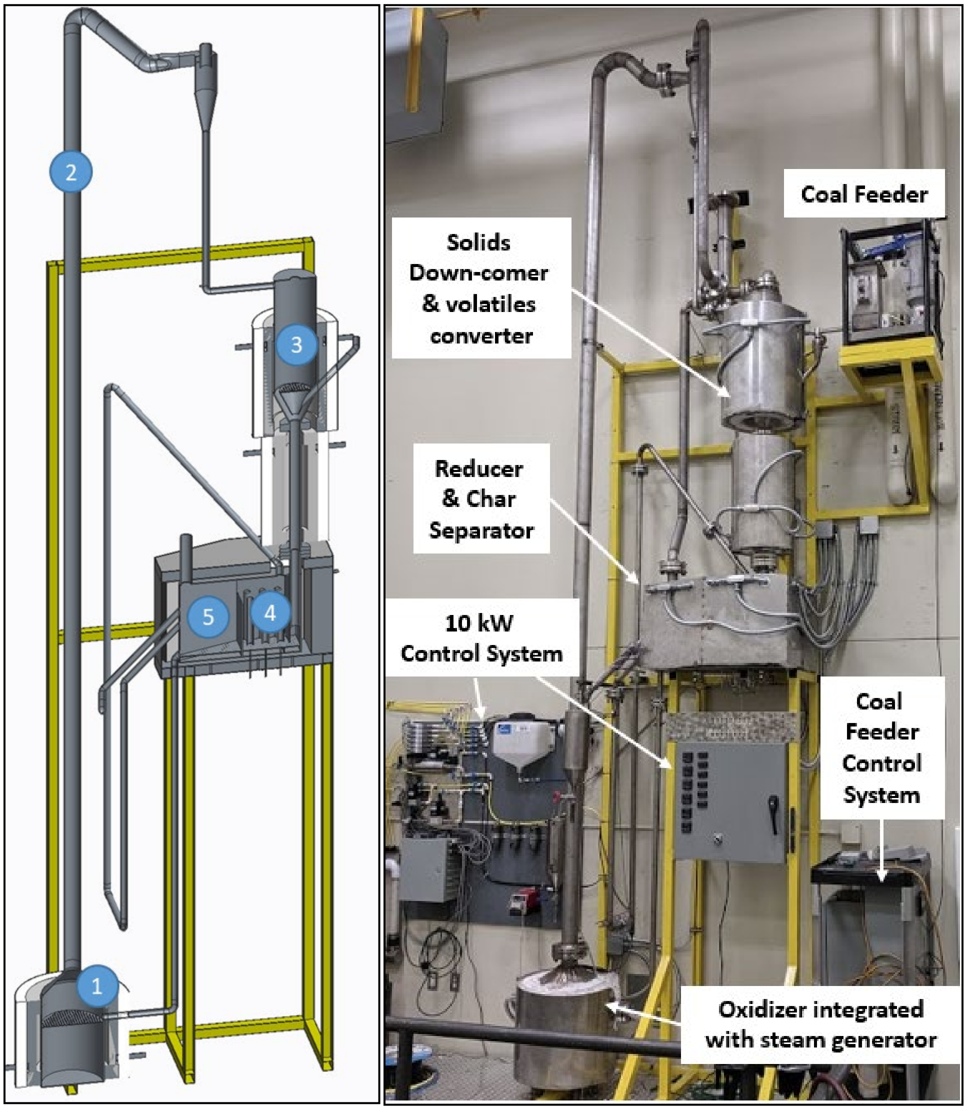

Figure 3-38 Left: Cross-sectional 3D model of CLC unit, Right: Simplified CLC schematic

Oxidizer, Riser and Cyclone

The oxidizer unit underwent design modifications during shakedown of the $10kW_{th}$ system to improve the temperature profile of the system and minimize heat losses. Final system design adopted was an 8 cm diameter (3 inch) oxidizer, identical to the riser. The oxidizer and riser are a total length of 7 meters from the oxidizer to the cyclone inlet. The expected operating velocity of the riser is 4 to 6 m/s for a total residence time of 1 - 2 seconds.

A classic Lapple cyclone design was used to construct the cyclone for a minimum inlet velocity of 10 m/s (33 feet per second) and a particle density of 4500 kg/m^3. The cyclone is considered a low efficiency cyclone focused on capturing the denser OC and recycling back to the system.

Down Comer

Figure 3-39 is a 3-D image of the downcomer design for returning solids from the cyclone to the reducer. The downcomer consists of a top section known as the volatile converter. The cone reducer contains a coal funnel with a grid plate on the top side for improving volatile conversion as discussed below. Two 3 kW heaters and two 1.75 kW external ceramic heaters are located on the top and bottom sections, respectively. The downcomer was designed to address the limitations of the bench scale unit to simulate the proposed full-scale design of converting volatiles by fluidizing through a moving column of OC.

In the 10 kW_{th} system downcomer, the oxidized oxygen carrier enters the top section via a loop seal located between a cyclone and the downcomer. The coal, fed using a loss-in-weight screw feeder, is fed into a funnel with a grid plate on the top surrounded by hot OC. The coal funnel serves as the coal devolatilization unit with released volatiles flowing up through the grid plate of the funnel into the bed of OC in the top section of the downcomer. This allows for counter current flow between the oxygen carrier and the volatiles and ensures that the volatiles are fully oxidized before exiting through the top exhaust. The bed height of the material in the top section controls the residence time of the volatiles. The coal char remaining in the funnel flows down and out of the funnel by gravity and mixes with the hot and partially reduced OC and is then transported to the reducer's loopseal and subsequently the reducer section.

Reducer

The reducer, Figure 3-40, is a modular spout-fluid bed design consisting of multiple high velocity spouts surrounded by a low velocity annular region. The reducer includes draft tubes located above the high velocity spout nozzles. The high velocity in the draft tubes creates a rising column of ilmenite that falls back in the annulus creating a circular motion that ensures good mixing. The oxygen carrier/coal mixture enters the reducer through an opening between the loop seal and reducer. The Loop seal, reducer and char separator are heated using external ceramic heaters with a total rating of 13 kW. During operation, the fluidization gas mixture (H_2O and CO_2/N_2) is fed through the annulus, and the spouting gas is controlled by nine rotameters. The steam generated in the oxidizer heated zone is regulated using a peristaltic pump. The material exits the reducer at the top of the weir located at the top of the bed before entering the char stripper. The Reducer unit has pressure taps and thermocouples located within the bed are used to determine and maintain bed height.

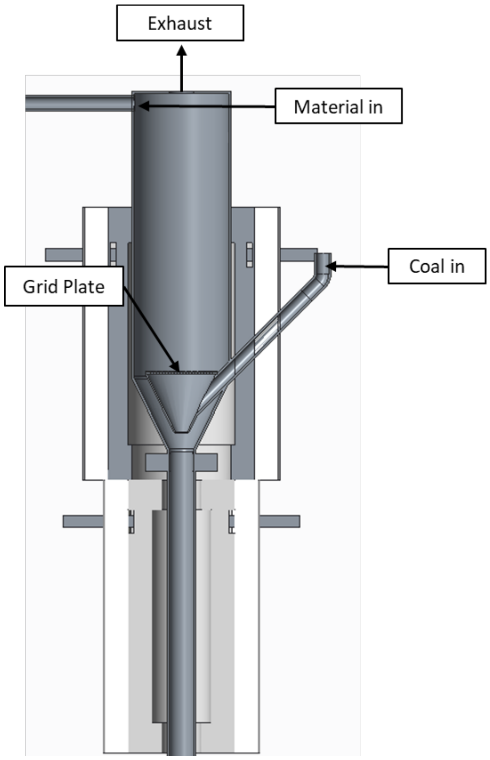

Figure 3-39 Cross-sectional 3D model of down comer

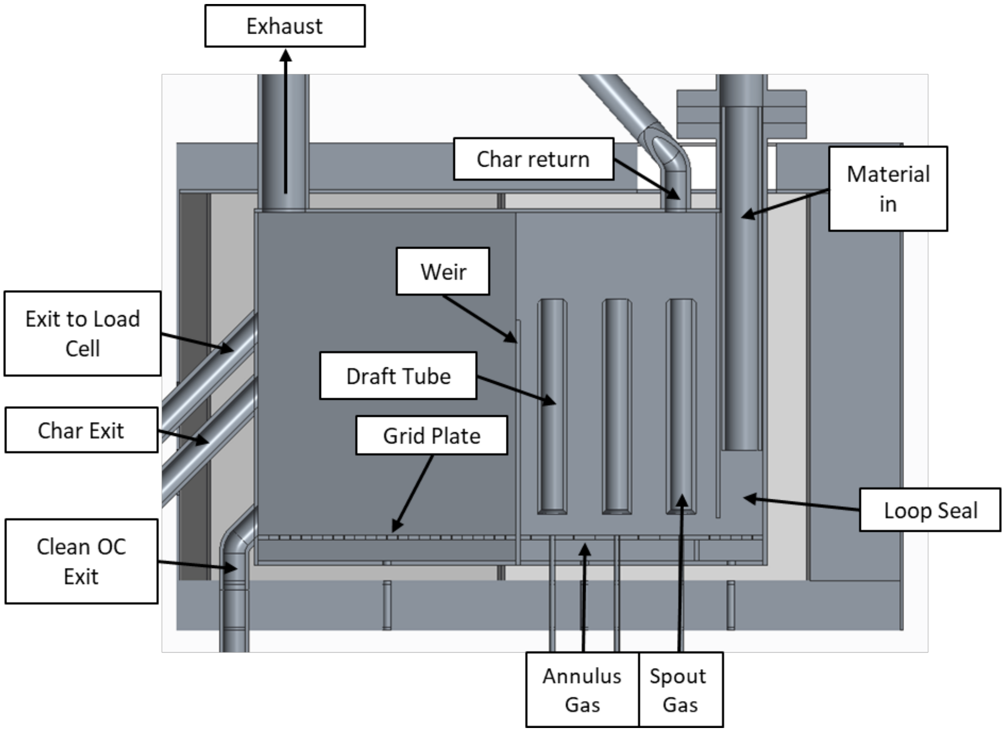

Figure 3-40 Cross-sectional 3D model of reducer/char stripper

L-Valves and Loop Seals

The L-valves and loop seals shown in orange in Figure 3-41 are used to control solid flow and isolate gases between beds. The down comer, reducer, and oxidizer L-valves are all constructed out of 2.5 cm (1-inch) diameter pipe and have a 30 cm (12-inch) long horizontal section. The reducer loop seal, described earlier, connects the downcomer to the reducer, and ensures a uniform feed material distribution into the reducer by having the same width as the reducer. The height of the vertical stand pipe for each loop seal was determined based on the minimum material head needed to maintain the pressure drop requirement between beds. All loop seals were designed to operate optimally at a material feed rate of 150 kg/hr. and are capable of feeding up to 500 kg/hr. Pressure taps located at the top of each loop seal are used to detect material height in the vertical portion of the loop seals in order to ensure that sufficient material head is maintained.

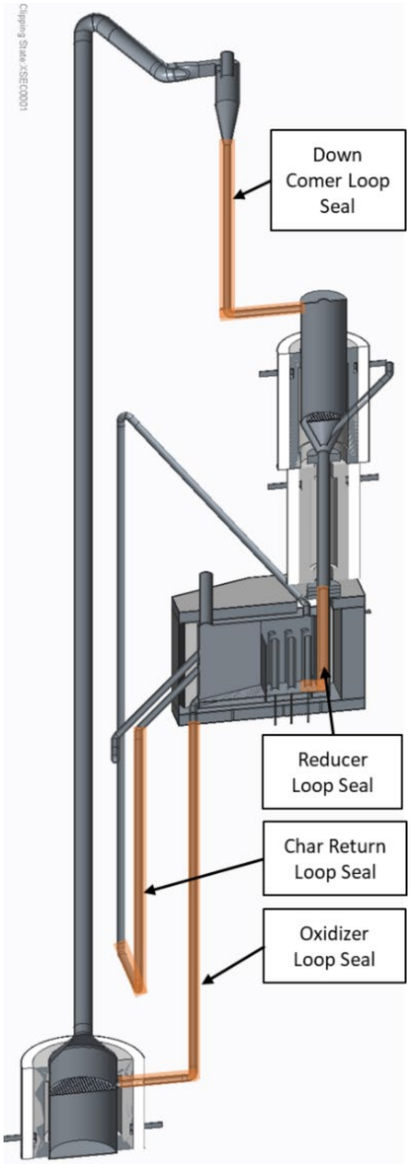

Figure 3-41 Loop seal locations

Commissioning of the 10 kW_{th} System

After installation of the system, multiple shake downs were performed to evaluate operation of the design and to make necessary modifications to account for heat losses and gas flow requirements. Additional instrumentation was added to capture changes in the system during operation as well as additional heating was needed to minimize heat losses from the oxidizer to reducer during low circulation rates. Once final shakedown was completed, performance testing of the system commenced.

Task 5 – Scaled-up OC Manufacturing

This task focused on manufacturing sufficient engineered OC to operate the 10 kW_{th} system. Initial production target of 1000 kg was reduced down to ~200 kg following modifications to the 10-kW_{th} system that reduced inventory requirements. The alternate down-selected OC – FEH31, was selected for scaled up manufacturing due to unsuccessful attempts to secure sufficient waste iron which is the key ingredient of FEL3.

All raw materials for manufacture of FEH31 were procured in sufficient quantities for the manufacture of up to 500 kg of OC. A jet mill system was procured to mill oversized raw materials to the desired particle size. A Lancaster K-1 mixer was leased for pelletization of the oxygen carrier to the desired particle range. Figure 3-42 is an image showing the mixer and jet mill setup. The mean particle size desired for the oxygen carrier is 300 µm. A coarser particle size, Table 3-13 and Figure 3-44, was targeted to minimize losses during OC processing. Approximately 200 kg of FEH31 was prepared using this mixer after final screening.

Figure 3-42 Lancaster K-1 mixer for preparing oxygen carrier formulations

Table 3-13 Target particle size of pelletization process including losses

Size Range (µm)	**Mass (lb)**	**Distribution**
4,760 – 841	**148**	26%
841 - 400	**363**	64%
- 400 (Discarded)	28	5%
Discarded (pan scrapings)	19	3%
Lost	10	2%
Total	**568**	**100%**

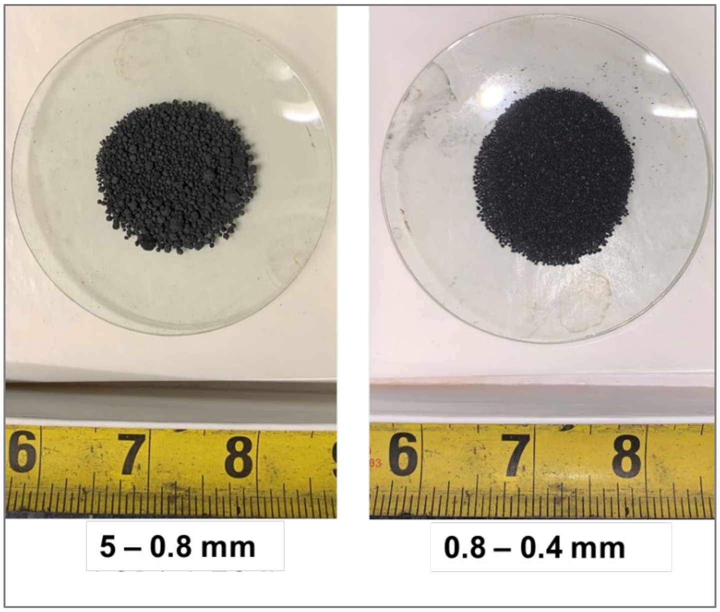

Figure 3-43 Size conditioned FEH31 sample

The final processing step was a curing step using a UNDEERC rotary kiln equipped with a volumetric screw feeder. Preliminary cold flow tests were performed to calibrate feed rates and verify flowability, see Figure 3-44. Figure 3-45 is an image showing samples of the cured material. Final recovery post curing was 162 kg with 23 kg rejected during the curing process.

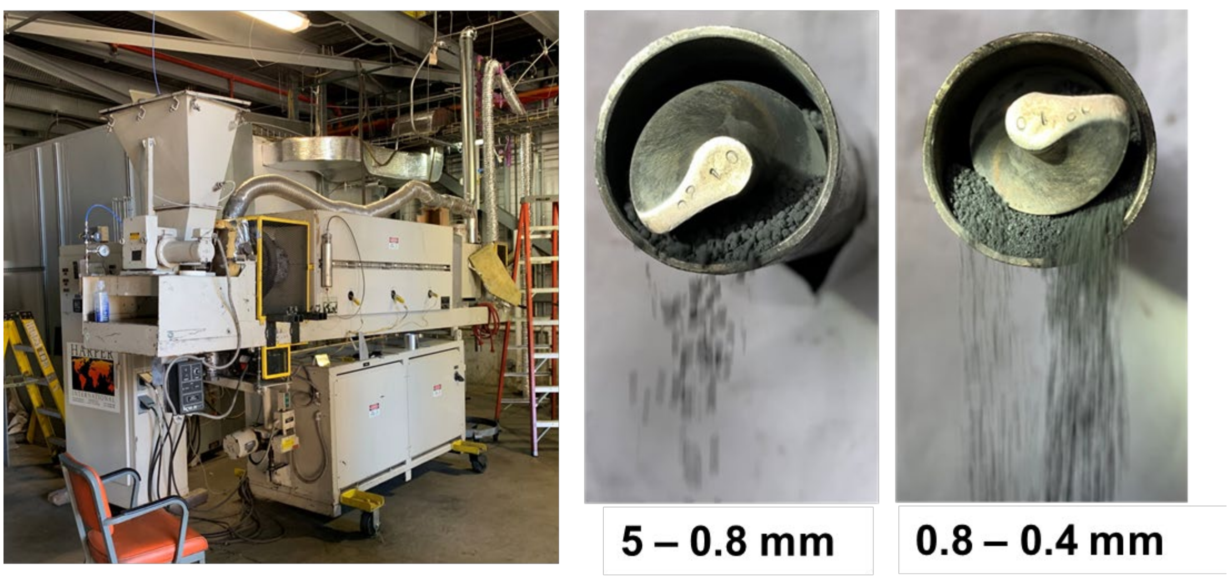

Figure 3-44 Kiln for curing FEH31 material (left) and flowability verification of particle sizes (right)

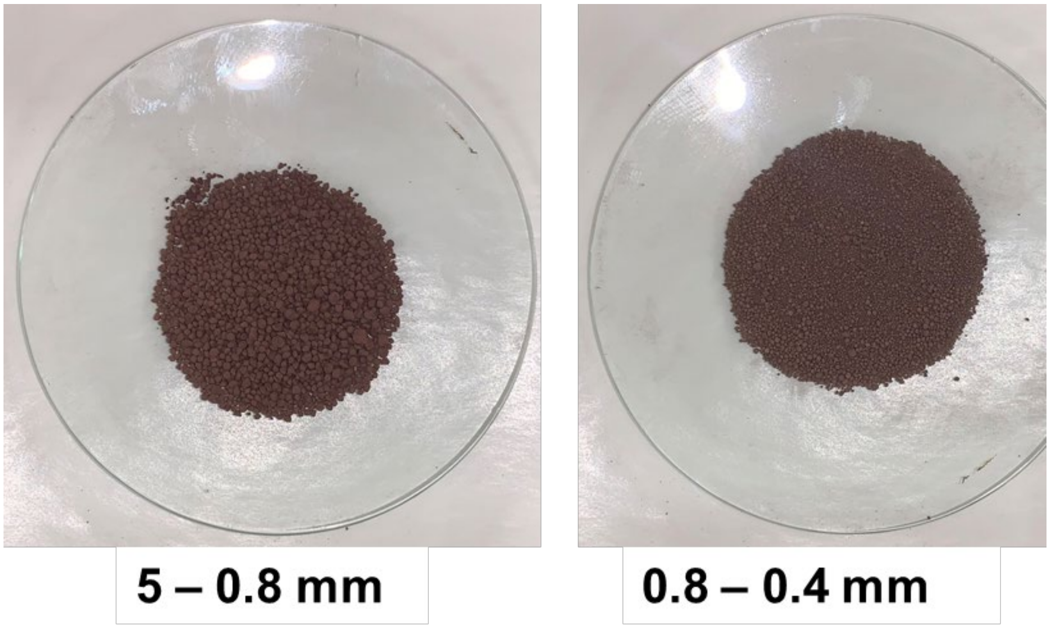

Figure 3-45 Final Prepared FEH31

Once the final cured material was received, the bulk density and fluidization parameters – minimum fluidization and minimum transport velocity, were used to determine the optimal particle size for operation in the 10 kW_{th} system. Table 3-14 shows a summary of the most important material properties. Cold-flow test results showed good agreement with models for both the minimum fluidization based on Wen & Yu correlations cited in Kinii and Levenspiel (1977)[7] and transport velocities based on terminal velocity models determined by Haider & Levenspiel (1989)[8] at 20˚C. These models were used to predict the fluidization and transport velocities for the material at elevated temperatures as indicated in Figure 3-46. Final conditioning of the particles was performed to obtain the desired particle size distribution for the 10 kW_{th} testing.

Table 3-14 10 kW_{th} test material (FEH31) properties

Property	Value
Average particle size (µm)	335
Minimum Fluidization Velocity (cm/s @ 20˚C)	7
Minimum Transport Velocity (cm/s @ 20˚C)	280

[7] Kunii, D., & Levenspiel, O. (1977). *Fluidization engineering*. Huntington, N.Y: R.E. Krieger Pub. Co
[8] Haider, A. and Levenspiel, O. (1989) Drag Coefficient and Terminal Velocity of Spherical and Nonspherical Particles. Powder Technology, 58, 63-70. http://dx.doi.org/10.1016/0032-5910(89)80008-7

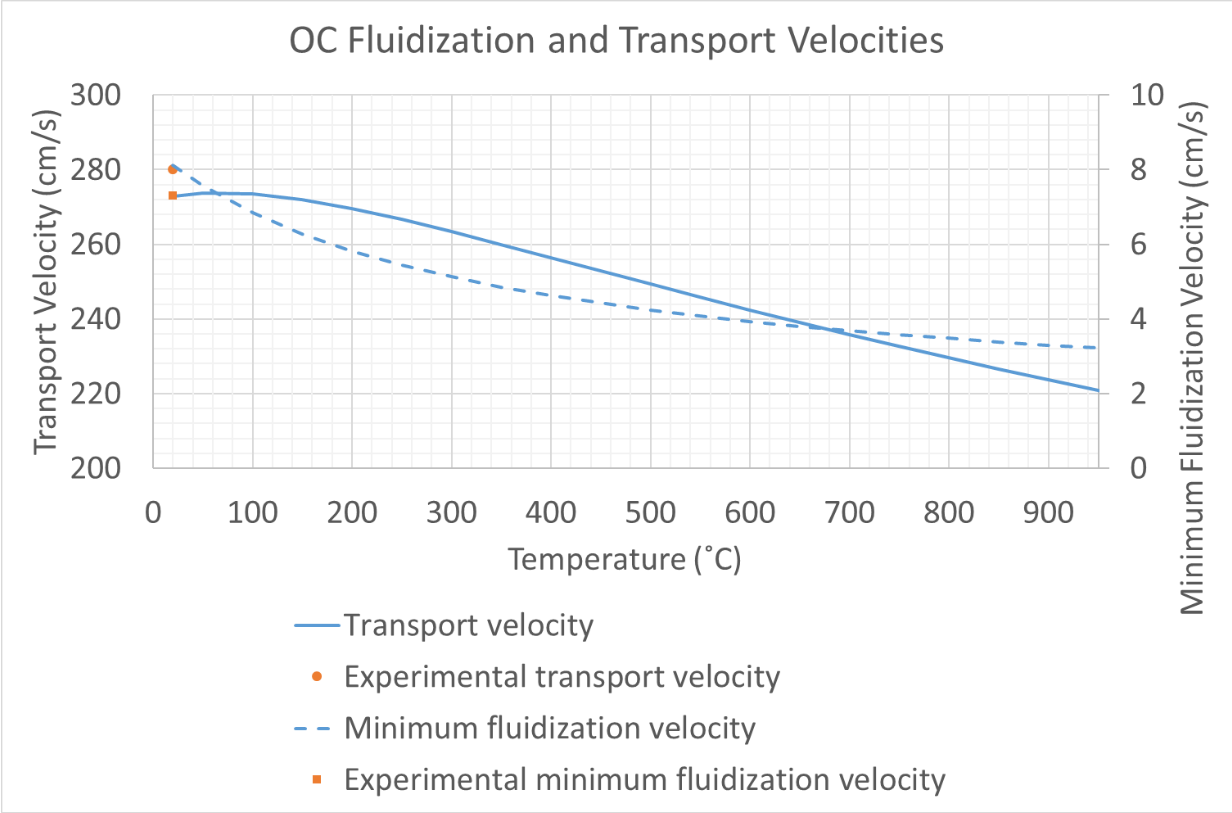

Figure 3-46 Modeled OC fluidization and transport velocities as a function of temperature.

Task 5 Conclusions

Close to 140 kg of oxygen carrier (FEH31) was successfully prepared by scaling up the production process developed by UND and Envergex. The scale-up procedures proved viable and can be implemented at an even larger scale. One key aspect that would need to be addressed for future applications would be better compaction of material during the pelletization equipment did not compact the material as strong as was observed in the smaller laboratory-scale OC manufacturing tests (discussed in Task 2.1). Higher compaction would likely result in a more dense, higher strength OC particle.

Task 6 – 10 kW_{th} Testing

Task 6 focused on testing of the spout fluid bed to verify performance as a fluidized bed reactor. The main objectives of Task 6 testing were 1) Investigate effect of temperature, OC/coal ratio, residence time, 2) compare performance of spout-fluid bed and a fluidized bed hydrodynamics, 3) fuel conversion and 4) recyclability of fines. Three tests were completed and results are summarized in Table 3-15 and Table 3-16. The energy content of the coal is 20,700 MJ/kg (8900 Btu/lb). The coal proximate and ultimate analysis were provided in Table 3-8.

Testing Methodology

The unit was preheated by a combination of electric heaters and a direct propane feed into the oxidizer. It takes approximately 24 hours for the oxidizer to attain set point. The coal was then fed into the downcomer where it flowed co-currently with the OC from the oxidizer. Flue gas from the oxidizer, reducer and the downcomer combined into a polishing chamber to complete combustion of any unconverted volatiles / CO / H_2 / char. Exhaust gases from the polishing chamber are then routed to a baghouse. Two Laser Gas Analyzers (LGA), were used to monitor the gas composition after the polishing chamber and of either the reducer exhaust or oxidizer exhaust. Reducer performance was determined by performing a carbon mass balance on the reducer/oxidizer and the entire system to determine char conversion.

Table 3-15 Coal Specifications for 10 kW Testing

	Coal	Particle Size (µm)	kW_{th}
Test 1	Sub-bituminous, Absaloka coal mine	620	3.7
Test 2		1016	6.5
Test 3			5.9

Table 3-16 Results for Preliminary Testing on 10 kW System

	Test 1	Test 2	Test 3
OC Circulation (lb/h)	1400		700
Coal Feed Rate (lb/h)	1.4	2.5	2.2
Oxidizer			
Temp (°C)	814	844	851
Vel. (m/s)	4.4	4.5	4.0
Gas Flow (slpm)	355	355	315
CO_2 (%)	0.75	1.78	NA
Reducer – Fluidized[1]			
Temp (°C)	773	794	782
Gas Flow (slpm)	40	33	34
U_o / U_{mf}	1.3	1.3	1.3
U_{sp} / U_{mf}	4.6	1.4	1.4
CO_2 (%)	4.3	10.0	9.3
CO (%)	0.52	0.47	0.77
H_2 (%)	<0.1	0.65	0.74
Carbon Conversion	**54%**	**57%**	**62%**

Reducer – Spouting[1]			
Temp (°C)	769	791	778
Gas Flow (slpm)	42	34	34
U_o / U_{mf}	0.4	0.4	0.4
U_{sp} / U_{mf}	14.9	11.1	11.2
CO_2 (%)	4.6	10.4	9.6
CO (%)	0.50	0.47	0.56
H_2 (%)	<0.1	0.67	0.84
Carbon Conversion	**61%**	**57%**	**68%**

[1]The operating conditions for the reducer are reported in multiples of the minimum fluidization (U_{mf}). U_o is annulus operating velocity, U_{sp} is spouting operating velocity.

Results Description

The minimum fluidization (U_{mf}) of the ilmenite was determined to be ~3 cm/s in the reducer at operating temperatures, agreeing with previous results from the batch experiments (Table 3-7). Figure 3-47 shows U_{mf} results obtained during test 2.

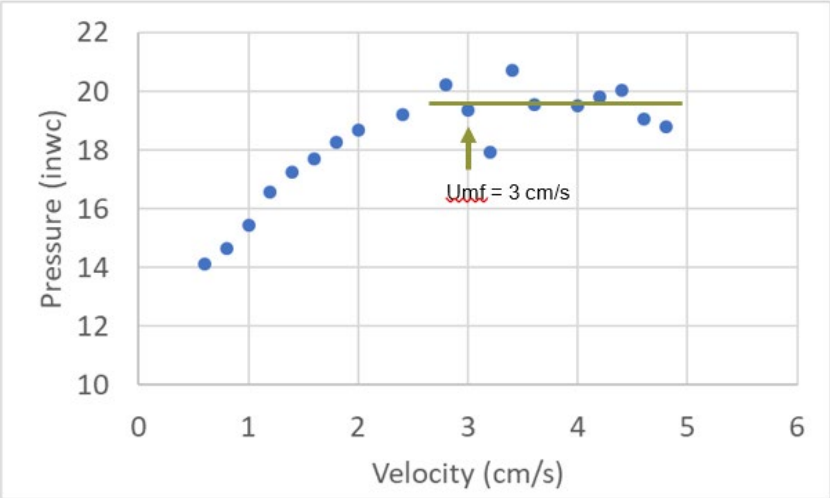

Figure 3-47 Figure showing minimum fluidization of velocity during 10 kW unit testing

Test 1 Gas Composition

Figure 3-48 shows the gas composition during Test 1. Total CO_2% refers to CO_2 concentration after the polishing scrubber and is used to verify the coal feed rate. Region A shows gas composition in the oxidizer and includes the CO_2 from propane combustion. Prior to coal addition, CO_2 concentration from propane alone was ~6%. An increase of ~1% was observed once coal was at steady state.

Region B shows the reducer gas composition under fluidizing conditions. As the analyzer is switched from the oxidizer to reducer, the O_2 composition drops to zero. The reducer gas composition steadily decreases with time, even though the total CO_2 stays constant implying coal/volatiles are bypassing reducer.

Region C shows the reducer after a switch to spouting mode in the reducer. A similar downward trend is observed, along with a small increase in CO_2 concentration.

In both region B and C, unconverted CO and H_2 are detected at compositions of 0.5% and <0.1% respectively. Two factors could explain this – poor ilmenite-char contact if char segregates to top of the bed and low temperatures in reducer affecting conversion rates.

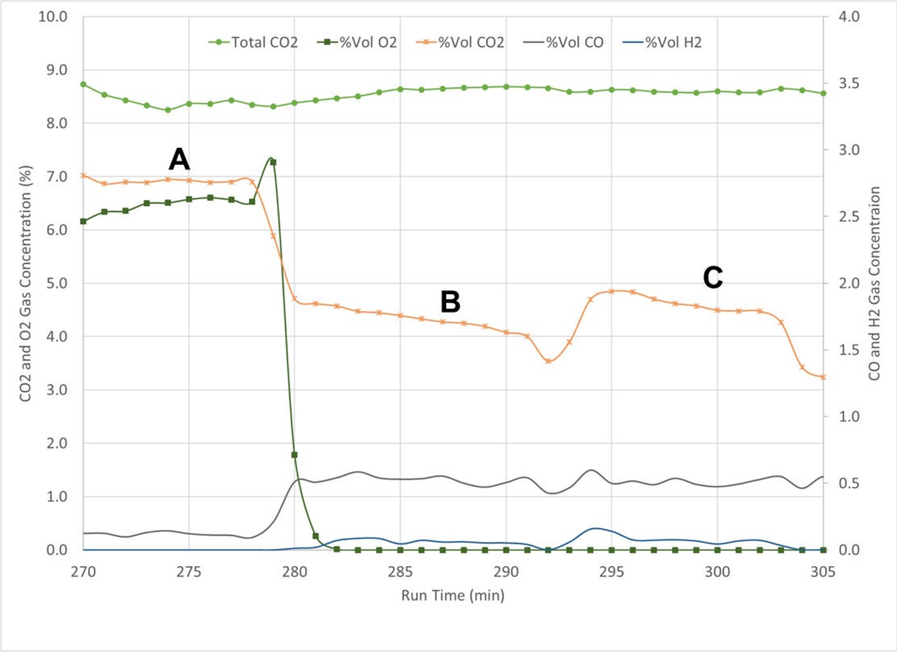

Figure 3-48 **Composition trends of Test 1 including char slip to oxidizer (A), coal conversion in reducer during regular fluidization (B) and spouting (C) bed hydrodynamics**

Test 2 Gas Composition

Figure 3-49 shows the oxidizer gas composition at the start of coal feeding. The CO_2 in the oxidizer includes the propane. Region A is the oxidizer composition before coal. At 140 min, coal feeding starts, and an increase in total CO_2 of the whole system is observed. It takes approximately 4 minutes for CO_2 to be observed in oxidizer, suggesting a residence time of 3 to 4 minutes in reducer and char separator. The CO_2 increase is due to a char slip from the reducer. In region B the reducer is operated in a fluidized mode (150 to 185 min) before a switch to spouting mode (Region C). No significant changes are observed from the plots. Note the %vol CO_2 is the CO_2 composition in the oxidizer including CO_2 from propane combustion.

Figure 3-50 and Figure 3-51 show reducer gas compositions during Test 2 at high and low circulation rates respectively. For Figure 3-50, Region D is the reducer operating in spouting mode and region E shows reducer operating in bubbling bed mode. No significant difference is observed between different bed operation mode. Note, that for both cases, reducer is operated at very low velocities. For Figure 3-51, circulation rate of 700 lb/hr, the oxidizer gas flow is decreased by 10% to achieve the drop in circulation. This results in an increase in oxidizer temperature and corresponding increase in concentration of the total %CO_2 leaving the system (less dilution). In the reducer, the CO_2 concentration should be unchanged thanks to hydraulic isolation of both systems, however, an increase in CO_2 is observed in the reducer. This is calculated as an increase in carbon conversion of 62% and 68%. The improvement is attributed to longer bed residence time. Additionally, the unconverted H_2 and CO drop slightly.

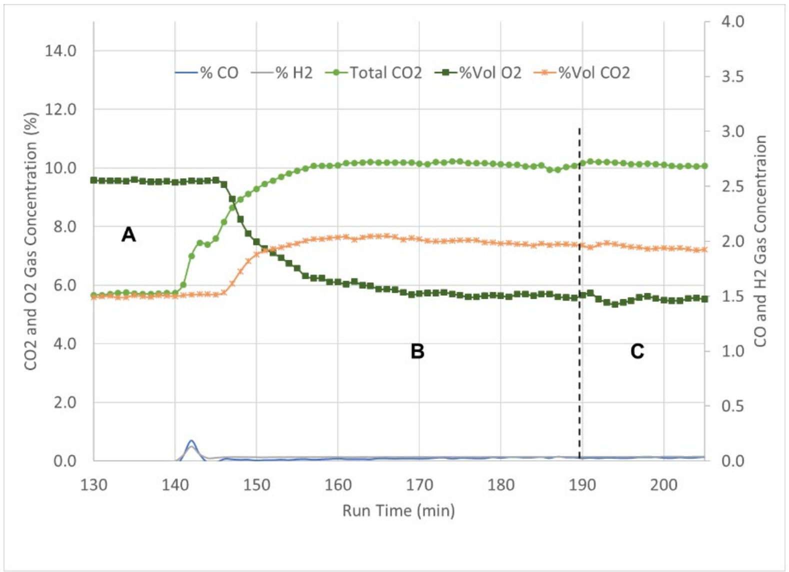

Figure 3-49 **Composition trends of test 2 including start of coal feeding (140 min), steady state operation in fluidizing (B) and spout-fluid (C) bed hydrodynamics**

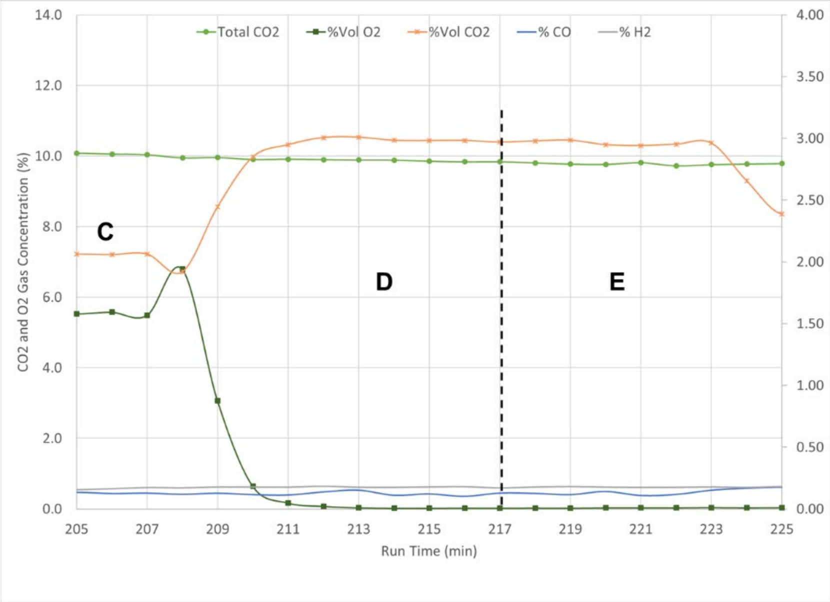

Figure 3-50 **Reducer gas composition at 1400 lb/hr circulation for both bubbling (D) and spouting (E) bed hydrodynamics**

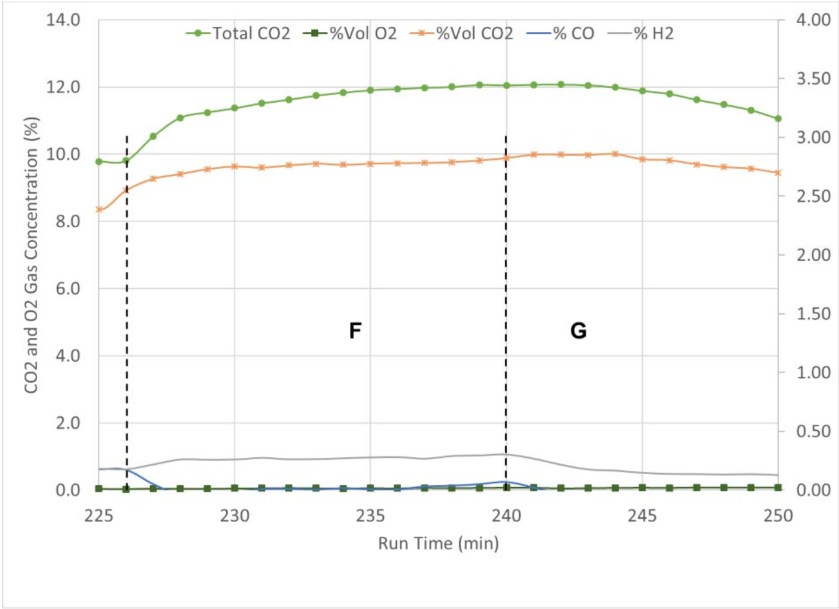

Figure 3-51 **Reducer gas composition at 700 lb/hr circulation for both bubbling (F) and spouting (G) bed hydrodynamics**

Effect of Coal Particle Size and Circulation Rate on Conversion Efficiency

The circulation rate of 1400 lb/hr corresponds to a residence time in the reducer of approximately 3 minutes. The circulation rate of 700 lb/hr corresponds to a residence time of approximately 6 minutes. At steady state, any unconverted char in the reducer slips to the oxidizer. To verify this, at the start of every test, the gas flow to the oxidizer is kept constant, and any change in the oxidizer CO_2 concentration confirms carbon slipping (Figure 3-49).

Test 2 had a higher slip than test 1, 1.78% vs 0.75% respectively (not including propane CO_2 in oxidizer). This is attributed to the difference in particle size of the coal. A larger coal particle size requires longer residence times in the oxidizer for conversion. The char slip for Test 3 was not determined as the unit ran out of coal before this step could be verified.

The carbon conversion efficiency for Test 3 showed the best performance of 68% and 62%. This higher performance is partly explained by previous batch testing (Subtask 3) that showed longer conversion times for larger particles. In the batch experiments, 6 minutes was sufficient for over 90% conversion. In Test 3, even though the residence time is 6 minutes, conversion is lower than 90% because the fuel is coal meanwhile char was used in the batch test. Consequently, incomplete volatile conversion from the coal due to poor solid-gas contacting results in lower conversion efficiency.

Test 2 conversion efficiency (57%) was better than Test 1 (47% and 51%), even though test 2 had a bigger char slip. The smaller particle size of the coal in Test 1 is expected to have resulted in higher carbon losses during de-volatilization in the downcomer and reducer.

Effect of Spouting

During spouting, the annulus region approximates to a moving bed, while the spouts ensure solid circulation and mixing. This results in improved solids contacting and is expected to yield better performance. For Test 1 and 2, a small improvement in coal conversion is noticeable. This supports the claim that the spout fluid bed has improved bed hydrodynamics for solid-solid interactions (heat transfer to coal particle). It is also important to point out that the velocities in question are very low, ~1 U_{mf}, permitting the use of milled (2,000 – 400 μm) and not pulverized (<100 μm) coal particles. Most bubbling beds operate at higher velocities and require the use of larger coal particles.

Alternatively, finer coal particles will require the use of circulating fluidized beds to maintain adequate gas-solid contacting.

Next Steps

Due to scheduling challenges, other factors of interest – effect of steam, char separator performance; and other oxygen carrier FEH31 where not investigated. The manufactured oxygen carrier – FEH31, was loaded into the reactor and circulated, however additional equipment heating limitations impacted the ability to operate the reducer at temperatures over 700°C during project period of performance.

Task 7 – Process Design and Techno-Economic Analysis

This task focused on performing a techno-economic analysis based on NETL's CLC Reference Plant Designs and Sensitivity Studies (NETL, 2014)[9] with data obtained from this project. An Aspen Plus® process model was developed to include the novel components (OC recycling and reformulation) of our process. The technical and economic impact of our novel OC technology were then compared to the baseline cases outlined in the NETL report.

It is essential to know how to produce sufficient quantities of OC and the associated manufacturing costs involved with the production of OC before a viable TEA can be performed. For this reason, we determined the best approach was to conduct two separate TEAs. The first TEA details the results from an assessment of the cost to construct a 1 million tonne per year OC production facility for our novel, low cost material. The second TEA details the results from an assessment of the cost to construct a CLC power plant using the novel, low cost OC.

Project partner Barr Engineering led the completion of this task. The total project cost estimate, divided into 6 different code-of-accounts, corresponds to a Class 5 estimate class (AACE International Recommended Practice No. 18R-97) for the process industries and the range of accuracy of -50- +100% accuracy. Under this costing scenario, the major pieces of equipment were identified, sized, and costed, using vendor quotations when possible. The aggregate purchase cost was used as the basis for all other factors, such as minor equipment, site work, buildings, engineering, construction, and contingency.

Task 7 Conclusions

The results of the TEA are presented in two appendices below, with a summary each respective TEA's highlights included below. Part I is dedicated to the OC manufacturing facility and Part II to the CLC power plant. The two complete TEAs are attached as appendices, with Appendix A containing the OC manufacturing plant TEA and Appendix B the CLC power plant TEA.

Part I – Key findings for the OC manufacturing facility

In this TEA, we decided to assess the cost of OC produced at a 1 million metric tonne per year system. The baseline was chosen to compare the TEA results to NETL's OC production cost (NETL, 2014) for the same sized setup. The OC cost plays an essential role in CLC, and our novel, low cost oxygen carrier can improve the viability of CLC drastically. The uniqueness of our process is vested in the use of low-cost raw materials and the exothermicity of the main production step to decrease the overall OC cost. For this reason, the production process was designed with heat recovery to increase thermal efficiency and to reduce green-house gas emissions. Two separate heat recovery processes are employed: heat exchange for internal heat recycle, and steam generation for captive power requirements. The detailed assumptions, cost breakdowns, and process description are given in Appendix A. The key approach and findings from this TEA were:

- The Total Overnight Costs were found by approximating the capital costs, O&M costs, and owners costs. Based on a finance model from NETL, the Total As-Spent Costs and Total Annual Costs were determined. The sale price required for OC (based on annual OC production) was determined to be approximately **$152/tonne** (with the heat recovery system) and **$149/tonne** (without the heat recovery system).

- Through a sensitivity analysis, the sale price of OC varies most greatly with the O&M costs and the Fixed Rate Charge (FRC). This is affected the most by the cost of the main iron-containing feed streams.

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[9] Keairns, D., Kuehn, N., Newby, R., & Shah, V. (2014). Guidance for NETL's Oxycombustion R&D Program: Chemical Looping Combustion Reference Plant Designs and Sensitivity Studies (No. DOE/NETL-2014/1643). NETL.

Part II – Key findings for the 585 MWe CLC power plant

This TEA developed a CLC power plant design using the novel, low cost OC with performance and cost estimates. The power plant's main design basis is the same as that documented in NETL's CLC Reference Plant Designs and Sensitivity Studies report by NETL (2014)[9]. The TEA study covered the most essential parts of the design basis, and the detailed assumptions, cost breakdowns and process description as given in Appendix B. The key approach and finding from this TEA were:

- By adopting a low cost and recyclable OC, the impact of OC price on electricity price is reduced significantly. The replacement costs of the OC are determined by the attrition rate, amount of attrited OC recycled and OC price. In our process, we assumed 53 percent recycle and an attrition rate of 0.02 percent of solid circulation. In this scenario, the impact of OC cost is negligible; a 100 percent increase in OC price results in less than one percent increase in the electric sale price of $0.104/kWh. The low cost and recyclable process significantly reduces the impact of OC price on the process.

- Addition of an air separation unit for gas polishing in the reducer would result in a 11% increase in electricity prices to $111/MWh.

- Adoption of a spout-fluid bed is feasible per a preliminary high-level evaluation of constructability that focused on comparing reactor volume to current commercial circulating fluidized beds. Adoption of a shorter vessel height compensated for the larger hydraulic diameter.

- The energy penalty for use of 100 percent steam to fluidize the reducer is feasible

The results of this TEA confirm that CLC is a promising carbon dioxide removal technology if the technical challenges involved can be resolved. With key process features such as high OC circulation rates and the interconnected fluidized beds already considered significant technical challenges, it is important to reduce the impact of all other factors on the overall technology risk profile, specifically OC replacement costs and coal conversion performance.

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5 Appendix A – TEA 1 MMTPY OC Facility

Technical Economic Analysis

Oxygen Carrier Production Plant – With Optional Heat Recovery System

Final

Author List:

University of North Dakota
Junior Nasah, Johannes Van der Watt, Mark Musich

Envergex LLC
Srivats Srinivasachar, Aaron Koenig

Barr Engineering Co.

Technical Economic Analysis

Oxygen Carrier Production Plant – With Optional Heat Recovery System

Final

Final Revision by: University of North Dakota

February 2022

UND UNIVERSITY OF NORTH DAKOTA
LEADERS IN ACTION

TECHNICAL ECOMONIC ANALYSIS

Oxygen Carrier Production Plant – With Optional Heat Recovery System

Final

FEBRUARY 2022

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Acronyms

AACE	Association for the Advancement of Cost Engineering
AFUDC	Allowance for Funds Used During Construction
ANSI	American National Standards Institute
ATWACC	After-Tax Weighted Average Cost of Capital
BEC	Bare Erected Cost
BOP	Balance of Plant
CCW	Closed Cycle Water
CLC	Chemical Looping Combustion
CF	Capacity Factor
CRF	Capital Recovery Factors
DOE	Department of Energy
EPC	Engineering, Procurement, and Construction
EPCM	Engineering, Procurement, and Construction Management
FA Fans	Fresh Air Fans
FBR	Fluidized Bed Reactor
FCR	Fixed Charge Rate
FD Fans	Forced Draft Fans
FEED	Front-End Engineering Design
FGD	Flue Gas Desulfurization
HRSG	Heat Recovery Steam Generator
HVAC	Heating, Ventilation, and Air Conditioning
ID Fans	Induced Draft Fans
IOU	Investor-Owned Utility
KW	Kilowatt
LTE	Low Temperature Economizer
MMBTU	One Million British Thermal Units
MW	Megawatt
NETL	National Energy Technology Laboratory
OC	Oxygen Carrier
OEM	Original Equipment Manufacturer
QGESS	Quality Guidelines for Energy Systems Studies
SCR	Selective Catalytic Reduction
SO_2	Silicon Dioxide
STG	Stream Turbine Generator
TASC	Total As Spent Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
UND-IES	University of North Dakota Institute for Energy Studies
ZLD	Zero Liquid Discharge

1 OC Production Process

1.1 Design Basis

The design annual OC production rate at a capacity factor of 0.85 is approximately 990,000 tonne, or, 133 tonne per hour. The production plant will be located in the Gary/East Chicago, Indiana region of the Great Lakes in order to be close to suitable iron ore reserves.

Proprietary binder and additives comprise approximately 20-30 wt.% of the green OC.

1.2 Process Description

A simplified block flow diagram is shown in Figure 1-1. The process consists of four areas: green OC preparation, OC calcination, heat recovery, and particulate and SO_2 control.

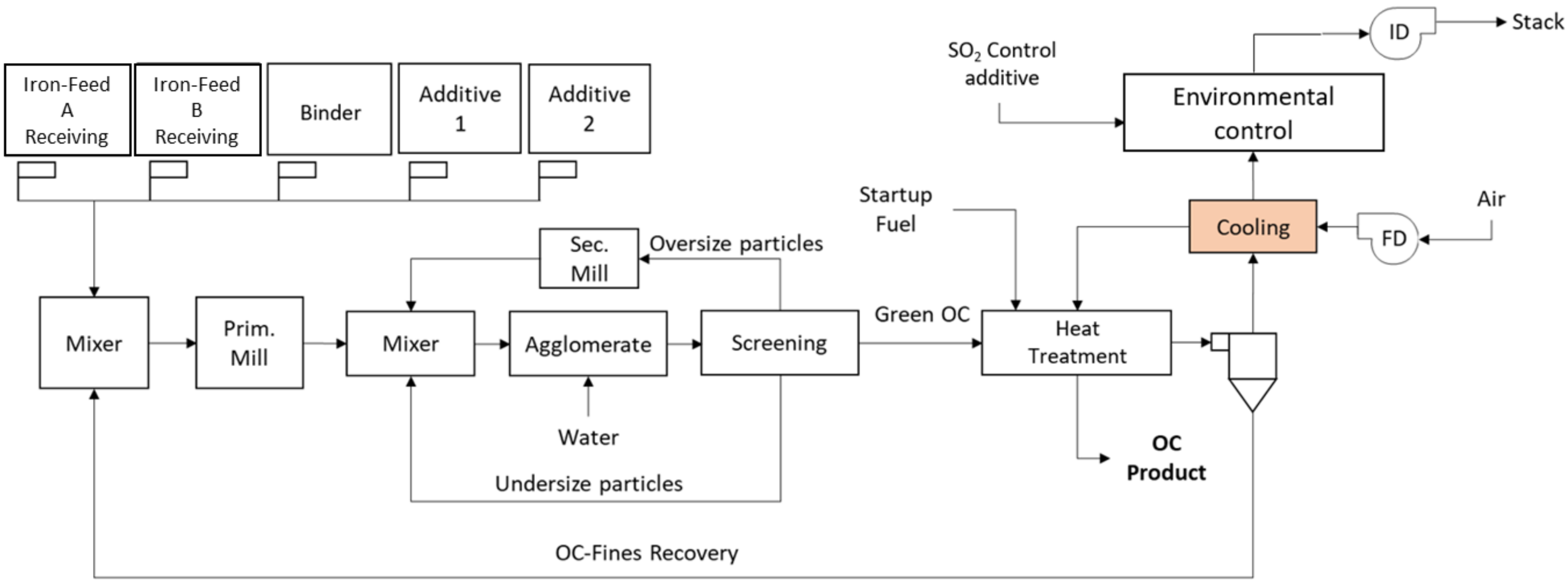


Figure 1-1 Block flow diagram of novel oxygen carrier production process without heat recovery

1.2.1 Green OC Preparation

Principal raw materials (iron-feed A, iron-feed B, binder, and additives) for green OC preparation are stored in silos with 72 hours capacity. Recycle streams from the OC calcination, and particulate and SO_2 control process are blended with the material.

Iron-feed A is fed to the first-stage paddle-mixer (Mixer 1). The nominal Mixer 1 output is 124 tonne per hour at an approximate moisture content of 1 wt.%. Mixer 1 product is discharged to the primary mill (Mill 1), a double roll crusher. The homogeneous, mm top-size product from Mill 1 is blended with recycled green OC screenings (oversize and undersize).

Mixer 2 product is discharged to a drum pelletizer (Granulator) where water is added to facilitate material cohesion and densification. Spherical agglomerates from the Granulator, are discharged to a classifier (Screen 1) to generate a green OC product in the target size range. Undersize green OC is returned directly to Mixer 2. Oversize green OC is processed in a double roll crusher (Mill 2) to produce a size suitable for recycle to Mixer 2. Green OC product from Screen 1, at a mass rate of 134 tonne per hour, is conveyed to OC calcination.

1.2.2 OC Curing

Curing is conducted in a low-velocity, bubbling fluid-bed to minimize elutriation and attrition. 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 Fe_2O_3.

The slightly oxygen depleted flue gas from the Calciner contains OC dust, with the coarse fraction captured in the High-Temperature (refractory-lined) Cyclone. The cyclone underflow solid stream is returned to Mixer 1.

During the calcination the exothermic reaction produces an excess of 35 MW_{th} that will need to be removed to maintain operating temperatures. Two engineering approaches were considered to address this requirement:

- In bed cooling accomplished through internal water pipes. The excess heat will convert the water to steam that will be sent to the Steam Turbine. The calciner equipment adopted was the commercial scale Fluidized Bed Reactor from SCHWING Technologies who proposed a modular design based on their commercial-scale FBR array. Current commercial design of the FBR systems add heat to the process, consequently, limited development work will be needed to modify the system to allow for heat extraction.

- 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.

1.2.3 Heat Recovery

The UND-IES and Envergex LLC's OC production process was designed with heat recovery to increase thermal efficiency and to reduce green-house gas emissions. Two separate heat recovery processes are employed: heat exchange for internal heat recycle, and steam generation for captive power requirements.

The flue gas exits the High-Temperature (refractory-lined) Cyclone and enters Cooler 1, which cools the gas before it enters the gas-to-gas heat exchanger (Heat Exchanger 2).

Through indirect gas-to-gas heat exchange (Heat Exchanger 2), partially cooled flue gas from Cooler 1 preheats fresh air for use in fluid-bed Calciner 1. Flue gas Heat Exchanger 2 effluent is maintained at a temperature high enough to prevent moisture condensation in downstream processes.

A secondary method for heat recovery is the generation of steam from the hot OC discharged from the Calciner. Two methods are proposed. 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 genset, with the resulting power captively used by motors (conveyors, blowers, mixers, mills, pumps, and fans) to support facility electrical and mechanical needs.

1.2.4 Particulate and SO_2 Control

Heat Exchanger 2 effluent flue gas is directed to particulate and SO_2 control. 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.

2 Methodology and Approach

2.1 Cost Estimation Qualifications

The Class 5 constructed cost estimate provided in this report is based on BARR Engineering's experience and qualifications and represents their best judgment as experienced and qualified professionals familiar with the project. This opinion is based on project-related information available to the team at this time, current information about probable future costs, and a concept-level design of the project. The construction cost opinion will likely change as more information becomes available and more design completed. In addition, because there is no control over the eventual cost of labor, materials, equipment, or services furnished by others; the contractor's methods of determining prices; competitive bidding; or market conditions, there is no guarantee that proposals, bids, or actual construction costs will not vary from the opinion of probable construction cost presented in this report. Greater assurance as to the probable construction costs can be achieved through additional design to provide a more complete project definition. Qualifying assumptions and exclusions on which the estimate is based are included in Section 4.5.

The following guidelines were used in evaluation and preparation of this cost report:

- Quality Guidelines for Energy Systems Studies (QGESS).
- The capital and O&M costs have been reported at a level of detail similar to that found in DOE/NETL Baseline studies for Cost and Performance Baseline for Fossil Energy Plants. Volume 1: Bituminous Coal and Natural Gas to Electricity.

2.2 Estimate Type

The cost estimate corresponds to a Class 5 estimate (AACE International Recommended Practice No. 18R-97) for the process industries. This estimate classification is characterized by limited project definition and the wide-scale use of scaling and power industry experience to calculate costs. A Class 5 has an end use for concept screening, with a lower bound accuracy range of -20% to -50% and an upper bound accuracy range of +30% to +100%. These parameters for a Class 5 estimate are in the table below.

Table 2-1 AACE Generic Cost Estimate Classification Matrix

		Primary Characteristic	Secondary Characteristics			
		Level of Project Definition	End Usage	Methodology	Low Range Expected Cost	High Range Expected Cost
Estimate Class	ANSI Classification	Expressed as % of complete project definition	Typical purpose of estimate	Typical estimating method	Typical +/- range relative to best range index	Typical +/- range relative to best range index
Class 5	Order of Magnitude	0% to 2%	Concept Screening	Capacity Factored, Parametric Models, Judgment, or Analogy	-30%- -50%	+30% - +100%

2.3 Cost Estimate Scope

The scope of the cost estimate is completed for a theoretical OC Production Plant with a 16 MW_e Heat Recovery System in the midwestern United States.

The capital cost estimate provided is considered an order of magnitude, or parametric type, estimate with historical/actual cost curves based on historical data from other projects.

The operating and maintenance costs were evaluated using the mass balance calculations from the model. Costs of labor, consumables, and waste disposal were estimated from the "Mineral Commodity Summaries 2020" by the United States Geological Survey (USGS) as well as estimates from similar projects.

2.4 System Code-of-Accounts

Table 2-2 includes a description of the code of accounts used to break down the cost evaluation.

Table 2-2 Description of OC Code of Accounts

Item	Description
1	Bulk Material Storage & Handling
2	Oxygen Carrier Processing
3	Heat Recovery System
4	Environmental Controls
5	Electrical Systems
6	Building & Facilities

2.4.1 Code of Accounts Detailed Breakdown

Class 5 cost estimates are presented for the following construction features required for the project:

1. Bulk Material Storage & Handling
    1. Iron-feed A storage silo (72hrs of storage)
    2. Iron-feed B storage silo (72hrs of storage)
    3. Binder 1 storage silo (72hrs of storage)
    4. Additive 1 storage silo (72hrs of storage)
    5. Additive 2 storage silo (72hrs of storage)
    6. OC Product storage silo (72hrs of storage)
    7. Conveyors (x6)
2. Oxygen Carrier Processing
    1. Mill 1
    2. Mill 2
    3. Mixer 1
    4. Mixer 2
    5. Granulator
    6. Screens
    7. Fluid-Bed Calciner

8. Heat Exchanger
9. High Temperature Cyclone
10. Flue Gas Cooler
11. Belt Conveyors (belt with dust control)
12. Forced Draft Fans

3. Heat Recovery System
   1. OC Solid-to-Air Heat Recovery Heat Exchangers
   2. HRSG/Heat Recovery Boiler
   3. Feed Water Pumps
   4. Fresh Air Fans
   5. 16 MW_e Steam Turbine and Generator
   6. Deaerator
   7. Condenser
   8. Condenser Pump
   9. Cooling Tower

4. Environmental Controls
   1. Baghouse
   2. Packed Bed Scrubber
   3. Instrumentation
   4. Slaker
   5. Sorbent Injection
   6. Forced Oxidizer
   7. Water Treatment / Zero Liquid Discharge System (ZLD)
   8. Induced Draft Fans

5. Electric Systems
   1. Power Auxiliaries
   2. Instrumentation and Control

6. Buildings and Structures
   1. Rigid steel frame with steel roof and walls, foundation and flooring, insulation, electrical, mechanical, HVAC, and plumbing
   2. Site preparation & improvements to site

The summary and detail tables of the total plant cost (TPC) estimate prepared for the project are included in Appendix B.

2.5 Assumptions and Exclusions

Key assumptions are included below and summarized in Appendix C.

2.5.1 Base Case Assumptions

1. 85% capacity factor (O&M Base Case)
2. Greenfield site / Midwest.
3. 100-acre site
4. The power produced by the heat recovery system supplies the OC system and auxiliary loads
5. The substation is provided by the utility company providing additional power (for case without heat recovery)
6. The natural gas supply is assumed to be available at the site boundary
7. Water is available from local municipality
8. Prices may fluctuate due to the varying costs of material and equipment that are driven by multiple market variables. Vendor quotes were either provided in 2021 or in 2019 (and scaled to 2021 dollars). The quotes provided by the vendor may vary over time and as the scope and design becomes more defined.

Base Case Exclusions

The following items are excluded from the project cost estimate:

1. Hazardous, contaminated materials and remediation
    a. Asbestos
    b. Lead abatement
    c. PCBs
    d. Contaminated soils
        i. Contaminated ground water
    e. Site conditions
        i. Piles or caissons
        ii. Rock removal
        iii. Excessive dewatering
        iv. Expansive soil considerations
        v. Excessive seismic considerations
        vi. Extreme temperature considerations
        vii. Demolition or relocation of existing structures
        viii. Unforeseen conditions
        ix. Sub-surface conditions
        x. Existing unknown conditions
    f. Fees and Permits
        i. State licenses
        ii. Local license
        iii. Environmental permits

    iv. Building permits

    v. Third party, professional fees, material testing, and inspections

g. Leasing of off-site land for parking or laydown

h. Busing of craft to site

i. Costs of off-site storage

j. Furnishings and special items

    i. Any furniture, window treatments, or other furnishings

k. Transportation and storage (T&S) is not considered in the capital cost, owner's costs, or O&M. T&S includes items such as:

    i. New access roads and railroad tracks

    ii. Upgrades to existing roads to accommodate increased traffic

    iii. Makeup water pipe outside the "fence line"

    iv. Landfill for onsite waste (slag) disposal

    v. Backup fuel provisions

    vi. Plant switchyard

    vii. Electrical transmission lines outside of plant boundary

    viii. Carbon unloading, sequestration, or transport pipeline

2.6 Cost of Mature Technologies and Designs

The cost estimates of mature technologies and designs are based on vendor quotes procured for this cost estimate. These quotes were used in a capital cost estimate conducted by Barr with a process contingency of 10%. Original equipment manufacturer (OEM) Quotes were obtained for the major equipment listed in Table 2-3.

Table 2-3 List of Major Equipment and Vendors

Equipment	Vendor
Silos and Screen	Pearson Arnold
Dust Control	Donaldson Inc.
FA Fans, FD Fans, and ID Fans	Howden
Steam Turbine (with auxiliaries)	Siemens Energy
HRSG / Heat Recovery Boiler	Hurst
Deaerator	Hurst
OC Solid to Air Heat Recovery Heat Exchangers	Solex
Cooling Tower	Evaptech
Condenser	Babcock Power Thermal Engineering International (TEI)
Condensate Pumps	Somesnick Sales Company, INC.
Circulating Water Pumps	Somesnick Sales Company, INC.
Water Treatment System and ZLD	WesTech

For these readily-available systems, a process contingency of 10% was considered in the cost estimate as shown in Table 2-3. These systems have been proven in full-scale commercial applications. Some unknowns contributing to the uncertainty are water/steam quality, as well as control of this plant to accommodate the high ramp rates and turndowns.

2.7 Costs of Emerging Technologies, Designs, and Trends

There are some areas where the technology is not common or commercially available. Table 4-5 lists one of these areas. The cost was obtained from the OEM for each of these areas. A higher process contingency of 20% is included to account for the emerging technology status as shown in Table 2-4.

Table 2-4 List of Emerging Technologies

Equipment	Vendor
Fluid Bed Calciner	Schwing Technologies
High Temp Cyclone	TBD
Environmental Controls	TBD

While the equipment listed is available on the market, additional engineering costs will be required to integrate the equipment into the proposed concept. These costs are taken into consideration in the TPC. The potential factors which may affect the capital cost of each of these technologies follow:

- ***Fluid Bed Calciner:*** The FBR system proposed by SCHWING Technologies for calcining is commercially available. A higher contingency was proposed due to the higher gas-to-solid ratio requirements for calcining the OC and the need for in-bed heat extraction.
- ***High Temperature Cyclones:*** The high exhaust temperatures and abrasive nature of the OC requires adoption of a high temperature cyclone. Even though cyclone manufacturing is pretty standard, the

operating conditions required are not. Consequently, a higher process contingency was provided to account for operating conditions.

- ***Environmental Controls:*** The scrubbing system itself is a straightforward concept that poses little uncertainty. Factors that are undefined are the levels of SO_2, required limits, ability to oxidize the scrubber water (as lime forced oxidation is a newer process), and make-up water quality going to the zero liquid discharge system (ZLD). These factors will influence the design of the packed bed scrubber, wastewater treatment, and the ZLD.

2.7.1 Project Contingency

Project contingency compensates for cost uncertainties and construction risk associated with final design and construction of the project until the project is completed. Uncertainty in early stages of project planning and design, especially during the feasibility study phase, are greater due to factors such as limited project definition, design and analysis assumptions, unforeseen constraints and constructability issues, construction schedule, and other construction risk factors. In general, uncertainty will decrease as greater definition is developed and more detailed information becomes available.

At this stage in the project, the design is less than 2% complete and constructability has not been evaluated due to insufficient design detail. Therefore, the range of uncertainty of total project cost is high. AACE 16R-90 states that project contingency for a “budget-type” estimate (AACE Class 4 or 5) should be 15% to 30% of the sum of BEC, EPC fees, and process contingency.

The project contingency was determined by taking various percentages of the bare erected costs plus the costs up through process contingency. The project contingency will be reduced as engineering progresses further in later phases and potential further cost reduction with value engineering, standardization, and modularization strategies.

3 Capital Cost Estimate

The Total Plant Cost (TPC) was determined to estimate the project's cost. The TPC is the sum of the Bare Erected Cost (BEC) for the project, plus the cost of the engineering, procurement, and construction (EPC) contractor, as well as process and project contingencies. The TPC is an overnight cost calculated in 2021 dollars.

The BEC consists of the cost of equipment and materials. The major equipment vendors provided Original Equipment Manufacturer (OEM) costs to be used to estimate the BEC. The BEC also contains new onsite facilities, site infrastructure, and balance-of-plant equipment necessary to support the process. The facilities and site infrastructure were estimated based on other Barr-related projects. The direct and indirect construction labor required for installation is included.

The Engineering, Procurement, and Construction Management (EPCM) costs include detailed design, building-related permits obtained by the contractor, as well as project and construction management costs. EPC costs are based on a construction management approach utilizing a prime contractor with multiple subcontractors. This approach provides the owner with greater scope control and flexibility, while mitigating the risk premium typically included in a traditional EPC lump-sum pricing structure.

3.1 Quantities and Allowances

High-level quantity takeoffs for major system components such as sorbent regeneration, flue gas clean-up, and conveyance systems were developed from the general process and instrument diagrams.

3.2 Escalation

Escalation was not considered for the TPC. Therefore, the TPC is in the current dollar amount (for 2021).

3.3 Labor Cost Basis

The estimate was not adjusted for local area labor rates. Labor rates reflect a burden rate, including: worker's compensation, state and federal unemployment taxes, fringe benefits, medical insurances, and other typical burdens. The labor rates are based upon a work week of 40 hours: 8 hours per day / 5 days per week. The average labor rate is $100 per hour with burdens.

3.4 Freight and Shipping Costs

The estimate provided includes freight and shipping cost, duties for all major items of equipment.

3.5 Contingency

Contingency represents an allowance to cover unknowns, uncertainties, and/or unanticipated conditions that are not possible to evaluate adequately from the information at hand at the time the cost estimate is prepared but must be represented by a sufficient cost to cover the identified risks. Contingency relates to a known defined project scope and is not used to predict future project scope or schedule changes. Contingency will normally decrease as more design information is known. This section summarizes important cost-estimating considerations related to cost contingency.

Contingencies, as used in this estimate, are intended to help identify an estimated construction cost amount for the items included in the current project scope. The contingency percentage includes process contingency and project contingency. These contingency amounts are based on AACE guidelines and professional judgment considering the level of design completed, the complexity of the work, and uncertainties in quantities and unit prices. The contingency includes the estimated cost of ancillary items not currently identified in the quantity estimates and allowances, but commonly identified in more detailed design and required for completeness of the work.

Contingencies are assigned to the cost estimate of each project feature based on engineering judgment and on the relative completeness of project definition. Contingency, as used in this cost estimate, will decrease with future design efforts.

The contingency provided with the estimate does not account for:

- changes in labor availability or productivity
- delays in equipment deliveries
- changes in current industry standards or regulations
- major changes in quantities
- major changes in unit pricing
- major changes in scope during detailed design or construction
- major changes or revisions to the design basis
- costs that may result from actual site conditions differing from generic site conditions assumed in this estimate
- costs that result from construction change orders
- costs that result from sequencing or expediting work to avoid critical path slippage
- costs that result from possible project schedule slippage
- costs that result from differing economic conditions or future cost growth
- costs related to plant performance during and after start-up
- force majeure

The contingency included in the cost estimate is based upon the Risk Management or Estimating Judgement process.

3.5.1 Process Contingency

Process contingency provides for uncertainty in the cost estimate related to the technology's maturity development. The configuration of the key technology pieces of this concept is currently unproven at the commercial scale in manufacturing facility applications. However, many aspects of the project use current proven and accepted technology for balance-of-plant and structural aspects. Therefore, process contingencies are applied to individual aspects of the cost estimate based on the current status of the technology for individual aspects of the cost estimate. AACE recommends the following guidelines in Table 3-1 for process contingency to apply.

Table 3-1 AACE Guidelines for Process Contingency

Technology Status	Process Contingency (% of Associated Process Capital)
New concept with limited data	40+
Concept with bench-scale data	30-70
Small pilot plant data	20-35
Full-sized modules have been operated	5-20
Process is used commercially	0-10

Process contingencies used in this estimate were assigned as shown in Table 3-2.

Table 3-2 Process Contingency for Technology

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

While the equipment listed is readily available on the market, additional engineering costs will be required to integrate the equipment into the proposed concept. These costs are taken into consideration in the TPC. The potential factors which may affect the capital cost of each of these technologies are discussed in Section 2.7.

3.5.2 Project Contingency

Project contingency, as described in Section 4.7, was applied to all code of accounts. Project contingencies used in this estimate were assigned as shown in Table 3-3.

Table 3-3 Project Contingency for Technology

Technology	Project Contingency (% of Associated Process Capital)
Bulk Material Storage & Handling	20%
Oxygen Carrier Processing	20%
Heat Recovery System	20%
Environmental Controls	20%
Electrical Systems	20%
Building & Facilities	20%

3.6 Capital Cost Results

The TPC cost for the OC system is summarized in Table 3-4.

Table 3-4 Capital Cost Summary

Item	Category	Bare Erected Cost (BEC) ($)	Engineering, Procurement & Construction ($)	Process Contingency ($)	Project Contingency ($)	Total Project Cost (TPC) ($)
1	Bulk Material Storage & Handling	2,984,000	448,000	344,000	755,000	4,531,000
2	Oxygen Carrier Processing	49,954,000	7,494,000	9,845,000	13,457,000	80,750,000
3	Heat Recovery System	35,892,000	5,383,000	4,129,000	9,082,000	54,486,000
4	Environmental Controls	27,944,000	4,192,000	6,427,000	7,712,000	46,275,000
5	Electrical Systems	21,628,000	3,245,000	2,487,000	5,472,000	32,832,000
6	Building & Facilities	19,914,000	2,988,000	2,290,000	5,038,000	30,230,000
	Total with Heat Recovery (2021)	158,316,000	23,750,000	25,522,000	41,516,000	249,104,000
	Total without Heat Recovery (2021)	143,560,000	21,537,000	23,824,000	37,782,000	226,703,000
	Total with Heat Recovery (2024)	**170,506,000**	**25,579,000**	**27,487,000**	**44,713,000**	**268,285,000**
	Total without Heat Recovery (2024)	**154,614,000**	**23,195,000**	**25,658,000**	**40,691,000**	**244,159,000**

4 Owner's Costs

The owner's costs were estimated by factoring the values provided in the B12B case in the NETL report. This report estimated the costs based on the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." In this document, the total owner's costs consist of preproduction (startup) costs, inventory capital, land, financing cost, and other owner's costs. Prepaid royalties and working capital are not included in the owner's costs.

The preproduction costs include six months of operating labor, six months maintenance materials at full capacity, one-month non-fuel consumables at full capacity, one-month waste disposal, 25% of one month's fuel cost at full capacity, and 2% of TPC. The six months of operating labor includes the cost of training the plant operators, participation in startup, and occasionally involving them in the design and construction of the power plant.

The inventory capital includes 0.5% of the TPC for spare parts, a 60-day supply (at full capacity) of fuel, and a 60-day supply (at full capacity) of non-fuel consumables (e.g., chemicals and catalysts) that are stored on site. The cost for a 60-day supply (at full capacity) of fuel is not applicable for natural gas. The 60-day supply (at full capacity) of non-fuel consumables does not include catalysts and adsorbents that are batch replacements (such as selective catalytic reduction catalysts).

The cost of land includes a 100-acre site with a $3,000/acre price (based on the site being located in a rural area).

The financing cost is based on 2.7% of the TPC and covers the cost of securing financing, fees, and closing costs. It does not include interest during construction (or AFUDC).

Other owner's costs are estimated using 15% of the TPC. This includes:

1. Preliminary feasibility studies (including a Front-End Engineering Design (FEED) study)
2. Economic development (costs for incentivizing local collaboration and support)
3. Construction and/or improvement of roads and/or railroad spurs outside of site boundary
4. Legal fees
5. Permitting costs
6. Owner's engineering (staff paid by owner to give third-party advice and to help the owner oversee/evaluate the work of the EPC contractor and other contractors)
7. Owner's contingency (sometimes called "management reserve"—these are funds to cover costs relating to delayed startup, fluctuations in equipment costs, unplanned labor incentives in excess of a five-day/ten-hour-per-day work week; owner's contingency is not a part of project contingency)

The owner's costs do not include:

1. EPC risk premiums (costs estimates are based on an EPCM approach utilizing multiple subcontracts, in which the owner assumes project risks for performance, schedule, and cost).
2. Transmission interconnection: the cost of interconnecting with power transmission infrastructure beyond the plant busbar.
3. Taxes on capital costs: all capital costs are assumed to be exempt from state and local taxes.
4. Unusual site improvements: normal costs associated with improvements to the plant site are included in the BEC, assuming that the site is level and requires no environmental remediation; unusual costs associated with the following design parameters are excluded: flood plain considerations, existing soil/site conditions, water discharges and reuse, rainfall/snowfall criteria, seismic design, buildings/enclosures, fire protection, local code height requirements, and noise regulations.

The factors used to adjust the costs were taken from the 2019 revision of the QGESS document "Capital Cost Scaling Methodology: Revision 4 Report."

4.1.1 Owner's Cost Results

The Owner's costs for the OC with and without the heat recover system are summarized in Table 4-1.

Table 4-1 Owner's Costs

Owners Costs		
Description	**Cost (with HR)**	**Cost (without HR)**
Pre-Production Costs		
6 Months All Labor	$7,289,000	$6,282,000
6 Month Maintenance Materials	$906,000	$870,000
1 Month Non-fuel Consumables	$7,972,000	$8,131,000
1 Month Waste Disposal	$59,000	$59,000
25% of 1 Month's Fuel Cost at 100% CF	$170	$30
2% of TPC	$5,366,000	$4,883,000
Total	**$21,592,000**	**$20,225,000**
Inventory Capital		
0.5% of TPC (Spare Parts)	$1,341,000	$1,221,000
60-day Supply of fuel at 100% CF	$1,300	$250
60-day Supply of consumables at 100% CF	$15,715,000	$16,028,000
Total	**$17,057,300**	**$17,249,000**
Land		
Cost (Based on 100 Acres)	$300,000	$300,000
Total	**$300,000**	**$300,000**
Financing Cost		
2.7% of TPC	$7,244,000	$6,592,000
Total	**$7,244,000**	**$6,592,000**
Other Costs		
15% of TPC	$40,243,000	$36,624,000
Total	**$40,243,000**	**$36,624,000**
Total Overnight Costs (TOC)	**$354,721,000**	**$294,789,000**
TASC/TOC Multiplier (IOU, high-risk, 5 year)	1.242	1.242
Total As-Spent Cost (TASC)	**$440,563,000**	**$366,128,000**

5 Operation and Maintenance Costs

The yearly operating and maintenance costs associated with the proposed power plant were calculated. The main components of the yearly operating cost are:

- Operating labor
- Maintenance material and labor
- Administrative and support labor
- Consumables
- Waste handling
- Co-products and saleable by-products
- Fuel

The operating and maintenance labor is estimated using methods similar to those contained in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." However, consumable produce prices were taken from "Mineral Commodity Summaries 2020" by the United States Geological Survey (USGS).

5.1 Auxiliary Power Consumption

A full load case based on auxiliaries was not completed but it is estimated that the power produced by the heat recovered from the produced OC excess heat would cover or exceed the load required for the auxiliary power consumption. Therefore, the auxiliary power consumption does not represent a financial cost to the project. A more detail analysis can be completed once design and value-added engineering is added to the process.

5.1.1 Operating Labor

The OC system will require highly skilled operating and maintenance personnel. Personnel will be required to understand the requirements for:

- Iron-feed A handling and processing
- Steam turbine and auxiliaries
- Environmental Controls
- Water treatment and ZLD

It is assumed that the number of personnel at this manufacturing facility with integrated power generation will be similar to manufacturing facilities of similar size and complexity. For this plant, the personnel includes: one plant manager, one operations manager, one maintenance engineer, one senior engineer, one junior engineer, one engineering technician, two financial accountants, two procurement and warehouse managers, two control room operators per shift, five outside operators per shift, two boiler operators per shift, two train unloading operators at two shifts per weekday, three maintenance mechanics, one I&C technician, two maintenance electricians, four general laborers, and one full-time security person. The fully burdened rates are based on estimated costs associated with an employee. This includes salary, benefits, overhead, and other costs.

5.1.2 Maintenance Material and Labor

Maintenance materials were also estimated using similarly sized projects. The maintenance required throughout the plant involves:

- Annual outages to service the steam turbine, calciner, and other major components
- Outages to inspect and maintain the steam turbine and generator
- Maintenance of the iron-feed A and limestone handling equipment, such as conveyers, crushers, mills, and dust collectors

- Occasional maintenance of the ZLD and water treatment system components, including vapor compressors, centrifuge, and demisting pads
- Maintenance of the packed bed-scrubber, including seal and nozzle replacements
- Maintenance of the pumps, heaters, and BOP
- Improvements to the buildings, pavement, and railing system
- Spares

5.1.3 Consumables

Consumable rates were provided by heat, water, and mass balances. The estimated cost of these consumables was derived from "Mineral Commodity Summaries 2020" by the United States Geological Survey (USGS) as well as factoring based on costs of consumables provided in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." Table 5-1 provides the estimated prices of the elements required for the OC process.

Table 5-1 Estimated Consumables Prices

Stream	Consumable	Price ($/tonne)
1	Iron-feed A	$100
2	Iron-feed B	$30
3	Binder	$300
4	Additive 1	$150
5	Additive 2	$130

5.1.4 Waste Disposal

Waste production rates were provided by equipment vendors or calculated from the heat, water, and mass balances. The cost estimate for removing or disposing of waste was derived from factoring based on costs of consumables provided in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance."

5.1.5 Co-Products and Saleable By-Products

Co-products and by-products production rates were provided by calculated values from the heat, water, and mass balances.

The forced oxidizer would produce gypsum that can be used in many applications including fluxing agencies, fertilizers, fillers in paper and textiles, and retarder in Portland cement. It is assumed that this can be sold at $35/tonne.

The value engineering was not considered for this project based on the progress in technology and current economic considerations.

5.1.6 Fuels

The consumption of natural gas for the fluid bed calciner and HRSG startup is based on calculations and vendor quotes. The price of natural gas is assumed to be $3.30/ MMBTU based on average 2021 natural gas prices (from Henry Hub Natural Gas Spot Price).

5.2 O&M Cost Results

The operating and maintenance costs for the HGCC system are summarized in Table 5-2. The resulting O&M costs are approximately $111,000,000 per year.

Table 5-2 Operating and Maintenance Summary

O&M Costs		
	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**

6 OC Sale Price

The method for calculating the OC sale Price is based on the model and methods described in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." This report makes assumptions provided in Section 8.1. This is used to develop the finance structure in Section 8.2. Both are used to calculate the required OC sale price in Section 8.3.

6.1 Global Economic Assumptions

The 2019 revision of the QGESS document "Cost Estimation Methodology for NETL" makes the following assumptions:

1. Taxes
    a. The Federal Income Tax Rate is 21%, the State Income Tax Rate is 6%, and the Effective Tax Rate (ETR) is 25.74%.
    b. Capital depreciation over 20 years is 150% (declining balance).
    c. There is no Investment Tax Credit.
    d. There is no Tax Holiday.
2. Contracting and Financing Terms
    a. The Contracting Strategy consists of Engineering Procurement Construction Management (owner assumes project risks for performance, schedule, and cost).
    b. Debt Financing is Non-recourse (collateral that secures debt is limited to the real assets of the project).
    c. The Repayment Term of Debt is equal to operational period in formula method.
    d. There is no grace period on debt repayment.
    e. There is no debt reserve fund.
3. Analysis Time Periods
    a. The capital expenditure period is 3 years for natural gas plants and 5 years for coal plants.
    b. The operational period is 30 years.
    c. The economic analysis period is 33 years for natural gas plants or 35 years for coal plants (capital expenditure period plus operational period).
4. Treatment of Capital Costs
    a. The capital cost escalation during the capital expenditure period is 0% real (or 3% nominal).
    b. The distribution of Total Overnight Capital over the capital expenditure (before escalation) is 10%, 60%, 30% for a 3-year period and 10%, 30%, 25%, 20%, 15% for a 5-year period.
    c. There is no working capital.
    d. 100% of the Total Overnight Capital depreciates (actual amounts are likely lower and do not influence results significantly).
5. Escalation of Operating Costs and Revenues
    a. Escalation of revenue, O&M Costs is 0% real (3% nominal).

b. Fuel costs are based on the Quality Guidelines for Energy Systems Studies Fuel Prices for Selected Feedstock in NETL Studies.

6.2 Finance Structure

To evaluate the economic feasibility of the project, a financial structure is established based on market and ownership risks. The cost analysis is developed for both commercial technology in 2021 and advancing technology projected to become commercial in 15 years or more. It can be assumed that they are commercially ready and that there are no risks or tax subsidies associated with any of the technology. The same structure should use real dollars and be applied to all scenarios to compare the technologies. Nominal dollars should be used to evaluate the technologies in various cash flow analyses. The structure will assume a large, financially stable, investor-owned utility (IOU) or merchant plant.

Table 6-1, taken from the 2019 revision of the QGESS document “Cost Estimation Methodology for NETL Assessment of Power Plant Performance,” provides the rates that are used for the financial structure. The nominal case will be used throughout the cost analysis. The nominal case for an IOU estimates an equity of 10%.

Table 6-1 Nominal and Real Rates Financial Structure for Investor-Owned Utility

Type of Security	% Total	Current Dollar Cost	Weighted Average Cost of Capital	After-Tax Weighted Average Cost of Capital
Nominal				
Debt	55%	5%	2.75%	2.04%
Equity	45%	10%	7.25%	4.50%
Total			**10%**	**6.54%**
Real (based on 2.01% average real GDP deflator, 1990-2018)				
Debt	55%	2.94%	1.61%	1.20%
Equity	45%	7.84%	3.53%	3.53%
Total			**5.14%**	**4.73%**

The TASC is expressed in mixed-year, current or real dollars over the entire capital expenditure period. It is calculated from the total overnight cost (TOC) by using the following factors taken from the 2019 revision of the QGESS document “Cost Estimation Methodology for NETL Assessment of Power Plant Performance” shown in Table 6-2. The TASC/TOC chosen for this study was a nominal three-year ratio.

Table 6-2 TASC/TOC Factors

Finance Structure	BBB+[3] or Higher Company	
Capital Expenditure Period	Three Years	Five Years
TASC/TOC *nominal*	1.242	1.289
TASC/TOC *real*	1.093	1.154

The TOC includes "overnight" depreciable and non-depreciable capital expenses that are incurred during the capital expenditure period. This does not include escalation and interest during construction. The factor of TASC to TOC is calculated by adding the cost of escalation to the cost of funding.

Table 6-3 Fixed Charge Rate

Finance Structure	IOU – 30 Years	
Capital Recovery Periods	Three Years	Five Years
FCR Nominal	0.0886	0.0886
FCR Real	0.0707	0.0707

6.3 Selling Price

The annual cost of OC production with heat recovery is shown in Table 6-4. Based on an annual OC production of approximately 990,000 tonnes, the estimated sale price (for 10% equity, 30-year operation) is $152/tonne with heat recovery and $149/tonne without heat recovery.

Table 6-4 Annual Cost of Production and OC Sale Price

Cost of Production		
	With HR	Without HR
Plant Capacity, %	85	85
Total Annual Operation, (hrs.)	7,451	7,451
Total As Spent Cost (TASC), ($)	440,563,000	366,128,000
Fixed Rate Charge (FRC), ($)	0.0886	0.0886
First Year Capital Charge, ($)	39,034,000	32,439,000
First Year Fixed Operating Costs, ($)	17,786,000	15,736,000
First Year Variable Operating Costs, ($)	93,596,000	99,067,000
First Year Fuel Costs, ($)	8,000	1,500
Total Annual Cost, ($)	150,424,000	147,244,000
Annual OC Production, tonne	989,000	989,000
OC Sale Price ($/tonne)	152	149

7 Risk Factors

7.1 Risk Factors

A risk analysis is currently being evaluated for the OC production system and will highlight the areas where the total plant cost is most probable.

As discussed in Section 1 of this report, the contingencies of areas that are considered emerging technologies include higher-process contingencies and, in some areas, engineering compared to the common commercialized technologies. We also included costs for several systems noted in the risk management discussions. The following list describes a summary of cost considerations based on risk management:

- High temperatures and flow rates in currently existing equipment that require evaluation or possible pilots (ex. high-temp fluid bed calciner)
- Unknow purities of consumables that could affect process efficiency
- Unknown water quality that could affect ZLD design
- Levels of SO_2 that increase size and cost of environmental controls

8 Sensitivity Analysis

8.1 Effects of Capital Costs

Figure 8-1 and Figure 8-2 provide illustrations of how the sale price of OC varies with capital costs. The sale price of OC varies from $110/tonne to $230/tonne when capital costs vary from $0 to approximately $800 million.

Figure 8-1 OC Sale Price Sensitivity Based on Capital Cost with Heat Recovery

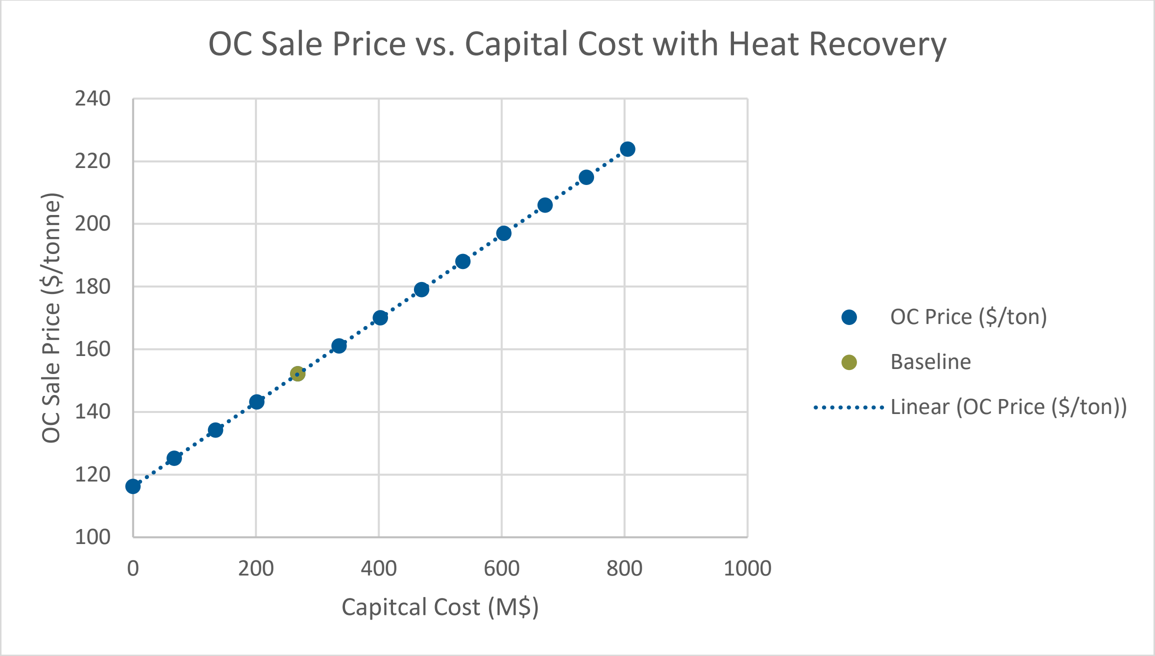
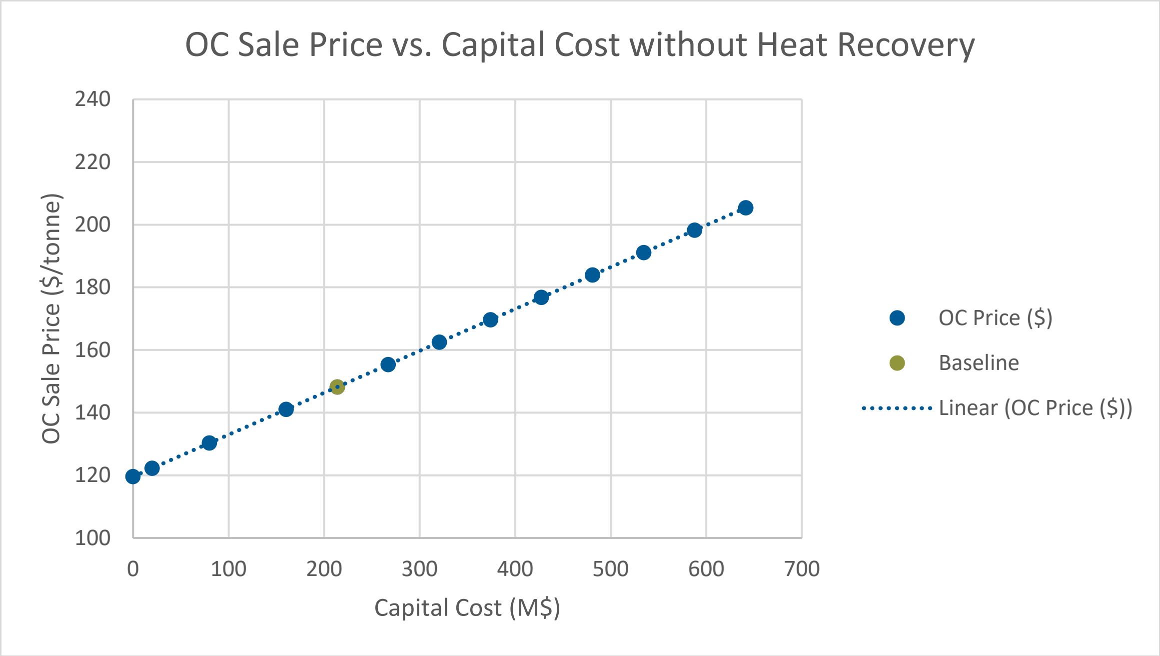


Figure 8-2 OC Sale Price Sensitivity Based on Capital Cost without Heat Recovery



8.2 Effects of O&M Costs

Figure 8-3 and Figure 8-4 provide illustrations of how the sale price of OC varies with O&M costs. The sale price of OC varies from $30/tonne to $390/tonne when O&M costs vary from $0 to approximately $350 million.

Figure 8-3 OC Sale Price Sensitivity Based on O&M Cost with Heat Recovery

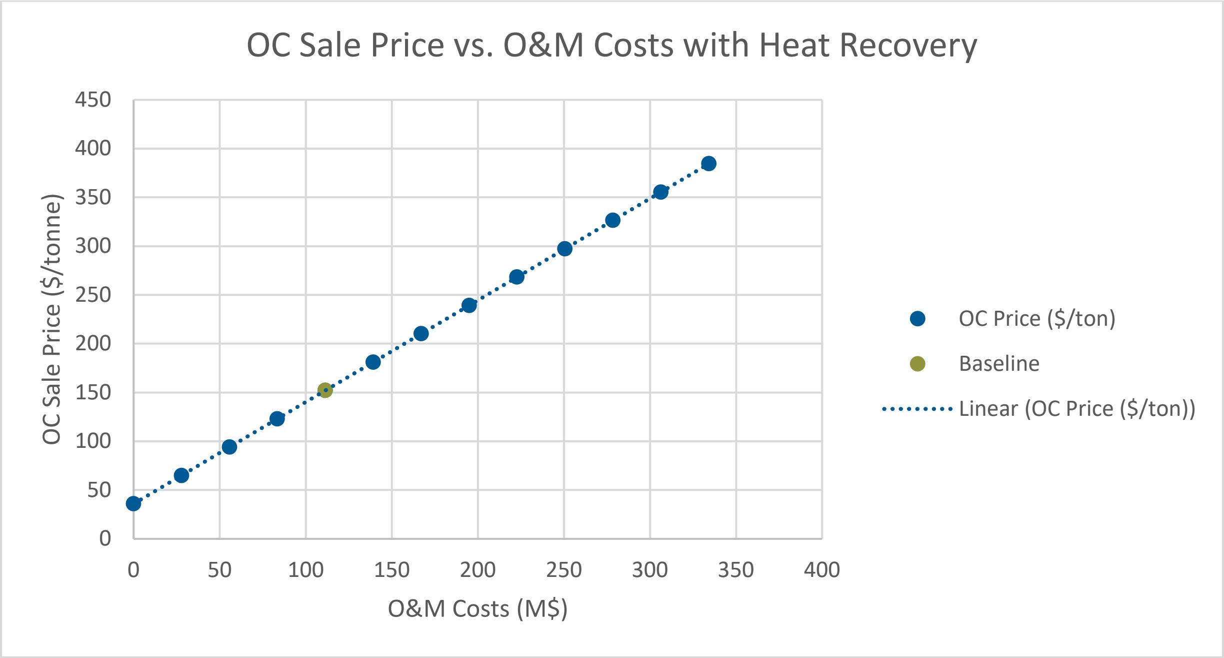
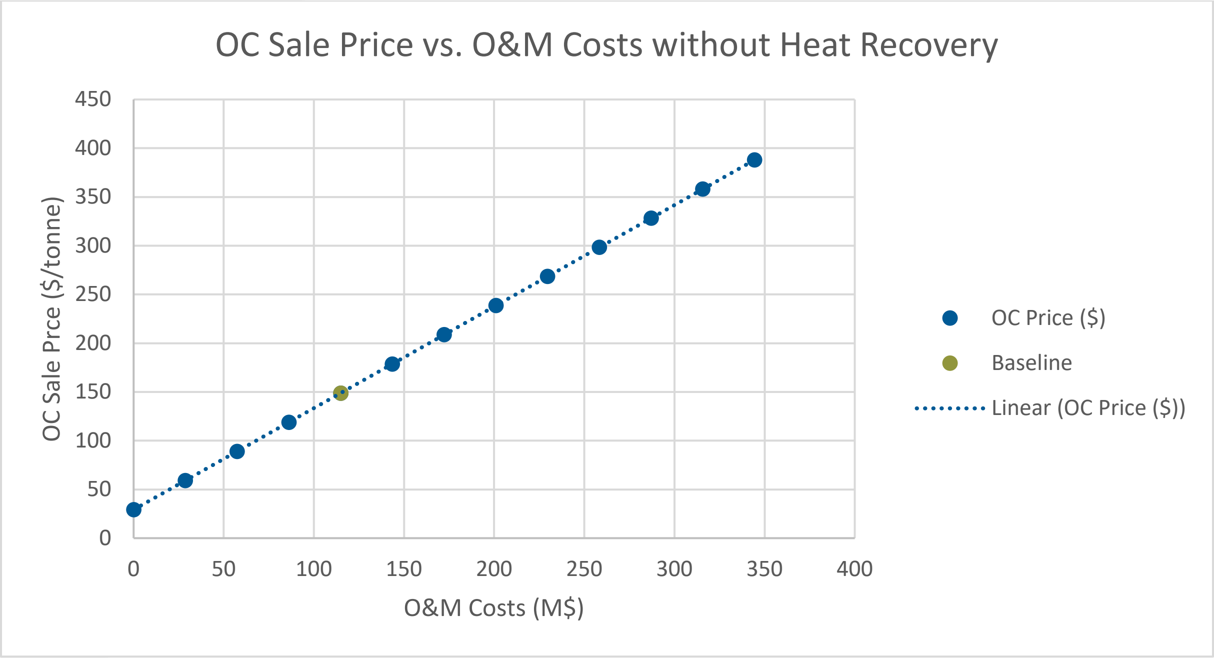


Figure 8-4 OC Sale Price Sensitivity Based on O&M Cost without Heat Recovery



8.3 Effects of Stream 1 Costs

Figure 8-5 and Figure 8-6 provide illustrations of how the sale price of OC varies with the sale price of stream 1. The sale price of OC varies from $120/tonne to $230/tonne when the price of stream 1 varies from $0/tonne to approximately $300/tonne.

Figure 8-5 OC Sale Price Sensitivity Based on Stream 1 Cost with Heat Recovery

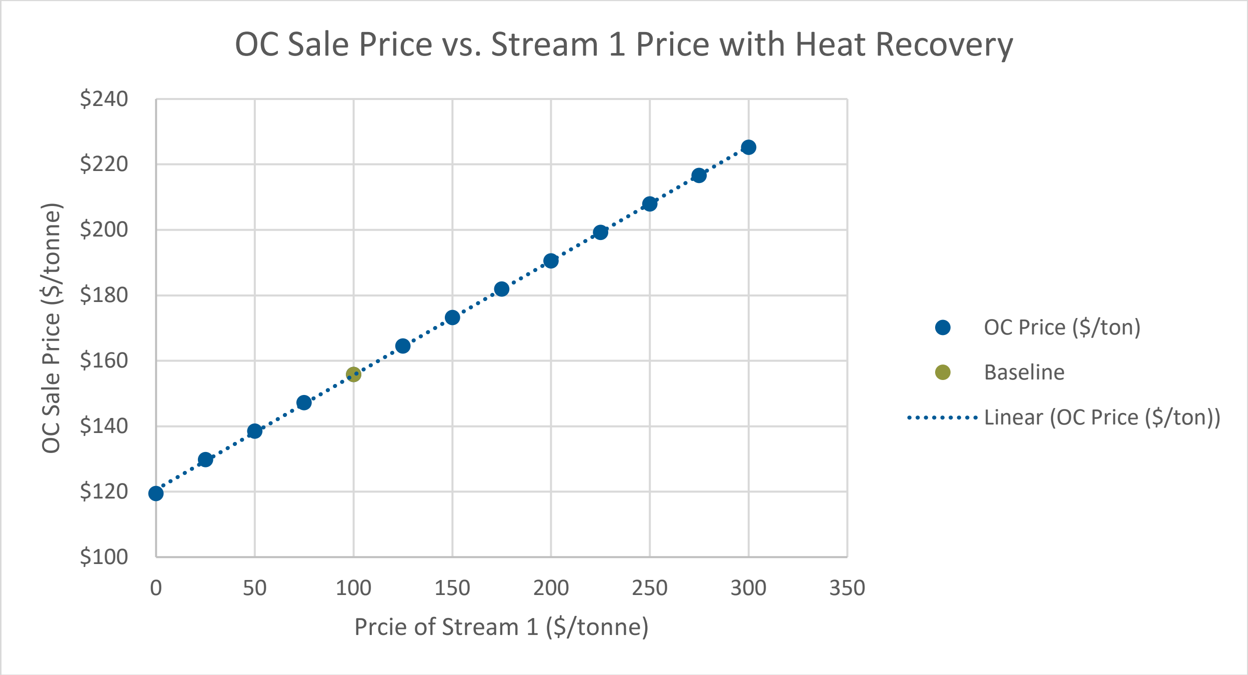
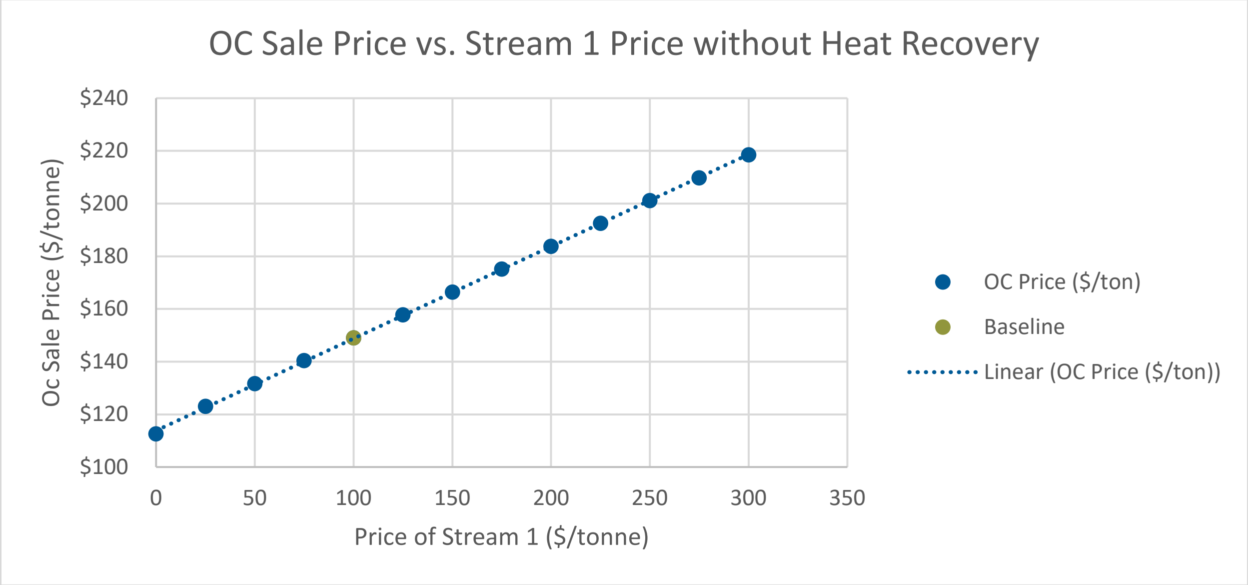


Figure 8-6 OC Sale Price Sensitivity Based on Stream 1 Cost without Heat Recovery



8.4 Effects of Stream 2 Costs

Figure 8-7 and Figure 8-8 provide illustrations of how the sale price of OC varies with the sale price of stream 2. The sale price of OC varies from $130/tonne to $180/tonne when the price of stream 2 varies from $0/tonne to approximately $90/tonne.

Figure 8-7 OC Sale Price Sensitivity Based on Stream 2 Cost with Heat Recovery

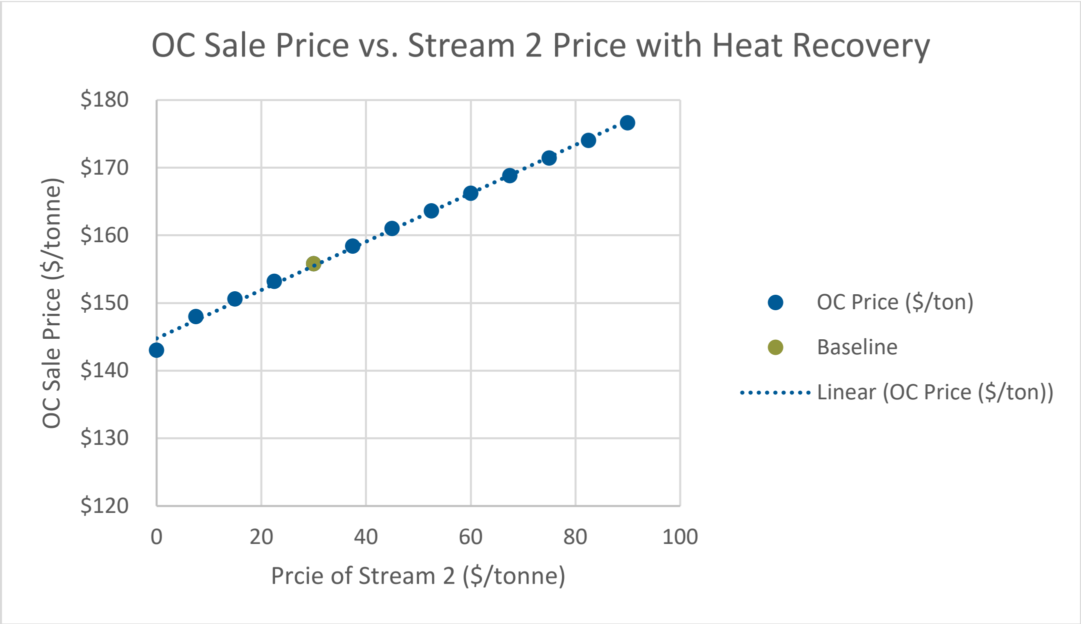
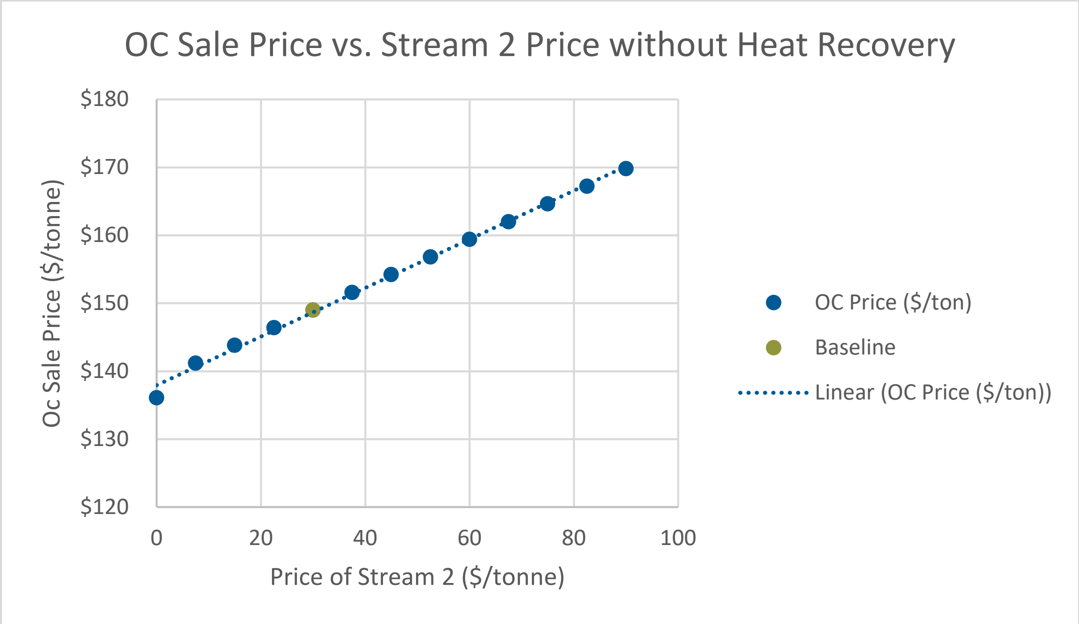


Figure 8-8 OC Sale Price Sensitivity Based on Stream 2 Cost without Heat Recovery



8.5 Effects of Stream 3 Costs

Figure 8-9 and Figure 8-10 provide illustrations of how the sale price of OC varies with the sale price of stream 3. The sale price of OC varies from $120/tonne to $220/tonne when the price of stream 3 varies from $0/tonne to approximately $900/tonne.

Figure 8-9 OC Sale Price Sensitivity Based on Stream 3 Cost with Heat Recovery

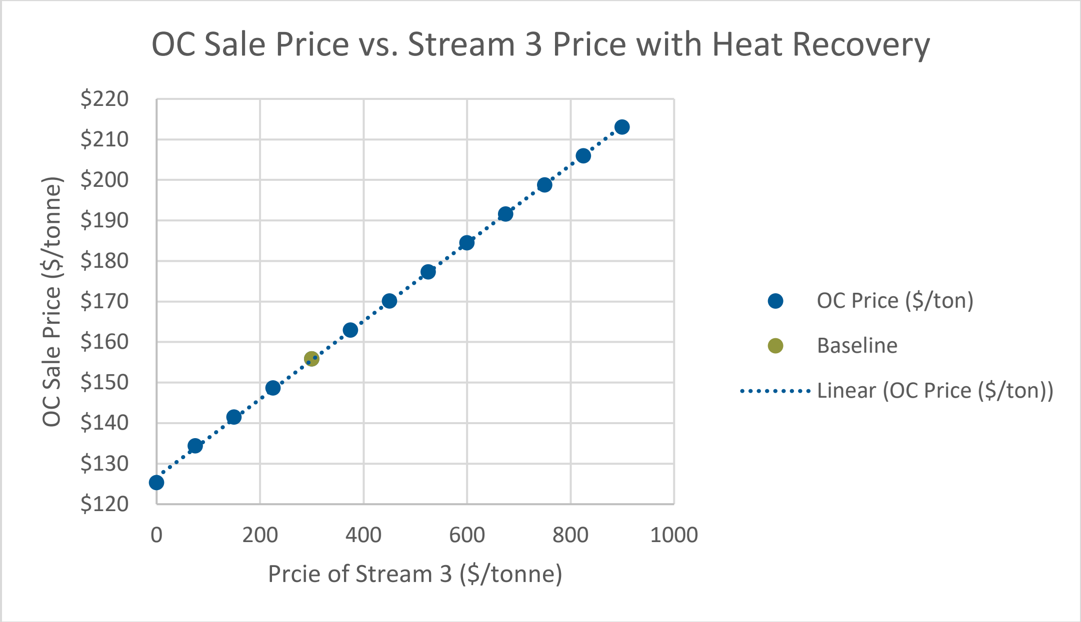
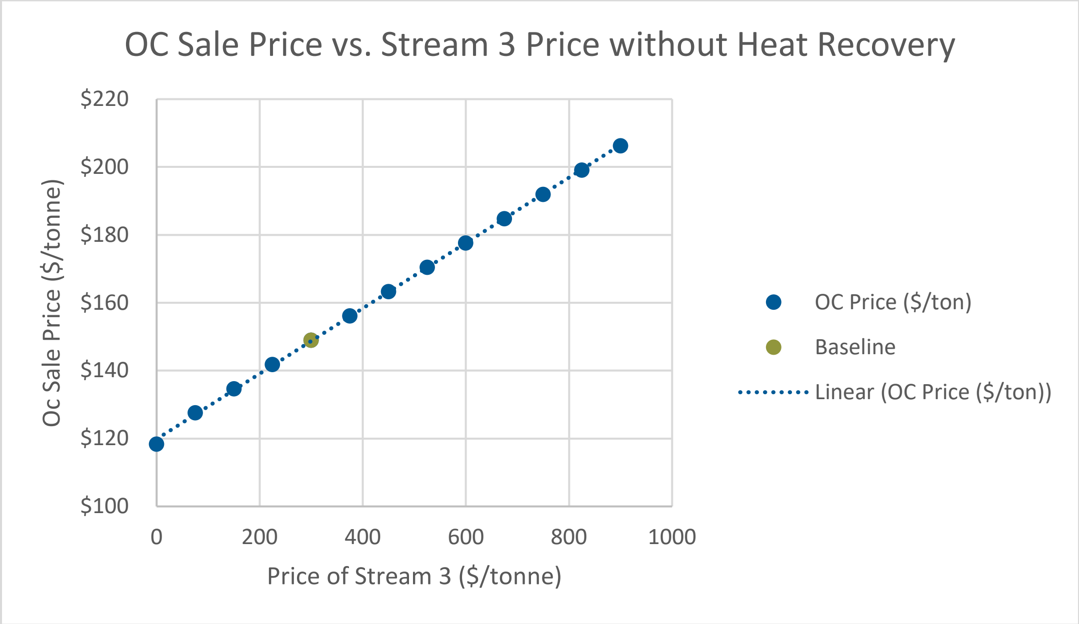


Figure 8-10 OC Sale Price Sensitivity Based on Stream 3 Cost without Heat Recovery



8.6 Effects of Stream 4 Costs

Figure 8-11 and Figure 8-12 provide illustrations of how the sale price of OC varies with the sale price of stream 4. The sale price of OC varies from $140/tonne to $160/tonne when the price of stream 4 varies from $0/tonne to approximately $450/tonne.

Figure 8-11 OC Sale Price Sensitivity Based on Stream 4 Cost with Heat Recovery

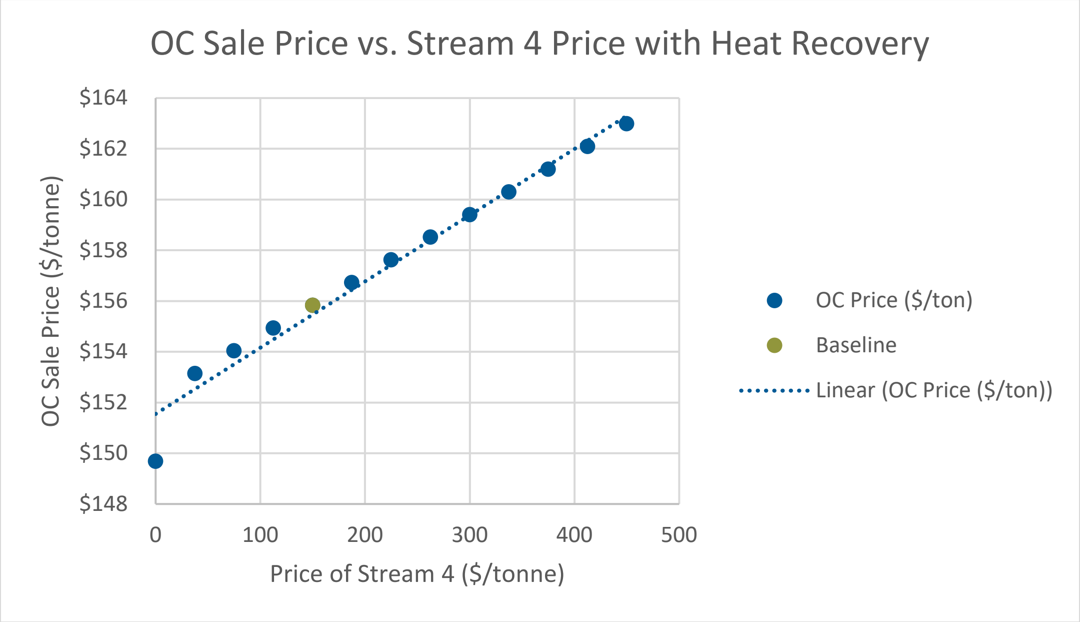
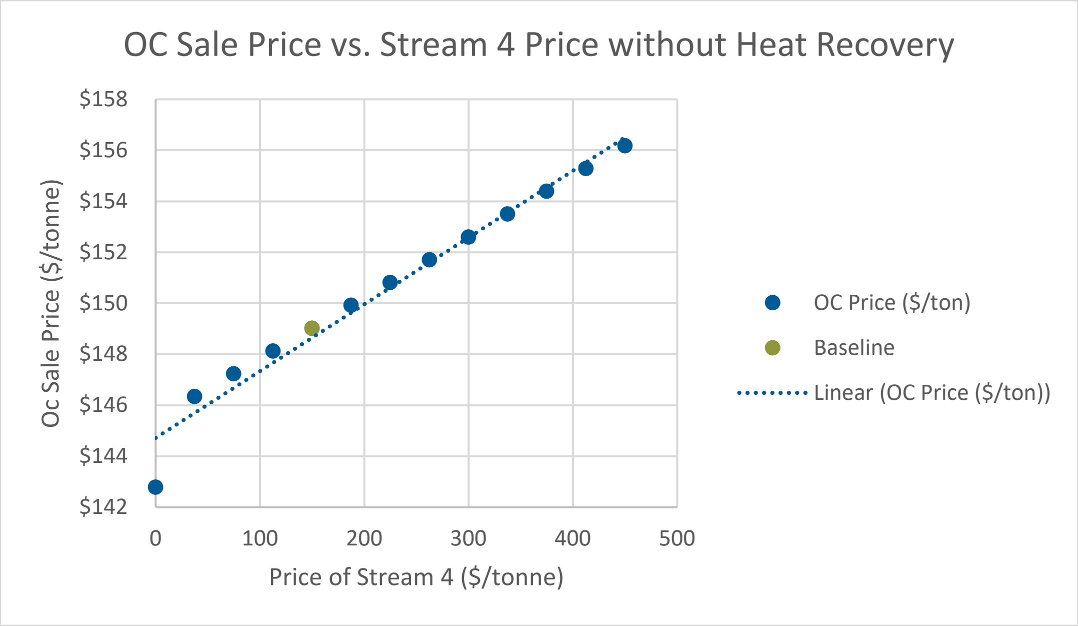


Figure 8-12 OC Sale Price Sensitivity Based on Stream 4 Cost without Heat Recovery



8.7 Effects of Stream 5 Costs

Figure 8-13 and Figure 8-14 provide illustrations of how the sale price of OC varies with the sale price of stream 5. The sale price of OC varies from $130/tonne to $180/tonne when the price of stream 5 varies from $0/tonne to approximately $400/tonne.

Figure 8-13 OC Sale Price Sensitivity Based on Stream 5 Cost with Heat Recovery

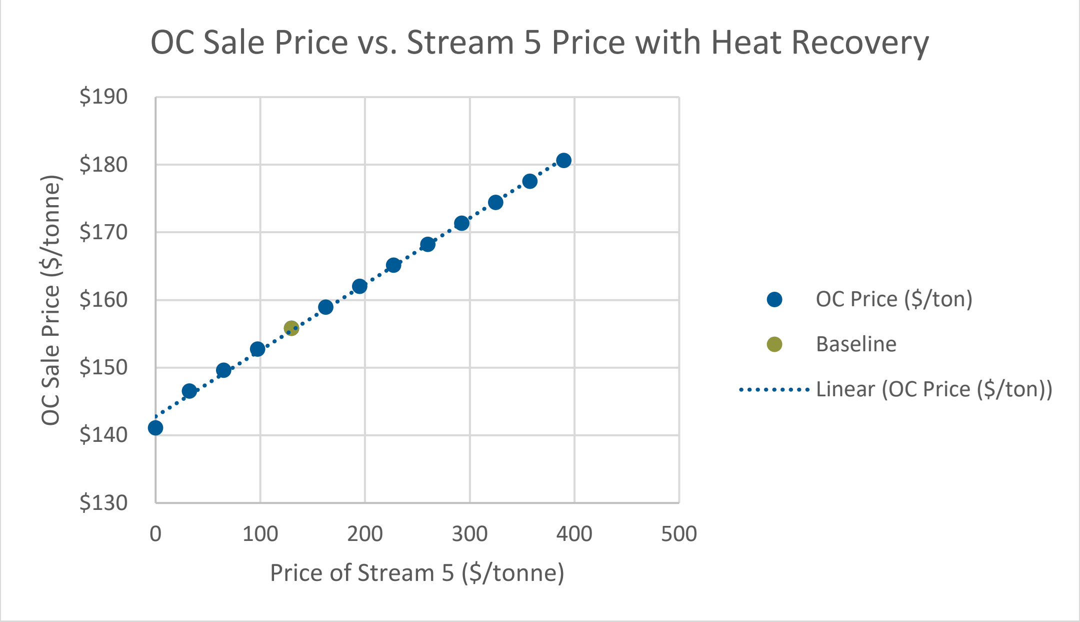
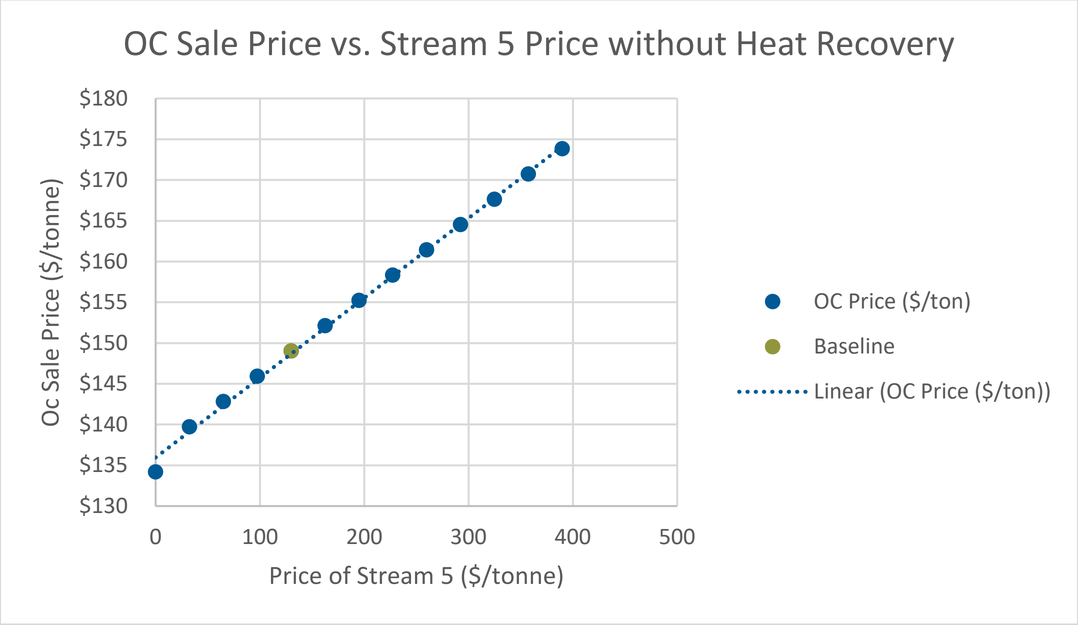


Figure 8-14 OC Sale Price Sensitivity Based on Stream 5 Cost without Heat Recovery



8.8 Effects of FRC

Figure 8-15 and Figure 8-16 provide illustrations of how the sale price of OC varies with the Fixed Rate Charge (FRC). The sale price of OC varies from $110/tonne to $230/tonne when the FRC varies from 0 to approximately 0.027.

Figure 8-15 OC Sale Price Sensitivity Based on FRC with Heat Recovery

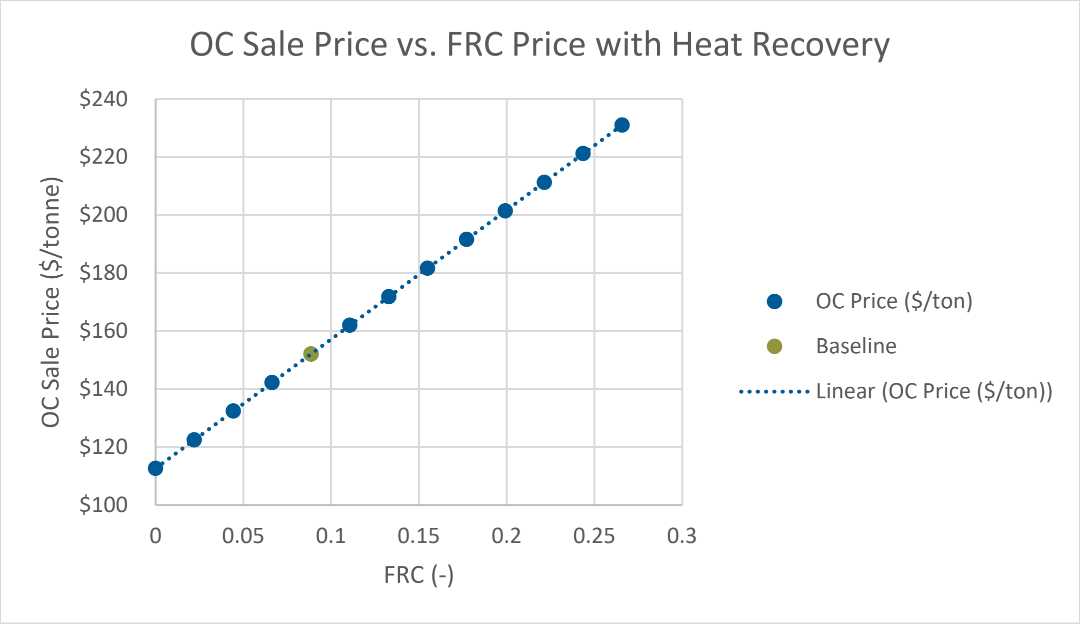
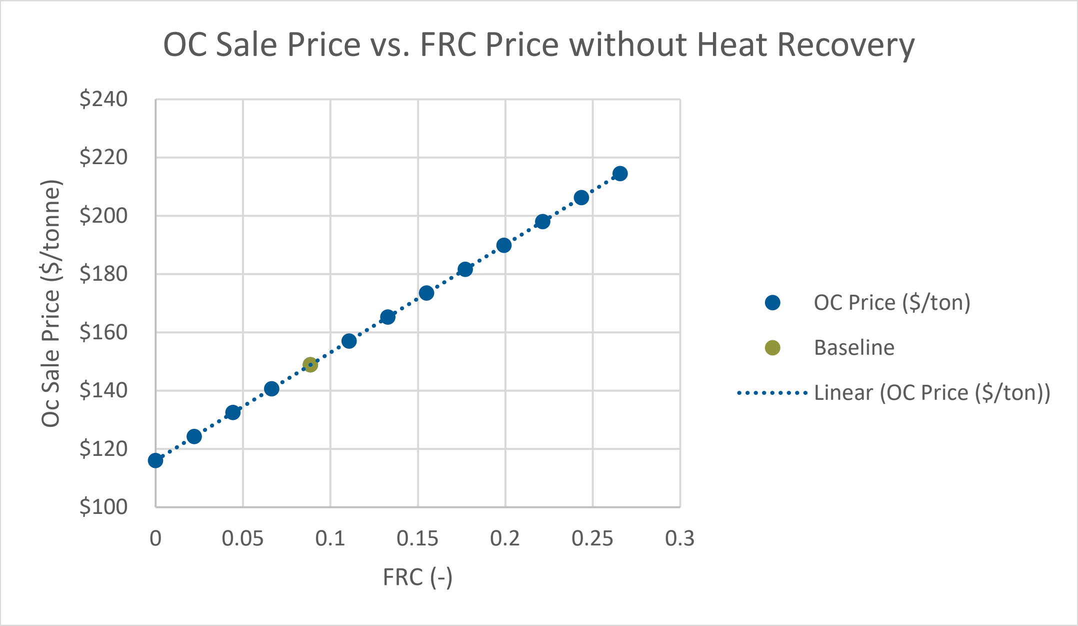


Figure 8-16 OC Sale Price Sensitivity Based on FRC without Heat Recovery



8.9 Effects of Costs on OC Sale Price

Figure 8-17 and Figure 8-18 provide illustrations of how the sale price of OC varies (in percentage of base price) with the percent change in other costs. The sale price of OC varies most greatly with the change in O&M costs, to which the price of stream 1 and stream 2 contribute more than then the price of stream 3, stream 4, and stream 5.

Figure 8-17 OC Sale Price Sensitivity Based on Percent Differences with Heat Recovery

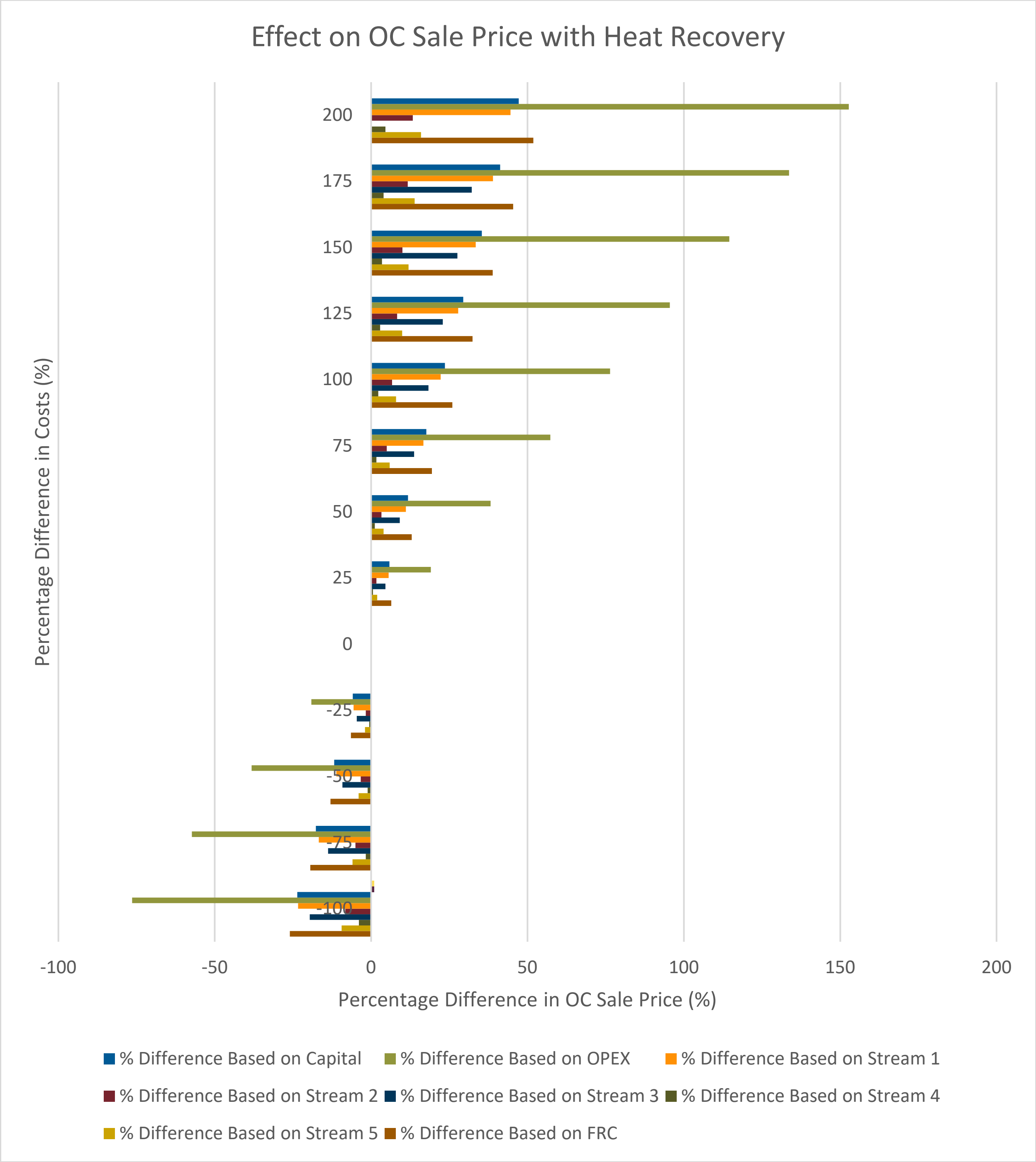
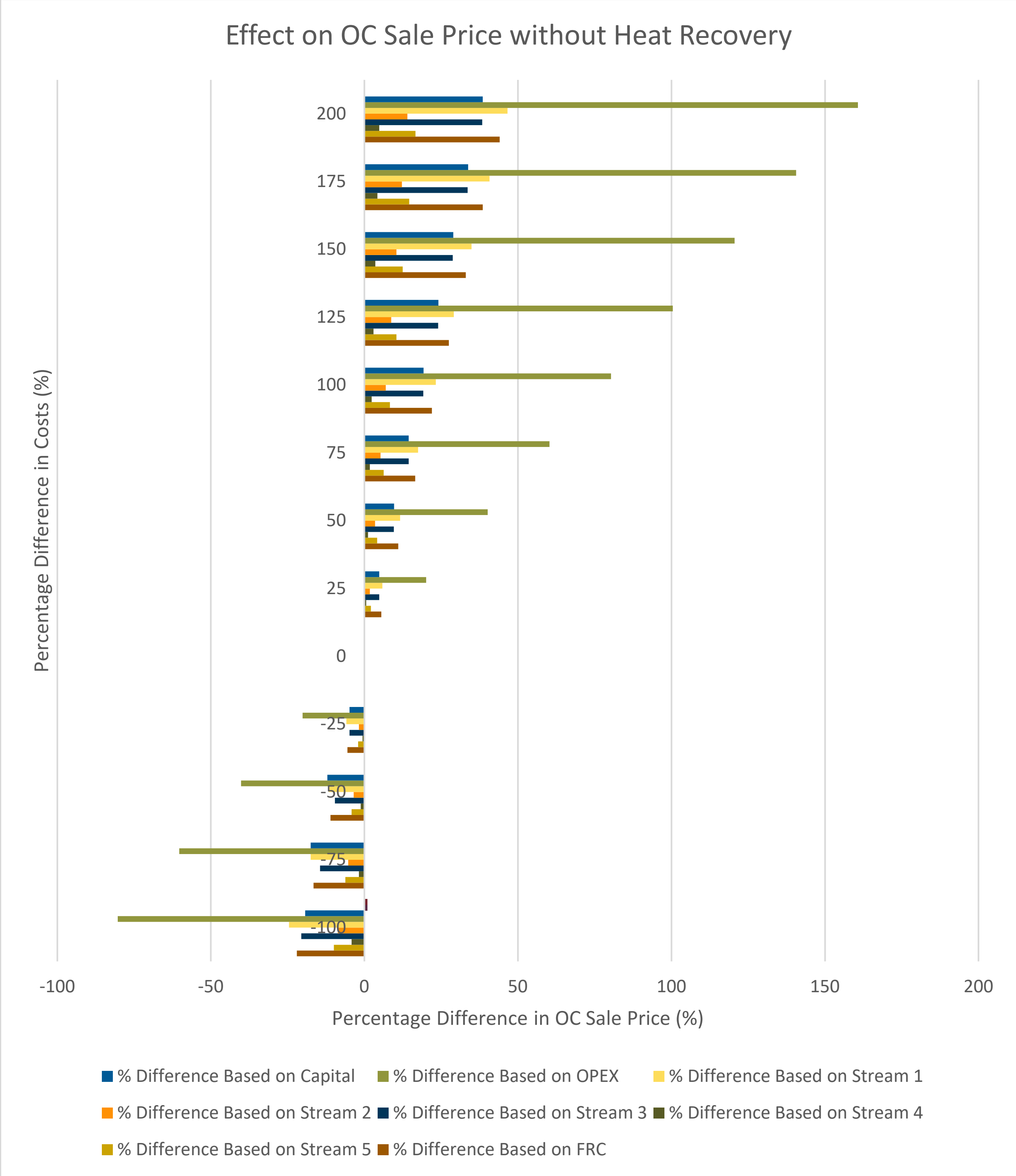


Figure 8-18 OC Sale Price Sensitivity Based on Percent Differences without Heat Recovery



UNIVERSITY OF NORTH DAKOTA

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6 Appendix B – TEA of 585 MW_e CLC Power Plant

Technical Economic Analysis

Oxygen Carrier Production Plant – With Optional Heat Recovery System

Final

Author List:

University of North Dakota
Junior Nasah, Johannes Van der Watt, Mark Musich

Envergex LLC
Srivats Srinivasachar, Aaron Koenig

Barr Engineering Co.

Technical Economic Analysis

Oxygen Carrier Production Plant – With Optional Heat Recovery System

Final

Final Revision by: University of North Dakota

February 2022

TECHNICAL ECOMONIC ANALYSIS

Oxygen Carrier Production Plant – With Optional Heat Recovery System

Final

FEBRUARY 2022

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Acronyms

AACE	Association for the Advancement of Cost Engineering
AFUDC	Allowance for Funds Used During Construction
ANSI	American National Standards Institute
ATWACC	After-Tax Weighted Average Cost of Capital
BEC	Bare Erected Cost
BOP	Balance of Plant
CCW	Closed Cycle Water
CLC	Chemical Looping Combustion
CF	Capacity Factor
CRF	Capital Recovery Factors
DOE	Department of Energy
EPC	Engineering, Procurement, and Construction
EPCM	Engineering, Procurement, and Construction Management
FA Fans	Fresh Air Fans
FBR	Fluidized Bed Reactor
FCR	Fixed Charge Rate
FD Fans	Forced Draft Fans
FEED	Front-End Engineering Design
FGD	Flue Gas Desulfurization
HRSG	Heat Recovery Steam Generator
HVAC	Heating, Ventilation, and Air Conditioning
ID Fans	Induced Draft Fans
IOU	Investor-Owned Utility
KW	Kilowatt
LTE	Low Temperature Economizer
MMBTU	One Million British Thermal Units
MW	Megawatt
NETL	National Energy Technology Laboratory
OC	Oxygen Carrier
OEM	Original Equipment Manufacturer
QGESS	Quality Guidelines for Energy Systems Studies
SCR	Selective Catalytic Reduction
SO_2	Silicon Dioxide
STG	Stream Turbine Generator
TASC	Total As Spent Cost
TOC	Total Overnight Cost
TPC	Total Plant Cost
UND-IES	University of North Dakota Institute for Energy Studies
ZLD	Zero Liquid Discharge

1 OC Production Process

1.1 Design Basis

The design annual OC production rate at a capacity factor of 0.85 is approximately 990,000 tonne, or, 133 tonne per hour. The production plant will be located in the Gary/East Chicago, Indiana region of the Great Lakes in order to be close to suitable iron ore reserves.

Proprietary binder and additives comprise approximately 20-30 wt.% of the green OC.

1.2 Process Description

A simplified block flow diagram is shown in Figure 1-1. The process consists of four areas: green OC preparation, OC calcination, heat recovery, and particulate and SO_2 control.

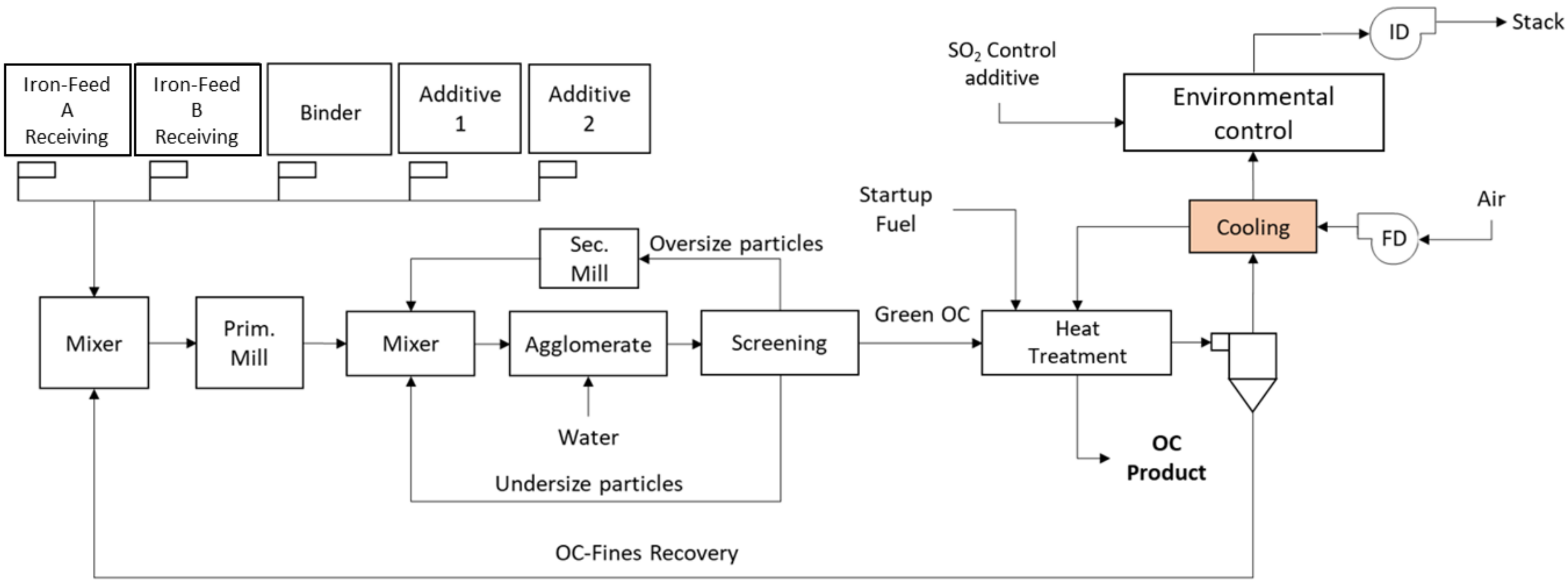


Figure 1-1 **Block flow diagram of novel oxygen carrier production process without heat recovery**

1.2.1 Green OC Preparation

Principal raw materials (iron-feed A, iron-feed B, binder, and additives) for green OC preparation are stored in silos with 72 hours capacity. Recycle streams from the OC calcination, and particulate and SO_2 control process are blended with the material.

Iron-feed A is fed to the first-stage paddle-mixer (Mixer 1). The nominal Mixer 1 output is 124 tonne per hour at an approximate moisture content of 1 wt.%. Mixer 1 product is discharged to the primary mill (Mill 1), a double roll crusher. The homogeneous, mm top-size product from Mill 1 is blended with recycled green OC screenings (oversize and undersize).

Mixer 2 product is discharged to a drum pelletizer (Granulator) where water is added to facilitate material cohesion and densification. Spherical agglomerates from the Granulator, are discharged to a classifier (Screen 1) to generate a green OC product in the target size range. Undersize green OC is returned directly to Mixer 2. Oversize green OC is processed in a double roll crusher (Mill 2) to produce a size suitable for recycle to Mixer 2. Green OC product from Screen 1, at a mass rate of 134 tonne per hour, is conveyed to OC calcination.

1.2.2 OC Curing

Curing is conducted in a low-velocity, bubbling fluid-bed to minimize elutriation and attrition. 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 Fe_2O_3.

The slightly oxygen depleted flue gas from the Calciner contains OC dust, with the coarse fraction captured in the High-Temperature (refractory-lined) Cyclone. The cyclone underflow solid stream is returned to Mixer 1.

During the calcination the exothermic reaction produces an excess of 35 MW_{th} that will need to be removed to maintain operating temperatures. Two engineering approaches were considered to address this requirement:

- In bed cooling accomplished through internal water pipes. The excess heat will convert the water to steam that will be sent to the Steam Turbine. The calciner equipment adopted was the commercial scale Fluidized Bed Reactor from SCHWING Technologies who proposed a modular design based on their commercial-scale FBR array. Current commercial design of the FBR systems add heat to the process, consequently, limited development work will be needed to modify the system to allow for heat extraction.

- 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.

1.2.3 Heat Recovery

The UND-IES and Envergex LLC's OC production process was designed with heat recovery to increase thermal efficiency and to reduce green-house gas emissions. Two separate heat recovery processes are employed: heat exchange for internal heat recycle, and steam generation for captive power requirements.

The flue gas exits the High-Temperature (refractory-lined) Cyclone and enters Cooler 1, which cools the gas before it enters the gas-to-gas heat exchanger (Heat Exchanger 2).

Through indirect gas-to-gas heat exchange (Heat Exchanger 2), partially cooled flue gas from Cooler 1 preheats fresh air for use in fluid-bed Calciner 1. Flue gas Heat Exchanger 2 effluent is maintained at a temperature high enough to prevent moisture condensation in downstream processes.

A secondary method for heat recovery is the generation of steam from the hot OC discharged from the Calciner. Two methods are proposed. 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 genset, with the resulting power captively used by motors (conveyors, blowers, mixers, mills, pumps, and fans) to support facility electrical and mechanical needs.

1.2.4 Particulate and SO_2 Control

Heat Exchanger 2 effluent flue gas is directed to particulate and SO_2 control. 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.

2 Methodology and Approach

2.1 Cost Estimation Qualifications

The Class 5 constructed cost estimate provided in this report is based on BARR Engineering's experience and qualifications and represents their best judgment as experienced and qualified professionals familiar with the project. This opinion is based on project-related information available to the team at this time, current information about probable future costs, and a concept-level design of the project. The construction cost opinion will likely change as more information becomes available and more design completed. In addition, because there is no control over the eventual cost of labor, materials, equipment, or services furnished by others; the contractor's methods of determining prices; competitive bidding; or market conditions, there is no guarantee that proposals, bids, or actual construction costs will not vary from the opinion of probable construction cost presented in this report. Greater assurance as to the probable construction costs can be achieved through additional design to provide a more complete project definition. Qualifying assumptions and exclusions on which the estimate is based are included in Section 4.5.

The following guidelines were used in evaluation and preparation of this cost report:

- Quality Guidelines for Energy Systems Studies (QGESS).
- The capital and O&M costs have been reported at a level of detail similar to that found in DOE/NETL Baseline studies for Cost and Performance Baseline for Fossil Energy Plants. Volume 1: Bituminous Coal and Natural Gas to Electricity.

2.2 Estimate Type

The cost estimate corresponds to a Class 5 estimate (AACE International Recommended Practice No. 18R-97) for the process industries. This estimate classification is characterized by limited project definition and the wide-scale use of scaling and power industry experience to calculate costs. A Class 5 has an end use for concept screening, with a lower bound accuracy range of -20% to -50% and an upper bound accuracy range of +30% to +100%. These parameters for a Class 5 estimate are in the table below.

Table 2-1 AACE Generic Cost Estimate Classification Matrix

		Primary Characteristic	Secondary Characteristics			
		Level of Project Definition	End Usage	Methodology	Low Range Expected Cost	High Range Expected Cost
Estimate Class	ANSI Classification	Expressed as % of complete project definition	Typical purpose of estimate	Typical estimating method	Typical +/- range relative to best range index	Typical +/- range relative to best range index
Class 5	Order of Magnitude	0% to 2%	Concept Screening	Capacity Factored, Parametric Models, Judgment, or Analogy	-30%- -50%	+30% - +100%

2.3 Cost Estimate Scope

The scope of the cost estimate is completed for a theoretical OC Production Plant with a 16 MW_e Heat Recovery System in the midwestern United States.

The capital cost estimate provided is considered an order of magnitude, or parametric type, estimate with historical/actual cost curves based on historical data from other projects.

The operating and maintenance costs were evaluated using the mass balance calculations from the model. Costs of labor, consumables, and waste disposal were estimated from the "Mineral Commodity Summaries 2020" by the United States Geological Survey (USGS) as well as estimates from similar projects.

2.4 System Code-of-Accounts

Table 2-2 includes a description of the code of accounts used to break down the cost evaluation.

Table 2-2 Description of OC Code of Accounts

Item	Description
1	Bulk Material Storage & Handling
2	Oxygen Carrier Processing
3	Heat Recovery System
4	Environmental Controls
5	Electrical Systems
6	Building & Facilities

2.4.1 Code of Accounts Detailed Breakdown

Class 5 cost estimates are presented for the following construction features required for the project:

1. Bulk Material Storage & Handling
    1. Iron-feed A storage silo (72hrs of storage)
    2. Iron-feed B storage silo (72hrs of storage)
    3. Binder 1 storage silo (72hrs of storage)
    4. Additive 1 storage silo (72hrs of storage)
    5. Additive 2 storage silo (72hrs of storage)
    6. OC Product storage silo (72hrs of storage)
    7. Conveyors (x6)
2. Oxygen Carrier Processing
    1. Mill 1
    2. Mill 2
    3. Mixer 1
    4. Mixer 2
    5. Granulator
    6. Screens
    7. Fluid-Bed Calciner

8. Heat Exchanger
9. High Temperature Cyclone
10. Flue Gas Cooler
11. Belt Conveyors (belt with dust control)
12. Forced Draft Fans

3. <u>Heat Recovery System</u>
   1. OC Solid-to-Air Heat Recovery Heat Exchangers
   2. HRSG/Heat Recovery Boiler
   3. Feed Water Pumps
   4. Fresh Air Fans
   5. 16 MW_e Steam Turbine and Generator
   6. Deaerator
   7. Condenser
   8. Condenser Pump
   9. Cooling Tower

4. <u>Environmental Controls</u>
   1. Baghouse
   2. Packed Bed Scrubber
   3. Instrumentation
   4. Slaker
   5. Sorbent Injection
   6. Forced Oxidizer
   7. Water Treatment / Zero Liquid Discharge System (ZLD)
   8. Induced Draft Fans

5. <u>Electric Systems</u>
   1. Power Auxiliaries
   2. Instrumentation and Control

6. <u>Buildings and Structures</u>
   1. Rigid steel frame with steel roof and walls, foundation and flooring, insulation, electrical, mechanical, HVAC, and plumbing
   2. Site preparation & improvements to site

The summary and detail tables of the total plant cost (TPC) estimate prepared for the project are included in Appendix B.

2.5 Assumptions and Exclusions

Key assumptions are included below and summarized in Appendix C.

2.5.1 Base Case Assumptions

1. 85% capacity factor (O&M Base Case)
2. Greenfield site / Midwest.
3. 100-acre site
4. The power produced by the heat recovery system supplies the OC system and auxiliary loads
5. The substation is provided by the utility company providing additional power (for case without heat recovery)
6. The natural gas supply is assumed to be available at the site boundary
7. Water is available from local municipality
8. Prices may fluctuate due to the varying costs of material and equipment that are driven by multiple market variables. Vendor quotes were either provided in 2021 or in 2019 (and scaled to 2021 dollars). The quotes provided by the vendor may vary over time and as the scope and design becomes more defined.

Base Case Exclusions

The following items are excluded from the project cost estimate:

1. Hazardous, contaminated materials and remediation
   a. Asbestos
   b. Lead abatement
   c. PCBs
   d. Contaminated soils
      i. Contaminated ground water
   e. Site conditions
      i. Piles or caissons
      ii. Rock removal
      iii. Excessive dewatering
      iv. Expansive soil considerations
      v. Excessive seismic considerations
      vi. Extreme temperature considerations
      vii. Demolition or relocation of existing structures
      viii. Unforeseen conditions
      ix. Sub-surface conditions
      x. Existing unknown conditions
   f. Fees and Permits
      i. State licenses
      ii. Local license
      iii. Environmental permits

iv. Building permits

v. Third party, professional fees, material testing, and inspections

g. Leasing of off-site land for parking or laydown

h. Busing of craft to site

i. Costs of off-site storage

j. Furnishings and special items

   i. Any furniture, window treatments, or other furnishings

k. Transportation and storage (T&S) is not considered in the capital cost, owner's costs, or O&M. T&S includes items such as:

   i. New access roads and railroad tracks

   ii. Upgrades to existing roads to accommodate increased traffic

   iii. Makeup water pipe outside the "fence line"

   iv. Landfill for onsite waste (slag) disposal

   v. Backup fuel provisions

   vi. Plant switchyard

   vii. Electrical transmission lines outside of plant boundary

   viii. Carbon unloading, sequestration, or transport pipeline

2.6 Cost of Mature Technologies and Designs

The cost estimates of mature technologies and designs are based on vendor quotes procured for this cost estimate. These quotes were used in a capital cost estimate conducted by Barr with a process contingency of 10%. Original equipment manufacturer (OEM) Quotes were obtained for the major equipment listed in Table 2-3.

Table 2-3 List of Major Equipment and Vendors

Equipment	Vendor
Silos and Screen	Pearson Arnold
Dust Control	Donaldson Inc.
FA Fans, FD Fans, and ID Fans	Howden
Steam Turbine (with auxiliaries)	Siemens Energy
HRSG / Heat Recovery Boiler	Hurst
Deaerator	Hurst
OC Solid to Air Heat Recovery Heat Exchangers	Solex
Cooling Tower	Evaptech
Condenser	Babcock Power Thermal Engineering International (TEI)
Condensate Pumps	Somesnick Sales Company, INC.
Circulating Water Pumps	Somesnick Sales Company, INC.
Water Treatment System and ZLD	WesTech

For these readily-available systems, a process contingency of 10% was considered in the cost estimate as shown in Table 2-3. These systems have been proven in full-scale commercial applications. Some unknowns contributing to the uncertainty are water/steam quality, as well as control of this plant to accommodate the high ramp rates and turndowns.

2.7 Costs of Emerging Technologies, Designs, and Trends

There are some areas where the technology is not common or commercially available. Table 4-5 lists one of these areas. The cost was obtained from the OEM for each of these areas. A higher process contingency of 20% is included to account for the emerging technology status as shown in Table 2-4.

Table 2-4 List of Emerging Technologies

Equipment	Vendor
Fluid Bed Calciner	Schwing Technologies
High Temp Cyclone	TBD
Environmental Controls	TBD

While the equipment listed is available on the market, additional engineering costs will be required to integrate the equipment into the proposed concept. These costs are taken into consideration in the TPC. The potential factors which may affect the capital cost of each of these technologies follow:

- ***Fluid Bed Calciner:*** The FBR system proposed by SCHWING Technologies for calcining is commercially available. A higher contingency was proposed due to the higher gas-to-solid ratio requirements for calcining the OC and the need for in-bed heat extraction.
- ***High Temperature Cyclones:*** The high exhaust temperatures and abrasive nature of the OC requires adoption of a high temperature cyclone. Even though cyclone manufacturing is pretty standard, the

operating conditions required are not. Consequently, a higher process contingency was provided to account for operating conditions.

- ***Environmental Controls:*** The scrubbing system itself is a straightforward concept that poses little uncertainty. Factors that are undefined are the levels of SO_2, required limits, ability to oxidize the scrubber water (as lime forced oxidation is a newer process), and make-up water quality going to the zero liquid discharge system (ZLD). These factors will influence the design of the packed bed scrubber, wastewater treatment, and the ZLD.

2.7.1 Project Contingency

Project contingency compensates for cost uncertainties and construction risk associated with final design and construction of the project until the project is completed. Uncertainty in early stages of project planning and design, especially during the feasibility study phase, are greater due to factors such as limited project definition, design and analysis assumptions, unforeseen constraints and constructability issues, construction schedule, and other construction risk factors. In general, uncertainty will decrease as greater definition is developed and more detailed information becomes available.

At this stage in the project, the design is less than 2% complete and constructability has not been evaluated due to insufficient design detail. Therefore, the range of uncertainty of total project cost is high. AACE 16R-90 states that project contingency for a “budget-type” estimate (AACE Class 4 or 5) should be 15% to 30% of the sum of BEC, EPC fees, and process contingency.

The project contingency was determined by taking various percentages of the bare erected costs plus the costs up through process contingency. The project contingency will be reduced as engineering progresses further in later phases and potential further cost reduction with value engineering, standardization, and modularization strategies.

3 Capital Cost Estimate

The Total Plant Cost (TPC) was determined to estimate the project's cost. The TPC is the sum of the Bare Erected Cost (BEC) for the project, plus the cost of the engineering, procurement, and construction (EPC) contractor, as well as process and project contingencies. The TPC is an overnight cost calculated in 2021 dollars.

The BEC consists of the cost of equipment and materials. The major equipment vendors provided Original Equipment Manufacturer (OEM) costs to be used to estimate the BEC. The BEC also contains new onsite facilities, site infrastructure, and balance-of-plant equipment necessary to support the process. The facilities and site infrastructure were estimated based on other Barr-related projects. The direct and indirect construction labor required for installation is included.

The Engineering, Procurement, and Construction Management (EPCM) costs include detailed design, building-related permits obtained by the contractor, as well as project and construction management costs. EPC costs are based on a construction management approach utilizing a prime contractor with multiple subcontractors. This approach provides the owner with greater scope control and flexibility, while mitigating the risk premium typically included in a traditional EPC lump-sum pricing structure.

3.1 Quantities and Allowances

High-level quantity takeoffs for major system components such as sorbent regeneration, flue gas clean-up, and conveyance systems were developed from the general process and instrument diagrams.

3.2 Escalation

Escalation was not considered for the TPC. Therefore, the TPC is in the current dollar amount (for 2021).

3.3 Labor Cost Basis

The estimate was not adjusted for local area labor rates. Labor rates reflect a burden rate, including: worker's compensation, state and federal unemployment taxes, fringe benefits, medical insurances, and other typical burdens. The labor rates are based upon a work week of 40 hours: 8 hours per day / 5 days per week. The average labor rate is $100 per hour with burdens.

3.4 Freight and Shipping Costs

The estimate provided includes freight and shipping cost, duties for all major items of equipment.

3.5 Contingency

Contingency represents an allowance to cover unknowns, uncertainties, and/or unanticipated conditions that are not possible to evaluate adequately from the information at hand at the time the cost estimate is prepared but must be represented by a sufficient cost to cover the identified risks. Contingency relates to a known defined project scope and is not used to predict future project scope or schedule changes. Contingency will normally decrease as more design information is known. This section summarizes important cost-estimating considerations related to cost contingency.

Contingencies, as used in this estimate, are intended to help identify an estimated construction cost amount for the items included in the current project scope. The contingency percentage includes process contingency and project contingency. These contingency amounts are based on AACE guidelines and professional judgment considering the level of design completed, the complexity of the work, and uncertainties in quantities and unit prices. The contingency includes the estimated cost of ancillary items not currently identified in the quantity estimates and allowances, but commonly identified in more detailed design and required for completeness of the work.

Contingencies are assigned to the cost estimate of each project feature based on engineering judgment and on the relative completeness of project definition. Contingency, as used in this cost estimate, will decrease with future design efforts.

The contingency provided with the estimate does not account for:

- changes in labor availability or productivity
- delays in equipment deliveries
- changes in current industry standards or regulations
- major changes in quantities
- major changes in unit pricing
- major changes in scope during detailed design or construction
- major changes or revisions to the design basis
- costs that may result from actual site conditions differing from generic site conditions assumed in this estimate
- costs that result from construction change orders
- costs that result from sequencing or expediting work to avoid critical path slippage
- costs that result from possible project schedule slippage
- costs that result from differing economic conditions or future cost growth
- costs related to plant performance during and after start-up
- force majeure

The contingency included in the cost estimate is based upon the Risk Management or Estimating Judgement process.

3.5.1 Process Contingency

Process contingency provides for uncertainty in the cost estimate related to the technology's maturity development. The configuration of the key technology pieces of this concept is currently unproven at the commercial scale in manufacturing facility applications. However, many aspects of the project use current proven and accepted technology for balance-of-plant and structural aspects. Therefore, process contingencies are applied to individual aspects of the cost estimate based on the current status of the technology for individual aspects of the cost estimate. AACE recommends the following guidelines in Table 3-1 for process contingency to apply.

Table 3-1 AACE Guidelines for Process Contingency

Technology Status	Process Contingency (% of Associated Process Capital)
New concept with limited data	40+
Concept with bench-scale data	30-70
Small pilot plant data	20-35
Full-sized modules have been operated	5-20
Process is used commercially	0-10

Process contingencies used in this estimate were assigned as shown in Table 3-2.

Table 3-2 Process Contingency for Technology

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

While the equipment listed is readily available on the market, additional engineering costs will be required to integrate the equipment into the proposed concept. These costs are taken into consideration in the TPC. The potential factors which may affect the capital cost of each of these technologies are discussed in Section 2.7.

3.5.2 Project Contingency

Project contingency, as described in Section 4.7, was applied to all code of accounts. Project contingencies used in this estimate were assigned as shown in Table 3-3.

Table 3-3 Project Contingency for Technology

Technology	Project Contingency (% of Associated Process Capital)
Bulk Material Storage & Handling	20%
Oxygen Carrier Processing	20%
Heat Recovery System	20%
Environmental Controls	20%
Electrical Systems	20%
Building & Facilities	20%

3.6 Capital Cost Results

The TPC cost for the OC system is summarized in Table 3-4.

Table 3-4 Capital Cost Summary

Item	Category	Bare Erected Cost (BEC) ($)	Engineering, Procurement & Construction ($)	Process Contingency ($)	Project Contingency ($)	Total Project Cost (TPC) ($)
1	Bulk Material Storage & Handling	2,984,000	448,000	344,000	755,000	4,531,000
2	Oxygen Carrier Processing	49,954,000	7,494,000	9,845,000	13,457,000	80,750,000
3	Heat Recovery System	35,892,000	5,383,000	4,129,000	9,082,000	54,486,000
4	Environmental Controls	27,944,000	4,192,000	6,427,000	7,712,000	46,275,000
5	Electrical Systems	21,628,000	3,245,000	2,487,000	5,472,000	32,832,000
6	Building & Facilities	19,914,000	2,988,000	2,290,000	5,038,000	30,230,000
	Total with Heat Recovery (2021)	158,316,000	23,750,000	25,522,000	41,516,000	249,104,000
	Total without Heat Recovery (2021)	143,560,000	21,537,000	23,824,000	37,782,000	226,703,000
	Total with Heat Recovery (2024)	**170,506,000**	**25,579,000**	**27,487,000**	**44,713,000**	**268,285,000**
	Total without Heat Recovery (2024)	**154,614,000**	**23,195,000**	**25,658,000**	**40,691,000**	**244,159,000**

4 Owner's Costs

The owner's costs were estimated by factoring the values provided in the B12B case in the NETL report. This report estimated the costs based on the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." In this document, the total owner's costs consist of preproduction (startup) costs, inventory capital, land, financing cost, and other owner's costs. Prepaid royalties and working capital are not included in the owner's costs.

The preproduction costs include six months of operating labor, six months maintenance materials at full capacity, one-month non-fuel consumables at full capacity, one-month waste disposal, 25% of one month's fuel cost at full capacity, and 2% of TPC. The six months of operating labor includes the cost of training the plant operators, participation in startup, and occasionally involving them in the design and construction of the power plant.

The inventory capital includes 0.5% of the TPC for spare parts, a 60-day supply (at full capacity) of fuel, and a 60-day supply (at full capacity) of non-fuel consumables (e.g., chemicals and catalysts) that are stored on site. The cost for a 60-day supply (at full capacity) of fuel is not applicable for natural gas. The 60-day supply (at full capacity) of non-fuel consumables does not include catalysts and adsorbents that are batch replacements (such as selective catalytic reduction catalysts).

The cost of land includes a 100-acre site with a $3,000/acre price (based on the site being located in a rural area).

The financing cost is based on 2.7% of the TPC and covers the cost of securing financing, fees, and closing costs. It does not include interest during construction (or AFUDC).

Other owner's costs are estimated using 15% of the TPC. This includes:

1. Preliminary feasibility studies (including a Front-End Engineering Design (FEED) study)
2. Economic development (costs for incentivizing local collaboration and support)
3. Construction and/or improvement of roads and/or railroad spurs outside of site boundary
4. Legal fees
5. Permitting costs
6. Owner's engineering (staff paid by owner to give third-party advice and to help the owner oversee/evaluate the work of the EPC contractor and other contractors)
7. Owner's contingency (sometimes called "management reserve"—these are funds to cover costs relating to delayed startup, fluctuations in equipment costs, unplanned labor incentives in excess of a five-day/ten-hour-per-day work week; owner's contingency is not a part of project contingency)

The owner's costs do not include:

1. EPC risk premiums (costs estimates are based on an EPCM approach utilizing multiple subcontracts, in which the owner assumes project risks for performance, schedule, and cost).
2. Transmission interconnection: the cost of interconnecting with power transmission infrastructure beyond the plant busbar.
3. Taxes on capital costs: all capital costs are assumed to be exempt from state and local taxes.
4. Unusual site improvements: normal costs associated with improvements to the plant site are included in the BEC, assuming that the site is level and requires no environmental remediation; unusual costs associated with the following design parameters are excluded: flood plain considerations, existing soil/site conditions, water discharges and reuse, rainfall/snowfall criteria, seismic design, buildings/enclosures, fire protection, local code height requirements, and noise regulations.

The factors used to adjust the costs were taken from the 2019 revision of the QGESS document "Capital Cost Scaling Methodology: Revision 4 Report."

4.1.1 Owner's Cost Results

The Owner's costs for the OC with and without the heat recover system are summarized in Table 4-1.

Table 4-1 Owner's Costs

Owners Costs		
Description	**Cost (with HR)**	**Cost (without HR)**
Pre-Production Costs		
6 Months All Labor	$7,289,000	$6,282,000
6 Month Maintenance Materials	$906,000	$870,000
1 Month Non-fuel Consumables	$7,972,000	$8,131,000
1 Month Waste Disposal	$59,000	$59,000
25% of 1 Month's Fuel Cost at 100% CF	$170	$30
2% of TPC	$5,366,000	$4,883,000
Total	**$21,592,000**	**$20,225,000**
Inventory Capital		
0.5% of TPC (Spare Parts)	$1,341,000	$1,221,000
60-day Supply of fuel at 100% CF	$1,300	$250
60-day Supply of consumables at 100% CF	$15,715,000	$16,028,000
Total	**$17,057,300**	**$17,249,000**
Land		
Cost (Based on 100 Acres)	$300,000	$300,000
Total	**$300,000**	**$300,000**
Financing Cost		
2.7% of TPC	$7,244,000	$6,592,000
Total	**$7,244,000**	**$6,592,000**
Other Costs		
15% of TPC	$40,243,000	$36,624,000
Total	**$40,243,000**	**$36,624,000**
Total Overnight Costs (TOC)	**$354,721,000**	**$294,789,000**
TASC/TOC Multiplier (IOU, high-risk, 5 year)	1.242	1.242
Total As-Spent Cost (TASC)	**$440,563,000**	**$366,128,000**

5 Operation and Maintenance Costs

The yearly operating and maintenance costs associated with the proposed power plant were calculated. The main components of the yearly operating cost are:

- Operating labor
- Maintenance material and labor
- Administrative and support labor
- Consumables
- Waste handling
- Co-products and saleable by-products
- Fuel

The operating and maintenance labor is estimated using methods similar to those contained in the 2019 revision of the QGESS document “Cost Estimation Methodology for NETL Assessment of Power Plant Performance.” However, consumable produce prices were taken from “Mineral Commodity Summaries 2020” by the United States Geological Survey (USGS).

5.1 Auxiliary Power Consumption

A full load case based on auxiliaries was not completed but it is estimated that the power produced by the heat recovered from the produced OC excess heat would cover or exceed the load required for the auxiliary power consumption. Therefore, the auxiliary power consumption does not represent a financial cost to the project. A more detail analysis can be completed once design and value-added engineering is added to the process.

5.1.1 Operating Labor

The OC system will require highly skilled operating and maintenance personnel. Personnel will be required to understand the requirements for:

- Iron-feed A handling and processing
- Steam turbine and auxiliaries
- Environmental Controls
- Water treatment and ZLD

It is assumed that the number of personnel at this manufacturing facility with integrated power generation will be similar to manufacturing facilities of similar size and complexity. For this plant, the personnel includes: one plant manager, one operations manager, one maintenance engineer, one senior engineer, one junior engineer, one engineering technician, two financial accountants, two procurement and warehouse managers, two control room operators per shift, five outside operators per shift, two boiler operators per shift, two train unloading operators at two shifts per weekday, three maintenance mechanics, one I&C technician, two maintenance electricians, four general laborers, and one full-time security person. The fully burdened rates are based on estimated costs associated with an employee. This includes salary, benefits, overhead, and other costs.

5.1.2 Maintenance Material and Labor

Maintenance materials were also estimated using similarly sized projects. The maintenance required throughout the plant involves:

- Annual outages to service the steam turbine, calciner, and other major components
- Outages to inspect and maintain the steam turbine and generator
- Maintenance of the iron-feed A and limestone handling equipment, such as conveyers, crushers, mills, and dust collectors

- Occasional maintenance of the ZLD and water treatment system components, including vapor compressors, centrifuge, and demisting pads
- Maintenance of the packed bed-scrubber, including seal and nozzle replacements
- Maintenance of the pumps, heaters, and BOP
- Improvements to the buildings, pavement, and railing system
- Spares

5.1.3 Consumables

Consumable rates were provided by heat, water, and mass balances. The estimated cost of these consumables was derived from "Mineral Commodity Summaries 2020" by the United States Geological Survey (USGS) as well as factoring based on costs of consumables provided in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." Table 5-1 provides the estimated prices of the elements required for the OC process.

Table 5-1 Estimated Consumables Prices

Stream	Consumable	Price ($/tonne)
1	Iron-feed A	$100
2	Iron-feed B	$30
3	Binder	$300
4	Additive 1	$150
5	Additive 2	$130

5.1.4 Waste Disposal

Waste production rates were provided by equipment vendors or calculated from the heat, water, and mass balances. The cost estimate for removing or disposing of waste was derived from factoring based on costs of consumables provided in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance."

5.1.5 Co-Products and Saleable By-Products

Co-products and by-products production rates were provided by calculated values from the heat, water, and mass balances.

The forced oxidizer would produce gypsum that can be used in many applications including fluxing agencies, fertilizers, fillers in paper and textiles, and retarder in Portland cement. It is assumed that this can be sold at $35/tonne.

The value engineering was not considered for this project based on the progress in technology and current economic considerations.

5.1.6 Fuels

The consumption of natural gas for the fluid bed calciner and HRSG startup is based on calculations and vendor quotes. The price of natural gas is assumed to be $3.30/ MMBTU based on average 2021 natural gas prices (from Henry Hub Natural Gas Spot Price).

5.2 O&M Cost Results

The operating and maintenance costs for the HGCC system are summarized in Table 5-2. The resulting O&M costs are approximately $111,000,000 per year.

Table 5-2 Operating and Maintenance Summary

O&M Costs		
	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**

6 OC Sale Price

The method for calculating the OC sale Price is based on the model and methods described in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." This report makes assumptions provided in Section 8.1. This is used to develop the finance structure in Section 8.2. Both are used to calculate the required OC sale price in Section 8.3.

6.1 Global Economic Assumptions

The 2019 revision of the QGESS document "Cost Estimation Methodology for NETL" makes the following assumptions:

1. Taxes
    a. The Federal Income Tax Rate is 21%, the State Income Tax Rate is 6%, and the Effective Tax Rate (ETR) is 25.74%.
    b. Capital depreciation over 20 years is 150% (declining balance).
    c. There is no Investment Tax Credit.
    d. There is no Tax Holiday.
2. Contracting and Financing Terms
    a. The Contracting Strategy consists of Engineering Procurement Construction Management (owner assumes project risks for performance, schedule, and cost).
    b. Debt Financing is Non-recourse (collateral that secures debt is limited to the real assets of the project).
    c. The Repayment Term of Debt is equal to operational period in formula method.
    d. There is no grace period on debt repayment.
    e. There is no debt reserve fund.
3. Analysis Time Periods
    a. The capital expenditure period is 3 years for natural gas plants and 5 years for coal plants.
    b. The operational period is 30 years.
    c. The economic analysis period is 33 years for natural gas plants or 35 years for coal plants (capital expenditure period plus operational period).
4. Treatment of Capital Costs
    a. The capital cost escalation during the capital expenditure period is 0% real (or 3% nominal).
    b. The distribution of Total Overnight Capital over the capital expenditure (before escalation) is 10%, 60%, 30% for a 3-year period and 10%, 30%, 25%, 20%, 15% for a 5-year period.
    c. There is no working capital.
    d. 100% of the Total Overnight Capital depreciates (actual amounts are likely lower and do not influence results significantly).
5. Escalation of Operating Costs and Revenues
    a. Escalation of revenue, O&M Costs is 0% real (3% nominal).

b. Fuel costs are based on the Quality Guidelines for Energy Systems Studies Fuel Prices for Selected Feedstock in NETL Studies.

6.2 Finance Structure

To evaluate the economic feasibility of the project, a financial structure is established based on market and ownership risks. The cost analysis is developed for both commercial technology in 2021 and advancing technology projected to become commercial in 15 years or more. It can be assumed that they are commercially ready and that there are no risks or tax subsidies associated with any of the technology. The same structure should use real dollars and be applied to all scenarios to compare the technologies. Nominal dollars should be used to evaluate the technologies in various cash flow analyses. The structure will assume a large, financially stable, investor-owned utility (IOU) or merchant plant.

Table 6-1, taken from the 2019 revision of the QGESS document “Cost Estimation Methodology for NETL Assessment of Power Plant Performance,” provides the rates that are used for the financial structure. The nominal case will be used throughout the cost analysis. The nominal case for an IOU estimates an equity of 10%.

Table 6-1 Nominal and Real Rates Financial Structure for Investor-Owned Utility

Type of Security	% Total	Current Dollar Cost	Weighted Average Cost of Capital	After-Tax Weighted Average Cost of Capital
Nominal				
Debt	55%	5%	2.75%	2.04%
Equity	45%	10%	7.25%	4.50%
Total			**10%**	**6.54%**
Real (based on 2.01% average real GDP deflator, 1990-2018)				
Debt	55%	2.94%	1.61%	1.20%
Equity	45%	7.84%	3.53%	3.53%
Total			**5.14%**	**4.73%**

The TASC is expressed in mixed-year, current or real dollars over the entire capital expenditure period. It is calculated from the total overnight cost (TOC) by using the following factors taken from the 2019 revision of the QGESS document “Cost Estimation Methodology for NETL Assessment of Power Plant Performance” shown in Table 6-2. The TASC/TOC chosen for this study was a nominal three-year ratio.

Table 6-2 TASC/TOC Factors

Finance Structure	BBB+[3] or Higher Company	
Capital Expenditure Period	Three Years	Five Years
TASC/TOC *nominal*	1.242	1.289
TASC/TOC *real*	1.093	1.154

The TOC includes "overnight" depreciable and non-depreciable capital expenses that are incurred during the capital expenditure period. This does not include escalation and interest during construction. The factor of TASC to TOC is calculated by adding the cost of escalation to the cost of funding.

Table 6-3 Fixed Charge Rate

Finance Structure	IOU – 30 Years	
Capital Recovery Periods	Three Years	Five Years
FCR Nominal	0.0886	0.0886
FCR Real	0.0707	0.0707

6.3 Selling Price

The annual cost of OC production with heat recovery is shown in Table 6-4. Based on an annual OC production of approximately 990,000 tonnes, the estimated sale price (for 10% equity, 30-year operation) is $152/tonne with heat recovery and $149/tonne without heat recovery.

Table 6-4 Annual Cost of Production and OC Sale Price

Cost of Production		
	With HR	Without HR
Plant Capacity, %	85	85
Total Annual Operation, (hrs.)	7,451	7,451
Total As Spent Cost (TASC), ($)	440,563,000	366,128,000
Fixed Rate Charge (FRC), ($)	0.0886	0.0886
First Year Capital Charge, ($)	39,034,000	32,439,000
First Year Fixed Operating Costs, ($)	17,786,000	15,736,000
First Year Variable Operating Costs, ($)	93,596,000	99,067,000
First Year Fuel Costs, ($)	8,000	1,500
Total Annual Cost, ($)	150,424,000	147,244,000
Annual OC Production, tonne	989,000	989,000
OC Sale Price ($/tonne)	152	149

7 Risk Factors

7.1 Risk Factors

A risk analysis is currently being evaluated for the OC production system and will highlight the areas where the total plant cost is most probable.

As discussed in Section 1 of this report, the contingencies of areas that are considered emerging technologies include higher-process contingencies and, in some areas, engineering compared to the common commercialized technologies. We also included costs for several systems noted in the risk management discussions. The following list describes a summary of cost considerations based on risk management:

- High temperatures and flow rates in currently existing equipment that require evaluation or possible pilots (ex. high-temp fluid bed calciner)
- Unknow purities of consumables that could affect process efficiency
- Unknown water quality that could affect ZLD design
- Levels of SO_2 that increase size and cost of environmental controls

8 Sensitivity Analysis

8.1 Effects of Capital Costs

Figure 8-1 and Figure 8-2 provide illustrations of how the sale price of OC varies with capital costs. The sale price of OC varies from $110/tonne to $230/tonne when capital costs vary from $0 to approximately $800 million.

Figure 8-1 OC Sale Price Sensitivity Based on Capital Cost with Heat Recovery

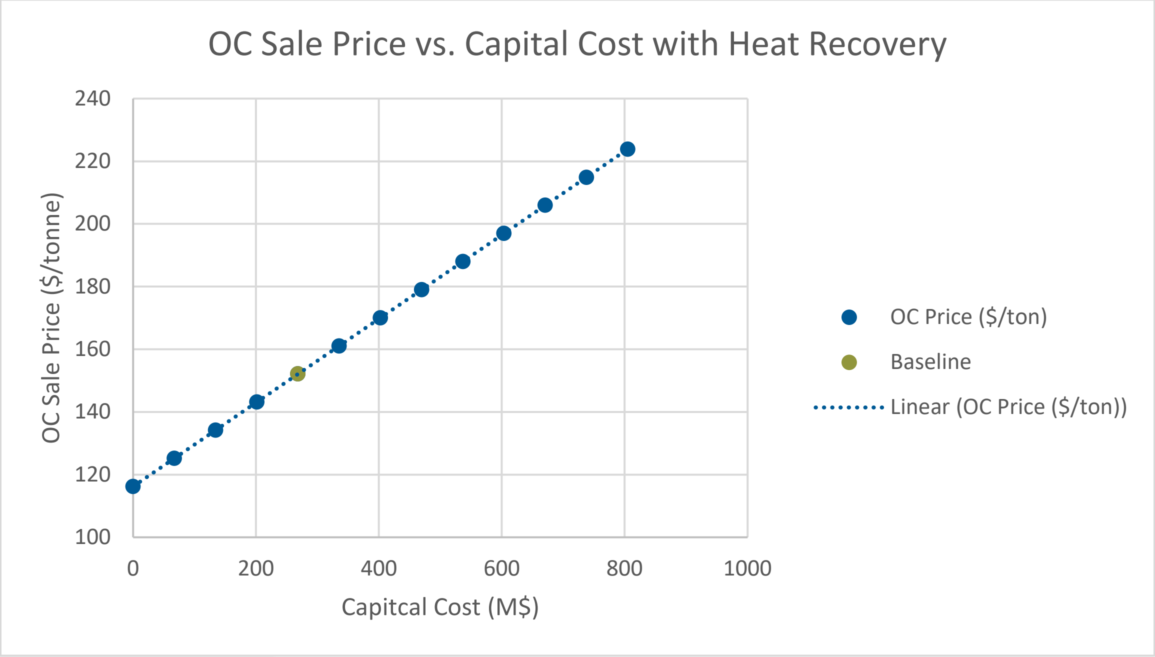
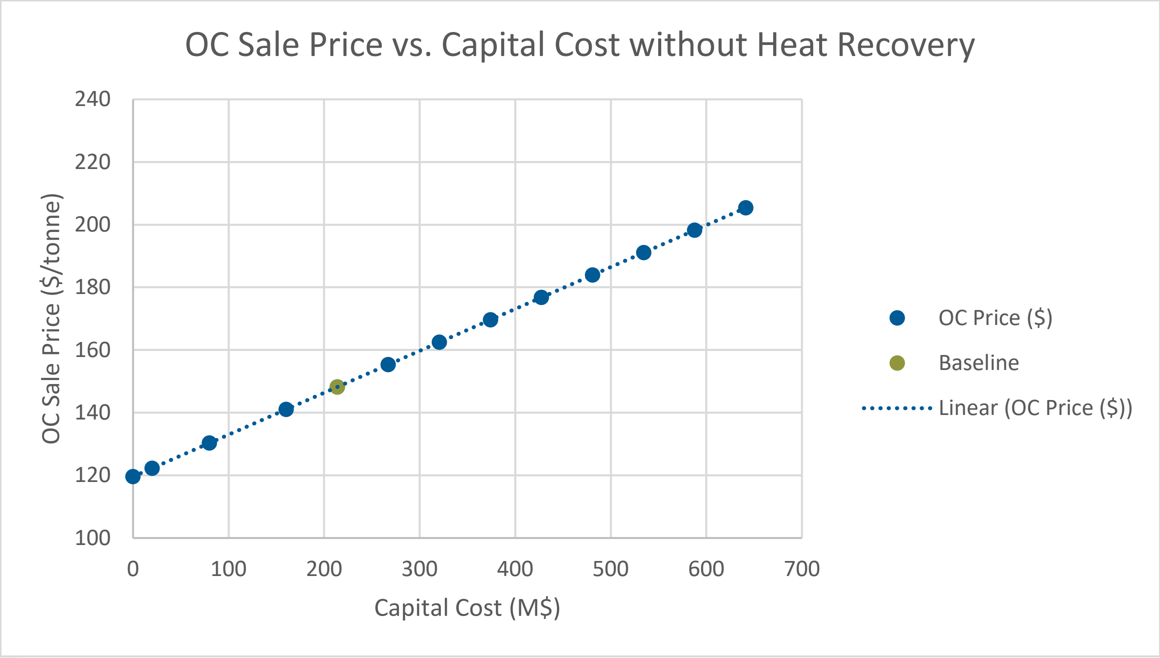


Figure 8-2 OC Sale Price Sensitivity Based on Capital Cost without Heat Recovery



8.2 Effects of O&M Costs

Figure 8-3 and Figure 8-4 provide illustrations of how the sale price of OC varies with O&M costs. The sale price of OC varies from $30/tonne to $390/tonne when O&M costs vary from $0 to approximately $350 million.

Figure 8-3 OC Sale Price Sensitivity Based on O&M Cost with Heat Recovery

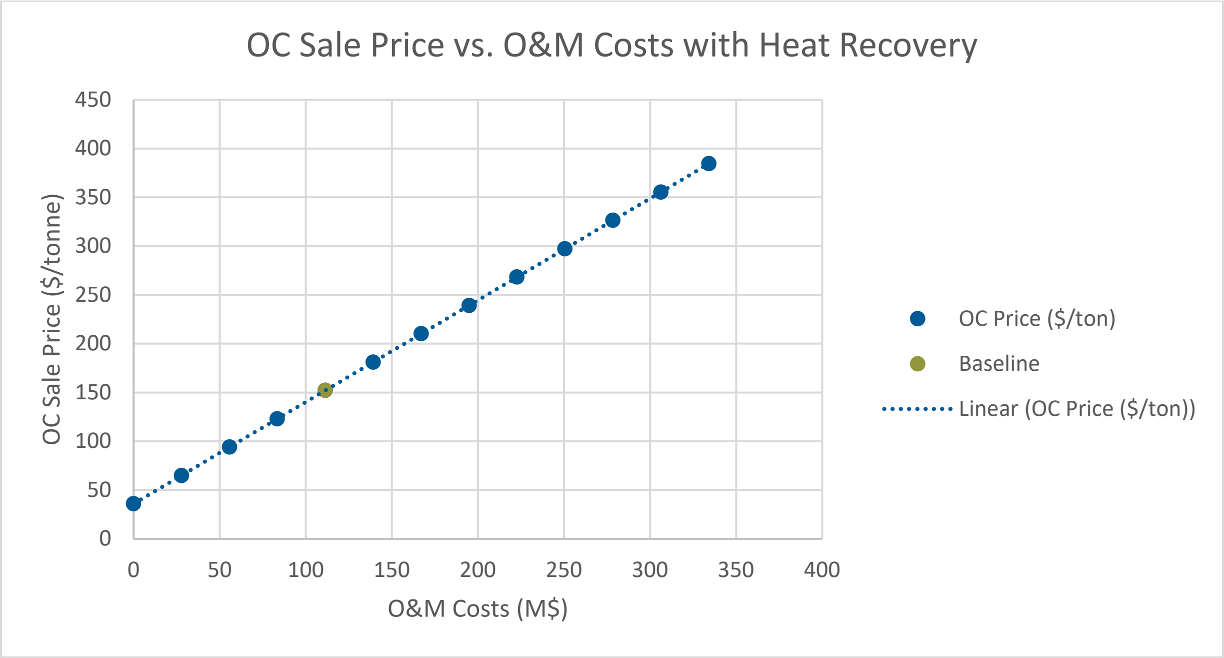
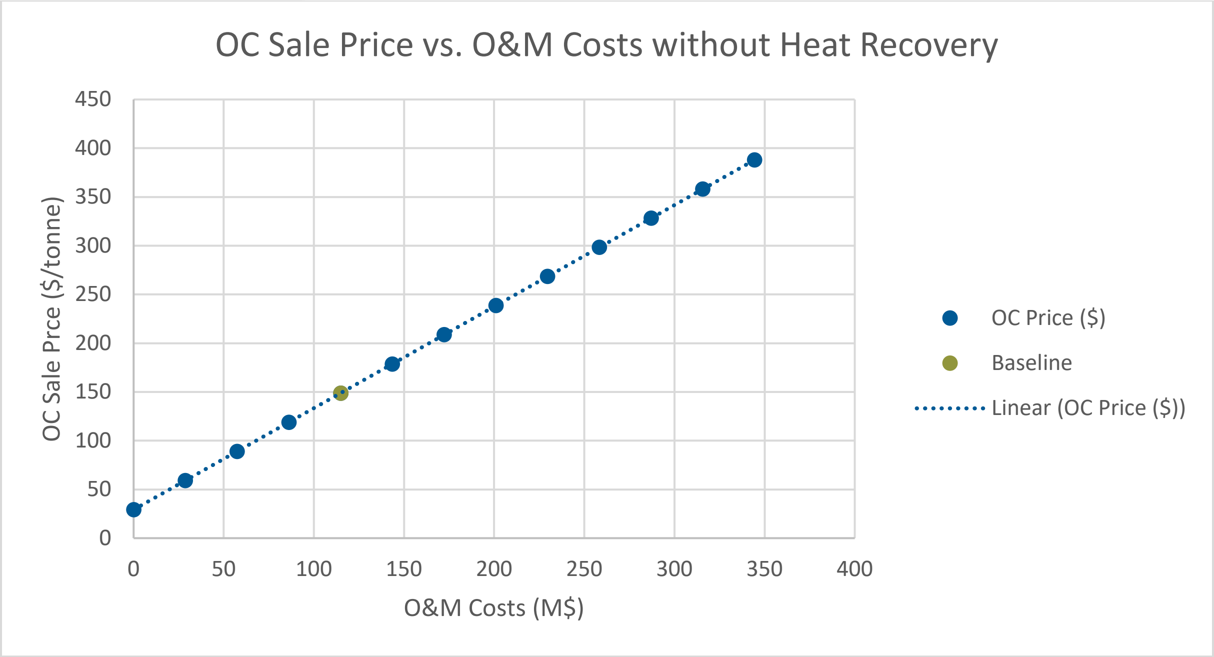


Figure 8-4 OC Sale Price Sensitivity Based on O&M Cost without Heat Recovery



8.3 Effects of Stream 1 Costs

Figure 8-5 and Figure 8-6 provide illustrations of how the sale price of OC varies with the sale price of stream 1. The sale price of OC varies from $120/tonne to $230/tonne when the price of stream 1 varies from $0/tonne to approximately $300/tonne.

Figure 8-5 OC Sale Price Sensitivity Based on Stream 1 Cost with Heat Recovery

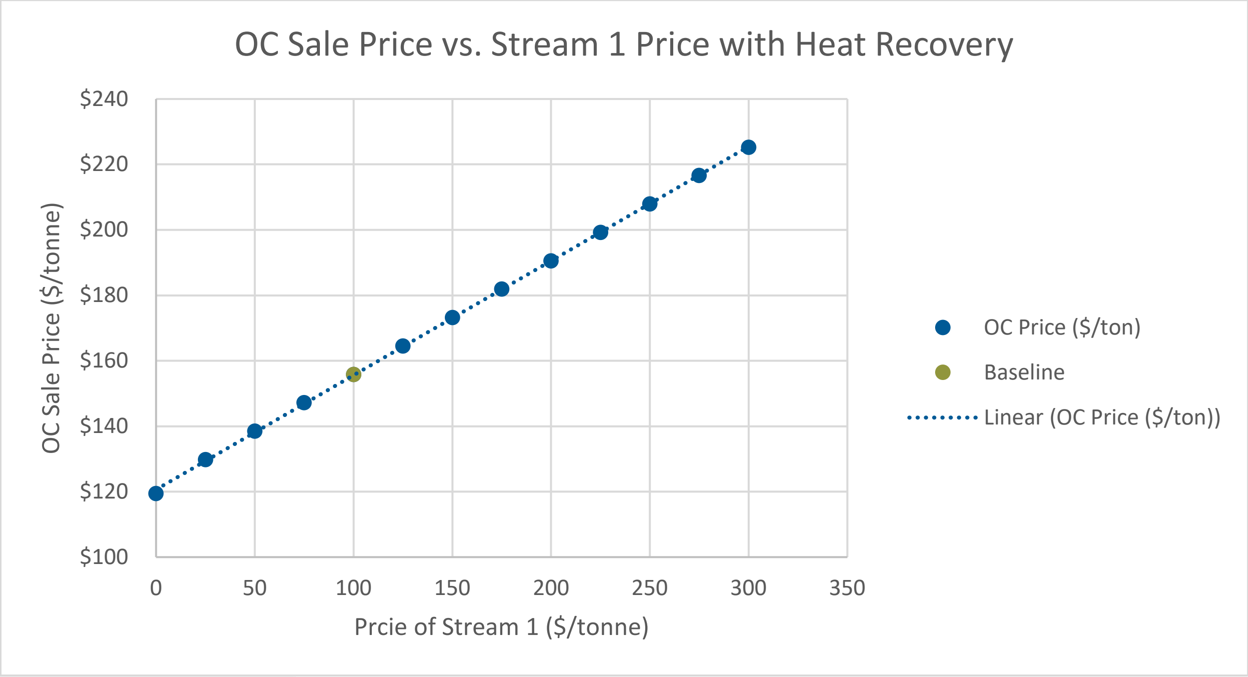
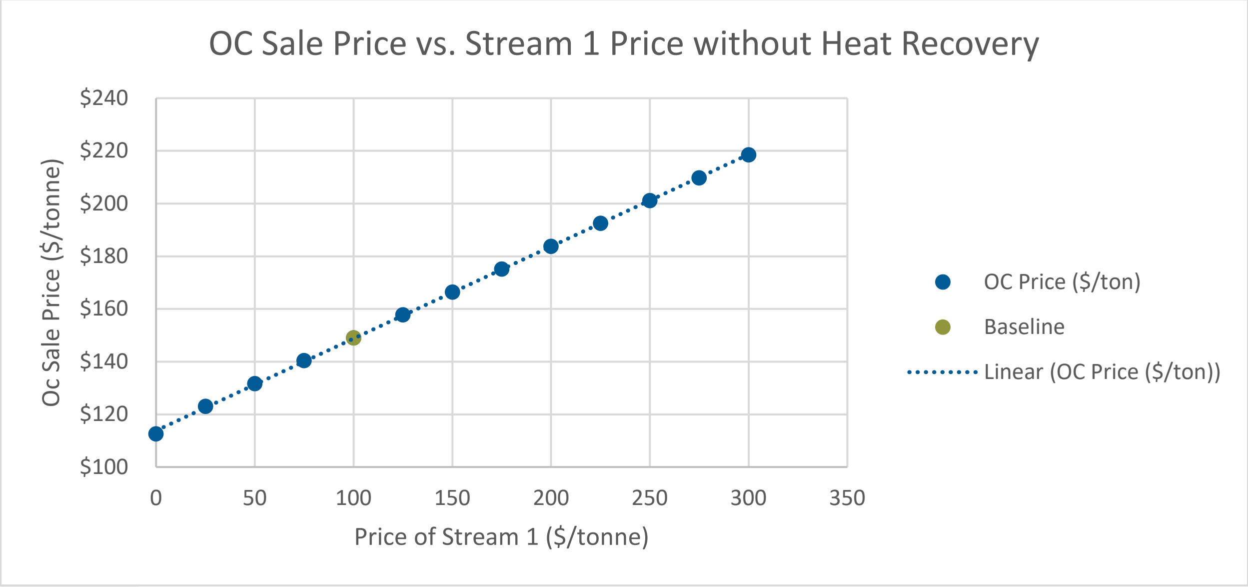


Figure 8-6 OC Sale Price Sensitivity Based on Stream 1 Cost without Heat Recovery



8.4 Effects of Stream 2 Costs

Figure 8-7 and Figure 8-8 provide illustrations of how the sale price of OC varies with the sale price of stream 2. The sale price of OC varies from $130/tonne to $180/tonne when the price of stream 2 varies from $0/tonne to approximately $90/tonne.

Figure 8-7 OC Sale Price Sensitivity Based on Stream 2 Cost with Heat Recovery

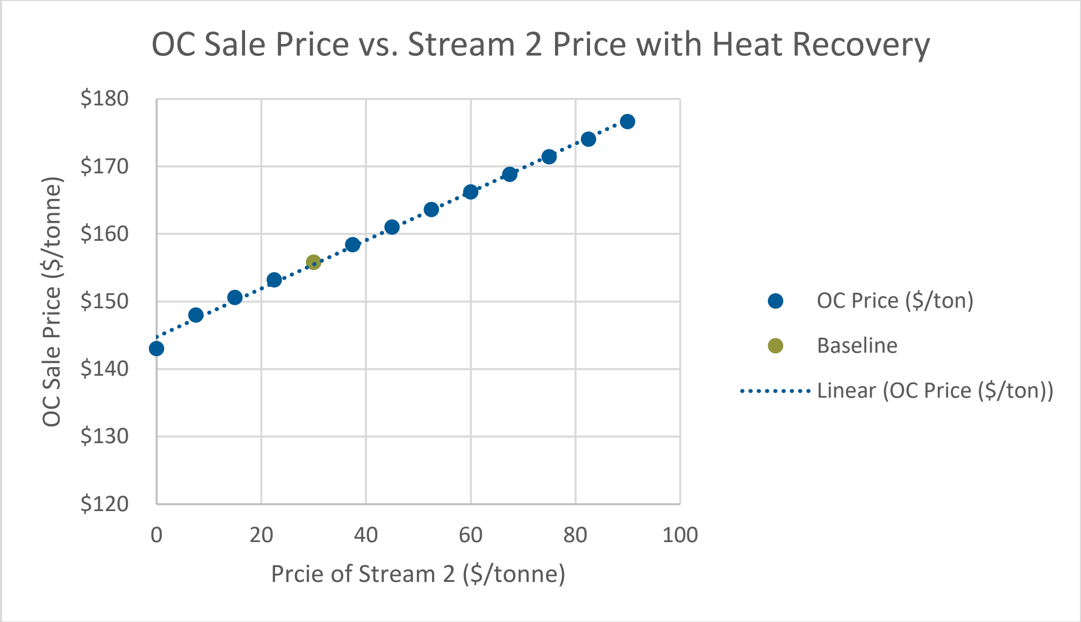
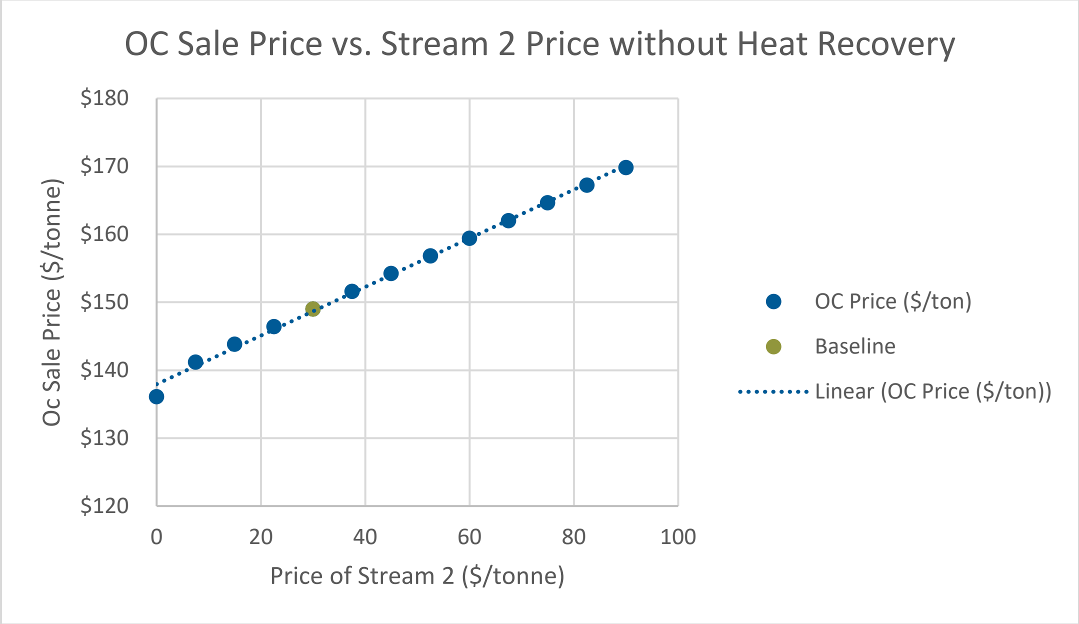


Figure 8-8 OC Sale Price Sensitivity Based on Stream 2 Cost without Heat Recovery



8.5 Effects of Stream 3 Costs

Figure 8-9 and Figure 8-10 provide illustrations of how the sale price of OC varies with the sale price of stream 3. The sale price of OC varies from $120/tonne to $220/tonne when the price of stream 3 varies from $0/tonne to approximately $900/tonne.

Figure 8-9 OC Sale Price Sensitivity Based on Stream 3 Cost with Heat Recovery

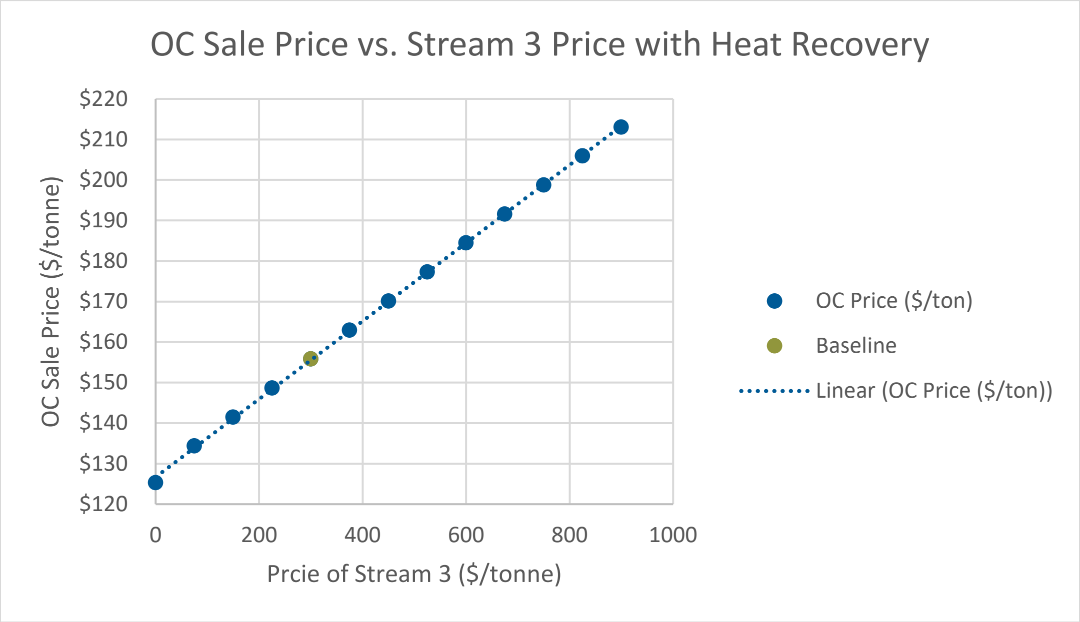
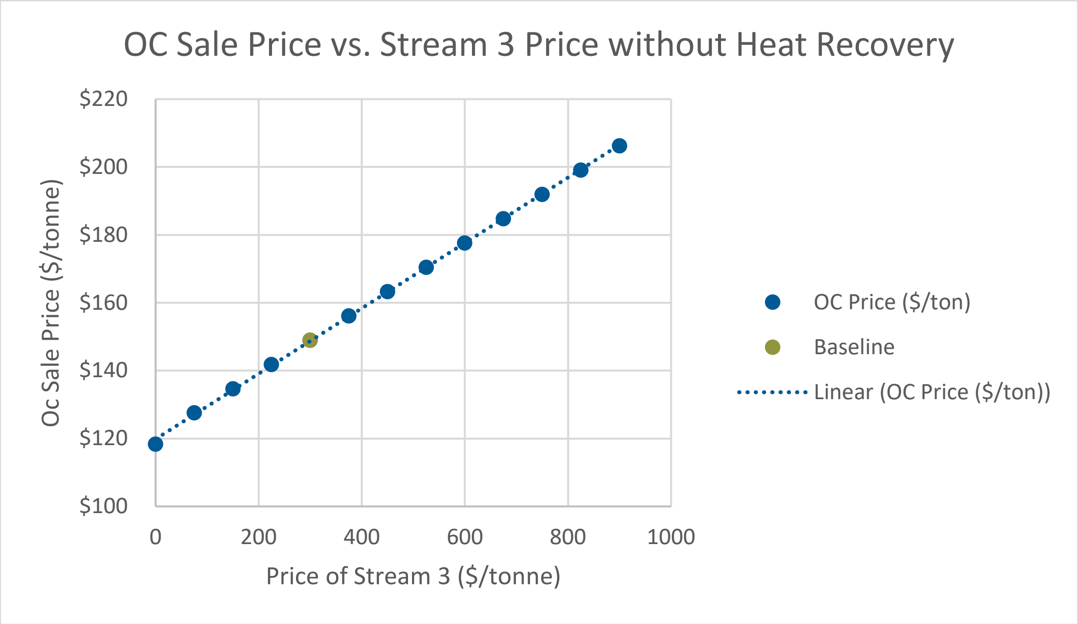


Figure 8-10 OC Sale Price Sensitivity Based on Stream 3 Cost without Heat Recovery



8.6 Effects of Stream 4 Costs

Figure 8-11 and Figure 8-12 provide illustrations of how the sale price of OC varies with the sale price of stream 4. The sale price of OC varies from $140/tonne to $160/tonne when the price of stream 4 varies from $0/tonne to approximately $450/tonne.

Figure 8-11 OC Sale Price Sensitivity Based on Stream 4 Cost with Heat Recovery

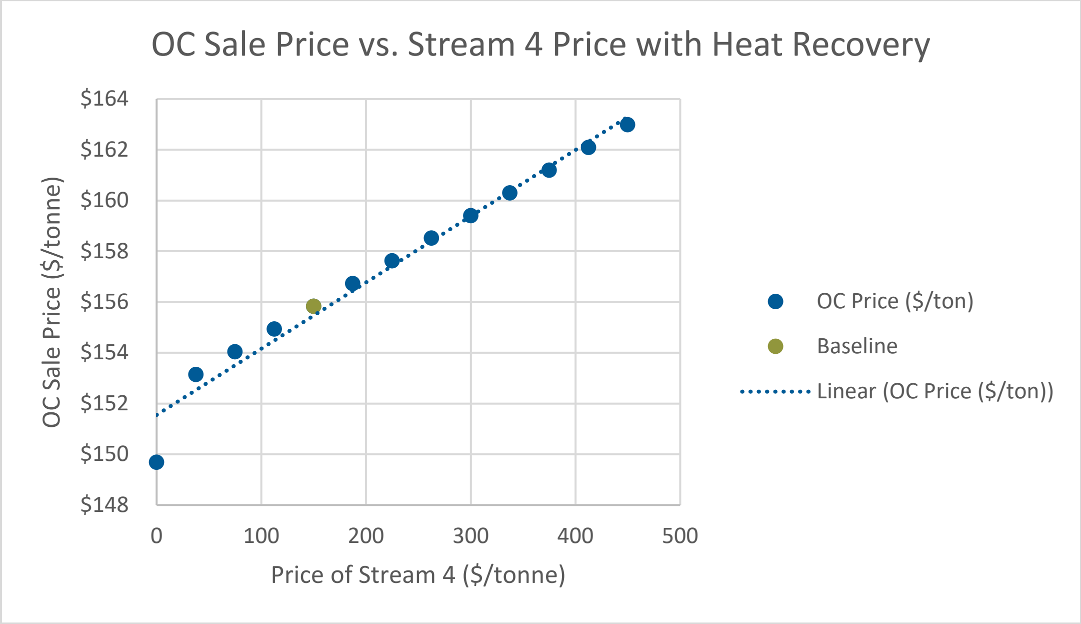
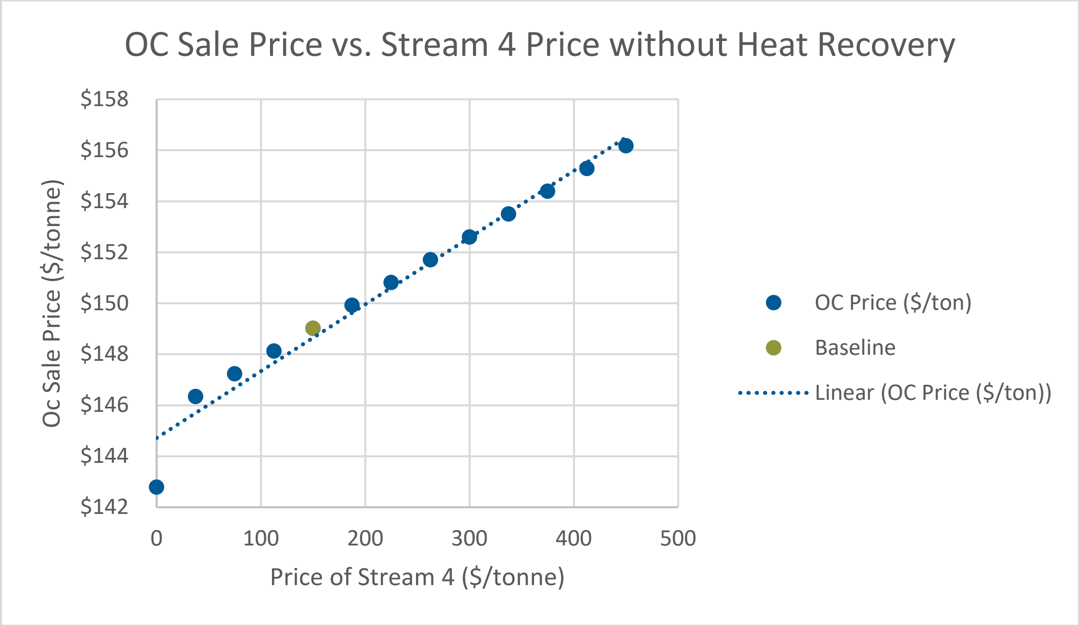


Figure 8-12 OC Sale Price Sensitivity Based on Stream 4 Cost without Heat Recovery



8.7 Effects of Stream 5 Costs

Figure 8-13 and Figure 8-14 provide illustrations of how the sale price of OC varies with the sale price of stream 5. The sale price of OC varies from $130/tonne to $180/tonne when the price of stream 5 varies from $0/tonne to approximately $400/tonne.

Figure 8-13 OC Sale Price Sensitivity Based on Stream 5 Cost with Heat Recovery

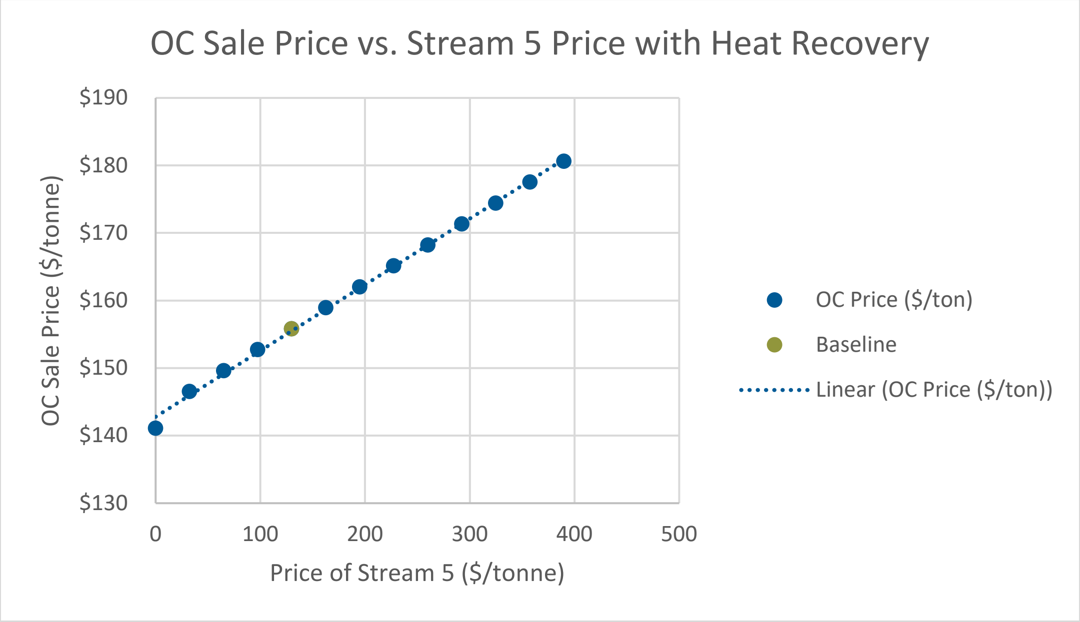
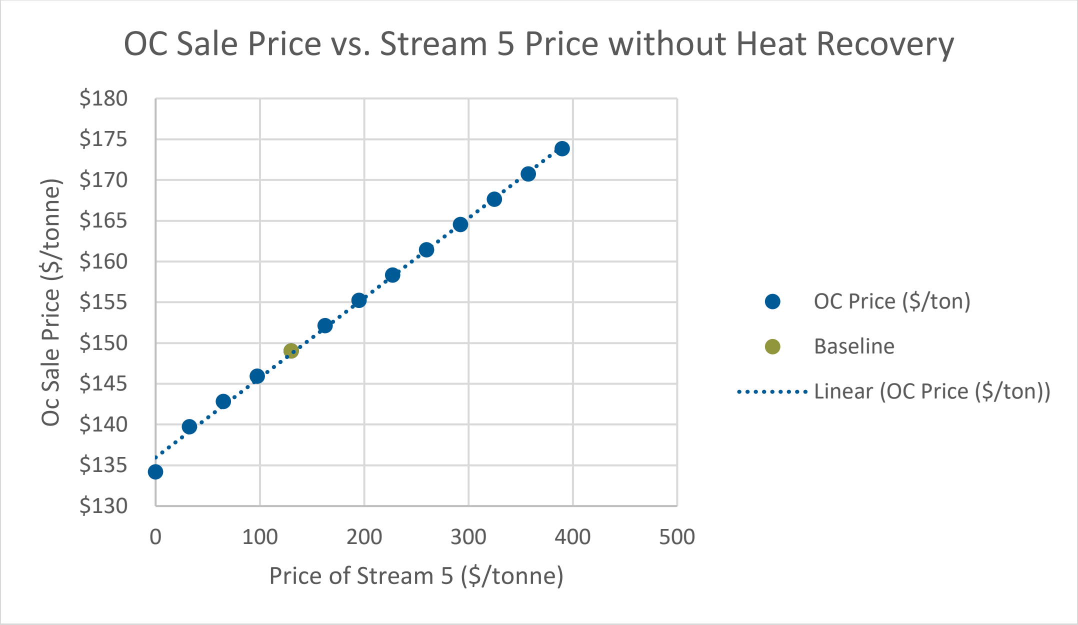


Figure 8-14 OC Sale Price Sensitivity Based on Stream 5 Cost without Heat Recovery



8.8 Effects of FRC

Figure 8-15 and Figure 8-16 provide illustrations of how the sale price of OC varies with the Fixed Rate Charge (FRC). The sale price of OC varies from $110/tonne to $230/tonne when the FRC varies from 0 to approximately 0.027.

Figure 8-15 OC Sale Price Sensitivity Based on FRC with Heat Recovery

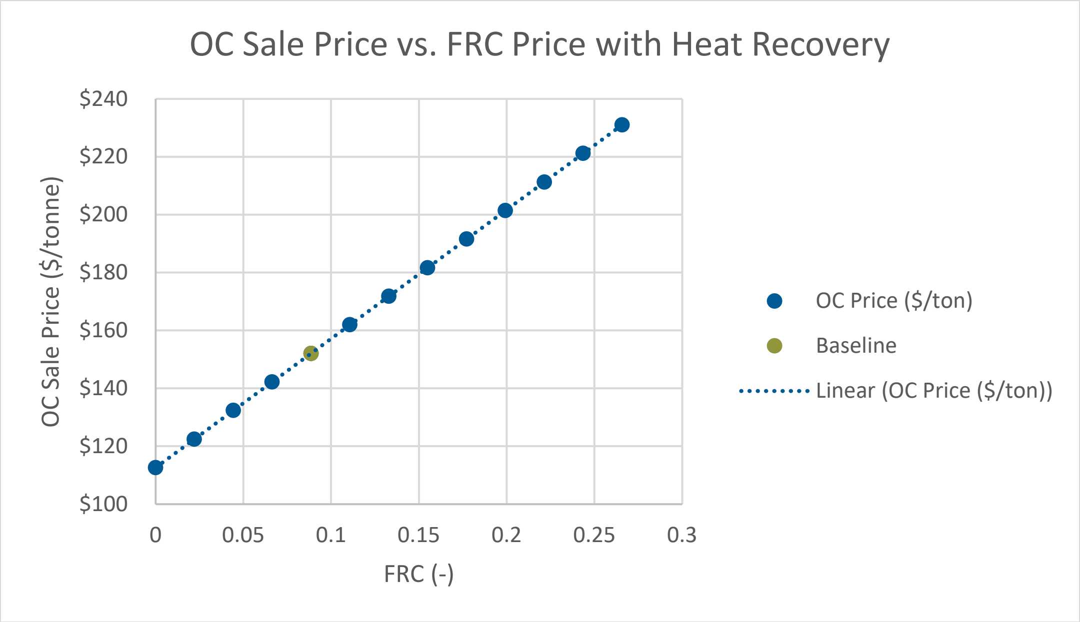
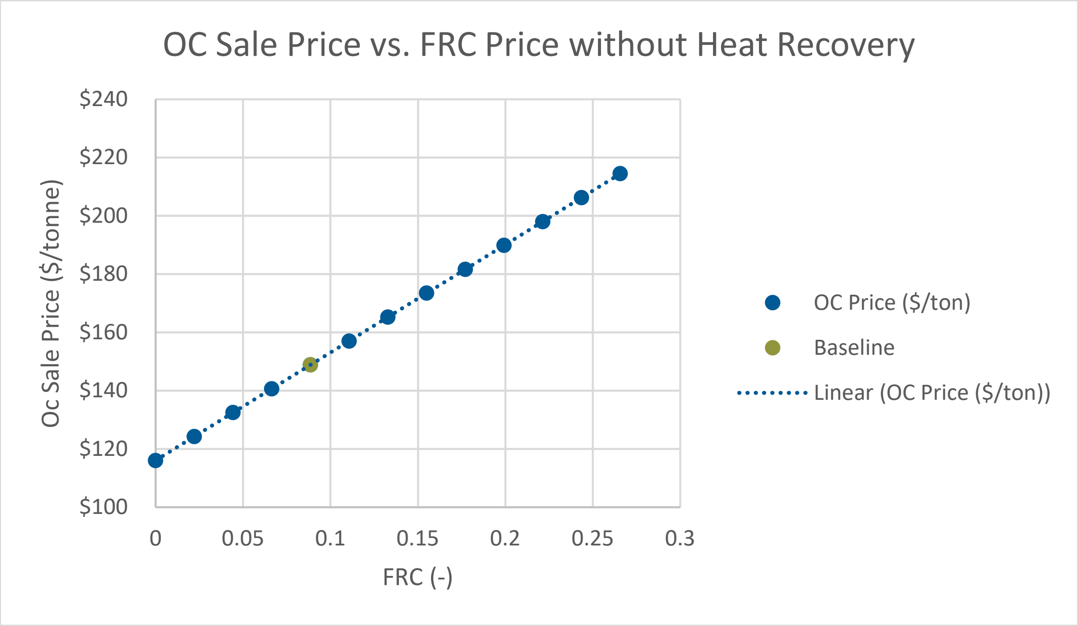


Figure 8-16 OC Sale Price Sensitivity Based on FRC without Heat Recovery



8.9 Effects of Costs on OC Sale Price

Figure 8-17 and Figure 8-18 provide illustrations of how the sale price of OC varies (in percentage of base price) with the percent change in other costs. The sale price of OC varies most greatly with the change in O&M costs, to which the price of stream 1 and stream 2 contribute more than then the price of stream 3, stream 4, and stream 5.

Figure 8-17 OC Sale Price Sensitivity Based on Percent Differences with Heat Recovery

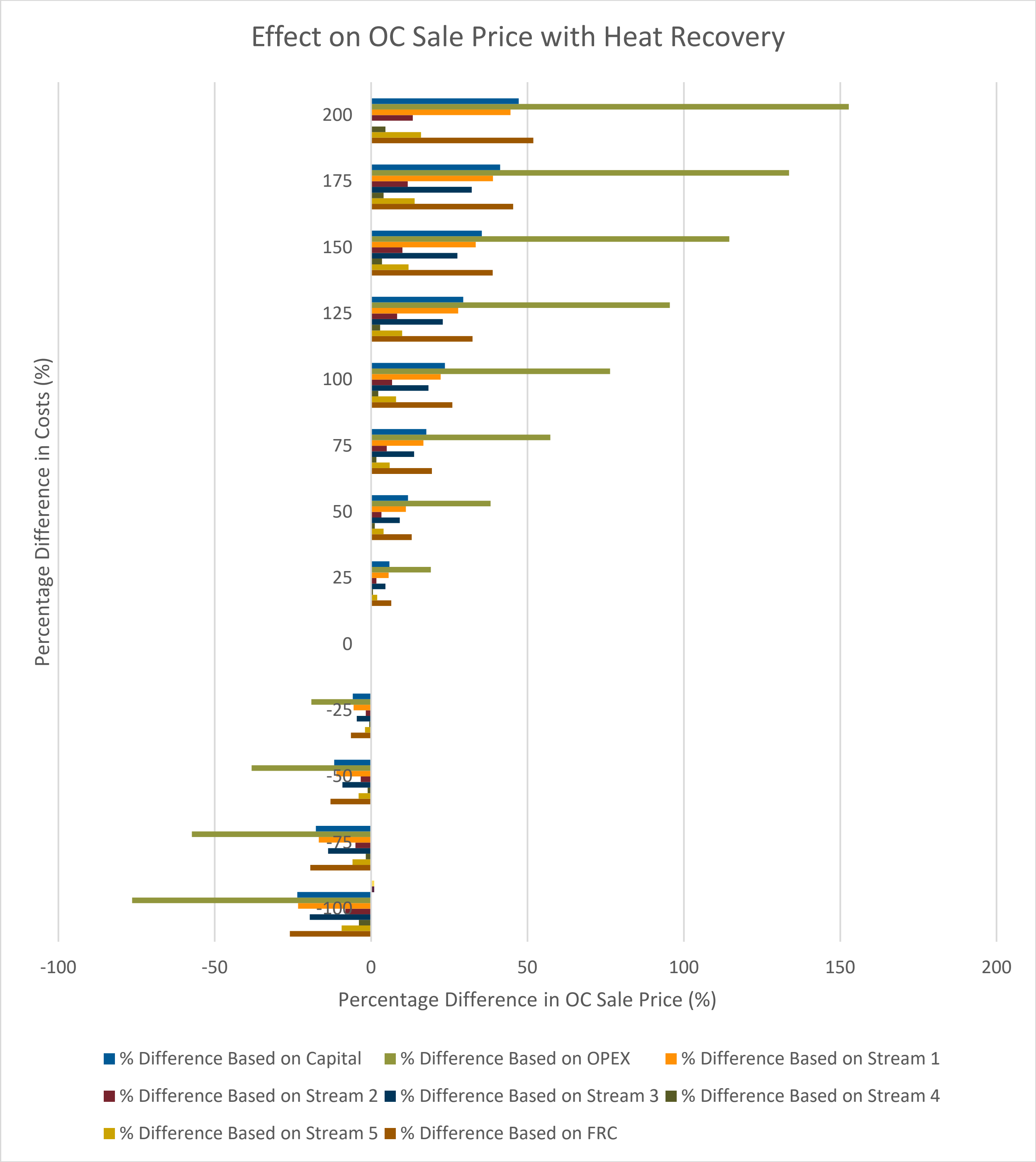
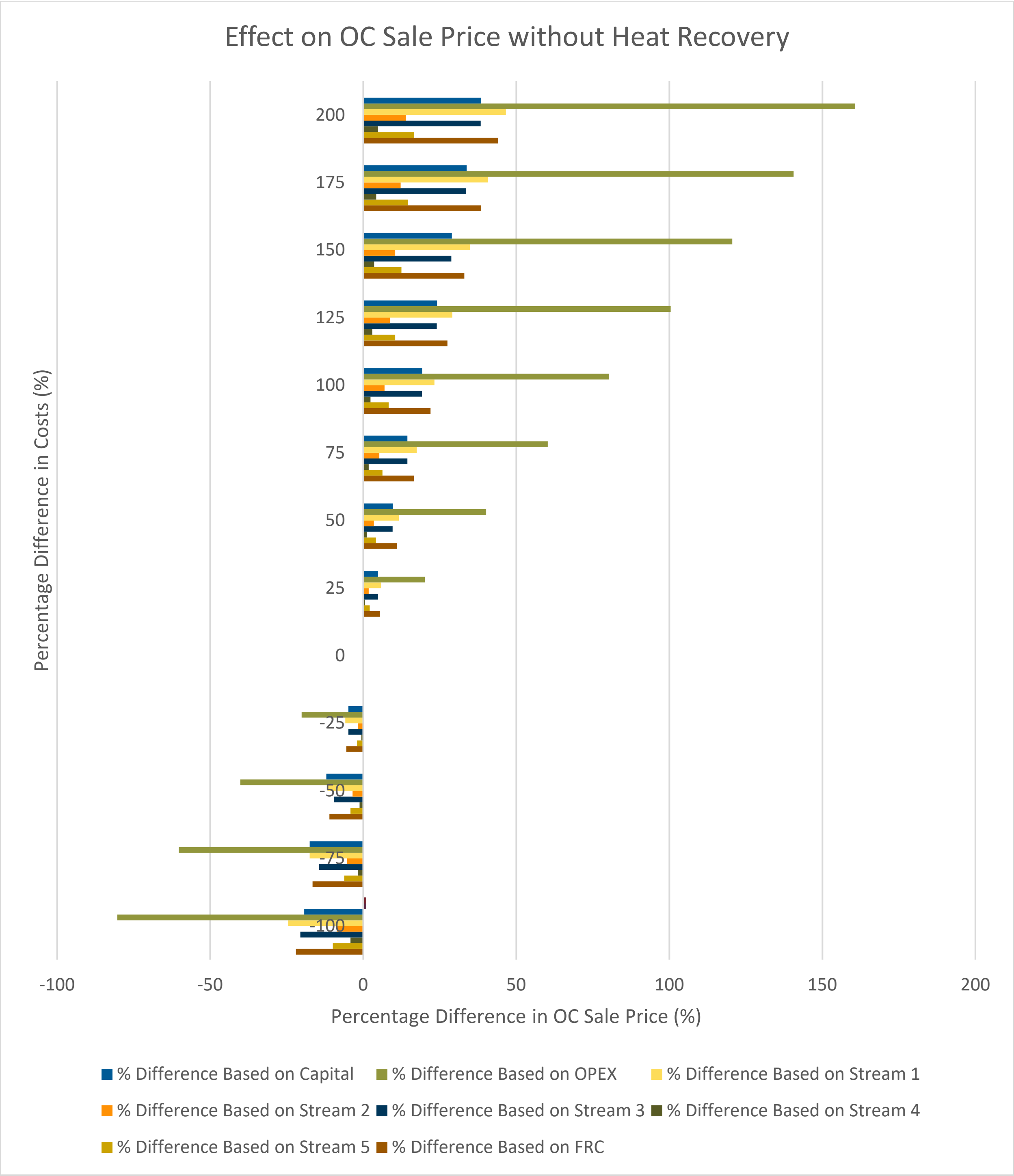


Figure 8-18 OC Sale Price Sensitivity Based on Percent Differences without Heat Recovery



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