MODELING, OPTIMIZATION AND ECONOMIC EVALUATION

OF RESIDUAL BIOMASS GASIFICATION

A Thesis

by

ADAM MICHAEL GEORGESON

Submitted to the Office of Graduate Studies of
Texas A&M University
in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

December 2010

Major Subject: Chemical Engineering

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Approved by:

Chair of Committee,	Mahmoud El-Halwagi
Committee Members,	M. Sam Mannan
	Sergiy Butenko
Head of Department,	Michael Pishko

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ABSTRACT

Modeling, Optimization and Economic Evaluation of Residual Biomass Gasification.

(December 2010)

Adam Michael Georgeson, B.S., University of Alberta

Chair of Advisory Committee: Dr. Mahmoud El-Halwagi

Gasification is a thermo-chemical process which transforms biomass into valuable synthesis gas. Integrated with a biorefinery it can address the facility's residue handling challenges and input demands. A number of feedstock, technology, oxidizer and product options are available for gasification along with combinations thereof.

The objective of this work is to create a systematic method for optimizing the design of a residual biomass gasification unit. In detail, this work involves development of an optimization superstructure, creation of a biorefining scenario, process simulation, equipment sizing & costing, economic evaluation and optimization. The superstructure accommodates different feedstocks, reactor technologies, syngas cleaning options and final processing options. The criterion for optimization is annual worth.

A biorefining scenario for the production of renewable diesel fuel from seed oil is developed; gasification receives the residues from this biorefinery. Availability of Soybeans, Jatropha, Chinese Tallow and woody biomass material is set by land use within a 50-mile radius. Four reactor technologies are considered, based on oxidizer

type and operating pressure, along with three syngas cleaning methods and five processing options.

Results show that residual gasification is profitable for large-scale biorefineries with the proper configuration. Low-pressure air gasification with filters, water-gas shift and hydrogen separation is the most advantageous combination of technology and product with an annual worth of $9.1 MM and a return on investment of 10.7%. Low-pressure air gasification with filters and methanol synthesis is the second most advantageous combination with an annual worth of $9.0 MM.

Gasification is more economic for residue processing than combustion or disposal, and it competes well with natural gas-based methanol synthesis. However, it is less economic than steam-methane reforming of natural gas to hydrogen. Carbon dioxide credits contribute to profitability, affecting some configurations more than others. A carbon dioxide credit of $33/t makes the process competitive with conventional oil and gas development.   Sensitivity analysis demonstrates a 10% change in hydrogen or electricity price results in a change to the optimal configuration of the unit. Accurate assessment of future commodity prices is critical to maximizing profitability.

DEDICATION

I would like to devote my academic work to my family and especially my parents Michael & Martha Georgeson. They have always encouraged me to pursue my dreams to the fullest extent. I am very grateful for their assistance and support.

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I would like to thank the Chevron Corporation, both for the insight into biorefinery design problems and for the provision of funding to the research group. I also want to extend my gratitude to my former co-workers at OPTI Canada for teaching me greatly about the engineering profession.

Finally, I wish to extend my thanks to my classmates and friends who have helped me through the program and made my time in College Station both enjoyable and fulfilling.

NOMENCLATURE

A_C	Annualized capital cost
A_D	Annual depreciation
A_F	Annual after-tax profit
A_R	Annual revenue
A_P	Annual expenditure
A_T	Annual tax
A_W	Annual worth
A_{WC}	Annual working capital
C_i^S	Capital cost of solids handling section i
C_j^T	Capital cost of reactor section j
C_j^U	Capital cost of gas cleaning section j
C_k^V	Capital cost of final processing option k
C	Capital cost, total
Diff	Differential between electricity and natural gas cost and selling price
F_k^E	Flowrate to electrical generation from processing option k
F_k^F	Flowrate to fuel gas from processing option k
$F_{j,q}^A$	Flowrate of gasifying agent q to reactor j
F_j^P	Flowrate of steam formed by heat recovery from reactor j
F_i^S	Flowrate from source i

$F_{i,q}^{S}$	Flowrate of component q from source i
$F_{i,j}^{ST}$	Flowrate of a stream connecting source i to reactor j
$F_{i,j,q}^{ST}$	Flowrate of component q between source i and reactor j
F_{j}^{T}	Flowrate to reactor j
$F_{j,q}^{T}$	Flowrate of component q to reactor j
F_{j}^{U}	Flowrate from reactor j
$F_{j,q}^{U}$	Flowrate of component q from reactor j
$F_{k,m}^{UV}$	Flowrate of a stream connecting cleaning section j to processing option k
$F_{k,m,q}^{UV}$	Flowrate of component q from cleaning section j to processing option k
F_{k}^{V}	Flowrate to final processing option k
$F_{k,q}^{V}$	Flowrate of component q to final processing option k
F_{k}^{W}	Flowrate from final processing option k
$F_{k,q}^{W}$	Flowrate of component q from final processing option k
$F_{j,q}^{Z}$	Flowrate of component q removed by the cleaning section j
$G_{j,q}^{T}$	Molar flowrate of component q to reactor & cleaning section j
$G_{j,q}^{U}$	Molar flowrate of component q from reactor & cleaning section j
HHV	Higher heating value
h	Scale up factor
i	Index variable for source
I	Maximum for index variable i
IR	Interest rate

j	Index variable for reactor and gas cleaning
J	Maximum for index variable j
k	Index variable for final processing option
K	Maximum for index variable k
LHV	Lower heating value
M_q	Price of raw material, commodity or utility
MP	Mole percent oxygen in air
MW_q	Molecular weight of component q
η_e	Efficiency of energy capture
η_t	Efficiency of electrical generation turbine
N	Project life
NR	Inflation rate
NPV	Net present value
os	On-stream factor
PSA	Pressure swing adsorption
$P^{o\&m}$	Operating cost for plant operations and maintenance
P_i^S	Operating cost for source
P_j^T	Operating cost for reactor
P_j^U	Operating cost for cleanup
P_k^V	Operating cost for final processing option
q	Index variable for chemical species
Q	Maximum for index variable q

r	Index variable for chemical element or species in feed
R_i^S	Operating revenue from source
R_j^T	Operating revenue from reactor
R_j^U	Operating revenue from cleanup
R_k^V	Operating revenue from final processing option
ROI	Return on investment
RPP	Payback period
S_i^E	Binary variable to indicate existence of source i
$T_{j,O2}^A$	Binary variable to indicate pure oxygen as the gasifying agent
$T_{j,H2O}^A$	Binary variable to indicate steam as the gasifying agent
T_j^D	Oxygen demand for reactor j
T_j^E	Binary variable to indicate existence of reactor j
TR	Tax rate
$U_{j,z}^E$	Binary variable to indicate cleaning type z for cleaning step j
V_k^E	Binary variable to indicate existence of processing option k
WC	Working capital
$X_{i,q}^S$	Mass fraction of component q in source i outlet flow
$X_{j,q}^T$	Mass fraction of component q in reactor j outlet flow
$Y_{j,q}^T$	Yield of component q from reactor j
$Y_{j,q}^U$	Yield of component q from cleaning step j
Y_{H2}^R	Yield of hydrogen from steam shift reactor in option k

Y_{H2}^{PSA} Yield of hydrogen from PSA in option k

Y_{Me}^{D} Yield of methanol from the distillation column in option k

Y_{Me}^{R} Yield of methanol from the reactor in option k

z Index variable for cleaning method

Z Maximum for index variable z

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1. INTRODUCTION

Gasification is a thermo-chemical conversion of solids into a synthetic gas, which has value as a fuel and as a chemical feedstock. Biomass gasifiers have been used for decades in small-scale applications, to provide synthetic fuel gas for heating and electrical production. A number of large-scale biomass gasifiers, those with capacity over 50 t/d, have been built and are currently operational. However many facilities of all scales require subsidies or specific legislative policies to operate at desirable economic performance[1]. As stand-alone facilities, they cannot leverage process integration to improve their performance.

By combining a biorefinery and a gasification unit together, waste residue can be transformed into needed products, potentially at a major economic and environmental benefit. The economics of an integrated biorefinery are expected to be better than a stand-alone facility because such synergies can be captured. The integrated biorefinery produces transportation fuels and helps address the urgent issue around the exhaustion of fossil-energy resources. Finally, biofuels are a promising option to reduce carbon dioxide emissions from transportation fuels. Certain gasification configurations further advance this promise by yielding a concentrated stream of carbon dioxide which can be easily sequestered rather than vented into the atmosphere.

With this opportunity, there is a need to design the most economic biorefinery. Economics are dependent on capacity, feedstock, technology and product selections

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This thesis follows the style of the *AIChE Journal*.

within the gasification unit. However, the best choice to make for each selection is not intuitive for the design engineer, which could lead to suboptimal configurations and lower economic returns, or even an avoidance of gasification altogether.

To begin, this thesis investigates the design of biorefineries and gasification units and reviews literature relevant to the optimization of such processes. Next, the specific gasification optimization problem is defined and a superstructure created for its solution. Details and equations for the targeting, process simulation, equipment sizing & costing and mathematical optimization steps are provided. A subsequent economic analysis identifies key variables for profitability and provides a comparison to other biorefining options. A realistic case study is developed for a biorefinery located in Texas and producing biodiesel or green diesel from seed oil. The case study is solved using the optimization procedure to identify possible outcomes and generate relevant financial data. This illustrates the power of the optimization approach to find the most economic process design under changing commodity prices, carbon dioxide regimes and facility sizes.

2. BACKGROUND

2.1 BIOREFINING

A biorefinery contains a number of conversion processes to convert biomass into fuels, chemicals and electricity. Around 170 biorefineries are operating in the United States, producing approximately 9 billion gallons of liquid biofuels annually[2]. The number of bio-refineries and the quantity of biofuels produced is expected to greatly increase in the near future due to declining fossil fuel supplies, increased production costs for remaining fossil fuels, and a federal mandate to produce 36 billion gallons of liquid fuels annually by the year 2022.

One common pathway in the biorefinery extracts oil from specially grown oil seeds using the process shown in Figure 1. Through hydrogenation, the oil can be processed into renewable diesel and jet fuels. Alternatively the oil can be processed through transesterfication to make biodiesel, a pathway which requires methanol. In both cases, unwanted residue is composed of leftover seeds and plant material from the oil seed extraction unit.

Another common pathway in the biorefinery is to use digestion and fermentation to transform sugars and starches into methane and alcohols, respectively. Here, residue consists of cellulose, hemicelluloses, lignin and other materials which the active cultures cannot consume[3]. As fermentation residues have higher water content and lower energy content than oil seed extraction residues, both of which are undesirable, emphasis is placed on the application of gasification to oil seed-based biorefineries.
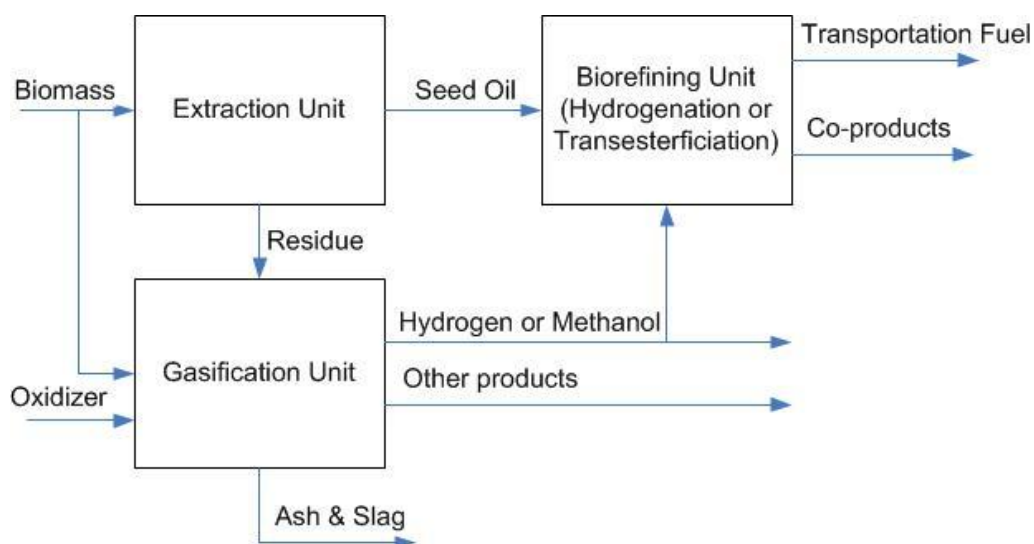
Figure 1. Sample seed oil biorefinery.

2.2 FEED

The feed stock to the gasification unit is primarily residue from the extraction unit. To more definitively characterize this residue, it is necessary to investigate the kind of biomass delivered to the biorefinery. Certain crops have high oil seed yields and characteristics that allow them to grow prolifically in the climate of the United States. Soybeans are the most commercially successful example, as they account for 90% of US seed oil production and have annual sales of almost $30 billion[4]. Based on this information alone, it is clear that soybeans should be considered as a biorefinery feedstock.

Jatropha and Chinese Tallow are promising oil seeds, with higher yields and lower water consumption per acre than soybeans. Researchers predict that high-yield species like Jatropha and Chinese Tallow will fuel the second generation of biorefineries

due to constraints on land availability and the desire to avoid using edible crops as fuel sources[2]. Many other oil seeds such as varieties of cottonseed, peanut, canola and safflower are also suitable biorefinery feed stocks but these have somewhat weaker cases for production and yield. Other feedstock alternatives for gasification might include algae residue, municipal solid waste, fermentation waste and animal manure. Virgin biomass (not a residue from other process units) can also be used as feedstock, either as a make-up stream to meet a desired mass flow rate or to influence the carbon-hydrogen ratio in a desired direction.

The extraction process most commonly used for oil seeds is crushing, hexane-based solvent extraction, and desolventizing-toasting to recover solvent from the residue. A block flow diagram of the extraction process is provided in Figure 2. Typically the desolventizer-toaster is operated near atmospheric pressure and slightly above the boiling point of hexane[5], which is insufficient to remove most water from the residue. Far from being a flaw, this is by design, since a desolventizer-toaster that removes a large amount of water requires more energy and has higher operating costs. Excessive water content can impede the gasification feed system, however this problem is solved by designing the feed system to the same standards as the desolventizer-toaster.
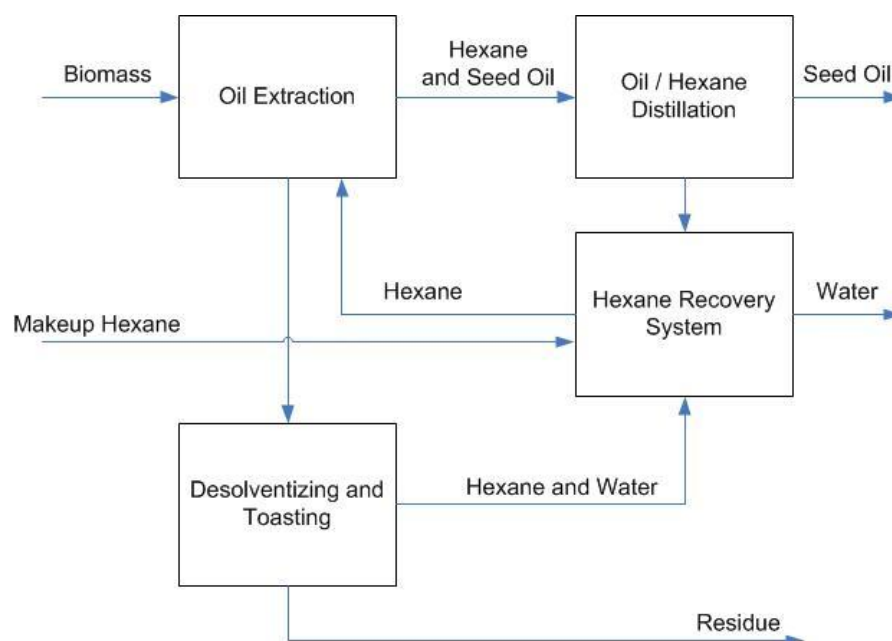
Figure 2. Flow diagram of the conventional solvent extraction process.

2.3 GASIFICATION

A gasification unit consists of three main blocks: the main reactor and associated heat transfer equipment, syngas cleaning and treating, and the final processing option to create or separate products out of the syngas stream. The configuration of these blocks is shown in Figure 3 along with feeds and product streams.

A recent survey indicates that thirteen different gasification reactors are currently available on a commercial basis from design and manufacturing firms[6]. Numerous other technologies are in the development phase, offering decreased formation of slag and tar and better handling of any slag and tar that does form as an unwanted byproduct of the reactions. Avoiding slagging reactions and facilitating the set of reactions[7] shown in Table 1 is the key aspect of reactor development.
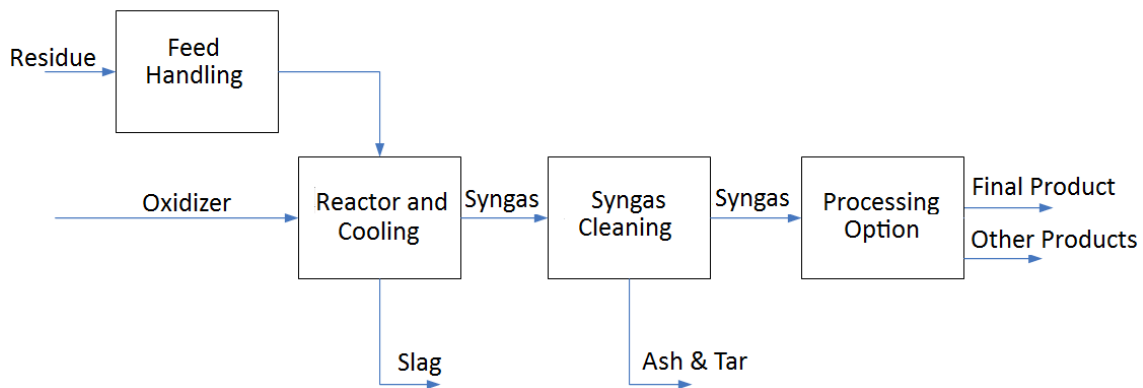
Figure 3. Block flow diagram for a sample gasification unit.

Table 1. Main Gasification Reactions

Type	Reaction	
Carbon Monoxide Formation	$C + 0.5O_2 \rightarrow CO$	(1)
Carbon Dioxide Formation	$CO + 0.5O_2 \rightarrow CO_2$	(2)
Carbon Equilibrium	$C + CO_2 \leftrightarrow 2CO$	(3)
Carbon Equilibrium	$C + H_2O \leftrightarrow CO + H_2$	(4)
Methane Formation	$C + 2H_2 \leftrightarrow CH_4$	(5)
Methane Equilibrium	$CH_4 + H_2O \leftrightarrow CO + 3H_2$	(6)
Water Formation	$H_2 + 0.5O_2 \rightarrow H_2O$	(7)
Water Gas Shift	$CO + H_2O \leftrightarrow CO_2 + H_2$	(8)

A number of different reactor designs include vertical updraft, vertical downdraft, horizontal, fluidized bed and entrained flow. The first three designs are not well suited to scale-up and the entrained flow design requires significant equipment to reduce the particle size of feed. Thus, fluidized bed is most common for large-scale reactors[8] and the basis of comparison used in this assessment. A number of different

setups using a fluidized bed reactor can be imagined and categorized based on oxidizer type and system operating pressure.

Possible oxidizers for use in gasification include air, pure oxygen and steam. Air, while free, introduces large quantities of inert nitrogen, requiring larger size equipment and increasing capital cost. Pure oxygen eliminates nitrogen but is more costly to obtain. Steam gasification boosts hydrogen yield but requires expensive steam and is net endothermic. Since each oxidizer has advantages and disadvantages, the best oxidizer can only be determined based on final economic results. Carbon dioxide is also suited for use, but is generally unavailable at high concentration in large quantities.

Excess air drives reactor temperature because of its cooling effect, and temperature drives outlet composition from equilibrium reactions. Desirable ratios for excess air have been calculated for different setups[9]. So, when evaluating each possible setup, the excess air ratio and an operating temperature must be established. Increasing temperatures reduce the formation of unwanted tar; studies show a 74% reduction in tar when the reactor temperature is increased from 700 ^{o}C to 850 ^{o}C [10]. However higher temperatures require more specialized, expensive metallurgy and insulation to prevent mechanical failure.

2.4 GAS CLEANUP

Gas cleanup is the removal of solids such as ash, heavy metals and tar, such that the clean syngas is suitable for further downstream operations. Technologies to perform cleanup are cyclones, filters, water wash and cracking. With no moving parts and low maintenance requirements, cyclones are of particular interest and their performance in connection with fluidized bed gasifiers can be modeled on solids removal or particle size[11]. Multiple cyclones may be used to handle large volumetric flowrates such as those found on the outlet of a low-pressure gasification reactor.

Filters can remove a high percentage of ash and tar but they are difficult operationally and represent a poor choice to capture a large quantity of solids. Typically bulk cyclones and fine filters are used together to clean syngas for even the most demanding applications.

Water wash is also effective at removing ash and tar. A continuous process, water wash does not have the same operational limitations of filters. On the other hand, the injection of liquid water cools the syngas stream and impacts thermal efficiency. Some form of water treatment is required for the spent water and the costs of this treatment can escalate quickly if heavy metals are present.

High temperature cracking can be used to break tars into lighter hydrocarbons rather than removing the tars. A catalyst and temperatures approaching 950 $^{\circ}$C are required for cracking, necessitating a secondary source of heat to supplement the gasification reactor[12] and adding complexity to the system. Overall carbon efficiency is improved when tars are cracked into light hydrocarbons which can later be used as fuels.

2.5 GAS PROCESSING

Gas processing includes steps to convert the syngas to final usable or saleable products. The first option is to sell clean syngas as a natural gas replacement. In this situation the energy content of syngas is compared to that of natural gas to determine the value of the syngas.

The second option is to burn clean syngas to produce electricity using an externally-fired gas turbine, the most promising energy conversion device for this service[13]. Revenue is generated by the sale of electricity to the utility grid.

The third option is to separate the hydrogen from the clean syngas. Hydrogen can be sold as a chemical commodity while the remaining syngas is burned to produce electricity, again using an externally-fired gas turbine. The fourth option is to perform water-gas shift reaction with the syngas to increase the amount of hydrogen produced. This requires a specific reactor vessel and is somewhat exothermic, as in equation (9). The hydrogen is then separated from the clean syngas and sold as a commodity.

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad \Delta H = \text{-41.16 kJ/kmol} \qquad (9)$$

In options three and four, the separation can be performed by pressure-swing adsorption, cryogenic distillation and membranes in order of decreasing hydrogen purity[14]. Also in either option, compression may be needed if the hydrogen pressure does not meet the delivery specifications of the biorefinery.

The fifth option is to include a reaction of hydrogen and carbon monoxide to form methanol. Designs to carry out this reaction have been in industrial use for decades with many applications worldwide,[15] although performance improvements and cost

savings have been achieved in more recently[16]. This reaction is highly exothermic and typical reactors are equipped with cooling systems to prevent unwanted temperature rise. After production, a low temperature distillation column separates the produced methanol from gaseous products and water.

$$CO + 2H_2 \rightarrow CH_3OH \qquad \Delta H = \text{-90.7 kJ/kmol} \qquad (10)$$

Syngas can also be used for ammonia synthesis, ethane synthesis, higher-order alcohol synthesis, methanation and Fisher-Tropsch conversion to liquid fuels[17]. For the purpose of meeting the requirements of an integrated seed oil biorefinery these options do not offer a particular advantage, as their products are not feedstocks to another process unit. Thus, in this study, attention is focused on the first five options which offer points of integration. A company wishing to exhaust all possibilities for gasification plant design could investigate and mathematically model other options as well.

Not all syngas must be directed to a single product; a combination of products might be the most economic solution. This is especially true for large facilities where the maximum equipment size is reached and either copies or alternatives must be built, or where demand for a particular product has been fully satisfied. Similarly, multiple reactors may connect to a single processing option to capture greater economies of scale.

Final processing may include equipment to reduce the impact of streams discharged to the environment. An example is that a waste stream containing carbon monoxide needs to be flared, not vented. With growing regulations and costs associated to carbon dioxide emissions, capture and storage of this byproduct may also be

economic. Some processing options provide a stream of nearly pure carbon dioxide, eliminating the challenge and cost of capturing carbon dioxide from a flue gas stream.

2.6 GASIFICATION UNIT DESIGN

A number of recent research publications have investigated biomass gasification plants in the context of feedstock, technology and product selection. Bridgwater[13] demonstrates the approach to gasification unit design without the benefit of computational power. Technology choices are provided with qualitative assessments along with curves for installation cost and efficiency. Factors such as operating pressure are discussed qualitatively along with identification of particular problem areas such as pretreatment and gas cleaning. The expectation is that a design engineer can identify a reasonably good combination of equipment based on the provided information.

Spath et al[18] analyzed hydrogen production from biomass gasification with low-pressure air and high-pressure oxygen combinations, along with pyrolysis. In all cases water-gas shift followed by a PSA unit was the downstream design. Unique to this research effort, the authors use a probability distribution rather than fixed quantities for a number of variables, yielding economics results weighted for the risk and uncertainty that comes with building a novel, large-scale gasification unit. Also investigated are transportation costs if hydrogen is not consumed on-site. A hydrogen selling price of $1.24 to $2.40 per kilogram is predicted based on a feedstock cost between 1.6 ¢/kg and 4.6 ¢/kg. Pyrolysis was the best technology of the three, with low-pressure air gasification slightly behind, and high-pressure oxygen the least economic.

Hamelinck et al[19,20] investigated process configurations and production costs for a standalone biomass gasifier aiming to minimize the cost of transportation fuel. In one study, two oxidizers, two reactor types, and two processing options (hydrogen and methanol) are considered. Another study considers three oxidizers, three operating pressures and Fischer-Tropsch synthesis as the sole processing option. In both studies the authors use a generic biomass source with an unlimited quantity of material available. Each possible setup is simulated and evaluated separately to find its final economics, without consideration for combinations of options. At the end of the study one setup is identified as the winner for offering the best economics.

Using this methodology, the authors conclude that biomass-derived methanol and hydrogen are likely to become competitive, and that a low-pressure air-blown gasifier with steam shift and pressure swing adsorption is the most economic setup. The authors also conclude that the Fischer-Tropsch option is uneconomic and requires either tax subsidies or a technological breakthrough. A high-pressure oxygen-blown gasifier is the best choice to connect with a Fischer-Tropsch reactor. This myriad of combinations reinforces that selection of gasification technologies and products is not intuitive. It also reinforces the need to have an automated routine to evaluate between alternatives, especially when inclusion of multiple feed stocks exponentially increases the number of setups to evaluate.

Brown et al[21] investigated gasification design using a multi-objective optimization program, with the objective functions being exergy efficiency and capital cost. This model included different oxidiziers, operating pressures and operating

temperatures. As drawbacks, it included only one feedstock, one product (electricity) and had a fixed 20 MW capacity. In carrying out 15000 point evaluations the authors arrive at a Pareto curve between the two objective functions. The study authors conclude that steam gasification has the lowest overall capital cost, although with lower operating costs, air gasification yielded the lowest cost per unit of electricity produced. The study also indicated an economic benefit for the internal combustion engine for power generation method over gas turbines in the same service.

Cameron et al.[22] investigate biomass combustion and gasification for the production of electricity in Canada. The authors identify a trade-off as combustion has lower capital costs but higher feedstock costs than a gasification plant of the same capacity. Only considering electrical production, however, the authors conclude than neither combustion nor gasification is competitive without subsidies.

2.7 OPTIMIZATION OF BIOREFINERIES

Biorefining has greatly matured from the conceptual to the commercial over the past two decades. There is now suitable conversion and cost data to apply mathematical modeling to the biorefinery in search of improved process configurations and detailed analysis of economic performance, and a number of researchers are active in this area. Unlike the aforementioned gasification models, these biorefinery models tend to have less detail on any single process unit, but also the ability to include many refining pathways and products in their respective superstructures.

Methods for arriving at an optimal solution for a network of feedstocks and products are of particular interest. Logically, if an optimized gasification unit can be fit in a superstructure of feed and product lines, then the mathematical optimum can be achieved for the entire biorefining system. On the feed side, Elms and El-Halwagi[23] developed a procedure for identifying the best processing schedule using a procedure of simulation, mass and energy integration, cost estimation and mathematical optimization. In doing so, they demonstrate than an optimum solution can indeed be obtained for feedstock problems. Sammons et al[24] use a similar procedure to solve optimization problems with multiple products. With a model for the gasification block, it becomes possible the ultimate biorefining optimization problem from farm to distribution point. An alternate method for biorefinery optimization using carbon cascade is proposed by Ng[3] which can be easily scaled up to a large number of pathways. In this method, gasification is treated as a black box with a user-specified conversion. The best-performing gasification unit can implemented in place of the black box and the overall economics of biorefinery improved.

A different approach is taken by Stuart and Janssen in their investigation of forestry-related biorefineries[25]. Design choices are made not on solely economics, but using multi-criteria decision making. These criteria include return on investment but also technology risk, feedstock flexibility, potential for new products and environmental impact. The weighting of each criterion is determined by a panel of experts, thus, the evaluation results include business acumen and social responsibility.

3. PROBLEM STATEMENT

Research on the technical performance of gasification equipment and the overall configuration of a biorefinery does not by itself identify the best configuration for economic performance. This effort seeks to bring together performance data and apply the concepts of process integration to create a procedure applicable to any biorefinery gasification situation. The overall goal of this work is a decision-making tool for optimizing gasification plant design from an economic perspective. Specifically, the problem addressed in this work is stated as follows:

Given are:

- A set of biomass feedstocks $\{i|i = 1,2,\ldots,I \}$ which includes fresh as well as residue biomass

- A set of gasification technologies $\{j|j = 1,2,\ldots,J\}$ along with their heat-recovery and gas-cleaning ancillary units

- A set of syngas-processing technologies $= \{k|k=1,2,\ldots,K\}$ that can produce a variety of chemical and fuels

It is desired to develop a systematic approach to the analysis, optimization, and screening of the process alternatives that can guide the decision makers in selecting gasification pathways under different conditions.

The solution approach begins with the analysis of biorefinery residues and other biomass available as feedstocks to the gasification unit. Composition, cost and an upper bound on availability are established for each feedstock. Next, a set of candidate

gasification technologies is built, each with relevant information for conversion, cost, and maximum size. Performance and cost data are also found for downstream gas cleaning and conversion processes. The optimal configuration(s) are selected on the basis of maximizing economic worth while remaining within equipment performance constraints. Annual worth is the economic measure selected as the optimization variable.

4. THEORETICAL APPROACH

4.1 TARGETING

Targeting is a technique to find the maximum possible performance of a process unit, both in terms of material balance and economics. It is useful to find whether the unit has a chance at profitability and for comparison purposes with the final outcome[26]. Complete conversion of carbon to carbon monoxide and of hydrogen to hydrogen gas are assumed, along with capture of all energy produced. Furthermore, 100% efficiency in reaction, conversion and separation are assumed. Targeting is applied to the same reactor and product choices as the optimization routine.

A key measure of the utility of a gasificaton unit is whether it can wholly meet the biorefinery's demand for hydrogen or methanol. If it cannot, a separate steam-methane reforming unit or methanol synthesis unit is required, adding cost and complexity to the overall project. Therefore targets are set for hydrogen and/or methanol demands from the biorefinery. Demands for hydrogen and methanol are based on the structure of triglyceride obtained from the oil seed. An example triglyceride is shown in Figure 4.
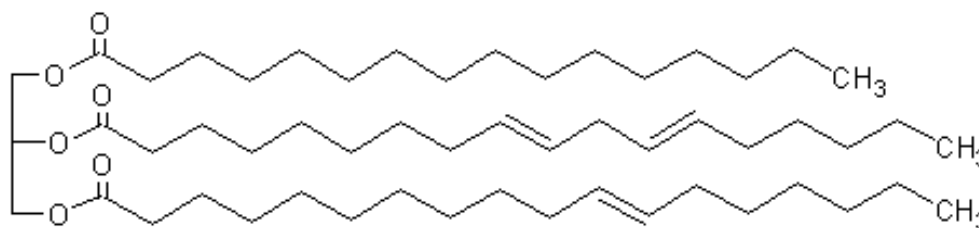
Figure 4. Typical triglyceride in soybean oil.

First, consider hydrogen demand created by the conversion of triglycerides to diesel length paraffins. There are two basic mechanisms for this reaction, hydrodeoxygenation and decarboxylation. Research in the catalysis field indicates a combination of both mechanisms occurs[27]. The variable π used in the reaction equations represents the number of unsaturated pi-bonds in the triglyeride.

Hydrodeoxygenation Reaction

$$1 \, triglyceride + (12 + \pi)H_2 \rightarrow C_3H_8 + 6H_2O + 3 \, paraffins \tag{11}$$
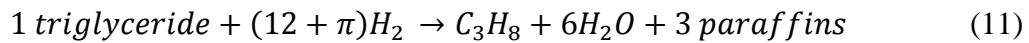
Decarboxylation Reaction

$$1 \, triglyceride + (3 + \pi)H_2 \rightarrow C_3H_8 + 3CO_2 + 3 \, paraffins \tag{12}$$
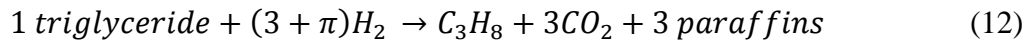
Methanol demand is established through the transesterification reaction. The transesterification reaction is accomplished by catalytically reacting the triglyceride with an excess of alcohol, in this case methanol, to produce glycerol and methyl esters which form biodiesel[28]. If desired, minimum flow rates for hydrogen and methanol can be established as constraints in the optimization routine to meet the objective of supplying the biorefinery.

Transesterfication

$$1 \, triglyceride + 3 \, CH_4OH \rightarrow 1C_3H_8O_3 + 6H_2O + 3 \, methyl \, esters \tag{13}$$
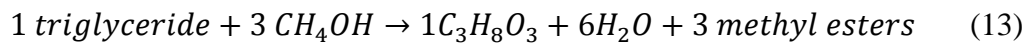
4.2 SIMULATION

ASPEN Plus is software for simulating unit operations including the gasification process. Simulations demonstrate feasibility and allow the material and energy balances to be obtained. Generally the simulation is adept at modeling the reactant and syngas flow along with the water-gas shift reaction and methanol distillation, if present. The simulation is limited in calculating the combustion of biomass; literature values for conversion are therefore input for the necessary reactions using the preferred dual-reactor design[7]. More rigorous ASPEN Plus models have been developed[29] but offer little improvement in calculation of outlet compositions in comparison to the conversion reactor. Similarly for separation equipment, the simulation is not able to estimate performance of membranes and adsorbents, thus literature values are used.

There are cases where the material being simulated is actually a blend of components and no data is available about the nature or ratio of these components. For example tar consists of a wide range of heavy hydrocarbons formed by chance in the reactor. Simplifying assumptions with regards to components are made to permit simulation; these assumptions are provided in Table 2.

Table 2. **Components Used in Simulation**

Actual Component	Representation
Triglycerides	Triolein
Ash	Calcium Oxide
Soot	Carbon
Tar	Phenanthrene ($C_{14}H_{10}$)
2-, 3- and 4-Carbon Hydrocarbons	Ethane

4.3 COST AND REVENUE ANALYSIS

Capital cost estimates are obtained from literature values or the ASPEN ICARUS

Process Evaluator. For the basis of comparison, all costs are updated to 2009 US dollars

using the Chemical Engineering Plant Cost Index and current exchange rates.

Operating revenues and costs are determined from current market prices for

commodities and labor. The model allows market prices to be easily adjusted since they

are expected to fluctuate in the future. A cost is also included for greenhouse gas

emissions representing the cost of carbon dioxide under a cap-and-trade system or a

carbon tax system.

4.4 OPTIMIZATION

The problem follows the structural parameter approach that is well-documented

in process synthesis[30] and recently applied to the biorefinery facilities such biodiesel

production[23]. A superstructure is created to embed the potential configurations of

interest for connecting feedstocks, processing units and products. Figure 5 provides the

superstructure employed for the optimization problem. This superstructure can be

extended to new feed stocks, reactor designs and products, as needed.
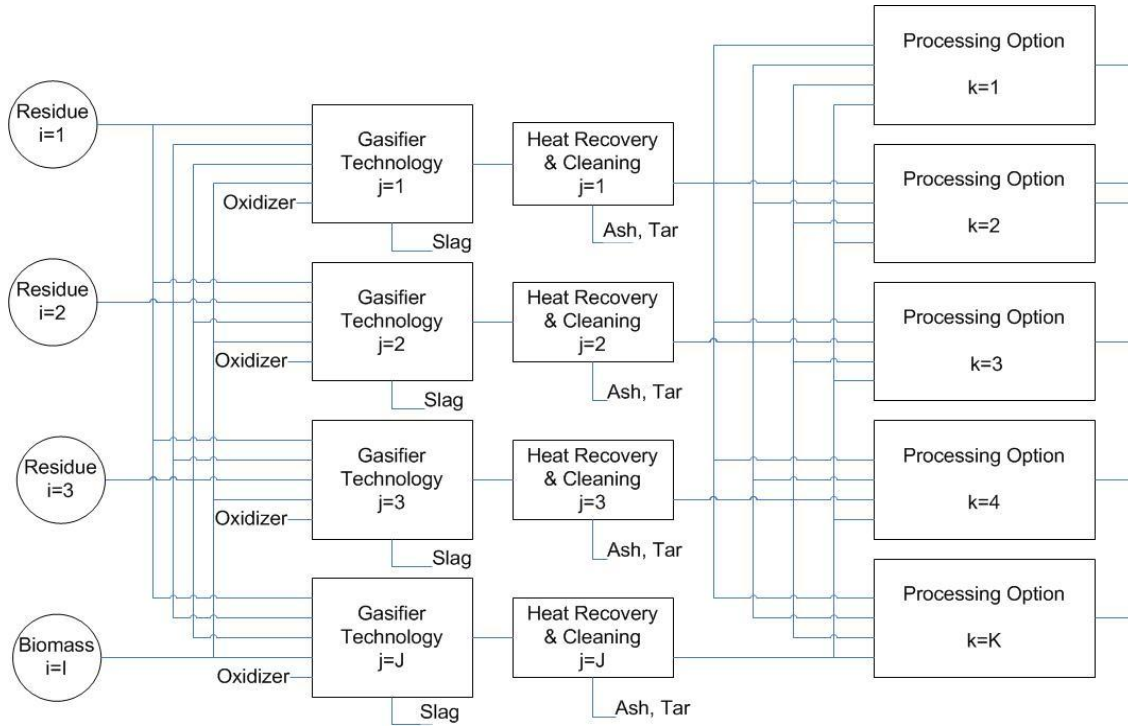
Figure 5. **Superstructure of the optimization problem.**

Based on the superstructure and the available equipment and economic components, a system model is built. The model is classified as mixed-integer non-linear or MINLP and is solved using LINGO software. LINGO is mathematical software that specializes in solving numerical optimization problems through a proprietary solver. The criterion of optimization is annual worth.

Beginning at the source, the biomass flow from each feedstock source i is split into a number of flows to the gasification reactors. The term $F_{i,j}^{ST}$ represents the flowrate assigned from source i to reactor j.

$$F_i^S = \sum_{j=1}^{J} F_{i,j}^{ST} \quad \forall i \tag{14}$$

For each source, a maximum flowrate $F_i^{S\,max}$ is set based on agricultural limits and the extraction unit design.

$$F_i^S \leq F_i^{S\,max} \quad \forall\, i \tag{15}$$

The source operating cost is the cost of feedstock from a particular source i. It is calculated by multiplying the source flowrate F_i^S with the delivered cost per unit mass.

$$P_i^S = F_i^S * P_i^{S\,ref} \quad \forall\, i \tag{16}$$

Source revenue is obtained if a gasification unit is paid or subsidized to accept a certain feedstock. It is calculated by multiplying the source flowrate F_i^S with the subsidy per unit mass.

$$R_i^S = F_i^S * R_i^{S\,ref} \quad \forall\, i \tag{17}$$

The source capital cost is a ratio of source i flowrate to the reference flowrate, adjusted by the scale up factor h. This ratio is multiplied by the reference capital cost $C_i^{S\,ref}$.

$$C_i^S = \left(\frac{F_i^S}{F_i^{S\,ref}}\right)^h * C_i^{S\,ref} \quad \forall\, i \tag{18}$$

To complete the first mass exchange network, the component flow is found by multiplying the total flow F_i^S by the known source composition $X_{i,r}^S$. This calculation is applied for each component r present in the streams connecting source i to reactor j.

$$F_{i,j,r}^{ST} = F_i^S * X_{i,r}^S \quad \forall\, i,j,r \tag{19}$$

The next set of constraints relate to the reactor section of the gasification unit. At the inlet of reactor j, mass flow is found through summation of incoming flows $F_{i,j}^{ST}$ from all sources i. If reactor j does not exist, then $T_j^E = 0$, and reactor inlet flow is set at zero.

$$F_j^T = \sum_{i=1}^{I} F_{i,j}^{ST} * T_j^E \quad \forall\, j \tag{20}$$

Owing to the limit on the feasible size for constructing a gasification reactor, a maximum inlet flowrate is set for each reactor j.

$$F_j^T \leq F_j^{T\,max} \quad \forall\, j \tag{21}$$

The reactor inlet component mole flow for each reactor j is the summation of incoming component flows $F_{i,j,r}^{ST}$ for each component r, divided by the molecular weight for that component to convert from mass basis to molar basis.

$$G_{j,r}^T = \sum_{j=1}^{J} \frac{F_{i,j,r}^{ST}}{MW_r} \quad \forall\, i,r \tag{22}$$

Reactor performance follows a conversion model for the majority of components. Outlet mole flows are found using the conversion values $Y_{j,q}^T$ obtained in literature, as

given in equation (23). Water and slag are the free components in the hydrogen and carbon mole balances, respectively. That is, the amount of water produced is equal to the total amount of hydrogen that entered the reactor minus that amount converted into hydrogen, methane, tar and ethane as given in equation (24). Similarly, the amount of slag produced is equal to the total amount of carbon that entered the reactor minus the amount converted into carbon monoxide, carbon dioxide, methane or ethane as provided in equation (25).

$$G_{j,q}^{U} = G_{j,r}^{T} * Y_{j,q}^{T} \quad \forall j, q \neq h_2o, slag \tag{23}$$

$$G_{j,q}^{U} = 0.5G_{j,H2}^{T} + G_{j,H2O}^{T} - G_{j,H2}^{U} - 2G_{j,CH4}^{U}$$

$$-5G_{j,Tar}^{U} - 3G_{j,C2H6}^{U} + \frac{F_{j,H2O}^{A}}{MW_{H2O}} \quad \forall j, \ q = h_2o \tag{24}$$

$$G_{j,q}^{U} = G_{j,C}^{T} - G_{j,CO}^{U} - G_{j,CO2}^{U} - G_{j,CH4}^{U} - 2\,G_{j,C2H6}^{U} \quad \forall j, q = slag \tag{25}$$

Reactor oxygen demand is the quantity of oxidizer delivered to the reactor in order to gasify all of the feed. It is calculated from the mole balance of oxygen entering and exiting the reactor system.

$$T_{j}^{D} = G_{j,CO}^{U} + 2G_{j,CO2}^{U} + G_{j,H2O}^{U} - G_{j,O}^{T} - G_{j,H2O}^{T} \quad \forall j \tag{26}$$

Next, the mass flowrate of the oxidizer is determined. If reactor j receives air as the oxidizer, then $T_{j,O2}^{T} = T_{j,steam}^{T} = 0$ and equation (27) finds the mass flowrate of the oxidizer by multiplying the molecular oxygen demand by the mole weight of pure

oxygen and dividing by the fraction oxygen in air. If reactor j receives pure oxygen as the oxidizer, then $T^T_{j,O2} = 1$ and equation (28) is used to find the oxidizer mass flowrate by multiplying oxygen demand by the molecular weight of oxygen. Steam is used in excess quantities and leaves the reactor as a product. If reactor j receives steam as the oxidizer, then $T^T_{j,steam} = 1$ and equation (29) is used to find the oxidizer mass flowrate based on the ratio of steam to feed.

$$F^A_{j,q} = T^D_j * \left(1 - T^T_{j,O2}\right) * \left(1 - T^T_{j,H2O}\right) * \frac{MW_{O2}}{2MP} \quad \forall \, j, \ q = air \tag{27}$$

$$F^A_{j,q} = T^D_j * T^T_{j,O2} * \frac{MW_{O2}}{2} \quad \forall \, j, \ q = O2 \tag{28}$$

$$F^A_{j,q} = \frac{T^F_j * T^T_{j,steam}}{4.340 \ kg \ steam \ / \ kg \ biomass} \quad \forall \, j, \ q = H2O \tag{29}$$

A significant amount of the gasification unit revenue comes from energy production and the quantity of energy produced is dependent on the feedstock composition. An equation developed by Sheng and Alvarado is used to find biomass energy content in units of MJ/kg [31].

$$HHV = 0.32559 * \frac{G^{in}_{j,C} * MW_C * 100}{F^{T\,in}_j} + 3.4597 \quad \forall \, j \tag{30}$$

Reactor energy production is the higher heating value multiplied by the mass flowrate $F^{T\,in}_j$ and the gasifier thermal efficiency η_c. By dividing energy production by the specific and latent heat required to vaporize boiler feed water into steam, the mass flowrate of produced steam is calculated.

$$F_j^P = \frac{HHV * F_j^{T\,in} * \eta_c}{2.690 + 0.313\ MJ/kg} \quad \forall\, j \tag{31}$$

Slag produced in the reactor is caught in the slag trap and removed from the system; no slag carries forward with the syngas leaving the reactor. Therefore the removal of slag is established as the molar flowrate of slag being produced, $G_{j,slag}^U$, multiplied by the molecular weight of slag.

$$F_{j,slag}^Z = G_{j,slag}^U * MW_{slag} \quad \forall\, j,\ q = slag \tag{32}$$

The reactor capital cost is a ratio of the actual flowrate of reactor j to the reference flowrate, adjusted by the scale up factor h. An additional value of \$0.1MM is added to cover the cost of initial technology acquisition. This ratio is then multiplied by the reference capital cost $C_j^{T\,ref}$.

$$C_j^T = 0.1 T_j^E + C_j^{T\,ref} * \left(\frac{F_j^T}{F_j^{T\,ref}} \right)^h \quad \forall\, j \tag{33}$$

Reactor operating cost is based on flowrate F_j^T multiplied by reference operating cost $P_j^{T\,ref}$. Additional operating costs for the purchase of oxygen and steam oxidizers are added to achieve the final operating cost value.

$$P_j^T = F_j^T * P_j^{T\,ref} * + F_{j,O2}^A * M_{O2} + F_{j,H2O}^A * M_{Steam} \quad \forall\, j \tag{34}$$

Reactor operating revenue is obtained from the multiplication of the quantity of produced steam, F_j^P, and the steam price.

$$R_j^T = F_j^P * M_{Steam} \quad \forall j \tag{35}$$

Three possible cleaning methods can be associated with reactor j, namely cyclones, filters and water wash. This constraint prevents multiple cleaning methods from existing concurrently for the same reactor.

$$\sum_{z=1}^{Z} U_{j,z}^E = 1 \quad \forall j \tag{36}$$

The quantity of ash and tar removed by the cleaning system is the product of two terms. The first term is removal efficiency $Y_{j,z}^U$ is multiplied by $U_{j,z}^E$ so that it becomes zero when the cleaning method does not exist. The second term is mass flowrate, found by multiplying the reactor outlet molar flowrate $G_{j,q}^U$ by the molar weight of component q.

$$F_{j,q}^Z = (U_{j,z}^E * Y_{j,z}^U) * (G_{j,q}^U * MW_q) \quad \forall j,z \quad q = ash, tar \tag{37}$$

It is assumed that the cleaning step removes only solids and does not remove other species. Thus, the outlet flow for each species is equal to the molar flow produced in the reactor $G_{j,q}^U$ multiplied by the molar weight, as shown in equation (38). For ash and tar, the outlet flows are found the same way, but with subtraction of the quantity of ash and tar removed by the cleaning system, as given in equation (39).

$$F_{j,q}^U = G_{j,q}^U * MW_q \quad \forall j, \quad q = CO, CO_2, CH_4, H_2, N_2, C_2H_6, H_2O \tag{38}$$

$$F_{j,q}^U = G_{j,q}^U * MW_q - F_{j,q}^Z \quad \forall j, \quad q = ash, tar \tag{39}$$

The total outlet flowrate from reactor and cleaning section j is the summation of the mass flowrates of individual component flowrates $F_{j,q}^U$ from the same section.

$$F_j^U = \sum_{q=1}^{Q} F_{j,q}^U \tag{40}$$

To acquire the mass fraction of each component leaving the reactor, the mass flowrate of an individual component q from the reactor and cleaning section is divided by the total mass flowrate leaving the reactor and cleaning section. Unstable results can potentially occur at very small flowrates and mass fractions approaching unity. Therefore, and in agreement with experimental results for gasification systems, mass fractions are limited to 0.9 and below for all components.

$$X_{j,q} * F_j^U = F_{j,q}^U \quad \forall j, \ q \tag{41}$$

$$X_{j,q} < 0.9 \quad \forall j, q \tag{42}$$

The cleaning method capital cost is found using a ratio of the actual flowrate of mass through cleaning section j to the reference flowrate, adjusted by the scale up factor h. An additional value of \$0.1MM is added to cover the cost of initial technology acquisition. This ratio is multiplied by the reference capital cost $C_j^{U\ ref}$.

$$C_j^U = U_{j,z}^E * C_{j,z}^{U\ ref} * \left(\frac{F_j^U}{F_{j,z}^{U\ ref}} \right)^h + 0.1 U_j^E \quad \forall j, z \tag{43}$$

The cleaning operating cost is the reference operating cost $P_{j,z}^{U\,ref}$ and the mass

flowrate through the cleaning device F_j^U. The operating cost is set to zero if the cleaning

option does not exist using the term $U_{j,z}^E$.

$$P_j^U = U_{j,z}^E * P_{j,z}^{U\,ref} * F_j^U \quad \forall\, j, z \tag{44}$$

The cleaning operating revenue is calculated by multiplying the reference

operating revenue $R_{j,z}^{U\,ref}$ and the mass flowrate through the cleaning device F_j^U. The

operating revenue is set to zero if the cleaning option does not exist using the familiar

term $U_{j,z}^E$.

$$R_j^U = U_{j,z}^E * R_{j,z}^{U\,ref} * F_j^U \quad \forall\, j, z \tag{45}$$

The second mass flow network is formulated in the same way as the first network

was formulated. The outlet flowrate from reactor j is equal to the sum of the flows from

reactor j to final processing option k.

$$F_j^U = \sum_{k=1}^K F_{j,k}^{UV} \quad \forall\, j$$

Component flow is the product of total flow F_j^U and the known source

composition $X_{j,k}^U$. This calculation is applied for each component q in the streams

connecting reactor j to processing option k.

$$F_{j,k,q}^{UV} = F_j^U * X_{j,q} \quad \forall\, j, k, q \tag{46}$$

The next set of constraints deal with the final processing option within the gasification unit. At the inlet of processing option k, the mass flow is the summation of incoming flows $F_{j,k}^{UV}$ from all reactors j. If processing option k does not exist, then $V_k^E = 0$ and the processing option inlet flow is set to zero.

$$F_k^V = \sum_{j=1}^{J} F_{j,k}^{UV} \quad \forall \, k \tag{47}$$

A maximum capacity $F_k^{V \, max}$ is established for each processing option based on the feasibility of construction. In most situations the processing option capacity is much higher than the feedstock availability and reactor size constraints and does not affect the final solution. It is included for the sake of completeness and to allow the model to be generally applicable.

$$F_k^V < F_k^{V \, max} * V_k^E \quad \forall \, k \tag{48}$$

The processing option inlet component mole flow for each processing option k is calculated through the summation of incoming component flows $F_{j,k,q}^{UV}$ for each component q, divided by the molecular weight for that component to convert from mass basis to molar basis.

$$F_{k,q}^V = \sum_{k=1}^{K} F_{j,k,q}^{UV} \quad \forall \, q \tag{49}$$

Fuel gas is the primary product from the first processing option. In this case, all combustible species entering the option are accounted for in the fuel gas flow.

$$F_{k,q}^F = F_{k,q}^V \quad k = 1, q = CO, H_2, CH_4, C_2H_6 \tag{50}$$

Fuel gas is a secondary product from the methanol synthesis option. Based on the performance of the distillation column in simulation, it is assumed that all light combustible gases are recovered in the fuel gas stream and none are mixed with the liquid methanol and water streams. For methane and ethane, which do not take part in the methanol synthesis reactions, the flowrate entering the option is accounted for in the fuel gas flow as shown in equation (51). For carbon monoxide and hydrogen, the flow to fuel gas is equal to the inlet flow minus the quantity converted to methanol as provided in equations (52) and (53) respectively.

$$F_{k,q}^F = F_{k,q}^V \quad k = 5, q = CH_4, C_2H_6 \tag{51}$$

$$F_{k,q}^F = F_{k,q}^V * (1 - Y_{Me}^R) \quad k = 5, q = CO \tag{52}$$

$$F_{k,q}^F = F_{k,q}^V * (1 - Y_{Me}^R) * (1 - Y_{Me}^D)^{-1} * \frac{MW_{H2}}{MW_{Me}} \quad k = 5, q = H2 \tag{53}$$

Electricity is the primary product of the second processing option. In this case, all combustible species entering the option are accounted for in the flow of syngas to the power plant.

$$F_{k,q}^E = F_{k,q}^V \quad k = 2, q = CO, H_2, CH_4, C_2H_6 \tag{54}$$

Electricity is the co-product of the third processing option. Pressure swing adsorption leaves the carbon-based gases present in the flow of syngas to the power plant as given in equation (55). The quantity of hydrogen in the flow of syngas to the power plant, found in equation (56) is the inlet quantity of hydrogen minus the quantity removed by the PSA for separate sales.

$$F_{k,q}^E = F_{k,q}^V \quad k = 3, q = CO, CH_4, C_2H_6 \tag{55}$$

$$F_{k,q}^E = F_{k,q}^V * (1 - Y_{H2}^{PSA}) \quad k = 3, q = H_2 \tag{56}$$

Staying with the third processing option, the quantity of hydrogen removed and available for sales is found by multiplying the inlet hydrogen flowrate by the PSA efficiency Y_{H2}^{PSA}.

$$F_{k,q}^W = F_{k,q}^V * Y_{H2}^{PSA} \quad k = 3, q = H_2 \tag{57}$$

Hydrogen is the only product of the fourth processing option. The flow rate of produced hydrogen is the sum of the inlet hydrogen flow rate $F_{k,q}^V$ and the amount generated in the steam shift reactor, as determined by the inlet rate of carbon monoxide $F_{k,CO}^V$, multiplied by the conversion Y_{H2}^R and finally adjusted for the PSA efficiency Y_{H2}^{PSA}.

$$F_{k,q}^W < (F_{k,q}^V + F_{k,CO}^V * Y_{H2}^R) * Y_{H2}^{PSA} \quad k = 4, q = H_2 \tag{58}$$

Finally, methanol is the primary product of the fifth processing option. The flowrate of methanol leaving the distillation column is calculated from the initial flowrate of carbon monoxide $F^V_{k,CO}$ multiplied by the conversion Y^R_{Me} and the distillation efficiency Y^D_{Me} and adjusted for molecular weight.

$$F^W_{k,q} < F^V_{k,CO} * Y^R_{Me} * Y^D_{Me} * \frac{MW_{Me}}{MW_{CO}} \quad k = 5, q = Me \tag{59}$$

For the fourth and fifth processing options, the secondary reactor requires an elevated inlet temperature to overcome the activation energy for the desired reaction. The water wash cleaning method cools the syngas below this temperature. A constraint is established to prevent combinations including water wash and these options.

$$U^E_{j,water\ wash} + V^E_k < 1 \quad k = 4,5 \tag{60}$$

For these same processing options, the catalyst is sensitive to tar and ash carried in the syngas. To prevent fouling the catalyst with these solids, a maximum tar and ash constraint is established

$$F^W_{k,ash} + F^W_{k,tar} < F^{max}_{k,ash+tar} \quad k = 4,5 \tag{61}$$

Carbon dioxide emissions are found for each processing options to permit analysis of the impact of carbon taxes and credits. For the first and fifth options (fuel gas and methanol/fuel gas), the entire product mass flow is pipelined and thus there are no carbon dioxide emissions.

$$F_{k,q}^W = 0 \quad k = 1, q = CO_2 \tag{62}$$

For the second, third and fourth options, there is no product containing carbon. All carbon ends up in the atmosphere as carbon dioxide whether it is directly released, flared or burned in a combustion process. Therefore, the carbon dioxide emissions are the sum of incoming carbon dioxide flow with the molar weight-adjusted methane, ethane and carbon monoxide flows.

$$F_{k,q}^W = F_{j,q}^V + F_{j,CH4}^V * \frac{MW_{co2}}{MW_{CH4}} + 2F_{j,C2H6}^V * \frac{MW_{co2}}{MW_{C2H6}} + F_{j,CO}^V * \frac{MW_{co2}}{MW_{CO}}$$

$$k = 2,3,4 \quad q = CO_2 \tag{63}$$

The processing option capital cost is calculated using a ratio of the actual flowrate of option k to the reference flowrate, adjusted by the scale up factor h. An additional value of \$0.1MM is added to cover the cost of initial technology acquisition. This ratio is multiplied by the reference capital cost $C_k^{U\ ref}$.

$$C_k^V = 0.1V_k^E + C_k^{V\ ref} * \left(\frac{F_k^V}{F_k^{V\ ref}}\right) \quad \forall\ k \tag{64}$$

Processing option operating cost is evaluated based on flowrate F_k^V multiplied by reference operating cost $P_k^{V\ ref}$. The cost of carbon dioxide emissions is added based on the flowrate $F_{k,CO2}^W$ found earlier.

$$P_k^V = P_k^{V\ ref} * F_k^V + F_{k,CO2}^W * M_{CO2} \quad \forall\ k \tag{65}$$

Processing option operating revenue is obtained by adding together the revenue

from a number of different products. For electrical production, the energy term is the

summation of the flow rate of each fuel gas product $F_{k,q}^E$ with its heating value LHV_q.

This energy term is multiplied by the electrical price $M_{elect} * Diff$, the conversion

factor and the electrical turbine efficiency η_t to determine revenue. For fuel gas, the

energy term is again obtained by summation of the flow rate of each fuel gas product

$F_{k,q}^F$ with its heating value LHV_q. The energy term is multiplied by the selling price of

natural gas $M_{NG} * Diff$ and conversion factor to determine the revenue. Revenues from

hydrogen and methanol sales are found by multiplying mass flowrate with product price.

$$R_k^V = \left(\sum_q^Q F_{k,q}^E * LHV_q\right) * M_{elect} * \frac{Diff}{3.6} * \eta_t + \left(\sum_q^Q F_{k,q}^F * LHV_q\right) * M_{NG} * \frac{Diff}{1000}$$

$$+ F_{k,H2}^W * M_{H2} + F_{j,Me}^W * M_{Me} \quad \forall\, k \tag{66}$$

The fixed capital cost of the entire gasification unit is the sum of the capital costs

from the source, reactor, gas cleaning and processing option sections.

$$C = \sum_{i=1}^I C_i^S + \sum_{j=1}^J C_j^T + \sum_{j=1}^J C_j^U + \sum_{k=1}^K C_k^V \tag{67}$$

The annual value of capital cost is obtained using the standard formula for

annualizing a one-time cost[32] with the project interest rate IR and the project life N.

$$A_C = \left(C * \frac{IR(1+IR)^N}{(1+IR)^N - 1}\right) \tag{68}$$

The annual operations and maintenance cost is established to be 6.6% of the fixed capital cost following the guidelines of Ulrich and Vasudevan[33].

$$P^{o\&m} = 0.066 * C \tag{69}$$

Annual revenue is the summation of the source R_i^S, reactor R_j^T, cleaning R_j^U and processing option R_k^V revenues, annualized with the onstream factor os.

$$A_R = os\left(\sum_{i=1}^{I}\left(R_i^S\right) + \sum_{j=1}^{J}\left(R_j^T\right) + \sum_{j=1}^{J}\left(R_j^U\right) + \sum_{k=1}^{K}\left(R_k^V\right)\right) \tag{70}$$

Annual expenditure is set by adding the operating and maintenance cost to the summation of the source P_i^S, reactor P_j^T, cleaning P_j^U and processing option P_k^V operating costs, again annualized with the onstream factor os.

$$A_P = P^{o\&m} + os\left(\sum_{i=1}^{I}\left(P_i^S\right) + \sum_{j=1}^{J}\left(P_j^T\right) + \sum_{j=1}^{J}\left(P_j^U\right) + \sum_{k=1}^{K}\left(P_k^V\right)\right) \tag{71}$$

Annual depreciation is calculated using the straight line method over the project life.

$$A_D = C/N \tag{72}$$

Annual income tax is paid as a tax rate *TR* against the difference between annual revenue and annual operating cost and depreciation.

$$A_T = (A_R - A_P - A_D) * TR \tag{73}$$

Annual after-tax cash flow is the subtraction of the annual income tax and annual operating costs from the annual revenue.

$$A_F = A_R - A_P - A_T \tag{74}$$

Working capital is 15% of the total capital cost[33], which translates into about 17.6% of the fixed capital costs.

$$WC = C * \frac{0.15}{0.85} \tag{75}$$

Working capital is recovered at the end of the project with adjustments for inflation. The net present value of working capital can be found using the following equation[32]. Note that if the interest rate IR matches the inflation rate NR then the net present value of working capital is zero and the term has no impact on the annual worth.

$$NPV_{WC} = C * \frac{0.15}{0.85} + (-1 + (1 + IR)^N * (1 + NR)^{-N}) \tag{76}$$

If needed, the annualized working capital cost is found using the same formula as was applied for the annualized fixed capital cost.

$$A_{WC} = \left(WC * \frac{IR(1+IR)^N}{(1+IR)^N - 1}\right) \tag{77}$$

Finally, the gasification unit annual worth is the annual revenue minus annual operating costs, annualized capital costs, and annualized working capital if any.

$$A_W = A_R - A_P - A_C - A_{WC} \tag{78}$$

A number of metrics beyond the financial could also be used for process optimization. Hot-gas and cold-gas efficiency and carbon conversion efficiency metrics have been used to quantify the performance of a gasification design[34]. However there is always a trade-off between achieving high values on these metrics and the capital cost of the facility. The goal of this work is top economic performance, not necessarily top performance on an intermediate metric.

4.5 ECONOMIC EVALUATION

Three frequently used project metrics, payback period, return on investment and net present value are defined with equations (76), (77) and (78)[32]. The first two of these metrics are not suitable for the optimization routine because neither is guaranteed to produce a result that is the maximum economic benefit. Instead they are calculated to benefit the business community and for comparison to other unrelated projects which frequently use such metrics. Net present value is suited for optimization but was not selected in favor of annual worth.

$$ROI = \frac{A_R - A_P}{C} \tag{77}$$

$$RPP = \frac{C}{A_R - A_E} \tag{78}$$

$$NPV = -C + \sum_{n=1}^{N} A_F (1 + IR)^{-N} \tag{79}$$

Tax implications are considered when finding annual worth and these three project metrics. Once full engineering is complete for the unit, it is likely that some

equipment can be identified as being part of an accelerated depreciation class, providing a slightly higher annual worth than predicted here.

On the feed side, gasification is compared against options such selling the residue for animal feed, combustion of the residue for energy, and disposal of the residue as solid waste. On the produced hydrogen side, gasification is compared against the industry norm for hydrogen production, steam-methane reforming. These comparisons identify if gasification is better than competing options and thus should be used in the biorefinery.

Sensitivity analysis finds how the project economics change as a function of price changes in the feeds and products. In this case, sensitivity analysis is applied to the hydrogen price, electricity price and feedstock availability. A sensitivity analysis will also be performed for reactor conversion to identify the impact of the reactor being unable to meet the desired performance.

5. EXAMPLE STUDY

5.1 BIOREFINERY SETUP

The scenario proposed is a hydrogenation- or transesterfication-based biorefinery with a nominal capacity of 5000 bbl/day. This biorefinery is proposed in an agricultural region near Beaumont, TX. The region has high rainfall, warm temperatures to support plant growth year-round, close proximity of other industrial facilities and access to all modes of transportation. Looking at the economics of a biorefinery, transportation costs are a major factor because of the low energy density of biomass. A realistic 50-mile transportation radius[35] can be established based on trucking costs to the biorefinery[36-38]. The facility location and surrounding transportation radius are illustrated in Figure 6.
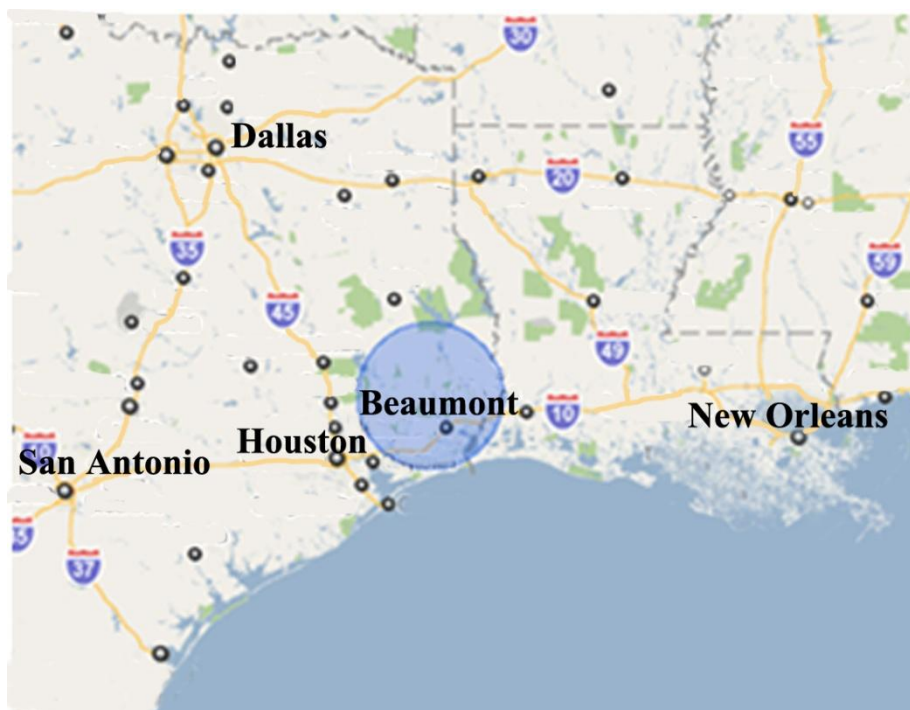

Figure 6. Location of biorefinery in Texas.

Since soybeans, Jatropha and Chinese Tallow are suitable feedstocks to the biorefinery then corresponding residues are available. The composition of these residues is based on published research into each oil seed[38,39]. These selections also demonstrate flexibility in the model to handle multiple feedstocks.

The extraction unit follows the conventional hexane extraction process[40], commonly used for extraction of soybean oil from raw soybeans and was shown earlier in Figure 2. Raw feed is processed by the oil extractor, with extracted flakes and hexane travelling to the desolventizer-toaster, which removes the entrained hexane. The residual biomass is obtained from the desolventizer-toaster and transported to the gasification unit on conveyors.

Additionally to the residual feedstock, the gasification unit can import wood as a virgin biomass feedstock. The composition of this feedstock is selected to fall within the normal range for willow and poplar species[41]. Because this material is not processed by the extraction unit, it has a high water content consistent with a freshly-cut tree. Table 3 provides the composition of feedstock used in the case study.

The cost of the feed handling system is found using literature data[20] and scaled as appropriate to the size of facility under consideration. The operating cost of the feed handling system is combined with the cost of the residue material yielding a combined feed cost. For this case study, no subsidy is paid for disposal of the residue; such a subsidy might exist if sewage sludge or municipal solid waste was being considered.

Table 3. Composition of Biomass Feedstock to Biorefinery

	Soybeans	Chinese Tallow	Jatropha	Wood	Unit
Oil	19.9%	41.0%	47.3%	0.0%	wt%
Non-Oil	80.1%	59.0%	52.8%	100.0%	wt%
Protein	36.5%	11.3%	24.6%	6.0%	wt%
Fibre	9.3%	9.9%	10.1%	12.0%	wt%
Moisture	8.5%	9.0%	5.5%	55.0%	wt%
NC	4.9%	4.0%	4.5%	3.0%	wt%
Carbohydrate	20.9%	24.8%	8.0%	24.0%	wt%
Total	100.0%	100.0%	100.0%	100.0%	wt%
Availability	6815	32843	8119	5000	kg/hr

5.2 REACTOR SETUP

For the purpose of this study, four configurations are selected:

    a) Air at low pressure

    b) Air at high pressure

    c) Oxygen at high pressure

    d) Steam at low pressure

These configurations are of interest for a high capacity gasifier and operate with the conditions shown in Table 4. The model of the process is expandable to many possible configurations of both operating conditions and design. The model can also be expanded to include multiple trains of the same design; such as if the total amount of feed stocks exceeds the maximum size of a single gasifier.

Reactors are not able to scale to an infinite size due to the feasibility of construction and operations. Fluidized bed reactors for air-blown, low-pressure biomass gasification can be built for 20 t/h and are noted to have easy scale up beyond this size[13]. A maximum 50 t/h is established for a reactor of this design. As noted by Hamelinck and Faaij, designs that pressurize the oxidizer and eliminate inert nitrogen in the oxidizer can handle a greater throughput for the same reactor volume[20]. Based on constant mass feed, elimination of nitrogen in the feed can increase throughput by 27%. Pressurization has less of an effect, as the feedstock is solid phase, and here will be assumed to increase throughput by 10%. In the interest of comparison, the largest gasifier at SASOL's world-class facility in South Africa can process 75 t/h of coal[42].

Table 4. Operating Conditions and Design Assumptions for the Reactor Section

	LP Air	HP Air	Oxygen	Steam	Units
Oxidizer	25	25	-20	500	^{o}C
Oxidizer	101	101	3400	700	kPaa
Reactor	850	850	865	847	^{o}C
Reactor	130	3395	3395	365	kPaa
Maximum Throughput	50 000	60 000	66 000	60 000	kg/hr

5.3 GAS CLEANING

Three options for gas cleaning are considered; cyclones, cyclones paired with filters, and cyclones paired with water wash. For the simulation and optimization routine, a number of assumptions for solids removal efficiency and pressure drop are made as provided in Table 5.

Table 5. Design Assumptions for the Gas Cleaning Section

Equipment	Value	Units or Parameter
Cyclone	65.7	% solids removal
Cyclone	5	ΔP, kPa
Filters	95	% solids removal
Filters	5	ΔP, kPa
Water wash	25	^{o} C inlet for water
Water wash	5	ΔP, kPa
Water wash	10 : 1	Water:solids ratio in effluent

5.4 PRODUCTS

In this study five processing options are examined based on their ability to integrate with a biorefinery and compete with fossil energy:

- Fuel gas sales

- Pure electricity

- Electricity from CO; product sales from H2

- Product sales from H2 including steam shift

- Methanol synthesis

For each option examined, specific design and costing assumptions have been made as provided in Table 6 [43,44]. Complex commercial agreements typically cover sales from a major facility. To simplify for this study, for each product a maximum quantity that the bio-refinery is willing to purchase may be established as a constraint. Below this maximum quantify all production is sold at full price and above this maximum quantity, no sales are achieved.

Table 6. Design Assumptions for the Processing Options

Equipment	Value	Units or Parameter
Water-Gas Shift Reactor	87	% conversion of CO
Water-Gas Shift Reactor	425	^{o}C
Water-Gas Shift Reactor	5	ΔP, kPa
Water-Gas Shift Reactor	1	% maximum solids
PSA	14	% hydrogen losses [14]
PSA	99.5	% hydrogen purity
PSA	CO, H2O	Impurity species
Methanol Synthesis Reactor	94	% conversion of CO
Methanol Synthesis Reactor	260	^{o}C
Methanol Synthesis Reactor	5	ΔP, kPa
Methanol Synthesis Reactor	2	% maximum solids
All options	250 000	kg/hr maximum

5.5 ALTERNATIVES FOR THE BIOREFINERY

One possibility for residue disposal is to sell the residue for animal feed. This has the highest potential value, as edible food has a premium price compared with

energy. The sale of soybean residual for animal feed is currently practiced. A lengthy Food and Drug Administration approval is needed for any residues not currently being used as animal meal, in this case Jatropha and Chinese Tallow.

Residues can also be burnt in a wood-fired furnace to generate electricity, following the standard design of a waste-to-energy plant. Combustion minimizes technological risk but does not efficiency use carbon, releasing it all into the environment as carbon dioxide.

Finally, residues can be returned to agricultural areas for spreading on fields, which imposes a transportation cost similar to what was paid to acquire the residue. This is a straight disposal cost and does not offer any economic benefit to the biorefinery, other than simply permitting the biorefinery to operate.

Hydrogen, required for the hydrogenation-based biorefinery, can be created through steam-methane reforming. Reforming is a mature process with a much lower capital cost than gasification but considerable operating costs due to natural gas requirements.

Methanol, required for the transesterfication-based biorefinery, can be purchased from a number of bulk suppliers. Typically methanol is derived from natural gas however biomass-based methanol can be obtained at higher cost if the objective is a wholly renewable process.

5.6 COMMON ECONOMIC ITEMS

A number of assumptions are used for the economic evaluation of all gasification designs and competing options.   A project life of 25 years is assumed with the unit having no scrap value at the end of this life. An interest rate of 5% and an inflation rate of 5% are assumed where necessary. A 20% corporate tax rate is assumed along with straight-line depreciation over the life of the project. Common product and utility prices are provided in Table 7.

Table 7. **Product and Utility Prices**

	Cost ($)	Cost units per
Natural Gas	3.50	GJ
Hydrogen	2.00	kg
Carbon Dioxide Emissions	10	t
Oxygen	29.07	t
Steam	13.3	t
Electricity	0.09	kWh
Boiler Feed Water	0.073	m^3
Methanol	366	t

On-stream factors for large operating biomass gasification units tend to be of dubious merit, as the few units that do exist are demonstration or one-of-a-kind plants. The coal gasification industry is more technologically mature and SASOL reports availability of any one of their coal gasifiers to be over 91%[42].   For this analysis a 90% on-steam factor is assumed.

While equipment and direct costs are obtained from literature or ICARUS, indirect costs are estimated using the factors provided in Table 8[33].

Table 8. Factors for the Calculation of Indirect Costs

Indirect Cost	Value ($/$100 of delivered equipment cost)
Engineering & Supervision	Lang Factor 32
Construction Expenses	Lang Factor 34
Legal Expenses	Lang Factor 4
Contractor's Fees	Lang Factor 19

There is no established carbon tax or credit scheme in the United States at the present, and thus, the base-case regime will consider no carbon taxes or subsidies. Due to uncertainty in this area, a company planning a biofuel facility must evaluate it under a number of different carbon regimes. One possible regime is carbon taxes, where the gasification unit has to pay for carbon dioxide emitted into the atmosphere, at a price of $10/t. This tax regime assumes the seed grower or extractor has already taken any available credits. Another possible regime is carbon subsidies; taxpayer funded credits for producing fuels with lower overall emissions than the petroleum-based fuels currently in service. A life-cycle analysis of biofuels production[45] is used to identify carbon dioxide emissions for the agricultural, extraction hydrogenation and transesterfication processes. Added to the emissions from gasification, the result is compared against the emissions from fossil diesel and the difference credited to the final economics at $10/t.

6. RESULTS AND DISCUSSION

6.1 TARGETING

Applying targeting to the biorefinery scenario, the required hydrogen and methanol demands are provided in Tables 9 and 10. In addition to the demand from the reaction itself, additional hydrogen and methanol are lost in the recycle loop, creating a larger demand for these reactants.

The maximum cash flow for each combination can also be found. This target cash flow does not include annualized capital cost and the related operations & maintenance cost, since the target is an upper bound on performance, and it occurs when such costs are zero. The cash flow targets, given in Table 11, show that steam gasification to produce hydrogen is theoretically the best configuration for the unit. Logically, with no capital cost impediment, all biomass sources are used at their maximum throughput to achieve the maximum cash flow.

Table 9. Targeted Hydrogen Demand

	Mass Flow	Units
Reaction hydrogen demand Hydrodexygenation route	1639	kg/h
Reaction hydrogen demand Decarboxylation route	830	kg/h
Reaction hydrogen demand Average	1234	kg/h
Hydrogen demand including loop losses	1851	kg/h
Hydrogen supply from gasification	5378	kg/h

Table 10. Targeted Methanol Demand

	Mass Flow	Units
Reaction methanol demand Transesterficiation route	4276	kg/h
Methanol demand including loop losses	4704	kg/h
Methanol supply from gasification	43077	kg/h

Table 11. Targeting Results: Cash Flow (in Millions) from Feedstocks & Products

	Fuel Gas	Electricity	Electricity & H2	H2	Methanol
Air	$41	$84	$100	$119	$92
Oxygen	$38	$81	$97	$115	$89
Steam	$22	$119	$168	$181	$175

6.2 EQUIPMENT SIZING AND COSTING

Mass and energy balances acquired from ASPEN Plus simulation are used to perform equipment sizing and costing; these balances are provided in Appendix A. ASPEN ICARUS estimates cost for conventional equipment such as heat exchangers. Literature values are used for the cost estimates of specialized equipment such as gasification vessels and the pressure swing adsorption vessels[20,46,47]. The results from capital cost analysis are provided in detail in Appendix B and summarized in Tables 12, 13, 14 and 15 for solids handling, the gasification reactor, syngas cleaning and final processing option respectively. Operating costs and revenues, calculated from the given product and utility prices and a fraction of capital cost allocated for operations and maintenance, are also provided in the Appendix and tables.

Table 12. Cost Results for Solids Handling

	Soybeans	Jatropha	Chinese Tallow	Woody Biomass
Reference Size (kg/h)	6832	32896	8119	5000
Capital Cost ($MM)	4.5	11.7	5.0	4.8
Operating Cost (¢/kg)	0.30	0.30	0.30	6.80
Operating Revenue (¢/kg)	0	0	0	0

Table 13. Cost Results for the Gasifier Reactor

	LP Air Gasifier	HP Air Gasifier	Oxygen Gasifier	Steam Gasifier
Reference Size (kg/h)	47779	47779	47779	47779
Capital Cost ($MM)	96.1	87.1	70.6	80.3
Operating Cost (¢/kg)	1.81	3.194	1.249	4.408
Operating Revenue (¢/kg)	1.15	1.381	0.585	0.734

Table 14. Cost Results for Syngas Cleaning

	Cyclones	Cyclones & Filters	Water Wash
Reference Size (kg/h)	115091	62988	111788
Capital Cost ($MM)	1.6	2.2	5.2
Operating Cost (¢/kg)	0.017	0.044	0.050
Operating Revenue (¢/kg)	0.001	0.006	0.001

Table 15. **Cost Results for the Gas Processing Option**

	Electricity	Electricity & H2	H2	Fuel Gas	Methanol
Reference Size (kg/h)	112522	107158	59148	112522	52699
Capital Cost ($MM)	162.0	188.9	49.0	0.2	5.8
Operating Cost (¢/kg)	1.205	1.488	1.126	0.002	1.308
Operating Revenue (¢/kg)	3.567	4.638	7.466	1.690	8.875

6.3 OPTIMIZATION

The specific superstructure used in solving the case study is provided in Figure 7. From the material balance, energy balance and cost data the optimization and economic evaluation is performed for each technology separately within the optimization program. Optimization results are first compared with hand-calculated results for the main flow sheets to ensure accuracy. Figure 8 demonstrates this accuracy is indeed achieved with only minor differences due to rounding in the hand calculation.

Each of the technological and processing options is evaluated independently as provided in Table 16 and Table 17. To acquire this independent data, constraints are established in the optimization program to limit choices to the area of interest. When all choices are enabled in the optimization program, the best solution is to use only low-pressure air gasification, as given in Table 18.
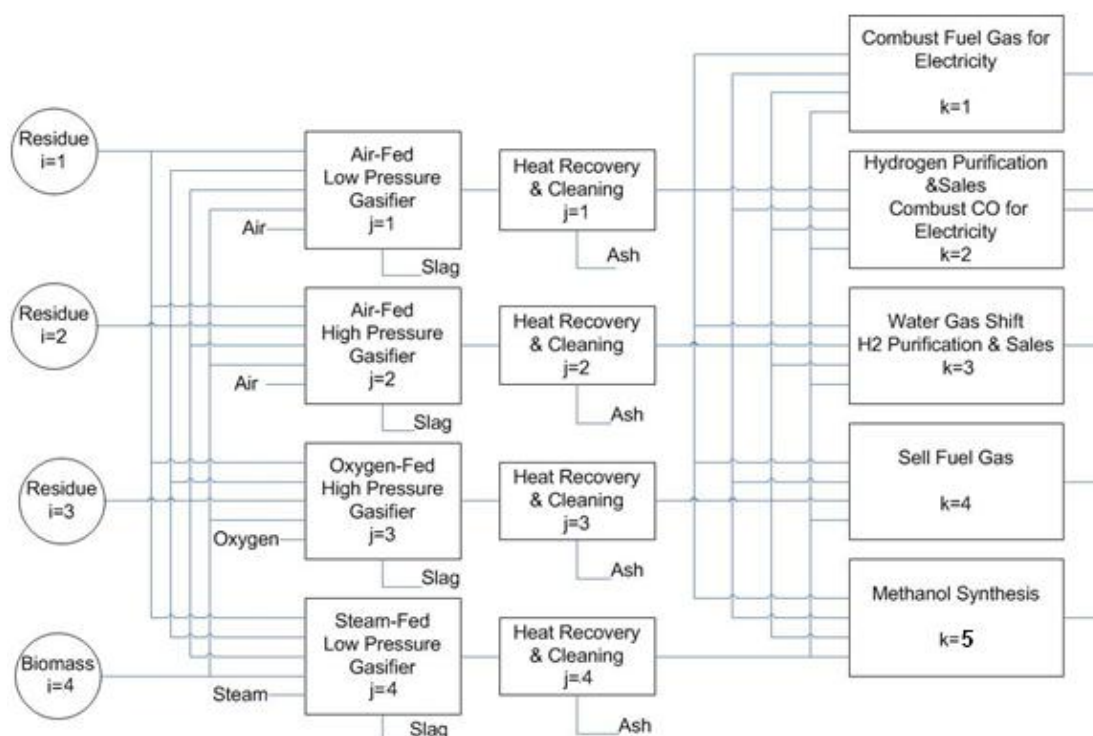
Figure 7. Optimization superstructure, as applied to the case study.
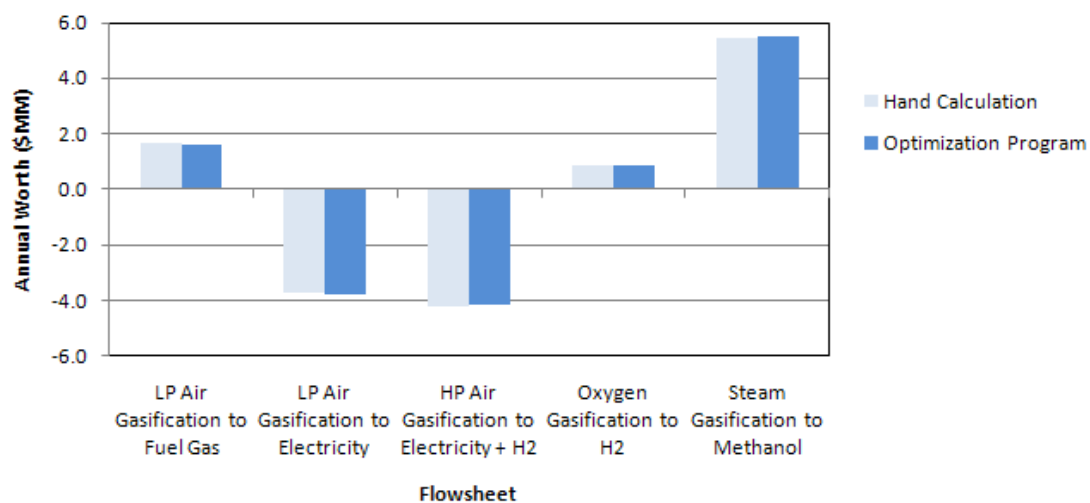

Figure 8. Verification of the accuracy of the optimization program.

The optimal gasification unit consumes all biorefinery residues but does not import fresh biomass as makeup. It produces 2310 kg/hr of hydrogen, more than enough to supply a biorefinery hydrotreating unit, allowing process integration without an extra hydrogen plant. The second-best choice is low-pressure air gasification to produce methanol, with an annual worth only about 1.1% lower than the optimum. Producing 9863 kg/h methanol, it fully satisfies the transesterficiation unit, allowing process integration without an external methanol synthesis plant. Combinations of multiple technologies did not yield a better economic outcome, or in other words, economies of scale outweighed any potential economies from combining technologies.

Table 16. Economic Evaluation with Technologies Considered Separately

	LP Air	HP Air	Oxygen	Steam	Units
Biomass Flow Rate	47777	47777	47777	47777	kg/hr
Preferred Gas Cleaning	Filter	Filter	Filter	Filter	
Preferred Processing Option(s)	Hydrogen	Hydrogen	Hydrogen	Electricity & Hydrogen	
Capital Cost	201	190	148	224	$MM
Annual Op. Cost	13.8	17.8	22.1	25.2	$MM
Annual Revenue	40.1	37.0	34.8	51.4	$MM
Annual After-Tax Cash Flow	23.4	16.9	11.4	22.8	$MM
Annual Worth	9.1	3.4	0.8	6.8	$MM

Table 16 continued

	LP Air	HP Air	Oxygen	Steam	Units
Net Present Value	129	47	12	10	$MM
Rate of Return	9.4%	6.7 %	5.6%	7.6%	
Payback Period	8.6	11.3	13.0	10.3	years

Table 17. Economic Evaluation with Processing Options Considered Separately

	Fuel Gas	Electricity	Electricity & Hydrogen	Hydrogen	Methanol	Units
Biomass Flow Rate	47777	47777	47777	47777	47777	kg/hr
Preferred Oxidizer	LP Air	Steam	Steam	LP Air	LP Air	
Preferred Gas Cleaning	Cyclone	Filter	Filter	Filter	Filter	
Capital Cost	117	203	223	201	127	$MM
Annual Op. Cost	7.7	23.8	25.2	13.8	16.6	$MM
Annual Revenue	18.9	42.1	51.4	40.1	37.9	$MM
Annual After-Tax Cash Flow	9.9	16.2	22.8	23.4	18.0	$MM
Annual Worth	1.6	1.8	6.9	9.1	9.0	$MM
Net Present Value	23	26	97	129	127	$MM
Rate of Return	6.3%	5.9%	8.0%	9.4%	11.7%	
Payback Period	11.8	12.5	9.8	8.6	7.0	years

Table 18. Economic Evaluation with a Fully Optimized Unit

	Optimal Configuration	Units
Biomass Flow Rate	47777	kg/hr
Preferred Oxidizer(s)	LP Air	
Preferred Gas Cleaning	Filters	
Preferred Processing Option(s)	Hydrogen	
Capital Cost	201	$MM
Annual Operating Cost	13.8	$MM
Annual Revenue	40.9	$MM
Annual After-Tax Cash Flow	23.4	$MM
Annual Worth	9.1	$MM
Net Present Value	129	$MM
Rate of Return	9.4%	
Payback Period	8.6	years

To aid business managers and investors removed from the engineering profession, financial indicators obtained by analyzing the model results are provided in Figure 9. As the financial results are detached from the specifics of gasification, they can be used for comparison against a whole range of potential projects and investments that a business might have.
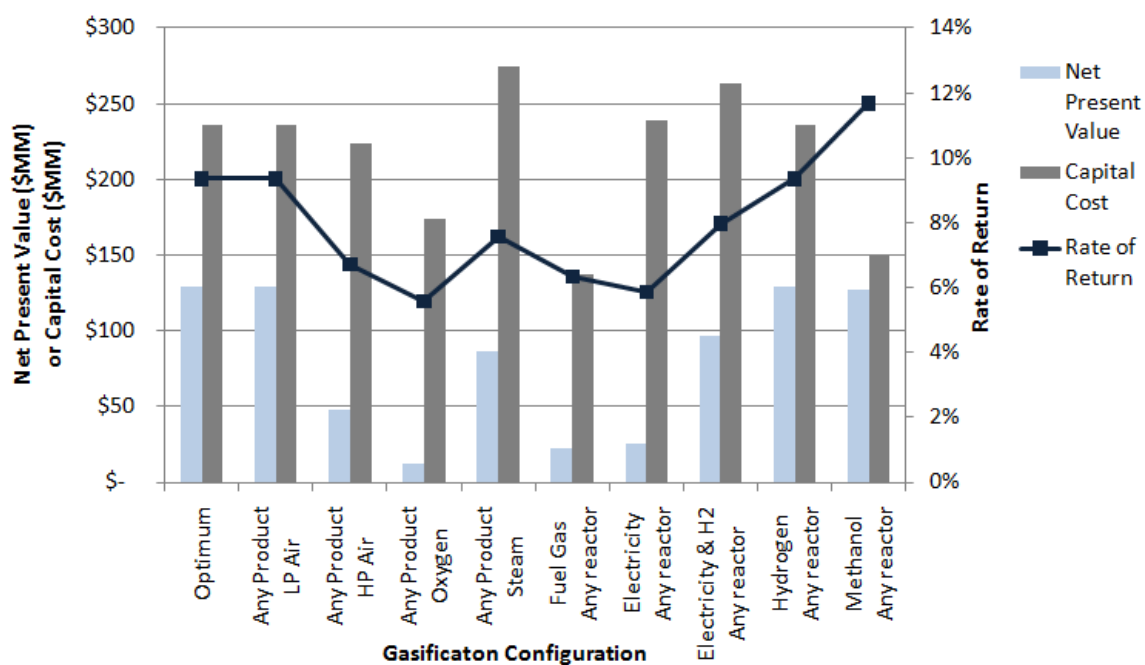
Figure 9. Key financial indicators for various gasification configurations.

It is also useful to compare the results obtained by this optimization routine with literature values and such a comparison is provided in Figure 10, on the basis of equivalent hydrogen price and updated to year 2009[48]. A 10% rate of return is set when calculating hydrogen price for this case study. To linearize the information shown in Figure 10, the data is sorted by the type of reactor rather than study author, and plotted on a log-log graph seen in Figure 11. There is a good correlation between the results of this case study and literature values, as indicated by the position of the current study on the linear trend line in Figure 11, which builds confidence in the method used here. Each author has used different assumptions in calculating the final hydrogen price, therefore, these figures should be understood as a comparison between estimates for hydrogen price and not as a proof of optimality.
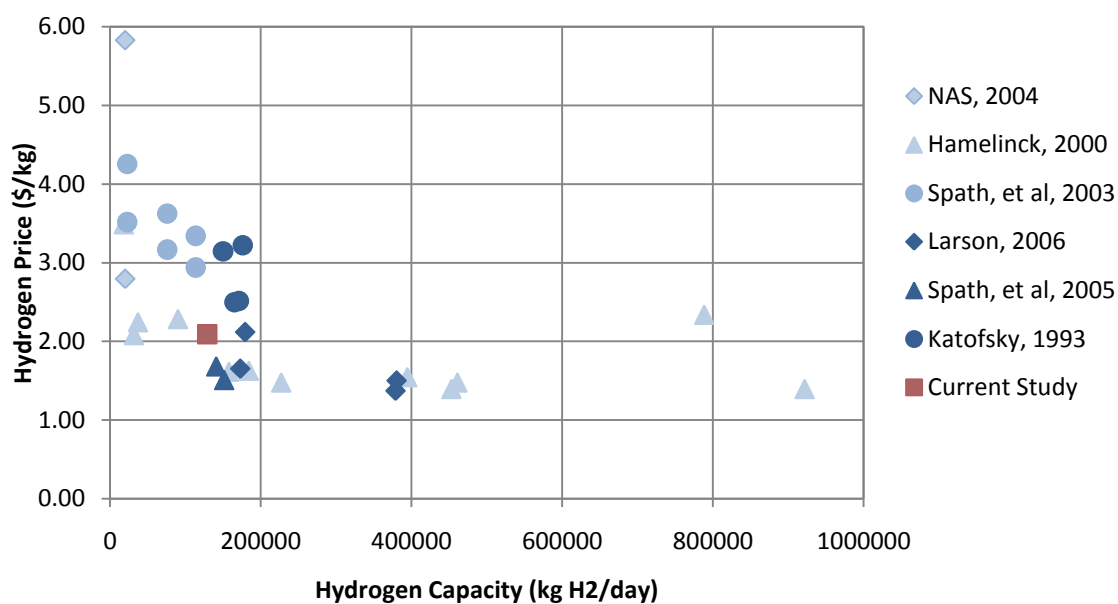
Figure 10. Comparison of calculated and literature values for hydrogen cost, sorted by author.
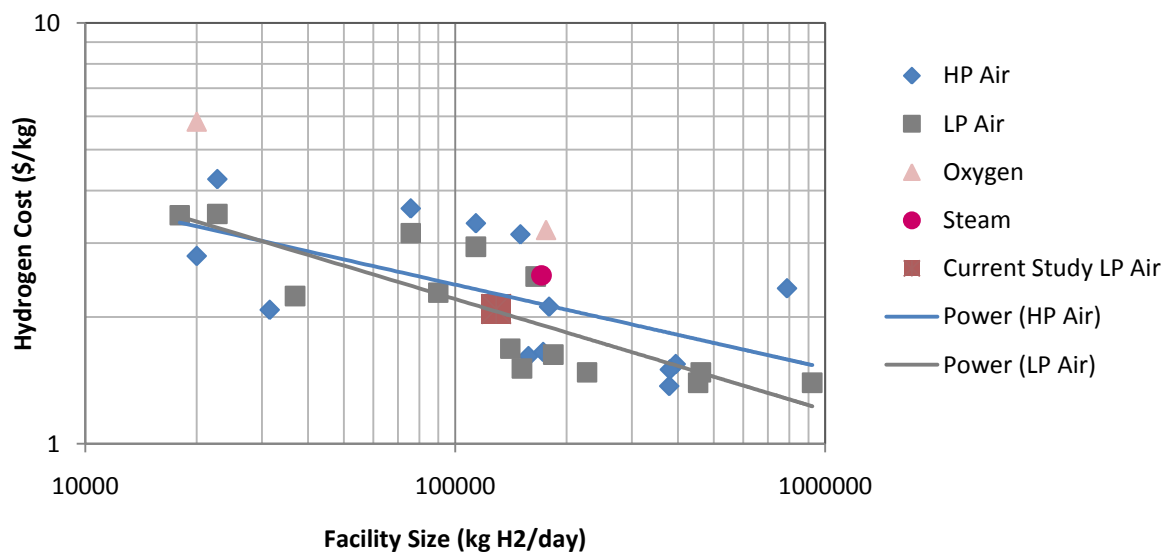

Figure 11. Linearization of calculated and literature values for hydrogen cost, sorted by oxidizer type.

6.4 ALTERNATIVE DISPOSAL OPTIONS AND CARBON DIOXIDE REGIMES

The alternatives to gasification can be evaluated on their own merits and then compared against the economic results that gasification provides. Table 19 shows that animal feed sales is the best competing option, however, such sales are dependent on federal approval for crops like Jatropha and Chinese Tallow that are not currently used for animal feed. Secondly, large-capacity biorefineries could flood the animal meal market and fail to achieve the desired sales prices. Figure 12 shows that a single 5000 bbl/day biorefinery would supply a fifth of the animal meal used throughout a wide region of the United States. Two or three biorefineries would have a flooding effect.

Table 19. Economic Results of Competing Options

|	Combustion	Meal Sales	Disposal
Biomass Flow Rate (kg/hr)	47777	47777	47777
Selling Price ($/t)		102	-7
($/kWh)	0.081		
Capital Cost ($MM)	158	57	57
Annual Op. Cost ($MM)	14.9	16.8	2.7
Annual Revenue ($MM)	33.8	42.3	0
Annual Worth ($MM)	-2.7	54	7.5
Rate of Return	9.3%	36.7%	n/a
Payback Period (years)	9.6	2.7	n/a
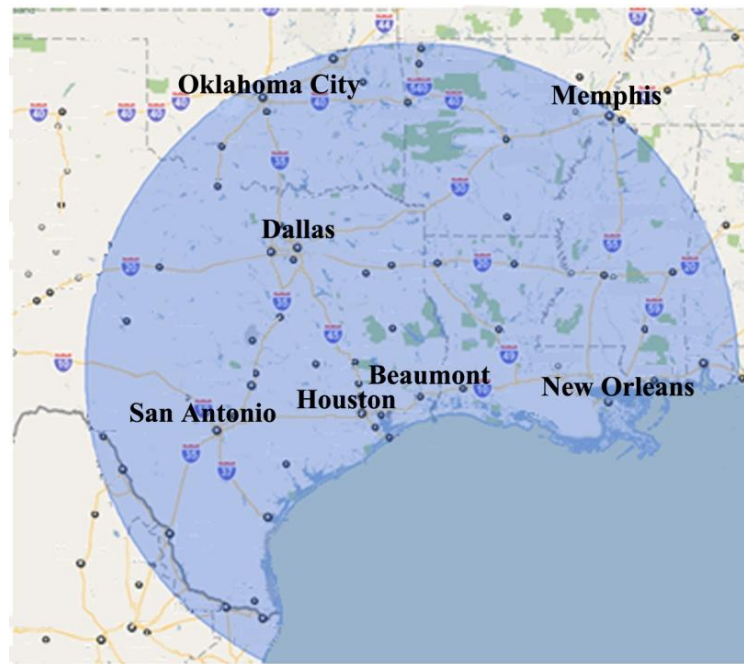
Figure 12. Depiction of animal meal sales region.

Compared with other options to dispose of residue, gasification has a substantially higher capital cost, as illustrated in Figure 13. Without federal approval for meal sales, the option with the highest annual worth is the optimal gasification configuration, exceeding combustion and disposal. However, the carbon dioxide regime has a major impact on the selection of the residual disposal option, because as shown in Figure 14, carbon taxes and carbon credits do not impact the disposal options equally. Carbon taxes have the largest effect on gasification plants producing hydrogen and electricity, along with combustion and meal sales, as all carbon in the residue is eventually released to the environment using these options. Fuel gas production, methanol production and disposal are less impacted as carbon is contained in their product streams. The opposite is true for carbon credits; the credits primarily benefit

fuel gas and methanol production using gasification, along with disposal. When carbon

taxes or credits are applied the optimal solution changes from the production of

hydrogen to the production of methanol.

Also possible within the carbon regimes is a move beyond the $10/t basis

considered thus far. Figure 15 shows the impact of increasing carbon credits on the

optimum solution. The optimum uses low-pressure air gasification, but switches from

hydrogen to methanol once credits are applied. Profitability for other configurations also

increases substantially but they do not ever become the optimum.

In the base case, the optimal gasification unit has a 10.7% rate of return. By

comparison, ExxonMobil has posted an average 23% rate of return in the years between

2002 and 2009[49]. To make a residual biomass project economically competitive with

projects at a major oil company like ExxonMobil, a subsidy of $33/t is required.
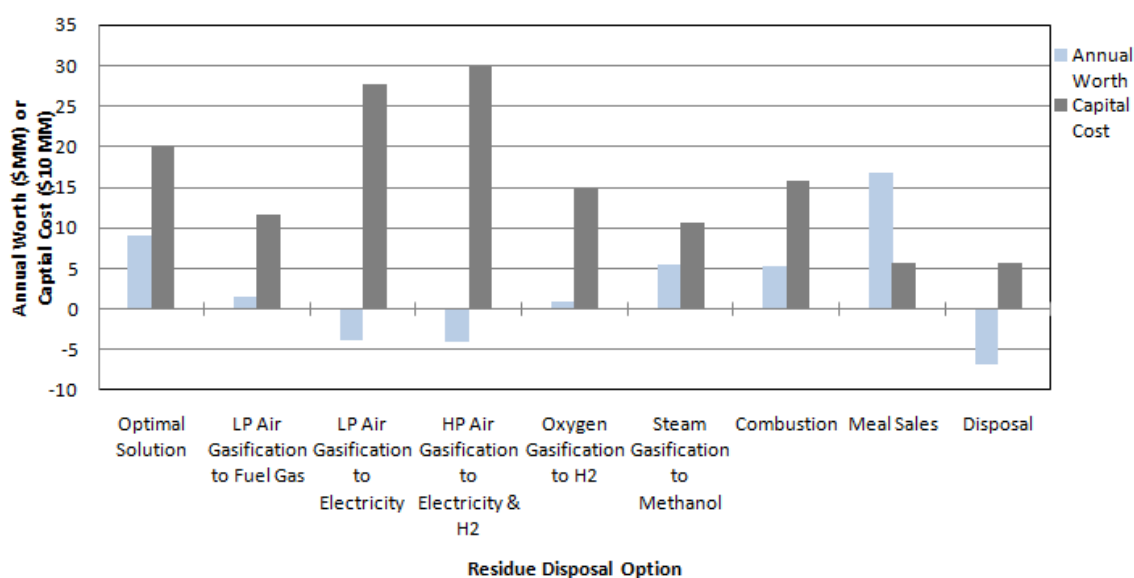
**Figure 13. Comparison of residue disposal options.**
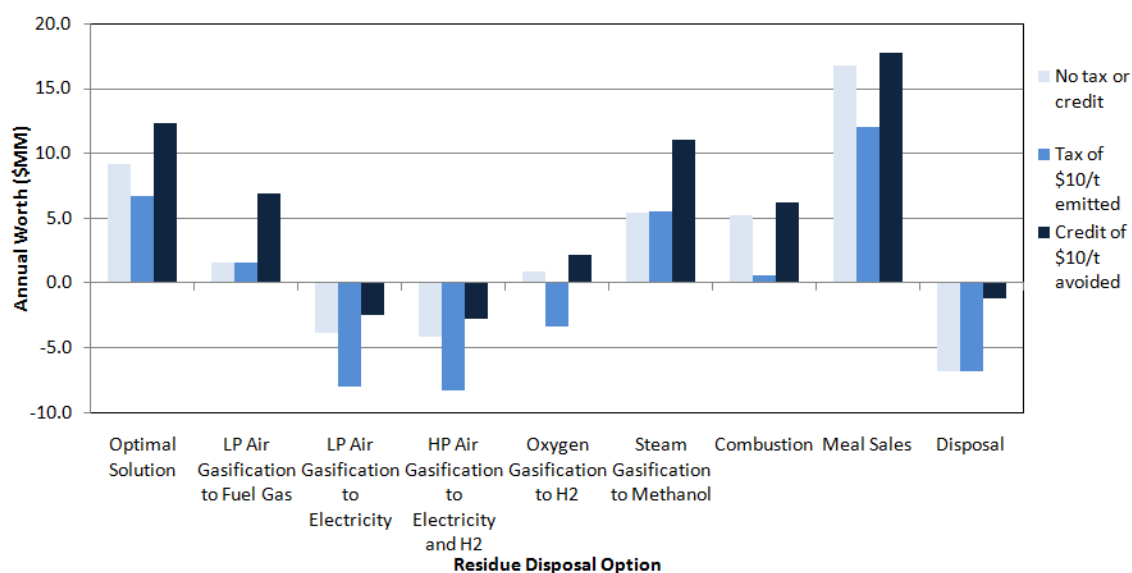
Figure 14. Comparison of residue disposal options under carbon dioxide regimes.
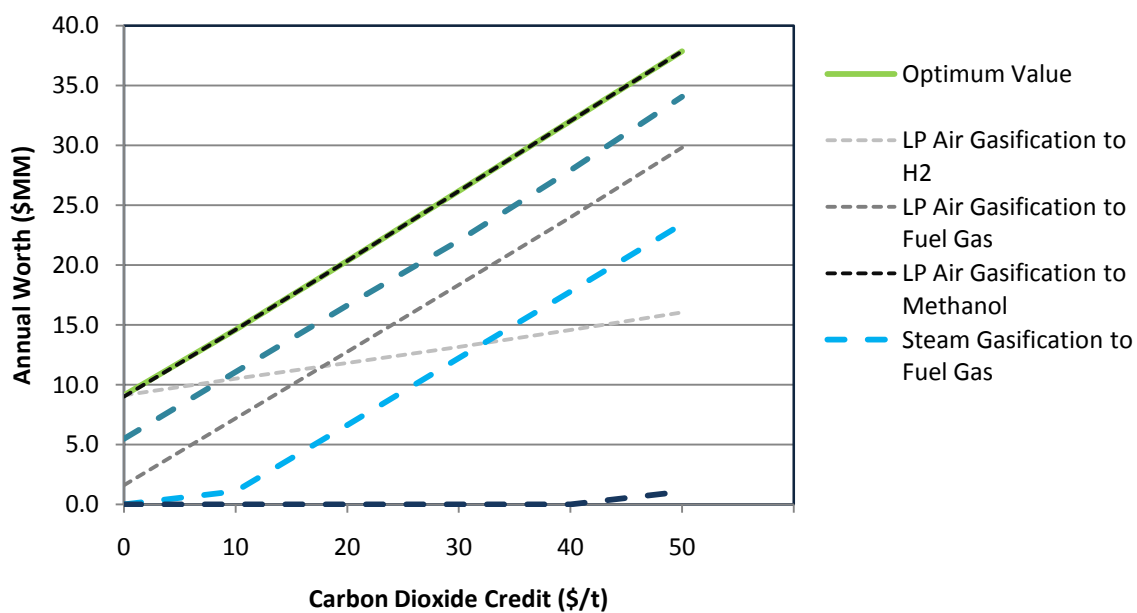

Figure 15. Impact of carbon dioxide credits on annual worth.

At current natural gas prices, steam-methane reforming is more economic than gasification as a hydrogen production technology. Assuming a 10% rate of return, the minimum hydrogen selling price for gasification is $2.09/kg, higher than the target price for the case study. Figure 16 illustrates that the gap between gasification and steam-methane reforming is heavily dependent on natural gas price and only moderately dependent on capacity[50]. Gasification becomes favored at high natural gas prices. Upon increasing size, the unit is able to capture economies of scale but eventually reaches the equipment capacity limit. At the capacity limit, the optimum switches to a higher-capacity single train. Multiple trains are needed to deliver large quantities of hydrogen which prevents major gains from economies of scale.
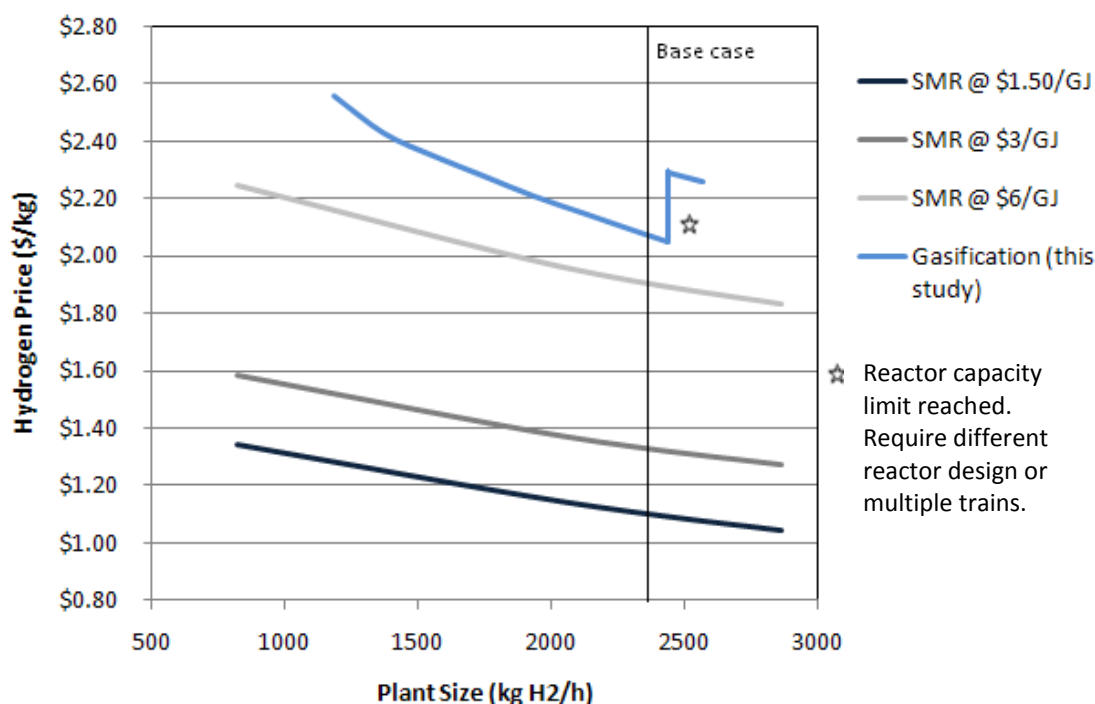

Figure 16. Comparison of gasification to steam-methane reforming.

It is also relevant to compare gasification to methanol synthesis. World-scale methanol plants produce approximately 5000 t/d, about 20 times larger than this biomass gasification plant. Therefore it is illogical to make a direct comparison between the two on the basis on plant size as was done for hydrogen. Instead, Figure 17 shows the impact of gasification plant size against the historical prices for methanol[51]. Methanol prices are very volatile but in general, gasification is competitive at the base case and reduced flowrate case. If the biomass availability is large enough, two reactor trains can be paired with a single methanol synthesis option to capture enhanced economies of scale for the second half of the process.
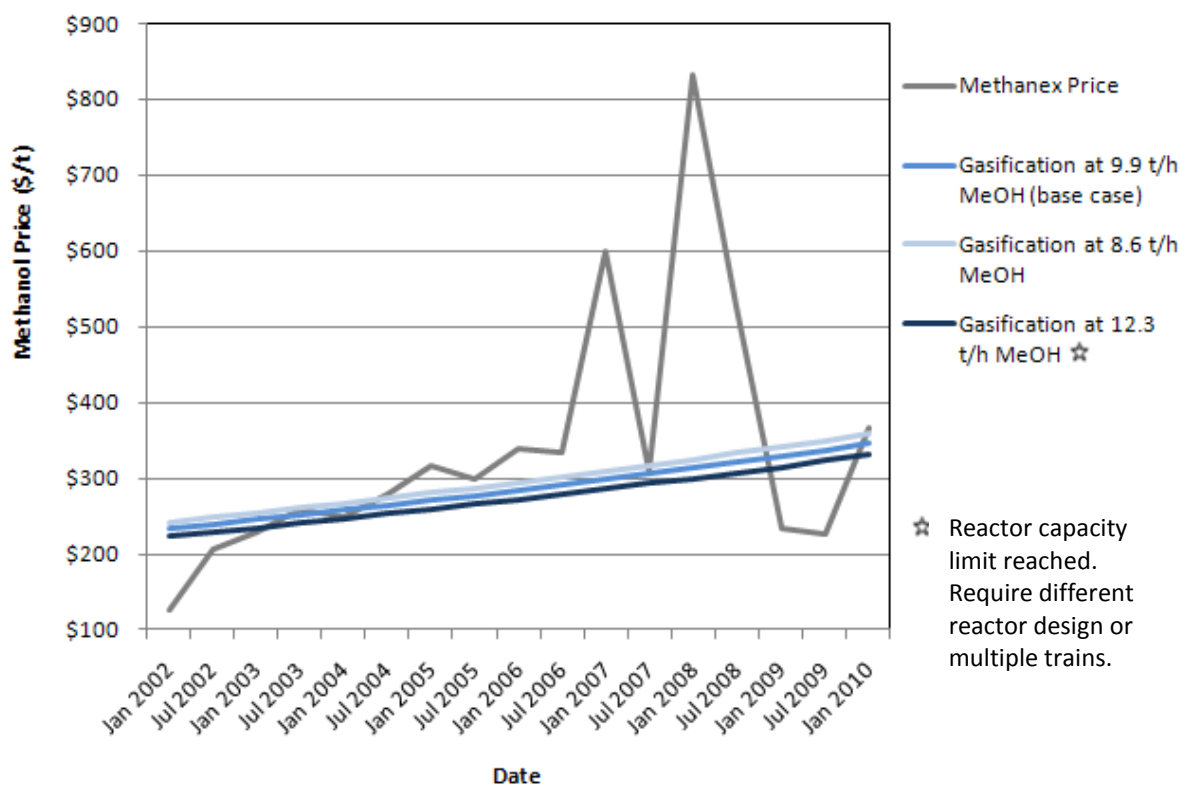

Figure 17. Comparison of gasification to methanol synthesis.

6.5 SENSITIVITY

Being a capital-cost intensive process, it is necessary to operate a gasification unit for many years to achieve maximum economic results. Sensitivity analyses demonstrate the robustness of gasification economics to changes in product prices which are likely to occur over the lifespan of the project. Figure 18 shows that hydrogen price has a significant impact on economics. Hydrogen is the most economic end product for a number of technologies and is already the most profitable choice; price increases further compound this advantage. When the hydrogen price drops even slightly, it begins to become more advantageous to produce methanol. On the other hand, electricity is initially uneconomic, and price decreases continue to give a result that is uneconomic. Price increases favor steam gasification to produce hydrogen and electricity as seen in Figure 19. A number of other configurations also become economic at higher electricity prices but pale in comparison to this option. These sensitivity plots demonstrate that the best configuration and the final economic results are highly dependent on product prices.

Feedstock flow could vary due to crop failure and increase with agricultural improvements. Figure 20 shows the sensitivity to changes in the availability of Chinese Tallow, which is less pronounced than the sensitivity to product prices. Even with a reduced quantity of Chinese Tallow residue, it does not become economic to begin importing virgin biomass, meaning low cost residue is more important than additional economies of scale. With increased quantities of Chinese Tallow residue, the system scales as expected until the maximum size limitation is reached for the low pressure air

systems. The excess residue beyond the maximum is not gasified as it is uneconomic to build a reactor for much a small quantity. Note that, for the sake of readability, obvious inferior choices are omitted from these sensitivity plots.

Finally, the sensitivity to changes in conversion is given in Figure 21. This analysis demonstrates the outcome if the low pressure air reactor exceeds or falls below the performance specified in literature. Other reactors are not included as their conversion values stay constant. Conversion has a substantial impact on final economics; however, it is less likely to vary over time than hydrogen price, electricity price or feed availability.
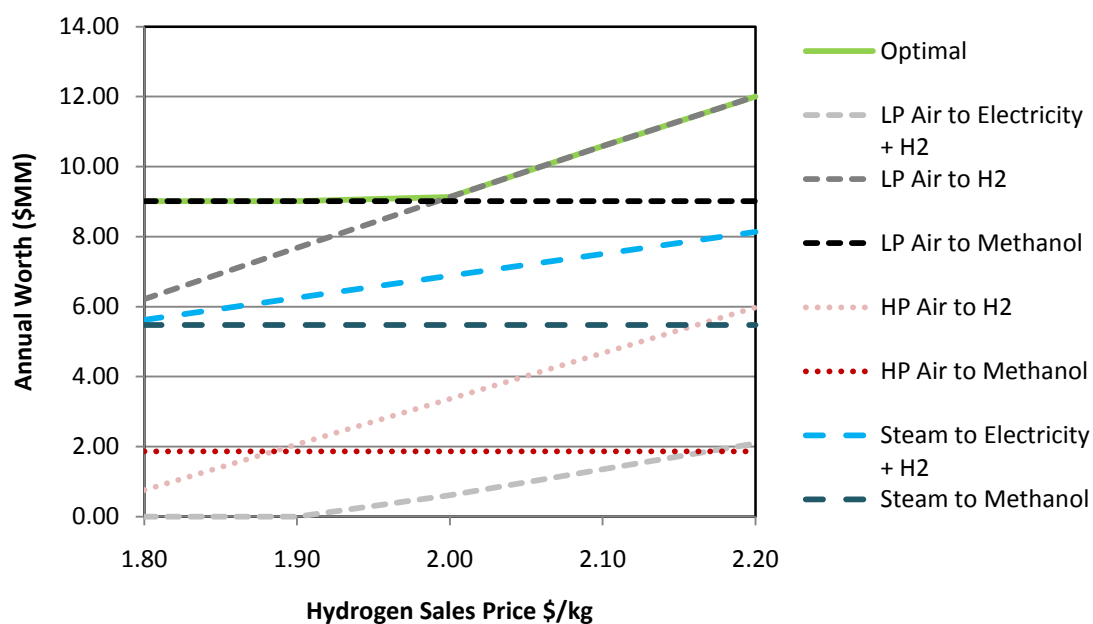

Figure 18. **Sensitivity analysis for changes to hydrogen sales price.**
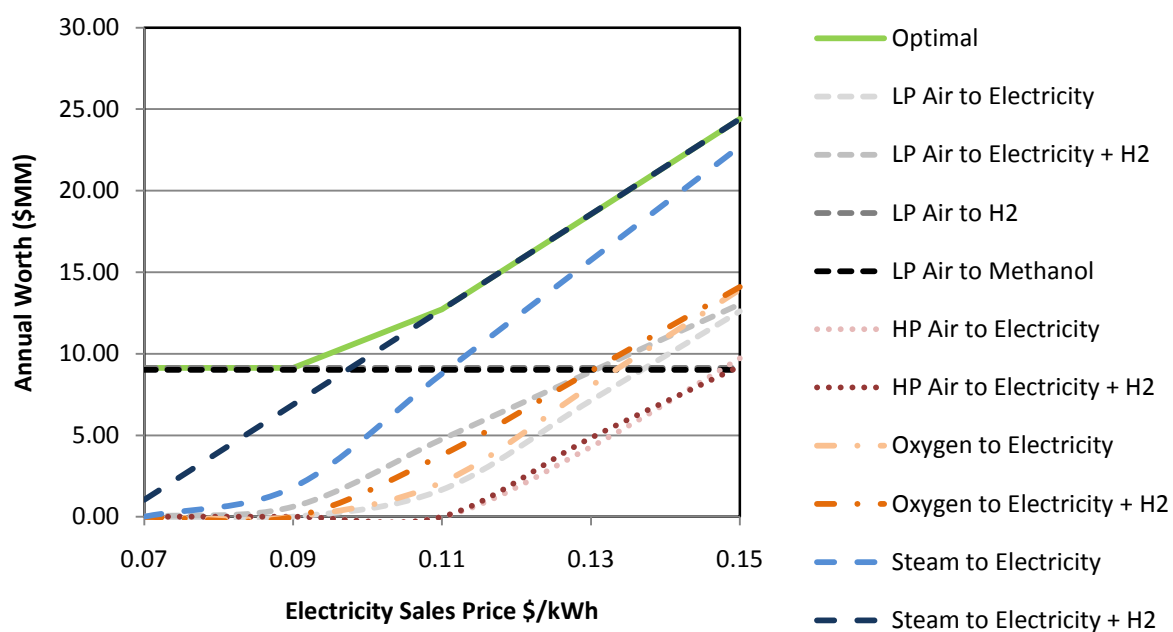
Figure 19. **Sensitivity analysis for changes to electricity price.**
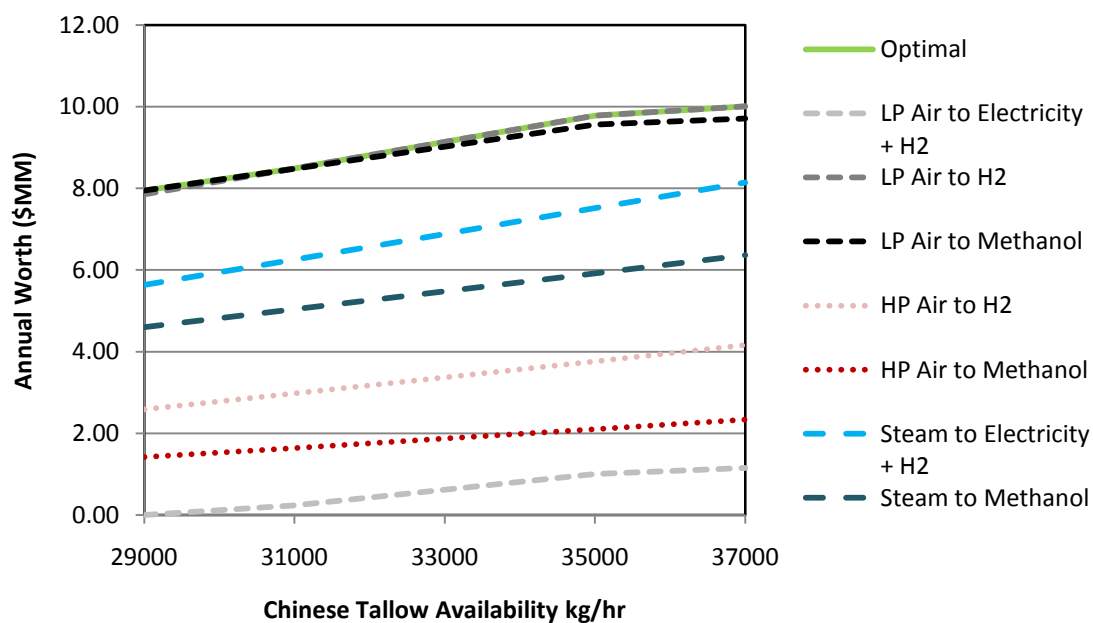

**Figure 20. Sensitivity analysis for changes in Chinese Tallow residue quantity.**
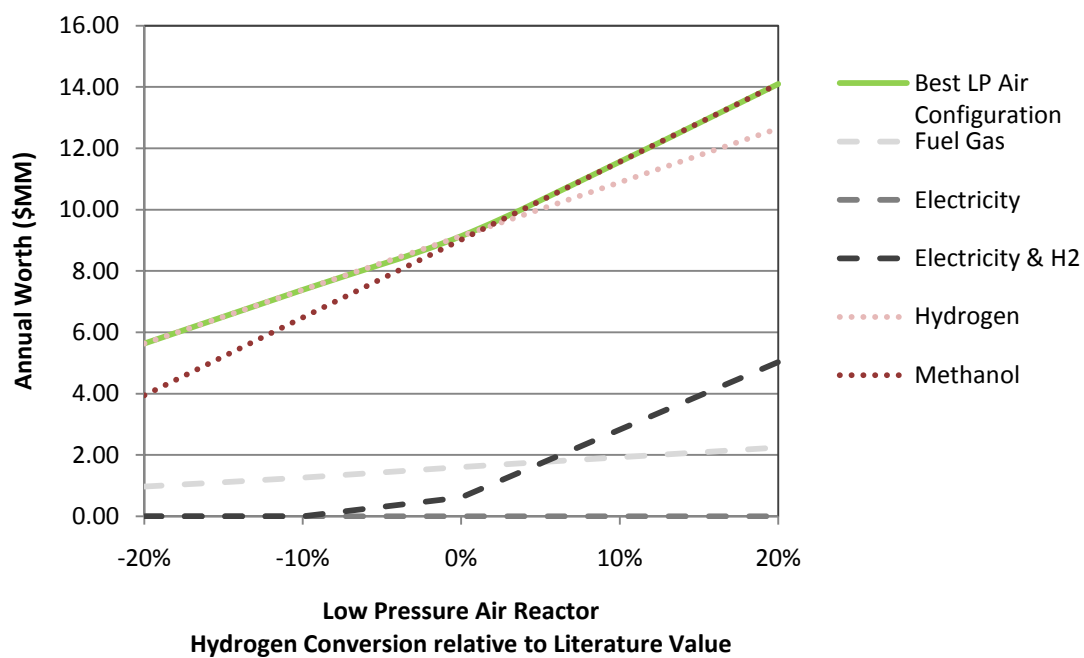
Figure 21. Sensitivity analysis for the change in hydrogen conversion.

7. CONCLUSION

The procedure is successful at optimizing the design of a gasification unit. From the optimization model, interested parties can gather important engineering and financial measures for a wide range of configurations, information that would be extraordinarily time-consuming to obtain through manual iteration.

The results show that very different levels of profitability dependent on the choice of feedstock, technology and product produced. Thus it is important for companies considering gasification products to conduct an investigation and optimization for their unique feedstock and product opportunities and commercially available technologies.

When applied to a representative large-scale scenario under current conditions, low-pressure air gasification for the production of hydrogen was the most economic choice. Low-pressure air gasification for the production of methanol was the second-best choice. In both situations, cyclones paired with filters provided the best syngas cleaning. In all profitable cases, use of the three residue streams was maximized but the addition virgin biomass was not imported due to high cost. This emphasizes gasfication's role as a residual disposal method in a biorefinery; not as a separate biorefinery in its own right.

Gasification is preferable to combustion and disposal for eliminating residue from the biorefinery. It is slightly more expensive than steam-methane reforming as a source of hydrogen, and it compares well against methanol synthesis from natural gas.

Thus it can be concluded that low-pressure air gasification should be strongly considered for inclusion in next-generation biorefineries.

Economies of scale are moderately important to the gasification process; large reactors and large processing options improve economics and make flow splitting between many combinations of technologies a poor choice. Because of the reactor capacity limit, however, economies of scale cannot be exploited indefinitely. Completely new trains have a high cost, but it is possible to combine multiple reactors with a single processing option and deliver enhanced economic performance.

The price of hydrogen and electricity had a critical impact on profitability and configuration, with even small changes of 10% impacting the optimum setup. Thus it can be concluded that a detailed investigation of these product prices must be undertaken before beginning on a gasification project. Carbon dioxide regimes can significantly improve economics; if credit values similar to those available in Europe become available in the USA, residual gasification would be competitive with conventional oil & gas exploration.

The model results confirm the initial prediction that gasification has better returns when configured to compete against petroleum-derived commodities rather than coal-derived electricity. With the challenges of petroleum exhaustion and environmental impact mitigation, it is expected that the next generation of biorefining facilities will be of a size which makes gasification the preferred choice.

Based on these conclusions, there may be interest in expanding the gasification model to include other configurations and details. On the feedstock side, algae residues

could be included, to allow consideration of the growing algae-based biofuels industry. For the reactor section, additional configurations might include supercritical gasification and the use of carbon dioxide as the gasifying agent, the latter of which offers unique possibilities from a carbon trading perspective. Fischer-Tropsch synthesis of liquid fuels could be modeled for an additional processing option. Owing to the risk posed by high-temperature, high-pressure gasification process in comparison with combustion or meal sales, the model could be enhanced with an additional factor characterizing process risk.

Another area of potential future work is to integrate this model with biorefinery operating and scheduling models. This would allow a more comprehensive economic evaluation and optimization of a wider range of designs, including fermentation-based designs which occupy a large portion of US market share.

Finally, since availability is a major problem for gasification, the model could be developed to include different availability factors for each source, reactor, cleaning system and process option. This would then allow the combination of multiple trains in different configurations to achieve on-stream performance comparable with the refining and power generation industries.

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APPENDIX A

FLOW DIAGRAMS, MASS AND ENERGY BALANCES

F-300	C-300	P-300	T-300	X-300	R-300	V-300	E-302	C-301	E-303	X-301	X-302
Air Filter	Air Compressor	BFW Pump	Residue Hopper	Feed Conveyor	Gasification Reactor	Slag Trap	Heat Recovery Exchanger	Cyclone	Product Gas Cooler	Slag Conveyor	Ash Conveyor

**LP Air Gasification
Cyclone cleaning
Electricity and Fuel Gas Products**

Process Flow Diagram

Biomass

Air

BFW

Hydrogen

Syngas to Power Plant

Steam

Ash

Slag

300 302 303 304 305 306 307 301 350 330 331 332 340 310 311 312 320 321 323 324 342 343 349 351 352 354 355 356 357 358 334

F-300
C-300A
C-300B
E-301A
E-301B
T-300
X-300
R-300
V-300
E-302
C-301
V-301
V-303
V-304A
V-304B
V-304C
X-301
X-302
K-300
P-300

F-300 Air Filter	C-300 Air Compressor	E-301 Intercooler	T-300 Residue Hopper	P-300 BFW Pump	X-300 Feed Conveyor	R-300 Gasification Reactor	V-300 Slag Trap	E-302 Heat Recovery Exchanger	C-301 Cyclone	V-301 Water Wash	K-300 Water Filtration Package	X-301 Slag Conveyor	X-302 Ash Conveyor	V-303 Buffer Vessel	V-304 Pressure Swing Adsorption

HP Air Gasification
Cyclone cleaning with water wash
H2 Product with CO to Electricity

Process Flow Diagram
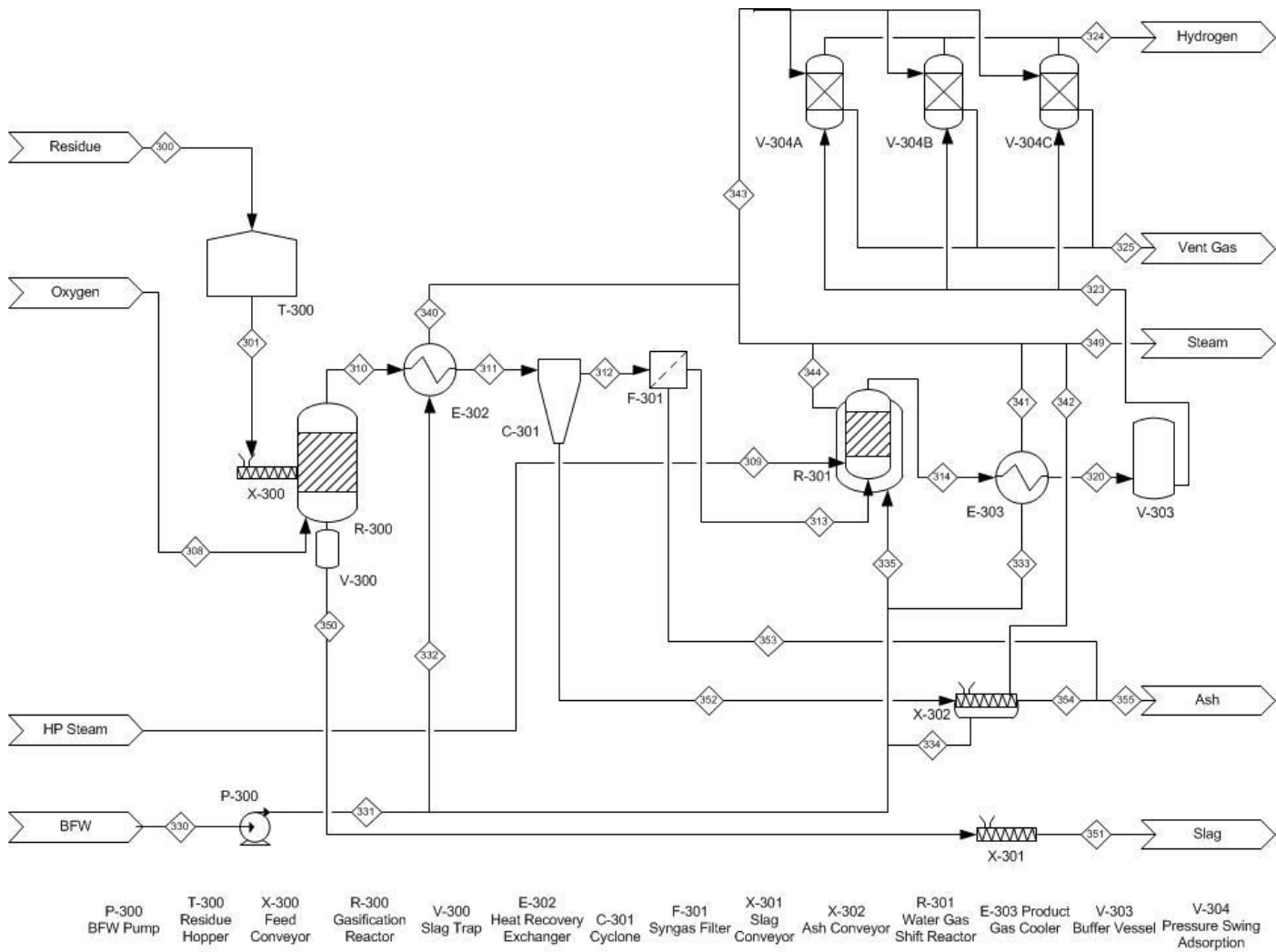
Oxygen Gasification
Cyclone cleaning with Filter
H2 Product

Process Flow Diagram

P-300	T-300	X-300	R-300	V-300	E-302	C-301	F-301	X-301	X-302	R-301	E-303 Product	V-303	V-304
BFW Pump	Residue Hopper	Feed Conveyor	Gasification Reactor	Slag Trap	Heat Recovery Exchanger	Cyclone	Syngas Filter	Slag Conveyor	Ash Conveyor	Water Gas Shift Reactor	Gas Cooler	Buffer Vessel	Pressure Swing Adsorption

Steam Gasification
Cyclone cleaning with filters
Methanol Product

Process Flow Diagram

P-300 BFW Pump	T-300 Residue Hopper	X-300 Feed Conveyor	R-300 Gasification Reactor	V-300 Slag Trap	E-302 Heat Recovery Exchanger	C-301 Cyclone	F-301 Syngas Filter	X-301 Slag Conveyor	X-302 Ash Conveyor	R-302 Methanol Synthesis Reactor	E-303 Product Cooler	V-305 Methanol Separation Column	E-304 Condenser	E-305 Reboiler

Mass Balance: LP Air Gasification to Electricity / Fuel Gas

	300	301	302	303	304	305	306	307	308	309	310	311	312	313	314	320	321	322	323	324	325	327	328
Temperature C	25	25	25	25				56			850	200			200		128						
Pressure kPa	140	140	101	101				130			125	120			115		115						
Mass Vapor Frac	0	0	1.00	0				1.00			0.97	0.97			0.99		0.99						
Mass Solid Frac	0.78	0.78	0	1.00				0			0.03	0.03			0.01		0.01						
Mass Flow kg/hr	47779	47779	68300	68300				68300			115091	115091			112522		112522						
O2	0	0	19807	19807				19807			52	52			52		52						
N2	0	0	48493	48493				48493			50553	50553			50553		50553						
CO	0	0	0	0				0			21612	21612			21612		21612						
CO2	0	0	0	0				0			24360	24360			24360		24360						
CH4	0	0	0	0				0			3767.2	3767			3767		3767						
H2	0	0	0	0				0			1352	1352			1352		1352						
H2O	6463	6463	0	0				0			9485	9485			9485		9485						
STARCH	935	935	0	0				0			0	0			0		0						
TRIGLYC	7704	7704	0	0				0			0	0			0		0						
PROTEIN	12916	12916	0	0				0			0	0			0		0						
CELLULOS	3298	3298	0	0				0			0	0			0		0						
LIGNIN	13194	13194	0	0				0			0	0			0		0						
NC	3269	3269	0	0				0			0	0			0		0						
ASH	0	0	0	0				0			3269	3269			1121		1121						
SOOT	0	0	0	0				0			0	0			0		0						
TAR	0	0	0	0				0			641	641			220		220						
METHANOL	0	0	0	0				0			0	0			0		0						
ETHANE	0	0	0	0				0			0	0			0		0						

	329	330	331	332	333	334	335	340	341	342	343	344	349	350	351	352	353	354	355	356	357	358
Temperature C		25	25	25	25	25		149	149	149			149	850	850	200			100			
Pressure kPa		101	466	466	466	466		466	466	466			466	125	125	115			115			
Mass Vapor Frac		0	0	0	0	0		1.00	1.00	0.97			1.00	0.02	0.02	0			0			
Mass Solid Frac		0	0	0	0	0		0	0	0			0	0.98	0.98	1.00			1.00			
Mass Flow kg/hr		45500	45500	41400	4000	100		41400	4000	100			45500	996	996	2569			2569			
O2		0	0	0	0	0		0	0	0			0	0	0	0			0			
N2		0	0	0	0	0		0	0	0			0	0	0	0			0			
CO		0	0	0	0	0		0	0	0			0	0	0	0			0			
CO2		0	0	0	0	0		0	0	0			0	0	0	0			0			
CH4		0	0	0	0	0		0	0	0			0	0	0	0			0			
H2		0	0	0	0	0		0	0	0			0	0	0	0			0			
H2O		45500	45500	41400	4000	100		41400	4000	100			45500	0	0	0			0			
STARCH		0	0	0	0	0		0	0	0			0	5	5	0			0			
TRIGLYC		0	0	0	0	0		0	0	0			0	39	39	0			0			
PROTEIN		0	0	0	0	0		0	0	0			0	65	65	0			0			
CELLULOS		0	0	0	0	0		0	0	0			0	16	16	0			0			
LIGNIN		0	0	0	0	0		0	0	0			0	66	66	0			0			
NC		0	0	0	0	0		0	0	0			0	0	0	0			0			
ASH		0	0	0	0	0		0	0	0			0	0	0	2148			2148			
SOOT		0	0	0	0	0		0	0	0			0	806	806	0			0			
TAR		0	0	0	0	0		0	0	0			0	0	0	421			421			
METHANOL		0	0	0	0	0		0	0	0			0	0	0	0			0			
ETHANE		0	0	0	0	0		0	0	0			0	0	0	0			0			

Mass Balance: HP Air Gasification to Electricity + Fuel Gas

	300	301	302	303	304	305	306	307	308	309	310	311	312	313	314	320	321	322	323	324	325	327	328
Temperature C	25	25	25	25	301	70	368	70			850	200	200			150	128		150	150			
Pressure kPa	3400	3400	101	101	600	595	3405	3400			3395	3390	3385			3375	466		3375	3375			
Mass Vapor Frac	0	0	1.00	100	1.00	1.00	1.00	1.00			0.97	0.97	0.99			1	1		1	1			
Mass Solid Frac	0.78	0.78	0	0	0	0	0	0			0	0	0.01			0	0		0	0			
Mass Flow kg/hr	47779	47779	65400	65400	65400	65400	65400	65400			111788	111788	109219			107158	107252		107158	907			
O2	0	0	18966	18966	18966	18966	18966	18966			16	16	16			16	16		16	0			
N2	0	0	46434	46434	46434	46434	46434	46434			48494	48494	48494			48492	48492		48492	0			
CO	0	0	0	0	0	0	0	0			21612	21612	21612			21661	21661		21661	4			
CO2	0	0	0	0	0	0	0	0			21407	21407	21407			21407	21407		21407	0			
CH4	0	0	0	0	0	0	0	0			4305	4305	4305			4305	4305		4305	0			
H2	0	0	0	0	0	0	0	0			1048	1048	1048			1047	146.6		1047	901			
H2O	6463	6463	0	0	0	0	0	0			10996	10996	10996			10685	11685		10685	2			
STARCH	935	935	0	0	0	0	0	0			0	0	0			0	0		0	0			
TRIGLYC	7704	7704	0	0	0	0	0	0			0	0	0			0	0		0	0			
PROTEIN	12916	12916	0	0	0	0	0	0			0	0	0			0	0		0	0			
CELLULOS	3298	3298	0	0	0	0	0	0			0	0	0			0	0		0	0			
LIGNIN	13194	13194	0	0	0	0	0	0			0	0	0			0	0		0	0			
NC	3269	3269	0	0	0	0	0	0			0	0	0			0	0		0	0			
ASH	0	0	0	0	0	0	0	0			3269	3269	1121			0	0		0	0			
SOOT	0	0	0	0	0	0	0	0			0	0	0			0	0		0	0			
TAR	0	0	0	0	0	0	0	0			641	641	220			0	0		0	0			
METHANOL	0	0	0	0	0	0	0	0			0	0	0			0	0		0	0			
ETHANE	0	0	0	0	0	0	0	0			0	0	0			0	0		0	0			

	329	330	331	332	333	334	335	340	341	342	343	344	349	350	351	352	353	354	355	356	357	358
Temperature C		25	25	25		25		149		149	149		149	850	850	200		100	120	25	158	158
Pressure kPa		101	466	466		466		466		466	466		466	3395	3395	3385		3385	3375	3400	3375	3375
Mass Vapor Frac		0	0	0		0		1.00		0.97	1.00		1.00	0.01	0.01	0		0	0	0	0	0
Mass Solid Frac		0	0	0		0		0		0	0		0	0.98	0.98	1.00		1	1	0	0.09	1
Mass Flow kg/hr		40300	40300	40200		100		40200		100	1000		39300	1399	1399	2569		2569	3910	13200	15261	1341
O2		0	0	0		0		0		0	0		0	0	0	0		0	0	0	0	0
N2		0	0	0		0		0		0	0		0	0	0	0		0	0	0	130	0
CO		0	0	0		0		0		0	0		0	0	0	0		0	0	0	59	0
CO2		0	0	0		0		0		0	0		0	0	0	0		0	0	0	226	0
CH4		0	0	0		0		0		0	0		0	0	0	0		0	0	0	21	0
H2		0	0	0		0		0		0	0		0	0	0	0		0	0	0	0	0
H2O		40300	40300	40200		100		40200		100	1000		39300	0	0	0		0	0	13200	13483	0
STARCH		0	0	0		0		0		0	0		0	5	5	0		0	0	0	0	0
TRIGLYC		0	0	0		0		0		0	0		0	39	39	0		0	0	0	0	0
PROTEIN		0	0	0		0		0		0	0		0	65	65	0		0	0	0	0	0
CELLULOS		0	0	0		0		0		0	0		0	16	16	0		0	0	0	0	0
LIGNIN		0	0	0		0		0		0	0		0	66	66	0		0	0	0	0	0
NC		0	0	0		0		0		0	0		0	0	0	0		0	0	0	0	0
ASH		0	0	0		0		0		0	0		0	0	0	2148		2148	3269	0	1121	1121
SOOT		0	0	0		0		0		0	0		0	1209	1209	0		0	0	0	0	0
TAR		0	0	0		0		0		0	0		0	0	0	421		421	641	0	220	220
METHANOL		0	0	0		0		0		0	0		0	0	0	0		0	0	0	0	0
ETHANE		0	0	0	0	0		0		0	0		0	0	0	0		0	0	0	0	0

Mass Balance: Oxygen Gasification to Hydrogen

	300	301	302	303	304	305	306	307	308	309	310	311	312	313	314	320	321	322	323	324	325	327	328
Temperature C	25	25							-20	241	865	370	370	370	425	163			163	163	95		
Pressure kPa	3400	3400							3400	3400	3395	3390	3385	3380	3375	3370			3370	3370	300		
Mass Vapor Frac	0	0							1	1	0.94	0.94	0.98	1.00	1.00	0.97			0.97	1.00	0.95		
Mass Solid Frac	0.78	0.78							0	0	0.06	0.06	0.02	0.00	0.00	0.00			0.00	0.00	0.00		
Mass Flow kg/hr	47779	47779							16300	17000	62988	62988	60380	59143	76143	76143			76143	2397	75058		
O2	0	0							16300	0	62	62	62	62	62	62			62	0	62		
N2	0	0							0	0	2060	2060	2060	2060	2060	2060			2060	0	2060		
CO	0	0							0	0	25885	25885	25885	25885	3365	3365			3365	2	3363		
CO2	0	0							0	0	13582	13582	13582	13582	48965	48965			48965	0	48965		
CH4	0	0							0	0	5112	5112	5112	5112	5112	5112			5112	0	5112		
H2	0	0							0	0	777	777	777	777	2398	2398			2398	2062	336		
H2O	6463	6463							0	17000	11600	11600	11600	11600	14115	14115			14115	10	15105		
STARCH	935	935							0	0	0	0	0	0	0	0			0	0	0		
TRIGLYC	7704	7704							0	0	0	0	0	0	0	0			0	0	0		
PROTEIN	12916	12916							0	0	0	0	0	0	0	0			0	0	0		
CELLULOS	3298	3298							0	0	0	0	0	0	0	0			0	0	0		
LIGNIN	13194	13194							0	0	0	0	0	0	0	0			0	0	0		
NC	3269	3269							0	0	0	0	0	0	0	0			0	0	0		
ASH	0	0							0		3269	3269	1089	54	54	54			54	0	54		
SOOT	0	0							0	0	0	0	0	0	0	0			0	0	0		
TAR	0	0							0	0	641	641	213	11	11	11			11	11	0		
METHANOL	0	0							0	0	0	0	0	0	0	0			0	0	0		
ETHANE	0	0							0	0	0	0	0	0	0	0			0	0	0		

	329	330	331	332	333	334	335	340	341	342	343	344	349	350	351	352	353	354	355	356	357	358
Temperature C		25	25	25	25	25	25	149	149	149	149	149	149	865	865	370	370	100	192			
Pressure kPa		101	466	466	466	466	466	466	466	466	461	461	461	3395	3395	3385	3385	3385	3385			
Mass Vapor Frac		0	0	0	0	0	0	0.99	1.00	0.97	0.99	0.98	0.99	0.02	0.02	1.00	1.00	1.00	1.00			
Mass Solid Frac		0	0	0	0	0	0	0	0	0	0	0	0	0.98	0.98	1.00	1.00	1.00	1.00			
Mass Flow kg/hr		43200	43200	21010	14800	290	7100	21010	14800	290	1000	7100	42200	1097	1097	2608	1237	2608	3844			
O2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
N2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
CO		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
CO2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
CH4		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
H2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
H2O		43200	43200	21010	14800	290	7100	21010	14800	290	1000	7100	42200	0	0	0	0	0	0			
STARCH		0	0	0	0	0	0	0	0	0	0	0	0	5	5	0	0	0	0			
TRIGLYC		0	0	0	0	0	0	0	0	0	0	0	0	39	39	0	0	0	0			
PROTEIN		0	0	0	0	0	0	0	0	0	0	0	0	65	65	0	0	0	0			
CELLULOS		0	0	0	0	0	0	0	0	0	0	0	0	16	16	0	0	0	0			
LIGNIN		0	0	0	0	0	0	0	0	0	0	0	0	66	66	0	0	0	0			
NC		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
ASH		0	0	0	0	0	0	0	0	0	0	0	0	0	0	2180	1034	2180	3215			
SOOT		0	0	0	0	0	0	0	0	0	0	0	0	907	907	0	0	0	0			
TAR		0	0	0	0	0	0	0	0	0	0	0	0	0	0	427	203	427	630			
METHANOL		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			
ETHANE		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0			

Mass Balance: Steam Gasification to Methanol

	300	301	302	303	304	305	306	307	308	309	310	311	312	313	314	320	321	322	323	324	325	327	328
Temperature C	25	25								500	849	260	260	260	260	137						5	5
Pressure kPa	370	370								370	365	360	355	350	345	340						300	300
Mass Vapor Frac	0	0								1.00	0.93	0.93	0.98	1.00	1.00	1.00						0	1
Mass Solid Frac	0.78	0.78								0	0.07	0.07	0.02	0.00	0.00	0.00						0	0
Mass Flow kg/hr	47779	47779								11000	56373	56373	53917	52699	52699	52699						9598	34836
O2	0	0								0	0	0	0	0	0	0						0	0
N2	0	0								0	2060	2060	2060	2060	2060	2060						2	2058
CO	0	0								0	7892	7892	7892	7892	474	474						1	473
CO2	0	0								0	21775	21775	21775	21775	21775	21775						483	21292
CH4	0	0								0	5919	5919	5919	5919	5919	5919						19	5901
H2	0	0								0	1148	1148	1148	1148	80	80						0	80
H2O	6463	6463								11000	9174	9174	9174	9174	9174	9174						1184	15
STARCH	935	935								0	0	0	0	0	0	0						0	0
TRIGLYC	7704	7704								0	0	0	0	0	0	0						0	0
PROTEIN	12916	12916								0	0	0	0	0	0	0						0	0
CELLULOS	3298	3298								0	0	0	0	0	0	0						0	0
LIGNIN	13194	13194								0	0	0	0	0	0	0						0	0
NC	3269	3269								0	0	0	0	0	0	0						0	0
ASH	0	0								0	3269	3269	1121	56	56	56						0	0
SOOT	0	0								0	0	0	0	0	0	0						0	0
TAR	0	0								0	470	470	161	8	8	8						0	0
METHANOL	0	0								0	0	0	0	0	8486	8486						7759	502
ETHANE	0	0								0	4665	4665	4665	4665	4665	4665						151	4514

	329	330	331	332	333	334	335	340	341	342	343	344	349	350	351	352	353	354	355	356	357	358
Temperature C	130	25	25	25	25	25	25	149	149	149		191	154	849	849	260	260	100	155			
Pressure kPa	300	101	466	466	466	466	466	461	466	466		466	461	365	365	355	355	355	355			
Mass Vapor Frac	0.00	0	0	0	0	0	0	0.99	1.00	0.90		1.00	1.00	0.01	0.01	0	0	0	0			
Mass Solid Frac	0.01	0	0	0	0	0	0	0	0	0		0	0	0.99	0.99	1.00	1.00	1.00	1.00			
Mass Flow kg/hr	8264	40000	40000	26340	4000	160	9500	26340	4000	160		9500	40000	2406	2406	2456	1218	2456	3675			
O2	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
N2	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
CO	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
CO2	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
CH4	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
H2	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
H2O	7975	40000	40000	26340	4000	160	9500	26340	4000	160		9500	40000	0	0	0	0	0	0			
STARCH	0	0	0	0	0	0	0	0	0	0		0	0	5	5	0	0	0	0			
TRIGLYC	0	0	0	0	0	0	0	0	0	0		0	0	39	39	0	0	0	0			
PROTEIN	0	0	0	0	0	0	0	0	0	0		0	0	65	65	0	0	0	0			
CELLULOS	0	0	0	0	0	0	0	0	0	0		0	0	16	16	0	0	0	0			
LIGNIN	0	0	0	0	0	0	0	0	0	0		0	0	66	66	0	0	0	0			
NC	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
ASH	56	0	0	0	0	0	0	0	0	0		0	0	0	0	2148	1065	2148	3213			
SOOT	0	0	0	0	0	0	0	0	0	0		0	0	2216	2216	0	0	0	0			
TAR	8	0	0	0	0	0	0	0	0	0		0	0	0	0	309	153	309	462			
METHANOL	225	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			
ETHANE	0	0	0	0	0	0	0	0	0	0		0	0	0	0	0	0	0	0			

Energy Balance: LP Air Gasification to Electricity or Fuel Gas

Equipment	Description	Duty (MW)	Stream Type	Temperature IN (degrees C)	Temperature OUT (degrees C)	Other Information
C-300	Air Compressor	0.6	n/a	n/a	n/a	Electrical
E-302	Heat Recovery Exchanger	30.6	hot	850	200	
E-303	Product Gas Cooler	3.0	hot	200	128	
P-300	BFW Pump	0.01	n/a	n/a	n/a	Electrical
X-300	Feed Auger	0.01	n/a	n/a	n/a	Electrical
X-301	Slag Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Cooling	0.1	hot	200	100	

Energy Balance: HP Air Gasification to Electricity + Fuel Gas

Equipment	Description	Duty (MW)	Stream Type	Temperature IN (degrees C)	Temperature OUT (degrees C)	Other Information
C-300	Air Compressor	5.0	n/a	n/a	n/a	Electrical
C-301	Air Compressor	3.8	n/a	n/a	n/a	Electrical
E-301A	Intercooler	4.3	hot	301	70	
E-301B	Intercooler	5.6	hot	367	70	
E-302	Heat Recovery Exchanger	29.7	hot	850	200	
K-300	Water Filtration Package	0.03	n/a	n/a	n/s	Electrical
P-300	BFW Pump	0.01	n/a	n/a	n/a	Electrical
X-300	Feed Auger	0.01	n/a	n/a	n/a	Electrical
X-301	Slag Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Cooling	0.1	hot	200	100	

Energy Balance: Oxygen Gasification to Hydrogen

Equipment	Description	Duty (MW)	Stream Type	Temperature IN (degrees C)	Temperature OUT (degrees C)	Other Information
E-302	Heat Recovery Exchanger	15.5	hot	865	370	
E-303	Product Gas Cooler	11.0	hot	425	163	
P-300	BFW Pump	0.01	n/a	n/a	n/a	Electrical
R-301	Water Gas Shift Reactor	5.2	hot	425	425	Exothermic Rxn.
X-300	Feed Auger	0.01	n/a	n/a	n/a	Electrical
X-301	Slag Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Cooling	0.2	hot	370	100	

Energy Balance: Steam Gasification to Methanol

Equipment	Description	Duty (MW)	Stream Type	Temperature IN (degrees C)	Temperature OUT (degrees C)	Other Information
E-302	Heat Recovery Exchanger	19.4	hot	847	260	
E-303	Product Gas Cooler	3.0	hot	260	137	
E-304	Condenser	31.5	cold	91	5	Refrigeration
E-305	Reboiler	20.9	hot	120	130	
P-300	BFW Pump	0.01	n/a	n/a	n/a	Electrical
R-300	Gasification Reactor	43.2	cold	850	850	Endothermic Rxn.
R-302	Methanol Synthesis Reactor	7.3	hot	260	260	Exothermic Rxn.
X-300	Feed Auger	0.01	n/a	n/a	n/a	Electrical
X-301	Slag Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Conveyor	0.01	n/a	n/a	n/a	Electrical
X-302	Ash Cooling	0.1	hot			

APPENDIX B

CAPITAL COST, OPERATING COST AND REVENUE

Capital Cost: LP Air Gasification to Fuel Gas - Reactor Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-300	Air Compressor	57492 m3/h from 0 kPag to 29 kPag	$ 1,228,400	$ 1,499,700	Icarus	2007	$ 1,282,175	$ 1,565,351
E-302	Heat Recovery Exchanger	Comparison with existing exchanger ($6.99M for 39.2 kg/s steam)	$ 1,674,535	$ 3,349,070	Hamelinck et al Chem Eng	2001	$ 2,328,975	$ 4,657,951
F-300	Air Filter	57492 m3/h bag filters	$ 184,800	$ 231,000	Process Design	2004	$ 253,361	$ 316,701
P-300	BFW Pump	12.31 L/s with 37.2 m head	$ 5,600	$ 37,000	Icarus	2007	$ 5,845	$ 38,620
R-300	Gasification Reactor	Comparison with existing low pressure gasifier ($44.3M for 68.8 dry t/h)	$ 28,476,256	$ 35,595,320	Hamelinck	2001	$ 39,605,322	$ 49,506,653
V-700	Slag Trap	Included in reactor cost						
X-300	Feed Conveyor	Triple 9 inch screw, up to 5.5 kg/s, 60 foot	$ 126,000	$ 541,800	Plant Design and Economics	2002	$ 174,667	$ 751,070
X-302	Slag Conveyor	9 inch screw, up to 5.5 kg/s, 60 foot	$ 42,000	$ 180,600	Plant Design and Economics	2002	$ 58,222	$ 250,357
Total							$ 43,708,568	$ 57,086,702

Capital Cost: LP Air Gasification to Fuel Gas - Gas Cleaning Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-301	Cyclone	Seven 1500mm cyclones at 200 C	$ 247,800	$ 739,700	Icarus	2007	$ 258,648	$ 772,081
X-302	Ash Conveyor, Jacketed	9 inch screw, up to 5.5 kg/s, 60 foot	$ 96,200	$ 288,600	Plant Design and Economics	2002	$ 114,795	$ 363,084
Total							$ 373,443	$ 1,135,166

Capital Cost: LP Air Gasification to Fuel Gas -Process Option Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
E-303	Product Cooler	47 m2 at 230 C, 534 kPag tube and 323 kPag shell	$ 22,400	$ 79,400	Icarus	2007	$ 23,381	$ 82,876
Total							$ 23,381	$ 82,876

Capital Cost: LP Air Gasification to Fuel Gas Indirect Costs

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Engineering & Supervision	Lang Factor of 32	Plant Design and Economics	2009	$ 13,986,742	$ 119,502	$ 7,482
Construction Expenses	Lang Factor of 34	Plant Design and Economics	2009	$ 14,860,913	$ 126,971	$ 7,949
Legal Expenses	Lang Factor of 4	Plant Design and Economics	2009	$ 1,748,343	$ 14,938	$ 935
Contractor's Fees	Lang Factor of 19 + $100000	Plant Design and Economics	2009	$ 8,404,628	$ 170,954	$ 104,442
Total				$ 39,000,626	$ 432,364	$ 120,809

Capital Cost: LP Air Gasification to Fuel Gas Cost Summary

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Fixed Capital	Sum of Direct & Indirect	n/a	2009	$ 96,087,328	$ 1,567,530	$ 203,685
Working Capital	15% of Total	Plant Design and Economics	2009	$ 16,956,587	$ 276,623	$ 35,944
Total			2009	$ 113,043,915	$ 1,844,153	$ 239,629

Capital Cost: LP Air Gasification to Electricity - Reactor Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-300	Air Compressor	57492 m3/h from 0 kPag to 29 kPag	$ 1,228,400	$ 1,499,700	Icarus	2007	$ 1,282,175	$ 1,565,351
E-302	Heat Recovery Exchanger	Comparison with existing exchanger ($6.99M for 39.2 kg/s steam). Equip cost	$ 1,674,535	$ 3,349,070	Hamelinck et al Chem Eng	2001	$ 2,328,975	$ 4,657,951
F-300	Air Filter	57492 m3/h bag filters	$ 184,800	$ 231,000	Process Design	2004	$ 253,361	$ 316,701
P-300	BFW Pump	12.31 L/s with 37.2 m head	$ 5,600	$ 37,000	Icarus	2007	$ 5,845	$ 38,620
R-300	Gasification Reactor	Comparison with existing low pressure gasifier ($44.3M for 68.8 dry t/h).	$ 28,476,256	$ 35,595,320	Hamelinck et al	2001	$ 39,605,322	$ 49,506,653
V-700	Slag Trap	Included in reactor cost						
X-300	Feed Conveyor	Triple 9 inch screw, up to 5.5 kg/s, 60 foot	$ 126,000	$ 541,800	Plant Design and Economics	2002	$ 174,667	$ 751,070
X-302	Slag Conveyor	9 inch screw, up to 5.5 kg/s, 60 foot	$ 42,000	$ 180,600	Plant Design and Economics	2002	$ 58,222	$ 250,357
Total							$ 43,708,568	$ 57,086,702

Capital Cost: LP Air Gasification to Electricity - Gas Cleaning Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-301	Cyclone	Seven 1500mm cyclones at 200 C	$ 247,800	$ 739,700	Icarus	2007	$ 258,648	$ 772,081
X-302	Ash Conveyor, Jacketed	9 inch screw, up to 5.5 kg/s, 60 foot	$ 96,200	$ 288,600	Plant Design and Economics	2002	$ 114,795	$ 363,084
Total							$ 373,443	$ 1,135,166

Capital Cost: LP Air Gasification to Electricity - Process Option

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
E-303	Product Cooler	47 m2 at 230 C, 534 kPag tube and 323 kPag shell	$ 22,400	$ 79,400	Icarus	2007	$ 23,381	$ 82,876
K-301	Power Plant	Comparison against gas turbine with HRSG (26.3 MW for $35.5 MM)	$ 48,150,239	$ 60,187,799	Hamelinck et al	2001	$ 66,968,275	$ 83,710,344
Total							$ 87,767,381	$ 83,793,220

Capital Cost: LP Air Gasification to Electricity Indirect Costs

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Engineering & Supervision	Lang Factor of 32	Plant Design and Economics	2009	$ 13,986,742	$ 119,502	$ 28,085,562
Construction Expenses	Lang Factor of 34	Plant Design and Economics	2009	$ 14,860,913	$ 126,971	$ 29,840,909
Legal Expenses	Lang Factor of 4	Plant Design and Economics	2009	$ 1,748,343	$ 14,938	$ 3,510,695
Contractor's Fees	Lang Factor of 19 +$100000	Plant Design and Economics	2009	$ 8,404,628	$ 170,954	$ 16,775,802
Total				$ 39,000,626	$ 432,364	$ 78,212,969

Capital Cost: LP Air Gasification to Electricity Cost Summary

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Fixed Capital	Sum of Direct & Indirect	n/a	2009	$ 96,087,328	$ 1,567,530	$ 162,006,189
Working Capital	15% of TCI	Plant Design and Economics	2009	$ 16,956,587	$ 276,623	$ 28,589,327
Total Capital	Fixed + Working	n/a	2009	$ 113,043,915	$ 1,844,153	$ 190,595,516

Capital Cost: HP Air Gasification to Electricity and H2 - Reactor Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-300A	Air Compressor	55051 m3/h from 0 kPag to 499 kPag	$ 3,244,400	$ 3,522,700	Icarus	2007	$ 3,386,427	$ 3,676,910
C-300B	Air Compressor	10790 m3/h from 494 kPag to 3304 kPag	$ 2,116,700	$ 2,325,000	Icarus	2007	$ 2,209,361	$ 2,426,780
E-301A	Intercooler	142 m2 at 330 C, 413 kPag tube, 668 kPag shell	$ 34,100	$ 120,500	Icarus	2007	$ 35,593	$ 125,775
E-301B	Intercooler	94.7 m2 at 397 C, 2398 kPag tube, 3648 kPag shell	$ 35,500	$ 153,900	Icarus	2007	$ 37,054	$ 160,637
E-302	Heat Recovery Exchanger	Comparison with existing exchanger ($6.99M for 39.2 kg/s steam	$ 1,645,242	$ 3,290,483	Hamelinck	2001	$ 2,288,234	$ 4,576,467
F-300	Air Filter	55051 m3/h bag filters	$ 179,520	$ 224,400	Chem Eng Process Design	2004	$ 246,122	$ 307,652
P-300	BFW Pump	12.31 L/s with 37.2 m head	$ 5,500	$ 35,000	Icarus	2007	$ 5,741	$ 36,532
R-300	Gasification Reactor	Comparison with existing high pressure gasifier ($38.1M for 68.1 dry t/h)	$ 22,583,833	$ 28,229,791	Hamelinck	2001	$ 31,410,028	$ 39,262,535
V-700	Slag Trap	Included in reactor cost						
X-300	Feed Conveyor	Triple 9 inch screw, up to 5.5 kg/s, 60 foot	$ 126,000	$ 541,800	Plant Design and Economics	2002	$ 174,667	$ 751,070
X-302	Slag Conveyor	9 inch screw, up to 5.5 kg/s, 60 foot	$ 42,000	$ 180,600	Plant Design and Economics	2002	$ 58,222	$ 250,357
Total							$ 39,851,449	$ 51,574,715

Capital Cost: HP Air Gasification to Electricity and H2 - Gas Cleaning Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-301	Cyclone	Single 1295mm cyclone at 200 C	$ 28,600	$ 46,600	Icarus	2007	$ 29,852	$ 48,640
K-300	Water Filtration Package	Primary Treatment 13.2 m3/hr, Pressure Factor 3	$ 1,050,000	$ 2,205,000	Chem Eng Process Design	2004	$ 1,439,550	$ 3,023,055
V-301	Water Wash	1.67 m diameter, 5 tray vessel, pressure 3618 kPag	$ 72,300	$ 223,700	Icarus	2007	$ 75,465	$ 233,493
X-302	Ash Conveyor, Jacketed	9 inch screw, up to 5.5 kg/s, 60 foot	$ 96,200	$ 288,600	Plant Design and Economics	2002	$ 114,795	$ 363,084
Total							$ 1,659,662	$ 3,668,272

Capital Cost: HP Air Gasification to Electricity and H2 - Process Option Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
V-303	Buffer Vessel	5 minutes (389 m3) at 3618 kPag	$ 879,300	$ 1,325,700	Icarus	2007	$ 917,792	$ 1,383,734
V-304	PSA	$24MM per kmol/s	$ 11,856,000	$ 29,640,000	T Kreutz	2004	$ 16,254,576	$ 40,636,440
K-301	Power Plant	Comparison against gas turbine with HRSG (26.3 MW for $35.5 MM)	$ 41,956,883	$ 52,446,103	Hamelinck et al	2001	$ 58,354,437	$ 72,943,046
Total							$ 82,980,368	$ 114,963,220

Capital Cost: HP Air Gasification to Electricity and H2 Indirect Costs

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Engineering & Supervision	Lang Factor of 32	Plant Design and Economics	2009	$ 12,752,464	$ 531,092	$ 26,553,718
Construction Expenses	Lang Factor of 34	Plant Design and Economics	2009	$ 13,549,493	$ 564,285	$ 28,213,325
Legal Expenses	Lang Factor of 4	Plant Design and Economics	2009	$ 1,594,058	$ 66,386	$ 3,319,215
Contractor's Fees	Lang Factor of 19 +$100000	Plant Design and Economics	2009	$ 7,671,775	$ 415,336	$ 15,866,270
Total				$ 35,567,790	$ 1,577,099	$ 73,952,528

Capital Cost: HP Air Gasification to Electricity and H2 Cost Summary

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Fixed Capital	Sum of Direct & Indirect	n/a	2009	$ 87,142,505	$ 5,245,371	$ 188,915,748
Working Capital	15% of TCI	Plant Design and Economics	2009	$ 15,378,089	$ 925,654	$ 33,338,073
Total Capital	Fixed + Working	n/a	2009	$ 102,520,594	$ 6,171,025	$ 222,253,821

Capital Cost: Oxygen Gasification to H2 - Reactor Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
E-302	Heat Recovery Exchanger	Comparison with existing exchanger ($6.99M for 39.2 kg/s steam). Equip cost	$ 520,337	$ 1,040,674	Hamelinck et al	2001	$ 723,695	$ 1,447,389
P-300	BFW Pump	12.31 L/s with 37.2 m head	$ 5,600	$ 37,000	Icarus	2007	$ 5,845	$ 38,620
R-300	Gasification Reactor	Comparison with existing high pressure gasifier ($38.1M for 68.1 dry t/h)	$ 22,583,833	$ 28,229,791	Hamelinck et al	2001	$ 31,410,028	$ 39,262,535
V-700	Slag Trap	Included in reactor cost						
X-300	Feed Conveyor	Triple 9 inch screw, up to 5.5 kg/s, 60 foot	$ 126,000	$ 541,800	Plant Design and Economics	2002	$ 174,667	$ 751,070
X-302	Slag Conveyor	9 inch screw, up to 5.5 kg/s, 60 foot	$ 42,000	$ 180,600	Plant Design and Economics	2002	$ 58,222	$ 250,357
Total							$ 32,372,458	$ 41,749,970

Capital Cost: Oxygen Gasification to H2 - Gas Cleaning Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-301	Cyclone	Single 1295mm cyclone at 4189 m3/h	$ 28,600	$ 46,400	Icarus	2007	$ 29,852	$ 48,431
F-301	Syngas Filter	Plate and Frame	$ 701,600	$ 961,800	Icarus	2007	$ 732,313	$ 1,003,904
X-302	Ash Conveyor, Jacketed	9 inch screw, up to 5.5 kg/s, 60 foot	$ 96,200	$ 288,600	Plant Design and Economics	2002	$ 114,795	$ 363,084
Total							$ 876,960	$ 1,415,419

Capital Cost: Oxygen Gasification to H2 - Process Option

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
E-303	Product Gas Cooler	89 m2, 3618 kPag shell, 2378 kPag tube	$ 40,400	$ 162,900	Icarus	2007	$ 42,169	$ 170,031
R-301	WGS Reactor	Comparison with existing WGS reactor ($36.9M for 15.6 Mmol CO+H2/h). Equip	$ 8,342,057	$ 10,427,571	Hamelinck et al	2001	$ 11,602,292	$ 14,502,865
V-303	Buffer Vessel	5 minutes (314 m3) at 3618 kPag	$ 879,300	$ 1,325,700	Icarus	2007	$ 545,500	$ 998,100
V-304	PSA	$24MM per kmol/s. Equip cost estimated at 50% of direct cost.	$ 11,998,800	$ 23,997,600	T Kreutz	2004	$ 16,450,355	$ 32,900,710
Total							$ 16,995,855	$ 33,898,810

Capital Cost: Oxygen Gasification to H2 Indirect Costs

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Engineering & Supervision	Lang Factor of 32	Plant Design and Economics	2009	$ 10,359,186	$ 280,627	$ 5,438,674
Construction Expenses	Lang Factor of 34	Plant Design and Economics	2009	$ 11,006,636	$ 298,167	$ 5,778,591
Legal Expenses	Lang Factor of 4	Plant Design and Economics	2009	$ 1,294,898	$ 35,078	$ 679,834
Contractor's Fees	Lang Factor of 19 + $100000	Plant Design and Economics	2009	$ 6,150,767	$ 166,622	$ 3,229,212
Total				$ 28,811,487	$ 780,495	$ 15,126,311

Capital Cost: Oxygen Gasification to H2 Cost Summary

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Fixed Capital	Sum of Direct & Indirect	n/a	2009	$ 70,561,458	$ 2,195,914	$ 49,025,120
Working Capital	15% of TCI	Plant Design and Economics	2009	$ 12,452,022	$ 387,514	$ 8,651,492
Total Capital	Fixed + Working	n/a	2009	$ 83,013,479	$ 2,583,429	$ 57,676,612

Capital Cost: Steam Gasification to Methanol - Reactor Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
E-302	Heat Recovery Exchanger	Comparison with existing exchanger ($6.99M for 68.1 dry t/h)	$ 652,341	$ 1,304,681	Hamelinck et al	2001	$ 907,288	$ 1,814,576
P-300	BFW Pump	12.2 L/s with 37.2 m head	$ 5,500	$ 35,000	Icarus	2007	$ 5,741	$ 36,532
R-300	Gasification Reactor	Comparison with existing low pressure gasifier ($44.3M for 68.8 dry t/h)	$ 25,628,630	$ 32,035,788	Hamelinck et al	2001	$ 35,644,790	$ 44,555,988
V-700	Slag Trap	Included in reactor cost						
X-300	Feed Conveyor	Triple 9 inch screw, up to 5.5 kg/s, 60 foot	$ 126,000	$ 541,800	Plant Design and Economics	2002	$ 174,667	$ 751,070
X-302	Slag Conveyor	9 inch screw, up to 5.5 kg/s, 60 foot	$ 42,000	$ 180,600	Plant Design and Economics	2002	$ 58,222	$ 250,357
Total							$ 36,790,708	$ 47,408,521

Capital Cost: Steam Gasification to Methanol - Gas Cleaning Section

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
C-301	Cyclone	Dual 1500mm cyclones at 200 C	$ 70,800	$ 139,700	Icarus	2007	$ 73,899	$ 145,816
F-301	Syngas Filter	Plate and Frame	$ 701,600	$ 961,800	Icarus	2007	$ 732,313	$ 1,003,904
X-302	Ash Conveyor, Jacketed	9 inch screw, up to 5.5 kg/s, 60 foot	$ 96,200	$ 288,600	Plant Design and Economics	2002	$ 114,795	$ 363,084
Total							$ 921,008	$ 1,512,804

Capital Cost: Steam Gasification to Methanol - Process Option

Item	Description	Sizing Criteria	Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
E-303	Product Cooler	31 m2 at 290 C, 534 kPag tube and 413 kPag shell	$ 112,800	$ 207,100	Icarus	2007	$ 117,738	$ 216,166
E-304	Condenser + Accumulator +	725 m2 at 125 C, 243 kPag tube and shell. 19m3 Accumulator. Pump at 56 L/s.	$ 158,200	$ 454,100	Icarus	2007	$ 165,125	$ 473,979
E-305	Reboiler	492 m2 at 290 C, 757 kPag tube and 471 kPag shell	$ 86,800	$ 159,364	Icarus	2007	$ 90,600	$ 166,341
R-302	Methanol Synthesis	Comparison with existing gaseous methanol synthesis reactor ($7.0M for 87.5	$ 1,380,090	$ 1,725,112	Hamelinck et al	2001	$ 1,919,456	$ 2,399,319
V-304	Methanol Separation	5 tray column, diameter of 3.05m and height of 6.25 m	$ 97,100	$ 318,300	Icarus	2007	$ 101,351	$ 332,234
Total							$ 2,394,269	$ 3,588,039

Capital Cost: Steam Gasification to Methanol Indirect Costs

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Engineering & Supervision	Lang Factor of 32	Plant Design and Economics	2009	$ 11,773,027	$ 294,723	$ 766,166
Construction Expenses	Lang Factor of 34	Plant Design and Economics	2009	$ 12,508,841	$ 313,143	$ 814,052
Legal Expenses	Lang Factor of 4	Plant Design and Economics	2009	$ 1,471,628	$ 36,840	$ 95,771
Contractor's Fees	Lang Factor of 19 +$100000	Plant Design and Economics	2009	$ 7,090,235	$ 274,991	$ 554,911
Total				$ 32,843,731	$ 919,697	$ 2,230,900

Capital Cost: Steam Gasification to Methanol Cost Summary

Item	Sizing Criteria	Reference	Cost Date	Reactor Section	Gas Cleaning Section	Option Section
Fixed Capital	Sum of Direct & Indirect	n/a	2009	$ 80,252,252	$ 2,432,501	$ 5,818,938
Working Capital	15% of TCI	Plant Design and Economics	2009	$ 14,162,162	$ 429,265	$ 1,026,871
Total Capital	Fixed + Working	n/a	2009	$ 94,414,414	$ 2,861,766	$ 6,845,810

Capital Cost: Feed Handling Equipment for Soybeans

Equipment	Sizing Criteria	FOB Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
Conveyers	Compare to 33.5 wet t/h	$ 134,825	$ 417,957	Hamelinck and Faaij	2001	$ 187,517	$ 581,303
Grinding	Compare to 33.5 wet t/h	$ 157,938	$ 489,607	Hamelinck and Faaij	2001	$ 219,663	$ 680,954
Storage	Compare to 33.5 wet t/h	$ 385,214	$ 1,194,163	Hamelinck and Faaij	2001	$ 535,763	$ 1,660,864
Total	Referense 6832 kg/hr					$ 942,942	$ 2,923,121

Capital Cost: Feed Handling Equipment for Chinese Tallow

Equipment	Sizing Criteria	FOB Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
Conveyers	Compare to 33.5 wet t/h	$ 346,029	$ 1,072,691	Hamelinck and Faaij	2001	$ 481,264	$ 1,491,920
Grinding	Compare to 33.5 wet t/h	$ 405,349	$ 1,256,581	Hamelinck and Faaij	2001	$ 563,767	$ 1,747,677
Storage	Compare to 33.5 wet t/h	$ 988,656	$ 3,064,832	Hamelinck and Faaij	2001	$ 1,375,041	$ 4,262,627
Total	Reference 32896 kg/hr					$ 2,420,072	$ 7,502,224

Capital Cost: Feed Handling Equipment for Jatropha

Equipment	Sizing Criteria	FOB Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
Conveyers	Compare to 33.5 wet t/h	$ 149,535	$ 463,558	Hamelinck and Faaij	2001	$ 207,976	$ 644,726
Grinding	Compare to 33.5 wet t/h	$ 175,170	$ 543,025	Hamelinck and Faaij	2001	$ 243,629	$ 755,250
Storage	Compare to 33.5 wet t/h	$ 427,243	$ 1,324,452	Hamelinck and Faaij	2001	$ 594,217	$ 1,842,074
Total	Reference 8119 kg/hr					$ 1,045,822	$ 3,242,050

Capital Cost: Feed Handling Equipment for Wood

Equipment	Sizing Criteria	FOB Equip. Cost	Direct Cost	Reference	Cost Date	2009 Equip Cost	2009 Direct Cost
Conveyers	Compare to 33.5 wet t/h	$ 111,795	$ 346,565	Hamelinck and Faaij	2001	$ 155,487	$ 482,009
Grinding	Compare to 33.5 wet t/h	$ 130,960	$ 405,976	Hamelinck and Faaij	2001	$ 182,142	$ 564,640
Storage	Compare to 33.5 wet t/h	$ 319,415	$ 990,186	Hamelinck and Faaij	2001	$ 444,248	$ 1,377,170
Iron Removal	Compare to 33.5 wet t/h	$ 118,183	$ 366,369	Hamelinck and Faaij	2001	$ 164,372	$ 509,553
Loading Dock	n/a	$ 100,000	$ 100,000	estimated	2009	$ 100,000	$ 139,082
Total	Reference 5000 kg/h					$ 1,046,249	$ 3,072,454

Capital Cost: Feed Handling Indirect Costs

Item	Sizing Criteria	Reference	Cost Date	Soybeans	Chinese Tallow	Jatropha	Wood
Engineering & Supervision	Lang Factor of 32	Plant Design and Economics	2009	$ 301,742	$ 774,423	$ 334,663	$ 334,800
Construction Expenses	Lang Factor of 34	Plant Design and Economics	2009	$ 320,600	$ 822,825	$ 355,580	$ 355,725
Legal Expenses	Lang Factor of 4	Plant Design and Economics	2009	$ 37,718	$ 96,803	$ 41,833	$ 41,850
Contractor's Fees	Lang Factor of 19	Plant Design and Economics	2009	$ 179,159	$ 459,814	$ 198,706	$ 198,787
Contingency	Lang Factor of 10	specified	2009	$ 94,294	$ 242,007	$ 104,582	$ 104,625
Total				$ 933,513	$ 2,395,872	$ 1,035,364	$ 1,035,786

Capital Cost: Feed Handling Cost Summary

Item	Sizing Criteria	Reference	Cost Date	Soybeans	Chinese Tallow	Jatropha	Wood
Fixed Capital	Sum of Direct & Indirect	n/a	2009	$ 3,856,634	$ 9,898,096	$ 4,277,414	$ 4,108,240
Working Capital	15% of Total	Plant Design and Economics for	2009	$ 680,583	$ 1,746,723	$ 754,838	$ 724,984
Total			2009	$ 4,537,217	$ 11,644,819	$ 5,032,252	$ 4,833,223

Operating Cost: LP Air Gasification to Fuel Gas

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Cost to Reactor	Cost to Cleaning	Cost to Option
Boiler Feed Water	Supply to unit	$ 0.071	m3	2009	$ 0.07	Chemical Engineering: Process Design and Economics A Practical Guide	$ 23,262	$ 2,248	$ -
Slag	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ 17,227	$ -	$ -
Ash	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ 44,423	$ -
Electricity	For rotating equipment	$ 0.09	kWh	2009	$ 0.09	Source: Energy Information Administration, April 2010	$ 444,894	$ 7,096	$ -
Operations	Supplies and lab expenses	0.6% of capital	n/a	2009	0.6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 576,524	$ 9,405	$ 1,222
Maintenance	Salaries, parts, tools,	6% of capital	n/a	2009	6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 5,765,240	$ 94,052	$ 12,221
Total							$ 6,827,146	$ 157,223	$ 13,443

Operating Cost: LP Air Gasification to Electricity

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Cost to Reactor	Cost to Cleaning	Cost to Option
Boiler Feed Water	Supply to unit	$ 0.071	m3	2009	$ 0.07	Chemical Engineering: Process Design and Economics A Practical Guide	$ 23,262	$ 2,248	$ -
Slag	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ 17,227	$ -	$ -
Ash	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ 44,423	$ -
Electricity	For rotating equipment	$ 0.09	kWh	2009	$ 0.09	Source: Energy Information Administration, April 2010	$ 444,894	$ 7,096	$ -
Operations	Supplies and lab expenses	0.6% of capital	n/a	2009	0.6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 576,524	$ 9,405	$ 972,037
Maintenance	Salaries, parts, tools,	6% of capital	n/a	2009	6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 5,765,240	$ 94,052	$ 9,720,371
Total							$ 6,827,146	$ 157,223	$ 10,692,408

Operating Cost: HP Air Gasification to Electricity and H2

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Cost to Reactor	Cost to Cleaning	Cost to Option
Boiler Feed Water	Supply to unit	$ 0.07	m3	2009	$ 0.07	Chemical Engineering: Process Design and Economics A Practical Guide	$ 23,040	$ 2,292	$ -
Slag	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ 24,195	$ -	$ -
Ash	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ 67,614	$ -
Electricity	For rotating equipment	$ 0.09	kWh	2009	$ 0.09	Source: Energy Information Administration, April 2010	$ 6,233,485	$ 28,382	$ -
LP Steam	Supply to Unit	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ -	$ 104,956
Operations	Supplies and lab expenses	0.6% of capital	n/a	2009	0.6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 522,855	$ 31,472	$ 1,133,494
Maintenance	Salaries, parts, tools,	6% of capital	n/a	2009	6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 5,228,550	$ 314,722	$ 11,334,945
Total							$ 12,032,124	$ 444,484	$ 12,573,395

Operating Cost: Oxygen Gasification to H2

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Cost to Reactor	Cost to Cleaning	Cost to Option
Boiler Feed Water	Supply to unit	$ 0.07	m3	2009	$ 0.07	Chemical Engineering: Process Design and Economics A Practical Guide	$ 11,904	$ 164	$ 4,023
Slag	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ 18,968	$ -	$ -
Ash	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ 66,488	$ -
Oxygen	Supply to unit	$ 101.64	t	2004	$ 106.09	SLS Partners submission to the Federal Trade Commission, July 2004	$ 13,633,465	$ -	$ -
Electricity	For rotating equipment	$ 0.09	kWh	2009	$ 0.09	Source: Energy Information Administration, April 2010	$ 18,449	$ 7,096	$ -
LP Steam	Supply to Unit	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ -	$ 104,956
HP Steam	Supply to Unit	$ 0.0142	kg	2009	$ 0.0142	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ -	$ 1,907,372
Operations	Supplies and lab expenses	0.6% of capital	n/a	2009	0.6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 423,369	$ 13,175	$ 294,151
Maintenance	Salaries, parts, tools,	6% of capital	n/a	2009	6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 4,233,687	$ 131,755	$ 2,941,507
Total							$ 18,339,841	$ 218,678	$ 5,252,009

Operating Cost: Steam Gasification to Methanol

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Cost to Reactor	Cost to Cleaning	Cost to Option
Boiler Feed Water	Supply to unit	$ 0.07	m3	2009	$ 0.07	Chemical Engineering: Process Design and Economics A Practical Guide	$ 15,115	$ 2,295	$ 9,452
Natural Gas	Supply to unit to meet heat	$ 3.50	GJ	2009	$ 3.50	Source: Toronto Market Exchange, http://www.ngx.com/natgas.html, April 2010	$ 10,728,547	$ -	$ -
Wastewater	Treatment of effluent leaving	$ 0.116	m3	2009	$ 0.116	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ -	$ 7,583
Slag	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ 41,613	$ -	$ -
Ash	Disposal of solid waste	$ 0.0022	kg	2009	$ 0.0022	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ 63,550	$ -
Electricity	For rotating equipment	$ 0.09	kWh	2009	$ 0.09	Source: Energy Information Administration, April 2010	$ 18,449	$ 7,096	$ -
LP Steam	Supply to Unit	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ -	$ 3,799,398
HP Steam	Supply to Unit	$ 0.0142	kg	2009	$ 0.0142	Chemical Engineering: Process Design and Economics A Practical Guide	$ 1,235,705	$ -	$ -
Refrigerant	Supply to Unit	$ 1.38	GJ	2009	$ 1.38	Chemical Engineering: Process Design and Economics A Practical Guide	$ -	$ -	$ 1,235,102
Operations	Supplies and lab expenses	0.6% of capital	n/a	2009	0.6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 481,514	$ 14,595	$ 34,914
Maintenance	Salaries, parts, tools,	6% of capital	n/a	2009	6% of capital cost	Chemical Engineering: Process Design and Economics A Practical Guide	$ 4,815,135	$ 145,950	$ 349,136
Total							$ 17,320,963	$ 233,486	$ 5,435,585

Operating Cost: Feed Handling System

Item	Description	Reference	Soybeans	Chinese Tallow	Jatropha	Wood
Handling Equipment	All feed handling equipment	Estimated	$ 161,590	$ 778,056	$ 192,030.59	$ 118,260.00
Biomass Producrement	Buy material	Hamerlink and Faaji	$ -	$ -	$ -	$ -
Transportation	Hauling to site	Mahmudi and Flynn	$ -	$ -	$ -	$ 281,913
Total			$ 161,590	$ 778,056	$ 192,031	$ 400,173

Operating Revenue: LP Air Gasification to Fuel Gas

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Value to Reactor	Value to Cleaning	Value to Option
Fuel Gas	Produced	3.15	GJ	2009	3.15	Chevron, 2009	$ -	$ -	$ 14,570,340.97
LP Steam	Produced	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ 4,345,168	$ 10,496	$ 419,823
Total							$ 4,345,168	$ 10,496	$ 14,990,164

Operating Revenue: LP Air Gasifiation to Electricity

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Value to Reactor	Value to Cleaning	Value to Option
Electricity	Produced	0.09	kWh	2009	0.09	Source: Energy Information Administration, April 2010	$ -	$ -	$ 31,222,159.23
LP Steam	Produced	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ 4,345,168	$ 10,496	$ 419,823
Total							$ 4,345,168	$ 10,496	$ 31,641,982

Operating Revenue: HP Air Gasification to Electricity and H2

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Value to Reactor	Value to Cleaning	Value to Option
Hydrogen	Produced	$ 2.00	kg	2009	$ 2.00	Source: Toronto Market Exchange, http://www.ngx.com/natgas.html, April 2010	$ -	$ -	$ 14,202,874
Electricity	PRoduced	$ 0.09	kWh	2009	$ 0.09	Source: Energy Information Administration, April 2010	$ -	$ -	$ 24,979,843
LP Steam	Produced	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ 4,219,221	$ 10,496	$ -
Energy Source	Intercoolers	$ 3.5000	GJ	2009	$ 3.5000	Natural Gas Price	$ 983,450	$ -	$ -
Total							$ 983,450	$ 10,496	$ 39,182,717

Operating Revenue: Oxygen Gasification to H2

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Value to Reactor	Value to Cleaning	Value to Option
Hydrogen	Produced	$ 2.00	kg	2009	$ 2.00	Chevron, 2009	$ -	$ -	$ 32,518,050
LP Steam	Produced	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ 2,205,120	$ 30,437	$ 2,298,531
Total							$ 2,205,120	$ 30,437	$ 34,816,581

Operating Revenue: Steam Gasification to Methanol

Material	Description	Original Cost	Unit	Cost Date	2009 Cost	Reference	Value to Reactor	Value to Cleaning	Value to Option
Methanol	Produced	$ 133	t	2009	$ 133	Source: Methanex Market Update	$ -	$ -	$ 22,388,029
Fuel Gas	Produced	$ 3.15	GJ	2009	$ 3.15	Source: Toronto Market Exchange, http://www.ngx.com/natgas.html, April 2010	$ -	$ -	$ 13,067,763
LP Steam	Produced	$ 0.0133	kg	2009	$ 0.0133	Chemical Engineering: Process Design and Economics A Practical Guide	$ 2,764,535	$ 16,793	$ 1,416,903
Total							$ 2,764,535	$ 16,793	$ 36,872,694

Note: there is no operating revenue for the feed handling system for this case study.

APPENDIX C

OPTIMIZATION PROGRAM

```
!------Standard Variables------------------------------------------;
n           = 25;        Onstream   = 7884;       IR          = 0.05;
h           = 0.6;       Tax        = 0.2;        Diff        = 0.9;
MW_C        = 12.00;     MW_H       = 1.01;       MW_H2       = 2.02;
MW_O        = 16.00;     MW_O2      = 32.00;      MW_N        = 14.0;
MW_N2       = 28.0;      MW_CO      = 28.0;       MW_CO2      = 44.00;
MW_CH4      = 16.04;     MW_H2O     = 18.02;      MW_Me       = 32.04;
MW_Tar      = 178.24;    MW_Et      = 30.08;      LHV_CO      = 10.9;
LHV_CH4     = 50.1;      LHV_H2     = 120.1;      LHV_Et      = 47.80;
MP          = 0.29;      MW_Air     = MW_O2 * MP + MW_N2 * (1-MP);


!------Sets and Data-------------------------------------------;
SETS:
M/1/:       M_NG, M_H2, M_CO2, M_O2, M_Steam, M_Elect, M_Me;
S/1..4/:    SE, SF, SF_Max,
            SX_C, SX_H, SX_O, SX_N, SX_H2O, SX_NC,
            SP, SP_ref, SC, SC_ref, SC_size;

T/1..4/:    TE, TF, TF_Max, TF_NC,
            TG_C, TG_H, TG_O, TG_N, TG_H2O, TT_O2, TT_H2O,
            TY_CO, TY_CO2, TY_CH4, TY_H2, TY_Tar, TY_Et,
            TD, T_HHV, T_HHVe, T_Steam,
            TP, TP_ref, TS, TC, TC_ref, TC_size,
            TW_Air, TW_O2, TW_H2O,
            UE_Cyclone, UE_Filter, UE_Wash,
            UF, UF_Ash, UF_Tar,
            UG_CO, UG_CO2, UG_CH4, UG_H2, UG_N2, UG_H2O, UG_Slag,
            UG_Tar, UG_Et, UG_H2Oc,
            UX_CO, UX_CO2, UX_CH4, UX_H2, UX_N2, UX_H2O, UX_Slag,
            UX_Tar, UX_Ash, UX_Et,
            UY, UP, UP_ref, UC, UC_ref, UC_size,
            UZ_Slag, UZ_Ash, UZ_Tar;

V/1..5/:    VE, VF, VF_CO, VF_CO2, VF_CH4, VF_H2, VF_H2O, VF_N2,
            VF_Tar, VF_Ash, VF_Et,
            WF, WF_COe, WF_CO2, WF_CH4e, WF_H2e, WF_H2, WF_H2O,
            WF_Steam,
            WF_N2, WF_Tar, WF_Ash, WF_Me, WF_COf, WF_CH4f, WF_H2f,
            WF_Ete, WF_Etf,
            VC, VC_ref, VC_size, VP, VP_ref, VS,
            VY_H2, VY_PSA, VY_Me, VY_Ds;

Fst(S,T):   FstF, FstF_C, FstF_H, FstF_O, FstF_N, FstF_NC, FstF_H

Ftv(T,V):   FtvF, FtvF_CO, FtvF_CO2, FtvF_CH4, FtvF_H2,
            FtvF_H2O, FtvF_N2, FtvF_Ash, FtvF_Tar, Ftv_Et; ENDSETS
```

```
DATA:
M_H2          = @FILE('Data.ldt');    M_CO2         = @FILE('Data.ldt');
M_O2          = @FILE('Data.ldt');    M_Steam       = @FILE('Data.ldt');
M_Elect       = @FILE('Data.ldt');    M_Me          = @FILE('Data.ldt');
M_NG          = @FILE('Data.ldt');    SF_Max        = @FILE('Data.ldt');
SX_C          = @FILE('Data.ldt');    SX_H          = @FILE('Data.ldt');
SX_O          = @FILE('Data.ldt');    SX_N          = @FILE('Data.ldt');
SX_NC         = @FILE('Data.ldt');    SX_H2O        = @FILE('Data.ldt');
SP_ref        = @FILE('Data.ldt');    SC_ref        = @FILE('Data.ldt');
SC_size       = @FILE('Data.ldt');    TF_Max        = @FILE('Data.ldt');
TY_CO         = @FILE('Data.ldt');    TY_CO2        = @FILE('Data.ldt');
TY_H2         = @FILE('Data.ldt');    TY_CH4        = @FILE('Data.ldt');
TY_Tar        = @FILE('Data.ldt');    TY_Et         = @FILE('Data.ldt');
TT_O2         = @FILE('Data.ldt');    TT_H2O        = @FILE('Data.ldt');
TP_ref        = @FILE('Data.ldt');    TC_ref        = @FILE('Data.ldt');
TC_size       = @FILE('Data.ldt');    T_HHVe        = @FILE('Data.ldt');
UY            = @FILE('Data.ldt');    UP_ref        = @FILE('Data.ldt');
UC_ref        = @FILE('Data.ldt');    UC_size       = @FILE('Data.ldt');
VP_ref        = @FILE('Data.ldt');    VC_ref        = @FILE('Data.ldt');
VC_size       = @FILE('Data.ldt');    VY_H2         = @FILE('Data.ldt');
VY_PSA        = @FILE('Data.ldt');    VY_Me         = @FILE('Data.ldt');
VY_Ds         = @FILE('Data.ldt');    ENDDATA


INIT:
TF      = 1000, 1000, 1000, 1000;    SF      = 2000, 2000, 2000, 2000;
AW      = 10000000;                  AR      = 10000;            ENDINIT


!-----Source Model-------------------------------------------------;
@FOR(S(i):@BIN(SE(i)));
@FOR(S(i):SF(i)      = @SUM(Fst(i,j):FstF(i,j)));
@FOR(S(i):SF(i)      < SF_Max(i)*SE(i));
@FOR(S(i):SP(i)      = SP_ref(i) * SF(i));
@FOR(S(i):SC(i)      = SE(i)*0.1 + SC_ref(i)*((SF(i)/SC_size(i))^h));


!-----Mass Flow Model----------------------------------------------;
@FOR(Fst(i,j):FstF(i,j)        >0);
@FOR(Fst(i,j):FstF_C(i,j)      = SX_C(i)*FstF(i,j)/100);
@FOR(Fst(i,j):FstF_H(i,j)      = SX_H(i)*FstF(i,j)/100);
@FOR(Fst(i,j):FstF_O(i,j)      = SX_O(i)*FstF(i,j)/100);
@FOR(Fst(i,j):FstF_N(i,j)      = SX_N(i)*FstF(i,j)/100);
@FOR(Fst(i,j):FstF_H2O(i,j)    = SX_H2O(i)*FstF(i,j)/100);
@FOR(Fst(i,j):FstF_NC(i,j)     = SX_NC(i)*FstF(i,j)/100);


!-----Reactor Inputs ----------------------------------------------;
@FOR(T(j):@BIN(TE(j)));
@FOR(T(j):TF(j)         < TF_max(j)*TE(j));
@FOR(T(j):TF(j)         = @SUM(Fst(i,j):FstF(i,j)));
@FOR(T(j):TG_C(j)       = @SUM(Fst(i,j):FstF_C(i,j))/MW_C*.999);
@FOR(T(j):TG_H(j)       = @SUM(Fst(i,j):FstF_H(i,j))/MW_H*.999);
@FOR(T(j):TG_O(j)       = @SUM(Fst(i,j):FstF_O(i,j))/MW_O*.900);
@FOR(T(j):TG_N(j)       = @SUM(Fst(i,j):FstF_N(i,j))/MW_N*.999);
@FOR(T(j):TG_H2O(j)     = @SUM(Fst(i,j):FstF_H2O(i,j))/MW_H2O*.999);
@FOR(T(j):TF_NC(j)      = @SUM(Fst(i,j):FstF_NC(i,j)));
```

```
!------Reactor Conversion-----------------------------------------;
@FOR(T(j):@FREE(TD(j)));
@FOR(T(j):@FREE(UG_H2Oc(j)));
@FOR(T(j):@BIN(UE_Cyclone(j)));
@FOR(T(j):@BIN(UE_Filter(j)));
@FOR(T(j):@BIN(UE_Wash(j)));


@FOR(T(j):UG_CO(j)      = TG_C(j)*TY_CO(j)/100);
@FOR(T(j):UG_CO2(j)     = TG_C(j)*TY_CO2(j)/100);
@FOR(T(j):UG_CH4(j)     = TG_C(j)*TY_CH4(j)/100);
@FOR(T(j):UG_H2(j)      = (TG_H(j)+2*TG_H2O(j))*TY_H2(j)/2/100);
@FOR(T(j):UG_Tar(j)     = TG_C(j)*TY_Tar(j)/14/100);
@FOR(T(j):UG_Et(j)      = TG_C(j)*TY_Et(j)/2/100);
@FOR(T(j):UG_N2(j)      = TG_N(j)/2+TW_Air(j)*(1-MP)/MW_N2);
@FOR(T(j):UG_H2Oc(j)    = TG_H(j)/2+TG_H2O(j)-UG_H2(j)-2*UG_CH4(j)
                         -5*UG_Tar(j)-3*UG_Et(j));
@FOR(T(j):TD(j)         =(UG_CO(j)+2*UG_CO2(j)+UG_H2Oc(j)-TG_O(j)
                         -TG_H2O(j))*1.05);
@FOR(T(j):TW_O2(j)      = TD(j)*TT_O2(j)*MW_O2/2);
@FOR(T(j):TW_H2O(j)     = TF(j)*TT_H2O(j)/4.340);
@FOR(T(j):TW_Air(j)     = TD(j)*(1-TT_O2(j))*
                         (1-TT_H2O(j))*MW_O2/(2*MP));
@FOR(T(j):UG_Slag(j)    = TG_C(j)-UG_CO(j)-UG_CO2(j)-UG_CH4(j)
                         -UG_Tar(j)*14);
@FOR(T(j):UZ_Slag(j)    = UG_Slag(j)*MW_C);
@FOR(T(j):T_HHV(j)      = ((0.3259*TG_C(j)*MW_C*100 +3.4597*TF(j))));
@FOR(T(j):T_Steam(j)    = T_HHV(j)*T_HHVe(j)/100/(2.690+.313));
@FOR(T(j):UF(j)         = (UG_CO(j)*MW_CO+UG_CO2(j)*MW_CO2
                         +UG_CH4(j)*MW_CH4+UG_H2(j)*MW_H2
                         +UG_H2O(j)*MW_H2O+UF_Tar(j)+UF_Ash(j)
                         +UG_N2(j)*MW_N2+UG_Et(j)*MW_Et));
@FOR(T(j):UG_H2O(j)     = UG_H2Oc(j)+TW_H2O(j)/MW_H2O);
@FOR(T(j):TE(j)         = UE_Cyclone(j)+UE_Filter(j)+UE_Wash(j));
@FOR(T(j):UZ_Ash(j)     = (UY(1)*TF_NC(j)*UE_Cyclone(j)
                         +UY(2)*TF_NC(j)*UE_Filter(j)
                         +UY(3)*TF_NC(j)*UE_Wash(j))/100);
@FOR(T(j):UF_Ash(j)     = TF_NC(j)-UZ_Ash(j));
@FOR(T(j):UZ_Tar(j)     = (UY(1)*UG_Tar(j)*UE_Cyclone(j)
                         +UY(2)*UG_Tar(j)*UE_Filter(j)
                         +UY(3)*UG_Tar(j)*UE_Wash(j))/100*MW_Tar);
@FOR(T(j):UF_Tar(j)     = (UG_Tar(j)*MW_Tar)-UZ_Tar(j));


!-----Reactor and Cleaning Economic Evaluation----------------------;
@FOR(T(j):TP(j)    = TP_ref(j)*TF(j)+M_Steam(1)*TW_H2O(j)
                    +M_O2(1)*TW_O2(j));
@FOR(T(j):TS(j)    = M_Steam(1)*T_Steam(j));
@FOR(T(j):TC(j)    = TE(j)*0.1 + TC_ref(j)*((TF(j)/TC_size(j))^h));
@FOR(T(j):UP(j)    = (UP_ref(1)*UE_Cyclone(j)*UF(j))
                    +(UP_ref(2)*UE_Filter(j)*UF(j))
                    +(UP_ref(3)*UE_Wash(j)*UF(j)))
@FOR(T(j):UC(j)    = UE_Cyclone(j)*UC_ref(1)*((UF(j)/UC_size(1))^h)
                    +UE_Filter(j)*UC_ref(2)*((UF(j)/UC_size(2))^h)
                    +UE_Wash(j)*UC_ref(3)*((UF(j)/UC_size(3))^h));
```

```
!-----Outlet Compositions----------------------------------------;
@FOR(T(j):UX_CO(j)  *UF(j) = UG_CO(j)  *MW_CO);
@FOR(T(j):UX_CO2(j)*UF(j) = UG_CO2(j)*MW_CO2);
@FOR(T(j):UX_CH4(j)*UF(j) = UG_CH4(j)*MW_CH4);
@FOR(T(j):UX_H2(j)  *UF(j) = UG_H2(j)  *MW_H2);
@FOR(T(j):UX_H2O(j)*UF(j) = UG_H2O(j)*MW_H2O);
@FOR(T(j):UX_N2(j)  *UF(j) = UG_N2(j)*MW_N2);
@FOR(T(j):UX_Ash(j)*UF(j) = UF_Ash(j));
@FOR(T(j):UX_Tar(j)*UF(j) = UF_Tar(j));
@FOR(T(j):UX_Et(j)  *UF(j) = UG_Et(j)*MW_Et);
@FOR(T(j):UX_CO(j)        < 0.9*TE(j));
@FOR(T(j):UX_H2(j)        < 0.9*TE(j));
@FOR(T(j):UX_CH4(j)       < 0.9*TE(j));
@FOR(T(j):UX_N2(j)        < 0.9*TE(j));
@FOR(T(j):UX_CO2(j)       < 0.9*TE(j));
@FOR(T(j):UX_H2O(j)       < 0.9*TE(j));
@FOR(T(J):UX_Ash(j)       < 0.9*TE(j));
@FOR(T(j):UX_Tar(j)       < 0.9*TE(j));
@FOR(T(j):UX_Et(j)        < 0.9*TE(j));

!-----Second Mass Flow Model----------------------------------;
@FOR(T(j):UF(j)           = @SUM(Ftv(j,k):FtvF(j,k)));
@FOR(Ftv(j,k):FtvF(j,k)   > 0);
@FOR(V(k):VF(k)           = @SUM(Ftv(j,k):FtvF(j,k)));
@FOR(V(k):VF_CO(k)        = @SUM(Ftv(j,k):FtvF(j,k)*UX_CO(j)));
@FOR(V(k):VF_CO2(k)       = @SUM(Ftv(j,k):FtvF(j,k)*UX_CO2(j)));
@FOR(V(k):VF_CH4(k)       = @SUM(Ftv(j,k):FtvF(j,k)*UX_CH4(j)));
@FOR(V(k):VF_H2(k)        = @SUM(Ftv(j,k):FtvF(j,k)*UX_H2(j)));
@FOR(V(k):VF_H2O(k)       = @SUM(Ftv(j,k):FtvF(j,k)*UX_H2O(j)));
@FOR(V(k):VF_N2(k)        = @SUM(Ftv(j,k):FtvF(j,k)*UX_N2(j)));
@FOR(V(k):VF_Ash(k)       = @SUM(Ftv(j,k):FtvF(j,k)*UX_Ash(j)));
@FOR(V(k):VF_Tar(k)       = @SUM(Ftv(j,k):FtvF(j,k)*UX_Tar(j)));
@FOR(V(k):VF_Et(k)        = @SUM(Ftv(j,k):FtvF(j,k)*UX_Et(j)));

!-----Option Evaluation-----------------------------------------;
@FOR(T(j):UE_Wash(j)+VE(3)<=1);
@FOR(T(j):UE_Wash(j)+VE(5)<=1);
@FOR(V(k):@BIN(VE(k)));
@FOR(V(k):VF(k)<250000*VE(k));
@FOR(V(k):WF_Ash(k)=VF_Ash(k));
@FOR(V(k):WF_Tar(k)=VF_Tar(k));

!-----Option One:  Burn Gas for Electricity---------------------;
WF_Me(1)          = 0;
WF_COe(1)         = VF_CO(1);
WF_H2e(1)         = VF_H2(1);
WF_CH4e(1)        = VF_CH4(1);
WF_Ete(1)         = VF_Et(1);
WF_COf(1)         = 0;
WF_H2f(1)         = 0;
WF_CH4f(1)        = 0;
WF_Etf(1)         = 0;
WF_CO2(1)         = VF_CO2(1)+VF_CH4(1)*(MW_CO2/MW_CH4)
                    +VF_CO(1)*(MW_CO2/MW_CO)+2*VF_Et(1)*(MW_CO2/MW_Et);
```

```
WF_H2(1)              = 0;
WF_H2O(1)             = VF_H2O(1);
WF_Steam(1)           = 0;


!-----Option Two: Separate, Burn CO and CH4 for Energy, Sell H2------;
WF_Me(2)              = 0;
WF_COe(2)             = VF_CO(2);
WF_H2e(2)             = VF_H2(2)*((100-VY_PSA(2))/100);
WF_CH4e(2)            = VF_CH4(2);
WF_Ete(2)             = VF_Et(2);
WF_COf(2)             = 0;
WF_H2f(2)             = 0;
WF_CH4f(2)            = 0;
WF_Etf(2)             = 0;
WF_CO2(2)             = VF_CO2(2)+VF_CH4(2)*(MW_CO2/MW_CH4)
                        +VF_CO(2)*(MW_CO2/MW_CO)+2*VF_Et(2)*(MW_CO2/MW_Et);
WF_H2(2)              = VF_H2(2)*(VY_PSA(2)/100);
WF_H2O(2)             = VF_H2O(2);
WF_Steam(2)           = 0;


!-----Option Three: Steam Shift, Separate, Sell H2-------------------;
WF_Me(3)              = 0;
WF_COe(3)             = 0;
WF_H2e(3)             = 0;
WF_CH4e(3)            = 0;
WF_Ete(3)             = 0;
WF_COf(3)             = 0;
WF_H2f(3)             = 0;
WF_CH4f(3)            = 0;
WF_Etf(3)             = 0;
WF_CO2(3)             = VF_CO2(3)+VF_CH4(3)*(MW_CO2/MW_CH4)
                        +VF_CO(3)*(MW_CO2/MW_CO)+2*VF_Et(3)*(MW_CO2/MW_Et);
WF_H2(3)              = (VF_H2(3)+VF_CO(3)*(MW_H2/MW_CO)*
                        (VY_H2(3)/100))*(VY_PSA(3)/100);
WF_H2O(3)             = VF_H2O(3);
WF_Steam(3)           = 0.287*VF(3);
VF_Ash(3)+VF_Tar(3) < 0.01*VF(3);


!----Option Four: Sell Fuel Gas as Natural Gas Replacement-----------;
WF_Me(4)              = 0;
WF_COe(4)             = 0;
WF_H2e(4)             = 0;
WF_CH4e(4)            = 0;
WF_Ete(4)             = 0;
WF_COf(4)             = VF_CO(4);
WF_H2f(4)             = VF_H2(4);
WF_CH4f(4)            = VF_CH4(4);
WF_Etf(4)             = VF_Et(4);
WF_CO2(4)             = 0;
WF_H2(4)              = 0;
WF_H2O(4)             = VF_H2O(4);
WF_Steam(4)           = 0;
```

```
!----Option Five: Methanol-------------------------------------------;
WF_Me(5)            = VF_CO(5)*(MW_Me/MW_CO)
                      *(VY_Me(5)/100)*(VY_Ds(5)/100);
WF_COe(5)           = 0;
WF_H2e(5)           = 0;
WF_CH4e(5)          = 0;
WF_Ete(5)           = 0;
WF_COf(5)           = VF_CO(5)*(100-VY_Me(5))/100;
WF_H2f(5)           = VF_H2(5)-(2*WF_Me(5)*(MW_H2/MW_Me))/(VY_Ds(5)/100);
WF_CH4f(5)          = VF_CH4(5);
WF_Etf(5)           = VF_Et(5);
WF_CO2(5)           = 0;
WF_H2(5)            = 0;
WF_H2O(5)           = VF_H2O(5);
WF_Steam(5)         = 0;
VF_Ash(5)+VF_Tar(5) < 0.01*VF(5);


!------Option Economic Evaluation ------------------------------;
@FOR(V(k):VC(k)         =VE(k)*0.1+VC_ref(k)*
                          (((VF(k)+WF_steam(k))/VC_size(k))^h));
@FOR(V(k):VP(k)         =VF(k)*VP_ref(k)+M_CO2(1)*WF_CO2(k));
@FOR(V(k):VS(k)         =WF_H2(k)*M_H2(1)+WF_Me(k)*M_Me(1)
                          +((WF_COe(k)*LHV_CO+WF_H2e(k)*LHV_H2
                          +WF_CH4e(k)*LHV_CH4+WF_Ete(k)*LHV_Et)
                          *(0.3/3.6))*M_Elect(1)*Diff+((WF_COf(k)*LHV_CO
                          +WF_H2f(k)*LHV_H2+WF_CH4f(k)*LHV_CH4
                          +WF_Etf(k)*LHV_Et)*M_NG(1)*Diff/1000));


!------Economic Evaluation-------------------------------------------;
@FOR(V(k):@FREE(VP(k)));
@FREE(AP); @FREE(AW); @FREE(AT); @FREE(AD);

C   = (@SUM(S(i):SC(i))+@SUM(T(j):TC(j))
       +@SUM(T(j):UC(j))+@SUM(V(k):VC(k)));
AP  = (@SUM(S(i):SP(i))+@SUM(T(j):TP(j))+@SUM(T(j):UP(j))
       +@SUM(V(k):VP(k)))*Onstream/1000 + (0.066*C);
AR  = (@SUM(T(j):TS(j))+@SUM(V(k):VS(k)))*Onstream/1000;
AD  = C*1000000 / n;
AT  = (AR-AP-AD) * Tax;
AC  = C *(IR * ((1 + IR)^n)/(((1+IR)^n)-1))*1000000;
AW  = AR-AP-AC-AT;

max = AW;


!------User Selected Constraints---------------------------------;
!TE(1)=0; !TE(2)=0; !TE(3)=0; !TE(4)=0;
!VE(1)=0; !VE(2)=0; !VE(3)=0; !VE(4)=0; !VE(5)=0;
!@FOR(T(j):UE_Cyclone(j)=0);
!@FOR(T(j):UE_Filter(j)=0);
!@FOR(T(j):UE_Wash(j)=0);
!@SUM(V(k):WF_H2(k))> 1851;
!@SUM(V(k):WF_Me(k))> 4704;
```

VITA

Adam Michael Georgeson was born and grew up in Alberta, Canada. He received a Bachelor of Science degree in chemical engineering from the University of Alberta in 2005 and began work as a Process Control Engineer at the OPTI Canada oil sands upgrader near Fort McMurray. He joined the graduate program at the Department of Chemical Engineering in Texas A&M University in August 2008 and completed his master's degree in December 2010. While at Texas A&M University, he was involved in biofuel plant design and optimization projects with Chevron and other energy companies.

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