Production of High-Hydrogen Content Coal-Derived Liquids

Type of Report: Final Scientific/Technical Report

Reporting Period Start Date: March 30, 2006

Reporting Period End Date: March 30, 2011

Principal Authors:

Stephen P. Bergin, PhD, P.E.

Date Report Issued: June 2011

DOE Award Number: DE-FC26-06NT42449

Submitted: June 2011

VSE Corporation
41150 Technology Park Drive, Suite 101
Sterling Heights, MI 48313

Disclaimer

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

Abstract

The primary goal of this project has been to evaluate and compare the effect of the intrinsic differences between cobalt (Co) and iron (Fe) catalysts for Fischer-Tropsch (FT) synthesis using coal-derived syngas. The many findings from this project are grouped into three primary categories, presented first in this Abstract:

- Crude oil, especially heavy, high-sulfur crude, is no longer the appropriate source for the additional, or marginal, amounts of middle-distillate fuels needed to meet growing US and world demand for diesel and jet fuels
 - Only about 1/3 of the marginal crude oil barrel can be made into diesel and jet fuels
 - The remaining 2/3 contributes further to global surpluses of by-products
 - FT can produce these needed marginal, low-sulfur middle-distillate fuels more efficiently, with less environmental impact, and from abundant US domestic resources
- Cobalt FT catalyst is more efficient, and less expensive overall, than iron FT catalyst
- Mechanisms of cobalt FT catalyst functioning, and poisoning, have been elucidated

Each of these primary findings is then amplified by several secondary findings, and these are presented, and verified in detail, subsequently in this report.

The most effective step the United States can take to begin building toward improved long-term national energy security, and to reduce dependence, over time, on imported crude oil from unfriendly and increasingly unstable areas of the world, is to begin producing additional, or marginal amounts of, middle-distillate-type fuels, such as ultra-low sulfur diesel (ULSD) and jet fuel (not gasoline) from US domestic resources other than petroleum. FT synthesis of these middle distillate fuels offers the advantage of being able to use abundant and affordable US coal and biomass as the primary feedstocks.

Use of the cobalt FT catalyst system has been shown conclusively to be more effective and less expensive than the use of iron FT catalyst with syngas derived from coal, or from coal and biomass combined. This finding is demonstrated in detail for the initial case of a relatively small FT plant of about 2000 barrels per day based upon coal and biomass. The primary feature of such a plant, in the current situation in which no commercial FT plants are operating in the US, is that it requires a relatively modest capital investment, meaning that such a plant could actually be built, operated, and replicated in the near term. This is in contrast to the several-billion dollar investment, and accompanying risk, that would be required for a plant of more than an order of magnitude greater capacity, which has been referred to in the technical literature on fuel production as the capacity required to be considered “commercial-scale.”

The effects of more than ten different potential poisons for cobalt FT catalyst have been studied extensively and in detail using laboratory continuous-stirred tank reactors (CSTRs) and bottled laboratory syngas “spiked” with precisely controlled amounts of the poisons, typically at the levels of 10s or 100s of parts per billion. This data set has been generated and interpreted by world-renowned experts on FT catalysis at the University of Kentucky Center for Applied Energy Research (UK-CAER), and has enabled unprecedented insight regarding the many molecular-scale mechanisms that can play a role in the “poisoning” of cobalt FT catalyst.

Table of Contents

Disclaimer	2
Abstract	3
Table of Contents	5
Executive Summary	7
Introduction	9
Significance of the Research.....	9
Objective.....	11
Background.....	11
Approach	12
1. Initial Planning and Project Management	17
A. University of Kentucky-Center for Applied Energy Research	17
B. EnviRes LLC	17
C. Emerging Fuels Technology	18
D. Smith Management Group	18
E. University of Alaska-Fairbanks.....	19
2. Cobalt Catalyst Sensitivities to Impurities in Syngas	21
A. Studying the Poisoning of Cobalt Catalyst Used for Fischer-Tropsch Synthesis	21
Discussion/Results - Poisoning by Sulfur.....	21
Conclusion.....	37
3. Synthesis Gas Clean-up and Related Treatments	39
A. Clean-up and Water Gas Shift of Coal-Derived Syngas	39
Introduction.....	39
Discussion and Results	40
Conclusions	42
B. Hydrogen Recovery from Coke Oven Gas.....	42
Discussion	42
Conclusion.....	43
4. Real World FT Fuel Production Implications.....	44
A. Comparison of Cobalt and Iron FT Catalyst Systems for Use with Coal-Derived Syngas.....	44
Discussion/Results	44
Conclusion.....	54
B. Evaluation of FT Product-Upgrading Strategies and Equipment	54
Discussion/Results	54
Conclusion.....	56
C. CAER's R&D-Scale Pilot Plant.....	57
Discussion	57
D. CTL Plant Site Evaluation and Economic Feasibility Analysis.....	57
Introduction.....	57
Discussion/Results	58
Conclusion - Regulatory Burden.....	59
Conclusion - Carbon Management	59
Conclusion - Uncertainty.....	60
5. Testing and Demonstrating High Hydrogen Content FT Fuel.....	61
A. Sample Quantities of Coal-Derived Fuels	61
B. Demonstration of Fischer-Tropsch Fuel Produced From Simulated Coal-Derived Syngas in a 6x6 LASSO Military Utility Vehicle.....	62
Discussion/Results	62
Conclusion - User Impressions of the FT Fuel.....	64
C. Diesel Gen-Set and Reformer/Fuel Cell Demonstration	65
Discussion	65

6. Near Term Production Potential of High Hydrogen Content Coal-Derived Fuel	66
Discussion/Results	66
Conclusion.....	67
Figures.....	69
Tables	71
References	72
Acknowledgements.....	73
List of Appendices	74

Executive Summary

The reason that producing high-hydrogen content liquid fuels from coal and biomass is so effective in improving US energy security, and in reducing dependence on imported petroleum, is that additional, or marginal, quantities of FT fuels can directly contribute to US supplies of ULSD and jet fuel, and at the same time directly and immediately offset more than three barrels of crude oil that would otherwise be imported for each barrel of FT fuel so produced. Furthermore, by offsetting over three barrels of imported crude oil, production of FT fuel has less negative environmental impact, in every respect, than producing the same marginal barrel of middle distillate fuel needed to meet demand by running “3 more barrels” of imported crude oil through a petroleum refinery. This is true even for CO₂ emissions, and even in the most extreme case of using coal as the only feedstock (without including any biomass) and in making no attempt to capture CO₂ for application or sequestration. All other domestic feedstocks, and FT plant operating scenarios, will have even lower environmental impacts. This is in stark contrast to any approach that attempts to reduce gasoline consumption in the overall US, because “conserving” gasoline, which is already in the global-surplus by-product category, will not have any beneficial impact on reducing either US crude oil imports or global greenhouse gas emissions.

Because synthesis of FT fuel from US domestic resources is needed now to enhance US energy security, and since its environmental impact is, regardless of the approach taken, less than continuing to produce the marginal amounts of needed middle distillate fuels by running yet more barrels of imported crude oil through petroleum refineries, it is more important than ever to use the most advantageous technology in launching the new FT industry in the US. Analysis by project-partner Emerging Fuels Technology has shown unequivocally that the cobalt FT catalyst system is more efficient and effective, in terms of energy, cost and environmental impact, than the iron FT catalyst system. This has been demonstrated in detail for a small plant size of approximately 2000 barrels per day, which is the size that can make a near-term contribution to US energy security without incurring multi-billion dollar costs and risks.

Project-partner UK-CAER has also used laboratory CSTRs to delve into the functioning of cobalt FT catalyst at the most fundamental level, and provided new insight into the molecular-scale mechanisms by which low levels of more than ten potential “poisons” in the syngas derived from coal, or from coal and biomass, interact with the cobalt FT catalyst and thereby how they can possibly shorten the productive lifetime of cobalt FT catalyst. The project has also evaluated in detail the interrelationships between the clean-up and related preparation of syngas derived from coal and biomass, and the impacts of the syngas preparation strategy on the subsequent use of that syngas in the production of high-hydrogen content coal-derived liquid fuels.

The project has made important contributions in several additional areas directly related to the production and ultimate use of high-hydrogen content coal-derived liquid fuels:

- The basic design for a state of the art syngas clean-up and water-gas-shift system, appropriate for a 2000 B/d FT plant using coal and biomass as feedstocks, has been developed, its cost has been estimated by industry experts, and its performance has been modeled in detail using the most advanced predictive-analysis tools available.
- Under this project, two CSTRs with cobalt FT catalyst were installed at a commercial coal gasification facility that uses the same type of syngas clean-up system as described above, but at larger scale. Operation of one of the CSTRs on the actual cleaned-up, coal-derived syngas for over 3 continuous months showed nearly identical results to the other CSTR operated on bottled laboratory syngas, and thus verified that cobalt FT catalyst can indeed be used successfully with coal-derived syngas.
- The technical and economic feasibility of extracting additional coal-derived hydrogen for FT synthesis from coke-oven gas has been evaluated, based upon the close proximity of an operating coke-oven to the originally planned site for a prototype coal-to-liquid (CTL) plant projected to be built (before the economic recession and accompanying credit crunch in 2008) by industrial project-partner (and project cost-share provider) EnviRes, LLC.
- More than ten commercially available hydrocracking catalysts (used now in petroleum refining) have been evaluated by UK-CAER for their potential applicability to hydrocracking the heaviest product fraction obtained from FT synthesis, FT wax.
- The strategy, schedule and means to identify all the responsible jurisdictions, for obtaining the full slate of necessary permits needed to plan, design, build and operate a CTL plant, have been researched and described in detail by project-partner Smith Management Group, an Environmental-Legal firm with extensive hands-on experience in assisting industrial clients to obtain such permits.
- FT diesel fuel produced from simulated coal-derived syngas was obtained from Rentech, Inc., and used to demonstrate the military utility of FT fuel in a diesel all-terrain vehicle developed for the US Special Operations Command, at an Air Force Air Show attended by hundreds of high-ranking visiting military personnel and about 200,000 members of the public.
- The managers of industrial facilities on Cook Inlet in Alaska, including both a nitrogen fertilizer plant and an FT pilot-plant that had operated previously on increasingly scarce and expensive local supplies of natural gas, were contacted by the project team in 2008 with regard to the possibility of converting their feedstock to coal, and thus making possible the near-term production of significant quantities (hundreds of B/d) of FT jet fuel produced from domestic coal for then-ongoing evaluations by the US military. A report was compiled in 2008 describing the possibilities, and difficulties, of using these now-mothballed facilities in the manner described.

Introduction

Significance of the Research

The development of domestic sources of liquid transportation fuels is the most important element of US national energy security policy, and FT synthesis is the most promising method for accomplishing the goal of increasing diversity of supply. One of the main reasons that FT synthesis may also be the most effective way to reduce dependency on foreign oil is that FT's products are predominantly middle-distillate fuels, basically low-sulfur, high-quality diesel fuel and jet fuel. Demand for diesel and jet fuel directly determines the amount of imported crude oil needed to meet the nation's transportation fuel demands today. This reality is not widely recognized nor reported, despite the fact that the extent of foreign oil dependency in the U.S. is a direct consequence of the demand for middle distillates, not the demand for gasoline.

Contrary to popular belief, the amount of imported crude oil needed to meet U.S. transportation fuel demand, since about 2005, is no longer determined by the amount of gasoline consumed, but instead by the amount of middle-distillates consumed. Middle distillate fuel production could be enhanced if US refineries used a higher percentage of low-sulfur crude oils. However, low-sulfur crudes are more expensive and in much shorter supply than the heavier, higher sulfur crudes. Although the heavier, higher-sulfur crudes can be processed by many US refineries quite effectively into gasoline, they yield much less middle distillate of acceptable quality. In fact, gasoline is now actually a global surplus by-product that is unavoidably co-produced in quantities greater than needed to meet its demand.

The reason that additional gasoline production is unavoidable when producing additional, or marginal, middle distillate from petroleum is that there is a very limited range of flexibility in varying the yield of fuel products from a petroleum refinery, especially with respect to increasing the yield of middle distillates (Reference 1 describes the technical and economic reasons that refinery yield-flexibility is so limited). A barrel of crude oil can produce no more than about a 1/3 of a barrel of highly valued, and therefore profitable, middle distillates. But about 1/2 a barrel of gasoline must also be produced, and even though it is already in the surplus category, the gasoline must be sold at a price commensurate with the cost of the original crude oil, or the economics of refining would become simply unworkable in the US. The reason is that virtually all the remaining, or "bottom," 15-20% of the crude oil barrel is made into low-value bulk-products such as residual fuel oil, petroleum coke, asphalt, etc., that while useful for some purpose, must essentially be "disposed of" at a net loss, in terms of their volume compared to the original crude oil.

Each additional, or marginal, barrel of middle-distillate produced from petroleum to meet its demand will unavoidably result in about an additional barrel and a half of surplus gasoline on the global market, and proportional amounts of other low-value products

which customers must be induced, with low prices, to accept, or the refinery cannot continue to operate.

As an example, the European Union (who had much earlier moved toward maximizing its middle-distillate refinery production) is now faced with having to export far more excess gasoline than the US and other developed countries in our hemisphere can any longer absorb. It should be noted that relatively cheap surplus gasoline from Europe has already put several of the older, less-efficient, gasoline-oriented refineries on the US east coast out of business. European refiners do not make a profit on their excess exported gasoline; in fact they are often accused of “dumping” their excess exported gasoline at below-cost prices (Reference 2). As US demand for gasoline has been decreasing, European refiners must now literally “push” increasing amounts of their excess gasoline to Africa, the Middle East, and even to Asia, at lower and lower prices further down the trade routes, just to “get rid of it.” This increasing availability of low-cost gasoline is actually artificially encouraging the growth of long-term, indeed permanent, demand for gasoline in several relatively poor countries which have populations that are among the largest on earth (Reference 3).

European refiners continue to produce increasing volumes of excess gasoline, on which they are losing more and more money, and for which far-away customers must be “bought” with fire-sale pricing, because Europe, like the rest of the world, needs more and more middle distillate fuels. But despite their overproduction of gasoline, they still can’t make enough diesel and jet fuel to meet their European demand. The source for most of Europe’s required additional imported middle distillate fuel is Russia and the former Soviet Union.

Our Western European allies have greatly appreciated the opportunity they have had recently, in the wake of the recent recession, to purchase some of the additional middle distillate fuels they need from US refiners, since US demand for middle distillates has been reduced by the recession. This also allowed US refiners to maintain operations, despite reduced demand, of the \$8 to 10 billion worth of upgrades and new equipment that they have installed within the last few years to meet the 15 ppm maximum sulfur requirement for on-road diesel fuel in the US.

The increased costs associated with Europe’s imbalance, namely production of too much gasoline but not enough middle distillate, are ultimately borne by European consumers of middle distillate fuels, not by the oil industry. European consumers have effectively subsidized US consumer’s purchases of gasoline for at least the past decade. To the extent that the US expends resources in an attempt to aggressively “conserve” gasoline, the net effect will be that we in the US will join European consumers in subsidizing the purchase of the same gasoline, by consumers in other countries “further down” the trade routes.

Any overall US effort focused solely on conserving gasoline, which is in the global-surplus supply-category, will have no impact on reducing US dependence on foreign sources of crude oil, or on reducing global greenhouse gas emissions. However, by

producing middle-distillates, especially Ultra-Low Sulfur Diesel (ULSD) and jet fuel from domestic resources other than petroleum, the U.S. can curb its need for foreign oil and effectively meet its growing demand for these fuels. In fact, each barrel of FT fuel can directly and immediately offset, or “back-out,” over three barrels of imported crude oil. Only after significant quantities, at least several hundred thousand barrels per day, of imported crude oil have been backed-out by increased domestic production of FT middle-distillates, will aggressive conservation of gasoline in the overall US begin to make sense.

A barrel of domestically produced FT fuel would also eliminate the emissions and other environmental impacts that would otherwise be attributable to the production, transportation, refining, distribution, and combustion of over three barrels of imported crude oil. Obviously, the production of FT fuel from US domestic resources has some environmental impacts. However, because each barrel of FT middle distillate directly and immediately backs out over three barrels of imported crude oil, all the environmental impacts, including even the CO₂ production attributable to the most extreme case of producing FT fuel from coal without carbon capture and application or sequestration, can actually be lower than using additional crude oil as the source for an additional, or marginal, barrel of needed diesel or jet fuel. Although other alternative sources of middle-distillate fuels may offer similar benefits, the domestic resource base that can be used for FT synthesis is so large and varied that it has the long-term potential to produce fuel quantities comparable to those currently obtained from petroleum.

The potential for FT synthesis to produce middle-distillate fuels from abundant domestic resources, and thereby directly reduce the demand for imported crude oil in a “leveraged” ratio of 3:1, provides the impetus and justification for accelerated development and deployment of this technology. The results of this study provide clear insight and guidance for determining the most efficient type of FT synthesis catalyst to use with coal, or with combined coal and biomass feedstocks.

Objective

The objective of this project was to evaluate and compare intrinsic differences between cobalt (Co) and iron (Fe) catalysts for Fischer-Tropsch (FT) synthesis utilizing coal-derived synthesis gas. Specific parameters to be evaluated include: the effect of contaminants contained in coal-derived syngas on catalyst activity and lifetime, the required H₂/CO ratios for efficient conversion, the required level of syngas cleanup to maintain catalyst reactivity and raw product upgrading requirements to produce high-hydrogen content coal-derived liquids.

Background

The production of hydrocarbons by the hydro-polymerization of carbon monoxide was discovered by Fischer and Tropsch in 1923 and was practiced commercially in Germany

from 1933-1945 (Reference 4). Interestingly, that first commercial FT process employed cobalt based catalysts in fixed bed reactors using coal derived synthesis gas. More recently, the rapid rise in world-wide petroleum prices, coupled with high taxes per barrel, has caused an increase in interest in production of FT products from coal derived synthesis gas. More recently, iron based FT catalysts have often been used for coal derived synthesis gas applications. While iron catalysts are often preferred for high temperature FT processes, targeting chemical feedstocks and high octane gasoline, the choice of optimum catalyst type for low temperature FT synthesis, targeting middle distillates and other paraffinic products from coal derived syngas is not straightforward. Both cobalt and iron have significant advantages and disadvantages. This report shall address and quantify those advantages and disadvantages.

Since the mid-1950's, FT plants that have used coal as their feedstock have virtually all used iron catalyst systems for FT synthesis. When coal is gasified, it produces syngas that has much lower hydrogen content than syngas obtained by reforming other hydrocarbons. For example, methane typically produces syngas with a H_2/CO ratio of ~2, or more than double that typically obtained with coal.

Fe FT catalyst systems are generally considered to be much more tolerant of syngas with coal's low H_2/CO ratios because the iron system can catalyze a significant amount of water-gas shift, and thus increase the effective H_2/CO ratio at essentially the same time as it catalyzes FT synthesis. This means that the raw FT products synthesized from coal-derived syngas using iron catalysts need not be as deficient in hydrogen as might be implied by the initial H_2/CO ratio of the syngas.

Co FT catalyst systems need their incoming syngas to have a H_2/CO ratio of about two to produce the highest quality products. If coal is the feedstock, the H_2/CO ratio of the syngas must be increased for a cobalt catalyst system. This adds cost and complexity, but the cobalt catalyst system may provide other advantages that more than compensate.

This project combines and evaluates real-world, in-depth studies of all the primary or core elements required for the integration of the full-range of FT synthesis technologies (with both Co and Fe catalyst systems) using coal-derived syngas. Included are detailed comparisons between cobalt and iron catalysts, evaluations of the catalysts' sensitivity to various contaminants and poisons, assessments of possible gas cleanup scenarios, and real-world estimates of costs, environmental impacts, and trade-offs involved in producing commercial quantities of coal-derived FT liquid fuels.

Approach

The project consists of a single phase effort focused on the production and testing of coal-derived FT liquids as a transportation fuel (primarily as a diesel fuel).

VSE/ICRC has served as the prime contractor and provided administrative and management support as well as technical input and analysis to the project. The project

has included two major subcontractors: EnviRes LLC and the University of Kentucky CAER. EnviRes has provided gasification and gas cleanup support and the University of Kentucky has conducted the catalyst development and Fischer-Tropsch conversion research. This work has evaluated both cobalt and iron based catalyst compositions. The project has focused on:

- Technical analyses and process data on the following core elements required for the integration of FT synthesis technology with coal-derived syngas:
 - Sensitivity of cobalt FT catalysts to the types and amounts of impurities in coal derived syngas,
 - Analysis of impurity removal or syngas clean-up, to include identification of impurities and their expected concentration ranges, and evaluation of removal strategies and the equipment required, including economics of removal,
 - Technical analysis and economic evaluation of CO-water gas shift and subsequent CO₂ removal, including a study of the types and characteristics of process equipment required,
 - Evaluation of coke-oven gas clean-up strategy and equipment, and the cost-effectiveness of such equipment, as an additional or alternative (to water-gas shift) source of hydrogen to adjust the ratio of H₂/CO to the required value for the type of FT catalyst system to be used,
 - Evaluation of the pros and cons of both cobalt and iron FT catalyst systems, and
 - Evaluation of differences in FT product-upgrading strategies and the equipment required – based on commercial catalysts and processes.
- Initial assessments of the plant-siting constraints, carbon-capture strategies and related environmental issues that are virtually certain to accompany the planning, siting, design, construction, and operation of any U.S. domestic commercial coal-to-liquid fuel production project.
- Feasibility assessment of the use of existing industrial facilities, including a Research and Development (R&D) scale FT plant, as potential near-term domestic sources of FT transportation and jet fuel.

The analyses and assessments described above were broken into 13 contractual tasks which were then grouped into one of six functional thrust areas as described below:

1. Initial planning and Project Management
2. Cobalt FT catalyst sensitivities to impurities in syngas
3. Synthesis gas clean-up and related treatments
4. Real world FT fuel production implications
5. Testing and demonstrating high hydrogen content FT fuel
6. Near term production potential of High Hydrogen Content Coal-Derived Liquid Fuel

1. Initial Planning and Project Management

VSE/ICRC (Integrated Concepts and Research Corporation) immediately developed a planned work breakdown structure (Research Management Plan-Task 1.0), that concisely addressed the overall project as set forth in the cooperative agreement. The Research Management Plan provided a concise summary of the technical objectives and technical approach for each task and, where appropriate, for each subtask, including schedules, milestones, deliverables, decision points, and anticipated expenditures. The Plan has been used as a guide in the execution of all project activities.

2. Cobalt Catalyst Sensitivities to Impurities in Syngas

One of the major thrusts was the evaluation of the sensitivity of cobalt catalyst systems to real-world types and amounts of impurities that are likely to be present in coal-derived syngas, even after some levels of gas clean-up intensity (Experimental Determination of the Sensitivity of Cobalt Catalyst Systems to Coal-Derived Syngas-Task 2.0). Subtask evaluations were concerned with: cobalt catalyst preparation and baseline comparisons; impacts of H_2S level on poisoning rate; installation and operation of a one-liter CSTR on coal-derived syngas to produce upgraded transportation fuel; impact of CS_2 , NH_3 and AsH_3 on catalyst performance; impact of hydrogen cyanide upon catalyst life; impact of COS on catalyst life; and impact of COS and HCN concentrations on catalyst life.

3. Synthesis Gas Clean-up and Related Treatments

Impurities that exceed the maximum allowable threshold limit (MATL) in the coal-derived synthesis gas must be removed before the synthesis gas enters the FT reactor system. Therefore, the Evaluation of Impurity Removal and/or Syngas Clean-Up (Task 3.0) was conducted. Zinc oxide beds can accomplish final polishing, although other sorbents may be used. It was anticipated that H_2S would be the predominant catalyst poison in the coal-derived synthesis gas, and the process of removing H_2S to an acceptable level in the feed to the FT reactor would most likely lower all other catalyst poisons to below the MATL.

The types and characteristics of process equipment required to perform water gas shift to achieve the required H_2/CO ratio for Fischer-Tropsch synthesis was evaluated (Technical Analysis and Economic Evaluation of CO-Water Gas Shift and Subsequent CO_2 Removal-Task 4.0). Included in this evaluation was an assessment of quenching of the reactor effluent with water as a way to achieve partial conversion of CO to CO_2 (with or without catalytically promoted assistance).

In parallel with the systems analysis in Task 4, the process requirements and economics of a system to compress, clean up and recover hydrogen from approximately 40 million standard cubic feet (scf) per day of coke oven gas to provide supplemental coal-derived hydrogen as a feed to an FT unit was examined (Evaluation of Coke-Oven Gas as a Syngas Supplement-Task 5.0).

4. Real World FT Fuel Production Implications

A comprehensive analysis was conducted comparing Co and Fe FT Catalyst Systems. The effect of contaminants/impurities, differences in product quality, the economic impact of adding any necessary water gas shift and/or CO₂ removal reactors/equipment on the design and the commercial viability of the respective catalyst systems was considered to evaluate the net tradeoff between using cobalt or iron catalyst formulations for the production of high-hydrogen content liquid fuels from coal (Evaluation of Cobalt and Iron FT Catalyst Systems-Task 6.0).

Appropriate laboratory/bench scale experiments (in a fixed bed reactor) were conducted to evaluate upgrading of the raw FT products produced using cobalt and iron catalysts to final fuels (Evaluation of FT Product-Upgrading Strategies and Equipment-Task 7.0). Experiments were conducted using commercial catalysts and processes – and reaction rates (kinetics) and hydrogen consumption for this conversion process were evaluated. In addition, an assessment of the differences in the final high-hydrogen content fuel products produced after upgrading the raw products from FT synthesis, from coal-derived syngas, over cobalt or iron catalysts was conducted.

An existing UK-CAER preliminary design for a one barrel per day (b/d) CTL pilot plant facility that was intended to be capable of utilizing both Co and Fe FT catalysts developed and used by CAER under this project was to have been made available to the public, primarily as an indicator of the breadth of accessible operating conditions needed for fully evaluating both types of FT catalyst systems (Initial Design and Definition of CAER's R&D-Scale Pilot FT Plant-Task 8.0). The initial preliminary design was upgraded and superseded during the timeframe of this project by CAER to the detailed-engineering level, as an additional step toward actual construction of the pilot-scale facility, using State of Kentucky funds. Furthermore, the breadth of operating conditions required for the comprehensive evaluation and comparison of cobalt and iron FT catalyst systems has been covered in detail in the project team's milestone report on Task 6.

Additionally, a comprehensive, real world site evaluation and commercial feasibility analysis was conducted for an industry-scale coal-to-liquids plant (Coal-to-Liquids Plant Site Evaluation and Economic Feasibility Analysis-Task 9). The analysis focused on a multitude of elements covered by the permits required from state and local authorities to build and operate a commercial-scale plant.

5. Testing and Demonstrating High Hydrogen Content FT Fuel

Demonstrations and evaluations of high hydrogen content FT fuel were planned. If actual coal-derived FT fuel could not be obtained affordably, FT fuel produced using simulated coal-derived syngas was to be used instead (Sample Quantities of Coal-Derived FT Fuels-Task 10.0). The FT diesel fuel product obtained under this Task was utilized in the vehicle demonstration (Task 11.0) and planned for potential diesel gen-set and reformer/fuel cell demonstrations (Task 12.0).

A vehicle demonstration was conducted (Vehicle Demonstration-Task 11.0) using the FT fuel acquired under Task 10.0 in a diesel-powered all-terrain vehicle (ATV). The demonstration vehicle was a six-wheel, all-wheel-drive, military ATV previously developed and prototyped by VSE/ICRC for the U.S. Special Operations Command. A primary purpose of the demonstration was to highlight the military utility of coal-derived liquid fuels.

An existing diesel-engine/generator test-bed at the University of Alaska Fairbanks (UAF) was originally planned for evaluating the performance of the high-hydrogen-content coal-derived FT liquid fuel (Diesel Gen-Set and Reformer/Fuel Cell Demonstration-Task 12.0). Efficiency, emissions, and engine vibrations were to be measured and compared to baseline data collected previously by VSE/ICRC and a UAF faculty member on both conventional distillate diesel fuels and FT liquids derived from natural gas. Similarly, the same UAF faculty member was to conduct a direct comparison test using an existing reformer to produce hydrogen from the high-hydrogen-content coal-derived FT liquids produced under this project, for comparison to data obtained previously using the same reformer and FT fuel derived from natural gas. The hydrogen produced by the reformer would then be used again to operate an existing solid oxide fuel cell at UAF to produce electricity as part of the FT fuel evaluation. The departure of the faculty member from UAF during the early timeframe of the project, as well as the directly related decommissioning of the UAF reformer/fuel-cell laboratory facilities, precluded carrying out the originally planned FT fuel demonstrations at UAF.

6. Near Term Production Potential of High Hydrogen Content Coal-Derived Liquid Fuel
A feasibility study was conducted (Evaluation of Near-Term, Domestic FT Fuel Sources-Task 13.0) to evaluate the potential of obtaining commercial quantities of coal-derived FT transportation and jet fuel from U.S. domestic sources. The analysis focused on near term (less than 10 years) sources. In addition, the analysis considered the potential of converting existing industrial facilities in Alaska from natural gas to coal feedstock as a means to produce FT liquid fuel from coal.

This final report shall discuss the details, results, and relevant conclusions of the results of the above thrust areas. In selected areas, Appendices are used to capture detailed experimental methods, technical results, and in-depth analyses of applicable thrust areas, tasks and deliverables.

1. Initial Planning and Project Management

VSE/ ICRC was the recipient of the cooperative agreement with the US Department of Energy (DOE) National Energy Technology Laboratory (NETL). The cooperative agreement was amended on July 10, 2008 and the Project Management Plan was, therefore, modified correspondingly. Project history during the 2006-2008 timeframe is reviewed in Appendix 3. VSE/ICRC was the main point of contact, and provided overall project management and oversight to the project (Task 1), as well as technical input and analysis.

This project was intended to be a team effort (as will be described) with all team members making contributions as necessary to achieve the project objectives. In addition to project management, VSE/ICRC had the primary responsibility for the vehicle demonstration on coal-derived FT fuel (Task 11); the evaluation of near-term, domestic FT fuel sources (Task 13), and for obtaining sample quantities of coal-derived FT fuels from industrial sources (Task 10). VSE/ICRC also worked closely with project partners UK-CAER, EnviRes, and Emerging Fuels Technology (EFT) (see below) on the design, evaluation, and performance-prediction of both the impurity-removal or syngas clean-up system (Task 3), and of the water gas shift and CO₂ removal system (Task 4), as well as the Co and Fe catalyst comparison.

As stated above, VSE/ICRC's project team included primary technical partners UK-CAER and EnviRes LLC. These two team members were located in close proximity to each other in Lexington, Kentucky and were already working together on coal-to-liquid technical issues. The plant site evaluation and economic feasibility analysis (with project partner Smith Management Group, Lexington, KY), included an analysis of carbon dioxide capture and application strategies (Task 9). Other VSE/ICRC team members were the University of Alaska Fairbanks and several exceptionally well-qualified technical consultants in Alaska, Kentucky, and Michigan.

A. University of Kentucky-Center for Applied Energy Research

UK-CAER conducted the FT catalyst development and FT conversion research, including experimental measurements of the effects of a wide range of trace-level impurities (likely to still be present at some level in the coal-derived syngas even after clean-up) on cobalt catalyst performance and degradation (Task 2). This data has helped define the level of syngas clean-up required to maintain acceptable catalyst performance and lifetime. UK-CAER also played a major role in the overall comparison of cobalt and iron FT catalyst systems (Task 6), in the evaluation of FT product upgrading strategies for both cobalt and iron FT catalyst systems (Task 7), and in the initial design and definition of an R&D-Scale Pilot FT Plant that would be capable of using both cobalt and iron FT catalyst systems (Task 8).

B. EnviRes LLC

By leveraging a commercial coal-to-liquids project under development, EnviRes lead the basic design, cost analysis, and performance prediction of state-of-the-art systems both to

clean up coal-derived syngas (Task 3) and to increase hydrogen level in the syngas by means of water-gas shift (Task 4). Furthermore, in conjunction with foregoing Tasks, EnviRes evaluated coke-oven gas, a supplemental source of coal-derived hydrogen, and examined the process requirements and economics of a system to compress, clean up and recover hydrogen from coke-oven gas (Task 5). The real-world results of these design studies and performance-prediction calculations for state-of-the-art systems have been made available to the public to assist in future developments of other coal-to-liquid projects.

C. Emerging Fuels Technology

EFT was founded in 2008 by Ken Agee, who had previously founded Syntroleum Corporation, a publicly traded company that did extensive GTL development between 1990 and 2005. Mr. Agee's previous chemical engineering experience was in crude oil refining and natural gas processing, and he was directly responsible for much of the process and catalyst development at Syntroleum.

EFT's services are an integrated set of research tools available for use in the development of technologies primarily related to synthesis gas chemistry and related processes. These services are generally associated with EFT's core competency in the areas of Fischer-Tropsch and related synthesis gas and hydroprocessing chemistry. EFT offers a broad range of catalyst development and testing, process and reactor modeling, process simulation and optimization, cost estimation, and analytical services, as well as engineering and consulting services focused on novel fixed bed Fischer-Tropsch systems to enable small-scale projects. EFT's expertise was therefore applied in performing the overall comparison of cobalt and iron FT catalyst systems (Task 6).

D. Smith Management Group

Smith Management Group (SMG) is an environmental-management and legal firm headquartered in Lexington, Kentucky. SMG was founded by Scott Smith, former Chief of Staff for Kentucky's Environmental and Public Protection Cabinet, an umbrella organization reporting to the governor, with overall direction and coordination responsibility for several state agencies, each having a different mandate for protecting the public or some aspect of the environment.

Over more than 20 years, SMG has worked primarily for industrial clients on:

- Environmental management and engineering
- Industrial hygiene, health and safety
- Liaison with governmental units and jurisdictions
- Acquisition of required permits
- Related legal issues

SMG has worked for Kentucky and the surrounding states on a frequent basis, every Canadian Province, over 35 states, Mexico and several Central American countries, in addition to Taiwan and Indonesia. SMG therefore applied this broad-based experience to conduct a comprehensive Coal-to-Liquids Plant Site Evaluation and Economic Feasibility Analysis (Task 9).

E. University of Alaska-Fairbanks

The UAF was planned to use a diesel-engine/generator test-bed to evaluate the performance of a small volume of the high-hydrogen content coal-derived FT liquids produced by the project. UAF was also to use existing equipment to reform the high-hydrogen content FT liquids to produce hydrogen for testing in a solid oxide fuel cell (Task 12).

Figure 1. Project Schedule Showing Projected Timelines, Tasks, Milestones and Deliverables.

2. Cobalt Catalyst Sensitivities to Impurities in Syngas

A. Studying the Poisoning of Cobalt Catalyst Used for Fischer-Tropsch Synthesis

To study the poisoning of Fischer-Tropsch (FT) catalysts during synthesis is a demanding task. Fixed bed reactors have been utilized, however, for this type of reactor the top of the catalyst (or preheater solid, if used) may act as a getter for the poison. Slurry phase reactors may be utilized, but here pore filling by the heavy FT products needs to be taken into account. In general, the impact of the potential catalyst poison may be defined by both the reactor used and the reaction conditions. Thus, the following results apply for slurry phase operation of cobalt FT catalyst. The use of the continuous stirred-tank reactor (CSTR) means that all catalyst particles are subjected to the same average catalyst poison concentration and to the same average partial pressure of reactants and products. However, the size of the catalyst particle may impact the results and this was not fully appreciated at the start of this work.

This report is also the project team's response to Action Item 3 from a NETL Peer Review, conducted in Morgantown in October 2010, to "catalog the results of investigations into catalyst impurities." Specifically, this report addresses the following topics that were included in the original mutually agreed upon Statement of Project Objectives (SOP):

1. The impact of impurities examined on the Co catalysts.
2. The demonstration of the effects of varying levels of several known or anticipated contaminants in a coal-derived syngas by the use of laboratory CSTR reactors that simulate slurry bed reactors.

Discussion/Results - Poisoning by Sulfur

A number of runs were made with various levels of hydrogen sulfide used as the source of sulfur. The initial attempt was to obtain a general level of sulfur that would be needed to obtain a measurable rate of deactivation. Measurement of this low-level poisoning is complicated by the reduced "background" conversion due to the normal catalyst aging. To overcome some of the problems encountered with normal catalyst aging, a run was continued until a reasonable steady state conversion level was obtained. For most runs, the potential catalyst poison was not added until this pseudo steady data was reached. In general, the more active the catalyst is, the longer it takes to reach a pseudo-steady state conversion level.

A run was made with a 25%Co-0.5%Pt-alumina catalyst at 215°C, 253 psig (17.2 atm) and $H_2/CO = 2.03$. Five volume percent nitrogen was added to the syngas as an internal standard. Starting with the lowest level of H_2S (61 ppb) it is possible that there was a slight increase in CO conversion and the decline in CO conversion was slower at this level of H_2S than it was prior to adding the poison (Figure 2). After about 250 hours of

adding 61 ppb H_2S , the amount was increased to 242-276 ppb range of H_2S . Again there appears to be an increase in CO conversion after increasing the amount of added sulfur.

The space velocity was reduced to increase the conversion and held at the new flow rate without sulfur addition for about 100 hours. The H_2S level was then increased to 438 ppb, but again the decline in CO conversion was about what would be expected for the normal rate of decline due to the normal catalyst aging in the absence of added sulfur. The conclusion for this run is that a very low level of sulfur (250 ppb or less) may increase the CO conversion level and that a sulfur level up to 438 ppb will not exhibit a poisoning effect that is measurable over a short (a few days) testing period.

Figure 2. Impact of adding low levels of H_2S to the synthesis gas for FT synthesis with a cobalt-alumina catalyst (8.2 g catalyst; GHSV = 2.0 slph/g cat; H_2/CO = 2.03; 215°C; 253 psig).

A separate run, shown in Figure 3, was made initially using about the same level of H_2S , 276 ppb, as was used during the intermediate period of the run shown in Figure 2. Following the period with 276 ppb H_2S addition, the sulfur in the feed was increased to 611 ppb. At this higher sulfur level, the catalyst did show deactivation at a rate of 1.95% CO conversion per day (Figure 3). Another run was made with the H_2S level of 717 ppb and the rate of activity decline for the addition of 717 ppb H_2S is 4.52 % CO conversion per day (Figure 4). Another run was made where 825 ppb of H_2S was added and the rate of decline was 9.21% CO conversion per day (Figure 5).

Figure 3. Catalyst poisoning rate when adding 611 ppb H₂S (8.2 g catalyst; GHSV = 2.0 slph/g cat; H₂/CO = 2.03; 215°C; 253 psig).

Figure 4. Poisoning rate when adding 717 ppb H₂S to a cobalt catalyst (GHSV = 3.76 slph/g cat; H₂/CO = 2.06; T = 215°C; P = 255).

Figure 5. Impact of adding 825 ppb H_2S during FT synthesis with a cobalt catalyst ($P = 253$ psig; $227^\circ C$, $H_2/CO = 2.04$, $GHSV = 8.0$; 5% nitrogen).

Consider in Figure 5 the 72-hour period (from run hour 91 to run hour 163), of 825 ppb H_2S addition, during which the catalyst lost activity at the rate of about 9.21% CO conversion per day. If it is assumed that all of the surface cobalt (10% dispersion) is exposed, then one sulfur atom eliminated six cobalt surface atom's catalytic activity. However, many assumptions go into calculating the number of surface cobalt atoms that are poisoned by one sulfur atom, and the number should not be considered to be an absolute quantitative number.

The run shown in Figure 5 was continued beyond 163 hours, as shown in Figure 6. (In Figure 6 the H_2S level from run hours 91 to 163 is expressed as 1ppm rather than as the more accurate value of 825 ppb H_2S .) After the catalyst had been exposed to the H_2S for 72 hours, the catalyst was then, starting at run hour 163, exposed to syngas that did not contain the sulfur poison for a long period of about 18 days (~432 hours), during which the space velocity was decreased twice to maintain a reasonable conversion rate.

Then at about run hour 600 the catalyst was again exposed to syngas that contained H_2S , this time 0.7 ppm (or more accurately 717 ppb) H_2S . During the initial exposure to the 0.7 ppm H_2S , the catalytic activity declined at a rate of 4.52 % CO conversion per day (this portion of the run is also shown in Figure 4). When the conversion reached a low level, the space velocity was decreased (at about run hour 700) while continuing the flow of syngas containing 0.7 ppm H_2S . At the lower space velocity, the catalyst lost activity at 4.0 % CO conversion per day, essentially the same as obtained at the higher space velocity.

Figure 6. CO conversion with and without addition of H_2S ($T = 215^\circ C$; $P = 250$; $GHSV = 2.59$ lphr/g cat).

In Figure 7 the loss of CO conversion per day is plotted versus the amount of H_2S that was added (the 825 ppb H_2S poisoning run was made at an early time on stream and the measured value was corrected by the rate of activity decline obtained prior to adding sulfur). The activity loss is reasonably defined by a straight line that intercepts the x-axis at about 560 ppb H_2S , indicating that at sulfur levels below about 500 ppb a measurable poisoning rate is not observed for these short poisoning runs. Much longer catalytic runs would be needed to verify this tentative conclusion. However, it is an important result in that it indicates that one could utilize a synthesis gas that contains more sulfur than the 10 ppb that is obtained by the Rectisol cleaning process.

Figure 7. Decline in CO conversion related to the amount of H₂S added to the syngas for a cobalt-alumina catalyst.

An important question is whether the catalytic effect is restored when the addition of sulfur is terminated. Viewing Figure 8, the tentative answer is that at low levels of exposure to H₂S the activity, or at least a part of the activity, is restored after the addition of H₂S is terminated. This figure also illustrates the problem of defining the extent of poisoning and/or recovery when utilizing low levels of the poison. A few days after terminating the addition of H₂S the CO conversion was about what would be anticipated from the slow loss of catalytic activity due to other causes. Again, because of the slow rate of poisoning and recovery, very long run times will be needed to verify this.

Figure 8. Impact of adding 611 ppb H_2S and recovery following termination of addition of H_2S (GHSV = 2 slph/g cat; $227^\circ C$, 253 psig, $H_2/CO = 2.03$).

Other sulfur compounds are potential catalyst poisons - one is COS. The COS addition started at an early time on stream and corrections were not made to account for the normal decline due to catalyst break-in; thus, the poisoning rate is likely higher than it would have been if it was measured at a later time on stream. In the case of COS, the rate of loss of activity was 8.85% CO conversion per day (Figure 9).

Figure 9. Impact of addition of COS on CO conversion during Fischer-Tropsch synthesis with a cobalt-alumina catalyst (GHSV = 3 slph/g cat; $H_2/CO = 2.0$; $220^\circ C$; 280 PSIG).

Similar poisoning runs were made with CS₂, again at an early time on stream. In this case the rate of activity decline was 6.86 % CO conversion per day (Figure 10).

Figure 10. Impact of the addition of CS₂ upon the activity of a cobalt-alumina catalyst
2.04 slph/g cat; H₂/CO = 2.06; T = 220°C; 280 psig).

Based on the two data sets represented by Figures 9 and 10, it appears that COS and CS₂ are similar in their catalyst poisoning properties. These results imply that the sulfur in COS and CS₂ is retained by the catalyst in a different manner from that of H₂S. One way for this to occur would be that the SH bonding to the catalyst eliminates the strong bonding that is available for the S of COS and CS₂ and that the removal of S from COS and CS₂ occurs without the addition of hydrogen. Again, this would be an important observation that requires additional work.

To define whether the initial period of operation without the addition of sulfur impacted the final results, a run was made to reproduce an earlier run except that sulfur would be added from the beginning of the run. This should define whether adding sulfur after the initial break-in period impacted the data and the conclusions reached that were based on this data. The results in Figure 11 show that essentially the same results are obtained whether the sulfur is added during the break-in period or not.

Figure 11. Comparison of two runs with same catalyst and reaction conditions except for one run H_2S was added after 5 hours (O) and another run where sulfur was not added during the early period of the run (small squares).

Discussion/Results - Poisoning by Ammonia

Figure 12. CO conversion impact upon addition of ammonia as NH_4OH (2.48 SL/hr/g cat5, H_2/CO – 1.99, 7.47% nitrogen, $T = 220^\circ C$, 280 psig).

The data for the addition of 620 ppm ammonia indicates that this level of ammonia does not impact the conversion of CO. When the level of ammonia addition was increased to 1240 ppm it appears that there is a rapid drop in CO conversion (3.8% decrease) and then a slow decline of about 0.75% CO conversion per day (Figure 12). However, these measurements were made rather early in the run and the decline during the addition of 1240 ppm ammonia is also consistent with the decline being due to the normal catalyst aging. That this is the case is shown by the data in Figure 13 which was collected at a later time on stream with an even higher concentration of ammonia (1700 ppm) and does not show a change in conversion. Thus, it was concluded that even at a concentration of 1500 ppm (note that this is ppm, not ppb), ammonia does not impact the conversion of CO (Figure 13). The amount of ammonia added greatly exceeds that which will be retained in the reactor and exits the reactor to be collected in the water fraction (Figure 14). Thus, it was concluded that ammonia is not a catalyst poison for the cobalt-alumina catalyst.

Figure 13. Impact of the addition of 1700 ppm ammonia during Fischer-Tropsch synthesis with a Co-alumina catalyst (2.48 SL/hr/g cat5, $H_2/CO = 1.99$, 7.47% nitrogen, $T = 220^\circ C$, 280 psig).

Figure 14. Amount of ammonia in the water collected during the addition of 1700 ppm ammonia (note the feed is based on total syngas flow and the aqueous concentration is based on water collected which is smaller than the syngas feed due to the conversion of CO being about 35%).

Discussion/Results - Poisoning by Nitrate

A run was made in which an aqueous nitric acid solution was added after the Co-Pt – alumina catalyst had attained a stable conversion level (Figure 15). As with the addition of ammonia, the addition of nitrogen in the nitrate form did not impact the catalytic activity of the cobalt-alumina catalyst. The nitrate was converted to ammonia as about 3 ppm ammonia was detected in the gas phase; this is the amount expected since ammonia is soluble in water and most of the ammonia is found in the aqueous phase. Similar results were obtained for another run at a later time on stream (1480 to 1750 hours on stream).

Figure 15. Impact of the addition of an aqueous nitric acid solution (298 ppm NO_3^-) upon the CO Conversion (GHSV – 2.44 slph/g cat; $\text{H}_2/\text{CO} = 1.93$; 10.1% nitrogen, 224°C; P = 280 psig).

Discussion/Results - Poisoning by Ammonium Nitrate

In a run, nitrogen was added as an aqueous solution of ammonium nitrate. From the above results, it was expected that the addition of these forms of nitrogen would not have an impact on CO conversion and this was the case (Figure 16).

Figure 16. Impact of ammonium nitrate addition to a cobalt-alumina catalyst during FT synthesis (GHSV = 4.15; $\text{H}_2/\text{CO} = 2.03$; 227°C; 250 psig).

Discussion/Results - Poisoning with Potassium Nitrate

Since neither ammonia nor the nitrate ion caused catalyst poisoning as indicated by the above results, the impact of potassium nitrate was tested. Here there was a poisoning impact during the initial addition of the salt but at later times during the addition it appeared that the catalyst activity had stabilized at a lower value. The activity declined very rapidly (7.3% CO conversion per day) and then appeared to attain a steady-state activity that was only about 14% of the original activity (Figure 17).

Taking the time to go from the original activity when the KNO_3 addition started and the value it attained at 504 hours (near the steady state lower conversion level), it was calculated that about 0.0017 moles of K had been added. Based upon the amount of catalyst loaded (8.4 g), the cobalt loading (25 wt.%), and a dispersion of 10% (it is likely lower than this) it was calculated that there would be 0.0036 moles of Co on the surface.

On the basis of 10% dispersion, enough K had been added to cover half the surface sites; on the other hand, if the dispersion was only 5% then essentially all of the surface Co sites would be occupied by K. The latter view is more likely. In either case, the data show that potassium is a severe poison for the cobalt-alumina catalyst. Unlike the case for sulfur, there is no indication that any of the lost catalytic activity is recovered when the addition of potassium is terminated.

In another run, the shape of the poisoning curve was similar; the initial conversion of CO of 31.7% dropped during 78 hours to 13.0% CO conversion and then remained constant. In this run only 25% of the surface Co would have been covered by K until the steady state CO conversion was obtained and it then remained stable for an additional 200 hours. In another run in which 10 ppm potassium nitrate (based on total gas feed) was added, the conversion dropped by 13.7% overnight and then continued to drop at 1.32% per day during the next 12 days (Figure 18). Obviously, the poisoning by potassium is very severe but is also very complex. To quantitatively relate the amount of potassium added to the poisoning effect will require a much longer experimental program that utilizes longer poisoning runs and catalyst characterization.

Thus, potassium must be viewed as a permanent poison. The results do not permit us to speculate about the fate of the potassium that was added that was in excess of the amount that could occupy all of the surface cobalt sites. It is reasonable to speculate that the additional K resides on the alumina support but evidence for this is lacking at this time.

Figure 17. Impact of the addition of aqueous potassium nitrate solution to a cobalt-alumina catalyst (8.4 g catalyst, GHSV = 3.00; $H_2/CO = 2.00$; 220°C, 280 psig).

Figure 18. Impact of adding 10 ppm potassium nitrate to the syngas fed to a Co-alumina catalyst (8.2 g catalyst; GHSV = 2.50; $H_2/CO = 2.01$; 220°C; 280 psig).

Discussion/Results - Poisoning by Hydrogen Chloride

An aqueous solution containing various amounts of HCl was mixed with the syngas fed to a cobalt-alumina catalyst. The addition of 50 ppb HCl to the syngas for seven days did not have a measurable impact on the CO conversion. Following this addition, periods of

pure syngas and HCl addition were altered, first adding 200 ppb Cl and the 400 ppb Cl (Figure 19). During the periods of HCl addition, no impact was noted on the CO conversion. Thus, it appears that Cl is not a severe poison for the Co-alumina catalyst. The most obvious reason for this is that the HCl reacts with the alumina support. It appears that it will require months of continuous addition of HCl to obtain a measure of the potential of this compound as a catalyst poison.

Figure 19. Impact of the addition of HCl on the activity of a cobalt-alumina catalyst during FT synthesis (GHSV = 2.4 slph/g cat; 10% nitrogen, 220°C; 280 psig.).

Discussion/Results - Poisoning Impact of Potassium Chloride

Based upon the lack of poisoning by Cl and the rapid poisoning by KNO₃, it was expected that KCl would behave similarly to KNO₃. That was not what was found. The addition of 100 ppm KCl as an aqueous solution led to a slow decline in activity of about 0.9% CO conversion per day for a nine day period (Figure 20). With the addition of this much KNO₃, 90% or more of the CO conversion would have been lost in this time period but in the case of KCl the loss was less than 15%. This suggests that with the presence of chloride the support plays a role in defining the rate that the catalyst loses activity. Again, only with long term testing and with a variation of the support can this issue be resolved.

Figure 20. Impact of KCl on the catalytic activity of a cobalt-alumina catalyst for FT synthesis (GHSV = 2.48 slph/g cat; $H_2/CO = 2.05$, 220°C, 280 psig).

Discussion/Results - Poisoning with Arsine

Arsine is a challenge to work with because of its volatility, flammability and it being poisonous to humans. There is a hint of promotion of the CO conversion activity for low levels of arsine addition and then a period of rapid catalytic activity decline (Figure 21). With continued addition of arsine, the conversion levels-off at a conversion that was about one half that when the addition of arsine was initiated. Thus, it appears that the adsorption of this poison reaches a steady-state level and continuing the addition of arsine does not further alter the catalytic activity. Again, much more work would be needed to verify the promotion effect of very low levels of arsine addition, and of the saturation of the catalyst surface at a given partial pressure of arsine.

Figure 21. Poisonous nature of arsine for a cobalt-alumina catalyst under FT synthesis conditions (GHSV – 3.99; $H_2/CO = 2.01$; 280 psig; 220°C).

Conclusion

The definition of catalyst poisoning turned out to be a much more demanding task than initially thought. It was anticipated that the use of the CSTR would expose the catalyst to the same poison concentration. Other work has shown that this is not as easy as anticipated. Other CAER studies have shown that even in a CSTR, wax may accumulate in the catalyst particle and essentially eliminate some fraction of the core volume of the catalyst from being exposed to the reactant gases. To date, CAER does not have a means of defining what fraction of the catalyst used in these studies may have been impacted by wax accumulation.

The current studies have shown that the impact of sulfur on the catalytic activity of a cobalt-alumina catalyst may depend upon the poison used, the catalyst size and the reactor conditions used.

The studies where low levels of H_2S were added to the reactor indicate that sulfur, at low levels, may be a catalyst promoter. This effect of low levels of sulfur has been indicated by some reports in the literature but still needs to be verified. While this is of great academic interest, it appears that the level of sulfur addition where promotion may occur, if it does, is so low that controlling it to that level and maintaining a constant level of sulfur in the reactor will be a demanding task.

An important finding is that there appears to be a threshold level of sulfur before a poisoning effect is noted. Under the conditions used in these studies, it appears that sulfur levels up to about 500 ppb do not impact the catalytic activity; however, above this level it appears that there is a linear relationship between the level of sulfur in the syngas

and the rate of loss of catalytic activity. This finding, if further verified, could lead to a less expensive process for cleaning the syngas feed for a cobalt catalyst than is used at this time.

It appears that for a given level of sulfur addition, a pseudo-steady state will be attained where the continued addition of sulfur will have no impact on the CO conversion. More importantly, it appears that at least a significant fraction of sulfur adsorbed on the catalyst can be removed by desorption at the reaction conditions, provided the addition of the H₂S is terminated.

It has been demonstrated conclusively that neither ammonia nor nitrate, when added as nitric acid, is a catalyst poison. It appears that the ammonia is eliminated from the reactor in the gas phase and concentrates in the water phase in the sample collector. The nitrate, added as nitric acid, appears to be hydrogenated to produce ammonia which leaves the reactor.

Potassium is a catalyst poison. Moreover, it appears that the anion impacts the poisoning effect of the potassium. At the same concentration level of addition, potassium nitrate is a much more severe poison than potassium chloride is. This suggests that the alumina support is also chemisorbing the chloride (and the potassium), thereby decreasing the fraction of the added poison that impacts the cobalt.

For additional information see References 5 and 6.

Two complete Milestone reports can be seen in Appendix 2 and Appendix 3.

3. Synthesis Gas Clean-up and Related Treatments

A. Clean-up and Water Gas Shift of Coal-Derived Syngas

The technical objective dealing with synthesis gas clean-up and preparation was to evaluate impurity removal and/or cleanup of syngas prior to exposure to catalysts and entry into the FT reactor system, as well as determine what types and characteristics of process equipment are required to perform water gas shift to achieve the required H_2/CO ratio for Fischer-Tropsch synthesis, using a state-of-the-art syngas clean-up system.

The next few pages focus first on activities for Task 3.0 (“Evaluation of impurity removal and/or syngas clean-up”) and Task 4.0 (“Technical analysis and economic evaluation of CO-water gas shift and subsequent CO_2 removal”) of the subject cooperative agreement. EnviRes selected WorleyParsons (WP) to perform integrated process engineering and cost estimation activities for both tasks. EnviRes provided WP with the syngas composition and conditions as a starting point for this work. WP generated process flow diagrams, a mass balance and an energy balance for two cases. The first case provided chemical-grade hydrogen at the battery limits ready for use in a refinery or chemical process. The second case provided syngas suitable and ready for Fischer-Tropsch synthesis with a ratio of hydrogen to carbon monoxide of 2.1 to 1.

Introduction

EnviRes and WorleyParsons generated an optimized gas clean-up and CO-water gas shift design and simulated its performance. This design used conventional process equipment in widespread use throughout the world at the size used in this design or larger. This system should be compatible with virtually any high temperature ($>1371^\circ C$) gasifier and any FT reactor system. The throughput of this system equated to 907.2 mt/d (1,000 t/d) with 20%w of the feed comprised of biomass.

The simulations used in this report indicated that FT catalyst poisoning would be substantially avoided or mitigated with this design. This design would cost little or no more than a design that would be considerably more vulnerable to catalyst poisoning. This design exhibited very low levels of conventional emissions such as SO_x , NO_x , CO and particulates. Nominally 97% of the carbon dioxide produced by the process was captured at a purity of 99.86%v. This design compressed the CO_2 to a pressure compatible with pipeline transport or geologic sequestration. Potential FT catalyst poisons from trace elements occurring in coal and biomass were simulated, studied and found to be of no significant threat. This scale would best be characterized as a large demonstration or a small commercial facility in size. Clearly, larger gas clean-up and CO-water gas shift systems could be built with little risk. Several entrained flow gasifiers have been routinely operated on coal and petroleum coke at several times the throughput of this design. The largest commercial single train FT reactors employed by SASOL require approximately ten times as much throughput as was used in this design.

The utility of this design lies in demonstrating emerging gasifier and FT designs with the

capability of incorporating biomass as part of the feed. The FT reactors used by SASOL, described above, are 10.7 m in diameter. They are capable of producing nominally 3,000 mt/d (20,000 bbl/d) of liquid product. Designing complex pressure vessels of this size and larger is a daunting and expensive challenge. Such difficulties provide a strong incentive to find alternative technologies that can produce greater quantities of product from a smaller reactor vessel. SASOL uses substantially different gasification technology than is economically attractive for most opportunities in the USA.

If FT technology is to become a major factor in the supply of transportation fuels from coal/biomass feeds, it appears likely that not only will better FT technology be required, but better gasification technology optimized for FT production will also be required. Several of these improved gasifier and/or FT designs must be demonstrated at the scale of this design.

Discussion and Results

WP developed a Front End Loading (common industry term) FEL level 2 plant design and economics for the Syngas Treatment Area of a coal/biomass to liquids plant with a nominal feed rate of an 80% w coal 20% w biomass feed input at 907.2 mt/d (1,000 t/d). This generated nominally $1.67 \times 10^6 \text{ Nm}^3/\text{d}$ (62.3 MSCFD) of hot raw syngas. The Syngas Treatment Area had a nominal output of $1.54 \times 10^6 \text{ Nm}^3/\text{d}$ (57.3 MSCFD) of feed suitable and ready for a FT reactor with hydrogen to carbon monoxide ratio of 2.1 to 1. Additionally $0.040 \times 10^6 \text{ Nm}^3/\text{d}$ (1.5 MSCFD) of 99% v hydrogen were produced for hydrotreating and upgrading of the FT product.

Other products included 1145 mt/d (1262 t/d) of 99.64% v CO₂ compressed to 15.2 MPa (2,200 psig), ready for pipeline injection and suitable for sequestration. Also produced were 15 mt/d (16.5 t/d) of Claus sulfur and 24.4 GJ/hr (23.1 MBTU/hr) of fuel gas. More than 90% of the mercury in the raw syngas was captured. The capital cost for the Syngas Treatment Section was \$147,016,000 based on US Gulf Coast construction and 2nd quarter 2009 pricing. Major equipment capital costs are listed in Appendix 5 so that some “what if” studies could be performed as long as a reliable way of sizing or scaling the equipment is employed. Similar changes to the operating costs would also be necessary. WP also ran a second case where as much of the syngas was converted to hydrogen as was practical. This case produced $1.42 \times 10^6 \text{ Nm}^3/\text{d}$ (52.9 MSCFD) of 99.48% v Hydrogen at 9 1.79 MPa (260 psig) from the same quantity of syngas as the FT case. The capital cost for the Syngas Treatment Section was \$168,209,000 based on US Gulf Coast construction and 2nd quarter 2009 prices. The major difference with the hydrogen only case was that it had increased CO-water gas shift reactors and a much larger pressure swing absorption (PSA) unit than did the FT case.

WP used a direct contact two step water quench to cool the raw syngas from 1,500°C (2732°F) to 205°C (401°F). The first step mixed water with the hot raw syngas flowing downward in a vertical pipe inside the quench vessel. The outlet of the pipe was below the water level in the vessel. The second step contacted the effluent gas from the first vessel with additional water in a Venturi scrubber followed by water disengagement in a trayed contactor vessel with a demister pad at the top. This arrangement accomplishes

near total particulate removal while adding 90% of the water vapor required for the shift reactor and cooling the raw syngas to 205° C (401°F). The required inlet temperature for the shift reactor was 240°C (464°F). The heat necessary to get the syngas to the proper temperature was accomplished by heat exchange with the reactor effluent from the first shift reactor in a feed-effluent heat exchanger.

WP bypassed approximately 50% of the syngas being shifted around the first shift reactor, mixed the bypassed syngas with the cooled effluent from the first shift reactor, added steam and fed the mixture to the second shift reactor. WP did this because the syngas contained higher levels of CO than was typical for a shift reactor, thus the temperature rise across the reactor was very high. This arrangement affords an effective use of the shift reactor while minimizing the use of additional steam. For the hydrogen only case a third shift reactor was employed while for the FT case some of the quenched syngas bypassed the shift reactors completely and was subsequently mixed with the shifted syngas to attain the 2.1:1 H₂ to CO ratio. In either case the syngas was cooled to 38 °C (100°F), passed through a knock out drum, reheated to 43°C (110°F) and fed to a mercury removal adsorber. The subcooling and reheat was performed to preclude liquid water in the adsorber that markedly reduces its mercury adsorbing capability.

WP employed a Rectisol unit to remove H₂S and CO₂ from the shifted syngas. While Rectisol incurs significantly higher utility cost due to its lower operating temperature it offers several advantages not the least of which is its ability to remove H₂S and other sulfur compounds to less than 100 ppbv. Rectisol effectively removes COS without the need for hydrolysis. This is approximately the threshold of significant catalyst impairment from sulfur poisoning. Any commercial acid gas removal system for syngas currently in use employs a zinc oxide bed downstream of it to further reduce sulfur in the feed to a FT reactor. The use of Rectisol for acid gas removal reduces the consumption and change out frequency of zinc oxide. The use of Rectisol for acid gas removal also mitigates the consequences of breakthrough in the zinc oxide bed. The consequences of sulfur poisoning for a cobalt FT catalyst are much greater than for an iron FT catalyst due to the much greater cost of the cobalt catalyst.

WP included a booster compressor after acid gas removal to ensure that the syngas would be available to the FT reactor at a sufficiently high pressure in the event that the system pressure drop became too high or that for some other reason the system pressure was too low. WP also included a small PSA unit to produce a small 99% hydrogen stream for upgrading the FT products by wax hydrotreating and perhaps some other hydroprocessing step.

The CO₂ stream recovered by the Rectisol unit had less than 0.15% v total impurities. WP included a compression system to boost the CO₂ to 15.2 MPa (2,200 psig). It was decided not to provide a spare for this compressor assuming that the CO₂ could be vented in the event of nonavailability of the compression system. A thermal oxidizer was not provided for the vent gas. It was expected that the compression system would be on-stream at least 90% of the time. Higher CO₂ compression availability could be achieved by incorporating a spare train for this system and perhaps the thermal oxidizer could be eliminated.

Conclusions

Based on the work performed by WP and EnviRes the following conclusions can be reached:

1. An optimized gas clean up and CO-water gas shift system can be designed using components currently in use throughout the world at the size of this design or larger.
2. FT catalyst poisoning can be substantially avoided with this process design.
3. A high level of sulfur recovery with total sulfur emissions below minor source thresholds was demonstrated.
4. Virtually all CO₂ produced by the process was recovered at 99.86% v purity and compressed to pipeline pressure for subsequent use or sequestration.
5. Catalyst poisons such as COS, HCN, CS₂, Hg and other heavy metals were generated at such low levels that they should pose no threat to a FT catalyst when the current process scheme was used.
6. The process scheme described in this report affords significant flexibility for utilizing virtually any high temperature gasifier or FT reactor system.

The complete report can be seen in Appendix 4.

B. Hydrogen Recovery from Coke Oven Gas

Discussion

In parallel with the systems analysis of the last Tasks 3 and 4, process requirements and economics of a system were examined to compress, clean-up, and recover hydrogen from approximately 40 million standard cubic feet (scf) per day of coke oven gas to provide supplemental coal-derived hydrogen (Task 5) as a feed to an FT unit. The FT unit of a projected CTL plant was to be in close proximity to an operating coke-oven, and was projected to be built (before the economic recession and accompanying credit crunch in 2008) by EnviRes, LLC, the industrial project-partner and project cost-share provider to this project. A design basis was established, including target H₂/CO ratios and a detailed coke oven gas composition. Process and equipment requirements were evaluated to clean up and compress coke oven gas and extract high purity hydrogen as a supplemental feed to a FT reactor. Additionally, a preliminary process design was developed for the system.

Natco UK Limited was commissioned by EnviRes to carry out a feasibility and verification study on the recovery of Hydrogen from Coke Oven Gas (COG) at an existing AK STEEL process plant in Ashland Kentucky. The study reviewed the compression of COG, Pre-purification, Hydrogen pre-treatment, and reported on its findings. Additionally, the submission of Heat Material balances, Process Flow Diagrams (PFD's), Process Instrument Diagrams (P&ID's), Plot Plant and preliminary Datasheets for vessels and rotating machines were outputs of the study.

Also included within the study were budgetary costs for plant engineering, design and equipment supply for Pre-purification/COG conditioning, Pressure Swing Absorption Plant, and Hydrogen Pre-treatment. Indicative prices for the COG compressor were also identified.

Conclusion

This portion of the study concluded that with the Process design and equipment scheduled (in section 8.0 of the report seen in Appendix 6), operation with the defined Coke oven gas feedstock is feasible and practical. The recovery of pure hydrogen saturated with water, from coke oven gas will be 78% giving a total flow of 17mmscfd.

Pre-purification Units were included in the design to prevent large amounts of hydrocarbon and sulphur compounds poisoning the absorbents beds present in the PSA. Due to the large quantities of oxygen present, a Deoxo unit would need to be installed after the PSA to provide pure hydrogen (99.9%) at the battery limits of 375 psig and 100 degrees F.

Complete technical and economic feasibility reports can be seen in Appendix 6.

4. Real World FT Fuel Production Implications

This section of the report is focused on using FT catalysts to produce middle-distillate type fuels from coal- and biomass-derived syngas in the real world. This section includes: Comparison of cobalt and iron FT catalyst systems under the significantly different operating conditions that are optimum for each system; Effects of potential contaminants in the syngas on each FT catalyst system; Differences in product upgrading requirements for these two FT catalyst systems; and the overall feasibility of designing, building and operating a coal and biomass based FT fuel production plant, in terms of both economics and the environmental-permit requirements imposed by real-world governments and jurisdictions.

A. Comparison of Cobalt and Iron FT Catalyst Systems for Use with Coal-Derived Syngas

Discussion/Results

The objective of this task (Task 6) and the fundamental objective of the overall project as well, has been to evaluate and compare the effect of the intrinsic differences between cobalt and iron catalysts for FT synthesis using coal-derived syngas and using a realistic commercial design for a CTL plant. Thus, the ultimate goal is to consider all the relevant factors, and to evaluate the net tradeoff between using cobalt or iron catalyst formulations for the production of high-hydrogen content liquid fuels from coal.

Two additional objectives were added to this overall project as a result of Action Items 1 and 3 (see bullets below) from a NETL Peer Review of this project in October 2010:

- Demonstrate the tie-in between the syngas preparation system and the comparison of cobalt and iron FT catalyst systems, and
- Demonstrate the impact of syngas contaminant CO₂ on Reactor Design, the model FT processes, and system economics.

To demonstrate the tie-in between the syngas preparation system and the comparison of cobalt and iron FT catalyst systems, the plant production capacity considered is approximately 2,000 barrels per day (BPD) of FT liquid products, because this FT plant capacity corresponds to the capacity of the syngas preparation system modeled under Tasks 3 and 4 of this overall project. The syngas required is obtained from the co-gasification of 800 tons per day (t/d) of coal and 200 t/d of biomass (Reference 7). A feedstock mix such as this has many advantages compared to the use of coal as the only feedstock (Reference 8).

This task compared cobalt and iron FT catalyst systems under two conditions, namely with the CO₂ from the initial water-gas shift left in the syngas (as a contaminant), and the case where the CO₂ is removed from the syngas after the initial shift. The latter case

requires a more sophisticated and expensive syngas preparation system, such as the one described in the foregoing section (3) of this report and in Reference 7.

The results of the overall comparison show that the overwhelmingly decisive factor favoring cobalt over iron FT catalyst for this application is the much higher capital cost of the major equipment needed for the iron-based FT portion of the CTL plant, as compared to a cobalt-based FT plant with essentially the same liquid fuel production capacity. The higher capital cost of the iron-based FT plant is attributable primarily to the need for significantly larger FT reactors, and to the need for expensive internal recycle compressors. For cobalt, these compressors are not needed for the case where the CO₂ is not removed from the syngas after the shift reactor, but a smaller compressor (recycle ratio of 0.55) is needed only for the first stage for the case where the CO₂ is removed after the shift reactor.

Other factors are considered as well, but they are shown to have relatively little influence on the overall comparison of cobalt and iron, in view of the large difference in capital cost of the plant. For example, iron FT catalyst, while not cheap, costs less than cobalt FT catalyst. However, more than twice as much iron catalyst is needed for the same production of FT fuels. Furthermore, the productive lifetime of the cobalt catalyst is normally an order of magnitude longer than that of iron, meaning that effective catalyst cost per barrel of product is actually lower for cobalt.

Although iron FT catalyst is able to tolerate a somewhat higher sulfur level in the syngas, this does not necessarily translate into a significant saving in practice. The same expensive initial-stage syngas clean-up system, capable of reducing syngas sulfur level to about 200 parts per billion (ppb) is required for both catalyst types. Although cobalt's required final guard-bed type system for additional sulfur removal (down to ~40 ppb) might not be absolutely required with iron, its added cost is much lower than the initial-stage clean-up system. Furthermore, even with iron FT catalyst, the final guard-bed system can function as a valuable "insurance policy" to protect the catalyst charge in the event of upstream upsets in the plant.

Iron FT catalyst is able to promote the water-gas-shift (WGS) reaction in addition to catalyzing FT synthesis. This means that syngas with a lower H₂/CO ratio can be fed to an iron catalyst FT reactor. However, both the total CO₂ produced by both catalysts per barrel of products is similar, and ultimately the total amount of hydrogen required to produce high-hydrogen content coal-derived finished liquid fuels, post upgrading, is also similar for both catalysts.

While for cobalt the CO₂ is produced during the shift reaction before the cobalt-catalyst FT reactor, for iron a similar total amount of CO₂ is produced; this includes both some amount of the initial WGS ahead of the iron-catalyst FT reactor, and additional WGS shift inside the FT reactor. For iron it is therefore convenient to first evaluate a range of H₂/CO ratios in the syngas feed in order to assess the tradeoff--as the H₂/CO ratio of the feed increases, the FT reactor size decreases, but there is lower overall FT fuel production for a given initial flow rate of raw (i.e., pre-shift) syngas.

In terms of CO₂ management, two scenarios have been evaluated. One scenario assumes that CO₂ from the initial water-gas shift is “left in” the synthesis gas through the FT synthesis process, and that the CO₂ could be removed subsequently from the tail gas of the FT process. The other scenario assumes that CO₂ is removed (using a Rectisol-based syngas preparation system, see Reference 7) from the synthesis gas prior to entering the FT process. The results obtained under these two scenarios constitute the project team’s response to Action Item 3 of the Peer Review referred to previously, namely to demonstrate the impact of syngas contaminant CO₂ on Reactor Design, the model FT processes, and system economics.

1). Intrinsic Difference between Cobalt and Iron Based Fischer-Tropsch Catalysts

The commercially practical iron and cobalt-based catalysts are very different in composition, properties and behavior. This study summarizes those differences that have an impact on the configuration, cost and productivity of a CTL plant.

A summary of these differences is shown in Table 1 below.

Generally, a precipitated iron FT catalyst has a higher activity than a supported iron FT catalyst. The cobalt based catalyst, however, typically has a much higher activity than an iron based one. This difference in activity could be two or threefold in fixed-bed Fischer-Tropsch reactors appropriate for a plant capacity of nominal 2,000 BPD. (The plant production capacity considered in detail is approximately 2,000 BPD of FT liquid products, because this FT plant capacity corresponds to the capacity of the syngas preparation system modeled under Tasks 3 and 4 of this overall project.) Because of heat transfer considerations, however, the cobalt catalyst is not able to apply the whole of its activity in fixed-bed reactors. The iron catalyst, on the other hand, is more often limited by kinetics than by heat transfer.

One area where the iron catalyst has an advantage for producing liquid transportation fuels over a cobalt catalyst is methane (CH₄) selectivity. For iron catalysts, the CH₄ selectivity varies between 4 and 5 % while for cobalt catalysts this selectivity will vary between 6 to 10 %. This represents an advantage of between 2 to 5 % more of the most desired products via an iron-based FT catalyst.

This advantage, however, is lost due to the iron catalyst’s high activity for the water-gas-shift reaction.

At the operating conditions typical of low temperature FT, the WGS equilibrium constant is close to 100. This means that there is no net reverse reaction and the H₂/CO ratio will keep increasing as the CO is converted to CO₂.

On the other hand, the average hydrogen content of the Fischer-Tropsch products is about $\text{CH}_{2.14}$ and the Fischer-Tropsch reaction is:

This means that for the iron catalyst to be able to consume the CO in an optimum manner and avoid making more CO_2 than necessary, the H_2/CO ratio in the outlet should be as low as possible, without promoting deactivation or excessive oxygenates production. This is difficult to achieve because as the reactant's partial pressures decrease with conversion and more H_2O is present, the ratio of the FT/WGS rates decrease due to their respective kinetic behaviors. Therefore, the best H_2/CO ratio practically possible at the outlet of the iron second stage is about 2.5. This means that the iron catalyst is producing more CO_2 than the cobalt based catalyst, which has an H_2/CO ratio at the outlet of the second stage of about 1.3.

Because the best iron catalysts are prepared by the precipitation technique, they are mechanically weaker than the cobalt ones that are typically prepared by impregnating the metal precursor over a strong metal oxide such as alumina or silica. This means that the diameter for the iron catalyst extrudates has to be larger than that for the cobalt catalyst (or it will break during loading or upon deactivation by oxidation). This means that the effectiveness factor for iron catalysts is lower than that for cobalt ones and therefore the volumetric productivity will decrease even more than that due solely to the lower intrinsic activity of iron as compared to cobalt.

The iron catalysts are more prone to oxidation than cobalt catalysts. It has been reported that cobalt catalysts can stand up to 5 to 5.5 atm of $\text{P}_{\text{H}_2\text{O}}$, as compared to less than 1.3 to 1.5 for iron catalysts.

This is a very important difference. It means that at the same operating pressure, and because water is a product of the FT reaction, the iron catalyst is able to accept a much lower per pass conversion than a cobalt catalyst. A plant using an iron catalyst will therefore need both more stages and more reactors to achieve a high CO conversion (say, above 90 %), or the use of expensive compressors for internal recycle. The water is condensed and separated from the recycle stream, therefore allowing a higher per stage conversion.

Again, because the iron catalyst is prepared by the precipitation of, for example, iron nitrate and then mixed with a structural promoter such as silica or alumina, its physical strength is also a function of the degree of deactivation or oxidation of the catalyst. As the catalyst loses its FT activity by oxidation of the iron carbide to magnetite (Fe_3O_4 , active for WGS) this is accompanied by sintering and loss of surface area. Because of this, the deactivation of precipitated iron catalysts by oxidation is irreversible. On the other hand, because the cobalt metal is supported on an oxide such as alumina or silica, the oxidation of the surface of the cobalt crystallites is reversible by a re-reduction process. The regenerability of the cobalt based FT catalyst, together with its slow deactivation, facilitates a 2 to 3 years catalyst life (in the absence of irreversible poisons such as sulfur-based compounds).

The Fischer-Tropsch reaction is a hydrogenation reaction. This means that catalysts with a strong hydrogenation activity (e.g., cobalt catalysts), will have a faster FT reaction rate and will produce more hydrogenated products than catalysts with a lower hydrogenating activity (e.g., iron catalysts).

This also means that the higher the hydrogenation activity the more methane is produced and explains why cobalt methane selectivity is higher than that for iron. Further, for catalysts with a high hydrogenation activity it is easier to hydrogenate the growing hydrocarbon chain on the catalyst surface to paraffin therefore terminating its growth and producing shorter hydrocarbons. This explains why the iron catalyst's products are intrinsically "heavier" (longer carbon chain lengths) than those for cobalt catalysts. Because of its lower hydrogenating activity, the iron catalyst has a higher selectivity towards olefin and oxygenated hydrocarbon products. The treatment of water produced by iron FT catalysts is then more expensive than that for cobalt catalysts, and if hydrogenation of some of the hydrocarbon products is needed to produce high-hydrogen content fuels, more hydrogen will be needed.

Both catalysts react in a similar manner to the poisons present in the syngas feed and will need to have extensive syngas cleaning ahead of the FT reactors. While the iron based catalyst will need to have the sulfur contaminants levels reduced to about 200 ppb, the cobalt catalyst will need an extra step, probably a ZnO bed to lower it further from about 200 to about 40 ppb. The cost for lowering the contaminants in the coal-derived syngas down to about 200 ppb is much more expensive than its further lowering down to 40 ppb.

The reason for the higher maximum amount of contaminants in the iron case is not due to an intrinsic higher resistance to sulfur. There are two main reasons for this:

The precipitated iron catalyst has more iron sites per cc of catalyst (or reactor unit volume) than the supported cobalt catalyst.

The expected life time of the iron catalyst is measured in months while the cobalt is measured in years. At the same contamination levels, the cobalt catalyst will have more time (about 10 times more) than an iron catalyst to absorb contaminants.

The incremental syngas clean-up step for cobalt catalysts is shown in Figure 22. The lowering of sulfur contaminants from about 200 to about 40 ppb can be achieved by means of a guard bed such as the commercially available ZnO option.

Figure 22. Syngas cleaning for iron and cobalt Fischer-Tropsch catalysts

The iron catalyst is less expensive per pound of catalyst. Although we need about twice as much iron catalyst as cobalt catalyst on a mass basis, the initial catalyst cost for an iron catalyst is lower than that for cobalt. Because of the longer lifetime of the cobalt catalyst, the final cost of catalyst per barrel of products is lower for cobalt than for iron, as based upon the experience of Emerging Fuels Technology, Inc. and other industry experts.

Table 1. Comparison between iron and cobalt Fischer-Tropsch catalysts		
Item	Iron	Cobalt
Activity	Lower	Higher
Methane Selectivity	Lower	Higher
Diesel Yield	Similar	Similar
Olefins/Oxygenates in products	Higher	Lower
Catalyst life	Lower (months)	Higher (years)
Regeneration	No	Yes
Catalyst mechanical strength	Lower	Higher
Contaminants tolerance	Higher	Lower
Reactor volume (for same production)	Larger	Smaller
Optimum H ₂ /CO in inlet	~ 1.5 – 1.8	~ 2.1
CO ₂ selectivity in FT reactor	Higher	Lower
Catalyst cost, \$/lb cat	Lower	Higher
Catalyst cost, \$/barrel	Higher	Lower

2). Performance Comparison of Cobalt and Iron FT Catalyst Systems

Before proceeding to perform a comparison of the effect of the intrinsic activities of the Cobalt and Iron based FT catalysts, it was necessary to ascertain whether some of the current claims in the literature about the production superiority of one metal over the other were a function of the operating conditions.

It was determined that the operating conditions where one metal (in this case, iron) had a higher volumetric productivity than the other (cobalt) were outside those suitable for practical application. (Refer to full report on Task 6 in the Appendix 7)

Once this had been demonstrated, a commercial CTL plant was designed and optimized for each catalyst system type, based on the composition and amount of the raw synthesis gas described in the Milestone report for Tasks 3 and 4 of this project issued in February 2010 (Reference 7). This syngas is obtained by the high-temperature gasification (in molten iron) of a total of 1,000 t/d of feedstock, consisting of 80% coal and 20% biomass.

The operating conditions selected for the comparison of cobalt and iron FT catalyst systems were the optimum ones for each specific catalyst.

In response to the NETL Peer Review Action Items from October 2010, and to fully demonstrate both the:

Tie-in between the syngas preparation system and the FT catalyst system comparison, and

The effects of syngas contaminant CO₂ on Reactor Design, the model FT processes, and system economics,

the evaluation was performed for two scenarios: with and without the removal of CO₂ after the shift reactor.

a. Comparison without removal of CO₂ after the shift reactor

For the first case study, the CO₂ was not removed from the syngas after the shift took place (despite the fact that the equipment configuration described in Reference 7 implies otherwise). Leaving the CO₂ in the syngas provides some benefit because the CO₂ behaves as an inert in the Fischer-Tropsch reactor. As an inert, it helps to lower the partial pressure of water and therefore allows for a higher per-pass conversion for both the iron and cobalt based catalysts. This could also benefit the syngas preparation section of the plant by postponing, or possibly eliminating, the CO₂ removal step, depending on the overall configuration and planned operating scenario of the plant with respect to CO₂ management. The excess water not consumed during the shift has to be removed because it would increase the deactivation of both catalysts.

It was found that for the same amount of raw syngas (described in Reference 7), and after performing water gas shift at a level suitable for each case, the results favored a cobalt based FT catalyst. This is shown in Table 2 below. The specific iron case is a compromise selected after a study of nine (9) iron cases, each with a different H₂/CO ratio in the syngas feed (after shift) to the first Fischer-Tropsch stage.

Table 2 shows the plant production for the case where naphtha and diesel are the only products. The plant could, however, also include hydrotreating equipment suitable for the production of jet fuel. The carbon number range for the jet fuel overlaps with both the naphtha and diesel fractions. The wide cut jet fuels are comprised of highly isomerized hydrocarbons with an overall carbon range between C8 to C17 or C18, depending upon the particular specification. This approach will reduce the total amount of naphtha produced.

Table 2. Design of a CTL plant based on iron and cobalt FT catalysts (without CO ₂ removed after shift reactor)				
	Iron		Cobalt	
	1 st Stage	2 nd Stage	1 st Stage	2 nd Stage
FT reactor weight (tons)	635	254	240	158
Total plant FT reactor weight (tons)	889		398	
FT Reactor recycle, mol/mol	1.25	0.4	None	None
Catalyst Volume, ft ³	8200	2500	2380	1585
Total plant catalyst vol (ft ³)	10700		3965	
Plant production, BPD				
naphtha (C5-C9)	490	85	420	180
diesel (C10-C19)	1420	220	1085	465
Total per Stage	1910	305	1505	645
Total plant production (BPD)	2215		2150	
FT Equipment Capital Cost	\$51,550,000		\$21,778,000	

(Refer to full report on Task 6 in Appendix 7)

Notice that for the same amount of raw syngas, a cobalt based CTL plant will produce less barrels per day, the reactor volume will be about 1/3, and the reactors will weigh less than half as much, and the capital cost of the FT equipment will be less than half as much as for a CTL plant designed using an iron based catalyst. In addition, a cobalt based CTL plant will not need internal recycle compressors, if the CO₂ from the shift is left in the syngas. In the case of the iron catalyst, 82 % of the CO reacted in the first stage, 52 % to hydrocarbons and the rest to CO₂. Because of this, the feed to the second stage is even more diluted by the CO₂, affecting negatively the iron kinetics by considerably lowering the FT/WGS reaction ratio and therefore increasing the necessary FT reactor volume.

The ratio of FT reactor tons per bpd in the first iron stage is about 0.33 and in the second stage about 0.83, and the ft³ of catalyst per bpd in the first iron stage is about 4.3 and in the second stage about 8.2.

For cobalt, the ratio of FT reactor tons per bpd in the first stage is about 0.16 and in the second stage about 0.24, while the ft³ of catalyst per bpd in the first cobalt stage is about 1.6 and in the second stage about 2.5.

b. Comparison with removal of CO₂ after the shift reactor

In addition to its beneficial effect of allowing higher per-pass conversion, the diluent CO₂ from the initial shift also increases the total mass and volume of syngas that must be processed, which increases the required FT reactor sizes. Removing the CO₂ after the shift has a cost, but it reduces the size of FT reactors needed for both catalyst types, as well as capturing this CO₂ for subsequent application or sequestration, depending upon the carbon management strategy being used by the plant. Note that this approach can be particularly effective with cobalt FT catalyst, because hardly any additional CO₂ is produced within the cobalt FT reactors.

In this case, the results again favor the cobalt catalyst in terms of capital cost, FT reactor tonnage, catalyst volume and in the much lower recycle needs for the cobalt case. The total C₅+ production, however, favors the iron case by about 140 bpd for the same amount of raw syngas feed.

According to the results shown in Table 3, the reactor tonnage for the iron case is over 2 times that required for the cobalt case; the amount of catalyst is almost three times higher; and the recycle needs are again much higher for iron than for cobalt. For iron to be better than cobalt in this scenario, the difference in the investment cost for the two options would have to be justified by this difference in production.

The difference in capital investment for the iron and cobalt cases (excluding civil/utility and support equipment/structures) is almost \$24 million, or an investment of about \$170,000 per barrel produced per day for the extra 140 bpd. On a per-barrel-per-day basis, this incremental investment would be about an order of magnitude greater than the typical overall equipment cost for an FT plant. This means that in this scenario, where the CO₂ is removed after the shift reactor, the cobalt is again more economic than iron.

Both catalysts will need to have extensive syngas cleaning. While the iron-based catalyst will need to have the sulfur (S) contaminants levels reduced to about 200 ppb, the cobalt catalyst will need an extra step, probably a ZnO bed, to lower it further from about 200 to about 40 ppb. The cost for lowering the contaminants in the coal derived syngas down to about 200 ppb is much more expensive than its further lowering down to 40 ppb.

Table 3. Design of a CTL plant based on iron and cobalt FT catalysts (with CO ₂ removed after the shift reactor)				
	Iron		Cobalt	
	1 st Stage	2 nd Stage	1 st Stage	2 nd Stage
FT reactor weight (tons)	404	205	235	45
Total plant FT reactor weight (tons)	609		280	
FT Reactor recycle, mol/mol	1.6	1.2	0.55	None
Catalyst Volume, ft ³	5321	2752	2337	456
Total plant cat vol (ft ³)	8073		2793	
Plant production, BPD				
naphtha (C5-C9)	466	125	505	93
diesel (C10-C19)	1354	336	1306	239
Total per Stage	1820	461	1811	332
Total plant production (BPD)	2281		2143	
FT Equipment Capital Cost	\$40,580,000		\$16,617,000	

(Refer to full report on Task 6 in Appendix 7)

Conclusion

Therefore, a supported cobalt catalyst is more effective than an iron catalyst for the conversion of coal to high-hydrogen content liquid transportation fuels. This conclusion is based on the higher equipment cost, larger volume of the FT reactors, FT reactor weight, catalyst volume, and increased plant complexity--including recycle compressors (e.g., the iron option as compared to cobalt).

The complete report can be seen in Appendix 7.

B. Evaluation of FT Product-Upgrading Strategies and Equipment

Discussion/Results

The ultimate purpose of this task was to assess the differences, particularly with respect to hydrogen consumption, in upgrading the raw FT products (primarily wax) synthesized

from coal-derived syngas using iron and cobalt FT catalyst systems to produce finished-fuel products. The intent was that this information from Task 7.0 would be an input to Task 6.0, in which the overall commercial viability of the respective catalyst systems has been analyzed to evaluate the net tradeoff between using cobalt or iron catalyst formulations for the production of high-hydrogen content liquid fuels from coal.

The Task 6.0 analysis has already shown that the much higher capital cost of major equipment needed for the iron catalyst system is the overwhelming factor favoring cobalt in the comparison of iron and cobalt using coal-derived syngas, and that other factors, including the relative amounts of hydrogen needed for product upgrading, have relatively little effect on the overall comparison. Finished-fuel products from cobalt FT catalysts do need to consume less hydrogen during product upgrading, but they also require a much higher hydrogen content in the original syngas entering the cobalt FT reactor, resulting in little, if any, net difference between iron and cobalt FT catalyst systems in terms total hydrogen consumption to produce high-hydrogen content coal-derived liquid fuels.

Furthermore, the overall “state of the art” in using commercial hydrocracking catalysts (used commercially in petroleum refining) to upgrade FT wax to finished-fuel products, is not yet particularly well advanced, at least within the non-proprietary, open literature. Since attempting to quantify subtle differences for upgrading the waxes derived from iron and cobalt FT catalysts, respectively, is effectively beyond the current state of technology, several candidate commercial hydrocracking catalysts have been evaluated as to their ability to produce finished-fuel products from FT wax.

UK-CAER work on Task 7.0 in this project and on another project allowed work on conducting hydrocracking studies with CAER wax derived from cobalt and iron FT catalysts. Runs conducted during a two year period are summarized in Table 4. These analyses are ongoing, and the overall results will be published at a later date. In general, the narrow pore zeolites produced more gasoline than is desirable. A definition of the optimum catalyst physical and chemical properties that are needed to provide the maximum diesel and/or jet fuel will require studies utilizing an even wider range of catalysts than shown here. A systematic study needs to be conducted to define the optimum hydrogenation activity and cracking activity that is needed, as well as studies to define the best balance of the hydrogenating and cracking activities for the optimum fuel production.

Catalyst	Catalyst ID	Run ID	Kinetic Conditions
0.5 % Pt/HZSM-5	NASA 20	MAW145 MAW158	250-330 °C, 450-550 psig, H ₂ /Wax = 5-17, LHSV = 1-30 NL/g-cat/h
0.5% Pt/SiO ₂ -Al ₂ O ₃ Davison 135	NASA 21	MAW148	300-340 °C, 400-550 psig, H ₂ /Wax = 15-30, LHSV = 1.0 NL/g-cat/h
2009 3% CoO and 12 % MoO ₃ supported on (nonsilicate) alumina	NASA 22	MAW151	300-340 °C, 500 psig, H ₂ /Wax = 15, LHSV = 1.0 NL/g-cat/h
3.0 % Ni/Davison 135 SiO ₂ -Al ₂ O ₃	NASA 23	MAW155	300-340 °C, 500-700 psig, H ₂ /Wax = 17-30, LHSV = 1.0 NL/g-cat/h
15.0 % Ni/Davison 135 SiO ₂ -Al ₂ O ₃	NASA 24	MAW 186	300-340 °C, 400-550 psig, H ₂ /Wax = 15-30, LHSV = 1.0 NL/g-cat/h
0.5 % Pt/Al ₂ O ₃	NASA 25	MAW 203	300-380 °C, 400-550 psig, H ₂ /Wax = 15-30, LHSV = 1.0 NL/g-cat/h
T2753	Sud-chemie	MAW173	320-390 °C, 600 psig, H ₂ /Wax = 20, LHSV = 1.0 NL/g-cat/h
2010 T2753, PP8304	Sud-chemie	MAW176	300-375 °C, 600-800 psig, H ₂ /Wax = 20, LHSV = 1.0-5.0 NL/g-cat/h
T-2753, PP8323	Sud-chemie	MAW200	320-400 °C, 450 psig, H ₂ /Wax = 20, LHSV = 1.0 NL/g-cat/h
T-2753, PP8313	Sud-chemie	MAW206	320-400 °C, 450-600 psig, H ₂ /Wax = 20, LHSV = 1.0 NL/g-cat/h
T-2753, PP8305	Sud-chemie	MAW208	320-360 °C, 450 psig, H ₂ /Wax = 20, LHSV = 1.0-4.0 NL/g-cat/h

Table 4. Hydrocracking runs to obtain kinetic data and product distributions of the hydrocracked materials.

Conclusion

In summary, the raw FT products, such as wax, with boiling ranges above those of finished transportation fuels, must be cracked to yield finished fuel products in the desired boiling ranges. The results obtained to date show that, in general, the cracked products will be composed of at least 50% branched alkanes, and this improves the octane number of the gasoline fraction, but not to an adequate octane number, so additional processing would be required to produce gasoline. Cracking also improves the low-temperature properties of the diesel without reducing its cetane number significantly.

Some fraction of the raw products from FT synthesis are typically in the liquid range and do not necessarily require hydrocracking, but they may need other forms of upgrading, especially if the goal is to produce transportation fuels (as opposed to other products such as solvents or chemicals, for example). For use as transportation fuels, the straight-run products that fall within the boiling range of the desired fuel will define product quality. For the gasoline-range raw FT products (naphtha), the octane number is much too low and only by isomerization or blending can an adequately high octane number be obtained. The very low aromatic content and essentially zero-level of heteroatoms in these gasoline-range FT products make them an ideal blend stock, provided that their octane number can be increased sufficiently.

The straight chain nature of the diesel and jet fuel range raw FT products make them desirable because of their very high cetane number, but undesirable because of their low density and relatively poor low-temperature properties. Hydroisomerization is usually required to improve low-temperature properties, especially to produce acceptable jet fuel.

C. CAER's R&D-Scale Pilot Plant

Discussion

The objective was to make available to the public an existing preliminary design for a one barrel per day coal-to-liquids pilot plant facility that could utilize the cobalt and/or iron FT catalysts, developed by CAER, as an indicator of the breadth of accessible operating conditions needed for fully evaluating both types of FT catalyst systems.

Based upon hearsay evidence at the time this project was being restructured, the existing preliminary design was assumed to include all components of a functional pilot scale facility – from coal preparation to final product processing. However, the actual existing CAER design did not include any provisions for the use of coal, but rather had been intended from the beginning to be operated on bottled laboratory syngas. Furthermore, the initial preliminary design was upgraded, subsequently superseded, and effectively made obsolete, during the timeframe of this project by CAER as a step toward actual construction of the more comprehensive pilot-scale facility, paid for with State of Kentucky funds. In order to address and provide a thorough treatment of the breadth of operating conditions required for properly evaluating both iron and cobalt FT catalyst systems, this information has been incorporated into the previous section of this report on Task 6.

D. CTL Plant Site Evaluation and Economic Feasibility Analysis

A site evaluation and commercial feasibility analysis was conducted for an industry-scale coal-to-liquids plant (Coal-to-Liquids Plant Site Evaluation and Economic Feasibility Analysis-Task 9.0).

Introduction

Development of a robust CTL fuel industry is critically important in the move towards energy independence. The United States has large deposits of coal which have historically been mined at reasonable costs. In addition, the world-wide move toward control and reduction of greenhouse gases is driving the use of coal for energy production toward newer technologies which are better able to capture and store or sequester CO₂.

Any industrial development must comply with many layers of environmental and governmental regulation and control. The CTL industry will be one of the more heavily regulated industries, due to its potential emissions, especially of greenhouse gases. CTL projects will be further delayed by protests and legal action from groups ranging from NIMBYs (not in my back yard) to those interested in stopping the use of coal.

Development of a road map and analysis of the permitting requirements applicable to the development of a commercial-scale CTL facility should help prospective developers make good decisions about schedules, costs and the economic feasibility of their proposed projects. Understanding the likely time and cost requirements of planning, siting and permitting a CTL facility is critical in any developer's decision matrix.

Discussion/Results

This project has evaluated permitting issues and plant siting for development of a coal-to-liquids plant, resulting in a Permit Handbook which can be used as a guideline for development of a CTL project. The research has used a development site in Greenup County, Kentucky as a model. Therefore, while the Permit Handbook has been prepared to include Kentucky permitting references, each state will have parallel agencies and requirements. This work has provided perspective on the information, cost and time required for preparation of applications and required information, regulatory review and approval.

Work was also performed to identify the impact of the choice of plant site on the permitting requirements for a CTL plant; and the impact of permitting requirements on the design, layout and operation of such a plant. Discussion of the economic impact of the permitting process and environmental compliance on the feasibility of building and operating a commercial plant has been included.

Strategies to manage carbon dioxide were examined, utilizing the specific site in South Shore, Kentucky as a test case. Carbon sequestration was explored in more general terms as well and the current, albeit ever-changing, status of legislation and regulation on carbon management was explored.

The report provides analysis and introduction to several attached tools and reports and includes the following information and tools:

- Explanation of the CTL process and the major equipment and processes likely to be used.
- The resources needed or inputs to the process, and the outputs expected, including product, emissions and wastes.
- Site attributes that are either required or beneficial for a successful CTL project.
- A list of major environmental permits and authorizations and a chart identifying appropriate contacts within each state environmental, energy or economic development agencies for development of a CTL facility.
- A Permit Handbook that can be used as a manual for developers to guide permit preparation and scheduling.
- Analysis of the influences and impacts that each permitting initiative may have on choice of site and design and development of the project.
- The economic impact of environmental compliance on the viability of developing and operating a CTL facility.
- Analysis of geologic sequestration potential at the sample site.

- The current status of legislation and regulation applicable to geologic sequestration. Understandably, information in this section changes rapidly with developments at the US Environmental Protection Agency (EPA) and under state law.
- A report of a policy Workgroup which evaluated legal issues of geologic sequestration as applied to Kentucky.

Conclusion - Regulatory Burden

Any developer preparing to design, site and construct a coal-to-liquid facility should be prepared for a long, protracted effort with regard to developing a comprehensive permit package. Permits and authorizations will be required from dozens of state and federal agencies with applications that are among the most complex possible. Public input will be extensive and likely adversarial for the most part. Opportunities for delay will be found within every regulatory program.

Permitting programs are, for the most part, not integrated and are in constant flux. This increases the burden to the developer who is required to provide the same information repeatedly and in different forms to different agencies. The permitting authorities do not routinely communicate.

As the project used a site in Kentucky for research purposes, it is illustrative to review how Kentucky addresses permitting issue and integration. In 2007, Kentucky's legislature adopted a requirement that energy projects be provided with "one stop shopping" for permitting. Kentucky has made an effort to store permitting information in an electronic database which can be accessed throughout state regulatory offices. However, there has been no change in permitting process. As one example, applicants are required to provide full reports on the potential impact of cultural and historic resources within its application to the State Historic Protection Officer, within its solid waste permit application, within any federal application, within an Environmental Assessment or Environmental Impact Statement. Each of these agencies then conducts a new review of the submitted information. No agency, as a matter of course, accesses the information available in the state database to reduce the volume of material required or the number of times it is reviewed.

Conclusion - Carbon Management

Legal requirements for reduction, limitation, or capture and sequestration of carbon dioxide are still in development. While no federal legislation has been passed mandating the control of such emissions, US EPA has adopted requirements to inventory and report greenhouse gas emissions (GHG) and now is requiring emission reductions for the first few categories of facilities. Litigation over EPA actions may reverse or encourage those developments.

The information provided in this report and its attachments reflects this flux and the changes that have been recognized over the past couple of years. The necessary conclusion is that regulatory requirements, while to some extent still in development, are substantial, time consuming and costly to any coal-to-liquid project. Siting aspects and

technology design will equally drive the permitting requirements. Delays inherent in the permitting process and allowed through public input and litigation will add substantially to the time and cost of permit acquisition.

An important area to watch and encourage will be the alternative use of carbon dioxide. Limitations associated with a project site, or limitations of overall storage space, will require that portions, possibly substantial portions, of the carbon dioxide produced through the use of fossil fuels be captured and used other than through sequestration. The best scenario will be to find ways in which carbon dioxide becomes a valuable commodity, adding to the economic wealth of the country rather than subtracting from it.

Conclusion - Uncertainty

The development of a CTL industry in the United States is hampered by uncertainty. There is no uncertainty about the direction EPA is heading as they have issued regulations for permitting injection wells and regulations for permitting greenhouse gas emissions. There is uncertainty in where the permitting issues will end due to the litigation on these actions.

There is also the uncertainty that comes with any new industry or process. While the science of CTL appears to be well developed, there are no CTL facilities operating in the United States at this time. There are numerous test wells in the country and worldwide through which substantial information has been developed. There are, however, no operating commercial injection wells in the United States at this time.

The complete report can be seen in Appendix 8.

5. Testing and Demonstrating High Hydrogen Content FT Fuel

A. Sample Quantities of Coal-Derived Fuels

The objective of this task was to acquire, from available industrial sources, sample quantities of actual coal-derived FT fuels, produced using either cobalt or iron FT catalysts. The fuel products obtained under this task were to be utilized in the vehicle demonstration and planned for the diesel gen-set and reformer/fuel cell demonstration. Since actual coal-derived FT fuels could not be obtained affordably, FT fuels

produced using simulated coal-derived syngas were acquired from Rentech, Inc. Rentech has a nominal 10 barrel per day capacity pilot FT fuel plant. The pilot-scale Product Demonstration Unit (PDU) produces, and then processes, simulated coal-derived syngas to make finished FT diesel and jet fuels.

Rentech's PDU is located in Commerce City, Colorado at the Rentech Energy Technology Center (RETC). The PDU is believed to be the only fully-integrated synthetic transportation fuels production facility operating in the United States. This facility is designed to produce over 400 gallons per day of ultra-clean synthetic jet fuel, aviation fuel, ultra-low sulfur diesel, and specialty waxes and chemicals and is scalable for greater output.

Rentech designed the PDU FT pilot plant to operate on coal, with the potential future capability to include at least some fraction of biomass as feedstock at a later date. Accordingly, Rentech selected a plant site with railroad access for coal delivery, and procured and installed both coal handling equipment and the comprehensive syngas clean-up capability that would be required for coal-derived syngas. Rentech also procured, installed and attempted to operate a pilot-scale coal gasifier to provide actual coal-derived syngas to the pilot-scale syngas-processing and FT synthesis units downstream.

After installation, and several attempts to work-around gasifier deficiencies and make the plant operational, Rentech chose to obtain a small-scale commercial reformer system to first produce syngas from natural gas, and then, through the use of a membrane system, remove some of the hydrogen from the syngas so that it would more accurately simulate coal-derived syngas. The simulated coal-derived syngas could be processed in the downstream units and thus produce simulated coal-derived diesel and jet fuels for

evaluation. The hydrogen that had been removed from the syngas was recovered and used for upgrading the raw FT product to finished fuels.

Rentech has not given up on the possibility of incorporating the capability of using at least some fraction of coal in the feedstock for the “new” pilot-scale gasifier that is planned to be added to the existing PDU pilot plant in the near future. They have demonstrated that they intend to persevere in their goal of producing FT fuels from both coal and biomass, resources that the US has in abundance. Rentech agreed to provide the quantities of PDU FT fuel required for the demonstrations. The Rentech fuel was subsequently used to complete the Vehicle Demonstration as described below.

B. Demonstration of Fischer-Tropsch Fuel Produced From Simulated Coal-Derived Syngas in a 6x6 LASSO Military Utility Vehicle

Discussion/Results

On August 21-23, 2009, representatives from VSE Corporation’s Alternative Fuels Team participated in a demonstration of FT neat fuel, produced from simulated coal-derived syngas, at Selfridge Air National Guard Base near Mt. Clemens, MI in conjunction with their bi-annual Air Show. Selfridge hosts this public open house to allow the residents of the metro Detroit area a rare opportunity to interact with fighter pilots and their maintenance crews in an effort to demonstrate where their defense dollars are spent each year. Air Show organizers expected an estimated 175,000 people, providing free admission and free parking throughout the event.

Throughout the demonstration, the Rentech simulated coal-derived FT neat fuel was run in an off-road, 6x6, material handling vehicle, which was developed by VSE Corporation for the Special Operations community, called the LASSO® vehicle. LASSO stands for Land and Sea Special Operations, and the 6x6 version is the first in a planned series of vehicles. The three day

demonstration consisted of a one-day, ride and drive, soldier demonstration, and two public demonstration days. Throughout the soldier demonstration day, several active-duty base soldiers took the LASSO[®] vehicle for performance test drives, noting the performance characteristics of the vehicle as well as the fuel.

During the public demonstration days, the LASSO[®] vehicle was demonstrated statically to create auxiliary power supporting off-board booth electronics as well as on test drives throughout the show grounds to showcase the visibly reduced emissions, due to the cleaner burning FT fuel. VSE Corporation team members staffed a booth throughout the two public demonstration days that was located adjacent to the “Technology Hangar” which showcased other new and emerging military technologies including fuel cell powered vehicles, unmanned surveillance robots, and materials exhibits provided by NASA.

The simulated FT diesel fuel (discussed previously and as shown on the hood of the LASSO test vehicle) was produced in Rentech’s pilot-scale, nominal 10 barrel per day Product Demonstration Unit in Denver, CO.

To prepare the LASSO[®] vehicle for the three day demonstration, the fuel tank was emptied of all remaining diesel fuel and one gallon of the FT fuel produced from simulated coal-derived syngas, was used to completely flush the system prior to a full fill up. The fuel filter was also replaced. A one-gallon sample of the FT fuel was taken for future reference. The 14 gallon vehicle fuel tank was then filled with FT fuel and the engine was run for a few minutes at the shop in order to identify, in advance, any compatibility issues that might occur during the weekend.

Conclusion - User Impressions of the FT Fuel

Throughout the three day demonstration, exposure to military personnel with combat experience, as well as to the thousands of public visitors who visited the booth proved invaluable. Soldiers who test drove the vehicle remarked that they noticed a reduced exhaust smell as compared to conventional fuel. They were also impressed by the absence of particulate collecting on the cross frame member next to the vehicle exhaust, which are both characteristics of the clean burning FT fuel.

LASSO[®] designers who handled the fuel during changeover operations noticed the water-like clarity of the FT fuel that is in marked contrast to the amber color of conventional diesel fuel. They also mentioned the lack of a strong odor, (“a kerosene smell”) typically associated with conventional diesel, when handling the FT fuel. The FT fuel was almost odorless by comparison during flush and fill operations. When comparing the performance of the LASSO[®] vehicle driving on the FT diesel, to the test drives taken on conventional fuel, both designers remarked that they noticed no change.

The complete report can be found in Appendix 9.

C. Diesel Gen-Set and Reformer/Fuel Cell Demonstration

Discussion

The original task was to evaluate the performance of high-hydrogen-content coal-derived FT liquid fuel in an existing diesel-engine/generator test-bed at the University of Alaska Fairbanks (Diesel Gen-Set and Reformer/Fuel Cell Demonstration-Task 12.0). The use of coal-derived FT fuel was to be compared to baseline data collected previously by VSE/ICRC and a UAF faculty member (using the same diesel gen-set) on both conventional distillate diesel fuels and FT liquids derived from natural gas. The success of the diesel engine powered vehicle demonstration (task 11) coupled with the departure of the UAF faculty member during the early timeframe of the project, and the directly related decommissioning of the UAF reformer/fuel-cell laboratory facilities, precluded carrying out the originally planned FT fuel demonstrations at UAF.

6. Near Term Production Potential of High Hydrogen Content Coal-Derived Fuel

Discussion/Results

The purpose of this feasibility study was to evaluate the near-term potential of obtaining commercial quantities of high-hydrogen content, coal-derived transportation and jet fuel (FT liquid fuels) from US domestic sources. The National Energy Technology Laboratory had requested an assessment of existing Alaska industrial facilities that could be reconfigured relatively quickly for use in potential coal-to-liquids projects capable of producing large/commercial quantities of coal-derived Fischer-Tropsch liquid fuels within the next few years.

Conversion of appropriate existing industrial facilities was likely to be the fastest route to beginning the US domestic production of high-hydrogen content coal-derived liquid FT transportation fuels. There are existing industrial facilities in Alaska that have (or had approximately two years ago) the potential for being converted to near-term FT fuel production, and their comprehensive evaluation is a primary subject of this report.

Initially, relatively small quantities (a few hundred barrels per day) of such coal-derived liquid FT fuels are needed for validation testing. The validation testing process may also reveal that the initially selected fuel formulations need to be adjusted somewhat to improve the efficiency of the overall fuel-production and fuel-utilization systems. Therefore, a flexible fuel production facility that could be brought on-line quickly, with relatively low production capacity, would actually be preferable, as a first CTL plant, to a much larger commercial scale plant, even though the smaller initial plant would not have the benefit of economy of larger scale. Commercial-scale production of coal-derived FT fuel would then be required for any significant implementation of these fuels into real-world fuel distribution systems. Therefore, the status of projects targeting near-term commercial-scale FT fuel production from coal is also a primary subject of this report. Such projects targeting commercial-scale production could also be based upon conversion of existing industrial facilities, such as those in Alaska, as a means to get on-line sooner. However, it is more likely that larger, more expensive CTL plants would be based upon designing, permitting and building new dedicated facilities from the “ground-up” to maximize efficiency and economies of commercial-scale in producing several thousand barrels per day of FT fuel from coal.

While there are several proposals for large commercial-scale coal-to-liquids projects being advanced in the US, the concept explored in this report is for evaluation of a near term, smaller, domestic plant that could be available within a few years. Of interest also is the capability of producing lesser quantities of fuel to different specifications that could be used for qualification and validation testing. This report reviews the possibilities for such a plant in Alaska. And because there are large-scale coal-to-liquids plants under consideration in relatively close proximity to some of these existing industrial facilities, status reports on these projects are also included. The assessment begins with a

discussion of why Alaska has emerged as optimal for near term pursuit of coal-based FT fuel production.

A FT plant in Alaska should be considered for several reasons:

1. There is a small gas-based FT test plant, the only operating plant (in 2008) of this type in North America, in operation at Nikiski, near Kenai, Alaska. This plant could provide the platform for a small production plant that could produce FT fuels for testing;
2. Alaska has extremely large coal resources, one producing coal mine and a second coal mine at an advanced stage of development;
3. Industrial facilities and hardware already exist which could be converted to coal based FT fuel production;
4. There are large coal-to-liquids projects at advanced stages of consideration in Alaska; therefore it is possible that one of these projects could be developed fairly quickly;
5. Rural Alaska has a small population spread out over a huge expanse of inhospitable terrain and climate, with minimal supporting infrastructure. Although total rural Alaskan distillate fuel demand is low (compared to other states), there is a large demand per person in the Alaskan villages for distillate fuel with extreme low-temperature capability, since this fuel provides for virtually all electricity, space heat, transportation, mobility, construction, maintenance, snow removal, etc., energy needs in rural Alaska. Rural Alaskans already pay the highest prices for liquid fuels of any US location. Jet fuel, with typical sulfur level of several hundred ppm, which is produced locally by Alaska's low-tech, "topping-plant" refineries, has traditionally been used as diesel fuel in rural Alaska;
6. The looming disparity in acceptable sulfur levels between jet fuel (up to 3000 ppm) and diesel fuel (15 ppm maximum as of December 1, 2010 in all of rural Alaska; see link: <http://www.epa.gov/OMS/regs/fuels/diesel/420f06040.htm>), is causing even greater difficulty and will increase costs even further for providing relatively small volumes of an acceptable ultra-low-sulfur, low-temperature capable liquid fuel product to be used as "diesel fuel" in rural Alaska. Locally produced FT jet fuel with virtually no sulfur could be an extremely high-value product that could help meet rural Alaskan energy needs going forward.

Conclusion

The summary of this section of the report concludes with detailed analysis of five feasible options including:

- Near-term: Use of a 300 b/d Fischer-Tropsch plant at Nikiski, Alaska
- Medium-term: A CTL plant at or near the Usibelli Mine, in Interior Alaska.
- Long-term; A large 80,000 barrel-per-day CTL plant near the Beluga coal fields
- Conversion of the Agrium ammonia/urea plant to a FT plant

- A fast-paced development plan working directly with BP for nearly immediate test fuel acquisitions

The complete report, seen in Appendix 10, addresses, in detail, the pros and cons of harnessing existing resources to produce high Hydrogen content FT fuels from domestic coal/biomass resources.

Figures

Figure 1. Project Schedule Showing Projected Timelines, Tasks, Milestones and Deliverables.	20
Figure 2. Impact of adding low levels of H ₂ S to the synthesis gas for FT synthesis with acobalt-alumina catalyst (8.2 g catalyst; GHSV = 2.0 slph/g cat; H ₂ /CO = 2.03; 215°C; 253 psig).	22
Figure 3. Catalyst poisoning rate when adding 611 ppb H ₂ S (8.2 g catalyst;GHSV = 2.0 slph/g cat; H ₂ /CO = 2.03; 215°C; 253 psig).	23
Figure 4. Poisoning rate when adding 717 ppb H ₂ S to a cobalt catalyst (GHSV = 3.76 slph/g cat; H ₂ /CO = 2.06; T = 215°C; P = 255).....	23
Figure 5. Impact of adding 825 ppb H ₂ S during FT synthesis with a cobalt catalyst (P = 253 psig; 227°C, H ₂ /CO = 2.04, GHSV = 8.0; 5% nitrogen).	24
Figure 6. CO conversion with and without addition of H ₂ S (T = 215°C; P = 250; GHSV = 2.59 lphr/g cat).....	25
Figure 7. Decline in CO conversion related to the amount of H ₂ S added to the syngas for a cobalt-alumina catalyst.....	26
Figure 8. Impact of adding 611 ppb H ₂ S and recovery following termination of addition of H ₂ S (GHSV = 2 slph/g cat; 227°C, 253 psig, H ₂ /CO = 2.03).....	27
Figure 9. Impact of addition of COS on CO conversion during Fischer-Tropsch synthesis with a cobalt-alumina catalyst (GHSV = 3 slph/g cat; H ₂ /CO = 2.0; 220°C; 280 PSIG).....	27
Figure 10. Impact of the addition of CS ₂ upon the activity of a cobalt-alumina catalyst 2.04 slph/g cat; H ₂ /CO = 2.06; T = 220°C; 280 psig).	28
Figure 11. Comparison of two runs with same catalyst and reaction conditions except for one run H ₂ S was added after 5 hours (O) and another run where sulfur was not added during the early period of the run (small squares).	29
Figure 12. CO conversion impact upon addition of ammonia as NH ₄ OH (2.48 SL/hr/g cat5, H ₂ /CO – 1.99, 7.47% nitrogen, T= 220oC, 280 psig).	29
Figure 13. Impact of the addition of 1700 ppm ammonia during Fischer-Tropsch synthesis with a Co-alumina catalyst (2.48 SL/hr/g cat5, H ₂ /CO – 1.99, 7.47% nitrogen, T= 220°C, 280 psig).	30
Figure 14. Amount of ammonia in the water collected during the addition of 1700 ppm ammonia (note the feed is based on total syngas flow and the aqueous concentration is based on water collected which is smaller than the syngas feed due to the conversion of CO being about 35%).	31
Figure 15. Impact of the addition of an aqueous nitric acid solution (298 ppm NO ₃ ⁻) upon the CO Conversion (GHSV – 2.44 slph/g cat; H ₂ /CO = 1.93; 10.1% nitrogen, 224°C; P = 280 psig).....	32

Figure 16. Impact of ammonium nitrate addition to a cobalt-alumina catalyst during FT synthesis (GHSV = 4.15; H ₂ /CO = 2.03; 227°C; 250 psig).....	32
Figure 17. Impact of the addition of aqueous potassium nitrate solution to a cobalt-alumina catalyst (8.4 g catalyst, GHSV = 3.00; H ₂ /CO = 2.00; 220°C, 280 psig).....	34
Figure 18. Impact of adding 10 ppm potassium nitrate to the syngas fed to a Co-alumina catalyst (8.2 g catalyst; GHSV = 2.50; H ₂ /CO = 2.01; 220°C; 280 psig).....	34
Figure 19. Impact of the addition of HCl on the activity of a cobalt-alumina catalyst during FT synthesis (GHSV = 2.4 slph/g cat; 10% nitrogen, 220°C; 280 psig,).	35
Figure 20. Impact of KCl on the catalytic activity of a cobalt-alumina catalyst for FT synthesis (GHSV = 2.48 slph/g cat; H ₂ /CO = 2.05, 220°C, 280 psig).	36
Figure 21. Poisonous nature of arsine for a cobalt-alumina catalyst under FT synthesis conditions (GHSV = 3.99; H ₂ /CO = 2.01; 280 psig; 220°C).....	37
Figure 22. Syngas cleaning for iron and cobalt Fischer-Tropsch catalysts	49

Tables

Table 1. Comparison between iron and cobalt Fischer-Tropsch catalysts.....	50
Table 2. Design of a CTL plant based on iron and cobalt FT catalysts (without CO ₂ removed after shift reactor).....	52
Table 3. Design of a CTL plant based on iron and cobalt FT catalysts (with CO ₂ removed after the shift reactor).....	54
Table 4. Hydrocracking runs to obtain kinetic data and product distributions of the hydrocracked materials.	56

References

1. “Increasing Distillate Production at U.S. Refineries – Past Changes and Future Potential,” U.S. Energy Information Administration, Office of Petroleum, Gas, and Biofuels Analysis, Department of Energy, Office of Policy and International Affairs, October 2010. [ftp://ftp.eia.doe.gov/features/distillateprod.pdf](http://ftp.eia.doe.gov/features/distillateprod.pdf)
2. McGurty, Janet, “Analysis: U.S. refineries still need to trim capacity,” Reuters UK, July 28, 2010
<http://uk.reuters.com/article/2010/07/28/us-refinery-survivalofthefittest-idUKTRE66R4KA20100728>
3. Shore, Joanne, and John Hackworth, U.S. Energy Information Administration, “Atlantic Basin Refining Dynamics from U.S. Perspective,” page 11, Platts 4th Annual European Refining Markets Conference, September 2010
http://www.eia.gov/pub/oil_gas/petroleum/presentations/2010/atlanticbasin/atlanticbasin.pdf
4. LeViness, Steve, Heinz Robota and Rafael Espinoza, “Use of Cobalt Fischer-Tropsch Catalysts with Coal Derived Synthesis Gas”, International Pittsburgh Coal Conference; September 2006.
<http://www.engr.pitt.edu/pcc/2006/2006%20Abstract%20Booklet%20Final.pdf>
5. Davis, Burtron H., and Mario L. Occelli, editors “Fischer-Tropsch Synthesis, Catalysts and Catalysis,” CRC Press, Chemical Industries/128, 2010.
6. Davis, Burtron H., and Mario L. Occelli, editors, “Fischer-Tropsch Synthesis, Catalysts and Catalysis,” Elsevier, Studies in Surface Science and Catalysis 163, 2007.
7. Malone, Donald P., and Holcombe, Thomas C., “Design and Performance Simulation of a Clean-up and Water-Gas-Shift System for Syngas Derived from Coal and Biomass,” Project DE-FC26-06NT42449 Milestone Report for Tasks 3 and 4, February 2010. (This Reference is included in Appendix 4 of this report)
8. Tarka, Thomas J., P.E., et. al., “Affordable, Low-Carbon Diesel Fuel from Domestic Coal and Biomass,” US Department of Energy, National Energy Technology Laboratory Report DOE/NETL-2009/1349, January 14, 2009.
<http://www.netl.doe.gov/energy-analyses/pubs/CBTL%20Final%20Report.pdf>

Acknowledgements

This project has been a team effort, with all project team members making their appropriate contributions to its overall success. The members of the project team listed below are singled out as being primarily responsible, both for doing most of the work, and for most of the report writing, on the specific topics indicated.

Dr. Burtron H. (Burt) Davis and support staff members at the University of Kentucky Center for Applied Energy Research (UK-CAER) did the laboratory work and the analyses that are the basis for the “Studying the Poisoning of Cobalt Catalyst Used for FT Synthesis” section of this report. In addition, Burt and his staff carried out the “Evaluation of FT Product-Upgrading Strategies and Equipment.” Furthermore, Burt and CAER provided information to the project on CAER’s R&D Scale FT Pilot Plant, which included much of the background information required for the “Comparison of Cobalt and Iron FT Catalyst Systems for Use with Coal-Derived Syngas” to be performed on the basis of optimized operating conditions for each catalyst-system type.

Rafael Espinoza, Kym Arcuri and their co-workers at Emerging Fuels Technology (EFT) performed the detailed “Comparison of Cobalt and Iron FT Catalyst Systems for Use with Coal-Derived Syngas.”

Don Malone and Tom Holcombe of EnviRes LLC were primarily responsible for the “Synthesis Gas Clean-Up and Related Treatments” section of this report, including their oversight of the design and analyses carried out by engineering subcontractors Worley Parsons and NATCO.

Sara Smith and her colleagues at Smith Management Group (SMG) produced and assembled the “Coal to Liquids (CTL) Plant Site Evaluations and Economic Feasibility Analysis” section of the report, based upon the knowledge and experience gained by SMG in providing liaison for industrial clients with a broad range state and local governments and jurisdictions to obtain the permits required for industrial development projects.

Alaska business writer Tim Bradner marshaled his network of industry and government sources to compile the “Near Term Production Potential of High Hydrogen Content Coal-Derived Liquid Fuel” section of this report, which was focused on underutilized FT-related industrial assets (that were potentially available in 2008) in the Cook-Inlet Region of Alaska.

Hans Steiniger and Al Farkas of ICRC/VSE assisted in the overall technical management and execution of the project in a myriad of ways over several years, and particularly in the assembly and integration of its many technical segments into this final project report. Jim Vorrasso and Leslie Keyes provided guidance and assistance in the financial and administrative management of the project. As a consultant, Kevin Mulrenin continued his liaison efforts among the project team members, many of whom he had been instrumental in bringing to the project team originally, even after Kevin had departed from ICRC/VSE as an employee.

Steve Bergin, Project Director

List of Appendices

- Appendix 1 – List of Acronyms, Abbreviations and Nomenclatures
- Appendix 2 – Effects of Syngas Hydrogen Sulfide Level on Cobalt FT Catalyst
- Appendix 3 – On-Site Measurement of Cobalt FT Catalyst Deactivation Using Coal-Derived Syngas
- Appendix 4 - Performance Simulation of Syngas Cleanup, CO₂ Removal and CO-Water-Gas-Shift
- Appendix 5 - Syngas Treatment Study Documentation
- Appendix 6 – Hydrogen Recovery from Coke Oven Gas
- Appendix 7 - Comparison of Cobalt and Iron FT Catalyst Systems
- Appendix 8 – CTL Plant Site Evaluation & Feasibility Analysis
- Appendix 9 – Military Vehicle Demonstration on FT Fuel from Simulated Coal-Derived Syngas
- Appendix 10 – Conversion of US Domestic Industrial Facilities to Produce FT Fuel from Coal

Appendix 1

List of Acronyms, Abbreviations and Nomenclatures

ASTM	American Society for Testing and Materials
atm	Atmosphere
ATV	All Terrain Vehicle
Bbl/d	Barrel per day
BPD	Barrels per day
BTU	British Thermal Unit
CAER	Center for Applied Energy Research
cCO/ccCat/h	Conversion of CO/cc of Catalyst/hour
CH ₄	Methane
Cl	Chloride

Cm	Centimeters
CO	Carbon monoxide
Co	Cobalt
CO ₂	Carbon dioxide
Comp	Completed
conv	Conversion
COS	Carbonyl sulfide
CS ₂	Carbon disulfide
CSTR	Continuous-Stirred Tank Reactor
CTL	Coal-to-liquids
CX	Carbon chain length (paraffin)
°C	Degrees Celsius
°F	Degrees Fahrenheit
DOE	Department of Energy
EFT	Emerging Fuels Technology
EPA	Environmental Protection Agency
EST\$PRO	Gulf Publishing estimating software
Fe	Iron
Fe ₃ O ₄	Magnetite
FT	Fischer-Tropsch
Ft ³	Cubic feet
GHG	Green House Gas
GHSV	Gas hour space velocity
Gj/hr	Giga joules per hour
GTL	Gas-to-Liquid
H ₂	Hydrogen
H ₂ S	Hydrogen Sulfide
H ₂ S/N ₂	ratio of hydrogen sulfide to nitrogen
HCl	Hydrogen Chloride
HCN	Cyanide
HFTL	Heavy Fischer-Tropsch Liquids
HNO ₃	Nitric Acid
hrs	Hours
ICRC	Integrated Concepts and Research Corporation
ID	Inside Diameter
KCl	Potassium Chloride Sox oxides of sulfur
KNO ₃	Potassium Nitrate
LASSO	Land and Sea Special Operations
LFTL	Light Fischer-Tropsch Liquids

Lphr/g	Liters per hour per gram
LV	Linear Velocity
m	Meters
MATL	Maximum Allowable Threshold Limit
MBTU/hr	Million British thermal units per hour
MFTL	Medium Fischer-Tropsch Liquids
min	Minutes
MMscfd	Million standard cubic feet per day
MPa	Mega pascals
mt/d	Million tons per day
NETL	National Energy Technology Laboratory
NH ₃	Ammonia
NH ₄ NO ₃	Ammonium Nitrate
NH ₄ OH	Ammonium hydroxide
NIMBY	Not In My Backyard
Nm ₃ /d	Normal cubic meters per day
nm ³ /h	Normal Cubic Meters Per Hour
NO ₃	Nitrate
NO _x	Oxides of nitrogen
OD	Outside Diameter
P&ID	Process Instrument Diagram
PDU	Product demonstration Unit
PFD	Process flow diagram
PH ₂ O	Water partial pressure
ppb	Parts per billion
ppm	Parts per Million
PSA	Pressure Swing Absorption
psig	Pounds per Square Inch Gauge
Psig	Psi gage (vs. absolute)
R&D	Research and Development
RETC	Rentech Energy Technology Center
ROI	Return on investment
Rx	Reactor(s)
s	Seconds
S	Sulfur
Scf	Standard cubic feet
scfd	Standard Cubic Feet Per Day
SCFH	Standard Cubic Feet per Hour
Slph/g	Standard liters per hour per gram
SMG	Smith Management Group

SOPO	Statement of Project Objectives
t/d	tons per day
TEMA	Tubular Exchanger Manufacturers Association
tickn	Thickness
Tot	Total
UAF	University of Alaska Fairbanks
UK	University of Kentucky
ULSD	Ultra Low Sulfur Diesel
US	United States
WGS	Water-gas-shift reaction
WP	Worley Parsons
wt	Weight
ZnO	Zinc Oxide

APPENDIX 2
Effects of Syngas Hydrogen Sulfide Level on Cobalt FT Catalyst

**The Sensitivity of Cobalt-Based Fischer-Tropsch Catalyst
to the Level of Hydrogen Sulfide Present in Syngas**

Report Type: Technical

Submitted to:
National Energy Technology Laboratory (NETL)

Cooperative Agreement DE-FC26-06NT42449

Principal Author:

Dr. Burtron H. Davis

Report Period Start Date: July 15, 2008
Report Period Ending Date: April 30, 2009
Submitted: April 30, 2009

Submitting Organization:
Integrated Concepts & Research Corporation
Subsidiary of VSE Corporation
41150 Technology Park Drive
Suite 101
Sterling Heights, MI 48314

Subcontractor:
Dr. Burtron H. Davis
University of Kentucky,
Center for Applied Energy Research
2540 Research Park Dr.
Lexington, KY 40511

ACKNOWLEDGEMENT

This work was supported by the U.S. Department of Energy's National Energy Technology Laboratory (NETL) under Cooperative Agreement DE-FC26-06NT42449.

DISCLAIMER

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

ABSTRACT

The presence of sulfur compounds in Fischer-Tropsch (F-T) syngas has a detrimental effect on catalyst activity. It is frequently stated that cobalt-based F-T catalysts are much more sensitive to sulfur poisoning than iron-based catalysts, but tolerable levels of sulfur when using cobalt catalysts have not been well defined. One “Rule of Thumb” that is frequently cited states that the sulfur content must be reduced to a level of about 10 parts per billion (ppb). This requires severe syngas cleaning and introduces a very costly component for the synthesis reaction. For example, Sasol cleans their synthesis gas to about this level and it is generally stated that this gas cleaning represents about 50% of the cost of producing the syngas. The purpose of this research was to determine the level of hydrogen sulfide (H_2S) in the syngas feed that would cause an unacceptable poisoning of a cobalt F-T catalyst. The data shows that sulfur is a serious catalyst poison but that the level at which it becomes detrimental upon the CO conversion may be at least 10 times higher than has been indicated previously.

TABLE OF CONTENTS

ACKNOWLEDGEMENT.....	2
DISCLAIMER.....	3
ABSTRACT	4
TABLE OF CONTENTS	5
TABLES AND FIGURES.....	6
ACRONYMS, ABBREVIATIONS AND NOMENCLATURES	7
1.0 INTRODUCTION	8
2.0 EXPERIMENTAL.....	8
3.0 RESULTS	9
4.0 SUMMARY	18

TABLES AND FIGURES

FIGURE 1. SCHEMATIC OF THE CONTINUOUS STIRRED TANK REACTOR (CSTR) USED FOR THE POISONING STUDIES	9
TABLE 1. COMPARISON OF THE PHYSICAL PROPERTIES OF CATALYST BATCHES (ALL 0.5% PLATINUM / 25% COBALT / ALUMINA)	10
FIGURE 2. EFFECT ON CONVERSION OF 825 PPB H ₂ S	11
FIGURE 3. POISONING TEST WITH 717 PPB OF SULFUR	12
FIGURE 4. TEST WITH 114 PPB H ₂ S	13
FIGURE 5. TEST WITH 276 PPB H ₂ S	14
FIGURE 6. TEST WITH 611 PPM H ₂ S	15
FIGURE 7. TEST WITH 61 PPB H ₂ S	16
FIGURE 8. TEST AT 61, 242 AND 276 PPB H ₂ S	17
FIGURE 9. TEST AT 438 PPB H ₂ S	18
FIGURE 10. SUMMARY OF DATA FOR RUN AWP006 DURING THE ADDITION OF H ₂ S.	19

ACRONYMS, ABBREVIATIONS AND NOMENCLATURES

BET surface area, m ² /g	Brunauer, Emmett, and Teller, (inventors of the concept)
CAER	Center for Applied Energy Research
CSTR	Continuous Stirred Tank Reactor
GHSV	Gas Hourly Space Velocity: Volume of gas feed per hour per volume of catalyst
m ² /g	Square meters area per gram of catalyst
ppb	Parts Per Billion
ppm	Parts Per Million
Psig	Pounds per square inch gauge pressure
R19 / R20	Reactor 19 / Reactor 20
Slph/g	Standard Liters per Hour Per Gram
UHP	Ultra High Purity

1.0 INTRODUCTION

The presence of sulfur compounds in Fischer-Tropsch (F-T) syngas has a detrimental effect on catalyst activity. It is frequently reported that cobalt-based catalysts are much more sensitive to sulfur poisoning than iron-based catalysts, but tolerable levels of sulfur when using cobalt catalysts have not been well defined. One “Rule of Thumb” that is frequently cited states that the sulfur content must be reduced to a level of about 10 parts per billion (ppb). This requires severe syngas cleaning and introduces a very costly component for the synthesis reaction. For example, Sasol cleans their synthesis gas to about this level and it is generally stated that this gas cleaning represents about 50% of the cost of producing the syngas. The initial portion of the CAER research program was to determine the level of hydrogen sulfide (H_2S) in the syngas feed that would cause an unacceptable poisoning of a cobalt F-T catalyst. This report fulfills the Statement of Project Objectives milestone requirement of completing sub-task 2.2 (Impact of H_2S level on poisoning rate of cobalt F-T Catalyst). The report presents experimental data that quantifies the effect of sulfur level on the cobalt F-T catalyst.

2.0 EXPERIMENTAL

There are currently two one-liter continuous stirred tank reactors (CSTRs) that are committed to this poisoning study. Rather than risk potential contamination to the other 18 CSTRs that are already in service for various other F-T research projects, these two reactors (designated as R19 and R20) were built specifically for this study. The operation of these reactors has been described in many CAER publications (see, for example, a recent Reference at the end of this report).

The hydrogen (H_2) and the carbon monoxide (CO) used for all of the reactors are provided from two large tube trailers that are typically refilled about 2-3 times per year, helping to ensure a constant level of any extraneous background sulfur contamination from these two principal constituents of the syngas mixture. Frequently, a small amount of nitrogen (N_2) is added to the syngas (~5-10% of the total) as an aid for mass balance calculations, and this is supplied via a manifold of four high pressure ultra high purity (UHP) gas cylinders, with each cylinder sufficient for about four days of operation for all twenty CSTRs.

FIGURE 1. SCHEMATIC OF THE CONTINUOUS STIRRED TANK REACTOR (CSTR) USED FOR THE POISONING STUDIES.

For these first tests, individual gas cylinders containing various ppm levels of H_2S in N_2 were used to add the sulfur contaminant. After establishing a baseline conversion with a syngas mixture containing only H_2 , CO and N_2 , a fourth gas mass flow controller was used to add the $\text{H}_2\text{S}/\text{N}_2$ mixture, with the UHP N_2 feed then being reduced or eliminated. Typically, after several days of operation with declining activity, the $\text{H}_2\text{S}/\text{N}_2$ feed was stopped while the conversion continued to be monitored. If a steady conversion of greater than ~30-40% was obtained, another test with a different H_2S concentration was started. If the conversion was lower than this, the syngas feed rate was reduced to first increase conversion before proceeding with the next test.

3.0 RESULTS

A large amount of catalyst (designated as GJ452, and containing 0.5 wt% platinum (Pt) and 25 wt% cobalt (Co) on an alumina (Al_2O_3) support) had been produced so that all of the poisoning runs could be made using catalyst drawn from the same batch. When there was some question of its performance after the first few runs, 10g portions of this catalyst and the subsequent near identical GJ455 batch (which was intended for a different

project) were each tested with no H₂S addition in two different CSTRs to confirm that both batches exhibited acceptable conversion.

Both batches of catalyst used Condea Vista Sba-200 (high purity γ -alumina, $\sim 200 \text{ m}^2/\text{g}$) as the support. Both were prepared by a slurry impregnation method, and cobalt nitrate was used as the precursor. In this method, which follows a Sasol patent, the ratio of the volume of solution used to the weight of alumina was 1:1, such that approximately 2.5 times the pore volume of solution was used to prepare the loading solution. Two impregnation steps were used, each to load 12.5% of Co by weight. Between each step the catalyst was dried under vacuum in a rotary evaporator at 80°C and the temperature was slowly increased to 95°C. After the second impregnation/drying step, the Pt promoter was added by incipient wetness impregnation via an aqueous solution containing tetra-ammine platinum (II) nitrate. Following promoter incorporation, the catalyst was calcined in flowing air at 350°C.

There were however two presumably minor differences in the production of the two catalyst batches. Two different lots of the Condea Vista Sba-200 support were used to make the two batches. Also, GJ452 was calcined in flowing air at the CAER using a tubular reactor, while GJ455 was calcined in flowing air at the same temperature, but in a special oven at Süd-Chemie, Inc., Louisville, KY.

This technique has been used many times at the CAER to produce similar catalysts, albeit usually with an alumina support that has a lower BET surface area, producing a catalyst with a BET of $\sim 100 \text{ m}^2/\text{g}$.

property / catalyst number	(prior)	GJ452	GJ455
BET surface area, m^2/g	103.0	134.6	141.2
Single Point Pore Volume, cm^3/g	0.258	0.241	0.253
BJH Adsorption Pore Volume, cm^3/g	0.264	0.249	0.279
Single Point Average Radius, nm	~ 5.0	3.570	3.576
BJH Adsorption Average Radius, nm		3.440	3.621

TABLE 1. COMPARISON OF THE PHYSICAL PROPERTIES OF CATALYST BATCHES (ALL 0.5% PLATINUM / 25% COBALT / ALUMINA)

The first successful H₂S poisoning run was AWP006, which began in R20 on 18 December 2008 and ended 892 hours later on 29 January 2009. Two graphs are presented from this run showing rapid loss of activity with large amounts of H₂S in the feed. The first poisoning test during this run is illustrated by Figure 2. In this case, R20 was operating with 4% UHP N₂ added to the syngas for the first 92 hours on stream. At this point, this UHP N₂ feed was replaced by a similar amount of a 19 ppm H₂S/N₂ mixture, producing a syngas mixture that contained 825 ppb H₂S.

R20 Run AWP006 7.1g of GJ452 1st H₂S poisoning test
 feeding 7.967 slph (CO + H₂) per gram of catalyst, plus 0.375 slph/g UHP N₂
 (H₂ + CO + N₂) = 59.262 slph H₂/CO ratio = 2.037, 215 C, 253 psig

FIGURE 2. EFFECT ON CONVERSION OF 825 PPB H₂S

At the end of this first test, the H₂S/N₂ mixture was replaced with UHP N₂ and the overall syngas feed rate was reduced several times over the course of the next few days to increase the CO conversion. When a stable conversion of ~50% was achieved (as seen on the left side of figure 3), an additional 0.156 slph/g of the 19 ppm H₂S/N₂ was added to the syngas, without reducing or eliminating the UHP N₂ feed. This resulted in a slightly lower gas residence time in the reactor, but of course did not affect the (CO + H₂) syngas feed rate. The H₂S composition in the total gas feed was 717 ppb. This test also showed rapid loss of activity at these high H₂S levels.

FIGURE 3. POISONING TEST WITH 717 PPB OF SULFUR.

The second reactor to be dedicated to the poisoning study was R19, with construction completed in mid-December 2008. After a brief trial run that lasted only four days using a syngas that was heavily contaminated with sulfur compounds, this reactor was deemed ready to begin poisoning tests. Run AWP007 commenced on 25 December 2008 using the GJ452 catalyst, but was terminated on 04 January 2009 due to low CO conversion. Post run, it proved to be difficult to remove the head bolts. New head bolts made from a different alloy were ordered to replace the originals. Run AWP009 commenced on 12 February, but this run was also terminated after only one week due to low CO conversion. The reactor body was subsequently removed from the stand and the inside of the body was sandblasted to remove any possible sulfur contamination from the initial run. The third run in R19 was AWP010, which commenced on 27 February and terminated on 10 April after 1000 hours. Three successful poisoning trials were made during this run.

R19 Run AWP010 8.2g of GJ452 1st H₂S poisoning test
 feeding 4.01 slph (CO + H₂) per gram of catalyst, plus 0.21 slph/g UHP N₂
 (H₂ + CO + UHP N₂) = 34.597 slph H₂/CO ratio = 1.985, 215 C, 253 psig

FIGURE 4. TEST WITH 114 PPB H₂S

The first test is shown in Figure 4. The addition of 114 ppb H₂S was begun after six days on stream, which in retrospect was too early, as stable conversion without H₂S had not yet been achieved. The continuing loss of activity following the introduction of H₂S is believed to be attributable to the catalyst's normal performance during the break-in period, and that this break-in period was longer than usual due to the catalyst's greater surface area.

At the end of this test, the feed rate was cut in half and sufficient time was allowed to demonstrate a steady conversion level. After four days, the H₂S was again introduced, with the concentration increased to 276 ppb. No loss of activity was seen in the following seven days on stream at this level of contamination, further indicating that the loss of activity seen in the previous test was actually a result of a longer than anticipated break-in period.

R19 Run AWP010 8.2g of GJ452 2nd poisoning test
 feeding 1.863 slph (CO + H₂) per gram of catalyst, plus 0.115 slph/g UHP N₂
 (H₂ + CO + UHP N₂) = 16.230 slph H₂/CO ratio = 2.032, 215C, 253 psig

FIGURE 5. TEST WITH 276 PPB H₂S

After seven days with no loss of activity, the H₂S concentration was increased directly to 611 ppb with no intervening period without the contaminant. The left side of Figure 6 is a repetition of the last two days shown in Figure 5. The middle portion of the figure shows the declining activity with this increased H₂S concentration and the right side illustrates a partial recovery of activity when the H₂S was removed from the feed.

FIGURE 6. TEST WITH 611 PPM H₂S

R20's next run was AWP008, which began on 04 February, but this run was terminated after ~450 hours when a stable CO conversion of ~50% could not be achieved. Two possibilities for this were discussed and addressed. One possibility was that there had been a lingering sulfur contamination in the reactor from the previous run, so extra effort was applied in scrubbing the reactor before loading the catalyst for the next run.

A second possibility was that there might have been a problem with the catalyst itself, and so the next run (AWP011, started on 27 February), used catalyst from the GJ455 batch. Four separate poisoning tests were completed during the 1000 hours of this run.

FIGURE 7. TEST WITH 61 PPB H₂S

Figure 7 illustrates the first test conducted on this run. 61 ppb of H₂S was introduced after six days on stream. While this was once again later realized as being too soon after startup, the rate of loss of activity after introducing the H₂S was actually less than what was seen in the six days before sulfur was added.

While continuing to feed 61 ppb H₂S, the conversion leveled out at about 30%. The H₂S concentration was increased to 242 ppb and the conversion actually increased by 5% over the next four days. When the H₂S/N₂ cylinder was expended at this point, it was replaced with a new cylinder that had a slightly higher H₂S content and the conversion began declining, as seen in Figure 8. As a previous test at near this concentration had shown no loss of activity, it's presumed that the reported concentration in the supply cylinder was in error.

FIGURE 8. TEST AT 61, 242 AND 276 PPB H₂S

At the end of this test, the feed rate was reduced and the H₂S feed was discontinued to allow the conversion to stabilize. After six days, the conversion had begun to stabilize and so a third test was started. H₂S was again introduced, this time at a concentration of 438 ppb.

FIGURE 9. TEST AT 438 PPB H₂S

After comparing the results with the two different catalysts and going back over the accumulated data, it became more apparent that at least some of the rapid loss of activity observed over the first approximately ten days on stream on most of the runs was due to the properties of the catalyst itself, and not due to lingering sulfur contamination in the reactor from previous runs.

Eight separate poisoning tests using different concentrations of H₂S have been completed in the course of three different successful runs (AWP006, AWP010 and AWP011), while three runs were terminated, possibly prematurely, due to low catalyst activity (AWP007, AWP008 and AWP009).

4.0 SUMMARY

Using the dispersion of the cobalt catalyst and the decline in activity, one can calculate the impact of the addition of a sulfur atom in eliminating active Co surface atoms. For the period of the addition of high loading of sulfur where there was an apparent decline in activity, one calculates that each sulfur atom added eliminated 5 to 6 cobalt surface atom's catalytic activity. This implies that a site for the Fischer-Tropsch synthesis

requires a site that contains 5 to 6 cobalt atoms or, if the active site is one cobalt atom, then the impact of one sulfur atom extends beyond the cobalt upon which the sulfur is adsorbed.

FIGURE 10. SUMMARY OF DATA FOR RUN AWP006 DURING THE ADDITION OF H₂S.

The purpose of this portion of the research program was to establish the concentration of sulfur in the syngas that would impact the catalytic activity. The data shows that this level is about 400 ppb and therefore provides the level of sulfur that needs to be added to obtain the long-term poisoning data to establish a reliable impact of sulfur in the decline of catalytic activity. This data shows that sulfur is a serious catalyst poison but that the level that it becomes detrimental upon the CO conversion may be at least 10 times higher than has been indicated previously.

Reference

Davis, B.H. "Fischer-Tropsch Synthesis: reaction mechanisms for iron catalysts,"
Catalysis Today, 141, 25-33, 2009. **(2009 Jan/Feb)**

APPENDIX 3
On-Site Measurement of Cobalt FT Catalyst
Deactivation Using Coal-Derived Syngas

On-Site Measurement of Cobalt FT Catalyst Deactivation with Actual Coal-Derived Syngas at a Coal Gasification Facility

Type of Report: Technical

Reporting Period Start Date: March 30, 2006

Reporting Period End Date: January 31, 2011

Principal Authors:

Burtron H. Davis, University of Kentucky Center for Applied Energy Research

Stephen P. Bergin, VSE Corporation

Date Report Issued: January 2011

DOE Award Number: DE-FC26-06NT42449

Submitted: January 25, 2011

VSE Corporation

41150 Technology Park Drive, Suite 101

Sterling Heights, MI 48313

DISCLAIMER

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

ABSTRACT

This report shows that a data-set obtained using “bottled” laboratory syngas (primarily carbon monoxide and hydrogen, but also containing a range of low and controlled levels of expected syngas contaminants), to measure and map the laboratory performance and deactivation of a cobalt Fischer-Tropsch (FT) catalyst, can predict accurately the performance of the same cobalt FT catalyst when it is used with actual coal-derived syngas containing real-world contaminants. The predominant, recurrent theme in this report is the level of sulfur contained in syngas, as one of the contaminants of primary concern, and the effects of syngas sulfur on cobalt FT catalyst performance and deactivation.

When the goal is to produce FT fuels from syngas obtained by gasification of any solid resource, whether the resource is coal, biomass, petroleum coke, or some combination, the required basic syngas clean-up system will, of necessity, be expensive, complicated, and capable of removing virtually all contaminants of concern for potentially poisoning an FT catalyst, including sulfur, down to levels of parts per million. This report’s comparison, which demonstrates that a laboratory data-set can indeed predict cobalt FT catalyst performance with actual coal-derived syngas, is based upon the strategy that multi-stage syngas clean-up is required to reduce the sulfur level of the syngas to an even lower level than that obtained with the basic syngas clean-up system, to the minimum level that is practically achievable, in both the laboratory and at the real-world coal-gasification site.

Minimum syngas sulfur level is achieved under this strategy by specifying that guard-beds containing zinc oxide (to remove almost all sulfur down to a few parts per billion, maximum) must always be used, at both the laboratory scale, and at the commercial-plant scale. A commercial-scale guard-bed system adds significant costs to the overall syngas clean-up system of coal-to liquids plant, but in addition to minimizing syngas sulfur level under “normal” operating conditions, it also provides some level of “insurance” for protecting the downstream cobalt FT catalyst-charge from extreme spikes of syngas-sulfur under “abnormal” transient conditions that can be caused by upsets within the plant.

One of the most striking laboratory-based findings of this overall project is that cobalt FT catalyst may be able to tolerate a steady syngas sulfur level, in the form of hydrogen sulfide, H₂S, an order of magnitude greater than previously believed, and still deliver acceptable catalyst reactivity and life. If proven correct, this could allow a reduction in the cost, complexity, energy consumption, maintenance, etc., of a coal-derived syngas clean-up system for use with cobalt FT catalyst, possibly even eliminating the need for a second-stage guard-bed system.

An alternative, and more academically oriented, interpretation of this finding (that cobalt FT catalyst may be able to tolerate ~10X more sulfur) was that more fundamental research is needed. A mandatory peer review of this project was conducted by outside-experts contracted by DOE-NETL. These reviewers recommended strongly (so strongly

that follow-up on the recommendation became a project milestone) that the project team apply nanoscale instrumentation to show the amounts and sites of sulfur poisoning on the cobalt FT catalyst, an added project task that the project team was able to accomplish.

But the project team also sought to apply this finding (that ~10X sulfur may be acceptable) and further validate and expand upon the comparison made between laboratory and real-world cobalt FT catalyst performance data, but at significantly higher, known, and controllable, levels of hydrogen sulfide in the actual coal-derived syngas. Such a program would require variable clean-up intensity of the coal-derived syngas. Attempting to implement variable syngas cleanup would be so highly disruptive to a commercial coal-gasification site that a host-site could not be obtained for conducting this next-step program.

TABLE OF CONTENTS

DISCLAIMER	II
ABSTRACT	III
TABLE OF CONTENTS	V
ACRONYMS, ABBREVIATIONS AND NOMENCLATURES	VI
EXECUTIVE SUMMARY	2
INTRODUCTION	3
1.0 COBALT FT CATALYST DATA USING BOTH LABORATORY AND COAL- DERIVED SYNGAS	4
1.1. Syntroleum's Strategy: Minimize Syngas Sulfur Level for Cobalt FT Catalyst.....	4
1.2. Guard Beds	10
1.3. Comparison of Cobalt FT Catalyst Data with Coal-Derived and Laboratory Syngas	11
2.0 UNIVERSITY OF KENTUCKY-CAER FINDINGS: COBALT FT CATALYST MAY BE ABLE TO TOLERATE A SYNGAS-SULFUR LEVEL 10 TIMES HIGHER THAN THOUGHT PREVIOUSLY.....	16
2.1. Laboratory Data.....	16
2.2. Peer Review 2009 and Follow-up Laboratory Work at UK-CAER	18
2.3. Goals for Follow-on Field-Test with Controllable Sulfur Level of Coal-Derived Syngas	19
3.0 PROJECT HISTORICAL INFORMATION	20
3.1. The Textbook Approach to the Coal-To-Liquids Research Project.....	20
3.2. A Change in Syntroleum's Corporate Direction.....	21
3.3. Restructuring of the Project.....	22
REFERENCES	24
APPENDIX.....	25

ACRONYMS, ABBREVIATIONS AND NOMENCLATURES

NH ₃	Ammonia
As	Arsenic
CO	Carbon Monoxide
CAER	Center of Applied Energy Research
CEO	Chief Executive Officer
Cl	Chlorine
CTL	Coal-to-Liquid
Co	Cobalt
CSTRs	continuous stirred tank reactors
DOD	Department of Defense
DOE	Department of Energy
FT	Fischer-Tropsch
F	Fluorine
GTL	Gas-to-Liquid
H ₂ S	hydrogen sulfide
Hg	Mercury
NETL	National Energy Technology Laboratory
OEM	Original Equipment Manufacturer
ppb	parts per billion
PDU	Product Demonstration Unit
Sr	Strontium
syngas	synthesis gas
UK	University of Kentucky

EXECUTIVE SUMMARY

The purpose of this task is to compare the performance and deactivation of cobalt Fischer-Tropsch (FT) catalyst, using actual coal-derived syngas, to the performance of the same cobalt FT catalyst under controlled laboratory conditions intended to simulate the coal-derived syngas environment. The ultimate objective is to use the resulting comparison data to validate the laboratory-based approach for predicting cobalt FT catalyst performance in the real world. Continuous stirred tank reactors (CSTRs) are used for the cobalt FT catalyst testing because the constant mixing exposes all of the cobalt FT catalyst under test to the syngas equally, and to the same level of whatever impurities or contaminants may be present in the syngas.

The data show conclusively that laboratory results do indeed predict accurately the performance and deactivation of cobalt FT catalyst in a coal-derived syngas environment. These data, obtained by Syntroleum at Tennessee-Eastman using experimental apparatus installed in support of this project, also demonstrate the overall feasibility of using cobalt FT catalyst with coal-derived syngas.

The basic strategy used in all phases of this comparison (laboratory, field-test at the coal-gasification site, and ultimately at a future coal-based FT fuel production plant) would be to:

- Absolutely minimize the exposure of the cobalt FT catalyst to irreversible poisons in the syngas such as sulfur (i.e. to a few parts per billion, ppb, maximum),
- But accept a higher level (a few parts per million) of exposure to other potential cobalt catalyst poisons such as ammonia in the syngas that have been demonstrated to be reversible by using cobalt-catalyst regeneration technology.

This basic strategy is predicated on maximized, multi-stage clean-up (i.e. including guard beds) capable of removing almost all syngas sulfur, down to a continuous level no greater than about 20 ppb, even in a commercial-scale coal-based FT plant.

One of the most striking laboratory-based findings of this overall project is that cobalt FT catalyst may be able to tolerate syngas sulfur level, in the form of hydrogen sulfide, H_2S , an order of magnitude greater than previously believed, and still deliver acceptable catalyst reactivity and life. Specifically, an H_2S level of 200 ppb, or possibly even somewhat higher (~240 ppb), might prove to be acceptable based upon laboratory CSTR data. If proven correct, this could allow a reduction in the cost, complexity, energy consumption, maintenance, etc., of a coal-derived syngas clean-up system for use with cobalt FT catalyst.

Therefore, the project team sought to further validate and expand upon the type of comparison presented, but at significantly higher, known, and controllable, levels of hydrogen sulfide in the coal-derived syngas (i.e. with variable clean-up intensity of coal-

derived syngas). The ideal follow-on test program would be to make at least one (and preferably more) approximate-two-month CSTR run(s), with fresh cobalt FT catalyst, covering a range of syngas H₂S levels, at the same (Tennessee Eastman), or possibly at another, continuously operating (24/7/365) coal-gasification site.

ICRC and project-partner UK-CAER considered a range of potential options for the installation and operation of a laboratory CSTR at a coal gasification site where a controllable syngas-sulfur-level could be obtained. A continuous-operation commercial site would offer the major advantage of access to virtually uninterrupted “live” syngas, but the only two such sites in the US, Tennessee-Eastman and Dakota Gasification, both declined overtures from the project team to participate in a project of this type.

The project team also made contact with the managers of other potential coal gasification sites. Rentech was in the process of installing a small-scale gasifier system for commercial-type long-term operation within their Product Demonstration Unit (PDU) in Denver (<http://www.rentechinc.com/pdu.php>), and expressed interest in working with the project team on this task. However, the gasifier system would not operate acceptably on coal (despite its OEM rating), so the system was ultimately removed from the PDU and replaced by a natural-gas reformer system so that the rest of the pilot FT plant could still operate.

Other research-oriented gasifier sites were considered, but these could operate for only relatively short periods (typically about a week), between shutdowns of comparable duration, thus requiring compression and storage of coal-derived syngas and its contaminants, with attendant difficulties and uncertainties thus introduced into the catalyst testing results.

INTRODUCTION

This report covers one sub-task in an overall research project comprised of 13 total tasks that cover a broad range of topics related to the production of high hydrogen content coal-derived liquids. Accomplishment of this particular sub-task required a project team with the full range of FT competencies and capabilities, ranging from a thorough fundamental understanding of FT synthesis catalyst technology, to long-term, on-site access to extremely capital-intensive (and uncommon within the US) commercial coal-gasification and syngas clean-up facilities that operate with virtually no interruptions.

Laboratory-based systems are typically designed to provide the maximum latitude in controlled experimental conditions, but at both a scale and a cost that are orders of magnitude smaller than the commercial-scale systems to be simulated. Industrial systems, by contrast, are built to operate efficiently within a relatively narrow prescribed window of conditions and their capital and operating costs are orders of magnitude greater than those of laboratory systems. Therefore, while it might be desirable to compare results over the full range of obtainable laboratory conditions to commercial-scale system results over the same range; such a wide-range comparison is unlikely to be technically or economically feasible.

1.0 COBALT FT CATALYST DATA USING BOTH LABORATORY AND COAL-DERIVED SYNGAS

The purpose of this section is to review the performance and deactivation data for cobalt FT catalyst obtained in both the laboratory and at a coal-gasification site. However, it is necessary to understand the strategy being used in order to interpret the data.

1.1. Syntroleum's Strategy: Minimize Syngas Sulfur Level for Cobalt FT Catalyst

Syntroleum's strategy for using cobalt FT catalyst with coal-derived syngas includes minimizing the sulfur contamination of the syngas. This strategy is outlined in a presentation (Reference 1) made to the Pittsburgh Coal Conference in September 2006, as preparations were being made to operate the CSTRs at the Tennessee Eastman coal-gasification facility under this project (Figure 1). (As described in Section III of this report, Syntroleum would later withdraw from active participation at their originally forecasted levels; seek a less active supporting role in the project; and ultimately continue to operate the same test facility as originally planned.)

Figure 1. Cover Slide of Presentation Outlining Syntroleum's Strategy

The 2006 presentation made several key points regarding Syntroleum's strategy for using cobalt FT catalyst with coal-derived syngas, as shown in the following 10 Figures:

- Carbon efficiency, expressed as the % of carbon in the syngas converted to hydrocarbons, can be higher for cobalt FT catalyst than for iron in CTL (Figure 2)
- Poisons in the syngas feed can be reversible or irreversible (Figure 3)
- Nitrogen containing poisons can be "reversed" (Figures 4-6)
- Sulfur containing poisons cannot be reversed and must be minimized (Figures 7 & 8)
- Cobalt FT catalyst requires the use of guard beds as an extra clean-up step, at extra cost, to minimize the sulfur contamination of coal-derived syngas (Figures 9-11)

Figure 2.

POISONS IN THE FEED

Syntroleum

- Reversible poisons, e.g. nitrogen based compounds
- Irreversible poisons, e.g. sulphur based ones
- Other - Hg, Sr, As, Cl, F, etc

27

Figure 3.

NITROGEN CONTAINING POISONS AND SURFACE CONDENSATION

Syntroleum

- These poisons are reversible and can be dealt with using a rejuvenation and/or a regeneration process.
- At Syntroleum, both treatments have shown to be very effective in this case, as well as with deactivation due to surface condensation and in the extreme, also for the unlikely carbon deposition.
- The percentage of activity recovered was in excess of 98-99% (nearly 100% in the laboratory experiments).

28

Figure 4.

EXAMPLE OF CATALYST MAINTENANCE STRATEGY AT VARIOUS LEVELS OF NH_3 IN THE FEED

Syntroleum

ASSUMPTIONS

- Only 96% activity recovery for rejuvenation
- Regeneration recovers 99%
- Sulphur poisoning non-regenerable
- In case for 3000 ppb NH_3 (b), the rejuvenation capacity was increased by 40% and regeneration by 15%. Same frequency.
- Fresh catalyst added in the two year period: 65% of initial inventory: 3 year life

29

Figure 5.

SAME FIGURE, EXPANDING THE REGION OF INTEREST

Syntroleum

30

Figure 6.

SOME CONCLUSIONS ON REVERSIBLE AND IRREVERSIBLE CONTAMINANTS

Syntroleum

The previous figures were prepared based on demonstration scale regeneration procedures and laboratory data on the effect of the NH_3 level (and other N containing compounds) on the rate of deactivation.

They show that it is possible to deal with reversible deactivation mechanisms over a large range of concentrations.

Due to its irreversible nature, it is not practically possible to regenerate a cobalt based sulphur poisoned catalyst.

Because a cobalt catalyst is expected to have a life time of 2 or more years, it is imperative to remove the irreversible poisons from the syngas feed.

A generally accepted level of irreversible contaminants in the feed for cobalt application is less than 20 ppb. For iron application, this level is about an order of magnitude higher.

31

Figure 7.

DIFFERENCES BETWEEN Fe AND Co BASED PROCESSES DUE TO IRREVERSIBLE POISONS IN THE FEED

Syntroleum

32

Figure 8.

COST OF FEED(S) CLEANING FOR Fe AND Co CATALYSTS

Syntroleum

The S compounds in the feed for both catalysts have to be very low, but cobalt catalysts need an extra step, to go from the low 100's ppb (eg ~ 200) for Fe catalysts to the low 10's (eg ~ < 20) ppb.

33

Figure 9.

CONCLUSIONS FROM THE MATERIAL IN THIS PRESENTATION

Syntroleum

- Cobalt catalysts can deal with different H_2/CO ratio (low and high) from gasifiers as good or better than iron catalysts. Iron catalysts will have a lower carbon efficiency as the H_2/CO ratio approaches the stoichiometric value.
- Cobalt catalysts can deal with reversible poisons in a practical manner.
- The synthesis gas feed will have to be cleaned more extensively for cobalt catalysts as compared to iron catalysts. The feed for iron catalysts also has to be extensively cleaned.
- This extra cleaning cost (for Co catalysts) will have to be compensated by: lower reactor volume requirements, lower catalyst cost and ability to achieve higher overall co-conversions (e.g. middle 90's) in two or less stages.

34

Figure 10.

Figure 11.

1.2. Guard Beds

The guard beds used in Syntroleum's strategy are absorbing materials (normally zinc oxide) deployed and intended to remove "almost all of the last remaining" level of contamination, particularly sulfur contamination, before the syngas enters an FT reactor such as a CSTR, or a commercial FT reactor. When used in a bench-scale experiment, it is usually possible to provide sufficient guard-bed absorbent capacity, and to replace it as often as necessary, to obtain an extremely low final level of contamination, but still not down to absolute zero contamination.

The question regarding the capacity of a guard-bed system used in a bench-scale experiment is whether or not the same guard-bed capacity, on an equivalent basis per unit volume of syngas processed, would also be cost effective and practical in a full-scale installation. In Syntroleum's strategy, guard beds must always be used to minimize syngas sulfur contamination, whether in a small-scale experiment or in a commercial-scale FT plant. Although guard beds add some costs, they can also be considered "cheap insurance," especially in the event of an unanticipated upset in the coal-gasification process that could otherwise destroy a very expensive cobalt FT catalyst-charge.

1.3. Comparison of Cobalt FT Catalyst Data with Coal-Derived and Laboratory Syngas

Syntroleum set up two CSTRs for testing cobalt FT catalyst at the Tennessee Eastman coal gasification facility under this project. It should be noted that the constant mixing inside a CSTR, as in a slurry-type FT reactor, exposes all of the FT catalyst equally to the syngas, and to whatever impurities or contaminants it might contain. This is in contrast to a fixed-bed catalyst configuration, in which the top (or entry) layers of the catalyst bed can “filter-out” some of the contaminants, such that the “filtered-out” contaminants are localized, and therefore don’t immediately reach the lower (or interior) levels of the catalyst bed. CSTRs therefore represent the most severe test condition for the effects of impurities in the syngas on FT catalyst performance and deactivation.

One CSTR ran on “live” coal-derived syngas, and the other CSTR ran on “bottled” reference syngas for a back-to-back, on-site reference and comparison, both to the “live” syngas CSTR, and to laboratory CSTR data obtained previously. After the testing at the coal-gasification facility had been completed, Syntroleum made a presentation (Reference 2) of the results at the Gasification Technology Conference in San Francisco in October 2007 (Figure 12).

Figure 12. Cover slide of presentation describing the test results

The 2007 presentation of the Tennessee Eastman on-site data made several key points, as shown by the following 7 Figures:

- Runs of 2500 hours (over 3 months) were made (Figure 13)
 - Levels and types of (reversible) contaminants were measured
 - Steady state deactivation rate of cobalt FT catalyst was measured
 - Cobalt FT catalyst was regenerated, and its renewed activity measured
- Syntroleum's equipment was installed at Tennessee Eastman before the end of 2006
 - An on-site laboratory-type facility was used for testing (Figure 14)
 - CSTR reactors, guard beds and controls were installed (Figure 15)
- Previous Laboratory Trace Contaminant Study Data were presented (figure 16)
 - Un-identified "contaminant" at levels of 0 ppb, X, 10X and 100X
 - Compare to Figures 5 and 6 for NH₃ at 30, 300 and 3000 ppb levels
- Data from "live" syngas & reference CSTRs plotted with previous lab data (Figure 17)
 - Lab and "live syngas" cobalt FT catalyst deactivation rates are similar
 - On-site reference CSTR data provides additional verification
 - **Figure 17 demonstrates that laboratory data can accurately predict cobalt FT catalyst performance with actual coal-derived syngas**
- Data from Syntroleum's GTL pilot plant was also added to the same plot (Figure 18)
- Cobalt FT catalyst from Tennessee Eastman was regenerated and re-tested (Figure 19)
 - Un-identified "contaminant(s)" in the syngas thus shown to be reversible
 - Guard-beds minimized any sulfur contamination in the coal-derived syngas

Objectives

Syntroleum

- Conduct a long (2500 hour) run on live coal derived syngas from Eastman Chemical Company's commercial gasification facility in Kingsport, TN
- Measure levels and type of contaminants
- Measure steady state deactivation rate of Syntroleum cobalt FT catalyst
- Regenerate FT catalyst and measure initial rate

2

Figure 13.

Catalyst Testing Unit

Syntroleum

5

Figure 14.

Reactors, Guard Beds and Controls

Syntroleum

Figure 15.

Laboratory Trace Contaminant Study

Syntroleum

7

Figure 16.

Bench Scale CSTR Data

Syntroleum

8

Figure 17.

CSTR Data and Syntroleum Pilot Plant Data

Syntroleum

9

Figure 18.

Syntroleum FT Catalyst before and after Regeneration

Syntroleum

10

Figure 19.

The same data as is presented in Figures 13 through 19 above has also been available, along with some additional descriptive information, on Syntroleum's website since late 2007 (Reference 3).

2.0 UNIVERSITY OF KENTUCKY-CAER FINDINGS: COBALT FT CATALYST MAY BE ABLE TO TOLERATE A SYNGAS-SULFUR LEVEL 10 TIMES HIGHER THAN THOUGHT PREVIOUSLY

One of the most striking laboratory-based findings of this overall project is that cobalt FT catalyst may be able to tolerate syngas sulfur level, in the form of hydrogen sulfide, H_2S , an order of magnitude greater than previously believed, and still deliver acceptable catalyst reactivity and life. Specifically, an H_2S level of 200 ppb, or possibly even somewhat higher (~240 ppb), might prove to be acceptable based upon UK-CAER laboratory CSTR data. If proven correct, this could allow a reduction in the cost, complexity, energy consumption, maintenance, etc., of a coal-derived syngas clean-up system for use with cobalt FT catalyst.

2.1. Laboratory Data

As shown in Figure 20, the addition of H_2S showed that low levels (few parts per billion) did not cause a decrease of conversion at the lowest level, and may even have caused the conversion to increase as has been reported by many others. Thus, at the 240 ppb level

H₂S may have caused a slight increase in conversion, but at 276 ppb H₂S a decline in CO conversion is noted.

Figure 20. Addition of low levels of H₂S to cobalt FT catalyst

The laboratory data shown in Figure 20, which has been essentially repeated in other testing at UK-CAER, covers only a few days (after de-greening/stabilization of the cobalt FT catalyst), whereas the required lifetime for cobalt FT catalyst in a commercial application would need to be measured in terms of months, if not years. If it did prove to be possible for cobalt FT catalyst to operate acceptably in commercial practice at a syngas sulfur level of about 200 ppb, rather than being limited to only about 20 ppb, the syngas cleanup system would still be complex and expensive, but it might not need the additional capital cost, complexity, energy consumption, maintenance, etc., of guard beds.

Whether or not the cost saving from eliminating the guard beds would be sufficient to accept the risk of the potentially destructive effects on the cobalt FT catalyst charge of unanticipated gasification upsets, without the guard beds in place (as last-resort-type safety system), would also need to be determined.

Nevertheless, the laboratory data in Figure 20 suggest that a strategy different than the one used by Syntroleum, namely absolutely minimizing syngas sulfur level, could be viable and thus also worthy of testing further with actual coal-derived syngas. The intent here is neither to criticize, nor simply to repeat, the work that Syntroleum has already

done, but rather to learn from and build upon the previous work, and thus be in a better position to define logical “next steps” that could advance the state of the art, even if these next steps might include experiments that would be significantly (and perhaps prohibitively) more complicated, expensive and disruptive to perform.

The project team sought to further validate and expand upon the type of comparison presented above, but at significantly higher, known, and controllable, levels of hydrogen sulfide in the coal-derived syngas. This would almost certainly require that the clean-up system for coal-derived syngas be operated at different, stable levels of intensity and clean-up effectiveness, resulting in different stable levels of syngas sulfur concentration.

Depending upon the design of the basic syngas cleanup system, it might be possible to achieve one of the target syngas sulfur levels by simply removing or by-passing the guard beds used previously ahead of the “live” syngas CSTR. However, many additional factors, including coal characteristics, and variations in operating conditions caused by normal, and potentially abnormal, changes in plant-wide processing variables, could cause large variations in syngas sulfur levels over the test-run time period.

The ideal follow-on test program would be to make at least one (and preferably more) approximate-two-month CSTR run(s), each with fresh cobalt FT catalyst, covering a range of stable syngas H₂S levels, at the same (Tennessee Eastman), or possibly at another, continuously operating coal-gasification site. Making arrangements with a coal-gasification site to conduct such a test program would be a significant challenge for the project team under any circumstances. The project team’s focus was instead shifted to considering the sulfur on the cobalt FT catalyst at the nanoscale as a result of the Peer Review described next.

2.2. Peer Review 2009 and Follow-up Laboratory Work at UK-CAER

The 2009 Peer Review was a mandatory NETL-sponsored project event in February 2009. Outside experts were contracted by NETL to review projects and make recommendations. The Peer Review required both significant preparation and extensive follow-up, both of which added administrative and scientific/technical tasks to the project without providing any additional funding. Furthermore, a Peer Review recommendation resulted in the addition of a Milestone to the project, a significant laboratory-based analytical task which was completed by UK-CAER between mid 2009 and mid 2010.

The new Project Milestone was stated as: “The project needs to use modern nanoscale instrumentation, such as Auger and Raman spectroscopy, to show the amounts and sites of poisoning” of the cobalt FT catalyst by sulfur. The report on the results of this added Milestone (Reference 4) includes analytical results for “used” cobalt FT catalyst samples that had been exposed to syngas containing sulfur, with subsequent analyses of sulfur on the catalyst obtained by electron microscopy, energy dispersive X-ray analysis and electron energy loss spectroscopy. These techniques were used because it was determined that they would provide the highest possible resolution and sensitivity obtainable with today’s technology for this added task (it was also determined that Auger

and Raman spectroscopy, referred to by a Peer Review Committee member in the new Milestone's statement, would not be adequate).

The results of this additional testing were interesting (Reference 4), however, the project's focus was still the operation of laboratory CSTRs at a coal-gasification site. Furthermore, work done on the additional Project Milestone consumed a significant portion of the project's financial, scientific, technical, and managerial resources, without contributing materially to the project's original goals and objectives.

2.3. Goals for Follow-on Field-Test with Controllable Sulfur Level of Coal-Derived Syngas

ICRC and project-partner UK-CAER considered a range of potential options for the installation and operation of a laboratory CSTR at a coal gasification site where a controllable syngas-sulfur-level could be obtained. A continuous-operation commercial site would be by far the most desirable because it would offer the major advantage of access to virtually uninterrupted "live" syngas, but the only two such sites in the US, Tennessee-Eastman and Dakota Gasification, both declined overtures from the project team to participate in a project of this type.

The project team also made contact with the managers of other potential coal gasification sites. In the early stages of the project, Rentech was in the process of installing a small-scale gasifier system for commercial-type long-term operation within their Product Demonstration Unit (PDU) in Denver (<http://www.rentechinc.com/pdu.php>), and expressed interest in working with the project team on this task. However, after the gasifier system was installed, it would not operate acceptably on coal (despite its OEM rating), so the system was ultimately removed from the PDU and replaced by a natural-gas reformer system so that the rest of the pilot FT plant could still operate.

Other research-oriented gasifier sites were considered, but these could operate for only relatively short periods (typically about a week), between shutdowns of comparable duration, thus requiring compression and storage of coal-derived syngas and its low but critical (for this project) level of contaminants, with attendant difficulties and uncertainties thus introduced into the CSTR results.

To provide interpretable data, a CSTR must run virtually continuously (interrupted only by rare events, such as power outages, for example) for up to several months at a time. Therefore, it must have a continuous supply of syngas. If the coal gasifier is subject to shutdowns, syngas must be stored to "cover" the gasifier downtime. Storage of coal-derived syngas, with its low but non-zero level of contaminants, brings with it several additional issues that can affect the ultimate interpretability of the CSTR data.

An additional syngas compression step, with the potential to remove or modify some of the contaminants, would be necessary to reduce storage volume. Then there is the potential for interaction, over time, between the surfaces of the storage pressure-vessel with both:

- The low-level of contaminants remaining in the now-compressed, coal-derived syngas after clean-up, possibly lowering their level further, and
- With the compressed syngas itself, such as the formation of iron carbonyl, a known cobalt FT catalyst poison, from the interaction of CO with an iron-containing surface.

Note that the potential for low levels of contaminants in the syngas to be removed or modified during compression or storage would be much less of an issue if the goal were simply to repeat the work done previously at Tennessee Eastman in which guard beds were used ahead of the FT reactors to absolutely minimize sulfur contamination. Loss of syngas-sulfur contamination during compression or storage would not significantly influence the ultimate performance of the cobalt FT catalyst operating on the stored syngas, since the “lost” sulfur would have been removed by the guard beds anyway.

For the reasons cited above, arrangements could not be made with a suitable coal-gasification site that could provide un-interrupted “live” coal-derived syngas, with a higher, controlled and consistent level of sulfur contamination, for a sufficient time interval to obtain interpretable CSTR data on cobalt FT catalyst performance and deactivation. Furthermore, the budgeted funds originally intended for this testing were consumed in meeting the requirements of the new Milestone added to the project as a direct result of the 2009 Peer Review. Therefore, the follow-on testing of cobalt FT catalyst with coal-derived syngas at higher, controlled sulfur level could not be conducted.

3.0 PROJECT HISTORICAL INFORMATION

When this overall coal-to-liquids project began with the award of the NETL cooperative agreement in March 2006, ICRC’s primary partner was Syntroleum. ICRC and Syntroleum were then wrapping-up together a similar, joint, NETL-sponsored FT fuel production and demonstration gas-to-liquids project based on pipeline natural gas feedstock that had been underway for almost five years (since 2001). Syntroleum’s primary role in the new project was to provide both the full range of technical expertise necessary to integrate cobalt FT catalyst technology with coal-derived syngas, and to provide the required project cost-share.

3.1. The Textbook Approach to the Coal-To-Liquids Research Project

This new coal-based research project was structured with a virtual “textbook” approach:

- Laboratory experiments at the higher pressure and levels of reversible contaminants (i.e. not including sulfur) typical for coal-derived syngas would provide additional cobalt FT catalyst performance data (beyond the existing natural-gas-based data);
- The new data would be incorporated into existing cobalt FT catalyst synthesis process models to extend their range, and;

- Long-term field-data obtained at a coal-gasification site would be used for comparison and validation.

In addition, the cost effectiveness of various approaches for conditioning of coal-derived syngas, in terms of contaminant removal and adjustment of the H₂/CO ratio, would also be evaluated, and additional tasks would be performed, all with the ultimate goal of efficiently integrating cobalt FT synthesis catalyst technology with coal-derived syngas on an industrial scale.

As had been typical during the full duration of several previous government-sponsored joint projects with ICRC, Syntroleum began with diligent effort in all aspects of the new coal-based project. A new Syntroleum in-house catalyst laboratory with advanced state of the art capabilities was being brought on-line, and as described earlier, significant progress had been made in installing the required equipment for measurement of cobalt FT catalyst performance and deactivation at the Eastman Tennessee coal-gasification site.

3.2. A Change in Syntroleum's Corporate Direction

However, a change in Syntroleum's corporate direction became apparent in early September 2006 with the sudden closure of Syntroleum's natural-gas-based Catoosa, Oklahoma demonstration FT plant and the lay-off of almost half the company's employees. Syntroleum's explanation for this at the time was that the US DoD was no longer willing to pay the relatively high cost of the synthetic jet fuel produced at the Syntroleum demonstration plant that was being used in the DoD Assured Fuels Initiative, but instead DoD would purchase the FT jet fuel needed for ongoing military aircraft certification from lower-cost foreign suppliers. Since most of the Syntroleum engineers that ICRC had worked with for years previously were no longer employed by Syntroleum, official liaison with ICRC going forward was to be handled by Syntroleum's legal department.

Despite these ominous signs, Syntroleum assured ICRC that they planned to proceed with the project as originally planned, primarily because the project was focused on coal as the feedstock for FT fuel production. The Catoosa demonstration plant had proven Syntroleum's cobalt FT catalyst technology using (expensive) natural gas, but coal represented (according to Syntroleum at the time) a much more abundant and affordable feedstock.

Syntroleum had been working with Tennessee-Eastman since at least early 2006, and had kept both ICRC and Syntroleum shareholders generally informed of progress. The goal was to install and operate a continuous stirred tank reactor (CSTR) with cobalt FT catalyst using "live" coal-derived syngas to obtain catalyst performance data for comparison to, and validation of, laboratory catalyst performance data. Plans for obtaining this data at a then un-named commercial coal gasification site were also referred to publically in the abstract of a Syntroleum presentation at the Pittsburgh Coal Conference in September 2006 (Reference 1).

Despite the assurances to the contrary through the fall of 2006, Syntroleum's CEO notified ICRC's CEO by letter that Syntroleum had decided to "withdraw from participating" in the NETL/ICRC project in December 2006 (See Appendix). The letter made several noteworthy points:

- Although Syntroleum had worked successfully with ICRC in the past on broad-range research projects such as the current coal-based project, Syntroleum's focus going forward would be limited to providing the catalyst technology for commercial projects.
- Dakota Gasification had refused to act as the host coal-gasification site.
- The cost of constructing a new mobile laboratory facility had increased beyond the original budget. (Although not stated specifically in the letter, Syntroleum had previously informed ICRC that their laboratory CSTR equipment had been installed instead within an existing laboratory-type facility at the Tennessee-Eastman coal-gasification site.)
- Syntroleum had already taken several steps to prove its cobalt FT catalyst with coal-derived syngas for commercial application, and
- The possibility of future collaboration between ICRC and Syntroleum remained open.

The entire suite of bench-scale equipment required to operate Syntroleum's CSTRs had already been installed at Tennessee-Eastman under this project at the time of Syntroleum's notification in December 2006, as shown by the dated photographs included in Reference 2.

3.3. Restructuring of the Project

ICRC informed NETL immediately of Syntroleum's decision to withdraw from participating in the project. During telephone communications between ICRC and NETL it became clear that NETL favored a restructuring of the project, if possible, probably with additional partners, such that the ICRC project team could still carry out the same basic project with a continuing focus on integrating cobalt FT catalyst technology with coal-derived syngas.

Since Syntroleum would no longer be willing or able to provide the full range of technical support and expertise, other project-partners would be needed. For example, a university-based research organization (such as UK-CAER) could conduct the laboratory evaluations of cobalt catalyst performance with various levels of the contaminants found in coal-derived syngas, and industrial partners (such as EnviRes LLC and Worley Parsons) could evaluate the cost and effectiveness of various approaches to clean-up of coal-derived syngas for integration with FT synthesis using cobalt catalyst technology. Other partners (such as Smith Management) could evaluate the environmental aspects of coal-to-liquids technology and the costs associated with complying with all applicable environmental standards.

As ICRC, guided by NETL, was attempting in early 2007 to restructure the project as described above, ICRC maintained an awareness of Syntroleum's activities, and even held discussions with Syntroleum regarding an ongoing role in the project, albeit in a modified capacity going forward.

These discussions led to an invitation (See Appendix) from ICRC to Syntroleum to continue their activities at Tennessee-Eastman, intended to demonstrate commercial applicability of cobalt FT catalyst to coal-derived syngas, within the context of the original project, and thus Syntroleum could be eligible for reimbursement of some of their operating costs of this task going forward. At the same time, ICRC was very actively pursuing additional possibilities for partnerships with both UK-CAER and EnviRes, as described above, to undertake other project tasks suited to their expertise.

For more than a month during the critical project restructuring period, Syntroleum did not respond to ICRC's invitation, despite repeated attempts by ICRC personnel, including by ICRC's CEO, to elicit a Syntroleum response. The event that finally provoked an informal, but generally positive response from Syntroleum, by telephone and email, to ICRC's invitation was ICRC's making direct telephone contact with Tennessee-Eastman, as recommended by NETL, to propose that ICRC and Tennessee-Eastman could work together on the task at hand, with or without Syntroleum. In that conversation, Tennessee-Eastman confirmed to ICRC that they were indeed working with Syntroleum, and rejected the offer to work directly with ICRC.

Syntroleum's formal response was eventually received by ICRC several days later (See Appendix). This response enumerated several conditions that Syntroleum would demand if they were to return to participating actively in the project, including that some aspects of the data obtained could not be made public under this project for at least five years.

During the time-period of Syntroleum's non-responsiveness in the spring of 2007, ICRC was, of necessity, working closely with UK-CAER and EnviRes, attempting to restructure the project tasks and budget, including the required level of project cost-share that would be required from these additional partners. ICRC was led by events to conclude that, within the budget of the restructured project, it would not be in the project's nor the government's best interest to maintain Syntroleum's financial participation as an ongoing subrecipient.

As ICRC was working to restructure the project to fit within the additional constraints resulting from increasing the number of members on the project team, Syntroleum was continuing to conduct the same evaluation of cobalt FT catalyst in the CSTRs at the Tennessee-Eastman coal gasification site as originally planned under this project.

References

1. LeViness, Steve, Robota, Heinz, and Espinoza, Rafael, "Use of Cobalt Fischer-Tropsch Catalysts with Coal Derived Synthesis Gas," Pittsburgh Coal Conference, September 11-14, 2006, Session 8.
<http://www.engr.pitt.edu/pcc/2006/2006%20Abstract%20Booklet%20Final.pdf>
2. Agee, Ken, "Syntroleum Fischer Tropsch Long Term Catalyst Testing Using Live Coal-Based Syngas," Gasification Technology Conference, San Francisco, CA, October 17, 2007.
<http://www.gasification.org/uploads/downloads/Conferences/2007/58AGEE.pdf>
3. Syntroleum Corporation, Media Resources, Articles, "Fischer Tropsch Catalyst Test on Coal-Derived Synthesis Gas," November 2007.
<http://www.b2i.cc/Document/2029/110913.pdf>
4. Quarterly Progress Report, Project Number DE-FC26-06NT42449, "Production of High-Hydrogen Content Coal-Derived Liquids," Covering the period April 1, 2010 – June 30, 2010.

Appendix

Letter from Jack Holmes, CEO Syntroleum, to Jim Lexo, CEO ICRC, dated December 8, 2006.

Letter from Steve Bergin, ICRC, to Michael Taubman, Senior Counsel and Government Affairs Manager, Syntroleum, dated February 23, 2007.

Letter from Michael Taubman, Senior Counsel and Government Affairs Manager, Syntroleum, to Steve Bergin, ICRC, dated April 20, 2007.

Syntroleum

December 8, 2006

VIA FEDERAL EXPRESS

Mr. Jim Lexo, President and CEO
Integrated Concepts & Research Corporation
1033 North Fairfax Street, Suite 400
Alexandria, VA 22314

Re: *DOE/NETL Cooperative Agreement (DE-FC26-06NT42449) for "Production of High Hydrogen Content Coal-Derived Liquids"*

Dear Jim:

Syntroleum has enjoyed the opportunity to work with ICRC on several research projects over the years, examining many applications of our technology and its products in the marketplace. Our business development, engineering and research teams have been diligently working for months now to advance our commercial opportunities around the globe. At this time, Syntroleum must commit our personnel to achieve the business goals of our commercial efforts, so Syntroleum will withdraw from participating in this project effective immediately.

We've encountered several problems along the way on this research project, with the cost increases to construct the proposed mobile facility and the refusal by Dakota Gasification to act as our host site. In the time between our first discussions on this project and today, Syntroleum has taken several steps to prove our proprietary cobalt-FT catalyst for our commercial CTL business. Syntroleum would be taking a step back in the evolution of its commercial business by starting this project now and repeating work we've already performed at our own expense.

We appreciate ICRC's work in identifying research projects for our joint participation. I believe that Syntroleum and ICRC have done great things together, and I hope we can continue to do so in the future. Please do not hesitate to call me if you have additional questions about this matter or any other.

Sincerely,

John B. Holmes, Jr.
President & Chief Executive Officer

cc: Ken Roberts, Syntroleum
Michael Taubman, Syntroleum
Steve Bergin, ICRC

Advanced Vehicle Technologies
41150 Technology Park Drive
Suite 101
Sterling Heights, MI 48314
Tel: 586.799.1800
Fax: 586.991.0955

February 23, 2007

Mr. Michael Taubman
Senior Counsel and Government Affairs Manager
Syntroleum Corporation
4322 S. 49th W. Ave.
Tulsa, OK 74107

Dear Mr. Taubman:

Congratulations on your recent reveal in Washington. We applaud your continued clean fuels and energy security initiatives.

As we discussed on the phone earlier this week, Integrated Concepts and Research Corporation (ICRC) and Englobal Technical Services (ETS) have been in recent discussions regarding the DOE-NETL-ICRC "Production of High-Hydrogen Content Coal-Derived Liquids" program. DOE-NETL's willingness to modify the program seems genuine and ETS is working with ICRC to arrive at an acceptable solution for all participants. Since Syntroleum has already been working toward the commercial application of F-T technology to produce liquid fuels from coal, ICRC and ETS welcome Syntroleum's consideration of the following as a means for Syntroleum to participate in the project, and be compensated for participating, without disrupting other business priorities. Syntroleum's participation could include some coal-to-liquids project work already done, if, for example, that work was done after the project's cooperative agreement was awarded to ICRC on March 30, 2006.

FT Production Results—It is proposed that Syntroleum provide the project a summary with regards to general performance of the Syntroleum cobalt catalyst as functioning with cleaned synthesis gas derived from coal. This could include: conversion ratios, activity, etc. Syntroleum's requirements for intellectual property protection will be fully safeguarded and continuously maintained. Some negotiation, including DOE-NETL's input as the ultimate customer, may be required to arrive at a summary of results that would meet the requirements of all parties.

Finished Product—It is proposed that Syntroleum provide 50 gallons (or other amount TBD) of Fischer-Tropsch (F-T) fuel via cleaned synthesis gas derived from coal; F-T reactors (fixed tube or continuous stirred) and Syntroleum's cobalt catalyst. The fuels could include: S-2 and S-8. Syntroleum would be paid a negotiated price for said fuel(s). An alternative approach would be for Syntroleum to provide, at a lower negotiated price, F-T wax derived from actual coal-based syngas, with subsequent upgrading to finished fuels being done by a mutually agreeable third party.

Cost Share—A key element of the “original” program was cost share. This also needs to be an aspect of a modified program. It is possible that this could include (Based on successful Syntroleum work): equipment purchased/leased for the production of FT product; man-hours for the engineering and design work; operational expenses; etc., for the production of Fischer-Tropsch product via cleaned synthesis gas derived from coal; F-T reactors (fixed tube or continuous stirred) and Syntroleum’s cobalt catalyst. It is anticipated that the above could provide at least \$500,000 in cost share.

Process Component Evaluation—Should Syntroleum want to evaluate commercial process components (e.g., syngas guard beds, FT product filtration, etc.), those areas can be cost supplemented for a defined period.

Facilities—It is proposed that Syntroleum assist ICRC and Englobal in organizing a visit to the facility where the FT product was made via cleaned synthesis gas derived from coal. During our discussions DOE-NETL strongly recommended that we make arrangements to visit David Denton at Eastman Gasification Co. to determine their interest in participating in the project.

We appreciate Syntroleum’s desire to move swiftly to commerciality and to optimize available human and financial resources. It is anticipated that the modified program with DOE-NETL will both advance a commercial CTL facility sooner, rather than later, and will mitigate Syntroleum’s intellectual property risks and significantly reduce its resource requirements.

Thank you again for your time and consideration.

Sincerely,

Steve Bergin
ICRC Alternative Energy Program Manager

Syntroleum

April 20, 2007

Mr. Steve Bergin, Alternative Energy
Program Manager
Integrated Concepts and Research Corporation
41150 Technology Park Drive, Suite 101
Sterling Heights, MI 48314

Re: Invitation to participate in the *Production of High-Hydrogen Content Coal-Derived Liquids (DE-FC26-06NT42449)* program (the "Program") with Integrated Concepts and Research Corporation ("ICRC") and the U.S. Department of Energy, National Energy Technology Laboratory ("DOE-NETL").

Dear Steve,

I appreciated receiving your correspondence re-inviting Syntroleum Corporation to participate in the Program. As you know, we previously informed ICRC that Syntroleum intended to withdraw from participation in this Program to commit our personnel to achieve our commercial efforts. I presented your proposal to our management, and, upon review, they approved soliciting our Engineering and Research & Development groups for their recommendations.

At substantial private expense, Syntroleum developed its proprietary, cobalt catalyst for use in the Fischer-Tropsch ("FT") process. We have been demonstrating the Syntroleum cobalt FT catalyst for our commercial efforts in coal-to-liquids (CTL) with live coal-derived synthesis gas, accomplished by our work at the Eastman Chemical Company (Eastman) facility in Kingsport, TN. Syntroleum is currently conducting start up activities in order to commence a new demonstration run at Eastman, which is slated for approximately fifteen (15) weeks (~2,500 hours) of on-stream runtime. Syntroleum is willing to share the benefit of its knowledge with the Program, subject to our reaching agreement on certain conditions.

1. FT Production Results.

As part of ICRC's proposal, you requested that Syntroleum provide the Program a summary of the results with regard to the general performance of the Syntroleum cobalt catalyst functioning on coal-derived synthesis gas. Upon completion of the 2,500 hour catalyst testing run, Syntroleum proposes to share certain results of the program regarding catalyst performance, namely:

- the conversion ratio of the Syntroleum FT cobalt catalyst;
- the catalyst activity performance (deactivation) during the program; and
- if applicable, make general recommendations related to commercially available synthesis gas clean-up technologies.

These catalyst performance metrics will be analyzed, evaluated, summarized, and documented. Syntroleum requires that technical data provided to the Program by Syntroleum regarding the performance of the proprietary cobalt catalyst shall be treated as *Protected Data* in accordance with 10 CFR 600.325 and remain confidential, under the meaning of 5 U.S.C. 552(b)(4), for a minimum of five (5) years. A Topical Report summarizing the results will be prepared by Syntroleum for the Program.

In the Topical Report, Syntroleum would describe, in a block flow diagram (BFD), the inlet conditions (temperature & pressure) of the synthesis gas entering the system and the outlet results (products). Details about the Syntroleum FT process operating conditions would not be included.

Syntroleum receives the synthesis gas from Eastman on an "as-is, where-is" basis, at conditions suitable for Eastman's petrochemical manufacturing purposes. Syntroleum will include a description in the Topical Report of the gas pre-treatment steps (e.g. guard beds) taken by Syntroleum post-receipt of the synthesis gas from Eastman, but Syntroleum will not investigate, nor require Eastman, to impart its proprietary synthesis gas production and clean-up conditions, occurring upstream of the Syntroleum synthesis gas pre-treatment effort.

2. Products from the Syntroleum® Fischer-Tropsch process.

Syntroleum expects that the current catalyst testing program shall yield approximately forty pounds (~40 lbs.) of FT raw products post-filtration to remove catalyst fines (if any), Syntroleum proposes to deliver to ICRC approximately 40 lbs. of FT raw products produced from coal-derived synthesis gas during the catalyst performance testing. Syntroleum shall provide limited analysis on the FT raw products, in order to prepare information for the Topical Report related to the conversion ratio metric. Syntroleum requires that technical data provided to the Program by Syntroleum regarding the performance of the proprietary cobalt catalyst shall be treated as *Protected Data* in accordance with 10 CFR 600.325 and remain confidential, under the meaning of 5 U.S.C. 552(b)(4), for a minimum of five (5) years. Syntroleum proposes to share results of the current program regarding the FT raw products, namely analyses of:

- gravimetric and volumetric results;
- Schulz-Flory alpha (α) product distribution; and
- sulfur content in-situ in parts per million (ppm).

If requested by ICRC and sufficient raw FT products and funds are allocated to the effort to cover all costs and allow for hydro-processing, Syntroleum will contract with a third party laboratory (e.g. PARC or SwRI) to refine the FT products into synthetic diesel (Syntroleum® S-2) for further evaluation by ICRC. Syntroleum estimates that, using its Synfining® process conditions, 40 lbs. of FT products would convert to approximately 4.8 gallons of Syntroleum® S-2 synthetic diesel and 2.2 gallons of FT Naphtha (Syntroleum® S-1). The costs of this refining effort, both Syntroleum direct and third party costs, would be in addition to the amount requested for the catalyst performance testing. Further, Syntroleum anticipates that third party refining would require an up-front payment equal to 20% of the total estimated cost to cover the Syntroleum direct and third party cost to contract for production.

3. Syntroleum cost share & cost reimbursement.

In reaching this point of our efforts at Eastman, Syntroleum has incurred almost one million dollars (> \$995,000) in Syntroleum direct costs and third party costs, through December 31, 2006. Syntroleum would apply its prior costs incurred as the Syntroleum cost share basis required by the Program in the contract.

The current Syntroleum program for the 2500 hour catalyst testing run is estimated to cost approximately eight hundred seventy thousand dollars (~\$870,000) in Syntroleum direct costs and third party costs. Syntroleum proposes that the Program would reimburse Syntroleum for its actual expenses incurred, in exchange for sharing the results of its catalyst performance metrics with the Program.

4. Facilities.

Syntroleum would request, on behalf of ICRC and/or its designee (i.e. EnGlobal), to have a tour of the Eastman facility, subject to Eastman's approval and provided the tour is limited to the operating laboratory for the Syntroleum catalyst program.

April 20, 2007

Page 3 of 3

If our proposal to participate in the current Syntroleum catalyst testing program is of interest to ICRC, please let me know. Again, on behalf of Syntroleum, I appreciate the re-invitation from ICRC to participate in this Program. I believe that this is an opportunity for Syntroleum and ICRC to advance research and acceptance of Syntroleum Fischer-Tropsch technology. Please do not hesitate to call me if you have additional questions about this matter or any other.

Sincerely,

A handwritten signature in black ink, appearing to read "Michael P. Taubman", written in a cursive style.

Michael P. Taubman

Senior Counsel & Government Affairs Manager

cc: Gary Roth, Syntroleum
Crystal Green, Syntroleum
Jim Lexo, ICRC

APPENDIX 4
Performance Simulation of Syngas Cleanup,
CO₂ Removal and CO-Water-Gas-Shift

**Design and Performance Simulation of a Clean-up and Water-Gas-Shift
System for Syngas Derived from Coal and Biomass**

Milestone Report for Tasks 3 & 4

February 2010

Work Performed Under Cooperative Agreement No. DE-FC26-06NT42449

For

U.S. Department of Energy
National Energy Technology Laboratory
3610 Collins Ferry Road
P.O. Box 880
Morgantown, WV 26507- 0880

By

Donald P. Malone and Thomas C. Holcombe

EnviRes LLC
1500 Bull Lea Drive
Suite 116
Lexington, KY 40511

DISCLAIMER

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

ABSTRACT

This report describes activities for Task 3.0 (“Evaluation of impurity removal and/or syngas clean-up”) and Task 4.0 (“Technical analysis and economic evaluation of CO-water gas shift and subsequent CO₂ removal”) of the subject cooperative agreement. EnviRes selected WorleyParsons (WP) to perform integrated process engineering and cost estimation activities for both tasks. EnviRes provided WP with the syngas composition and conditions as a starting point for this work. WP generated process flow diagrams, a mass balance and an energy balance for two cases. The first case provided chemical-grade hydrogen at the battery limits ready for use in a refinery or chemical process. The second case provided syngas suitable and ready for Fischer–Tropsch synthesis (FT) with a ratio of hydrogen to carbon monoxide of 2.1 to 1.

TABLE OF CONTENTS

1. Executive Summary	1
2. Methods and Approach for Process Simulation	1
2.1. Background	2
2.2. Coal and Biomass Specification	3
2.3. Computational Simulation of Gasification Using FactSage	5
2.4. Process Simulation and Economic Evaluation	8
3. Results and Discussion	8
4. Conclusions	10
5. Graphical Materials	10
6. List of Acronyms and Abbreviations	13
7. Reference	15

1. Executive Summary

EnviRes and WorleyParsons generated an optimized gas clean-up and CO-water gas shift design and simulated its performance. This design used conventional process equipment in widespread use throughout the world at the size used in this design or larger. This system should be compatible with virtually any high temperature ($>1371^{\circ}\text{C}$) gasifier and any FT reactor system. The throughput of this system equated to 907.2 mt/d (1,000 t/d) with 20%w of the feed comprised of biomass.

The simulations used in this report indicated that FT catalyst poisoning would be substantially avoided or mitigated with this design. This design would cost little or no more than a design that would be considerably more vulnerable to catalyst poisoning. This design exhibited very low levels of conventional emissions such as SO_x , NO_x , CO and particulates. Nominally 97% of the carbon dioxide produced by the process was captured at a purity of 99.86%v. This design compressed the CO_2 to a pressure compatible with pipeline transport or geologic sequestration. Potential FT catalyst poisons from trace elements occurring in coal and biomass were simulated, studied and found to be of no significant threat.

This scale would best be characterized as a large demonstration or a small commercial facility in size. Clearly, larger gas clean-up and CO-water gas shift systems could be built with little risk. Several entrained flow gasifiers have been routinely operated on coal and petroleum coke at several times the throughput of this design. The largest commercial single train FT reactors employed by SASOL require approximately six times as much throughput as was used in this design.

The utility of this design lies in demonstrating emerging gasifier and FT designs with the capability of incorporating biomass as part of the feed. The FT reactors used by SASOL, described above, are 10.7 m in diameter. They are capable of producing nominally 3,000 mt/d (20,000 bbl/d) of liquid product. Designing complex pressure vessels of this size and larger is a daunting and expensive challenge. Such difficulties provide a strong incentive to find alternative technologies that can produce greater quantities of product from a smaller reactor vessel. SASOL uses substantially different gasification technology than is economically attractive for most opportunities in the USA.

If FT technology is to become a major factor in the supply of transportation fuels from coal/biomass feeds, it appears likely that not only will better FT technology be required, but better gasification technology optimized for FT production will also be required. Several of these improved gasifier and/or FT designs must be demonstrated at the scale of this design.

2. Methods and Approach for Process Simulation

While coal derived syngas clean-up, adjustment of hydrogen to carbon monoxide ratio and evaluation of FT processes are the focus of this project; we must start with a syngas that is representative of the gasifier outlet of a coal gasification process. The HyMelt® process developed by EnviRes can gasify nearly any carbonaceous feed that can be injected into the gasifier. The HyMelt process reacts feed and oxygen in a molten iron bath. The syngas

produced is similar to that from most entrained flow gasifiers using the same feed except that the HyMelt process produces very little CO₂ and consequently more CO compared to entrained flow gasifiers. Typically, a HyMelt gasifier effluent contains no more than 1% v CO₂, while entrained flow gasifiers typically contain 4% v to 10% v CO₂. The difference in CO₂ concentration between the two gasification processes results from the fact that HyMelt effluent compositions come closer to thermodynamic equilibrium than do the compositions for entrained flow gasifiers.

The feed composition and heating value must be specified before the syngas composition can be determined. For coal feeds, sulfur and ash content have a significant impact on the syngas composition. Trace impurities such as mercury, lead, cadmium, zinc, arsenic, chlorine and their oxides are volatile at gasifier temperatures. Ash constituents such as alumina, silica, calcium, sodium, potassium and titanium affect slag properties significantly. Flux agents added with the feed such as burnt lime can be used to adjust slag properties and capture some sulfur from the coal. EnviRes selected miscanthus giganteus (MG) as the biomass to be co-fed with coal. Properties for MG are given at <http://www.ecn.nl/phyllis/>, a database for biomass and waste.

EnviRes used a feed that was 80%w KY #11 coal and 20%w MG. It is possible to feed higher quantities of MG to the gasifier. Feeding more than 30%w of MG is difficult because it has a higher tendency to bridge (or “clog-up” during feeding) and increases CO₂ in the gasifier effluent. From a logistical standpoint, it may difficult to economically acquire enough MG to feed more than 20%w. These issues are no impediment to modeling higher levels of MG, but we chose the 20%w feed level as one with a reasonable chance of operational success without requiring extensive troubleshooting.

EnviRes determined the gasifier effluent composition and conditions using the feed described in this section and the FactSage thermo chemistry simulation program described below. Table 3 gives a summary of the results of this work. WorleyParsons used this information as the battery limits input for their process simulation and economic analysis for syngas treatment to produce a syngas suitable for input to a FT reactor.

Sulfur tolerant water gas shift catalysts require no syngas clean-up other than particulate removal, cooling to the proper temperature and addition of the appropriate amount of steam prior to the shift reactor. By using sulfur tolerant water gas shift catalysts all gas clean-up can be performed after the water gas shift reaction. Carbon dioxide produced from the water gas shift reaction constitutes the major contaminant removed by the gas clean-up section. Task #3 (Syngas clean-up) and Task #4 (CO-water gas shift and subsequent CO₂ removal) are closely interrelated and are treated as a single task for the purposes of this report.

2.1. Background

ICRC, a wholly owned subsidiary of VSE Corporation, entered into a cooperative agreement No. DE-FC26-06NT42449 with the National Energy Technology Laboratory (NETL) to study the integration of Fischer-Tropsch (FT) synthesis technology with coal-derived syngas. EnviRes LLC and the University of Kentucky Center for Applied Energy Research (CAER) are performing some of this work. Coal and Biomass gasification simulations were

employed to arrive at the starting gas composition and condition. The simulations used a total feed rate of 907.2 mt/d (1,000 t/d) of an 80%w coal 20%w biomass mixture.

2.2.Coal and Biomass Specification

EnviRes selected Kentucky No. 11 coal as the coal for the Basis of Design (BOD) because it represents one of the larger (in terms of reserves and production rate) high sulfur coal seams in Kentucky and the same seam is Illinois No. 6 north of the Ohio River. Together KY #11 and Illinois #6 represent one of the larger seams in terms of reserves of any seam east of the Mississippi River. The USGS Coal Database contains detailed analyses for 11 samples of this coal. The file KYTrace11.xls gives a comprehensive trace element analysis for these samples. Three samples were excluded because the ash content exceeded 19%. Such high ash coals are not good candidates for most high temperature gasification processes due to the increased thermal penalty of higher ash and increased slag handling expenses. Values for the remaining 8 samples were averaged to arrive at the composition and properties for the BOD coal. The file KY11AvgProx.xls gives selected trace elements, major ash constituents, the other coal elemental analyses, heat of combustion, ash fusion and Hargrove grindability index for these samples. The average moisture value for as received coal was 7.07%. Table 1 gives a summary of the proximate values.

Table 1. Summary of USGS Coal Properties for KY No. 11 Coal

SampleID	Moisture	Volitile Matter	Fixed Carbon	Standard Ash	Hydrogen	Carbon	Nitrogen	Oxygen	Sulfur	HHV BTU/lb	Slfate	Pyrite S	Organic S	Free Swelling Index	Ash Deformat ion Temperature, °F	Ash Softening Temperature, °F	Ash Fluid Temperature, °F
W225909	4.04%	39.51%	50.01%	6.44%	5.47%	71.12%	1.45%	12.24%	3.28%	12801	0.11%	1.11%	2.06%	3.00	2020	2080	2250
W229060	9.74%	37.34%	45.05%	7.87%	5.57%	65.13%	1.39%	16.78%	3.26%	11687	0.04%	1.32%	1.90%	2.00	2010	2050	2190
W229558	5.48%	41.22%	46.84%	6.46%	5.51%	70.09%	1.38%	12.90%	3.66%	12772	0.10%	1.58%	1.98%	4.00	2030	2080	2280
W229559	6.94%	42.66%	42.71%	7.69%	5.55%	66.15%	1.31%	14.27%	5.03%	12157	0.25%	2.93%	1.85%	3.50	2080	2110	2280
W232944	6.22%	35.12%	45.44%	13.22%	4.99%	62.70%	1.34%	12.81%	4.94%	11364	0.24%	3.56%	1.14%	4.50	2040	2070	2360
W233565	6.68%	40.76%	45.63%	6.93%	5.54%	68.64%	1.30%	13.50%	4.09%	12513	0.05%	1.88%	2.16%	3.50	1970	2070	2260
W233566	8.00%	38.85%	45.42%	7.73%	5.57%	65.95%	1.28%	15.49%	3.98%	12063	0.11%	1.80%	2.07%	3.50	2000	2070	2270
W233568	9.47%	35.60%	43.37%	11.56%	5.27%	61.20%	1.22%	16.51%	4.24%	11104	0.08%	2.69%	1.47%	3.00	1970	2040	2230
Avg	7.07%	38.88%	45.56%	8.49%	5.43%	66.37%	1.33%	14.31%	4.06%	12,058	0.12%	2.11%	1.83%	3.38	2,015	2,071	2,265

Coal as fed to the gasifier must be pulverized in a gas swept mill. This type of coal can easily be dried to 2% moisture in such a system¹. Drying the coal feed to a low moisture level has significant benefits for a dry feed gasifier. We used a value of 2% moisture for the coal fed to the gasifier. The elemental values for hydrogen and oxygen in the coal feed were adjusted to a moisture free basis. Moisture in the coal appears as a separate water stream. This facilitates combining the feed coal with other constituents that contain a different moisture level. The biomass tab of the file KY11AvgProx.xls gives the hourly feed rate for both the coal only case and the 80% coal-20% biomass case. Table 2 provides a summary of the trace elements used in this work.

Table 2. Total Elemental Composition for Coal and Miscanthus

	As Received	Coal @	Miscanthus	
	Coal@7.07%	2.0%	Dry Basis	5% Moisture
	moisture	moisture		
Carbon	67.17%	70.84%	48.42%	46.00%
Oxygen	15.81%	11.83%	43.93%	41.74%
Hydrogen	5.50%	5.19%	5.27%	5.01%
Sulfur	4.11%	4.33%	0.11%	0.11%
Nitrogen	1.35%	1.42%	1.06%	1.01%
H ₂ O	Included in H and O			5.00%
Fe	2.78%	2.93%	0.02%	0.02%
Si	1.68%	1.77%	0.17%	0.16%
Al	0.95%	1.00%	0.03%	0.03%
Ca	0.29%	0.31%	0.50%	0.48%
K	0.150%	0.158%	0.12%	0.12%
Ti	0.051%	0.054%	NA	
Cl	0.044%	0.046%	0.23%	0.22%
Mg	0.037%	0.039%	0.0561%	0.05326%
Na	0.029%	0.031%	6.566E-05	6.23737E-05
P	0.0069%	7.28E-05	0.022071%	0.020967%
B	0.0066%	6.96E-05	NA	
F	0.0053%	5.59E-05	NA	
Ba	0.0044%	4.64E-05	NA	
Zn	0.0039%	4.11E-05	0.004899%	0.004654%
Sr	0.0026%	2.74E-05	NA	
Rb	0.0017%	1.79E-05		
V	0.0016%	1.69E-05	0.000101%	0.000096%
Zr	0.0015%	1.58E-05		
Cr	0.0015%	1.58E-05	0.000939%	0.000892%
Ni	0.0011%	1.17E-05	0.000641%	0.000609%
Pb	0.00086%	9.07E-06	0.000152%	0.000144%
Nd	0.00086%	9.07E-06		
Ce	0.00080%	8.44E-06		
Pr	0.00076%	8.02E-06		
Br	0.00069%	7.28E-06		
Ge	0.00054%	5.7E-06		
As	0.00050%	5.27E-06	0.000076%	0.000072%
La	0.00043%	4.53E-06		
Sc	0.00026%	2.74E-06		
Dy	0.00017%	1.79E-06		
Se	0.00016%	1.69E-06		
Be	0.00013%	1.41E-06		
Os	0.00011%	1.19E-06		
Sm	0.00009%	9.49E-07		
Re	0.00008%	8.02E-07		
Tl	0.00008%	8.02E-07		
Cs	0.00007%	7.8E-07		
Ho	0.00005%	5.48E-07		
Hf	0.00005%	4.75E-07		
Cd	0.000012%	1.27E-07	0.000023%	0.000022%
Tb	0.000012%	1.27E-07		
Hg	0.000009%	9.49E-08	0.000009%	0.000009%

2.3.Computational Simulation of Gasification Using FactSage

FactSage is a robust multi phase thermodynamic simulator that has specific databases for the types of phases present in molten iron gasification. It is relatively easy to calculate equilibrium mixture compositions for a gas phase only system even if many impurities are present. When the system contains multiple liquid and solid phases such calculations become much more complex due to the very non-ideal behavior of key gasification species when dissolved in these phases. The FactSage databases contain interaction parameters for thousands of compounds and species that allow for these complex interactions to be determined. Additionally, FactSage uses a Gibbs Free Energy minimization module that facilitates equilibrium determination for a large number of species in multiple phases.

Equilibrium calculations for reactions taking place in a molten iron bath with a significant amount of slag are more complex even when the problems described above are resolved. The residence time of gaseous species in the reaction environment is on the order of a few seconds. The residence time of slag in the reaction environment is on the order of several hours. The residence time of metal in the reaction environment is on the order of days. What this means is that a calculation that assumes that all of the reactants are in equilibrium with all of the products may be seriously in error even if correct thermodynamic data and proper calculations are used. FactSage allows for the calculation of an “open” equilibrium. This means that with sequential feed injections the part of the feed that reports to the slag stays with the slag. The part of the feed that dissolves in the metal remains in the metal until or unless it reacts. The gaseous products thus formed are removed before the next increment of feed is added. If many feed steps are used, one usually finds that initially the composition of all three phases changes significantly from one injection to another. If a stable equilibrium is possible for the conditions used in the calculation, the change in the composition and in the amount of each phase from one injection to the next gets smaller until virtually no change takes place from one injection to the next. Typically each injection represents one hour of feed.

FactSage calculates the equilibrium composition of all three phases. It also determines if solid phases are present and their composition. This is important to the HyMelt process because the proper amount of burnt lime to be added with the feed to preclude the formation of a solid CaS phase or other solid phases can be calculated. The build up in the metal phase of chromium, nickel, vanadium and other trace metals in coal can be determined. Volatile salts and metals such as KCl, NaCl, Pb, Zn and others in the product gas can be quantified. This allows for the proper design of a quench system and perhaps down stream guard beds. We can determine the presence and composition of other liquid phases such as mattes (liquid metals such as iron or nickel with high levels of sulfur) that may affect refractory selection. Using FactSage, we can determine the fate of trace elements such as arsenic, lead, mercury, selenium and zinc that may have environmental impacts. Table 3 gives the amount and composition of the gas and dust in the gasifier effluent for 907.2 mt/d (1,000 t/d). EnviRes and WorleyParsons agreed to use a more conservative value of 1%v for CO₂ instead of the 0.396%v determined by FactSage to reflect HyMelt’s approach to equilibrium.

Mercury and other heavy metals are not shown in Table 3 because their levels are lower than the cut off concentration in the hot syngas for simulation, which was 1 ppmv. Lead was the heavy metal with the highest concentration in the hot syngas. Its concentration was 0.2

ppmv. Mercury was 6 ppbv. Such levels represent only a few hundred grams per hour. The initial water quench removes most of the heavy metals as particulate. Downstream processing precludes any of these metals existing in measurable concentrations.

Figures 1 and 2 show phase diagrams of slags that would likely be generated by a gasifier using the subject feed. Figure 1 is a tertiary diagram of the CaO-CaS-SiO₂ system. Figure 2 is a quaternary diagram of CaO-CaS-SiO₂ with a constant amount (10%w) of Al₂O₃. The isotherms are in °K. Diagrams such as these and others allowed us to determine optimum slag properties and flux addition rates that affect gas composition.

Table 3. Product Gas and Dust Composition

Page 40 - Step 40

Product Gas

Total gas flow 6792.78 lb mol/hr

	Mol wt	Mol fraction	Lb/hr
CO	28.010	0.67	126965.30
H ₂	2.016	0.31	4192.03
H ₂ O	18.015	6.64E-03	811.99
N ₂	28.014	5.79E-03	1101.97
H ₂ S	34.082	5.51E-03	1276.52
CO ₂	44.010	3.96E-03	1184.16
Ar	39.947	2.71E-03	735.61
CH ₄	16.043	9.33E-04	101.69
COS	60.076	4.19E-04	170.92
KCl	74.550	1.80E-04	91.10
HCl	36.460	8.32E-05	20.60
HCN	27.026	8.05E-05	14.78
HS	33.074	5.70E-05	12.81
KBO ₂	81.907	5.13E-05	28.57
H	1.008	2.65E-05	0.18
HF	20.006	2.36E-05	3.21
K	39.097	1.76E-05	4.68
NH ₃	17.030	1.44E-05	1.67
CS ₂	76.142	1.12E-05	5.82
HBO ₂	43.818	1.06E-05	3.16
S ₂	64.132	6.33E-06	2.76
C ₂ H ₂	26.038	6.28E-06	1.11
H ₂ CO	30.026	5.19E-06	1.06
NaCl	58.443	3.53E-06	1.40
CS	44.076	3.29E-06	0.98
KOH	56.105	2.72E-06	1.04
KCN	65.117	2.52E-06	1.12
CH ₃	15.035	2.00E-06	0.20
C ₂ H ₄	28.054	1.72E-06	0.33
Na	22.990	1.35E-06	0.21
H ₂ S ₂	66.145	1.16E-06	0.52
CHP	43.992	1.14E-06	0.34
HBO	27.817	1.12E-06	0.21

Dust

	lb/hr	wt. %
C	589.05	35.00%
S	76.58	4.55%
P	8.00	0.48%
Fe	925.65	55.00%
Zn	3.50	0.21%
Subtotal	1602.78	95.23%
Slag	80.22	4.77%
Total	1683.00	100.00%

Notes on gas composition

1) With exceptions noted below ignore any component

2) Assume CS, H & HS become CS₂, H₂ and H₂S

3) BaCl₂ may pose environmental problem

4) Pb, PbS, AsH₃, AsN, AsS, Cd, Hg, and PbH may remain totally or partially in the

5) For design purposes assume that the maximum H₂S level is 0.60% v, CO₂ is 1.0% v and Ar is 0.30% v

2.4.Process Simulation and Economic Evaluation

WP working with EnviRes developed a BOD (Basis of Design) for this study. The results of this study and the BOD can be found in the attached Appendix 5, Syngas Treatment Study Documentation, section of this overall report. The process simulation was performed using Aspen Plus as the simulation software. The Peng Robinson with Boston-Mathias modifications (PR-BM) equation of state was used to simulate streams where only the vapor phase was present. Where liquid water was in equilibrium with the vapor phase the ELECNTRL method was used. The Steam –TA method (ASME 1967 ASME steam table correlations) was used for pure water/steam. WorleyParsons defined the process scheme. This is described by the process flow diagrams (PFD) found in Appendix 5. The Aspen Plus simulator generated the heat and material balances based in input by WorleyParsons. These balances are included in Appendix 5. Justification of the processing technologies selected for the PFD and a comparison to those similar technologies not selected are given in Appendix 5.

Using the information described above, WP sized the major equipment for the PFD. Using vendor quotes and experience with similar equipment in similar environments WP developed the capital and operating costs for the CO-water gas shift and gas cleanup of syngas from the gasifier outlet to being ready to enter a FT reactor. This design includes process water clean-up, converting nearly all H₂S to elemental sulfur, capturing 90% of the mercury in the gasifier effluent, adjusting the temperature and pressure of the syngas to be ready for feeding a FT reactor, producing enough high purity hydrogen to facilitate wax hydrocracking and upgrading after FT reaction, capturing 95% of the CO₂ generated by the process and compressing it to 15.2 MPa (2200 psig) ready for injection into a CO₂ pipeline or for geologic sequestration.

3. Results and Discussion

WP developed a Front End Loading (common industry term) FEL level 2 plant design and economics for the Syngas Treatment Area of a coal/biomass to liquids plant with a nominal feed rate of an 80%w coal 20%w biomass feed input at 907.2 mt/d (1,000 t/d). This generated nominally $1.67 \times 10^6 \text{ Nm}^3/\text{d}$ (62.3 MSCFD) of hot raw syngas. The Syngas Treatment Area had a nominal output of $1.54 \times 10^6 \text{ Nm}^3/\text{d}$ (57.3 MSCFD) of feed suitable and ready for a FT reactor with hydrogen to carbon monoxide ratio of 2.1 to 1. Additionally $0.040 \times 10^6 \text{ Nm}^3/\text{d}$ (1.5 MSCFD) of 99%v hydrogen were produced for hydrotreating and upgrading of the FT product.

Other products included 1145 mt/d (1262 t/d) of 99.64%v CO₂ compressed to 15.2 MPa (2,200 psig), ready for pipeline injection and suitable for sequestration. Also produced were 15 mt/d (16.5 t/d) of Claus sulfur and 24.4 GJ/hr (23.1 MBTU/hr) of fuel gas. More than 90% of the mercury in the raw syngas was captured. The capital cost for the Syngas Treatment Section was \$147,016,000 based on US Gulf Coast construction and 2nd

quarter 2009 pricing. Major equipment capital costs are listed in Appendix 5 so that some “what if” studies could be performed as long as a reliable way of sizing or scaling the equipment is employed. Similar changes to the operating costs would also be necessary.

WP also ran a second case where as much of the syngas was converted to hydrogen as was practical. This case produced $1.42 \times 10^6 \text{ Nm}^3/\text{d}$ (52.9 MSCFD) of 99.48% v Hydrogen at 1.79 MPa (260 psig) from the same quantity of syngas as the FT case. The capital cost for the Syngas Treatment Section was \$168,209,000 based on US Gulf Coast construction and 2nd quarter 2009 prices. The major difference with the hydrogen only case was that it had increased CO-water gas shift reactors and a much larger pressure swing absorption (PSA) unit than did the FT case.

WP used a direct contact two step water quench to cool the raw syngas from 1,500°C (2732°F) to 205°C (401°F). The first step mixed water with the hot raw syngas flowing downward in a vertical pipe inside the quench vessel. The outlet of the pipe was below the water level in the vessel. The second step contacted the effluent gas from the first vessel with additional water in a Venturi scrubber followed by water disengagement in a trayed contactor vessel with a demister pad at the top. This arrangement accomplishes near total particulate removal while adding 90% of the water vapor required for the shift reactor and cooling the raw syngas to 205°C (401°F). The required inlet temperature for the shift reactor was 240°C (464°F). The heat necessary to get the syngas to the proper temperature was accomplished by heat exchange with the reactor effluent from the first shift reactor in a feed-effluent heat exchanger.

WP bypassed approximately 50% of the syngas being shifted around the first shift reactor, mixed the bypassed syngas with the cooled effluent from the first shift reactor, added steam and fed the mixture to the second shift reactor. WP did this because the syngas contained higher levels of CO than was typical for a shift reactor, thus the temperature rise across the reactor was very high. This arrangement affords an effective use of the shift reactor while minimizing the use of additional steam. For the hydrogen only case a third shift reactor was employed while for the FT case some of the quenched syngas bypassed the shift reactors completely and was subsequently mixed with the shifted syngas to attain the 2.1:1 H₂ to CO ratio. In either case the syngas was cooled to 38 °C (100°F), passed through a knock out drum, reheated to 43°C (110°F) and fed to a mercury removal adsorber. The subcooling and reheat was performed to preclude liquid water in the adsorber that markedly reduces its mercury adsorbing capability.

WP employed a Rectisol unit to remove H₂S and CO₂ from the shifted syngas. While Rectisol incurs significantly higher utility cost due to its lower operating temperature it offers several advantages not the least of which is its ability to remove H₂S and other sulfur compounds to less than 100 ppbv. Rectisol effectively removes COS without the need for hydrolysis. This is approximately the threshold of significant catalyst impairment from sulfur poisoning. Any commercial acid gas removal system for syngas currently in use employs a zinc oxide bed downstream of it to further reduce sulfur in the feed to a FT reactor. The use of Rectisol for acid gas removal reduces the consumption and change out frequency of zinc oxide. The use of Rectisol for acid gas removal also

mitigates the consequences of breakthrough in the zinc oxide bed. The consequences of sulfur poisoning for a cobalt FT catalyst are much greater than for an iron FT catalyst due to the much greater cost of the cobalt catalyst.

WP included a booster compressor after acid gas removal to ensure that the syngas would be available to the FT reactor at a sufficiently high pressure in the event that the system pressure drop became too high or that for some other reason the system pressure was too low. WP also included a small PSA unit to produce a small 99% hydrogen stream for upgrading the FT products by wax hydrotreating and perhaps some other hydroprocessing step.

The CO₂ stream recovered by the Rectisol unit had less than 0.15% v total impurities. WP included a compression system to boost the CO₂ to 15.2 MPa (2,200 psig). This compressor was not spared (i.e. provided with a back-up) assuming that the CO₂ could be vented in the event of nonavailability of the compression system. The system was expected to be on-stream at least 90% of the time. Higher CO₂ compression availability could be achieved by incorporating a spare train for this system.

4. Conclusions

Based on the work performed by WP and EnviRes the following conclusions can be reached:

1. An optimized gas clean up and CO-water gas shift system can be designed using components currently in use throughout the world at the size of this design or larger.
2. FT catalyst poisoning can be substantially avoided with this process design.
3. A high level of sulfur recovery with total sulfur emissions below minor source thresholds was demonstrated.
4. Virtually all CO₂ produced by the process was recovered at 99.86% v purity and compressed to pipeline pressure for subsequent use or sequestration.
5. Catalyst poisons such as COS, HCN, CS₂, Hg and other heavy metals were generated at such low levels that they should pose no threat to a FT catalyst when the current process scheme was used.
6. The process scheme described in this report affords significant flexibility for utilizing virtually any high temperature gasifier or FT reactor system.

5. Graphical Materials

CaO - CaS - SiO₂ - Al₂O₃
Projection (ASlag-liq), mole Al₂O₃/(CaO+CaS+SiO₂) = 0.00

Figure 1

CaO - CaS - SiO₂ - Al₂O₃
 Projection (ASlag-liq), mole Al₂O₃/(CaO+CaS+SiO₂) = 0.10

Figure 2

6. List of Acronyms and Abbreviations

Table of Acronyms and Abbreviations

\$	Dollars, US
%v	volume percent
%w	weight percent
°C	degrees centigrade
°F	degrees Fahrenheit
°K	degrees Kelvin
ASME	American Society of Mechanical Engineers
ASU	air separation unit
bar	10 ⁵ Pascal
BOD	basis of design
CAER	Center for Applied Energy Research-operated by the State of Commonwealth of Kentucky
cm	centimeter
CO	carbon monoxide
CO ₂	carbon dioxide
d	day
DOE	United States Department of Energy
eaf	electric arc furnace
EPC	engineering procurement contractor
FEL	Front End Loading – a design and cost estimating methodology
FT	Fischer - Tropsch
G	giga - 10 ⁹
gal	gallons, US
hr	hour
ID	inside diameter

IGCC	integrated gasification combined cycle
J	Joule – SI unit of energy
k	Kilo - 10^3
kSCF	thousand standard cubic feet
min	minute
mm	millimeter
MG	miscanthus giganteus
M	mega - 10^6
mt	metric ton, 1000 kg
NETL	National Energy Technology Laboratory
Nm ³	normal cubic meter @ 0°C and 1 ATM
NO _x	oxides of nitrogen
OD	outside diameter
Pa	Pascal, a unit of pressure
PFD	process flow diagram
ppbv	parts per billion, volume basis
ppm	parts per million, volume unless otherwise noted
ppmvd	parts per million, volume dry
PSA	pressure swing adsorption
psig	pounds per square inch, gauge
Rectisol	an acid gas removal technology developed by Lurgi that uses cold methanol
SASOL	South African Synthetic Oil Limited
SCFD	standard cubic feet per day
SO _x	Oxides of Sulfur
syngas	synthesis gas – a gas containing predominately CO and H ₂
t	short ton, US

UOP	a petroleum and chemical technology company formerly Universal Oil Products
USGS	United States Geologic Survey
USA	United States of America
WP	WorleyParsons

7. Reference

1. “Topical Report T-14, Coal Run No. 11 with Illinois No. 6 Coal and Kentucky No. 9 Coals in the Syncrude Mode” Choksey, P. J.; Harris, E. C.; Morrison, J. A.; Newby, D. V.; Richardson, F. E.; and Young, F. E.; Ashland Synthetic Fuels, Inc. under DOE contract DE-AC05-76ET10143, pp4-8 to 4-9, 1981.

Systems Integration Study

Introduction

The purpose of this supplemental report is to document that compatibility has been achieved in the individual designs of complex sub-systems designed separately, but ultimately intended to be fabricated, and then integrated into a high-value, industrial-scale system that can function seamlessly. This task is motivated by the realization that mistakes, miscalculations or interface incompatibilities are much less difficult and expensive to correct “on paper” than “in the field.” The goal is to avoid the stresses and disruptions associated with learning too late that extremely expensive, long-lead-time pieces of specialized industrial equipment are somehow incompatible, and therefore must be either “sent back for re-work” or modified on-site under circumstances that are always far less than ideal.

This research project is sponsored by DOE-NETL and is focused on the production of high-hydrogen content liquid transportation fuels, such as diesel and jet fuels, from coal and biomass. The NETL research project leverages both the technical expertise and the cost-share obtainable from a commercial project. The commercial project is being conducted by EnviRes, LLC, and is focused on the design, construction and operation of a demonstration-scale plant centered on the gasification of coal and biomass to produce synthesis gas, or syngas, and the production of downstream products, including Fischer-Tropsch (FT) liquid fuels.

The individual systems that ultimately need to be integrated are those used for:

- Syngas cleanup
- Water gas shift (to increase hydrogen/carbon monoxide ratio of the syngas)
- FT Synthesis, including both cobalt and iron FT catalyst systems
- Separation of hydrogen from coke-oven gas (an auxiliary method to produce hydrogen)

The first three of the foregoing systems are shown schematically in Figure 1. Note that the system for separating hydrogen from coke oven gas is not included in Figure 1. In accordance with “Amendment 01, Additional Scope, Interface Compatibility”, a review of Project Tasks 3 through 6 (which cover the systems noted above) of Subagreement Number ICRC-AVT-08-011 was completed.

Summary of Findings

A thorough review of the operating variables of the systems and their interfaces did indeed identify two significant interface incompatibilities. The first was a difference in assumed design operating pressure between units that would have been destined to be connected under a previously possible, but no longer likely, plant configuration. Hydrogen separated from coke-oven gas is no longer a likely component of the syngas intended to be used for FT synthesis. The site originally considered for a prototype FT plant to be built in the future by project-partner EnviRes, which is adjacent to an

operating coke oven, is no longer a viable site for the future phase of the commercial project that has provided both industrial expertise and cost-share to this research project.

Specifically, the system for separating hydrogen from coke oven gas was originally designed to deliver 17 million standard cubic feet per day (SCFD) of hydrogen at 375 psig. However, according to the design of the main-stream syngas conditioning system, this hydrogen would need to have been added to 57 million SCFD of syngas at a nominal pressure of 400 psig. This discrepancy in design operating pressures can now be rectified relatively easily since it has been recognized, and it would be acted upon long before any process units would be fabricated and delivered to the plant site for assembly.

The second more subtle finding of incompatibility was conceptual rather than purely physical, but it too could have caused serious problems if it had not been identified. Specifically, different assumptions were made by the designers of the combined syngas clean-up and water gas shift (WGS) systems, than the assumptions made by the designers of the downstream FT synthesis plant, with regard to the fate of CO₂ produced during the initial WGS. This subject has been treated in detail in Reference 1, the Report covering Task 6 of this project, the overall comparison of cobalt and iron FT catalyst systems, so it will be described below relatively briefly. However, it is instructive to consider all the factors that ultimately led to this second mostly conceptual, rather than purely physical, incompatibility.

Figure 1. System Schematic Diagram

Source of Conceptual Incompatibility

The root cause of the second, mostly conceptual, incompatibility referred to above derives primarily from the particular view one takes of the CO₂ that would inevitably be produced by an FT plant using coal as a feedstock. This view is significantly different in the US, or perhaps more precisely it is significantly different in the view of the US mass-media, than it is in most of the rest of the world. The politically correct US position on CO₂ is that its level in the atmosphere is already intolerably high, and therefore if one were to even consider using coal as a feedstock for FT synthesis, virtually all the CO₂ produced would need to be captured and sequestered as an absolute necessity.

The view in many other areas of the world is entirely different, especially given the recent realization in many quarters that petroleum is no longer the appropriate source for the additional, or marginal, quantities of middle distillate fuels (diesel and jet fuel) needed to meet their growing demand worldwide. The simple facts are:

- Petroleum currently (spring/summer 2011) costs about \$100 per barrel
- More than 3 marginal barrels of crude oil must be run through a refinery to produce 1 marginal barrel of valuable, and profitable, diesel & jet fuels
- On the “other 2 barrels” of products, the refiner can break-even at best
 - Gasoline is a minimal-profit, global-surplus by-product
 - Other bulk products (residual fuel oil, petroleum coke, asphalt, etc.) all sell for less than the original crude oil on a volume basis
- “Economics” of FT diesel and jet fuels from coal are no longer in doubt
 - Capital is already flowing to coal-FT projects outside the US
 - Capital is extremely unlikely to flow to US projects, due to the near certainty of protracted litigation over CO₂ emissions

The design philosophy behind the syngas clean-up and water gas shift systems of Task 3 and 4 of this project is aligned with the “politically correct” US view referred to above. Namely, the highest design priority is given to removing, and capturing for sequestration, virtually all the CO₂ produced by both the coal gasification process (relatively little CO₂), and by the initial water gas shift process (significantly more CO₂) required to increase the H₂/CO ratio of coal-derived syngas. As discussed in detail in Reference 1, the cobalt FT catalyst system has many advantages, including the fact that virtually all the CO₂ produced by the FT plant can be removed in a single acid-gas removal step “up-front” using the highly effective, but expensive, system designed under Tasks 3 and 4 of this project.

However, for many coal-FT plants currently being designed for installation and operation outside the US, the top priority is to minimize plant capital and operating costs, regardless of the effects on CO₂ emissions. In this case, it may make sense to use a less expensive syngas clean-up system without the capability of removing CO₂. Furthermore, the CO₂ produced by the water gas shift process can be “left-in” the syngas after shift. This CO₂ acts a diluent, which can have both beneficial and detrimental effects; these are covered in detail in Reference 1.

Briefly, the diluent CO₂ requires somewhat larger FT reactors, but for the cobalt system, which uses much smaller FT reactors than the iron system, this may be acceptable, especially since the diluent can also eliminate the need for any expensive recycle compressors for the cobalt FT catalyst system.

Further Considerations

It is important to note that erroneous conclusions could result if the results of this overall project are applied to significantly different circumstances (e.g. different technologies for individual process steps such as water gas shift or acid gas removal or different process schema). The results of comparing the intrinsic differences between cobalt- and iron-based Fischer-Tropsch catalysts when processing a coal-derived syngas depend greatly on the basis of the comparison.

The following variables should be considered when making comparisons:

1. Capacity. The current study was based on 1,000 tons per day of an 80/20 coal/biomass feedstock blend. Depending upon yields, this feedstock would produce 1,600 –2,200 barrels per day (bpd) of Fischer-Tropsch liquids. This is a relatively small capacity for a commercial facility. World-class Fischer-Tropsch plants usually have much larger capacities. Sasol, for instance, reports that a single Sasol reactor can produce as much as 20,000 bpd, while Shell claims that the maximum capacity of a single fixed bed reactor is up to 8,000 bpd. Economies-of-scale can have a major impact on investment returns and heavily influence priorities in a process comparison.
2. Type of Fischer-Tropsch reactor. Some types of Fischer-Tropsch reactors, such as slurry and fluidized bed reactors, tend to have good economies of scale at larger capacities, while fixed and packed bed reactors tend to work better at smaller capacities. As an extreme example, the capital cost of Velocys' microchannel reactors is virtually proportional to capacity. Furthermore, physical attrition tends to shorten catalyst life in slurry and fluidized bed reactors, thereby increasing the importance of the relative cost and attrition resistance of iron and cobalt catalysts.
3. Type of gasifier. The current study was based on gasification reactions taking place in a molten iron bath (the HyMelt[®] Process). The predicted syngas composition was significantly different from the composition produced by conventional high-temperature, entrained-flow, slagging gasifiers. The H₂/CO ratio was lower and the concentrations of oxide impurities were lower. In addition, since approximately one-half of the sulfur was removed by slag, the sulfur content (predominantly H₂S) was also lower.
4. Processes used to cleanup the syngas. This study showed that the use of the Rectisol process (using methanol as a physical solvent) was able to reduce the sulfur content of the syngas to less than 20 ppb. A zinc oxide guard bed was added prior to the Fischer-Tropsch reactor to protect against upsets and should reduce the sulfur concentration to below 10 ppb. This level of sulfur would minimize sulfur poisoning of iron or cobalt catalysts.

5. H₂/CO ratios. Coal-derived syngas typically has a H₂/CO ratio of around 0.5. The target H₂/CO ratio for feeding a cobalt-based Fischer-Tropsch reactor is typically about 2.1, slightly more than the stoichiometric requirement for producing paraffins. Thus, the syngas requires a significant amount of external water-gas shift. In contrast, the H₂/CO ratio feeding an iron-based Fischer-Tropsch reactor is typically about 1.0 due to the large amount of water-gas shifting that occurs in parallel with the Fischer-Tropsch reactions and therefore less shifting is required.
6. Carbon range distribution of the Fischer-Tropsch liquids. The objective of the current study was to maximum diesel and jet fuel production. Cobalt and iron catalysts can have significantly different yields of diesel/jet fuel, wax, oxygenates, etc., especially if the iron FT catalyst is used in the high-temperature FT process that could be used to produce high octane gasoline. Therefore, the desired product distribution can affect the preferred type of catalyst.

In conclusion, extreme care must be taken when extrapolating the results of this study to other Fischer-Tropsch applications.

Reference

1. Espinoza, Rafael, Kym Arcuri and Daniel Montoto, "Overall Comparison of Iron and Cobalt based Fischer-Tropsch Catalyst Systems for the Production of High-Hydrogen Content Coal-Derived Liquid Transportation Fuels" Milestone Report, Task 6, Project No. DE-FC26-06NT42449, March 2011. (Included in Appendix 7 of the Final Project Report.)
-

APPENDIX 5
Syngas Treatment Study Documentation

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

Final Report - Syngas Treatment Study

For EnviRes Coal/Biomass Gasification Project

53869201 – PCS-ICRC-0-LI-011-0001

18 Dec 2009

Power

2675 Morgantown Road,
Reading, PA 19607
USA

Telephone: +1 610 855 2000

Facsimile: +1 610 855 2384

www.worleyparsons.com

© Copyright 2009 WorleyParsons

WorleyParsons

resources & energy

EcoNomics

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

Disclaimer

This report has been prepared on behalf of and for the exclusive use of EnviRes LLC, and is subject to and issued in accordance with the agreement between EnviRes LLC and WorleyParsons. WorleyParsons accepts no liability or responsibility whatsoever for it in respect of any use of or reliance upon this report by any third party.

"Copyright © 2009 EnviRes LLC and WorleyParsons. This paper is written with support of the U.S. Department of Energy under Contract No. DE-FC26-06NT42449. The Government reserves for itself and others acting on its behalf a royalty-free, nonexclusive, irrevocable, worldwide license for Governmental purposes to publish, distribute, translate, duplicate, exhibit, and perform this copyrighted paper".

PROJECT 53869201 - FINAL REPORT - SYNGAS TREATMENT STUDY

REV	DESCRIPTION	ORIG	REVIEW	WORLEY- PARSONS APPROVAL	DATE	CLIENT APPROVAL	DATE
0	Issued for client review (Case 1)	JLH / JCM	J. White	M. DeLallo	8-July-09	N/A	
1	Issue for client review (Case 1 & 2)	JLH / JCM	J. White	M. DeLallo	2-Sept-09		N/A
2	Issue	JLH / JCM	J. White	M. DeLallo	30-Nov-09

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

CONTENTS

1.	EXECUTIVE SUMMARY	1
2.	INTRODUCTION.....	2
2.1	Purpose	2
2.2	Scope	2
2.3	Definition of Syngas Treatment.....	2
3.	DESIGN BASIS.....	4
3.1	Basis of Design	4
3.2	Feed and Product Summary	4
3.3	Block Flow Diagram	6
4.	PROCESS DESCRIPTION.....	9
4.1	Case 1 – Hydrogen Production.....	9
4.1.1	Case 1 - Syngas Quench & Scrubbing – Unit 21	9
4.1.2	Case 1 - Shift Reaction & Cooling – Unit 22	10
4.1.3	Case 1 - Acid Gas Removal – Unit 23	12
4.1.4	Case 1 - Hydrogen Purification – Unit 24.....	14
4.1.5	Case 1 - Sulfur Recovery – Unit 25	15
4.1.6	Case 1 - CO ₂ Compression – Unit 26	16
4.2	Case 2 – Fischer-Tropsch Liquids Production.....	17
4.2.1	Case 2 - Syngas Quench & Scrubbing – Unit 21	17
4.2.2	Case 2 - Shift Reaction & Cooling – Unit 22	18
4.2.3	Case 2 - Acid Gas Removal – Unit 23	20
4.2.4	Case 2 - Hydrogen Purification – Unit 24.....	22
4.2.5	Case 2 - Sulfur Recovery – Unit 25	23
4.2.6	Case 2 - CO ₂ Compression – Unit 26	24

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

4.2.7	Case 2 – Syngas Compression – Unit 27	24
5.	PROCESS FLOW DIAGRAMS	26
6.	HEAT & MATERIAL BALANCE	28
7.	DESIGN CONSIDERATIONS	29
7.1	General	29
7.2	Case 1 Design Considerations	29
7.2.1	Case 1 - Syngas Quench & Scrubbing	29
7.2.2	Case 1 - Shift Reaction & Cooling	30
7.2.3	Case 1 - Hydrogen Purification	30
7.2.4	Case 1 - CO ₂ Compression	31
7.3	Case 2 Design Considerations	31
7.3.1	Case 2 - Syngas Quench & Scrubbing	31
7.3.2	Case 2 - Shift Reaction & Cooling	31
7.3.3	Case 2 - Acid Gas Removal	32
7.3.4	Case 2 - Hydrogen Purification	32
7.3.5	Case 2 - CO ₂ Compression	33
7.3.6	Case 2 - Syngas Compression	33
8.	UTILITIES, CATALYST, CHEMICALS & EFFLUENTS	34
8.1	Utility Summary	34
8.2	Catalyst Summary	35
8.3	Chemical Summary	35
8.4	Effluent Summary	36
9.	PLOT PLAN	37
10.	COST ESTIMATES	39
10.1	Capital Cost	39
10.1.1	Purpose	39

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

10.1.2	Description.....	39
10.1.3	Principles of Estimate.....	40
10.1.4	Cost Estimate Methodology.....	41
10.1.5	Process Units.....	41
10.1.6	Contingency.....	43
10.1.7	Taxes.....	43
10.2	Operating Cost.....	44
10.2.1	Personnel Costs.....	44
10.2.2	Utility Costs.....	44
10.2.3	Catalysts and Adsorbent Media.....	45
10.2.4	Chemical Costs.....	45
10.2.5	Maintenance Costs	46
10.2.6	Disposal Costs	46
10.2.7	Miscellaneous Operating Costs	46
11.	RESULTS AND RECOMMENDATIONS.....	47

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

List of Figures

Figure 3-1	Case 1 – Block Flow Diagram	7
Figure 3-2	Case 2 – Block Flow Diagram	8

List of Tables

Table 3-1	Feed and Product Summary.....	5
Table 9-1	Utility Summary.....	34
Table 9-2	Catalyst Summary.....	35
Table 9-3	Chemical Summary.....	36
Table 9-4	Effluent Summary	36

Appendices

APPENDIX 1 -	BASIS OF DESIGN
APPENDIX 2 -	PROCESS FLOW DIAGRAMS
APPENDIX 3 -	HEAT & MATERIAL BALANCE
APPENDIX 4 -	PLOT PLAN
APPENDIX 5 -	CAPITAL COST
APPENDIX 6 -	OPERATING COST
APPENDIX 7 -	UTILITY SUMMARY BY UNIT BREAKDOWN

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

TERMINOLOGY

AGR Acid Gas Removal	H₂S Hydrogen Sulfide
ASU Air Separation Unit	HCN Hydrogen Cyanide
BFD Block Flow Diagram	Hg Mercury
BoD Basis of Design	H&MB Heat and Material Balance
BOP Balance of Plant	HP High Pressure
CCS Carbon Capture and Sequestration	HP Horsepower
CH₄ Methane	Hr, hr hour
Cl⁻ Chloride ion	ICRC Integrated Concepts and Research Corporation
CO Carbon Monoxide	ISBL Inside Battery Limits
CO₂ Carbon Dioxide	KO Knock Out
COS Carbonyl Sulfide	kW, kWe Kilowatt electric
CTL Coal to Liquids	LNW Liquid Nitrogen Wash
DOE Department of Energy (United States)	LP Low Pressure
DP/DT Design Pressure & Temperature	LTPD long tons per day
EOR Enhanced Oil Recovery	MMSCFD million std cu ft per day
EOS Equation of State	MDEA Methyl Diethanolamine
EPRI The Electric Power Research Institute	MM Btu Million Btu
FEED Front End Engineering Design	Mol% Molar percent
FEL Front End Loading	MP Medium Pressure
FT, F-T Fischer-Tropsch	N₂ Nitrogen
GPM, gpm gallons per minute	NaOH Sodium Hydroxide (caustic)
GTL Gas to Liquids	NETL National Energy Technology Laboratory
H₂ Hydrogen	NH₃ Ammonia
H₂O Water

WorleyParsons

resources & energy

EcoNomics™

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

NO_x Nitrous Oxides

O₂ Oxygen

O&M Operating and Maintenance

OPEX Operating Expenses

OSBL Outside Battery Limits

PFD Process Flow Diagram

P&ID Piping and Instrumentation Diagram

ppm Parts per Million

ppmv Parts per Million, volume

ppmw Parts per Million, weight

PSA Pressure Swing Adsorption

RWU Rectisol Wash Unit

S, S_x Sulfur (elemental)

S/DG Steam to Dry Gas Ratio

SIS Safety Instrumented System

SO₂ Sulfur Dioxide

SOR Start of Run

SRU Sulfur Recovery Unit

STPD Short Tons per Day

SG, Syngas Synthetic Gas

TBL Technical Battery Limits

TEG Triethylene Glycol

TGTU Tail Gas Treatment Unit

TIC Total Installed Cost

TPC Total Plant Cost

US, U.S. United States

USD, US\$ the United States Dollar

Vol% Volume percent

Wt% Weight percent

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

1. EXECUTIVE SUMMARY

WorleyParsons has developed two Front End Loading Level 2 (FEL 2) plant designs and economics for the Syngas Treatment Area of a plant facility that would process 62.3 MMSCFD of Hot Raw Syngas from an EnviRes HyMelt Molten Iron Gasifier. In Case 1 of the Syngas Treatment Study the syngas is processed to produce Hydrogen suitable for the production of ammonia or other uses. In Case 2 the syngas is processed to produce feed with a 2.1 H₂:CO ratio for Fischer-Tropsch Liquids Production.

In Case 1 (Hydrogen Production) the syngas from the EnviRes HyMelt Gasifier is processed to produce 52.9 MMSCFD of Hydrogen suitable for the production of ammonia or other uses.

Other principal products for Case 1 are 2,324 STPD (40.1 MMSCFD) of CO₂ compressed to pipeline pressure and suitable for sequestration, 15 LTPD of Sulfur and 101.5 MM Btu/hr of Fuel Gas. Over 90% of the mercury in the syngas will be trapped in the Mercury Removal Drum. The CO₂ product contains 95% of the carbon that was gasified into syngas.

In Case 2 (Fischer-Tropsch Liquids Production) the syngas from the EnviRes HyMelt Gasifier is processed to produce 57.3 MMSCFD of feed with a 2.1 H₂:CO ratio for Fischer-Tropsch Liquids Production and 1.5 MMSCFD of Hydrogen for hydrotreating the F-T liquids.

Other principal products for Case 2 are 1,262 STPD (21.8 MMSCFD) of CO₂ compressed to pipeline pressure and suitable for sequestration, 15 LTPD of Sulfur and 23.1 MM Btu/hr of Fuel Gas. Over 90% of the mercury in the syngas will be trapped in the Mercury Removal Drum. The CO₂ product contains 52% of the carbon that was gasified into syngas.

WorleyParsons' scope was limited to the 'Syngas Treatment' areas including Syngas Quench & Scrubbing, Shift Reaction & Cooling, Acid Gas Removal (AGR), Hydrogen Purification, Sulfur Recovery Unit (SRU), Carbon Dioxide Compression, and Syngas Compression (Case 2 only). Coal/biomass receiving and preparation, Gasification, Slag Handling, Air Separation Unit (ASU), Balance of Plant (BOP), and the hydrogen/syngas consumer (Ammonia/Fischer-Tropsch Plant) are outside of WorleyParsons' scope.

The Case 1 Capital Cost for the Syngas Treatment Unit is \$168,209,000 based on US Gulf Coast / Houston, Texas and 2nd Quarter 2009.

The Case 2 Capital Cost for the Syngas Treatment Unit is \$147,016,000 based on US Gulf Coast / Houston, Texas and 2nd Quarter 2009.

The US Gulf Coast was selected as the basis because cost estimates for many US and foreign locations are based on ratios from the US Gulf Coast cost. This provides a standard for adjusting the estimate for a location in Kentucky, or elsewhere.

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

2. INTRODUCTION

2.1 Purpose

ICRC, a wholly owned subsidiary of VSE Corporation, is under contract to the National Energy Technology Laboratory (NETL) and the State of Kentucky to conduct an assessment of coal to liquids and chemicals. The project is a team effort encompassing 13 specific technical tasks. EnviRes LLC is providing the gasification technology and has contracted WorleyParsons' assistance with the definition of the syngas treatment processes and equipment.

2.2 Scope

WorleyParsons herein provides technical analyses, process data, and cost estimating on the following core elements required for the integration of technology with coal-derived syngas:

Analysis of impurity removal, clean-up of coal-derived syngas, including identification of impurities and their expected concentration ranges; evaluation of removal strategies and the equipment required, including state-of-the-art equipment performance syngas modeling; prediction of equipment effectiveness and economics of impurity removal. (ICRC Task 3)

Technical analysis and economic evaluation of CO shift (also referred to as water gas shift) and subsequent CO₂ removal, including study of the types and characteristics of process equipment required, state-of-the-art equipment performance, syngas modeling, prediction of equipment effectiveness and economics. (ICRC Task 4)

WorleyParsons has developed a Front End Loading Level 2 (FEL 2) design/economics for ICRC Tasks 3 and 4 to convert all of a coal-derived syngas to high-purity hydrogen (with quality suitable for ammonia synthesis) (Case 1). As an adjunct to Task 3 and 4, WorleyParsons has also developed a Front End Loading Level 2 (FEL 2) design/economics to convert the coal-derived syngas into feed suitable for Fischer-Tropsch Liquids Production (Case 2).

2.3 Definition of Syngas Treatment

WorleyParsons has developed FEL 2 designs (reference ICRC Tasks 3 and 4) to convert all of a coal-derived syngas into either Case 1 - high-purity hydrogen (with quality suitable for ammonia synthesis) or Case 2 – syngas feed for Fischer-Tropsch Liquids Production. The gas composition to the syngas treatment area has been jointly determined by EnviRes and WorleyParsons.

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

WorleyParsons has documented a single design condition in the Basis of Design document (Appendix 1) for each Case. From this basis, WorleyParsons has performed the following:

- Defined a typical process scheme.
- Performed process simulation in order to determine the heat and material balance together with equipment size and the corresponding utility requirement based on WorleyParsons' experience, in-house know-how associated with data provided by licensors, and engineering judgment.
- Selected and defined syngas treatment technologies.
- Defined all major process and equipment, block flow diagrams (BFD), process flow diagrams (PFDs), and heat and material balances (H&MB).
- Developed syngas treatment plant performance estimates including consumable rates, production rates, and auxiliary power requirements.
- Developed basic engineering design data including major process descriptions, equipment list, and general equipment arrangement on site plot plan.
- Developed capital and operating and maintenance (O&M) cost estimates for the syngas treatment plant, to an accuracy of $\pm 25\%$, based on in-house and vendor estimates of major equipment items.

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

3. DESIGN BASIS

3.1 Basis of Design

See Appendix 1 for the Basis of Design (BoD) for the Syngas Treatment Study for the EnviRes Coal/Biomass Gasification Project.

3.2 Feed and Product Summary

FactSage is a robust multi phase thermodynamic simulator that incorporates specific databases for the types of phases present in molten iron gasification. It is relatively easy to calculate an equilibrium mixture for a gas-phase-only system even if many impurities are present. When the system contains multiple liquid and solid phases such calculations become much more complex due to the very non-ideal behavior of key gasification species when dissolved in these phases. The FactSage databases contain parameters for thousands of compounds and species that allow for these complex interactions to be determined. Additionally, FactSage uses a Gibbs Free Energy minimization module that facilitates equilibrium determination for a large number of species in multiple phases.

Equilibrium calculations for reactions taking place in a molten iron bath with a significant amount of slag are more complex even when the problems described above are resolved. The residence time of gaseous species in the reaction environment is on the order of a few seconds. The residence time of slag in the reaction environment is on the order of several hours. The residence time of metal in the reaction environment is on the order of days. What this means is that a calculation that assumes that all of the reactants are in equilibrium with all of the products may be seriously in error even if correct data and proper calculations are used. FactSage allows for the calculation of an “open” equilibrium. This means that with sequential feed injections the part of the feed that reports to the slag stays with the slag. The part of the feed that dissolves in the metal remains in the metal until or unless it reacts. The gaseous products thus formed are removed before the next increment of feed is added. If many feed steps are used, one usually finds that initially the composition of all three phases changes significantly from one injection to another. If a stable equilibrium is possible for the conditions used in the calculation, the change in the composition and in the amount of each phase from one injection to the next gets smaller until virtually no change takes place from one injection to the next. Typically each injection represents one hour of feed.

FactSage calculates the equilibrium composition of all three phases. It also determines if solid phases are present and their composition. This is important to the HyMelt process because the proper amount of burnt lime to be added with the feed to preclude the formation of a solid CaS phase or other solid phases can be calculated. The build up in the metal phase of chromium, nickel, vanadium and other trace metals

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

in coal can be determined. Volatile salts and metals such as KCl, NaCl, Pb, Zn and others in the product gas can be quantified. This allows for the proper design of a quench system and perhaps down stream guard beds. We can determine the presence and composition of other liquid phases such as mattes (liquid metals such as iron or nickel with high levels of sulfur) that may affect refractory selection. Using FactSage, we can determine the fate of trace elements such as arsenic, lead, mercury, selenium and zinc that may have environmental impacts.

The following table summarizes the feed and product streams for Case 1 (Hydrogen Production) and Case 2 (Fischer-Tropsch Liquids Production).

Table 3-1 Feed and Product Summary

	Case 1	Case 2	Units
Feed:			
Hot Raw Syngas	62.3	62.3	MMSCFD
Products:			
Hydrogen	52.9	1.5	MMSCFD
Syngas to F-T (2.1 H ₂ :CO ratio)	NA	57.3	MMSCFD
Carbon Dioxide	2,324 (40.1)	1,262 (21.8)	STPD (MMSCFD)
Sulfur	15.0	15.0	LTPD
Fuel Gas (PSA off-gas, net)	101.5	23.1	MM Btu/hr

Characterization of the Hot Raw Syngas from EnviRes is defined in the Basis of Design document.

Please refer to the H&MBs (Appendix 3) for stream properties and compositions for the feed syngas and product streams.

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

3.3 Block Flow Diagram

Please see the Block Flow Diagram (BFD) for Case 1 (Hydrogen Production) and the BFD for Case 2 (Fischer-Tropsch Liquids Production) on the next pages. The BFDs shows the major plant blocks and key feed, intermediate and product streams. The diagram also shows the responsibility breaks between WorleyParsons and other parties.

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Figure 3-1 Case 1 – Block Flow Diagram

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Figure 3-2 Case 2 – Block Flow Diagram

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

4. PROCESS DESCRIPTION

The process descriptions in this section describe each of the major process units for Case 1 (Hydrogen Production) and for Case 2 (Fischer-Tropsch Liquids Production) in the Syngas Treatment area of the plant. To aid in the understanding of the descriptions, please refer to the PFDs and the H&MBs in Appendices 2 and 3, respectively. All necessary utilities are provided to the Syngas Treatment area from the Balance of Plant (BOP) area.

Although the major process units for Case 1 and Case 2 are similar, separate descriptions are provided to avoid confusion between the Cases and to refer to the PFDs and equipment numbers for each Case.

4.1 Case 1 – Hydrogen Production

4.1.1 Case 1 - Syngas Quench & Scrubbing – Unit 21

Reference PFD: 21-PR-PFD-0010 and 0020

The hot raw syngas from the EnviRes HyMelt Gasification system enters the Quench & Scrubbing system of the Syngas Treatment Area where it is first quenched with water and then further scrubbed with additional water to remove solids and other impurities from the syngas. The resulting scrubbed syngas is suitable to be further processed in the Shift Reaction and Cooling system. The water stream containing the solids and other impurities is then treated such that most of the water can be reused in the scrubbing system.

The hot raw syngas from the EnviRes HyMelt Gasifier flows into the Quench Drum (V-2101). The transition piece between the Gasifier and the Quench Drum has a water-wall type lining. An internal pipe directs the syngas below the water level of the vessel while an internal spray arrangement quenches and cools the syngas. The water from this internal spray also protects the internal pipe from the hot syngas.

The syngas contains dust from the Gasifier. This solid dust material consists of small iron and carbon particles. Most of the dust material is removed with the water blowdown stream exiting the vessel through the Quench Drum level control valve. A separate remotely operated on-off valve located close to the vessel is part of the Safety Instrumented System (SIS). The function of this valve is to close on loss of level in the Quench Drum to prevent breakthrough of hot syngas to the water piping and downstream equipment.

The syngas is further scrubbed as it is mixed with additional water in the Venturi Mixer (Y-2101). The mixed three-phase stream (syngas, solids and water) is then routed into the Syngas Scrubber (V-2102) which acts as a final scrubbing step to remove the remaining particulate material. An internal pipe directs the inlet stream below the water level of the vessel. Disengaging space is provided above the water level

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

to allow larger droplets of water/solids to drop down into the water. Four wash trays use clean water (heated shift condensate) to scrub the syngas of any remaining particulate material. A vane-type mist eliminator at the top of the vessel removes entrained water before the clean raw syngas exits the Scrubber.

After exiting the Scrubber, the pressure of the syngas can be controlled by the pressure valve in the syngas piping to the flare. A SIS valve can be used to isolate the Quench and Scrubbing system from the downstream Shift system.

Heated gray water is routed to the Syngas Scrubber on level control. The scrubber bottoms water is pumped by the Syngas Scrubber Pumps (P-2101 A/B) back to the Quench Drum as spray water on flow control and to the Venturi Mixer also on flow control. It is critical to maintain the flow of water to the Quench Drum water sprays to prevent damage from the hot syngas. An emergency supply of water is available from the Boiler Feed Water system in the Utility Area.

A small amount of caustic is added to the quench water inlet stream. The caustic assists in raising the pH of the water and in removal of impurities (especially ammonia) from the syngas.

The quench bottoms water is flashed across the Quench Drum level control valve and flows into the MP Flash Drum (V-2103). The pressure is reduced and some water vapor and dissolved gas components exit the vessel in the overhead stream. The flash drum bottoms stream is further cooled in the Flash Bottoms Coolers (E-2101 A/B) and flows to the Solids Settler (T-2101) on level control. The Settler provides residence time for the solids material to concentrate. The overflow from the Settler flows into the Gray Water Tank (T-2102). The settler bottoms stream is pumped by the Settler Bottoms Pumps (P-2102 A/B) to the Solids Filters (Y-2102 A/B). The speed of the positive displacement type pumps is set by the settler interface level control.

The two Solids Filters are recessed plate and frame filters. One Filter will be operating while the filter cake from the other is being removed (dumped). After the filter cake is removed from the Filters, it is then routed to the Syngas Treatment battery limits for disposal. The filtrate flows into a small Filtrate Drum (V-2104). It is then pumped by the Filtrate Pumps (P-2105 A/B) to the Gray Water Tank (T-2102).

Most of the water from the Gray Water Tank is recycled back to the Scrubber and Quench Drum by the HP Gray Water Pumps (P-2103 A/B). A purge stream of water is blown down from the tank to the Syngas Treatment battery limits for control of chlorides and/or other impurities. The Blowdown Pumps (P-2104 A/B) are used to pump the blowdown purge stream on flow control to the treating system.

4.1.2 Case 1 - Shift Reaction & Cooling – Unit 22

Reference PFD: 22-PR-PFD-0010, 0020, 0030 and 0040

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

The saturated syngas flows from the Syngas Scrubber into the Shift Reaction & Cooling system. In the Shift Reactors the carbon monoxide (CO) in the syngas is reacted with water to form hydrogen (H₂) and carbon dioxide (CO₂). Sour shift catalyst shall be utilized. Three stages of shift are used to maximize the conversion of CO to H₂. Additional water (i.e. steam) is added to the syngas to the first shift reactor to improve conversion and reduce the reactor exit temperature. After shifting, the syngas is cooled in a series of heat exchangers. The syngas also passes through a mercury removal drum prior to the Acid Gas Removal system. Process condensate from the shifted syngas is recovered and reused in the Quench & Scrubbing system.

From the overhead of the Syngas Scrubber, the syngas flows into the Shift Inlet KO Drum (V-2201) where any condensed water is removed. MP steam is added to the saturated syngas from the KO Drum to provide adequate water for the shift reaction. The syngas then passes through the Shift Feed/Effluent Exchanger (E-2202) to be heated by the hot syngas stream leaving the First Shift Reactor (R-2201). A Shift Startup Heater (E-2201) is provided to preheat the shift catalyst.

Hot syngas is fed to the First Shift Reactor where carbon monoxide reacts with water to produce hydrogen and carbon dioxide. Carbonyl sulfide reacts with water to produce hydrogen sulfide and carbon dioxide. The following exothermic reactions take place in the presence of the catalyst:

The product stream from the first reactor is cooled by Shift Feed/Effluent Exchanger (E-2202) and then by the First MP Steam Generator (E-2203). Part of the syngas bypasses the first shift reactor on flow control and mixes with the partially shifted effluent from the first reactor. The blended stream is then fed to the Second Shift Reactor (R-2202) on temperature control. Again, similar exothermic reaction takes place in the second reactor to produce additional hydrogen. Shifted syngas from the Second Shift Reactor is cooled first by the Second MP Steam Generator (E-2204) and then by the First LP Steam Generator (E-2215).

The partially shifted stream is then fed to the Third Shift Reactor (R-2203) on temperature control. Again, similar exothermic reaction takes place in the third reactor to produce additional hydrogen. Shifted syngas from the Third Shift Reactor is cooled first by the Shift Condensate Heater (E-2205), the Gray Water Heater (E-2206), and the Second LP Steam Generator (E-2207). The syngas stream then splits and part of the stream is cooled by the Gray Water Preheater (E-2208) while the rest is cooled by the Utility Water Heater (E-2209).

Water in the syngas condenses as the gas is cooled. The condensate is separated from the shifted syngas in the Water Heater KO Drum (V-2203). Condensate from this knockout drum is routed to the HP

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Shift Condensate Drum (V-2202). The shift condensate is pumped by the HP Shift Condensate Pumps (P-2201 A/B) through the Shift Condensate Heater and on to the Syngas Scrubber.

The shifted syngas is further cooled by the Syngas Airfan Cooler (E-2210) and Syngas Trim Cooler (E-2211). The condensate is separated from the shifted syngas in the Syngas Airfan Cooler KO Drum (V-2204) and the Syngas Trim Cooler KO Drum (V-2205). Intermittent BFW wash streams can be added to the syngas upstream of both of these exchangers to wash out trace syngas components that might foul the exchangers.

Condensate from these two knockout drums is routed to the Ammonia Stripper Feed Drum (V-2207) on level control. The Feed Drum operates at much lower pressure than the Shift Cooling equipment so there is a high pressure drop across these control valves. A SIS valve is included at the water outlet of each KO Drum to prevent breakthrough of higher pressure gas to the downstream equipment upon loss of level.

Trace mercury is removed from the syngas before it is routed to the Acid Gas Removal system. The shifted syngas from the last knockout drum is heated in the Syngas Heater (E-2212) to prevent condensation on the downstream mercury adsorbent and flows through the Mercury Removal Drum (V-2206) (liquid moisture damages the adsorbant). The drum contains sulfur-impregnated activated carbon to remove mercury to very low levels. The shifted syngas then flows to the downstream Acid Gas Removal system.

The shift condensate from the Ammonia Stripper Feed Drum is pumped by the Ammonia Stripper Feed Pumps (P-2202 A/B) to the Ammonia Stripper (V-2208). Flash gas from the overhead of the MP Flash Drum in the Syngas Quench & Scrubbing system is used to provide part of the stripping vapor. The Ammonia Stripper Reboiler (E-2214) provides the remainder of the stripping vapor. The liquid from the Reboiler is pumped by the Ammonia Stripper Bottoms Pumps (P-2204 A/B) back to the HP Shift Condensate Drum.

The small amount of vapor from the Feed Drum combines with the overhead from the Ammonia Stripper. This combined overhead stream is cooled by the Ammonia Stripper Overhead Condenser (E-2213) and flows into the Ammonia Stripper Reflux Drum (V-2209). The liquid from the Reflux Drum is pumped by the Ammonia Stripper Reflux Pumps (P-2203 A/B) back to the Stripper. The sour gas from the Reflux Drum is routed to the downstream Claus Sulfur Recovery system.

4.1.3 Case 1 - Acid Gas Removal – Unit 23

Reference PFD: 23-PR-PFD-0010 and 0020

The shifted syngas is next treated in an Acid Gas Removal (AGR) system where acid gas components (CO_2 , H_2S and COS) are removed. A Rectisol Wash Package (Z-2301) has been used for this application.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Rectisol is a process co-licensed by Linde and Lurgi that utilizes methanol as a physical solvent at low temperatures and high pressures to selectively remove CO₂, sulfur compounds, and other impurities from raw feed gas. The bulk of the CO₂ removed from the syngas is concentrated into a CO₂ product stream. The H₂S and the remaining CO₂ are stripped from the solvent and leave the AGR in an acid gas stream.

The shifted syngas is first contacted with a small amount of lean solvent (methanol) to prevent freezing of any water present in the gas. The shifted syngas is cooled by cold syngas from the overhead of the Absorber and cold CO₂ gas from the CO₂ Column. After cooling, the syngas then enters the prewash section of the Main Absorber.

In the prewash section, the raw gas is washed with a small amount of cold CO₂-preladen methanol to remove ammonia, HCN, water and other expected trace impurities. This methanol is taken from the solvent stream fed to the H₂S removal section. The bottom product from the prewash section is routed to the Methanol Water Column.

The prewashed gas passes through the chimney tray separating the prewash section from the H₂S removal section. Here, H₂S and COS are removed while the CO₂ concentration remains practically constant. The solvent fed to the top of the section is already laden with CO₂ from the CO₂ absorption section. The top product of this section is almost sulfur free and flows to the CO₂ absorption section of the Absorber.

The feed gas from the H₂S removal section is cooled with cold, CO₂-laden methanol. In the upper section, the gas is finally washed with regenerated (lean) methanol coming from the Regenerator. The Absorber overhead is heated against the incoming feed syngas and is then routed to the Hydrogen Purification system.

The methanol leaving the CO₂ absorption sections at the bottom is partially fed to the top of the H₂S absorption section and partially to one of the Flash Drums where co-absorbed light valuable gases (CO, H₂, and CH₄) are released. The vapor stream from the Flash Drum is routed to a second Flash Drum. The methanol stream is then routed to the top of the CO₂ Column.

The methanol leaving the H₂S absorption section of the Absorber is fed to the second Flash Drum. Most of the co-absorbed light valuable gases are released here also. Together with the gases from the other Flash Drum, these gases are recompressed by the Recycle Compressor and recycled back to combine with the inlet feed gas. The methanol stream from the Flash Drum is then routed to an intermediate stage of the CO₂ Column.

A Reboiler system in the bottom of the CO₂ Column provides stripping vapor. The rich methanol collected in the Reboiler is pumped to the Methanol Regenerator. The CO₂ stream leaving the top of the CO₂ Column is cooled against the feed gas and routed downstream to the CO₂ Compression system.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

The sulfur enriched methanol from the CO₂ Column is fed to the top of the Methanol Regenerator. The overhead vapor stream, containing the released H₂S and CO₂ is cooled and fed to a reflux cooler/drum. The collected reflux is returned to the Methanol Regenerator. The overhead or acid gas stream is routed to the downstream Sulfur Recovery system. A methanol stream from the Methanol Water Column is also routed to the Regenerator for further stripping. The methanol is stripped with methanol vapor generated in the Regenerator Reboiler.

The fully regenerated methanol collected in the bottom of the Regenerator is pumped to the top of the Absorber. The regenerated lean methanol is cooled in a series of heat exchangers.

The methanol stream from the prewash section is routed to the Methanol Water Column. Distillation of water and methanol takes place, to keep the water content in the main methanol circuits at a low level. The vapor overhead from the Methanol Water Column is cooled and leaves the system as part of the Acid Gas stream. The bottom product is an impure water stream which is blown down to the battery limits for further treatment.

4.1.4 Case 1 - Hydrogen Purification – Unit 24

Reference PFD: 24-PR-PFD-0010

The clean raw hydrogen stream from the AGR is purified in the Hydrogen Purification system using a Pressure Swing Adsorption (PSA) unit. The impurities are captured on the adsorbent allowing the essentially pure hydrogen stream to flow through the beds. During the desorption part of this batch cycle, the impurities are desorbed from the beds at lower pressure and blended together as PSA off-gas. The off-gas stream is compressed and routed to the Syngas Treatment battery limits as fuel gas.

The PSA package (Z-2401) consists of multiple adsorbers, each operating in a batch mode. The operation of each adsorber is staggered to produce a constant product and off-gas flow. The number of needed adsorber beds will be determined by the PSA vendor. The PSA unit batch process operates in 5-steps:

- (1) Adsorption,
- (2) Co-current depressurization,
- (3) Counter-current depressurization,
- (4) Purge, and
- (5) Re-pressurization.

The hydrogen product from the PSA is routed to the Syngas Treating system battery limits.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

The PSA off-gas is used as a fuel gas and flows from the PSA Fuel Gas Drum to the Fuel Gas Compressor package (Z-2402). The fuel gas is compressed and routed to the battery limits. The Fuel Gas Compressor is not spared. If the compressor is not operating, the PSA fuel gas pressure can be raised so that the gas can be routed to the flare. The hydrogen recovery will decrease during this off-design mode of operation. The purity of the hydrogen product stream will remain on-spec.

4.1.5 Case 1 - Sulfur Recovery – Unit 25

Reference PFD: 25-PR-PFD-0010 and 0020

The acid gas stream is treated in a Claus Sulfur Recovery Unit (SRU) to produce elemental sulfur. The molten sulfur is degassed and pumped to the Syngas Treatment battery limits. Any truck or rail loading facilities needed are part of the Balance of Plant (BOP). The tail gas is treated and routed back to the inlet of the SRU.

In the Claus Package (Z-2501), the acid gas is converted into elemental sulfur by initial combustion and condensation followed by successive stages of gas reheating, catalytic conversion and condensation. HP steam is generated in the waste heat recovery boiler, while LP steam is generated in the successive sulfur condensers. The liquid sulfur drains from the condensers and is collected in the Sulfur Pit.

The Claus reaction consists of combining 2 moles of H_2S with 1 mole of SO_2 to form elemental sulfur. In the thermal stage of the process, a controlled amount of oxygen is used to oxidize some of the H_2S to SO_2 . This gives a high enough temperature to trigger the sulfur forming reaction thermally. This is depicted by the following reactions:

The acid gas stream from the upstream Rectisol Wash Package along with sour gas from the Ammonia Stripper is fed to a Knockout Drum. Any sour water or other liquid collected in the Knockout Drum is intermittently routed to treatment. The KO Drum's function is to prevent any liquid, resulting from an operating upset, being entrained in the acid gas passing to the Reaction Furnaces, thereby risking damage to the hot refractory and/or contamination of the downstream reactors' catalyst.

The Sulfur is degassed to <10 ppm (wt) H_2S in the Sulfur Degassing Package (Z-2502). Degassing of the sulfur is accomplished in a vertical vessel in which the undegassed sulfur is contacted with pressurized process air.

From the final sulfur condenser, the process gas flows to the Tail Gas Treating Package (Z-2503) for further processing to maximize sulfur recovery. The Tail Gas Treating system consists of a reducing or hydrogenation section and an amine treating section. In the hydrogenation section nearly all sulfur compounds in the SRU tail gas are converted to hydrogen sulfide.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Hydrogenation and hydrolysis reactions for the four primary sulfur constituents are as follows:

- $S + H_2 \rightarrow H_2S$
- $SO_2 + 3 H_2 \rightarrow H_2S + 2 H_2O$
- $CS_2 + 2 H_2O \rightarrow 2 H_2S + CO_2$
- $COS + H_2O \rightarrow H_2S + CO_2$

The amine treating section uses an MDEA solution to remove most of the hydrogen sulfide from the treated sulfur plant tail gas while minimizing the co-absorption of carbon dioxide. The treated tail gas from the overhead of the Amine Absorber then flows to the Thermal Oxidizer where the small amount of residual H_2S is oxidized to SO_2 . The rich amine solution from the bottom of the Absorber is then pumped and is heated in the Lean / Rich Exchanger before entering the Amine Regenerator. The rich solution is stripped counter currently with steam generated by the Amine Regenerator Reboiler. The amine acid gas exits in the overhead and is recycled to the front of the SRU.

The Thermal Oxidizer incinerates H_2S and other combustible compounds in the treated tail gas from the TGTU. The Oxidizer is typically a forced draft type incinerator. Fuel Gas is used with excess air to ensure sufficiently high temperature to achieve nearly complete oxidation of H_2S and trace sulfur compounds. Optionally, the hot gas produced in the Oxidizer can be used to superheat the High Pressure steam produced in the SRU Waste Heat Recovery Boiler.

The flue gas contains 3% oxygen and less than 10 ppm hydrogen sulfide. Flue gas is released to the atmosphere through the Stack.

4.1.6 Case 1 - CO_2 Compression – Unit 26

Reference PFD: 26-PR-PFD-0010

The raw CO_2 stream from the AGR is routed to the CO_2 Compression system. The CO_2 is compressed and routed to the Syngas Treatment battery limits. The CO_2 stream is suitable for use in EOR in a sweet field.

A package CO_2 Compressor (Z-2601) is used to compress the CO_2 from the Rectisol Wash Package to the required pipeline pressure. This multi-stage compressor package includes the compressor, motor, inlet knockout drum, and all interstage knockout coolers and drums. The CO_2 compressor is not spared.

A CO_2 Thermal Oxidizer Package is included to treat the CO_2 stream if the CO_2 compressor is not operating. Depending on the requirements of the site selected, it may be possible to delete this oxidizer package.

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

4.2 Case 2 – Fischer-Tropsch Liquids Production

4.2.1 Case 2 - Syngas Quench & Scrubbing – Unit 21

Reference PFD: 21-PR-PFD-0510 and 0520

The hot raw syngas from the EnviRes HyMelt Gasification system enters the Quench & Scrubbing system of the Syngas Treatment Area where it is first quenched with water and then further scrubbed with additional water to remove solids and other impurities from the syngas. The resulting scrubbed syngas is suitable to be further processed in the Shift Reaction and Cooling system. The water stream containing the solids and other impurities is then treated such that most of the water can be reused in the scrubbing system.

The hot raw syngas from the EnviRes HyMelt Gasifier flows into the Quench Drum (V-2151). The transition piece between the Gasifier and the Quench Drum has a water-wall type lining. An internal pipe directs the syngas below the water level of the vessel while an internal spray arrangement quenches and cools the syngas. The water from this internal spray also protects the internal pipe from the hot syngas.

The syngas contains dust from the Gasifier. This solid dust material consists of small iron and carbon particles. Most of the dust material is removed with the water blowdown stream exiting the vessel through the Quench Drum level control valve. A separate remotely operated on-off valve located close to the vessel is part of the Safety Instrumented System (SIS). The function of this valve is to close on loss of level in the Quench Drum to prevent breakthrough of hot syngas to the water piping and downstream equipment.

The syngas is further scrubbed as it is mixed with additional water in the Venturi Mixer (Y-2151). The mixed three-phase stream (syngas, solids and water) is then routed into the Syngas Scrubber (V-2152) which acts as a final scrubbing step to remove the remaining particulate material. An internal pipe directs the inlet stream below the water level of the vessel. Disengaging space is provided above the water level to allow larger droplets of water/solids to drop down into the water. Four wash trays use clean water (heated shift condensate) to scrub the syngas of any remaining particulate material. A vane-type mist eliminator at the top of the vessel removes entrained water before the clean raw syngas exits the Scrubber.

After exiting the Scrubber, the pressure of the syngas can be controlled by the pressure valve in the syngas piping to the flare. A SIS valve can be used to isolate the Quench and Scrubbing system from the downstream Shift system.

Heated gray water is routed to the Syngas Scrubber on level control. The scrubber bottoms water is pumped by the Syngas Scrubber Pumps (P-2151 A/B) back to the Quench Drum as spray water on flow control and to the Venturi Mixer also on flow control. It is critical to maintain the flow of water to the

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Quench Drum water sprays to prevent damage from the hot syngas. An emergency supply of water is available from the Boiler Feed Water system in the Utility Area.

A small amount of caustic is added to the quench water inlet stream. The caustic assists in raising the pH of the water and in removal of impurities (especially ammonia) from the syngas.

The quench bottoms water is flashed across the Quench Drum level control valve and flows into the MP Flash Drum (V-2153). The pressure is reduced and some water vapor and dissolved gas components exit the vessel in the overhead stream. The flash drum bottoms stream is further cooled in the Flash Bottoms Coolers (E-2151 A/B) and flows to the Solids Settler (T-2151) on level control. The Settler provides residence time for the solids material to concentrate. The overflow from the Settler flows into the Gray Water Tank (T-2152). The settler bottoms stream is pumped by the Settler Bottoms Pumps (P-2152 A/B) to the Solids Filters (Y-2152 A/B). The speed of the positive displacement type pumps is set by the settler interface level control.

The two Solids Filters are recessed plate and frame filters. One Filter will be operating while the filter cake from the other is being removed (dumped). After the filter cake is removed from the Filters, it is then routed to the Syngas Treatment battery limits for disposal. The filtrate flows into a small Filtrate Drum (V-2154). It is then pumped by the Filtrate Pumps (P-2155 A/B) to the Gray Water Tank (T-2152).

Most of the water from the Gray Water Tank is recycled back to the Scrubber and Quench Drum by the HP Gray Water Pumps (P-2153 A/B). A purge stream of water is blown down from the tank to the Syngas Treatment battery limits for control of chlorides and/or other impurities. The Blowdown Pumps (P-2154 A/B) are used to pump the blowdown purge stream on flow control to the treating system.

4.2.2 Case 2 - Shift Reaction & Cooling – Unit 22

Reference PFD: 22-PR-PFD-0510, 0520, 0530 and 0540

The saturated syngas flows from the Syngas Scrubber into the Shift Reaction & Cooling system. In the Shift Reactors the carbon monoxide (CO) in the syngas is reacted with water to form hydrogen (H₂) and carbon dioxide (CO₂). Sour shift catalyst shall be utilized. Two stages of shift perform the conversion of CO to H₂. Additional water (i.e. steam) is added to the syngas to the first shift reactor to improve conversion and reduce the reactor exit temperature. After shifting, the syngas is cooled in a series of heat exchangers. The syngas also passes through a mercury removal drum prior to the Acid Gas Removal system. Process condensate from the shifted syngas is recovered and reused in the Quench & Scrubbing system.

From the overhead of the Syngas Scrubber, the syngas flows into the Shift Inlet KO Drum (V-2251) where any condensed or entrained water is removed. Part of the syngas from the Shift Inlet KO Drum is mixed with MP steam to provide adequate water for the shift reaction. This syngas then passes through

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

the Shift Feed/Effluent Exchanger (E-2252) to be heated by the hot syngas stream leaving the First Shift Reactor (R-2251). A Shift Startup Heater (E-2251) is provided to preheat the shift catalyst.

The hot syngas is fed to the First Shift Reactor where carbon monoxide reacts with water to produce hydrogen and carbon dioxide. Carbonyl sulfide reacts with water to produce hydrogen sulfide and carbon dioxide. The following exothermic reactions take place in the presence of the catalyst:

The product stream from the first reactor is cooled by Shift Feed/Effluent Exchanger (E-2252) and then by the First MP Steam Generator (E-2253).

A second part of the syngas stream from the Shift Inlet KO Drum bypasses the first shift reactor on flow control and mixes with the partially shifted and cooled effluent from the First MP Steam Generator. The blended stream is then fed to the Second Shift Reactor (R-2252) on temperature control. Again, the above exothermic reactions take place in the second reactor to produce additional hydrogen. Shifted syngas from the Second Shift Reactor is cooled first by the Second MP Steam Generator (E-2254) and then by the Shift Condensate Heater (V-2255).

A third part of the syngas stream from the Shift Inlet KO Drum bypasses both shift reactors on flow control and mixes with the cooled syngas from the Shift Condensate Heater. Part of this blended stream is cooled by the Gray Water Heater (E-2256), then cooled by the Gray Water Preheater (E-2258), and then flows to the Water Heater KO Drum (V-2253). The rest of this blended syngas is cooled by the LP Steam Generator (E-2257), then cooled by the Utility Water Heater (E-2259), and then flows to the Water Heater KO Drum. The syngas flow to the Gray Water Heater is controlled by the temperature of the Gray Water from the Gray Water Preheater.

Water in the syngas condenses as the gas is cooled. The condensate is separated from the shifted syngas in the Water Heater KO Drum. Condensate from this knockout drum is routed to the HP Shift Condensate Drum (V-2252). The shift condensate is pumped by the HP Shift Condensate Pumps (P-2251 A/B) through the Shift Condensate Heater and on to the Syngas Scrubber.

The syngas from the Water Heater KO Drum is further cooled by the Syngas Airfan Cooler (E-2260) and Syngas Trim Cooler (E-2261). The condensate is separated from the shifted syngas in the Syngas Airfan Cooler KO Drum (V-2254) and the Syngas Trim Cooler KO Drum (V-2255). Intermittent BFW wash streams can be added to the syngas upstream of both of these exchangers to wash out trace syngas components that might foul the exchangers.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Condensate from these two knockout drums is routed to the Ammonia Stripper Feed Drum (V-2257) on level control. The Feed Drum operates at much lower pressure than the Shift Cooling equipment so there is a high pressure drop across these control valves. A SIS valve is included at the water outlet of each KO Drum to prevent breakthrough of higher pressure gas to the downstream equipment upon loss of level.

Trace mercury is removed from the syngas before it is routed to the Acid Gas Removal system. The shifted syngas from the Syngas Trim Cooler KO Drum is heated in the Syngas Heater (E-2262) to prevent condensation on the downstream mercury adsorbent and flows through the Mercury Removal Drum (V-2256) (liquid moisture damages the adsorbent). The drum contains sulfur-impregnated activated carbon to remove mercury to very low levels. The shifted syngas then flows to the downstream Acid Gas Removal system.

The shift condensate from the Ammonia Stripper Feed Drum is pumped by the Ammonia Stripper Feed Pumps (P-2252 A/B) to the Ammonia Stripper (V-2258). Flash gas from the overhead of the MP Flash Drum in the Syngas Quench & Scrubbing system is used to provide part of the stripping vapor. The Ammonia Stripper Reboiler (E-2264) provides the remainder of the stripping vapor. The liquid from the Reboiler is pumped by the Ammonia Stripper Bottoms Pumps (P-2254 A/B) back to the HP Shift Condensate Drum.

The small amount of vapor from the Feed Drum combines with the overhead from the Ammonia Stripper. This combined overhead stream is cooled by the Ammonia Stripper Overhead Condenser (E-2263) and flows into the Ammonia Stripper Reflux Drum (V-2259). The liquid from the Reflux Drum is pumped by the Ammonia Stripper Reflux Pumps (P-2253 A/B) back to the Stripper. The sour gas from the Reflux Drum is routed to the downstream Claus Sulfur Recovery system.

4.2.3 Case 2 - Acid Gas Removal – Unit 23

Reference PFD: 23-PR-PFD-0510 and 0520

The shifted syngas is next treated in an Acid Gas Removal (AGR) system where acid gas components (CO₂, H₂S and COS) are removed. A Rectisol Wash Package (Z-2351) has been used for this application.

Rectisol is a process co-licensed by Linde and Lurgi that utilizes methanol as a physical solvent at low temperatures and high pressures to selectively remove CO₂, sulfur compounds, and other impurities from raw feed gas. The bulk of the CO₂ removed from the syngas is concentrated into a CO₂ product stream. The H₂S and the remaining CO₂ are stripped from the solvent and leave the AGR in an acid gas stream.

The shifted syngas is first contacted with a small amount of lean solvent (methanol) to prevent freezing of any water present in the gas. The shifted syngas is cooled by cold syngas from the overhead of the

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Absorber and cold CO₂ gas from the CO₂ Column. After cooling, the syngas then enters the prewash section of the Main Absorber.

In the prewash section, the raw gas is washed with a small amount of cold CO₂-preladen methanol to remove ammonia, HCN, water and other expected trace impurities. This methanol is taken from the solvent stream fed to the H₂S removal section. The bottom product from the prewash section is routed to the Methanol Water Column.

The prewashed gas passes through the chimney tray separating the prewash section from the H₂S removal section. Here, H₂S and COS are removed while the CO₂ concentration remains practically constant. The solvent fed to the top of the section is already laden with CO₂ from the CO₂ absorption section. The top product of this section is almost sulfur free and flows to the CO₂ absorption section of the Absorber.

The feed gas from the H₂S removal section is cooled with cold, CO₂-laden methanol. In the upper section, the gas is finally washed with regenerated (lean) methanol coming from the Regenerator. The Absorber overhead is heated against the incoming feed syngas and is then routed to the Hydrogen Purification system.

The methanol leaving the CO₂ absorption sections at the bottom is partially fed to the top of the H₂S absorption section and partially to one of the Flash Drums where co-absorbed light valuable gases (CO, H₂, and CH₄) are released. The vapor stream from the Flash Drum is routed to a second Flash Drum. The methanol stream is then routed to the top of the CO₂ Column.

The methanol leaving the H₂S absorption section of the Absorber is fed to the second Flash Drum. Most of the co-absorbed light valuable gases are released here also. Together with the gases from the other Flash Drum, these gases are recompressed by the Recycle Compressor and recycled back to combine with the inlet feed gas. The methanol stream from the Flash Drum is then routed to an intermediate stage of the CO₂ Column.

A Reboiler system in the bottom of the CO₂ Column provides stripping vapor. The rich methanol collected in the Reboiler is pumped to the Methanol Regenerator. The CO₂ stream leaving the top of the CO₂ Column is cooled against the feed gas and routed downstream to the CO₂ Compression system.

The sulfur enriched methanol from the CO₂ Column is fed to the top of the Methanol Regenerator. The overhead vapor stream, containing the released H₂S and CO₂ is cooled and fed to a reflux cooler/drum. The collected reflux is returned to the Methanol Regenerator. The overhead or acid gas stream is routed to the downstream Sulfur Recovery system. A methanol stream from the Methanol Water Column is also routed to the Regenerator for further stripping. The methanol is stripped with methanol vapor generated in the Regenerator Reboiler.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

The fully regenerated methanol collected in the bottom of the Regenerator is pumped to the top of the Absorber. The regenerated lean methanol is cooled in a series of heat exchangers.

The methanol stream from the prewash section is routed to the Methanol Water Column. Distillation of water and methanol takes place, to keep the water content in the main methanol circuits at a low level. The vapor overhead from the Methanol Water Column is cooled and leaves the system as part of the Acid Gas stream. The bottom product is an impure water stream which is blown down to the battery limits for further treatment.

4.2.4 Case 2 - Hydrogen Purification – Unit 24

Reference PFD: 24-PR-PFD-0510

The clean and compressed shifted syngas from the Syngas Compression is purified in the Hydrogen Purification system using a Pressure Swing Adsorption (PSA) unit. The impurities are captured on the adsorbent allowing the essentially pure hydrogen stream to flow through the beds. During the desorption part of this batch cycle, the impurities are desorbed from the beds at lower pressure and blended together as PSA off-gas. The off-gas stream is compressed and routed to the Syngas Treatment battery limits as fuel gas.

The PSA package (Z-2451) consists of multiple adsorbers, each operating in a batch mode. The operation of each adsorber is staggered to produce a constant product and off-gas flow. The number of needed adsorber beds will be determined by the PSA vendor. The PSA unit batch process operates in 5-steps:

- (1) Adsorption,
- (2) Co-current depressurization,
- (3) Counter-current depressurization,
- (4) Purge, and
- (5) Re-pressurization.

The hydrogen product from the PSA is routed to the Syngas Treating system battery limits.

The PSA off-gas is used as a fuel gas and flows from the PSA Fuel Gas Drum to the Fuel Gas Compressor package (Z-2452). The fuel gas is compressed and routed to the battery limits. The Fuel Gas Compressor is not spared. If the compressor is not operating, the PSA fuel gas pressure can be raised so that the gas can be routed to the flare. The hydrogen recovery will decrease during this off-design mode of operation. The purity of the hydrogen product stream will remain on-spec.

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

4.2.5 Case 2 - Sulfur Recovery – Unit 25

Reference PFD: 25-PR-PFD-0510 and 0520

The acid gas stream is treated in a Claus Sulfur Recovery Unit (SRU) to produce elemental sulfur. The molten sulfur is degassed and pumped to the Syngas Treatment battery limits. Any truck or rail loading facilities needed are part of the Balance of Plant (BOP). The tail gas is treated and routed back to the inlet of the SRU.

In the Claus Package (Z-2551), the acid gas is converted into elemental sulfur by initial combustion and condensation followed by successive stages of gas reheating, catalytic conversion and condensation. HP steam is generated in the waste heat recovery boiler, while LP steam is generated in the successive sulfur condensers. The liquid sulfur drains from the condensers and is collected in the Sulfur Pit.

The Claus reaction consists of combining 2 moles of H_2S with 1 mole of SO_2 to form elemental sulfur. In the thermal stage of the process, a controlled amount of oxygen is used to oxidize some of the H_2S to SO_2 . This gives a high enough temperature to trigger the sulfur forming reaction thermally. This is depicted by the following reactions:

The acid gas stream from the upstream Rectisol Wash Package along with sour gas from the Ammonia Stripper is fed to a Knockout Drum. Any sour water or other liquid collected in the Knockout Drum is intermittently routed to treatment. The KO Drum's function is to prevent any liquid, resulting from an operating upset, being entrained in the acid gas passing to the Reaction Furnaces, thereby risking damage to the hot refractory and/or contamination of the downstream reactors' catalyst.

The Sulfur is degassed to <10 ppm (wt) H_2S in the Sulfur Degassing Package (Z-2552). Degassing of the sulfur is accomplished in a vertical vessel in which the undegassed sulfur is contacted with pressurized process air.

From the final sulfur condenser, the process gas flows to the Tail Gas Treating Package (Z-2553) for further processing to maximize sulfur recovery. The Tail Gas Treating system consists of a reducing or hydrogenation section and an amine treating section. In the hydrogenation section nearly all sulfur compounds in the SRU tail gas are converted to hydrogen sulfide.

Hydrogenation and hydrolysis reactions for the four primary sulfur constituents are as follows:

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

The amine treating section uses an MDEA solution to remove most of the hydrogen sulfide from the treated sulfur plant tail gas while minimizing the co-absorption of carbon dioxide. The treated tail gas from the overhead of the Amine Absorber then flows to the Thermal Oxidizer where the small amount of residual H_2S is oxidized to SO_2 . The rich amine solution from the bottom of the Absorber is then pumped and is heated in the Lean / Rich Exchanger before entering the Amine Regenerator. The rich solution is stripped counter currently with steam generated by the Amine Regenerator Reboiler. The amine acid gas exits in the overhead and is recycled to the front of the SRU.

The Thermal Oxidizer incinerates H_2S and other combustible compounds in the treated tail gas from the TGTU. The Oxidizer is typically a forced draft type incinerator. Fuel Gas is used with excess air to ensure sufficiently high temperature to achieve nearly complete oxidation of H_2S and trace sulfur compounds. Optionally, the hot gas produced in the Oxidizer can be used to superheat the High Pressure steam produced in the SRU Waste Heat Recovery Boiler.

The flue gas contains 3% oxygen and less than 10 ppm hydrogen sulfide. Flue gas is released to the atmosphere through the Stack.

4.2.6 Case 2 - CO_2 Compression – Unit 26

Reference PFD: 26-PR-PFD-0510

The raw CO_2 stream from the AGR is routed to the CO_2 Compression system. The CO_2 is compressed and routed to the Syngas Treatment battery limits. The CO_2 stream is suitable for use in EOR in a sweet field.

A package CO_2 Compressor (Z-2651) is used to compress the CO_2 from the Rectisol Wash Package to the required pipeline pressure. This multi-stage compressor package includes the compressor, motor, inlet knockout drum, and all interstage knockout coolers and drums. The CO_2 compressor is not spared.

A CO_2 Thermal Oxidizer Package is included to treat the CO_2 stream if the CO_2 compressor is not operating. Depending on the requirements of the site selected, it may be possible to delete this oxidizer package.

4.2.7 Case 2 – Syngas Compression – Unit 27

Reference PFD: 27-PR-PFD-0510

The sweet syngas stream from the AGR is routed to the Syngas Compression system. The sweet syngas is compressed and any residual sulfur is removed (i.e. polished) before it is routed to the Syngas Treatment battery limits. A small portion of the compressed syngas is cooled and routed to the Hydrogen Purification system.

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

A package Syngas Compressor (Z-2751) is used to compress the sweet syngas from the Rectisol Wash Package to the required pressure for the Fischer-Tropsch system. This single-stage compressor package includes the compressor, motor, inlet knockout drum, and spillback cooler. The Syngas Compressor is not spared.

The main portion of the compressed sweet syngas is heated in the Sulfur Guard Heater (E-2751) and is fed to the Sulfur Guard Vessel (V-2751). This vessel is used to ensure the syngas meets the Fischer-Tropsch sulfur requirements during minor upsets in the AGR by reacting sulfur components with Zinc Oxide (ZnO) adsorbent. The hot, polished syngas is routed to the Syngas Treatment battery limits.

A small part of the compressed sweet syngas is cooled in the PSA Feed Cooler (E-2752) and routed to the Hydrogen Purification system.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

5. PROCESS FLOW DIAGRAMS

Please see Appendix 2 for the Process Flow Diagrams (PFDs). The following PFD's provide the process configuration for Case 1 (Hydrogen Production) of the Syngas Treatment Study.

• ICRC-00-PR-PFD-0010-0	Legend Symbols
• ICRC-00-PR-PFD-0020-0	Legend Nomenclature
• ICRC-21-PR-PFD-0010-0	Syngas Quench & Scrubbing
• ICRC-21-PR-PFD-0020-0	Quench Water Handling
• ICRC-22-PR-PFD-0010-0	Shift Reactors
• ICRC-22-PR-PFD-0020-0	Shift Cooling - 1
• ICRC-22-PR-PFD-0030-0	Shift Cooling - 2
• ICRC-22-PR-PFD-0040-0	Ammonia Stripping
• ICRC-23-PR-PFD-0010-0	Rectisol Absorption
• ICRC-23-PR-PFD-0020-0	Rectisol Regeneration
• ICRC-24-PR-PFD-0010-0	Hydrogen Purification
• ICRC-25-PR-PFD-0010-0	Claus Sulfur Recovery
• ICRC-25-PR-PFD-0020-0	Tail Gas Treating
• ICRC-26-PR-PFD-0010-0	CO ₂ Compression

The following PFDs provide the process configuration for Case 2 (Fischer-Tropsch) of the Syngas Treatment Study.

• ICRC-00-PR-PFD-0510-0	Legend Symbols
• ICRC-00-PR-PFD-0520-0	Legend Nomenclature
• ICRC-21-PR-PFD-0510-0	Syngas Quench & Scrubbing
• ICRC-21-PR-PFD-0520-0	Quench Water Handling
• ICRC-22-PR-PFD-0510-0	Shift Reactors
• ICRC-22-PR-PFD-0520-0	Shift Cooling - 1
• ICRC-22-PR-PFD-0530-0	Shift Cooling - 2
• ICRC-22-PR-PFD-0540-0	Ammonia Stripping
• ICRC-23-PR-PFD-0510-0	Rectisol Absorption

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

- ICRC-23-PR-PFD-0520-0 Rectisol Regeneration
- ICRC-24-PR-PFD-0510-0 Hydrogen Purification
- ICRC-25-PR-PFD-0510-0 Claus Sulfur Recovery
- ICRC-25-PR-PFD-0520-0 Tail Gas Treating
- ICRC-26-PR-PFD-0510-0 CO₂ Compression
- ICRC-27-PR-PFD-0510-0 Syngas Compression

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

6. HEAT & MATERIAL BALANCE

Please see Appendix 3 for the Heat & Material Balances (H&MBs) for Case 1 (Hydrogen Production) and Case 2 (Fischer-Tropsch Liquids Production) of the Syngas Treatment Study.

WorleyParsons developed the H&MBs for the Syngas Treatment area which includes:

- Unit 21 – Syngas Quench & Scrubbing
- Unit 22 – Shift Reaction & Cooling
- Unit 23 – Acid Gas Removal
- Unit 24 – Hydrogen Purification
- Unit 25 – Sulfur Recovery
- Unit 26 – CO₂ Compression
- Unit 27 – Syngas Compression – only required for Case 2

AspenTech Aspen Plus software with electrolytic capability was used to produce the process simulation.

For both H&MBs Pages 1 to 5 of the H&MB show the component flows in lbmol/hr and mol %.

Pages 6 to 10 show the component flows in lb/hr and mass %.

The property data shown below the components flows are the same for both the molar and mass pages.

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

7. DESIGN CONSIDERATIONS

The following is a summary of the process design considerations developed during the Syngas Treatment Study.

7.1 General

Typical utilities (steam, power, cooling water, boiler feed water, etc.) will be available as needed by the process (including start-up), and excess steam generated, steam condensate, and waste heat recovered will be utilized by the Balance of Plant. However, this does not constitute a high level of process integration which might be achieved with a detailed design of the complete facility.

Air Fan cooling is used where practical (generally $>160^{\circ}\text{F}$) to minimize the load on Cooling Water. The definition of practical may change for sites with inexpensive water.

WorleyParsons developed the H&MB using Aspen to simulate the process. Three property methods are used in this simulation to calculate accurate physical properties:

- The ELECNTRL method is used where syngas and liquid water are expected. This method has the NRTL-RK property method for the liquid phase and the Redlich-Kwong Equation of State (EOS) for the vapor phase.
- The PR-BM method (Peng-Robinson with Boston-Mathias modifications) is used where only vapor phase syngas is present.
- The STEAM-TA method is used for pure water/steam (ASME 1967 steam table correlations).

Product recoveries based on a budgetary quote were used in the simulation for the PSA, and product recoveries based on similar units for other projects were used for the Rectisol unit.

7.2 Case 1 Design Considerations

7.2.1 Case 1 - Syngas Quench & Scrubbing

The Quench Drum effectively cools the Hot Raw Syngas from the gasifier and captures most of the solids. The Venturi Mixer and Syngas Scrubber remove the remainder of the solids producing a saturated, solids free Raw Syngas. Note that a minor portion of the solids (trace elements) in the Hot Raw Syngas are soluble in water and leave the Quench Drum with the electrolytes in the liquid phase in stream 103 in the H&MB.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

The blowdown from the Quench Drum was minimized by allowing a solids content of 3 wt% and a chloride ion (Cl⁻) content of 2,000 ppmw in the blow down water. These are reasonable and practical limits that have been successfully used in other facilities.

Heating of the Shift Condensate and Gray Water going to the Syngas Scrubber was maximized to raise the moisture content of the scrubbed Raw Syngas and minimize MP steam injection for the shift reactors. This gives Shift Cooling a more complex heat exchanger arrangement than might otherwise be required.

7.2.2 Case 1 - Shift Reaction & Cooling

Three shift reactors were selected to maximize CO conversion, but steam injection is minimized which constrains CO conversions (but lowers operating cost). Half of the syngas from the Syngas Scrubber is fed to the First Shift reactor along with just enough steam to control the temperature rise across this reactor. The outlet is cooled, mixed with the remaining syngas and fed to the Second Shift Reactor. This mixed stream contains adequate steam for the second and third shift reactors, and the hydrogen and carbon dioxide produced in the First Shift Reactor help limit the temperature rise across the second and third shift reactors. The third shift reactor is run as cool as allowed by the catalyst to maximize CO conversion.

Steam is generated by cooling the outlet from the First Shift Reactor and the Second Shift Reactor. This steam generation is used to control the inlet temperature to the Second Shift Reactor and the Third Shift Reactor. Because the inlet temperature to the Third Shift Reactor is relatively low, two levels of steam are generated to cool its inlet stream, MP steam and LP steam. This adds an additional exchanger and pressure drop to the syngas flow path, but it allows some steam to be generated at the more valuable 650 psig MP steam level. Two alternatives considered and rejected were: 1) to generate all the steam at less than 400 psig and 2) to raise the Third Shift Reactor inlet temperature which would decrease the CO conversion.

After LP Steam is generated from cooling the shifted syngas, some additional low level heat was assumed to be worth recovering before the Airfan Cooler rejects the cooling duty to atmosphere. This heat was not needed in the Syngas Treatment Area, so a utility water stream was assumed that could be heated. In the next phase, this could be one or more heating services that need heat at less than 300° F.

7.2.3 Case 1 - Hydrogen Purification

Higher hydrogen purity for the PSA was investigated (99.5 mol% H₂ versus 99.0 mol%), but this would lead to a significant increase of H₂ lost to the Fuel Gas / PSA Off-gas.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

7.2.4 Case 1 - CO₂ Compression

A Thermal Oxidizer was included to treat the CO₂ product when it is vented to the atmosphere due to a CO₂ compressor trip. There may be a possibility that some locations will not require this oxidizer for CO₂ purities which are still practical for a Rectisol unit, but this is by no means assured. Including this unit is the conservative choice.

An Integrally Geared Compressor is well suited for the multi-stage compression of CO₂. This type of compressor features a large bull gear rotating at motor speed and up to four high-speed pinions. One or two impellers (compressor stages) can be mounted on each pinion shaft. Impeller speed can be optimized for each stage of compression (or pair of stages) using the physical properties of CO₂ in each stage and the multiple case design allows the ability to intercool between stages to increase efficiency.

There are several manufacturers of Integrally Geared Compressors; Man Turbo AG, General Electric, and Mitsubishi Heavy Industries for example. Please refer to their web sites for further information on these compressors.

7.3 Case 2 Design Considerations

7.3.1 Case 2 - Syngas Quench & Scrubbing

The Quench Drum effectively cools the Hot Raw Syngas from the gasifier and captures most of the solids. The Venturi Mixer and Syngas Scrubber remove the remainder of the solids producing a saturated, solids free Raw Syngas. Note that a minor portion of the solids (trace elements) in the Hot Raw Syngas are soluble in water and leave the Quench Drum with the electrolytes in the liquid phase in stream 103 in the H&MB.

The blowdown from the Quench Drum was minimized by allowing a solids content of 3 wt% and a chloride ion (Cl⁻) content of 2,000 ppmw in the blow down water. These are reasonable and practical limits that have been successfully used in other facilities.

Heating of the Shift Condensate and Gray Water going to the Syngas Scrubber was maximized to raise the moisture content of the scrubbed Raw Syngas and minimize MP steam injection for the shift reactors. This gives Shift Cooling a more complex heat exchanger arrangement than might otherwise be required.

Overall, heat recovery in Case 2 was less than in Case 1 which gave slightly cooler Syngas Scrubber make-up water.

7.3.2 Case 2 - Shift Reaction & Cooling

Two shift reactors were selected because the third reactor is not needed to maximize H₂ production. Part of the syngas from the Syngas Scrubber is bypassed around both reactors to control the H₂:CO ratio at

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

the desired value of 2.1. A single shift reactor was not selected because it would use more steam for controlling the S/DG ratio than two shift reactors.

MP Steam is generated by cooling the outlet from the First Shift Reactor and the Second Shift Reactor. The First MP Steam Generator is used to control the inlet temperature to the Second Shift Reactor.

The first and second shift reactor operating conditions are the same as in Case 1 (using Unicat's proposal). The reactors are smaller because part of the syngas from the Syngas Scrubber is bypassed around both shift reactors.

Because there is less shifting occurring, less water is consumed to make H₂ and there is less steam generation than in Case 1. The cooling curve of the syngas changed which led to changes in the downstream arrangement of heat exchangers from Case 1.

After LP Steam is generated from cooling the shifted syngas, some additional low level heat was assumed to be worth recovering before the Airfan Cooler rejects the cooling duty to atmosphere. This heat was not needed in the Syngas Treatment Area, so a utility water stream was assumed that could be heated. In the next phase, this could be one or more heating services that need heat at less than 300° F.

7.3.3 Case 2 - Acid Gas Removal

The Rectisol Wash Package in Case 2 is smaller in than the Case 1. Because less CO was converted into H₂ and CO₂, there is only 82% as much Syngas going to the AGR.

The CO₂ product contains 52% of the carbon that was gasified into syngas. This is less than the 95% of the carbon in Case 1 as much of the carbon is converted into useful Fischer-Tropsch liquids.

7.3.4 Case 2 - Hydrogen Purification

The requirement for only 1.5 MMSCFD of 99.0% pure hydrogen gives a considerably smaller PSA package than in Case 1. Only 5.9% of the clean shifted syngas from the Rectisol package is needed to make this H₂.

The clean syngas feed to Hydrogen Purification was taken from the Syngas Compressor outlet instead of directly from the Rectisol package. This minimizes the PSA cost without significantly impact on the compressor.

The syngas to the PSA contains more CO than in Case 1 because less shifting was performed. The PSA process can still easily product the 99% pure H₂ product. The extra CO will give a proportionally larger PSA off-gas stream, but still small compared to Case 1.

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

7.3.5 Case 2 - CO₂ Compression

The CO₂ Compression is smaller than in Case 1 because less CO₂ is produced in CO Shift.

A Thermal Oxidizer was included to treat the CO₂ product when it is vented to the atmosphere due to a CO₂ compressor trip. There may be a possibility that some locations will not require this oxidizer for CO₂ purities which are still practical for a Rectisol unit, but this is by no means assured. Including this unit is the conservative choice.

An Integrally Geared Compressor is well suited for the multi-stage compression of CO₂. This type of compressor features a large bull gear rotating at motor speed and up to four high-speed pinions. One or two impellers (compressor stages) can be mounted on each pinion shaft. Impeller speed can be optimized for each stage of compression (or pair of stages) using the physical properties of CO₂ in each stage and the multiple case designs allows the ability to intercool between stages to increase efficiency.

There are several manufacturers of Integrally Geared Compressors; Man Turbo AG, General Electric, and Mitsubishi Heavy Industries for example. Please refer to their web sites for further information on these compressors.

7.3.6 Case 2 - Syngas Compression

A syngas compressor was added in order to supply the required syngas pressure to the Fischer-Tropsch unit. It was not practical to raise the HyMelt Molten Iron Gasifier pressure or significantly reduce the pressure drop through Syngas Treatment.

Because cobalt Fischer-Tropsch catalysts are very sensitive to sulfur, a Sulfur Guard Vessel was added to protect the downstream unit from upsets in the Rectisol Wash system per earlier discussions with the client. Rectisol is used reliably in many gasification plants (at one plant Rectisol is installed without a sulfur guard vessel), Rectisol reliably delivers ultra-low sulfur syngas to sweet shift catalyst which is also very sensitive to sulfur. However, adding the Sulfur Guard Vessel is the conservative approach.

Placing the ZnO Sulfur Guard Vessel after the Syngas Compressor will take advantage of the heat-of-compression. The Fischer-Tropsch unit operates near 400°F and the Sulfur Guard Vessel also needs this temperature to efficiently polish the syngas down to 10 ppb sulfur levels.

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

8. UTILITIES, CATALYST, CHEMICALS & EFFLUENTS

8.1 Utility Summary

Please see Appendix 7 or the detailed Utility Summary by Unit. The following table summarizes the utility requirements for both cases of the Syngas Treatment Study.

Table 8-1 Utility Summary

Utility	Case 1	Case 2	Units
MP Steam			
Consumed	26,210	28,690	lb/hr
Produced	49,355	31,117	lb/hr
Net Exported	23,145	2,427	lb/hr
LP Steam			
Consumed	28,730	24,765	lb/hr
Produced	20,786	30,902	lb/hr
Net Exported (Imported)	(7,944)	6,137	lb/hr
Electric Power (Operating)	12,664 16,983	9,787 13,125	kW hp
Cooling Water	11,750	8,890	gpm @ 20°F temp rise
Boiler Feed Water	71,584	63,201	lb/hr
Demin Water	86,370	59,940	lb/hr
Condensate Return	28,730	37,504	lb/hr
Utility Water for cooling / heat recovery (Note 1)	400	400	gpm

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Utility	Case 1	Case 2	Units
Fuel Gas			
Consumed	2.4	2.4	MM Btu/hr
Produced	103.9	25.5	MM Btu/hr
Net Exported	101.5	23.1	MM Btu/hr

8.2 Catalyst Summary

The following table summarizes the utility requirements for both cases of the Syngas Treatment Study.

Table 8-2 Catalyst Summary

Catalyst	Case 1	Case 2	Units
Shift Catalyst –			
First Shift Reactor	741	452	ft ³
Second Shift Reactor	1,412	861	
Third Shift Reactor	1,659	N/A	
Sulfur impregnated activated carbon for Mercury Removal	302	302	ft ³
Claus Catalyst - Alumina	103	103	ft ³
Claus Catalyst - Titania	34	34	ft ³
TGTU - Hydrogenation Catalyst	29	29	ft ³
PSA adsorbent	By PSA Vendor	By PSA Vendor	ft ³
Sulfur Guard catalyst (ZnO)	N/A	1,111	ft ³

8.3 Chemical Summary

The following table summarizes the chemical requirements for both cases of the Syngas Treatment Study.

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Table 8-3 Chemical Summary

Chemical	Case 1	Case 2	Units
Caustic (NaOH)	694	694	lb/hr of 20 wt% solution
Methanol (for Rectisol)	120,000	100,000	gal/year
MDEA (for TGTU)	2,000	2,000	lb/year

8.4 Effluent Summary

The following table summarizes the effluents for both cases of the Syngas Treatment Study.

Table 8-4 Effluent Summary

Effluent	Case 1	Case 2	Units
Filter Cake	3,360	3,360	lb/hr (50 wt% moisture)
Blowdown Gray Water	32,277	32,277	lb/hr (2,000 ppmw Cl ⁻)
Other Waste Water	1,384	1,172	lb/hr
Stack Gas from TGTU Thermal Oxidizer	1.5	1.5	MMSCFD; 3% excess O ₂ , 219 ppmv SO ₂

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

9. PLOT PLAN

Please see Appendix 4 for the Plot Plans for Case 1 and Case 2 of the Syngas Treatment Study.

The layout for the Case 1 preliminary plot plan includes the Syngas Quench & Scrubbing, the Shift Reaction & Cooling system, the Rectisol AGR, the PSA system, the Claus SRU & TGTU and the CO₂ Compression system. Many of the vessels and tanks that were sized during this study are shown individually on the plot plan. The other systems are shown as blocks.

The layout for the Case 2 preliminary plot plan is similar to the Case 1 plot plan with the Syngas Compression system added to the plot.

Access is shown for catalyst loading and unloading. Access is also shown for exchanger bundle removal. The gray water tank is located inside a diked area. The Solids Filters are elevated in a structure and paved access is provided to allow removal of the filter cake by truck or container. The Quench Drum is located immediately adjacent to the Gasification Area in order to minimize the length of special high temperature, water-wall piping.

In addition to the units with the WorleyParsons' scope, the following were also shown for illustrative purposes:

- Gasification Area
- Utility Area
- Cooling Tower
- Flare
- Office & Control Room
- Warehouse & Workshop
- Guard House

In developing this layout, the following assumptions were made:

- Greenfield site
- Level and clear of obstructions
- Available site is roughly square
- Independent facility (i.e. not part of a large facility)

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

- The ASU is not shown (over-the-fence oxygen supply)
- Coal and Biomass receiving, storage and preparation are not defined and therefore are not shown. These systems could be located across the plant road from the Gasification Area with supply to the process area by conveyors.
- Slag handling is assumed to be part of the Gasification area. Any additional plot space needed for slag storage outside of the Gasification Area is not included
- Highway, railroad, barge access to site are not defined and therefore are not shown.
- For Case 2, the Fischer-Tropsch unit is not shown and is assumed to be located on an adjacent plot.

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

10. COST ESTIMATES

10.1 Capital Cost

The estimated Capital Cost including material, labor, engineering and contingency for Case 1 (Hydrogen Production) is \$168,208,847 and for Case 2 (Fischer-Tropsch Liquids Production) is \$147,015,573.

Please see Appendix 5 for the Capital Cost details for both cases of the Syngas Treatment Study.

10.1.1 Purpose

WorleyParsons has prepared capital cost estimates for the syngas treatment portion of the gasification plant. The Case 1 facility produces hydrogen. The Case 2 facility produces feed for the Fischer-Tropsch Liquids Production along with a small amount of hydrogen for hydrotreating the F-T liquids. The estimates represent the total installed costs of the defined scope including the equipment and utilities required to support the above process installation. The OSBL (Outside Battery Limits) scope including the gasification, feed handling and ASU are by others and are excluded from WorleyParsons scope.

10.1.2 Description

SCOPE OF WORK

Facility Scope inside Battery Limits

The estimate scope includes the following units inside battery limits (ISBL)

- Unit 21 – Syngas Quench and Scrubbing
- Unit 22 – Shift Reaction and Cooling
- Unit 23 – Acid Gas Removal
- Unit 24 – Hydrogen Purification
- Unit 25 – Sulfur Recovery
- Unit 26 – CO₂ Compression
- Unit 27 – Syngas Compression (Case 2 only)

Catalysts are assumed to be installed during mechanical completion of the facility. It is therefore considered within the contractor's scope and included within the estimate.

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

Balance of Plant

The balance of estimate scope for utilities includes equipment and bulk materials necessary to directly support the operation of the above process units.

The balance of plant also includes electrical equipment and electrical bulk commodities directly associated with delivering power and control to the above process units, pipe rack, plant roads and site preparation and leveling.

Engineering, Procurement and Construction Management have been included as a percentage of total direct construction costs.

Exclusions from Scope

The OSBL scope excluded from the estimate comprises of interconnecting piping necessary to integrate the syngas treatment units with the remainder of the facility excluded from WorleyParsons scope such as Gasification, Feed Handling, and ASU. Utilities such as cooling water, flare system, wastewater, demineralization for boiler feed water, nitrogen supply for blanketing, sewage, plant operations and administrative facilities are also excluded.

Owner's costs are excluded from the estimate such as commissioning first fills of chemicals, vendor support for pre-commissioning, commissioning and start-up activities, operational and commissioning spares, preliminary engineering, site evaluations, owner's project management, operator training, office equipment, mobile equipment, & design development escalation.

10.1.3 Principles of Estimate

ESTIMATE BASIS AND DERIVATION

For estimating all syngas treatment units, an in-house data base with bare equipment costs was utilized. A current quotation for the PSA unit was solicited for Case 1 and for Case 2. For packages without current quotations, pricing was adjusted from quotations based upon relative capacities between the baseline and estimated units. The Nelson Farrar Refinery Index was used to escalate costs to the 2nd quarter of 2009.

CURRENCY AND LOCATION

The estimates are prepared in USD on a Gulf Coast basis for location, labor rates, and labor productivity. US Gulf Coast based capital costs are an industry recognized cost baseline and can be adjusted to reflect costs in most areas of the country with relatively simple manipulations.

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

CONTROLLING DOCUMENTATION

The following documents and data formed the basis of the Estimates:

- Equipment List
- Process Description
- Material Balance

10.1.4 Cost Estimate Methodology

The capital cost estimate is based on the engineered equipment list specifying size, design conditions, material selection and configuration for individual equipment items. Based upon these parameters, the bare cost delivered to the installation site is developed. The cost of insulation for hot/cold temperatures or personal protection is included in the bare cost for vessels and exchangers as required.

Installation factors are applied to cover material bulks within a perimeter of each equipment item. Material bulks include electrical hook-ups, tagged instruments, foundations, interconnecting piping and structural steel for support.

The installation factors also cover the direct labor necessary for field assembly and installation. Engineering and vendor support are excluded.

Process packages are generally shop fabricated with the majority of assembly performed before shipment to the installation site. The assemblies are shipped material bulks already connected. An installation factor of 1.5 or below is applied to cover the cost of installing the skid on a table top foundation and any field assembly and hook-ups with surrounding equipment.

Catalyst costs are developed separately based upon a quotation from Unicat. Labor costs are assumed to be included within the installation factors applied to the shift reactors and guard bed vessels.

Equipment and piping insulation is also included where required.

10.1.5 Process Units

The designs for Case 1 (Hydrogen Production) and Case 2 (Fischer-Tropsch Liquids Production) have much of the same equipment although frequently of different size. The same estimating method is used in both cases for similar equipment with adjustments for capacity.

The text below indicates the methodology for estimating the process units within the estimate scope.

Unit 21 – Syngas Quench and Scrubbing

Pump estimates are based on material specifications identified in the equipment list.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

The cost of duplex stainless steel tubing is included for the flash bottoms coolers.

The solids settler and syngas scrubber are adjusted from a quotation from in house cost data. The quench drum estimate includes additional cost for designing and fabricating a vessel with a bottom cone configuration.

The basis for the settler bottoms pump is rotary screw with progressive cavities which is suitable for handling liquids with solid fines.

The stainless steel Venturi mixer estimate is based on in house cost data.

The estimate for the solids filters are based on in house cost data. The filters are a plate and frame type filter press with polypropylene plates.

Unit 22 – Shift Reaction and Cooling

Pump estimates are based on material specifications identified in the equipment list.

The cost of high alloy material for the vessel/exchanger shells, exchanger tubes and cladding, is included in the estimate as specified.

The cost of internals is included for catalyst support and gas flow distribution for the shift reactors. The catalyst cost is added to the factored bare costs of the reactor shells and internals based on an in house vendor unit price.

The cost of a vane demister is included for the Syngas Cooler KO Drum and Shift Inlet KO Drum.

The mercury removal vessel costs are based upon estimating the shell and internals and adsorbent separately.

Unit 23 – Acid Gas Removal

The Rectisol process is the basis for acid gas removal. The costs are based on in house cost data and adjusted for the technical parameters of the project. The vendor licensing fee is included in the estimate.

Unit 24 – Hydrogen Purification

This unit includes a pressure swing absorption (PSA) skid and fuel gas compressor. A quotation for Case 1 and for Case 2 was provided directly for the PSA unit to account for differences in the feed composition, desired hydrogen purity and feed rates.

Quotations from UOP with a specified hydrogen purity of 99.0% are the basis for the cost estimates. The quoted scope includes absorber vessels, surge tank, and initial charge of adsorbent medium. It excludes the feed tank, analyzer, installation labor, piping and electrical wiring. An installation factor was included to cover the costs with input from UOP.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

In house estimating data was utilized for the fuel gas screw compressor taking tail gas from the PSA unit. An installation factor of 2.2 is applied to address the cost of intercoolers, interconnecting piping, instrumentation, electrical wiring, foundation, and field labor.

Unit 25 – Sulfur Removal

The Sulfur Removal is based on cost data from a two stage Claus unit and tail gas treatment unit with hydrogenation, amine treatment and incineration from a recent refinery project. Both baseline units are similar in size to the current unit. The sulfur degassing unit cost is adjusted from in house cost data.

Unit 26 – CO₂ Compression

The CO₂ compressor is based upon a recent quotation for a similar CO₂ compressor with an integral gear configuration, four stages and approximately the same suction and discharge pressures. The estimate is adjusted from the above quotation includes the compressor, synchronous motor, intercoolers and interconnecting piping. The foundation, additional instrumentation, field labor, engineering support etc. are included within the installation factor.

A vendor quote was obtained for the Case 1 CO₂ thermal oxidizer comprising of a unit with separate costs for the stack and burner management system. The Case 2 estimate was factored from the Case 1 quote.

Unit 27 – Syngas Compression (Case 2 - only)

Syngas Compression is included in the scope for Case 2. It includes a sulfur guard heater, PSA feed cooler, sulfur guard vessel, and syngas compressor package.

The syngas compressor is a single stage centrifugal unit and comprises installation on a table top foundation, auxiliary equipment including the after cooler, tagged instruments and electrical hook-ups.

A 30% allowance is included to cover the cost of internals necessary for installing and supporting the catalyst. The labor to install the catalyst is included with the guard bed reactor.

10.1.6 Contingency

An overall contingency of 25% has been applied reflecting the availability of information for the estimate and future uncertainty.

10.1.7 Taxes

Taxes are excluded from the estimate.

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

10.2 Operating Cost

The estimated annual Operating Cost including personnel, consumables, chemical, and disposal costs for Case 1 (Hydrogen Production) is \$18,049,000 and for Case 2 (Fischer-Tropsch Liquids Production) is \$15,526,000.

Please see Appendix 5 for the Capital Cost details for both cases of the Syngas Treatment Study.

The operating and maintenance costs are estimated as an annual average on a US Gulf Coast basis. All rates for utilities, consumables and waste disposal reflect Gulf Coast pricing. No tariffs or taxes are included in the OPEX estimate.

The following basis is used to generate the OPEX estimate:

10.2.1 Personnel Costs

- Management, supervisory and technical personnel will be scheduled 40 hours per week. They will work a normal day shift.
- Operating personnel will be assigned to one of five 12 hour shifts. Five crews will, therefore, work an average of 40 hours per week to cover normal hours and downtime for 10 working days vacation, 9 holidays and 10 sick days per year. A factor of 1.1 will be applied to straight time personnel costs to cover overtime costs.
- The overall facility will have Management/Finance/Office, Technical/Engineering and Maintenance/Warehouse personnel. 33% of their cost will be allocated to Syngas Treatment.
- Operation/Shift personnel designated for Syngas Treatment will have 100% of their cost allocated to Syngas Treatment.
- The facility will operate for 347 days per year. The remaining period will be downtime for shutdown and maintenance. All personnel will work their full schedule during this period.
- The local labor market will supply all operating personnel. Incentive pay does not apply.

10.2.2 Utility Costs

Utility costs are generated by multiplying the utility usage by the facility's internal price for that utility. The facility's internal utility prices were derived from recent project data.

Costs were estimated for the following utilities:

- Low and medium pressure steams are produced internally. An opportunity cost for consumption rather than export is applied.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

- Electric power is imported from a power grid servicing the total gasification facility.
- Cooling water requires the injection of chemical additives and pump horsepower expended outside the battery limits of the syngas treatment facility. Cooling water flows within a closed circuit between users in the syngas facility and the cooling tower outside the battery limits of the syngas facility.
- Boiler Feedwater and Demineralized water are generated outside the Syngas Treatment area.

10.2.3 Catalysts and Adsorbent Media

Process provided replacement rates for catalysts/adsorbents. Costs for maintenance and repair covers field labor required for installation. Shift catalyst costs were recently quoted by Unicat Inc. Other catalyst costs are based on previous projects.

- The First Shift Reactor catalyst is replaced every two years). The Case 1 First Shift Reactor catalyst volume is 741 ft³ and Case 2 is 452 ft³.
- The Second and Third Shift Reactor catalysts are replaced every four years . The Case 1 Second Shift Reactor catalyst volume is 1,412 ft³ and Case 2 is 861 ft³. The Case 1 Third Shift Reactor catalyst volume is 1,659 ft³; Case 2 does not have a third reactor.
- The Claus SRU Titania catalyst (34 ft³ for both cases) is replaced every four years.
- The Claus SRU Alumina catalyst (103 ft³ for both cases) is replaced every four years.
- The TGTU catalyst (29 ft³ for both cases) is replaced every six years.
- The adsorbent media for mercury removal (302 ft³ and 30 lb/ft³ for both cases) is replaced every 10 years.
- The PSA adsorbent does not require annual replacement.

10.2.4 Chemical Costs

The first fills for methanol in the Rectisol unit and MDEA in the Tail Gas Treatment unit are included in the Capital Cost Estimate. The following annual makeup costs apply to these chemicals. The per-unit costs are from previous projects.

- 120,000 gallons of methanol per year is required for Case 1.. Case 2 requires 100,000 gallons per year.
- The makeup rate for MDEA (for both cases) is 2,000 lb/year.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

In addition, caustic is consumed on an hourly basis in the process. The following annual makeup costs apply to caustic. The per unit costs are from previous projects.

- 578,160 gallons of 20 wt% caustic (for both cases) is consumed per year.

10.2.5 Maintenance Costs

For refineries and similar petrochemical-type plants, maintenance costs generally run between 2% and 8% of direct fixed capital costs of the plant facility. The annual expense for maintenance will depend upon the amount of rotating equipment installed as well as items that require frequent cleaning such as exchangers and change out of catalyst material and adsorbent media. An annual percentage charge of 2% was selected to account for the amount of rotating equipment, and required change outs for the shift catalyst, mercury removal adsorbent and Claus/TGTU catalysts.

10.2.6 Disposal Costs

A WorleyParsons Environmental manager provided the estimated cost for the disposal of Non-Hazardous Class 1 filter cake, and the disposal/transfer estimated cost for waste water at the unit battery limits.

10.2.7 Miscellaneous Operating Costs

Miscellaneous supplies include office supplies, office equipment, laboratory supplies, and operating supplies (non repair).

Miscellaneous contracts apply to analytical testing.

**ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT**

11. RESULTS AND RECOMMENDATIONS

The Syngas Treatment Area is capable of processing the HyMelt syngas to produce the required products at the specified conditions. These products for Case 1 (Hydrogen Production) are as follows:

- Hydrogen 52.9 MMSCFD
- Carbon Dioxide 2,324 STPD (40.1 MMSCFD)
- Sulfur 15.0 LTPD
- Fuel Gas 101.5 MM Btu/hr

The products for Case 2 (Fischer-Tropsch Liquids Production) are as follows:

- F-T Syngas Feed 57.3 MMSCFD with 2.1 H₂:CO ratio
- Hydrogen 1.5 MMSCFD
- Carbon Dioxide 1,262 STPD (21.8 MMSCFD)
- Sulfur 15.0 LTPD
- Fuel Gas 23.1 MM Btu/hr

The following is a list of recommendations for action or future study and consideration.

- After selection of site;
 - Revise Basis of Design and H&MB to reflect site conditions. Revise equipment sizing as needed.
 - Review need for CO₂ Thermal Oxidizer.
 - Investigate local carbon sequestration options and requirements.
 - Review need for Tail Gas Treating Unit (depending on local sulfur recovery requirements).
- After final determination of hydrogen user has been made, consider a Liquid Nitrogen Wash system instead of a PSA to purify the hydrogen for ammonia production.
- Investigate the potential market for pure CO and for syngas with low H₂:CO ratios close to the syngas from the HyMelt Molten Metal Gasifier. For example:
 - The carbonylation of methanol with CO to make acetic acid.

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

- The hydroformulation (Oxo process) of ethylene, CO, and H₂ to make propionaldehyde which is oxidized to make propionic acid.
 - The production of phosgene from CO and Cl₂ (used in making polycarbonates).
- After further development of Gasification Area startup and shutdown requirements, integrate these requirement with the downstream Syngas Treating system and the BOP.
- Review heat integration opportunities between the Gasification Area, the Syngas Treatment Area and the Balance of Plant. The low level heat currently used by the Utility Water Heater is one such opportunity for integration.
- Revise plot plan to include improved estimated allowance for Gasification Area and the Utility / Balance of Plant. Estimation of Gasification and BOP plot requirements by others.
- Investigate production of superheated steam in the SRU (i.e. steam from waste heat recovery boiler could be superheated in the Thermal Oxidizer).
- After selection of the Fischer-Tropsch catalyst and operating conditions, optimize the battery limit pressure and temperature of the clean, partially shifted syngas to Fischer-Tropsch.

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendices

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendix 1

ENVIRES LLC

Basis of Design (Rev. D)

Syngas Treatment Study for ICRC Coal Gasification to Hydrogen and Liquids

31-Aug-09

Hydrocarbons

WorleyParsons
6330 West Loop South
Bellaire, TX 77401, USA
Telephone: 713.407.5000
Facsimile: 713.407.5089
www.worleyparsons.com

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

SYNOPSIS

This document presents the Basis of Design (BoD) for the Syngas Treatment Study for ICRC Coal Gasification to Hydrogen and Liquids Project. This document is also used to define the scope of work of the study and to document the approach for completing the Syngas Treatment Study.

Disclaimer

This report has been prepared on behalf of and for the exclusive use of EnviRes LLC, and is subject to and issued in accordance with the agreement between EnviRes LLC and WorleyParsons. WorleyParsons accepts no liability or responsibility whatsoever for it in respect of any use of or reliance upon this report by any third party.

"Copyright © 2009 EnviRes LLC and WorleyParsons. This paper is written with support of the U.S. Department of Energy under Contract No. DE-FC26-06NT42449. The Government reserves for itself and others acting on its behalf a royalty-free, nonexclusive, irrevocable, worldwide license for Governmental purposes to publish, distribute, translate, duplicate, exhibit, and perform this copyrighted paper".

PROJECT 53869201 - BASIS OF DESIGN (REV. D)

REV	DESCRIPTION	ORIG	REVIEW	WORLEY- PARSONS APPROVAL	DATE	CLIENT APPROVAL	DATE
A	For Client Review	<u> </u> J. Hilton	<u> </u> J. White	<u> </u> S. Marr	09-Mar-09	<u> </u>	
B	Update	<u> </u> J. Hilton	<u> </u> J. White	<u> </u> S. Marr	10-Apr-09	<u> </u>	
C	Update	<u> </u> J. Hilton	<u> </u> J. White	<u> </u> S. Marr	24-Jun-09	<u> </u>	
D	Update for Case 2	<u> </u> J. Hilton	<u> </u> J. White	<u> </u> S. Marr	31-Aug-09	<u> </u>	

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

CONTENTS

CONTENTS	3
FIGURES	7
TABLES 7	
1. INTRODUCTION	8
1.1 Purpose	8
1.2 Project Background	8
1.3 Process Overview	8
1.4 Overall Process Description	9
1.4.1 Case 1 - Hydrogen Production	9
1.4.2 Case 2 – FT Liquids Production	11
2. SCOPE OF SERVICES	13
2.1 Design Approach	13
2.2 References & Standards	13
2.3 Special Resources	13
2.4 Deliverables to Client	13
2.5 Client Interface & Interdiscipline Coordination	14
2.6 Assumptions & Clarifications	14
3. SCOPE OF FACILITY	15
3.1 Feed Composition	15
3.2 Product Specifications	16
3.2.1 Hydrogen Product (for Case 1)	16
3.2.2 Clean Syngas to FT (for Case 2)	16
3.2.3 Hydrogen to FT (for Case 2)	17
3.2.4 Carbon Dioxide for EOR	17
3.2.5 Sulfur Product	18
3.2.6 Fuel Gas Product	19
3.3 Key Internal Stream Specifications	19
3.3.1 Scrubbed Syngas to CO Shift	19
3.3.2 Shifted Syngas to AGR	20

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

3.3.3	Acid Gas to SRU	20
3.3.4	Clean Syngas to Hydrogen Purification	21
3.4	Turndown Requirements	22
3.5	On-stream Factor	22
3.6	Design Life	22
3.7	Equipment Sparing Requirements	23
3.8	Driver Preferences	23
3.9	Special Client Requirements	23
3.10	Gasification Feedstock - Coal.....	24
4.	SITE CONDITIONS.....	26
4.1	Meteorological Data	26
4.1.1	Sources	26
4.1.2	Wind	26
4.1.3	Air Temperature	26
4.1.4	Relative Humidity (period of record – 96 years)	27
4.1.5	Barometric Pressure (Houston) (period of record – 21 years)	27
4.1.6	HVAC Design Conditions	27
4.1.7	Precipitation (Galveston)	27
4.1.8	Wind Load	29
4.1.9	Seismic.....	29
4.1.10	Additional Environmental Conditions	29
4.1.11	Soil Conditions.....	29
4.1.12	Topographical Condition	30
4.1.13	Tidal Data	30
4.1.14	Geotechnical.....	30
4.2	Environmental Data - Air Quality Standards.....	34
4.2.1	Ambient Air Quality Standards	34
4.2.2	Air Pollution Source Standards.....	34
4.2.3	Air Pollution Sources Subjected to Continuous Emission Monitoring.....	34
4.3	Emission Source Control	34
4.4	Fugitive Emission Monitoring.....	34

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

5.	UTILITY INFORMATION	35
5.1	Utility Battery Limit Conditions	35
5.1.1	Steam Systems	35
5.1.2	Condensate Systems	35
5.1.3	Feed Water Systems	35
5.1.4	Cooling Water Systems	36
5.1.5	Fuel Gas	36
5.1.6	Natural Gas	36
5.1.7	Air and Nitrogen	37
5.1.8	Chemicals	37
5.2	Natural Gas	37
5.3	Raw Water/Demineralized Water Analysis	38
5.4	Waste Water Pretreatment Requirements	39
5.5	Storm Water Quality Requirements	39
5.6	Electrical	39
6.	UNITS OF MEASUREMENT	40
6.1	Scope	40
6.2	Definition	40
6.3	English (U.S.) Units and Symbols	40
6.3.1	English (U.S.) Units	40
6.3.2	Prefixes	41
6.4	Dimensions	41
6.4.1	Drawings	41
6.5	Style	41
6.6	Mathematical Notation	42
7.	NUMBERING PROCEDURE	43
7.1	Scope	43
7.2	Unit Numbers	43
7.3	Equipment Numbering	44
7.3.1	Equipment Letter Designations	44
7.3.2	Equipment Number Examples	44

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

7.4	Drawing Numbering.....	45
7.4.1	Drawing Number Format.....	45
8.	ECONOMIC DATA.....	46
8.1	Feedstock & Product Values	46
8.2	Utility Costs.....	46
9.	ABBREVIATIONS AND ACRONYMS.....	47
	APPENDIX A - PREVAILING WINDS.....	50
	APPENDIX B - RAINFALL CURVES	51
	APPENDIX C - PRECIPITATION DEPTH.....	52

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

FIGURES

Figure 1-1 Case 1 Block Flow Diagram	10
Figure 1-2 Case 2 Block Flow Diagram	12

TABLES

Table 3-1 Syngas Feed Specification	15
Table 3-2 Hydrogen Product (Case 1) Specification	16
Table 3-3 Clean Syngas Product to FT (Case 2) Specification	16
Table 3-4 Hydrogen Product to FT (Case 2) Specification	17
Table 3-5 Carbon Dioxide Product Specification	18
Table 3-6 Sulfur Product Specification	18
Table 3-7 Fuel Gas Product Specification	19
Table 3-8 Scrubbed Syngas to CO Shift (Case 1 and 2)	19
Table 3-9 Shifted Syngas to AGR (Case 1 and 2)	20
Table 3-10 Acid Gas to SRU (Case 1 and 2)	20
Table 3-11 Clean Syngas to Hydrogen Purification (Case 1 and 2)	21
Table 3-12 Equipment & Piping Design Life	22
Table 3-13 Equipment Sparing Requirements	23
Table 3-14 EnviRes Coal Specification	24

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

1. INTRODUCTION

1.1 Purpose

This document establishes the Basis of Design (BoD) for the Syngas Treatment Study for the ICRC Coal Gasification to Hydrogen and Liquids Project. This document establishes a process design basis for use by all engineering disciplines, project management, and client personnel.

1.2 Project Background

ICRC, a wholly owned subsidiary of VSE Corporation, is under contract to the US Department of Energy, National Energy Technology Laboratory (NETL) and the State of Kentucky for a coal gasification to liquids and chemicals FEL level 2 Assessment. EnviRes LLC is providing the Gasification Technology (i.e. HyMelt Molten Iron Gasification) and the Syngas Treatment section. WorleyParsons shall provide technical analysis, process data and cost estimating for the Syngas Treatment section.

Two cases have been defined for the Syngas Treatment Study. In Case 1, Hydrogen is the main plant product. Other products are Sulfur, Carbon Dioxide for EOR and Fuel Gas. In Case 2, Fischer-Tropsch (FT) liquids are the main plant product. Other products are Sulfur, Carbon Dioxide for EOR and Fuel Gas.

1.3 Process Overview

The following description of the scope of facilities defines the Syngas Treatment section of the ICRC Coal Gasification to Hydrogen and Liquids Project. Refer to Figure 1-1 for the Block Flow Diagram for Case 1 and to Figure 1-2 for the Block Flow Diagram for Case 2.

The Syngas Treatment section consists of the following main sub-sections:

- Syngas Quench and Scrubbing
- Shift Reaction and Cooling
- Acid Gas Removal (AGR)
- Hydrogen Purification
- Sulfur Recovery Unit (SRU)
- CO₂ Compression
- Syngas Compression (Case 2 only)

Feedstock Handling, HyMelt Gasification and Slag Handling are outside of the WorleyParsons scope and are defined by EnviRes.

The Air Separation Unit (ASU) and Balance of Plant (i.e. Utilities and Offsites) are outside of the WorleyParsons scope and are defined by others.

For Case 2, FT Liquids, the FT Reaction and Upgrading section is outside of the WorleyParsons scope and is defined by others.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

1.4 Overall Process Description

1.4.1 Case 1 - Hydrogen Production

Hot, raw syngas from the EnviRes HyMelt Gasifier enters the Quench & Scrubbing system of the Syngas Treatment area. The syngas also contains dust from the Gasifier. This solid dust material consists of small iron and carbon particles. The hot raw syngas is first water quenched in a Quench Drum which cools and saturates the syngas. Most of the dust material is removed with the water blowdown stream. The syngas is further scrubbed as it is mixed with additional water in the Venturi Mixer. The Syngas Scrubber acts as a final scrubbing step to remove the remaining particulate material. The quench bottoms water is flashed to reduce the pressure and to remove dissolved gas components. This stream is routed to a settler where the particulate material is concentrated before being filtered. The filter cake is then routed to the Syngas Treatment battery limits for disposal. The overflow from the settler and the filtrate from the filters are combined in a tank. Most of the water is recycled back to the quench and scrubbing system. A purge stream of water is blown down from the tank to the Syngas Treatment battery limits for control of chlorides and/or other impurities.

The saturated syngas flows into the Shift Reaction & Cooling system. In the Shift Reactors the carbon monoxide (CO) in the syngas is reacted with water to form hydrogen (H₂) and carbon dioxide (CO₂). Sour shift catalyst shall be utilized. Three stages of shift shall be employed to maximize the conversion of CO to H₂. Additional water (i.e. steam) is added to the syngas to the first shift reactor to improve conversion and reduce the reactor exit temperature. After shifting, the syngas is cooled in a series of heat exchangers. The syngas also passes through a mercury removal drum before going to the AGR section. Process condensate from the shifted syngas is recovered and reused in Quench & Scrubbing system.

The shifted syngas is next treated in an Acid Gas Removal system where acid gas components (H₂S and CO₂) are removed. A Rectisol unit has been used for this application,. The syngas is first contacted with lean solvent (methanol) in an absorber column to remove the H₂S and CO₂. The bulk of the CO₂ removed from the syngas is concentrated into a CO₂ product stream. The H₂S and the remaining CO₂ are stripped from the solvent and leave the AGR in an acid gas stream.

The clean raw hydrogen stream from the AGR is purified in the Hydrogen Purification system using a Pressure Swing Adsorption (PSA) unit. The impurities are captured on the adsorbent allowing the essentially pure H₂ stream to flow through the beds. During the desorption part of the cycle, the impurities are desorbed from the beds and blending together as PSA off-gas. The off-gas stream shall be compressed and routed to the Syngas Treatment battery limits as fuel gas. This PSA fuel gas shall be suitable for combustion.

The acid gas stream is treated in a Claus Sulfur Recovery Unit (SRU) to produce elemental sulfur. The molten sulfur is degassed and pumped to the Syngas Treatment battery limits. Truck or rail loading facilities needed are part of the Balance of Plant (BOP) section. The residual tail gas is treated and routed to the inlet of the SRU.

The raw CO₂ stream from the AGR is routed to the CO₂ Compression system. The CO₂ shall be compressed and routed to the Syngas Treatment battery limits. The CO₂ stream shall be suitable for use in EOR in a sweet field. If necessary, the CO₂ stream shall be treated for removal of water and other impurities either before or during compression.

All necessary utilities shall be provided to the Syngas Treatment section from the Balance of Plant (BOP) section.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Figure 1-1 Case 1 Block Flow Diagram

**BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS**

1.4.2 Case 2 – FT Liquids Production

Most of the process flow is the same for Case 2. The main difference is that the scrubbed syngas stream is split and not all of the syngas is routed through the shift reactors. Part of the syngas bypasses the shift reactors to allow control of the H₂ to CO ratio of the syngas stream. Only two shift reactors are needed for Case 2. The shifted syngas to the AGR has a higher CO content for Case 2 than for Case 1.

The Rectisol Acid Gas Removal system removes the acid gas components (H₂S and CO₂) from the syngas. The acid gas stream is treated in a Claus Sulfur Recovery Unit (SRU) to produce elemental sulfur. The molten sulfur is degassed and pumped to the Syngas Treatment battery limits.

The clean syngas leaving the AGR is compressed and any residual sulfur is removed from the clean syngas in a final polishing step in the Syngas Compression system. This clean, “polished” syngas leaves the Syngas Treatment section as product and is routed to the FT Reaction and Upgrading section.

A small part of the compressed syngas is routed to the Hydrogen Purification system. A small PSA is used to produce H₂ needed to hydrotreat the FT liquids. As in Case 1, the PSA also produces a fuel gas stream for use outside of the Syngas Treatment section.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Figure 1-2 Case 2 Block Flow Diagram

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

2. SCOPE OF SERVICES

2.1 Design Approach

The following design approach outlines the engineering design services provided by WorleyParsons for the Syngas Treatment Study.

The following items are applicable to both Case 1 and Case 2.

- Develop Block Flow Diagram
- Evaluate removal strategies for syngas impurities
- Evaluate alternative flow schemes for CO Shift and AGR
- Develop Process Simulation
- Prepare Heat & Material Balance (H&MB) table
- Prepare Process Flow Diagrams (PFDs)
- Develop equipment sizing information
- Prepare Sized Equipment List
- Prepare estimate of Operating Cost (OpEx)
- Prepare Capital Cost Estimate (+/- 25 %)

2.2 References & Standards

For the Syngas Treatment Study, WorleyParsons practices and procedures shall be used unless otherwise specified in this document.

2.3 Special Resources

EnviRes shall provide simulation data for the quenched syngas developed using their FactSage software. WorleyParsons shall incorporate this data into the process simulations developed for the Syngas Treatment section.

2.4 Deliverables to Client

The following items shall be provided to the client in draft and final report format and are applicable to both Case 1 and Case 2.

- Block Flow Diagrams
- Process Descriptions
- Heat & Material Balance (H&MB) tables
- Process Flow Diagrams (PFDs)
- Sized Equipment Lists

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

- Estimate of Operating Cost (OpEx)
- Capital Cost Estimate

2.5 Client Interface & Interdiscipline Coordination

Refer to the Project Execution Plan (PEP) for this information.

2.6 Assumptions & Clarifications

No additional items have been identified.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

3. SCOPE OF FACILITY

The Project Name is **Syngas Treatment Study for ICRC Coal Gasification for Hydrogen and Liquids**. PFDs shall use the following in the title block: EnviRes Coal/Biomass Gasification Project. The plant location shall be shown as “US Gulf Coast” for this study.

3.1 Feed Composition

The feed to the Syngas Treatment System is the hot, raw syngas from the Gasification system. Its composition has been provided by EnviRes at the Kickoff Meeting and has been revised based on the latest coal selection. The syngas flow rate and composition are the same for both cases. The syngas also contains “dust” carried over from the Gasifiers and was specified by EnviRes.

Table 3-1 Syngas Feed Specification

Syngas Composition	vol %	ppmv
Hydrogen	30.38	
Carbon Monoxide	66.29	
Carbon Dioxide	1.00	
Methane	0.10	(961)
Hydrogen Sulfide	0.60	
Carbonyl Sulfide	0.04	(417)
Ammonia	0.00	(15)
Hydrogen Cyanide	0.01	(83)
Hydrogen Chloride	0.03	(266)
Argon	0.30	
Nitrogen	0.58	
Water	0.67	
Syngas Flow rate (Note 1,2)	6,837.41	lbmol/hr
Temperature	2732	°F
Pressure	500	psig
<p>Note 1: The syngas contains 1,683 lb/hr of solid “dust” material with a composition that is 65 wt% Iron and 35 wt% Carbon.</p> <p>Note 2: The syngas also contains 2.39 lbmol/hr (78.25 lb/hr) of trace elements that will be soluble in the quench water stream leaving the gasifier.</p>		

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

3.2 Product Specifications

3.2.1 Hydrogen Product (for Case 1)

A typical specification for Hydrogen for Ammonia Synthesis is shown. Additional inerts (nitrogen and argon) may be allowed but would impact the design of the ammonia synthesis and recycle. A sulfur guard system (typically ZnO adsorbent) may also be required in the Ammonia Facility to further protect the catalyst.

Table 3-2 Hydrogen Product (Case 1) Specification

Component	
Hydrogen, vol % min.	99.0
Nitrogen plus Argon, vol % max	1.0
Methane, ppmv max	100
Combined O ₂ , CO, CO ₂ , ppmv, max	5.0
Water, ppmv max	1.0
Sulfur, ppmv max as S	1
Temperature	80 - 120 °F
Pressure	250 psig (minimum)

3.2.2 Clean Syngas to FT (for Case 2)

A typical specification for Hydrogen for FT Synthesis is shown. Minimization of inerts (nitrogen, argon and methane) is important for the design of the FT system. A sulfur guard bed (typically ZnO) has been included in the WorleyParsons scope to further protect the catalyst. The client expects that a cobalt FT catalyst will be used.

Table 3-3 Clean Syngas Product to FT (Case 2) Specification

Component	
Hydrogen + Carbon Monoxide, vol % min	97.0
Hydrogen to Carbon Monoxide Ratio	2.10
Inerts (Nitrogen, Argon, Methane), vol % max	1.0
CO ₂ , vol %, max	1.0
Water, vol %, max	1.0
Sulfur, ppbv max as S	20.0

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Component	
Ammonia + HCN, ppmv max	10.0
Chloride, ppmv max	1.0
Temperature	400 °F
Pressure	475 psig

3.2.3 Hydrogen to FT (for Case 2)

A typical specification for Hydrogen for FT Synthesis is shown. Additional inerts (nitrogen and argon) may be allowed but would impact the design of the FT system. A base metal hydrotreating catalyst has been assumed. The flow rate has been provided by EnviRes.

Table 3-4 Hydrogen Product to FT (Case 2) Specification

Component	
Hydrogen, vol % min.	99.0
Nitrogen plus Argon, vol % max	1.0
Methane, ppmv max	100
Combined O ₂ , CO, CO ₂ , ppmv, max	5.0
Water, ppmv max	1.0
Sulfur, ppmv max as S	1
Flow rate	1.50 MMSCFD
Temperature	80 - 120 °F
Pressure	250 psig (minimum)

3.2.4 Carbon Dioxide for EOR

For this study, a carbon dioxide stream will be produced that is suitable for injection into sweet fields for enhance oil recovery (EOR). The carbon dioxide stream will be treated (i.e. dried) as needed and compressed to be transported by pipeline.

The carbon dioxide must also be suitable for venting to atmosphere if the CO₂ compression section is not operating. This may further limit the carbon monoxide and/or sulfur content of the stream.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Table 3-5 Carbon Dioxide Product Specification

Component	
Carbon Dioxide, vol % min	95.0
Nitrogen, vol % max	4.0
Methane and other HCs, vol % max	5.0
CO, vol %, max	1.0
Water, lb/MMscf max (no free water)	30.0
Sulfur, ppmv max as S	20.0
Oxygen, ppmv max	10.0
Glycol, ppmv max (no free glycol)	40.0
Temperature	80 - 120 °F
Pressure	2,200 psig

3.2.5 Sulfur Product

For this study, elemental sulfur will be produced. The SRU will produce a bright yellow, commercial grade sulfur to the following specification.

Table 3-6 Sulfur Product Specification

Component	Value
Purity	99.9 wt% (min on dry basis)
Appearance (as Solid)	Bright yellow
H ₂ S	10 ppmw (max)
Acidity as H ₂ SO ₄	50 ppmw (max)
Chlorides as Cl	10 ppmw (max)
Ash	500 ppmw (max)
Carbon matter	500 ppmw (max)
Arsenic	None if none in the feed
Selenium and Tellurium	None if none in the feed
Moisture	Nil

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

3.2.6 Fuel Gas Product

The offgas from the PSA system in the Hydrogen Purification section must be suitable for combustion in fired equipment (outside the scope of this study). Additional fuel gas or natural gas may be needed in order to combust the offgas.

Table 3-7 Fuel Gas Product Specification

Component	
Sulfur, ppmv max as S	10.0
Temperature	80 - 120 °F
Pressure	50 psig

3.3 Key Internal Stream Specifications**3.3.1 Scrubbed Syngas to CO Shift****Table 3-8 Scrubbed Syngas to CO Shift (Case 1 and 2)**

Syngas Composition	Case 1 and 2, vol %
Hydrogen	14.09
Carbon Monoxide	30.72
Carbon Dioxide	0.48
Methane	0.04
Hydrogen Sulfide	0.28
Carbonyl Sulfide	0.02
Ammonia	0.01
Hydrogen Cyanide	0.00
Hydrogen Chloride	0.00
Argon	0.14
Nitrogen	0.27
Water	53.95
Syngas Flow rate	14,478.17 lbmol/hr

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Temperature	401 °F
Pressure	475 psig

3.3.2 Shifted Syngas to AGR

Table 3-9 Shifted Syngas to AGR (Case 1 and 2)

Composition	Case 1, vol %	Case 2, vol %
Hydrogen	57.80	48.55
Carbon Monoxide	1.01	23.25
Carbon Dioxide	39.89	26.70
Methane	0.06	0.07
Hydrogen Sulfide	0.39	0.46
Carbonyl Sulfide	0.00	0.01
Ammonia	0.00	0.00
Hydrogen Cyanide	0.00	0.00
Hydrogen Chloride	0.00	0.00
Argon	0.18	0.22
Nitrogen	0.35	0.43
Water	0.32	0.30
Flow rate	11,235.72 lbmol/hr	9,203.17 lbmol/hr
Temperature	110 °F	110 °F
Pressure	312 psig	342 psig

3.3.3 Acid Gas to SRU

Table 3-10 Acid Gas to SRU (Case 1 and 2)

Composition	Case 1, vol %	Case 2, vol %
Hydrogen	1.52	1.02
Carbon Monoxide	0.08	1.38
Carbon Dioxide	60.00	60.00

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Methane	0.01	0.01
Hydrogen Sulfide	38.37	36.57
Carbonyl Sulfide	0.02	1.01
Ammonia	0.00	0.00
Hydrogen Cyanide	0.00	0.00
Hydrogen Chloride	0.00	0.00
Argon	0.00	0.00
Nitrogen	0.01	0.01
Water	0.00	0.00
Flow rate	112.90 lbmol/hr	115.45 lbmol/hr
Temperature	80 °F	80 °F
Pressure	45 psig	45 psig

3.3.4 Clean Syngas to Hydrogen Purification

Table 3-11 Clean Syngas to Hydrogen Purification (Case 1 and 2)

Composition	Case 1, vol %	Case 2, vol %
Hydrogen	97.10	67.00
Carbon Monoxide	1.69	31.89
Carbon Dioxide	0.22	0.12
Methane	0.09	0.10
Hydrogen Sulfide	0.00	0.00
Carbonyl Sulfide	0.00	0.00
Ammonia	0.00	0.00
Hydrogen Cyanide	0.00	0.00
Hydrogen Chloride	0.00	0.00
Argon	0.31	0.31
Nitrogen	0.59	0.59

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Water	0.00	0.00
Flow rate	6,681.97 lbmol/hr	374.42 lbmol/hr
Temperature	80 °F	110 °F
Pressure	270 psig	485 psig

3.4 Turndown Requirements

The plant shall be designed to operate at 50% rate for an indefinite time without flaring.

3.5 On-stream Factor

The design on-stream factor shall be 347 days per year of operation (8322 hours/year). The plant shall be designed recognizing the intent of having a full plant turnaround every two years. Key un-spared equipment should be designed to operate without a planned shutdown within this timeframe.

3.6 Design Life

The plant shall be designed basis the following table.

Table 3-12 Equipment & Piping Design Life

	Years
Major Equipment	
Columns, Reactors, Vessels, Integral Internals	20
Replaceable Internals without Heat Treat	10
Thick Alloy Equipment (Above 2")	
Reactors	30
Exchanger Shells and Channels	20
Tube Bundles and Furnace Tubes	10
Minor Vessels, less than 400 cu ft volume	
Ordinary Wall Thickness	10
Thick Walled (Above 2")	20
Heat Exchanger Bundles, Conventional Thickness	
Carbon Steel or Brass Alloy	5
Stainless or High Alloy	10
Piping	10

)

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

3.7 Equipment Sparing Requirements

The general sparing requirements are shown in the table below.

Table 3-13 Equipment Sparing Requirements

Equipment Service	Dedicated Spare	Shared Spare	No Spare
Process Centrifugal Compressors	-	-	X
Process Reciprocating Compressors	100%	-	-
Utility Compressors	100%	-	-
ISBL Process Pumps	100%	-	-
Offsite Feed Pumps	100%	-	-
Offsite Transfer Pumps	100%	-	-
BFW Pumps	50%	-	-
Utility Water Pumps	50%	-	-
Firewater Pumps	33%	-	-
Seawater Pumps	25%	-	-
Cooling Water Pumps	50%	-	-
Treated Water Pumps	[Hold]	-	-
Potable Water Booster Pumps	100%	-	-

For this project, the following compressors shall NOT be spared.

- CO₂ Compressor
- PSA Fuel Gas Compressor
- Syngas Compressor

3.8 Driver Preferences

The general philosophy for non-critical equipment is that in standard applications, electrical motors are preferred for main equipment and spares, and diesel engines for main firewater pumps. Steam turbines are preferred for services that are required to be kept on-line during power outages. There should be enough steam turbine drivers to safely shed and balance the plant steam systems during upsets and outages.

For the purposes of this study, electric drivers shall be assumed for all equipment in the Syngas Treatment section.

3.9 Special Client Requirements

The customer has designated several required 'givens' in the design, which will be incorporated in the design. These include:

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

- Maximize the recovery of Hydrogen
- Process condensate shall be reused to the maximum extent practical.
- Minimize fresh water usage
- There will be no routine flaring of sour gas. Sour gas may be flared during startup, shutdown and operational upsets or emergencies

3.10 Gasification Feedstock - Coal

Although the feed to the Gasification system is not the feed to the Syngas Treatment Study, it is an important factor in the design especially as related to the impurities in the syngas. Therefore, the coal feedstock selected for the Gasification system is listed in this section.

EnviRes' material balance for the syngas from the HyMelt gasifier was based on two percent moisture to the gasifier.

The normal coal selected for the project is a Kentucky #11 coal. The preliminary properties of the coal based on information from the USGS are as follows.

Table 3-14 EnviRes Coal Specification

Proximate Analysis	Kentucky #11 Coal Wt %
Moisture	7.07
Ash, as recv'd	8.49
Volatile, as recv'd	38.88
Fixed C, as recv'd	45.56

Ultimate Analysis	Wt %
Carbon	66.37
Hydrogen	5.43
Nitrogen	1.33
Sulfur	4.06
Ash	8.49
Oxygen (by difference)	14.31

Heating Value	Btu/lb
----------------------	---------------

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Gross, as recv'd	12,058
Hardgrove Grindability Index	
as recv'd	58
Ash Fusion Temperature	°F
Reducing - ID	2,015
Reducing - Softening	2,071
Reducing - Fluid	2,265

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

4. SITE CONDITIONS**4.1 Meteorological Data****4.1.1 Sources**

Local Climatological Data, Annual Summaries for 1993, NOAA
Climatic Averages and Extremes for U.S. Cities, June 1995, NOAA
Rainfall Frequency Atlas of U.S. Technical Paper No. 40, May 1961
1989 ASHRAE Handbook Fundamentals
ASCE 7-95 Minimum Design Loads for Buildings and Other Structures

4.1.2 Wind

Direction of Annual Prevailing Winds: SSE¹
Maximum Observed Sustained Velocity: 100 mph (Sept. 1900) (period of record - 122 yrs.)
Mean Annual Velocity: 11.0 mph (period of record - 93 yrs.)
Basic Wind Speed (3-sec. gust speed, 50-yr recurrence interval): 115 mph

4.1.3 Air Temperature

Annual Normal Mean (1961-1990):	69.7 °F
Coldest Month Average (1964-1993):	48.3 °F (Jan)
Annual Mean Daily Max. (1964-1993):	74.7 °F
Annual Mean Daily Min. (1964-1993):	65.0 °F
Max. Recorded (period of record - 123 yrs):	101 °F (July 1932)
Lowest Recorded (period of record - 123 yrs):	8 °F (Feb. 1899)
Mean Daily Max.: Summer (Jun-Aug):	86.7 °F
Mean Daily Min.: Summer (Jun-Aug):	78.5 °F
Design Basis: Maximum:	101 °F
Minimum:	8 °F
Winter (Dec-Feb):	60.2 °F
Winter (Dec-Feb):	49.3 °F

¹ See Appendix A – Prevailing Winds

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

4.1.4 Relative Humidity (period of record – 96 years)

Annual Average:	83%
High mean of monthly max:	86% (April)
Low mean of monthly max:	65% (October)
Design Basis: Max:	100%, Min: 30%

4.1.5 Barometric Pressure (Houston) (period of record – 21 years)

Design barometric pressure:	14.7 psia
Annual Average:	14.7 psia
High mean of monthly max:	14.75 psia (Jan)
Low mean of monthly max:	14.6 psia (May)
Max rate of change in barometric pressure:	N/A

4.1.6 HVAC Design Conditions
Galveston:

Average dry bulb:	Summer: 90 °F (mean 1%)	Winter: 31 °F (mean 99%)
Average wet bulb:	Summer: 79 °F (mean 1%)	Winter: N/A

Houston:

Average dry bulb:	Summer: 96 °F (mean 1%)	Winter: 27°F (mean 99%)
Average wet bulb:	Summer: 80 °F (mean 1%)	Winter: N/A

4.1.7 Precipitation (Galveston)
Rainfall

Normal Annual (1961-1990):	42.28 inches
----------------------------	--------------

Historical (period of record - 123 yrs)

Maximum rainfall in 24 hours:	14.35 inches (Jul 1900)
Maximum rainfall in 1 month:	26.01 inches (Sep 1885)

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

2 year frequency²

Maximum rainfall in 5 minutes:	0.58 inches
Maximum rainfall in 15 minutes:	1.22 inches
Maximum rainfall in 1 hour:	2.39 inches

100 year frequency²

Maximum rainfall in 5 minutes:	0.91 inches
Maximum rainfall in 15 minutes:	2.02 inches
Maximum rainfall in 1 hour:	4.62 inches

10 year frequency

Maximum rainfall in 1 hour:	3.50 inches
Maximum rainfall in 2 hours:	4.50 inches
Maximum rainfall in 3 hours:	4.90 inches
Maximum rainfall in 6 hours:	6.05 inches
Maximum rainfall in 12 hours:	7.60 inches
Maximum rainfall in 24 hours:	8.80 inches

Annual Average Number of Rainy Days (0.01 inches): 95.8

(period of record - 122 yrs)

Thunderstorm Frequency - Days per Year: 36

(period of record - 1 yr)

Snowfall:

Annual Average: 0 inches

(period of record - 122 yrs)

Maximum Monthly: 15.4 inches (Feb 1895)

(period of record - 122 yrs)

Snow on Ground: From: N/A To: N/A

Design Snow Loading: N/A lb/ft²

Depth of Frost Line: 0 inches

² See Appendix B - Rainfall curves. Compute additional times and frequencies in accordance with NWS Hydro-35.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

4.1.8 Wind Load

Refer to PIP STC01015 Structural Design Criteria, also ASCE 7-05 Minimum Design Loads for Buildings and Other Structures.

4.1.9 Seismic

Zone 0 per UBC. Seismic coefficient = 0

4.1.10 Additional Environmental Conditions

Specific atmospheric conditions at site which may cause corrosion and contamination.

Dust:	N/A
Fumes:	N/A
Climate:	Semi-Tropical
Saline Air:	Marine Exposure (Cloud Drift)
Sandstorm:	No
Temperature Inversion:	Possible
Max. 100 yr Flood Level:	N/A
Design Sea Water Temp.:	87 °F
Max Allowable Seawater Discharge Temp.:	°F
Susceptibility to Storm Surge:	None
Height of Storm Surge:	N/A

4.1.11 Soil Conditions

Soil investigation Reports	Available
Corrosivity (M&QS)_____Rural: X Industrial:_____Mine:_____Industrial/Marine:	
Buried Line Treatment	Required
Presence of Borers, Termites	Yes
0-bound Water Level	18 ft
Underground Obstructions	None
Nature and Description	Silty Clay to Silty Sand
Resistivity, ohm-cm @ depth	1980 @ 2 ft, 260 @ 10 ft
Soil Bearing	2,500 psf
Soil pH	7.0

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

4.1.12 Topographical Condition

Plant Map - Drawing No.	See Project Plot Plant - Later
Topographical Map - Drawing No.	Flat
Elevation Above Mean Sea Level	20 ft
Coordinates of Battery Limits for Project	See Project Plot Plant - Later
Nature of Terrain	Flat
Site Drainage	Available

4.1.13 Tidal Data

(Refer to Ft. VG VD Mean Sea Level = Mean Tide = +0.05' NG VD)	
Maximum High Tide, ft (Astronomical)	1.05
Normal High Tide, ft - Mean High High	0.63
Mean High Water, ft - Mean High	0.54
Normal Low Tide, ft - Mean Low	-0.43
Mean Low Water, ft - Mean Low Low	-0.78
Minimum Low Water, ft - (Astronomical)	-1.55
Hurricane Tidal Effect, ft (Highest recorded)	13.42
Hurricane Tidal Effect, ft (Lowest recorded)	- 6.38
Maximum Wave Height Depth - Limited Site Specific =	0.8 day, Max
Velocity & Direction of Current	TBD
100 Yr Base Flood Elevation, ft	14-16

4.1.14 Geotechnical

The following civil/structural assumptions will be used for the capital cost estimating. These assumptions are for conceptual level estimating and will be used until a formal geotechnical report is generated and formal design criteria are established:

Design of Foundations:

1. Soil-supported mat foundations and spread footings will be assumed for all equipment and structures including for pipe bridges, switchyard structures, pre-engineered buildings and

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

enclosures for electrical and mechanical equipment, sumps, and other structures with foundations having relatively low contact pressures.

2. Support for heavily loaded and/or settlement sensitive structures will assume an addition cost factored to be equivalent to deep foundation systems consisting of either 356 mm or 305 mm (14-inch or 12-inch) square precast prestressed concrete piles founded on the existing soil. Common reinforced concrete mats or individual concrete caps connected by concrete grade beams are assumed to connect and achieve load transfer from structures and equipment to the deep foundation system. The factored cost estimates assume the following allowable/design capacities for deep foundations:

- 1) 356 mm (14-Inch) Piles: To be used for boiler and turbine buildings, condensate storage tank, gasifier, F-T area, chimney, other large equipment and silo Vertical Loads
 - a) Compression: 890 kN (200 kips) per pile
 - b) Tension: 0 kN (0 kips) per pile except for extreme, temporary loads that will be handled on an individual basis as required by the Engineer
 - c) Lateral Loads

Structure or Area	Allowable Lateral Capacity kN (kips) per pile	
	Exterior Piles ³	Interior Piles ⁴
Boiler building and condensate storage tank	27 (6)	44 (10)
Turbine building	44 (10)	44 (10)
F-T and other areas	36 (8)	44 (10)
Gasifier Area	44 (10)	44 (10)
Chimney	7 (1.5)	53 (12)

³ Piles on the perimeter of a pile group, or piles which are greater than 1.8 m (6 feet) cc from adjacent piles.

⁴ Piles surrounded by at least 4 piles within 1.8 m (6 feet) cc.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

	Allowable Lateral Capacity kN (kips) per pile	
Silo	36 (8)	44 (10)

- 2) 305 mm (12-Inch) Piles: To be used for cooling tower, transformers, and other miscellaneous structures or equipment as determined by the Engineer Vertical Loads
- a) Compression: 623 kN (140 kips) per pile
 - b) Tension: 0 kN (0 kips) per pile except for extreme, temporary loads that will be handled on an individual basis as required by the Engineer.
 - c) Lateral Loads

	Allowable Lateral Capacity kN (kips) per pile	
Structure or Area	Exterior Piles ⁵	Interior Piles ⁶
Acid/Caustic Neutralized Waste Tanks	18 (4)	27 (6)
Cooling tower	31 (7)	36 (8)
Circulating Water Pump Structure	9 (2)	36 (8)
Diesel Generator Building	18 (4)	18 (4)
Limestone/Reagent/Fuel Prep Area	31 (7)	31 (7)

⁵ Piles on the perimeter of a pile group, or piles which are greater than 1.5 m (5 feet) cc from adjacent piles.

⁶ Piles surrounded by at least 4 piles within 1.5 m (5 feet) cc.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

	Allowable Lateral Capacity kN (kips) per pile	
Water Treatment Facility	31 (7)	36 (8)
Transformers	27 (6)	27 (6)

3. The effects of fluctuating ground water elevation will be not taken into account in this phase of estimate.
4. Uplift Conditions resulting in temporary uplift from environmental loads (wind, seismic) or permanent uplift from cantilevered construction will not be considered in this phase of estimate.
5. Frost protection for soil supported footings, pipes, and other frost-susceptible structures will not be considered for this application.
6. Minimum embedment in the soil for all foundations will be 760mm (30 inches). Minimum embedment for sidewalks, concrete pads at building doorways, and step off pads for ladders and stairways will be 152 mm (six inches).
7. The following soil parameters will be assumed in the design of foundations:

Unit weight for soils above water table	1762 - 2002 kg/m ³ (110-125 ⁷ pcf)
Unit weight of water	1000 kg/m ³ (62.4 pcf)
Unit weight of submerged soils	769 - 1003 kg/m ³ (48-62.6 ¹ pcf)
Passive earth pressure coefficient (k_p)	3.0
Active earth pressure coefficient (k_a)	0.33
At-rest pressure coefficient (k_o)	0.5
Coefficient of friction between concrete foundations and soil (for soil-supported foundations only)	0.5
Sub-Grade Modulus (for soil-supported foundations only)	1383995 kg/m ³ (50 pci)

⁷ Consider range of values and use worst case.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Shear wave velocity (soils above limestone)	244 m/s (800 ft/sec)
Poisson's ratio	0.2
Allowable bearing pressure (for soil-supported foundations only)	9764 kg/m ² (2.0 ksf)

8. Factors of safety for foundation design shall be as follows:

	<i>Normal</i>	<u>Transient or Construction</u>
Overturning:	1.5	1.25
Sliding:	1.5	1.25
Uplift:	1.25	1.1

4.2 Environmental Data - Air Quality Standards

4.2.1 Ambient Air Quality Standards

Title 30 of the Texas Administrative Code (30 TAC), Chapters 101-118

4.2.2 Air Pollution Source Standards

Title 30 of the Texas Administrative Code, Chapter 113

Title 40 of the Code of Federal Regulations (40 CFR), Parts 61 and 63

4.2.3 Air Pollution Sources Subjected to Continuous Emission Monitoring

Title 30 of the Texas Administrative Code, Chapters 101-118

Title 40 of the Code of Federal Regulations, Parts 70 and 71

4.3 Emission Source Control

Title 40 of the Code of Federal Regulations, Part 63

4.4 Fugitive Emission Monitoring

Title 30 of the Texas Administrative Code, Chapters 101-118

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

5. UTILITY INFORMATION

5.1 Utility Battery Limit Conditions

5.1.1 Steam Systems

The following table provides the steam conditions for the tabulated systems.

Minimum steam pressures and temperatures are the minimum conditions for each system which may be expected at any point in the area header during operation.

Normal steam pressures and temperatures are the average conditions which may be expected at any point in the area header during operation. Steam consumption calculations should be based on these values.

Design pressures and temperatures are the conditions specified for the mechanical design of piping and equipment excluding steam generation equipment.

System	Pressure psig				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
HP Steam	TBD	1,200	TBD		TBD	569	TBD	
MP Steam	TBD	650	TBD		TBD	497	TBD	
IP Steam	TBD	500	TBD		TBD	470	TBD	
LP Steam	TBD	75	TBD		TBD	320	TBD	

5.1.2 Condensate Systems

System	Pressure psig (1)				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
Condensate	TBD	45	TBD		TBD	TBD	TBD	
Note: (1) At battery limit at grade level								

5.1.3 Feed Water Systems

System	Pressure psig (1)				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
HP BFW	TBD	1,400	TBD		TBD	250	TBD	

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

System	Pressure psig (1)				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
LP BFW	TBD	800	TBD		TBD	250	TBD	
Utility Water								
Raw Water		50				80		
Potable Water								
Demin. Water	TBD	115	TBD		TBD	80	TBD	
Note: (1) At battery limit at grade level								

5.1.4 Cooling Water Systems

System	Pressure psig (1)				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
CW Supply	TBD	55	TBD		TBD	90	TBD	
CW Return	TBD	40	TBD		TBD	110	TBD	
Tempered Water								
Note: (1) At battery limit at grade level								

5.1.5 Fuel Gas

System	Pressure psig				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
Fuel Gas	TBD	50	TBD		TBD	100	TBD	

5.1.6 Natural Gas

System	Pressure psig				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
Pipe Line Natural Gas								
LP Natural Gas								

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS
5.1.7 Air and Nitrogen

System	Pressure psig				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
Instr. Air ⁸	TBD	110	TBD		TBD	80	TBD	
Service Air	TBD	100	TBD		TBD	80	TBD	
Plant Nitrogen	TBD	200	TBD		TBD	80	TBD	
Note : A higher pressure nitrogen stream may be needed for AGR stripping - TBD								

5.1.8 Chemicals

System	Pressure psig (1)				Temperature °F			
	Min.	Norm.	Max.	Design	Min.	Norm.	Max.	Design
Caustic – 20%	TBD	600	TBD		TBD	100	TBD	
TBD								
Note: (1) At battery limit at grade level								

5.2 Natural Gas

Lower Heating Value: TBD Btu/SCF

Composition: Expected to fluctuate within the range indicated in table below

Component	Vol%	Vol %
Oxygen		
Nitrogen		
Hydrogen		
CO ₂		
Methane		
Ethane		
Propane		
i-Butane		
n-Butane		

⁸ Dew Point -50 °F.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Component	Vol%	Vol %
i-Pentane		
n-Pentane		
C ₆ +		
H ₂ S, ppmv		
Normal		
Maximum		
Total Sulfur, ppmv		
Normal		
Maximum		
Total	100.00	100.00

5.3 Raw Water/Demineralized Water Analysis

Raw Water/Treated Water Analysis Table			
Analyzed for	Maximum Incoming Raw Water	Treated Water	Current Raw Water ⁹
pH			10.1
Conductivity, mmho/cm			n/r
Total Dissolved Solids, ppm			n/r
Total Alkalinity, ppm			31.5
Calcium as (Ca), ppm			n/r
Magnesium as (Mg), ppm			n/r
Sodium as (Na), ppm			n/r
Potassium as (K), ppm			n/r
Iron as (Fe), ppm			n/r
Phosphate as (P), ppm			n/r
Chloride as (Cl ⁻), ppm			n/r
Sulfate as (SO ₄ ²⁻), ppm			n/r
Silica as (SiO ₂), ppm			6.6
Residual Chlorine, ppm			0.3

⁹ Average values from CWA Nov 2008.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

Raw Water/Treated Water Analysis Table			
Analyzed for	Maximum Incoming Raw Water	Treated Water	Current Raw Water ⁹
Total Organic Carbon, ppm			n/r
Zinc as (Zn), ppm			n/r
Hardness as CaCO ₃ , ppm			n/r
Langelier Index			n/r
Turbidity			1.5
Total Hardness as CaCO ₃ , ppm			59.5
Notes:			

5.4 Waste Water Pretreatment Requirements

A purge stream of water shall be blown down from the Quench Water system. Before exiting the process area, the blowdown water stream shall be flashed to remove dissolved gases. Solid components shall be substantially removed by settling and filtration. Therefore, the blowdown stream shall consist of settler overflow and filtrate. The blowdown flow rate shall be set sufficient to purge chloride and other dissolved components from the process system.

5.5 Storm Water Quality Requirements

Requirements for Storm Water are outside of the scope of this study.

5.6 Electrical

Motor Size, hp	Voltage	Phase
<1/2	120	1
1/2 - 250	480	3
250 - 5000	4160	3
Above 5000	13.2 KV	3
Uninterruptible Power Supply:	TBD	
Motor Control Centers	TBD	

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

6. UNITS OF MEASUREMENT

6.1 Scope

This Specification defined the symbols to be used for the units for this project.

6.2 Definition

English (U.S.) Units will be used for this project.

6.3 English (U.S.) Units and Symbols

6.3.1 English (U.S.) Units

All descriptive and essential numerical quantities shall be expressed in the units and symbols defined in this document.

The selected units, prefixes and symbols for commonly used quantities are listed below. The units and symbols shall be used in all documents, drawings and specifications.

	English (U.S.)	Alternate Units
Temperature	°F	
Pressure	psig	in W.C. ¹⁰
Vacuum	mm Hg	in W.C. ⁵
Weight (Mass)	lb	
Density	lb/ft ³	
Volume, Liquids	gal	
Volume, Gases	SCF or ACF	
Flow, Liquids	lb/hr	
Flow, Gases	lb/hr	
Flow, Steam	lb/hr	
Heat	Btu	
Power	kW	
Electrical Potential	V	
Heat Transfer Coefficient	Btu/hr-ft ² -°F	
Viscosity	cP	
Equipment Dimensions and Pipe Length	ft-in	
Nominal Pipe Diameter	in	

¹⁰ Used for near atmospheric pressure streams, storage tank design.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

	English (U.S.)	Alternate Units
Tubing Sizes	in	
Tubing Wall Thickness	BWG	
Plot Plan Dimensions	ft	
Velocity	ft/s	
Vessel Nozzle Sizes	in	
Sound Pressure	dB(A)	
Sound Power	dB(A)	
Specific Heat	Btu/lb-°F	

6.3.2 Prefixes

When expressing a nominal numerical value in descriptive data, the following prefixes should be used.

Alternative Prefixes		
Factor	Prefix or Multiple	Symbol
10^6	Mega or million	MM
10^3	Kilo or thousand	M
10^{-2}	centi	c
10^{-3}	milli	m
10^{-6}	micro	μ

6.4 Dimensions

6.4.1 Drawings

Dimensions on product drawings shall be expressed only in feet and inches.

6.5 Style

When written in full, the names of units start with a lower case letter, except at the beginning of a sentence. Exceptions are when the unit is named for a person, such as Fahrenheit, or a country, such as British thermal unit.

The use of symbols is preferred instead of writing out the name of the unit. Unit symbols are written with lower case letters except that the first letter is upper case when the name of the unit is derived from the name of a person, such as Fahrenheit, or a country, such as British thermal unit. Unit symbols are typed in Arial non-italic characters.

All symbols for numerical prefixes are written with lower case letters except for MM (10^6) and M (10^3)

When writing symbols or names for units having prefixes, no space is left between letters making up the symbol or the name.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

When writing symbols after a number to which it refers, a space shall be left between the number and the symbol.

When written in full, the names of units are made plural when appropriate. Fractions, both common and decimal, are always singular.

Symbols for units are the same in singular and plural. No "s" is ever added to indicate a plural.

6.6 Mathematical Notation

When writing symbols for such units as square feet, write the symbol for the unit, followed by the superscript (for example: ft²).

A slant line, solid horizontal line or negative powers may be used to indicate division of symbols. Only one solid (or slant line) may be used. The word "per" may also be used in expressions to indicate divisions, but only one "per" may be used.

The product of two or more symbols shall be indicated by parentheses or brackets, or by inserting a dash or an asterisk or a space between the symbols.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

7. NUMBERING PROCEDURE

7.1 Scope

This document states the procedure for numbering equipment, drawings, lines and instrumentation. All equipment will be numbered according to Sections 7.3 and 7.3.2 below. The new equipment will start with the next available sequence number for that particular unit/area.

7.2 Unit Numbers

The unit numbers are shown below:

Unit ID Number	Unit
10's	Gasification (by others)
11	Gasification Train 1 (example)
12	Gasification Train 2
13	Feedstock Handling (common to all Gasification trains)
14	Slag Handling (common to all Gasification trains)
20's	Syngas Treatment
21	Syngas Quench & Scrubbing
22	Shift Reaction & Cooling
23	Acid Gas Removal
24	Hydrogen Purification
25	Sulfur Recovery
26	CO2 Compression
27	Syngas Compression (for Case 2)
30's	FT Reaction & Upgrading (by others)
40's	Auxiliary Units (by others)
	Air Separation
50 & 60's	Open
70, 80, 90's	Balance of Plant (Utilities, Offsites) (by others)

The second digit of the unit number can be used to distinguish between multiple trains in the Gasification or other units as needed.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

7.3 Equipment Numbering

7.3.1 Equipment Letter Designations

The following equipment letter designations will be used for equipment:

- C - Compressors, Blowers
- E - Exchangers
- F - Fired Equipment
- H - Vacuum Equipment
- P - Pumps
- R - Reactors
- S – Solids Handling Equipment (conveyors, etc)
- T – Tanks, Sumps
- V - Pressure Vessels, Columns, Knockout Drums
- X - Misc. Equipment
- Y - Filters, Mixers
- Z - Package Units

The equipment designation letters will be followed by a four digit number. The first two digits shall be the unit number (see Section 7.2 above); the last two digits are the equipment sequence number. Sequence numbers shall start with “01” (i.e. do not use “00”).

For Case 1, the sequence numbers shall start with “01” and for Case 2, the sequence numbers shall start with “51”.

Drivers shall carry the same number as the driven equipment except that the driver classification suffix will be added to the equipment designation letter.

- M - Motor
- T - Turbine

Spared equipment will be designated with a suffix letter A, B, C, etc.

7.3.2 Equipment Number Examples

P-2105B indicates a spared pump in Unit 21 with a sequence number of 05. Its corresponding motor will be PM-2105B.

V-1110 indicates a column in Unit 11 with a sequence number of 10.

E-7201C indicates the third exchanger in Unit 72 with a sequence number of 01. Exchangers E-7201A and E-7201B are in the same service.

V-2101 is the first vessel in unit 21 for Case 1 and V-2151 is the first vessel in unit 21 for Case 2.

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

7.4 Drawing Numbering**7.4.1 Drawing Number Format**

Drawing Numbers shall follow the following format:

a	b	c	d	e

23-PR-PFD-0010-0

- a Unit Number
- b Responsible Discipline (PR – Process, ME - Mechanical, etc)
- c Drawing type (PFD, etc))
- d Sequential drawing number
- e Issue (Revision) Number

For Case 1, drawings shall start with “0001” for general drawings such as the plot plan and with “0010” for PFDs in each unit.

For Case 2, drawings shall start with “0501” for general drawings such as the plot plan and with “0510” for PFDs in each unit.

**BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS**

8. ECONOMIC DATA

The following economic values should be used for design evaluations for this project.

8.1 Feedstock & Product Values

Later

8.2 Utility Costs

Power:	\$ 0.055 per Kwh
Natural Gas:	\$ TBD per MM Btu
Treated Water:	\$ TBD per 1,000 gallons
Nitrogen:	\$ TBD per M SCF
Steam	\$ TBD per M lb

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

9. ABBREVIATIONS AND ACRONYMS

AGR	acid gas removal (unit)
ASU	air separation unit
BFD	Block Flow Diagram
BoD, BOD	Basis of Design
BOP	Balance of Plant
CC	Combined Cycle
CCS	Carbon Capture & Sequestration (Storage)
CCT	Clean Coal Technology
CH ₄	Methane
CO	carbon monoxide
CO ₂	carbon dioxide
COS	carbonyl sulfide
CT	combustion turbine (also GT)
CTL	Coal to Liquids
DOE	Department of Energy (US)
EIA	Energy Information Administration (US)
EOR	enhanced oil recovery
EPRI	Electric Power Research Institute
FEED	Front-End Engineering Design
FEL	Front-End Loading
FGD	Flue Gas Desulfurization
FT, F-T	Fischer-Tropsch
GT	gas turbine (also CT)
GTL	Gas to Liquids
H&MB	Heat & Material Balance
H ₂	hydrogen
H ₂ O	water
H ₂ S	hydrogen sulfide

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

HC	hydrocarbon
Hg	mercury
HRSG	Heat Recovery Steam Generator
I	Interlock
IGCC	Integrated Gasification Combined Cycle
ISBL	Inside Battery Limits
MAF	moisture ash free
MMSCFD, MMSCF/D	million standard cubic feet per day
NETL	National Energy Technology Laboratory
NG	Natural Gas
NO _x	nitrogen oxides (typically NO and NO ₂)
NSPS/EG	New Source Performance Standards/Emission Guidelines
O ₂	Oxygen
OpEx, opex	Operating Expenses
OSBL	Outside Battery Limits
P&ID	Piping & Instrument Diagram
PEP	Project Execution Plan
PFD	Process Flow Diagram
PM ₁₀ , PM ₁₀	Particulate Material smaller than 10 microns
ppm	parts per million
ppmv	parts per million volume
ppmw	parts per million weight
PSA	Pressure Swing Adsorption
RWU	Rectisol Wash Unit
SCR	Selective Catalytic Reduction (of NO _x)
SG	syngas, synthesis gas
SIS	Safety Instrumented System
SNG	Synthetic (or substitute) Natural Gas
SO ₂	sulfur dioxide
SO _x	sulfur oxides
SRU	sulfur recovery unit

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

ST	steam turbine
STG	steam turbine-generator
TBD	To be determined
TBL	Technical Battery Limits
TGTU	Tail Gas Treating Unit
TIC	Total Installed Cost
Vol%	Volume percent
WPMP	WorleyParsons Project Management Process
Wt%	Weight percent
ZnO	Zinc Oxide

**BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS**

APPENDIX A - PREVAILING WINDS

**BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS**

APPENDIX B - RAINFALL CURVES

BASIS OF DESIGN (REV. D)
SYNGAS TREATMENT STUDY FOR ICRC COAL GASIFICATION FOR HYDROGEN AND LIQUIDS

APPENDIX C - PRECIPITATION DEPTH

Near Galveston, Texas				PRECIPITATION DEPTH (inches)					
				Return Period (years)					
				2	5	10	25	50	100
Input			Duration						
<u>HYDRO-35</u>			5 min	0.57	0.63	0.69	0.77	0.83	0.90
2-yr	100-yr		10	0.95	1.08	1.17	1.32	1.44	1.56
5-min	0.57	0.90	15	1.22	1.38	1.51	1.71	1.86	2.02
15-min	1.22	2.02	30	1.79	2.13	2.38	2.74	3.02	3.31
60-min	2.39	4.65	60	2.39	2.90	3.27	3.81	4.23	4.65
G&HES	Input <u>TP-40</u>		1 hr	2.94	3.75	4.50	5.07	5.65	6.32
			3	3.21	4.10	4.95	5.73	6.45	7.40
			6	3.86	5.05	6.02	7.02	8.00	9.00
			12	4.55	6.05	7.55	8.70	9.97	11.05
			24	5.45	7.40	8.90	10.20	11.97	13.30

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendix 2

	10	9	8	7	6	5	4	3	2	1		
A	<div>EQUIPMENT TAGS</div> <div><div>C</div><div>COMPRESSORS, BLOWERS</div></div> <div><div>E</div><div>EXCHANGERS</div></div> <div><div>F</div><div>FIRED EQUIPMENT</div></div> <div><div>H</div><div>VACUUM EQUIPMENT</div></div> <div><div>P</div><div>PUMPS</div></div> <div><div>R</div><div>REACTORS</div></div> <div><div>S</div><div>SOLIDS HANDLING EQUIPMENT (CONVEYORS ETC.)</div></div> <div><div>T</div><div>TANKS, SUMPS</div></div> <div><div>V</div><div>PRESSURE VESSELS, COLUMNS, KNOCKOUT DRUMS</div></div> <div><div>X</div><div>MISCELLANEOUS EQUIPMENT</div></div> <div><div>Y</div><div>FILTERS, MIXERS</div></div> <div><div>Z</div><div>PACKAGE UNITS</div></div>										NOTES:	
B												
C	<div>UTILITY & COMMODITY ABBREVIATIONS</div> <div><div>AG</div><div>ACID GAS, SOUR GAS</div></div> <div><div>BD</div><div>BLOWDOWN</div></div> <div><div>BFW</div><div>BOILER FEED WATER</div></div> <div><div>CWR</div><div>COOLING WATER RETURN</div></div> <div><div>CWS</div><div>COOLING WATER SUPPLY</div></div> <div><div>DMW</div><div>DEMINERALIZED WATER</div></div> <div><div>DR</div><div>DRAIN</div></div> <div><div>DW</div><div>DRINKING WATER</div></div> <div><div>FG</div><div>FUEL GAS</div></div> <div><div>FW</div><div>FIRE WATER</div></div> <div><div>H</div><div>HYDROGEN</div></div> <div><div>HN</div><div>HIGH PRESSURE NITROGEN</div></div> <div><div>HS</div><div>HIGH PRESSURE STEAM</div></div> <div><div>IA</div><div>INSTRUMENT AIR</div></div> <div><div>IS</div><div>INTERMEDIATE PRESSURE STEAM</div></div> <div><div>LN</div><div>LOW PRESSURE NITROGEN</div></div> <div><div>LS</div><div>LOW PRESSURE STEAM</div></div> <div><div>ME</div><div>METHANOL</div></div> <div><div>MS</div><div>MEDIUM PRESSURE STEAM</div></div> <div><div>N</div><div>NITROGEN</div></div> <div><div>NA</div><div>CAUSTIC</div></div> <div><div>NG</div><div>NATURAL GAS</div></div> <div><div>OX</div><div>OXYGEN</div></div> <div><div>PC</div><div>PROCESS CONDENSATE</div></div> <div><div>PR</div><div>PROPYLENE REFRIGERANT</div></div> <div><div>PW</div><div>PROCESS WATER</div></div> <div><div>RA</div><div>VENT TO ATMOSPHERE</div></div> <div><div>RF</div><div>RELIEF TO FLARE</div></div> <div><div>RW</div><div>RAW WATER</div></div> <div><div>SC</div><div>STEAM CONDENSATE</div></div> <div><div>SG</div><div>SYNTHESIS GAS</div></div> <div><div>SW</div><div>SERVICE WATER</div></div> <div><div>UW</div><div>UTILITY WATER</div></div> <div><div>WW</div><div>WASTE WATER</div></div>											
D	<div>GENERAL ABBREVIATIONS</div> <div><div>ACFM</div><div>ACTUAL CUBIC FEET PER MINUTE</div></div> <div><div>ACFS</div><div>ACTUAL CUBIC FEET PER SECOND</div></div> <div><div>AGR</div><div>ACID GAS REMOVAL (UNIT)</div></div> <div><div>ASU</div><div>AIR SEPARATION UNIT</div></div> <div><div>ATM</div><div>ATMOSPHERE</div></div> <div><div>BFD</div><div>BLOCK FLOW DIAGRAM</div></div> <div><div>BHP</div><div>BRAKE HORSEPOWER</div></div> <div><div>BOD</div><div>BASIS OF DESIGN</div></div> <div><div>BOP</div><div>BALANCE OF PLANT</div></div> <div><div>BTMS</div><div>BOTTOMS</div></div> <div><div>CAP</div><div>CAPACITY</div></div> <div><div>CIRC</div><div>CIRCULATION</div></div> <div><div>COND</div><div>CONDENSATE</div></div> <div><div>CR</div><div>CONTROL ROOM</div></div> <div><div>CS</div><div>CARBON STEEL</div></div> <div><div>CT</div><div>COMBUSTION TURBINE</div></div> <div><div>D</div><div>DAY</div></div> <div><div>DCS</div><div>DISTRIBUTED CONTROL SYSTEM</div></div> <div><div>DWG</div><div>DRAWING</div></div> <div><div>FT, F-T</div><div>FISCHER-TROPSCH</div></div> <div><div>H & MB</div><div>HEAT & MATERIAL BALANCE</div></div> <div><div>GPM</div><div>GALLONS PER MINUTE</div></div> <div><div>H, HR</div><div>HOURLY</div></div> <div><div>HP</div><div>HIGH PRESSURE, HORSEPOWER</div></div> <div><div>ID</div><div>INSIDE DIAMETER</div></div> <div><div>IP</div><div>INTERMEDIATE PRESSURE</div></div> <div><div>I</div><div>INTERLOCK</div></div> <div><div>ISBL</div><div>INSIDE BATTERY LIMITS</div></div> <div><div>LP</div><div>LOW PRESSURE</div></div> <div><div>KO</div><div>KNOCKOUT (DRUM)</div></div> <div><div>M</div><div>MINUTE</div></div> <div><div>MIN</div><div>MINIMUM, MINUTE</div></div> <div><div>MOL%</div><div>MOLE PERCENT</div></div> <div><div>MP</div><div>MEDIUM PRESSURE</div></div> <div><div>MMSCFD</div><div>MILLION STANDARD CUBIC FEET PER DAY</div></div> <div><div>MW</div><div>MOLECULAR WEIGHT</div></div> <div><div>NNF</div><div>NORMALLY NO FLOW</div></div> <div><div>OSBL</div><div>OUTSIDE BATTERY LIMITS</div></div> <div><div>P&ID</div><div>PIPING & INSTRUMENT DIAGRAM</div></div> <div><div>PFD</div><div>PROCESS FLOW DIAGRAM</div></div> <div><div>PLC</div><div>PROGRAMMABLE LOGIC CONTROLLER</div></div> <div><div>PPB</div><div>PARTS PER BILLION</div></div> <div><div>PPBV</div><div>PARTS PER BILLION VOLUME</div></div> <div><div>PPBW</div><div>PARTS PER BILLION WEIGHT</div></div> <div><div>PPH</div><div>POUNDS PER HOUR</div></div> <div><div>PPM</div><div>PARTS PER MILLION</div></div> <div><div>PPMV</div><div>PARTS PER MILLION VOLUME</div></div> <div><div>PPMW</div><div>PARTS PER MILLION WEIGHT</div></div> <div><div>PSA</div><div>PRESSURE SWING ADSORPTION</div></div> <div><div>RWU</div><div>RECTISOL WASH UNIT</div></div> <div><div>S, SEC</div><div>SECOND</div></div> <div><div>SD</div><div>SHUTDOWN</div></div> <div><div>SIS</div><div>SAFETY INSTRUMENTED SYSTEM</div></div> <div><div>SRU</div><div>SULFUR RECOVERY UNIT</div></div> <div><div>SS</div><div>STAINLESS STEEL</div></div> <div><div>ST</div><div>STEAM TURBINE</div></div> <div><div>STM</div><div>STEAM</div></div> <div><div>SU</div><div>START-UP</div></div> <div><div>TGTU</div><div>TAIL GAS TREATING UNIT</div></div> <div><div>T-T, T/T</div><div>TANGENT TO TANGENT</div></div> <div><div>VFD</div><div>VARIABLE FREQUENCY DRIVE</div></div> <div><div>VOL%</div><div>VOLUME PERCENT</div></div> <div><div>WT</div><div>WEIGHT, WATER TREATING</div></div> <div><div>WT%</div><div>WEIGHT PERCENT</div></div> <div><div>WWT</div><div>WASTE WATER TREATING</div></div>											
E	<div>INSTRUMENT ABBREVIATIONS</div> <div><div>AI</div><div>ANALYZER INDICATOR (ON-LINE)</div></div> <div><div>FC</div><div>FLOW CONTROL</div></div> <div><div>FE</div><div>FLOW ELEMENT</div></div> <div><div>FFC</div><div>FLOW RATIO CONTROLLER</div></div> <div><div>FI</div><div>FLOW INDICATOR</div></div> <div><div>FS</div><div>FLOW SWITCH</div></div> <div><div>FSH</div><div>FLOW SWITCH HIGH</div></div> <div><div>FSL</div><div>FLOW SWITCH LOW</div></div> <div><div>FT</div><div>FLOW TRANSMITTER</div></div> <div><div>FV</div><div>FLOW CONTROL VALVE</div></div> <div><div>HC</div><div>HAND CONTROL (FROM DCS)</div></div> <div><div>HS</div><div>HAND SWITCH (FROM DCS)</div></div> <div><div>LC</div><div>LEVEL CONTROL</div></div> <div><div>LI</div><div>LEVEL INDICATOR</div></div> <div><div>LT</div><div>LEVEL TRANSMITTER</div></div> <div><div>LV</div><div>LEVEL CONTROL VALVE</div></div> <div><div>PC</div><div>PRESSURE CONTROL</div></div> <div><div>PDI</div><div>PRESSURE DIFFERENTIAL INDICATOR</div></div> <div><div>PI</div><div>PRESSURE INDICATOR</div></div> <div><div>PT</div><div>PRESSURE TRANSMITTER</div></div> <div><div>PV</div><div>PRESSURE CONTROL VALVE</div></div> <div><div>SC</div><div>SPEED CONTROL</div></div> <div><div>TC</div><div>TEMPERATURE CONTROL</div></div> <div><div>TI</div><div>TEMPERATURE INDICATOR</div></div> <div><div>TT</div><div>TEMPERATURE TRANSMITTER</div></div> <div><div>TV</div><div>TEMPERATURE CONTROL VALVE</div></div>											
F												
G												
H												
	10	9	8	7	6	5	4	3	2	1		

NOTES:

0	06/25/09	FOR FINAL REPORT									
---	----------	------------------	--	--	--	--	--	--	--	--	--

REV	DATE	DESCRIPTION	DRAWN	CHECKED	DESIGNED	ENGINEER/LEADER	ENGINEER/LEADER	PROJECT MANAGER			
-----	------	-------------	-------	---------	----------	-----------------	-----------------	-----------------	--	--	--

PRELIMINARY STATUS	DATE	REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.
--------------------	------	--

APPROVED STATUS	DATE	REPRESENTS REVIEWED AND APPROVED DESIGN. ANY PORTION MARKED "HOLD" RETAINS PRELIMINARY STATUS.
-----------------	------	--

ORIGINATING PERSONNEL	PROFESSIONAL ENGINEER'S SEAL
-----------------------	------------------------------

DRAWN BY DJP	
CHECKED BY	
LEAD DESIGNER	
ENGINEER/TECH SPECIALIST J. HILTON	
PROJECT ENGINEERING MANAGER S. MARR	
PROJECT MANAGER M. DELALLO

Leadership No Incidents Safe Behavior	ORIGINALLY PREPARED UNDER THE RESPONSIBLE SUPERVISION OF PEL - STATE _____ DATE _____ LIC. NO. _____
---	--

WorleyParsons
resources & energy

CLIENT/PROJECT TITLE
ENVIRES COAL/BIOMASS
GASIFICATION PROJECT
CASE 1: HYDROGEN PRODUCTION

PROCESS FLOW DIAGRAM
UNIT 00
LEGEND NOMENCLATURE

SCALE NONE	DRAWING SIZE ARCH D (36" x 24")
---------------	------------------------------------

WORLDWIDE PARSONS' DWG. NO.	REV
-----------------------------	-----

ICRC-00-PR-PFD-0020 0

\\nrc00-pr-pfd-0020.dgn
10/26/2009 10:56:50 AM

NOTES:

1. THE EQUIPMENT SHOWN IS FOR ONE SYNGAS QUENCH & SCRUBBING TRAIN, ONLY ONE TRAIN IS REQUIRED.
2. EXCHANGER TYPE (DOUBLE PIPE OR SPIRAL WOUND) SHALL BE SUITABLE FOR HANDLING SMALL AMOUNT OF SOLIDS IN THE WATER STREAM.
3. FILTER CAKE WASH WATER.
4. SETTLER RAKE.
5. POSITIVE DISPLACEMENT PUMP.
6. TWO 100% EXCHANGERS.

[illegible][illegible]

REV	DATE	DESCRIPTION	RAWN	CHECKED	EAD	ENGINEER	ENGINEER	ECH, SPE	EAD DIS	ENGINEER	ROJ ENG	MANAGER
-----	------	-------------	------	---------	-----	----------	----------	----------	---------	----------	---------	---------

PRELIMINARY STATUS	DATE	REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.
--------------------	------	--

APPROVED STATUS	DATE	REPRESENTS REVIEWED AND APPROVE DESIGN. ANY PORTION MARKED 'HOLD' RETAINS PRELIMINARY STATUS.
1 DE		

ORIGINATING PERSONNEL	PROFESSIONAL ENGINEER'S SEAL
-----------------------	------------------------------

DRAWN BY	
----------	--

DJP
CHECKED BY

CHECKED BY	
------------	--

LEAD DESIGNER	
---------------	--

[illegible]

ENGINEER/TECH SPECIALIST	
--------------------------	--

J. HILTON	
PROJECT ENGINEERING MANAGER	

PROJECT ENGINEERING MANAGER
C. MARR

S. MARR	
PROJECT MANAGER	

M. DELALLO

Zero Leadership Initiative

Harm No Incidents Safe Behavior	ORIGINALLY PREPARED UNDER THE RESPONSIBLE SUPERVISION OF _____

PE: _____ STATE: _____
LIC. NO.: _____ DATE: _____

WILEY
WorleyParsons

 NATIONAL CENTER FOR THE STUDY OF THE HOLOCAUST,
 THE HOLOCAUST EDUCATION, REMEMBRANCE, AND RESEARCH CENTER

resources & energy

CLIENT/PROJECT TITLE
ENVIRES COAL /BIOMASS

ENVIRE3 COAL/BIO MASSGASIFICATION PROJECTCASE1: HYDROGEN PRODUCTION

CHSE1: HYDROGEN PRODUCTION

PROCESS FLOW DIAGRAM

PROCESS FLOW DIAGRAM

UNIT 21

QUENCH WATER HANDLING

SCALE	DRAWING SIZE
-------	--------------

SCALE NONE	DRAWING SIZE ARCH D (36" x 24")
---------------	------------------------------------

WORLEYPARSONS DWG. NO.	REV
1000 01 00 050 00000	0

ICRL-21-PR-PFD-0020	0
---------------------	---

[illegible]

1

...\\erc-21-pr-pfd-0020.dgn
5/23/2009 10:56:51 AM

NOTES:

- THE EQUIPMENT SHOWN IS FOR ONE ACID GAS REMOVAL TRAIN. ONLY ONE TRAIN IS REQUIRED.
- PACKAGED UNIT

0	06/25/09	FOR FINAL REPORT	DJP		JH	SM	MD
REV	DATE	DESCRIPTION	DRAWN	CHECKED	DESIGNER	ENGINEER/LEAD USER	PROJECT MANAGER
PRELIMINARY STATUS		DATE	REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.				
APPROVED STATUS		DATE	REPRESENTS REVIEWED AND APPROVED DESIGN. ANY PORTION MARKED "HOLD" RETAINS PRELIMINARY STATUS.				
ORIGINATING PERSONNEL			PROFESSIONAL ENGINEER'S SEAL				
DRAWN BY DJP							
CHECKED BY							
LEAD DESIGNER							
ENGINEER/TECH SPECIALIST J. HILTON							
PROJECT ENGINEERING MANAGER S. MARR							
PROJECT MANAGER M. DELALLO
Leadership No Incidents Safe Behavior			ORIGINALLY PREPARED UNDER THE RESPONSIBLE SUPERVISION OF PEL. LIC. NO. _____ STATE: _____ DATE: _____
WorleyParsons resources & energy							
CLIENT/PROJECT TITLE ENVIRES COAL/BIO MASS GASIFICATION PROJECT CASE 1: HYDROGEN PRODUCTION							
PROCESS FLOW DIAGRAM UNIT 23 RECTISOL ABSORPTION							
SCALE NONE		DRAWING SIZE ARCH D (36" x 24")					
WORLEYPARSONS' DWG. NO.		REV				0	
ICRC-23-PR-PFD-0010							

\\Nerc23-pr-pfd-0010.dgn
6/23/2009 10:56:52 AM

- NOTES:
1. THE EQUIPMENT SHOWN IS FOR ONE SULFUR RECOVERY TRAIN.
ONLY ONE TRAIN IS REQUIRED.
 2. PACKAGED UNIT.
 3. NUMBER OF SULFUR PREHEATERS, CONVERTERS, AND CONDENSERS TO BE DETERMINED BY LICENSOR.

[illegible]

<u>PRELIMINARY STATUS</u>	DATE	REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.
LDE		
<u>APPROVED STATUS</u>	DATE	REPRESENTS REVIEWED AND APPROVED DESIGN. ANY PORTION MARKED "HOLD" RETAINS PRELIMINARY STATUS.
LDE		

ORIGINATING PERSONNEL	PROFESSIONAL ENGINEER'S SEAL
DRAWN BY DJP	
CHECKED BY	
LEAD DESIGNER	
ENGINEER/TECH SPECIALIST J. HILTON	
PROJECT ENGINEERING MANAGER S. MARR	
PROJECT MANAGER M. DELALLO
Leadership No Incidents Safe Behavior	ORIGINALLY PREPARED UNDER THE RESPONSIBLE SUPERVISION OF PEJ: _____ STATE: _____ LIC: N/A _____ DATE: _____

<h1 data-bbox="2719 1622 2954 1639">WorlParsons</h1> <p data-bbox="2719 1639 2954 1649">resources & energy</p>
<p data-bbox="2636 1657 2954 1674">CLIENT / PROJECT TITLE</p>	
<p data-bbox="2636 1674 2954 1788"> ENVIRE COAL/BIOMASS GASIFICATION PROJECT CASE 1: HYDROGEN PRODUCTION </p>	
<p data-bbox="2636 1788 2954 1902"> PROCESS FLOW DIAGRAM UNIT 25 CLAUS SULFUR RECOVERY </p>	
<p data-bbox="2636 1902 2804 1927">SCALE</p> <p data-bbox="2636 1927 2804 1939">NONE</p>	<p data-bbox="2804 1902 2954 1927">DRAWING SIZE</p> <p data-bbox="2804 1927 2954 1939">ARCH D (36" x 24")</p>
<p data-bbox="2636 1939 2954 1956">WORLD PARSONS DWG. NO.</p>	
<p data-bbox="2636 1956 2954 1966">ICRC-25-PR-PFD-0010</p>	<p data-bbox="2954 1956 2990 1966">REV</p> <p data-bbox="2954 1966 2990 1974">0</p>

....\lerc-25-pr-pfd-0010.dgn

NOTES:

1. THE EQUIPMENT SHOWN IS FOR ONE SULFUR RECOVERY TRAIN. ONLY ONE TRAIN IS REQUIRED.
2. PACKAGED UNIT.
3. REGENERATOR INCLUDES REBOILER AND OVERHEAD CONDENSER AND REFLUX EQUIPMENT.

REV	DATE	DESCRIPTION	DRAWN	CHECKED	DESIGNED	ENGINEER/TECH	PROJECT
0	08/31/09	FOR FINAL REPORT	SMF		JH	SM	MD

PRELIMINARY STATUS	DATE	REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.
LDE		

APPROVED STATUS	DATE	REPRESENTS REVIEWED AND APPROVED DESIGN. ANY PORTION MARKED "HOLD" RETAINS PRELIMINARY STATUS.
LDE		

ORIGINATING PERSONNEL	PROFESSIONAL ENGINEER'S SEAL
DRAWN BY SMF	
CHECKED BY	
LEAD DESIGNER	
ENGINEER/TECH SPECIALIST J. HILTON	
PROJECT ENGINEERING MANAGER S. MARR	
PROJECT MANAGER M. DELALLO	

Zero Harm	Leadership No Incidents Safe Behavior	ORIGINALY PREPARED UNDER THE RESPONSIBLE SUPERVISION OF PEL - STATE, L.C. NO. DATE

WorleyParsons
resources & energy

CLIENT/PROJECT TITLE
ENVIRES COAL/BIOMASS
GASIFICATION PROJECT
CASE 2: FT LIQUIDS PRODUCTION

PROCESS FLOW DIAGRAM
UNIT 25
TAIL GAS TREATING

SCALE	DRAWING SIZE
NONE	ARCH D (36" x 24")

WORLEYPARSONS' DWG. NO.	REV
ICRC-25-PR-PFD-0520	0

\\wpc25-pr-pfd-0520.dgn
6/31/2009 6:56:22 AM

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendix 3

EnvReas Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)

Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	101	102	103	104	105	106	107	108	109	110
DESCRIPTION	Hot Raw Syngas from Gasification	Quenched Syngas to Venturi Mixer	Quench Bottoms Water to MP Flash Drum	Quench Spray Water to Quench Drum	Scrubbed Syngas to Shift Inlet KO Drum	Flash Gas to Ammonia Stripper	Flash Drum Bottoms to Cooler	Sellier Overflow to Gray Water Tank	Sellier Bottoms to Solids Filters	Filtrate to Gray Water Tank
V&L MIXTURE COMPONENTS										
H ₂	2,077.45	2,078.10	13.80	0.24	0.01	2,077.62	14.08	0.24	0.08	0.00
CO	4,532.85	4,532.24	30.09	0.32	0.01	4,531.04	30.72	0.32	0.10	0.00
CO ₂	68.37	68.54	0.45	0.07	0.00	67.10	0.48	0.09	0.03	0.00
CH ₄	6.57	6.57	0.04	0.00	0.00	6.57	0.04	0.00	0.00	0.00
H ₂ S	41.02	41.23	0.27	0.10	0.00	40.90	0.28	0.12	0.04	0.00
COS	2.85	2.85	0.02	0.00	0.00	2.85	0.02	0.00	0.00	0.00
NH ₃	0.10	0.08	0.00	0.05	0.00	0.85	0.01	0.02	0.01	0.00
H ₂ O	1.82	0.63	0.00	0.03	0.00	0.55	0.00	0.01	0.00	0.00
AR	20.51	20.51	0.00	0.00	0.00	20.51	0.00	0.00	0.00	0.00
N ₂	39.33	39.34	0.26	0.00	0.00	39.33	0.27	0.00	0.00	0.00
H ₂ O	45.97	8,273.48	54.92	2,995.81	99.57	7,956.66	53.95	302.85	99.55	99.55
ELECTROLYTES (Note 1)	0.00	0.00	0.00	12.08	0.40	0.00	0.00	12.07	0.45	3.73
Solids	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	9,837.41	100.00	15,063.86	3,005.69	100.00	14,748.17	100.00	303.64	100.00	743.68
V&L MIXTURE PROPERTIES										
Vapor Fraction	1.0	1.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0	0.0
Temperature	F	405.7	406.0	401.7	401.2	320.3	320.3	140.2	140.5	140.6
Pressure	psig	500.0	490.0	600.0	475.0	75.0	75.0	0.0	50.0	25.0
Mass Flow	lb/hr	136,671.0	286,631.0	54,421.0	281,202.0	5,473.9	48,947.1	33,800.1	15,147.0	13,464.0
Molecular Weight	MM Btu/lb	20.28	19.04	18.09	19.07	18.03	18.09	18.03	18.09	18.10
Enthalpy	MM Btu/lb	-92.817	-1,054.500	-351.370	-1,023.000	-30.976	-320.360	-227.340	-101.870	-90.491
Density	lb/ft ³	0.303	1.084	53.222	1.056	0.199	56.325	61.012	61.017	60.964
Vapor Phase Properties										
Sat. Volume Flow	MMSCFD	62.3	137.2		134.3	2.8				
Flowing Volume	ft ³ /s	127.1	73.5		74.0	7.6				
Mass Flow	lb/hr	136,671.0	286,631.0		281,202.0	5,473.9				
Density	lb/ft ³	0.303	1.084		1.056	0.199				
Heat Capacity	Btu/lb-F	0.402	0.438		0.435	0.473				
Conductivity	Btu/lb-F	0.119	0.031		0.031	0.017				
Viscosity	cP	0.060	0.022		0.022	0.015				
Ch/Cv		1.323	1.410		1.408	1.348				
Z-Factor		1.005	0.955		0.957	0.970				
MM		20.28	19.04		19.07	18.03				
Liquid Phase Properties										
Flowing Volume	gpm		127.5		468.8					
Mass Flow	lb/hr		54,421.0		201,809.0					
Density	lb/ft ³		53.22		53.67					
Heat Capacity	Btu/lb-F		1.07		1.08					
Conductivity	Btu/lb-F		0.37		0.37					
Viscosity	cP		0.12		0.11					
Surface Tension	dyne/cm		36		37					
SOLID PROPERTIES										
Mass Flow	lb/hr		1,663.0							
Enthalpy	MM Btu/hr		-5.454							
VELLS										
TOTAL Mass Flow	lb/hr	140,432.3	286,831.0	56,104.0	201,809.0	5,473.9	50,830.1	33,800.1	15,830.0	13,464.0
TOTAL Enthalpy flow	MM Btu/hr	-90.843	-1,054.500	-357.824	-1,023.000	-30.976	-328.842	-227.340	-102.406	-90.491

Note 1: Electrolytes are the sum of the ionic species and dissolved solids in the water phase.

Enviro Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)

Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFD Stream Number	111	112	113	114	115	201	202	203	204	205
DESCRIPTION	Filler Cake to Offsite Disposal	HP Gray Water to Gray Water Pretreater	Blowdown Water to Offsite Treating	Makeup Water to Water Tank	Caustic Addition to Quench Water	Bypass Syngas to Second Shift Reactor	Steam Addition to First Shift Reactor	Syngas/Steam to Shift Feed/Effluent Exchanger	Syngas/Steam to First Shift Reactor	
V&L MIXTURE COMPONENTS										
H ₂	0.00	0.00	0.00	0.00	0.00	1,038.81	0.00	1,038.81	1,038.81	11.77
CO	0.00	0.00	0.00	0.00	0.00	2,265.52	0.00	2,265.52	2,265.52	25.66
CO ₂	0.00	0.00	0.00	0.00	0.00	35.60	0.00	35.60	35.60	0.40
CH ₄	0.00	0.00	0.00	0.00	0.00	3.28	0.00	3.28	3.28	0.04
H ₂ S	0.00	0.00	0.00	0.00	0.00	20.45	0.00	20.45	20.45	0.23
COS	0.00	0.00	0.00	0.00	0.00	1.42	0.00	1.42	1.42	0.02
NH ₃	0.00	0.01	0.02	0.00	0.00	0.27	0.00	0.27	0.27	0.01
H ₂ N	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N ₂	0.00	0.00	0.00	0.00	0.00	10.25	0.14	10.25	10.25	0.12
H ₂ O	93.42	5,618.35	1,752.25	4,794.26	30.82	3,976.33	1,454.37	5,433.21	5,433.21	61.54
ELECTROLYTES (Note 1)										
Sulfur	0.00	3.83	8.24	0.46	18.38	0.00	0.00	0.00	0.00	0.00
TOTAL	93.42	5,622.39	1,833.52	4,794.26	37.76	7,374.08	1,454.38	8,828.96	8,828.96	100.00
V&L MIXTURE PROPERTIES										
Vapor Fraction	0.0	0.0	0.0	0.0	0.0	1.0	1.0	1.0	1.0	1.0
Temperature	F	140.5	140.5	80.0	100.0	401.2	488.3	416.8	464.0	484.0
Pressure	psig	0.0	50.0	50.0	600.0	475.0	650.0	475.0	485.0	485.0
Mass Flow	lb/hr	1,883.0	32,277.1	86,370.0	140,601.0	26,210.0	166,811.0	166,811.0	166,811.0	166,811.0
Molecular Weight	lb/lb-mole	18.02	18.02	18.02	18.38	18.07	18.02	18.89	18.89	18.89
Enthalpy	MM Btu/hr	-11,380	-217,040	-589,230	-4,475	-511,510	-147,720	-659,230	-855,590	-855,590
Density	lb/ft ³	61.370	61.002	62.233	75.839	1.056	1.398	1.045	0.960	0.960
Vapor Phase Properties										
Std. Volume Flow	MMSCFD					67.2	13.3	80.4	80.4	80.4
Flowing Volume	lb/hr					37.0	5.2	44.3	48.3	48.3
Mass Flow	lb/hr					140,601.0	26,210.0	166,811.0	166,811.0	166,811.0
Density	lb/ft ³					1.056	1.398	1.045	0.960	0.960
Heat Capacity	Btu/lb-F					0.435	0.594	0.457	0.454	0.454
Conductivity	Btu/hr-F					0.031	0.032	0.039	0.042	0.042
Viscosity	CP					0.022	0.014	0.017	0.018	0.018
Co/Cv						1.408	1.587	1.434	1.414	1.414
Z-Factor						0.957	0.935	0.941	0.952	0.952
MM						19.07	18.02	18.89	18.89	18.89
Liquid Phase Properties										
Flowing Volume	gpm		68.0	173.0	1.1					
Mass Flow	lb/hr	203.3	32,277.1	86,370.0	694.0					
Density	lb/ft ³	1,883.0	61.00	62.23	75.84					
Heat Capacity	Btu/lb-F	62.17	0.99	1.00	0.88					
Conductivity	Btu/hr-F	1.00	0.37	0.35	0.44					
Viscosity	CP	0.45	0.48	0.74	2.78					
Surface Tension	dynes/cm	66	86	72	81					
SOLID PROPERTIES										
Mass Flow	lb/hr	1,883.0			0.0					
Enthalpy	MM Btu/hr	-6,536								
TOTALS										
TOTAL Mass Flow	lb/hr	3,366.0	101,357.0	86,370.0	694.0	140,601.0	26,210.0	166,811.0	166,811.0	166,811.0
TOTAL Enthalpy flow	MM Btu/hr	-17,516	-888,710	-589,230	-4,475	-511,510	-147,720	-659,230	-855,590	-855,590

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)
Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	205	207	208	209	210	211	212	213	214	215
DESCRIPTION		Syngas from Second Shift Reactor	Syngas to Third Shift Reactor		Shifted Syngas from Shift Condensate Heater	Shifted Syngas from Gray Water Heater	Shifted Syngas from LP Steam Generator	Shifted Syngas from Water Heater KO Drum	Shifted Syngas from Water Heater KO Drum	Shifted Syngas from Airfin Cooler KO Drum
V&L MIXTURE COMPONENTS										
H ₂		lbmol/hr	lbmol/hr		lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr
CO		37.06	37.06		40.09	40.09	40.09	46.64	0.41	57.11
CO ₂		604.23	604.23		114.02	114.02	114.02	114.02	0.01	114.01
CH ₄		4,000.56	4,000.56		27.72	27.72	27.72	32.23	2.59	4,482.04
H ₂ S		6.57	6.57		6.57	6.57	6.57	0.05	0.00	6.57
COS		43.66	43.66		43.72	43.72	43.72	0.31	0.08	43.42
NH ₃		0.09	0.09		0.02	0.02	0.02	0.00	0.00	0.02
H ₂ O		1.80	1.80		1.50	1.50	1.50	0.01	0.14	0.01
HCN		0.00	0.00		0.00	0.00	0.00	0.00	0.00	0.00
AR		0.00	0.00		0.00	0.00	0.00	0.00	0.00	0.00
N ₂		20.51	20.51		20.51	20.51	20.51	0.15	0.00	20.51
H ₂ O		39.33	39.33		39.33	39.33	39.33	0.28	0.00	39.33
ELECTROLYTES (Note 1)		5,481.58	5,481.58		4,990.56	4,990.56	4,990.56	19.51	2,273.18	170.78
Sulfur		0.00	0.00		0.00	0.00	0.00	0.00	0.71	0.00
TOTAL		16,203.01	16,203.01		16,203.01	16,203.01	16,203.01	100.00	2,277.12	11,377.00
V&L MIXTURE PROPERTIES										
Vapor Fraction		1.0	1.0		1.0	1.0	1.0	1.0	0.0	1.0
Temperature		717.7	420.0		411.2	384.0	345.0	289.9	354.0	360.0
Pressure		425.0	405.0		345.0	375.0	365.0	354.0	354.0	343.0
Mass Flow		307,411.0	307,411.0		307,411.0	307,411.0	307,411.0	286,311.0	41,099.8	220,124.0
Molecular Weight		18.97	18.97		18.97	18.97	18.97	18.12	18.97	18.97
Enthalpy		-1,192,400	-1,234,200		-1,244,300	-1,250,900	-1,253,500	-1,025,200	-270,850	-775,920
Density		0.664	0.860		0.823	0.859	0.859	0.877	0.718	0.859
Vapor Phase Properties										
Std. Volume Flow		147.6	147.6		147.6	147.6	147.6	126.8	84.3	103.6
Flowing Volume		126.6	99.3		103.0	99.4	99.5	84.3	58.3	84.3
Mass Flow		307,411.0	307,411.0		307,411.0	307,411.0	307,411.0	286,311.0	41,099.8	220,124.0
Density		0.664	0.860		0.823	0.859	0.859	0.877	0.718	0.859
Heat Capacity		0.473	0.560		0.459	0.457	0.456	0.441	0.416	0.416
Conductivity		0.022	0.022		0.047	0.039	0.039	0.051	0.051	0.051
Viscosity		1.311	1.344		1.350	1.317	1.316	1.349	1.365	1.365
Gravimetric		0.995	0.991		0.979	0.979	0.979	0.986	0.972	0.983
Z-Factor		18.97	18.97		18.97	18.97	18.97	18.12	18.12	18.12
Liquid Phase Properties										
Flowing Volume										
Mass Flow										
Density										
Heat Capacity										
Conductivity										
Viscosity										
Surface Tension										
SOLID PROPERTIES										
Mass Flow										
Enthalpy										
V&L&S										
TOTAL Mass Flow		307,411.0	307,411.0		307,411.0	307,411.0	307,411.0	286,311.0	41,099.8	220,124.0
TOTAL Enthalpy flow		-1,192,400	-1,234,200		-1,244,300	-1,250,900	-1,253,500	-1,025,200	-270,850	-775,920

Envirotes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)
Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PPD Stream Number	216	217	218	219	220	221	222	223	224	225
DESCRIPTION	Shift Condensate from Airfin Cooler KO Drum	Shifted Syngas from Trim Cooler KO Drum	Shift Condensate from Trim Cooler KO Drum	Shifted Syngas from Mercury Removal Drum	Hot Gray Water to Syngas Scrubber	Hot Shift Condensate to Syngas Scrubber	Sour Gas to Ammonia Stripper Overhead Condenser	Shift Condensate to Ammonia Stripper	Sour Gas to Sulfur Recovery System	Shipped Condensate to HP Shift Condensate Drum
V&L MIXTURE COMPONENTS										
H ₂	0.48	0.02	0.02	0.02	0.00	0.00	0.48	0.02	0.74	0.00
CO	0.01	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.33	0.00
CO ₂	5.66	0.22	0.52	0.38	0.00	0.00	3.41	2.78	7.23	0.00
CH ₄	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ S	0.18	0.01	0.39	0.02	0.00	0.00	0.05	0.14	0.33	0.00
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH ₃	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ O	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ O	2,545.31	99.68	36.09	39.33	0.35	0.00	0.00	2,979.74	2.14	2,981.66
ELECTROLYTES (Note 1)	1.95	0.08	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00
Sulfur	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	2,553.59	100.00	11,235.72	100.00	5,822.39	100.00	4.31	100.00	11.41	2,982.06
V&L MIXTURE PROPERTIES										
Vapor Fraction	0.0	1.0	0.0	1.0	0.0	0.0	1.0	0.0	1.0	0.0
Temperature	160.0	100.0	100.0	109.6	353.0	400.0	157.1	157.2	200.0	286.2
Pressure	343.0	332.0	332.0	312.0	580.0	580.0	60.0	100.0	45.0	400.0
Mass Flow	48,187.6	217,673.0	2,450.7	217,673.0	101,357.0	94,822.3	158.0	48,480.3	389.8	53,722.3
Molecular Weight	18.09	19.37	18.12	19.37	18.03	18.03	37.48	18.03	34.17	34.17
Enthalpy	-310.470	-767.220	-16.599	-766.740	-682.830	-815.400	-0.603	-326.450	-1.466	-354.750
Density	60.694	1.130	61.375	1.059	55.566	55.668	0.429	60.933	0.292	57.488
Vapor Phase Properties										
Std. Volume Flow	102.3	102.3	102.3	102.3	102.3	102.3	0.0	0.0	0.1	0.0
Flowing Volume	53.5	53.5	53.5	53.5	53.5	53.5	0.0	0.0	0.1	0.0
Mass Flow	217,673.0	217,673.0	217,673.0	217,673.0	217,673.0	217,673.0	158.0	158.0	389.8	53,722.3
Density	1.130	1.130	1.130	1.059	1.059	1.059	0.429	0.429	0.292	0.292
Heat Capacity	0.410	0.410	0.410	0.410	0.410	0.410	0.293	0.293	0.293	0.293
Conductivity	0.046	0.046	0.046	0.046	0.046	0.046	0.017	0.017	0.017	0.017
Viscosity	0.015	0.015	0.015	0.015	0.015	0.015	0.017	0.017	0.017	0.017
Cp/Cv	1.380	1.380	1.380	1.380	1.380	1.380	1.301	1.301	1.297	1.297
Z-Factor	0.969	0.969	0.969	0.978	0.969	0.969	0.969	0.969	0.987	0.987
MW	19.37	19.37	19.37	19.37	19.37	19.37	37.48	37.48	34.17	34.17
Liquid Phase Properties										
Flowing Volume	94.9	94.9	94.9	94.9	94.9	94.9	94.9	94.9	94.9	94.9
Mass Flow	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6
Density	60.69	60.69	60.69	60.69	60.69	60.69	60.69	60.69	60.69	60.69
Heat Capacity	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Conductivity	0.31	0.31	0.31	0.31	0.31	0.31	0.31	0.31	0.31	0.31
Viscosity	0.38	0.38	0.38	0.38	0.38	0.38	0.38	0.38	0.38	0.38
Surface Tension	64	64	64	64	64	64	64	64	64	64
SOLID PROPERTIES										
Mass Flow	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6	48,187.6
Enthalpy	-310.470	-767.220	-16.599	-766.740	-682.830	-815.400	-0.603	-326.450	-1.466	-354.750
V&L &S										
TOTAL Mass Flow	48,187.6	217,673.0	2,450.7	217,673.0	101,357.0	94,822.3	158.0	48,480.3	389.8	53,722.3
TOTAL Enthalpy flow	-310.470	-767.220	-16.599	-766.740	-682.830	-815.400	-0.603	-326.450	-1.466	-354.750

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)
Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	301	302	303	304	401	402	501	502	601
DESCRIPTION	Sweet Syngas to PSA	CO2 to CO2 Compression	Acid Gas to SRU	AGR Waste Water to Offsite Treating	Hydrogen Product	Fuel Gas to Offsite	Sulfur Product	SRU Waste Water to Offsite Treating	CO2 Product
V&L MIXTURE COMPONENTS									
	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr
H ₂	6,488.11	97.10	1.71	1.52	5,774.42	99.48	713.69	81.35	0.00
CO	113.22	1.69	0.08	0.08	0.01	0.00	113.21	12.90	0.00
CO ₂	14.52	0.22	67.74	60.00	0.02	0.00	14.50	1.65	0.00
CH ₄	6.33	0.09	0.01	0.01	0.00	0.00	6.02	0.69	0.00
H ₂ S	0.01	0.00	43.31	38.37	0.00	0.00	0.00	0.00	0.00
CO ₂	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH ₃	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H ₂ O	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	20.51	0.31	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N ₂	39.28	0.59	0.00	0.00	10.25	0.18	10.25	1.17	0.00
H ₂ O	0.00	0.00	0.00	0.00	19.64	0.34	19.64	2.24	0.00
ELECTROLYTES (Note 1)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulfur	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	6,681.97	100.00	112.90	100.00	5,804.65	100.00	877.32	100.00	4,404.76
V&L MIXTURE PROPERTIES									
Vapor Fraction	1.0	1.0	1.0	0.0	1.0	1.0	0.0	0.0	0.0
Temperature	F	65.0	80.0	110.0	80.0	110.0	277.1	100.1	110.0
Pressure	psig	270.0	45.0	50.0	260.0	50.0	10.3	35.0	2,200.0
Mass Flow	lb/hr	18,911.0	4,465.2	650.1	12,606.5	6,304.5	1,398.9	734.1	193,647.0
Molecular Weight		2.83	38.55	18.02	2.17	7.19	32.07	18.02	43.96
Enthalpy	MM Btu/hr	-7,934	-11,855	-4,441	0.091	-7,834	0.070	-5,026	-762,250
Density	lb/ft ³	0.138	0.418	52.270	0.102	0.076	292.203	61.215	45.053
Vapor Phase Properties									
Std. Volume Flow	MMSCFD	60.9	1.0	52.9	34.2	8.0			
Flowing Volume	ft ³ /s	38.0	3.0	123.1	34.2	23.1			
Mass Flow	lb/hr	18,911.0	4,465.2	12,606.5	6,304.5	6,304.5			
Density	lb/ft ³	0.138	0.418	0.102	0.076	0.076			
Heat Capacity	Btu/lb-F	2.412	0.223	3.138	0.092	0.956			
Conductivity	Btu/hr-F	0.087	0.011	0.081	0.092	0.081			
Viscosity	cp	0.008	0.014	0.008	0.008	0.014			
Conc.	lb/hr	1,421	1,329	1,409	1,421	1,409			
2-Factor		1,006	0.978	1,006	1,006	1,002			
MW		2.83	39.35	2.17	2.17	7.19			
Liquid Phase Properties									
Flowing Volume	gpm			1.6			0.6	1.5	535.9
Mass Flow	lb/hr			650.1			1,398.9	734.1	193,647.0
Density	lb/ft ³			52.27			292.20	61.21	45.05
Heat Capacity	Btu/lb-F			1.08			0.20	1.08	0.89
Conductivity	Btu/hr-F			0.79					0.05
Viscosity	cp			0.83					0.07
Surface Tension	dyne/cm			69					0
SOLID PROPERTIES									
Mass Flow	lb/hr								
Enthalpy	MM Btu/hr								
V&L&S	lb/hr	18,911.0	4,465.2	650.1	12,606.5	6,304.5	1,398.9	734.1	193,647.0
TOTAL Mass Flow	lb/hr								
TOTAL Enthalpy flow	MM Btu/hr	-7,934	-11,855	-4,441	0.091	-7,834	0.070	-5,026	-762,250

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)
Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	101	102	103	104	105	106	107	108	109	110
DESCRIPTION	Hot Raw Syngas from Gasification	Quenched Syngas to Venturi Mixer	Quench Bottoms Water to MP Flash Drum	Quench Spray Water to Quench Drum	Scrubbed Syngas to Shift Inlet KO Drum	Flash Gas to Ammonia Stripper	Flash Drum Bottoms to Cooler	Seller Overflow to Gray Water Tank	Seller Bottoms to Solids Fillers	Filtrate to Gray Water Tank
VAL MIXTURE COMPONENTS										
H ₂	4,187.9	4,189.2	1.46	0.00	4,188.2	0.5	0.00	0.00	0.00	0.00
CO	128,967.0	128,950.0	44.26	8.9	128,916.0	45.13	0.00	0.00	0.00	0.00
CO ₂	3,005.0	3,016.4	1.05	3.1	3,133.4	1.11	0.00	0.00	0.00	0.00
CH ₄	105.4	105.4	0.04	0.00	105.4	0.04	0.00	0.00	0.00	0.00
H ₂ O	1,398.0	1,405.1	0.49	3.4	1,393.9	0.50	0.00	0.00	0.00	0.00
N ₂	171.2	171.2	0.06	0.00	171.2	0.06	0.00	0.00	0.00	0.00
NH ₃	1.7	6.4	0.00	0.00	16.1	0.01	0.00	0.00	0.00	0.00
HCN	13.4	17.0	0.01	0.00	14.9	0.01	0.00	0.00	0.00	0.00
HCL	68.4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HF	818.3	819.5	0.29	0.00	818.3	0.29	0.00	0.00	0.00	0.00
N ₂	1,101.8	1,101.9	0.36	0.00	1,101.7	0.39	0.00	0.00	0.00	0.00
H ₂ O	823.2	143,049.0	51.96	53,979.4	99.57	5,458.0	99.57	33,501.4	15,013.1	99.01
ELECTROLYTES (Note 1)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	138,571.2	285,831.3	100.00	54,421.0	281,201.1	5,473.9	100.00	33,800.1	15,147.0	100.00
VAL MIXTURE PROPERTIES										
Vapor Fraction	1.0	1.0	0.0	0.0	1.0	1.0	0.0	0.0	0.0	0.0
Temperature	F	405.7	405.0	401.7	401.2	320.3	320.3	16.2	140.5	140.5
Pressure	psig	500.0	490.0	600.0	475.0	75.0	75.0	0.0	50.0	50.0
Mass Flow	lb/hr	138,571.0	285,831.0	201,809.0	281,202.0	5,473.9	48,947.1	33,800.1	15,147.0	13,464.0
Molecular Weight	lb/lb-mole	20.28	19.04	18.02	18.07	18.03	18.09	18.09	18.09	18.09
Enthalpy	MM Btu/hr	-92,517	-1,054,500	-1,309,500	-1,023,000	-30,976	-320,360	-227,340	-101,870	-59,491
Density	lb/ft ³	0.303	1.084	53.658	1.055	0.199	56.325	51.012	61.017	60.984
Vapor Phase Properties										
MMSCFD	62.3	137.2			134.3	2.8				
Flowing Volume	ft ³ /s	127.1	73.5		74.0	7.6				
Mass Flow	lb/hr	138,571.0	285,831.0		281,202.0	5,473.9				
Density	lb/ft ³	0.303	1.084		1.055	0.199				
Heat Capacity	Btu/lb-F	0.402	0.438		0.435	0.473				
Conductivity	Btu/hr-ft-F	0.119	0.031		0.031	0.017				
Viscosity	cp	0.080	0.022		0.022	0.015				
Diffusivity	ft ² /s	1.323	1.410		1.406	1.348				
Z-Factor		1.005	0.965		0.957	0.970				
MMV	20.28	19.04			19.07	18.03				
Liquid Phase Properties										
Flowing Volume	gpm			488.8			108.2	59.1	30.9	27.5
Mass Flow	lb/hr			201,809.0			48,947.1	33,800.1	15,147.0	13,464.0
Density	lb/ft ³			53.67			56.32	51.01	61.02	60.98
Heat Capacity	Btu/lb-F			0.37			0.30	0.37	0.39	0.39
Conductivity	Btu/hr-ft-F			0.11			0.10	0.37	0.37	0.37
Viscosity	cp			0.12			0.17	0.48	0.48	0.48
Surface Tension	dyne/cm			37			47	66	66	66
SOLID PROPERTIES										
Mass Flow	lb/hr	1,761.3					1,683.0		1,683.0	
Enthalpy	MM Btu/hr	-5,926		-5,454			-6,482		-6,536	
VALS										
TOTAL Mass Flow	lb/hr	140,432.3	285,831.0	56,104.0	281,202.0	5,473.9	50,630.1	33,800.1	16,830.0	13,464.0
TOTAL Enthalpy flow	MM Btu/hr	-36,843	-1,054,500	-1,309,500	-1,023,000	-30,976	-328,842	-227,340	-108,406	-59,491

Note 1: Electrolytes are the sum of the ionic species and dissolved solids in the water phase.

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)
Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	111	112	113	114	115	201	202	203	204	205
DESCRIPTION	Filler Cake to Offsite Disposal	HP Gray Water to Gray Water Preheater	Blowdown Water to Offsite Treating	Makeup Water to Gray Water Tank	Caustic Addition to Quench Water	Bypass Syngas to Second Shift Reactor	Steam Addition for First Shift Reactor	Syngas/Steam to Shift Feed Effluent Exchanger	Syngas/Steam to First Shift Reactor	
V&L MIXTURE COMPONENTS										
H ₂	0.0	0.0	0.0	0.0	0.0	2,094.1	0.0	2,094.1	0.0	
CO	0.0	0.0	0.0	0.0	0.0	63,458.1	0.0	63,458.1	0.0	
CO ₂	0.0	0.0	0.0	0.0	0.0	1,566.7	0.0	1,566.7	0.0	
CH ₄	0.0	0.0	0.0	0.0	0.0	52.7	0.0	52.7	0.0	
H ₂ S	0.0	0.0	0.0	0.0	0.0	666.9	0.0	666.9	0.0	
COS	0.0	0.0	0.0	0.0	0.0	85.6	0.0	85.6	0.0	
NH ₃	0.0	0.0	0.0	0.0	0.0	8.1	0.0	8.1	0.0	
H ₂ O	0.0	0.0	0.0	0.0	0.0	7.4	0.0	7.4	0.0	
AR	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
N ₂	0.0	0.0	0.0	0.0	0.0	408.7	0.0	408.7	0.0	
H ₂ O	1,853.0	101,357.0	31,981.7	86,370.0	55.2	71,670.7	26,210.0	550.9	97,860.7	
ELECTROLYTES (Note 1)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Sulfur	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
TOTAL	1,853.0	101,357.2	32,277.1	86,370.0	694.0	140,601.0	26,210.0	166,610.0	166,610.0	
V&L MIXTURE PROPERTIES										
Vapor Fraction	0.0	0.0	0.0	0.0	0.0	1.0	1.0	1.0	1.0	
Temperature	120.5	50.4	120.6	80.0	100.0	401.2	498.3	416.8	464.0	
Pressure	0.0	600.0	50.0	50.0	600.0	475.0	650.0	475.0	465.0	
Mass Flow	1,853.0	101,357.0	32,277.1	86,370.0	694.0	140,601.0	26,210.0	166,610.0	166,610.0	
Molecular Weight	18.02	18.03	18.10	18.02	18.38	19.07	18.02	18.89	18.89	
Enthalpy	-11,360	-689,710	-217,040	-589,230	-4,475	-511,510	-147,720	-659,230	-659,590	
Density	61.370	62.173	61.022	62.233	75.839	1,066	1,398	1,045	0.960	
Vapor Phase Properties										
Std. Volume Flow						67.2	13.3	80.4	80.4	
Flowing Volume						37.0	5.2	44.3	44.3	
Mass Flow						140,601.0	26,210.0	166,610.0	166,610.0	
Density						1,056	1,398	1,045	0.960	
Heat Capacity						0.435	0.435	0.435	0.435	
Blurb-F						0.031	0.032	0.039	0.042	
Blurb-T-F						0.022	0.014	0.017	0.016	
Viscosity						1.408	1.367	1.434	1.444	
GrCv						0.857	0.855	0.841	0.852	
Z-Factor						19.07	16.02	18.89	18.89	
MMV										
Liquid Phase Properties										
Flowing Volume										
Mass Flow										
Density										
Heat Capacity										
Blurb-F										
Blurb-T-F										
Viscosity										
Surface Tension										
SOLID PROPERTIES										
Mass Flow										
Enthalpy										
TOTAL										
TOTAL Mass Flow										
TOTAL Enthalpy flow										

EnviroS Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)
Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	206	207	208	209	210	211	212	213	214	215
DESCRIPTION	Syngas from Second Shift Reactor	Syngas to Third Shift Reactor			Shifted Syngas from Shift Condensate Heater	Shifted Syngas from Gray Water Heater	Shifted Syngas from Steam Generator	Shifted Syngas from Water Heater KO Drum	Shift Condensate from Water Heater KO Drum	Shifted Syngas from Airfan Cooler KO Drum
V&L MIXTURE Components										
H ₂	12,103.8	3.94	12,103.8		13,093.5	4.26	13,093.5	4.26	13,093.7	4.02
CO	16,945.9	5.51	16,945.9		3,193.9	1.04	3,193.9	1.04	3,193.7	0.8
CO ₂	176,064.0	57.27	176,064.0		197,674.0	64.30	197,674.0	64.30	197,545.0	74.18
CH ₄	105.4	0.03	105.4		105.4	0.03	105.4	0.03	105.4	0.04
H ₂ S	1,488.1	0.48	1,488.1		1,480.2	0.48	1,480.2	0.48	1,486.7	0.56
COS	5.2	0.00	5.2		1.4	0.00	1.4	0.00	1.4	0.00
NH ₃	25.5	0.01	25.5		25.5	0.01	25.5	0.01	17.0	0.01
HCN	0.0	0.00	0.0		0.0	0.00	0.0	0.00	0.0	0.00
HCL	0.0	0.00	0.0		0.0	0.00	0.0	0.00	0.0	0.00
AR	818.3	0.27	818.3		819.3	0.27	819.3	0.27	819.3	0.31
N ₂	1,101.7	0.36	1,101.7		1,101.7	0.36	1,101.7	0.36	1,101.7	0.41
H ₂ O	98,752.2	32.12	98,752.2		89,906.3	29.25	89,906.3	29.25	48,948.3	18.38
ELECTROLYTES (Note 1)	0.0	0.00	0.0		0.0	0.00	0.0	0.00	0.0	0.00
Sulfur	0.0	0.00	0.0		0.0	0.00	0.0	0.00	27.7	0.07
TOTAL	307,411.1	100.00	307,411.1		307,411.2	100.00	307,411.2	100.00	266,311.1	100.00
V&L MIXTURE Properties										
Vapor Fraction	1.0		1.0		1.0		1.0000		1.0	
Temperature	717.7		420.0		364.0		345.0		299.9	
Pressure	425.0		405.0		375.0		365.0		354.0	
Mass Flow	307,411.0		307,411.0		307,411.0		307,411.0		266,311.0	
Molecular Weight	18.97		18.97		18.97		18.97		18.97	
Enthalpy	-1,192,400		-1,234,200		-1,250,900		-1,253,500		-1,025,200	
Density	0.664		0.860		0.859		0.858		0.877	
Vapor Phase Properties										
Std. Volume Flow	147.6		147.6		147.6		147.6		126.8	
Flowing Volume	128.6		99.3		99.4		99.5		84.3	
Mass Flow	307,411.0		307,411.0		307,411.0		307,411.0		266,311.0	
Density	0.664		0.860		0.859		0.858		0.877	
Heat Capacity	0.473		0.454		0.457		0.458		0.441	
Conductivity	0.058		0.050		0.056		0.055		0.051	
Viscosity	0.022		0.022		0.017		0.016		0.020	
Cp/Cv	1.311		1.344		1.300		1.303		1.349	
Z-Factor	0.995		0.891		0.974		0.972		0.986	
MW	18.97		18.97		18.97		18.97		19.12	
Liquid Phase Properties										
Flowing Volume										
Mass Flow										
Density										
Heat Capacity										
Conductivity										
Viscosity										
Surface Tension										
SOLID Properties										
Mass Flow										
Enthalpy										
V&L &S										
TOTAL Mass Flow	307,411.0	307,411.0	307,411.0		307,411.0	307,411.0	307,411.0	266,311.0	41,099.8	220,124.0
TOTAL Enthalpy flow	-1,192,400	-1,234,200	-1,234,200		-1,250,900	-1,253,500	-1,253,500	-1,025,200	-270,850	-775,920

Enviro Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)
Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	216	217	218	219	220	221	222	223	224	225
DESCRIPTION	Shift Condensate from Airfin Cooler KO Drum	Shifted Syngas from Trim Cooler KO Drum	Shift Condensate from Trim Cooler KO Drum	Shifted Syngas from Mercury Removal Drum	Hot Gray Water to Syngas Scrubber	Hot Shift Condensate to Syngas Scrubber	Sour Gas to Ammonia Stripper Overhead Condenser	Shift Condensate to Ammonia Stripper	Sour Gas to Sulfur Recovery System	Shifted Condensate to HP Shift Condensate Drum
V&L MIXTURE Components										
H ₂	1.0	13,091.7	6.01	13,091.7	6.01	0.0	0.0	0.0	1.5	0.0
CO	0.2	3,193.5	1.47	3,193.5	1.47	0.0	0.0	0.0	9.1	0.0
CO ₂	249.1	197,231.0	90.61	197,231.0	90.61	0.0	121.3	122.5	318.4	0.0
CH ₄	0.0	105.4	0.05	105.4	0.05	0.0	0.0	0.0	0.0	0.0
H ₂ S	6.2	1,479.5	0.88	1,479.5	0.88	0.0	1.7	4.7	11.2	0.0
COS	0.0	1.4	0.00	1.4	0.00	0.0	0.0	0.0	0.0	0.0
NH ₃	0.2	0.0	0.00	0.0	0.00	0.0	0.0	0.0	0.0	0.0
H ₂ O	0.0	0.0	0.00	0.0	0.00	0.0	0.0	0.0	10.7	6.4
HCL	0.0	0.0	0.00	0.0	0.00	0.0	0.0	0.0	0.0	0.0
AR	0.1	819.2	0.38	819.2	0.38	0.0	0.0	0.0	0.0	0.0
N ₂	0.1	1,101.7	0.51	1,101.7	0.51	0.0	0.0	0.0	0.1	0.0
H ₂ O	45,854.4	99,281.0	0.30	45,854.4	99,281.0	0.0	0.0	0.0	0.0	0.0
ELECTROLYTES (Note 1)	76.5	0.0	0.00	76.5	0.00	0.0	0.0	0.0	0.0	0.0
Sulfur	0.0	0.0	0.00	0.0	0.00	0.0	0.0	0.0	0.0	0.0
TOTAL	46,187.6	217,673.4	100.00	217,673.4	100.00	94,822.3	158.0	48,480.3	389.8	53,722.3
V&L MIXTURE Properties										
Vapor Fraction	0.0	1.0	0.0	1.0	0.0	0.0	1.0	0.0	1.0	0.0
Temperature	160.0	100.0	100.0	109.6	353.0	400.0	157.1	157.2	200.0	286.2
Pressure	343.0	332.0	332.0	312.0	580.0	590.0	60.0	100.0	45.0	400.0
Mass Flow	46,187.6	217,673.0	217,673.0	217,673.0	101,357.0	94,822.3	158.0	48,480.3	389.8	53,722.3
Molecular Weight	18.09	19.37	18.12	19.37	18.03	18.03	37.48	18.03	34.17	18.02
Enthalpy	-310.470	-767.220	-16.599	-766.740	-662.830	-615.400	-603.0	-326.450	-1.466	-354.750
Density	60.694	1.130	61.375	1.059	55.566	53.688	0.429	60.933	0.292	57.488
Vapor Phase Properties										
Std. Volume Flow	102.3	102.3	102.3	102.3	0.0	0.0	0.0	0.0	0.1	0.0
Flowing Volume	53.5	53.5	53.5	57.1	0.0	0.0	0.1	0.0	0.4	0.0
Mass Flow	217,673.0	217,673.0	217,673.0	217,673.0	101,357.0	94,822.3	158.0	48,480.3	389.8	53,722.3
Density	1.130	1.130	1.130	1.059	0.429	0.429	0.292	0.292	0.292	0.292
Heat Capacity	0.410	0.410	0.410	0.413	0.243	0.243	0.243	0.243	0.243	0.243
Blu/ft-F	0.048	0.048	0.048	0.040	0.017	0.017	0.017	0.017	0.017	0.017
Conductivity	0.015	0.015	0.015	0.014	0.017	0.017	0.017	0.017	0.017	0.017
Viscosity	1.380	1.380	1.380	1.383	1.301	1.301	1.297	1.297	1.297	1.297
Cp/Cv	0.989	0.989	0.989	0.978	0.986	0.986	0.986	0.986	0.987	0.987
Z-Factor	19.37	19.37	19.37	19.37	37.48	37.48	37.48	37.48	34.17	34.17
MM	19.37	19.37	19.37	19.37	37.48	37.48	37.48	37.48	34.17	34.17
Liquid Phase Properties										
Flowing Volume	94.9	94.9	94.9	94.9	227.4	220.2	99.2	99.2	99.2	116.5
Mass Flow	46,187.6	46,187.6	46,187.6	46,187.6	101,357.0	94,822.3	48,480.3	48,480.3	48,480.3	53,722.3
Density	60.69	60.69	60.69	60.69	55.57	53.69	60.93	60.93	60.93	57.49
Heat Capacity	1.00	1.00	1.00	1.00	1.05	1.08	1.00	1.00	1.00	1.03
Blu/ft-F	0.31	0.31	0.31	0.31	0.39	0.38	0.34	0.34	0.34	0.40
Conductivity	0.38	0.38	0.38	0.38	0.14	0.14	0.14	0.14	0.14	0.17
Viscosity	64	64	64	64	43	37	64	64	64	49
Surface Tension	64	64	64	64	43	37	64	64	64	49
SOLID Properties										
Mass Flow										
Enthalpy										
MM										
VALS										
TOTAL Mass Flow	46,187.6	217,673.0	217,673.0	217,673.0	101,357.0	94,822.3	158.0	48,480.3	389.8	53,722.3
TOTAL Enthalpy flow	-310.470	-767.220	-16.599	-766.740	-662.830	-615.400	-603.0	-326.450	-1.466	-354.750

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)

Case 1 - Hydrogen Production - Revision 0 (June 25, 2009)

PFID Stream Number	301	302	303	304	401	402	501	502	601
DESCRIPTION	Sweet Syngas to PSA	CO2 to CO2 Compression	Acid Gas to SRU	AGR Waste Water to Offsite Treating	Hydrogen Product	Fuel Gas to Offsite	Sulfur Product	SRU Waste Water to Offsite Treating	CO2 Product
V&L MIXTURE COMPONENTS									
H2	13,079.2 68.16	0.00 0.00	3.5 0.06	0.00 0.00	11,640.5 0.3	1,438.7 0.0	0.00 0.00	0.00 0.00	0.00 0.00
CO	3,171.4 16.77	0.00 0.00	2.4 0.05	0.00 0.00	0.00 0.00	3,171.1 0.0	0.00 0.00	0.00 0.00	0.00 0.00
CO2	639.0 3.38	193,647.0 99.98	2,591.2 0.00	0.00 0.00	0.00 0.00	638.3 0.0	0.00 0.00	0.00 0.00	193,647.0 99.98
CH4	101.6 0.54	0.00 0.00	0.00 0.00	0.00 0.00	5.1 0.04	96.5 0.0	0.00 0.00	0.00 0.00	0.00 0.00
H2S	0.2 0.00	0.00 0.00	1,476.2 0.00	0.00 0.00	0.00 0.00	0.2 0.00	0.00 0.00	0.00 0.00	0.00 0.00
COS	0.0 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00
NH3	0.0 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00
HCN	0.0 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00
HCL	0.0 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00
AR	819.2 4.33	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00
N2	1,100.3 5.82	0.00 0.00	0.00 0.00	0.00 0.00	499.6 0.0	409.6 0.0	0.00 0.00	0.00 0.00	0.00 0.00
H2O	0.0 0.00	0.00 0.00	0.00 0.00	0.00 0.00	550.2 0.0	550.2 0.0	0.00 0.00	0.00 0.00	0.00 0.00
ELECTROLYTES (Note 1)	0.0 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00
Sulfur	0.0 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00	0.00 0.00
TOTAL	18,911.0 100.00	193,646.3 100.00	4,465.2 100.00	650.1 100.00	12,606.5 100.00	5,304.5 100.00	1,398.9 100.00	734.1 100.00	193,646.3 100.00
V&L MIXTURE PROPERTIES									
Vapor Fraction	1.0	1.0	1.0	0.0	1.0	1.0	0.0	0.0	0.0
Temperature	80.0	65.0	80.0	110.0	80.0	110.0	277.1	100.1	110.0
Pressure	270.0	40.0	45.0	50.0	260.0	50.0	10.3	35.0	2,200.0
Mass Flow	18,911.0	193,647.0	4,465.2	650.1	12,606.5	5,304.5	1,398.9	734.1	193,647.0
Molecular Weight	2.83	43.96	38.55	18.02	2.17	7.19	3.0	6.02	43.96
Enthalpy	-7,934	-745,570	-11,855	-4,441	0.091	-7,834	0.070	-5,026	-162,250
Density	0.138	0.437	0.418	52.270	0.102	0.076	282.203	61.215	45.053
Vapor Phase Properties									
Sid. Volume Flow	60.9	40.1	1.0		52.9	8.0			
Flowing Volume	38.0	123.1	3.0		34.2	23.1			
Mass Flow	18,911.0	193,647.0	4,465.2		12,606.5	5,304.5			
Density	0.138	0.437	0.418		0.102	0.076			
Heat Capacity	2.412	0.223	0.223		3.138	0.956			
Conductivity	0.087	0.014	0.014		0.082	0.081			
Viscosity	0.008	0.014	0.014		0.008	0.014			
Cap/Cov	1.421	1.318	1.329		1.421	1.409			
Z-Factor	1.005	0.978	0.978		1.005	1.002			
MW	2.83	43.96	38.55		2.17	7.19			
Liquid Phase Properties									
Flowing Volume	gpm			1.6				1.5	555.9
Mass Flow	lb/hr			650.1				734.1	193,647.0
Density	lb/ft ³			52.27				61.21	45.05
Heat Capacity	Btu/lb-F			1.08				1.08	0.69
Conductivity	Btu/hr-ft-F			0.79				0.05	0.05
Viscosity	cp			0.93				0.07	0
Surface Tension	dynes/cm			69					
SOLID PROPERTIES									
Mass Flow	lb/hr								
Enthalpy	MM Btu/hr								
V&L&S									
TOTAL Mass Flow	lb/hr	18,911.0	193,647.0	650.1	12,606.5	5,304.5	1,398.9	734.1	193,647.0
TOTAL Enthalpy Flow	MM Btu/hr	-7,934	-745,570	-4,441	0.091	-7,834	0.070	-5,026	-162,250

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFID Stream Number	101	102	103	104	105	106	107	108	109	110	111
DESCRIPTION	Hot Raw Syngas from Gasification	Quenched Syngas to Venturi Mixer	Quenched Bottoms Water to MP Flash Drum	Quench Spray Water to Quench Drum	Scrubbed Syngas to Shift Inlet KO Drum	Flash Gas to Ammonia Stripper	Flash Drum Bottoms to Cooler	Settler Overflow to Gray Water Tank	Settler Bottoms to Solids Filters	Filtrate to Gray Water Tank	Filter Cake to Offsite Disposal
VAL MIXTURE COMPONENTS											
H ₂	2,077.45	2,078.10	13.79	0.24	0.01	0.89	0.01	2,077.94	14.08	0.24	0.08
CO	4,532.85	66.29	30.08	0.32	0.01	1.20	0.01	4,531.28	30.71	0.32	0.10
CO ₂	68.37	1.00	0.45	0.07	0.00	0.26	0.00	72.39	0.49	0.00	0.00
CH ₄	6.57	0.10	0.04	0.00	0.00	0.00	0.00	6.57	0.04	0.00	0.00
H ₂ S	41.02	0.60	0.27	0.10	0.00	0.38	0.00	41.05	0.28	0.12	0.04
COS	2.85	0.04	0.02	0.00	0.00	0.00	0.00	2.85	0.02	0.00	0.00
NH ₃	0.10	0.00	0.00	0.00	0.00	0.29	0.00	0.85	0.02	0.00	0.00
HCN	0.57	0.01	0.03	0.00	0.00	0.09	0.00	0.67	0.01	0.03	0.01
ACN	1.82	0.03	0.00	0.00	0.00	0.00	0.00	1.82	0.00	0.00	0.00
N ₂	20.51	0.30	0.14	0.00	0.00	0.00	0.00	20.51	0.14	0.00	0.00
H ₂ O	39.33	0.55	0.26	0.00	0.00	0.01	0.00	39.33	0.27	0.00	0.00
ELECTROLYTES (Note 1)	45.97	0.67	0.31	0.00	0.00	11,192.48	99.94	7,961.76	53.96	302.90	99.74
Sulfur	0.00	0.00	0.00	0.00	0.00	3.96	0.04	0.00	0.00	0.00	0.00
TOTAL	6,837.41	100.00	15,065.70	100.00	3,008.83	11,192.57	100.00	14,755.11	106.00	303.69	100.00
VAL MIXTURE PROPERTIES											
Vapor Fraction	1.0	1.0	0.0	0.0	1.0	1.0	0.0	1.0	0.0	0.0	0.0
Temperature	2,732.0	405.7	405.0	401.8	401.2	320.3	320.3	140.2	140.5	140.6	140.5
Pressure	500.0	480.0	480.0	600.0	475.0	475.0	475.0	33,801.8	15,147.0	13,464.0	13,464.0
Mass Flow	138,671.0	286,864.0	54,423.5	201,844.0	281,358.0	5,474.7	48,948.8	33,801.8	15,147.0	13,464.0	13,464.0
Molecular Weight	20.28	19.04	18.09	18.02	19.07	18.03	18.09	18.03	18.09	18.10	18.09
Enthalpy	-92,917	-1,054,600	-351,380	-1,309,700	-1,023,800	-30,381	-320,370	-227,350	-107,870	-90,497	-11,360
Density	0.303	1.084	53.222	53.662	1.056	0.189	56.325	61.012	61.017	60.864	61.370
Vapor Phase Properties											
Std. Volume Flow	62.3	137.2			134.4	2.8					
Flowing Volume	127.1	73.5			74.0	7.6					
Mass Flow	138,671.0	286,864.0			281,358.0	5,474.7					
Density	0.303	1.084			1.056	0.189					
Heat Capacity	0.402	0.438			0.35	0.35					
Conductivity	0.119	0.031			0.031	0.017					
Viscosity	0.060	0.022			0.022	0.015					
Cp/Cv	1.323	1.410			1.408	1.348					
Z-Factor	1.005	0.955			0.957	0.970					
MW	20.28	19.04			19.07	18.03					
Liquid Phase Properties											
Flowing Volume			127.5	469.0		105.3		69.1	30.8	27.5	3.4
Mass Flow			54,423.5	201,844.0		48,948.8		33,801.8	15,147.0	13,464.0	1,683.0
Density			53.22	53.66		56.33		61.01	61.02	60.86	61.37
Heat Capacity			1.07	1.08		1.03		0.99	0.99	0.89	1.00
Conductivity			0.37	0.37		0.40		0.37	0.37	0.37	0.37
Viscosity			0.12	0.11		0.17		0.48	0.48	0.45	0.45
Surface Tension			36	37		47		66	66	66	66
SOLID PROPERTIES											
Mass Flow			1,683.0			1,683.0			1,683.0		1,683.0
Enthalpy			-6,454			-6,462			-6,536		-6,536
VALS&S											
TOTAL Mass Flow	140,432.3	286,864.0	58,106.5	201,844.0	281,358.0	5,474.7		33,801.8	16,830.0	13,464.0	3,366.0
TOTAL Enthalpy flow	-96,843	-1,054,600	-357,834	-1,309,700	-1,023,800	-30,381		-227,350	-108,406	-90,497	-17,916

Note 1: Electrolytes are the sum of the ionic species and dissolved solids in the water phase.

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)
Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFID Stream Number	112	113	114	115	201	202	203	204	205	206	207
DESCRIPTION	HP Gray Water to Gray Water Preheater	Blowdown Water to Onsite Treating	Makeup Water to Gray Water Tank	Caustic Addition to Quench Water	Bypass Syngas to Second Shift Reactor	Steam Addition for First Shift Reactor	Syngas/Steam to Shift Exchanger	Syngas/Steam to First Shift Reactor		Total Syngas to Second Shift Reactor	
	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr	lbmol/hr		lbmol/hr	
	Mol %	Mol %	Mol %	Mol %	Mol %	Mol %	Mol %	Mol %		Mol %	
VAL MIXTURE Components											
H2	0.00	0.00	0.00	0.00	632.73	14.08	632.73	632.73		2,380.62	24.12
CO	0.00	0.00	0.00	0.00	1,379.78	30.71	1,379.78	1,379.78		1,644.57	16.66
CO2	0.00	0.00	0.00	0.00	22.04	0.49	22.04	22.04		1,160.07	11.75
CH4	0.00	0.00	0.00	0.00	0.00	0.04	0.00	0.04		4.00	0.04
H2S	0.00	0.00	0.00	0.00	12.50	0.28	12.50	12.50		25.83	0.26
COS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.90	0.01
NH3	0.01	0.02	0.00	0.00	0.26	0.01	0.26	0.26		0.69	0.01
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.17	0.00
HCL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		12.49	0.13
N2	0.00	0.00	0.00	0.00	11.98	0.27	11.98	11.98		23.95	0.24
H2O	4,151.46	99.54	3,327.18	30.62	2,424.36	53.96	885.39	100.00		4,617.94	46.78
ELECTROLYTES (Note 1)	3.83	0.09	8.24	0.46	0.00	0.00	0.00	0.00		0.00	0.00
Sulfur	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
TOTAL	4,155.30	100.00	3,327.18	37.76	4,492.33	100.00	885.39	5,378.32		9,871.26	100.00
VAL MIXTURE Properties											
Vapor Fraction	0.0	0.0	0.0	0.0	0.0	0.0	1.0	1.0		1.0	
Temperature	83.5	140.6	80.0	100.0	401.2	495.3	416.8	454.0		474.0	
Pressure	60.0	60.0	60.0	60.0	475.0	495.0	475.0	495.0		435.0	
Mass Flow	74,927.0	32,278.8	59,940.0	694.0	85,673.5	15,950.6	101,624.0	101,624.0		187,298.0	
Molecular Weight	18.03	16.10	18.02	18.38	18.02	18.02	18.02	18.02		18.02	
Enthalpy	-509,480	-217,050	-408,920	-4,475	-311,730	-39,897	-401,630	-399,410		-728,550	
Density	62.116	61.002	62.233	75.839	1.095	1.398	1.045	0.960		0.877	
Vapor Phase Properties											
MASS/FD											
Std. Volume Flow											
Flowing Volume											
Mass Flow											
Density											
Heat Capacity											
Blow-F											
Blow-F											
Blow-F											
Conductivity											
Viscosity											
CP											
Thermal Factor											
MM											
Liquid Phase Properties											
Flowing Volume											
Mass Flow	150.4	66.0	120.1	1.1	22.5	8.1	49.0	49.0		98.9	
Density	74,927.0	32,278.8	59,940.0	694.0	85,673.5	15,950.6	101,624.0	101,624.0		98.9	
Heat Capacity	62.12	61.00	62.23	75.84	1.095	1.398	1.045	0.960		0.877	
Blow-F	0.99	0.99	1.00	0.88	0.433	0.384	0.433	0.434		0.432	
Blow-F	0.36	0.37	0.35	0.44	0.031	0.032	0.039	0.042		0.051	
Conductivity	0.67	0.48	0.74	2.78	0.022	0.022	0.017	0.016		0.019	
Viscosity	71	66	72	81	1.095	1.398	1.045	0.960		0.877	
Surface Tension											
SOLID Properties											
Mass Flow											
Enthalpy											
MM											
VAL &S											
TOTAL Mass Flow	74,927.0	32,278.8	59,940.0	694.0	85,673.5	15,950.6	101,624.0	101,624.0		187,298.0	
TOTAL Enthalpy flow	-509,480	-217,050	-408,920	-4,475	-311,730	-39,897	-401,630	-399,410		-728,550	

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFID Stream Number	208	209	210	211	212	213	214	215	216	217	218
DESCRIPTION	Bypass Syngas to Shift Cooling	Shifted Syngas from Second MP Steam Generator	Shifted Syngas from Shift Condensate Heater	Shifted Syngas from Gray Water Heater	Shifted Syngas from Steam Generator	Shifted Syngas from LP Water Heater KO Drum	Shift Condensate from Water Heater KO Drum	Shifted Syngas from Airfan Cooler KO Drum	Shift Condensate from Airfan Cooler KO Drum	Shifted Syngas from Trim Cooler KO Drum	Shift Condensate from Trim Cooler KO Drum
VAL MIXTURE COMPONENTS											
H ₂	812.46	3,657.04	37.05	1,002.17	28.58	4,488.79	39.85	48.02	0.32	4,488.45	0.02
CO	1,771.73	30.71	368.33	3.73	479.85	1,368	1,660.21	13.68	0.25	2,139.87	0.02
CO ₂	28.31	2,437.34	24.69	552.85	15.76	1,912.71	15.76	2,461.53	3.67	2,139.87	0.02
CH ₄	2.57	4.00	4.00	0.04	1.47	5.10	0.04	6.57	0.00	2,457.69	0.00
H ₂ O	16.05	28.68	0.27	26.68	0.27	33.14	0.27	42.47	0.38	6.57	0.00
N ₂	1.11	0.02	0.05	0.00	0.28	0.01	0.81	0.00	0.17	42.27	0.01
NH ₃	0.33	0.01	0.01	0.01	0.28	0.01	0.81	0.00	0.17	0.01	0.01
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCL	8.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCN	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N ₂	3,113.05	53.96	23.95	33.84	1,447.03	41.26	5,006.43	18.13	4,420.31	27.36	99.87
Electrolytes (Note 1)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Solids	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	5,763.25	100.00	3,871.25	100.00	3,906.95	100.00	12,133.45	100.00	4,426.38	100.00	100.00
VAL MIXTURE PROPERTIES											
Vapor Fraction	1.0	1.0	1.0	0.945789000	0.8741	1.0	0.0	1.0	0.0	1.0	0.0
Temperature	401.2	550.0	410.0	356.4	356.4	289.9	289.9	160.0	160.0	100.0	100.0
Pressure	475.0	415.0	405.0	385.4	385.4	384.0	384.0	373.0	373.0	362.0	362.0
Mass Flow	110,011.0	187,298.0	187,298.0	66,663.4	230,645.0	217,556.0	79,852.3	182,992.0	34,464.2	181,138.0	1,854.5
Molecular Weight	19.07	18.97	18.97	19.01	19.01	19.01	19.01	19.01	19.01	19.01	19.01
Enthalpy	-400,290	-741,240	-753,300	-262,980	-924,780	-712,330	-526,390	-231,910	-520,290	-12,571	-12,571
Density	1.056	0.762	0.875	0.966	1.053	0.959	0.7221	1.151	0.6762	1.243	61.501
Vapor Phase Properties											
Std. Volume Flow	52.5	89.9	89.9	30.2	98.6	102.1	84.8	84.8	70.7	83.8	3.8
Flowing Volume	28.9	66.2	66.2	19.2	60.7	63.0	44.2	44.2	40.5	40.5	1.854.5
Mass Flow	110,011.0	187,298.0	187,298.0	63,221.7	203,105.0	217,556.0	182,992.0	182,992.0	34,464.2	181,138.0	1,854.5
Density	1.056	0.762	0.875	0.916	0.920	0.959	1.151	1.151	1.243	1.243	61.501
Heat Capacity	0.435	0.465	0.458	0.440	0.435	0.435	0.394	0.394	0.390	0.390	0.390
Conductivity	0.031	0.052	0.054	0.041	0.042	0.042	0.042	0.042	0.042	0.042	0.042
Viscosity	0.022	0.020	0.018	0.018	0.021	0.020	0.021	0.021	0.021	0.021	0.021
Opt/Ov	1.408	1.332	1.358	1.371	1.369	1.366	1.366	1.366	1.366	1.366	1.366
Z-Factor	0.957	0.987	0.975	0.973	0.977	0.989	0.989	0.989	0.989	0.989	0.989
MW	19.07	18.97	18.97	19.07	19.15	19.39	19.66	19.66	19.66	19.66	19.66
Liquid Phase Properties											
Flowing Volume				7.8	61.6	102.1	74.0	74.0	70.7	83.8	3.8
Mass Flow				3,441.7	27,540.5	217,556.0	79,852.3	79,852.3	34,464.2	181,138.0	1,854.5
Density				55.31	55.72	57.22	1.03	1.03	1.00	1.00	0.99
Heat Capacity				1.05	1.05	1.05	0.36	0.36	0.32	0.32	0.32
Conductivity				0.13	0.13	0.13	0.17	0.17	0.38	0.38	0.61
Viscosity				0.13	0.13	0.13	0.17	0.17	0.38	0.38	0.61
Surface Tension				42	43	43	49	49	64	64	70
SOLID PROPERTIES											
Mass Flow											
Enthalpy											
VAL &S											
TOTAL Mass Flow	110,011.0	187,298.0	187,298.0	66,663.4	230,645.0	217,556.0	79,852.3	182,992.0	34,464.2	181,138.0	1,854.5
TOTAL Enthalpy Flow	-400,290	-741,240	-753,300	-262,980	-924,780	-712,330	-526,390	-231,910	-520,290	-12,571	-12,571

Envires Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFID Stream Number	219	220	221	222	223	224	225	226	301	302	303
DESCRIPTION	Shifted Syngas from Mercury Removal Drum	Hot Gray Water to Syngas Scrubber	Hot Shift Condensate to Syngas Scrubber	Sour Gas to Ammonia Stripper Overhead Condenser	Shift Condensate to Ammonia Stripper	Sour Gas to Sulfur Recovery System	Stripper Condensate to HP Shift Condensate Drum	Mixed Bypass & Shifted Syngas	Sweet Syngas to Syngas Compression	CO2 to CO2 Compression	Acid Gas to SRU
V&L MIXTURE COMPONENTS											
H2	4,468.45	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO	2,139.67	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2	2,457.89	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH4	6.57	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C2H6	42.27	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C2H4	1.17	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C2H2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ar	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ne	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
He	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Electrolytes (Note 1)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulfur	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	9,203.17	4,153.30	8,335.51	100.00	2,007.86	100.00	2,397.12	100.00	6,663.13	2,397.06	115.45
V&L MIXTURE PROPERTIES											
Vapor Fraction	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Temperature	109.7	344.0	364.7	157.3	157.3	200.0	286.2	407.1	80.0	65.0	80.0
Pressure	342.0	580.0	580.0	60.0	60.0	40.0	41.863.3	297.309.0	70.868.8	105.147.0	45.0
Mass Flow	181,138.0	74,927.0	121,411.0	75.5	36,243.2	18.05	32.51	1,153.600	10.84	43.86	4,605.1
Molecular Weight	19.68	18.03	18.03	35.01	35.01	32.51	32.51	1,153.600	10.84	43.86	4,605.1
Enthalpy	-519.850	-450.400	-785.580	-0.278	-244.090	-0.825	-374.470	-1,153.600	-102.390	-404.050	-12,244
Density	1.166	55.804	53.853	0.400	60.922	0.277	57.489	0.864	0.576	0.436	0.421
Vapor Phase Properties											
Std. Volume Flow	83.8			0.0		0.1		142.4	60.7	21.8	1.1
Flowing Volume	43.2			0.1		0.2		93.4	34.2	67.0	3.0
Mass Flow	181,138.0			75.5		20.1		297,309.0	70,868.8	105,147.0	4,605.1
Density	1.166			0.400		0.277		0.864	0.576	0.436	0.421
Heat Capacity	0.383			0.253		0.275		0.864	0.576	0.436	0.421
Conductivity	0.036			0.019		0.018		0.018	0.018	0.011	0.011
Viscosity	0.014			0.017		0.017		0.018	0.018	0.011	0.011
CP/CV	1.406			1.310		1.304		1.310	1.436	1.318	1.328
Z-Factor	0.966			0.868		0.988		0.979	1.004	0.976	0.976
MW	19.68			35.01		32.51		18.01	10.64	43.86	35.39
Liquid Phase Properties											
Flowing Volume	167.4			74.2							
Mass Flow	74,927.0			36,243.2							
Density	55.80			60.92							
Heat Capacity	1.04			1.00							
Conductivity	0.38			0.35							
Viscosity	0.14			0.39							
Surface Tension	43			64							
SOLID PROPERTIES											
Mass Flow											
Enthalpy											
V&L&S											
TOTAL Mass Flow	181,138.0	74,927.0	121,411.0	75.5	36,243.2	20.1	41,563.3	297,309.0	70,868.8	105,147.0	4,605.1
TOTAL Enthalpy flow	-519.850	-450.400	-785.580	-0.278	-244.090	-0.825	-274.470	-1,153.600	-102.390	-404.050	-12,244

EnviroS Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Molar Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFD Stream Number	304	401	402	501	502	601	701	702
DESCRIPTION	AGR Waste Water to Offsite Treating	Hydrogen Product	Fuel Gas to Offsite	Sulfur Product	SRU Waste Water to Offsite Treating	CO2 Product	Treated Syngas to Fischer-Tropsch	Sweet Syngas to Hydrogen Purification
V&L MIXTURE COMPONENTS								
CH4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C2H6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
C3H8	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CH4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2S	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NH3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HCL	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
AR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H2O	27.36	99.41	0.00	0.00	0.00	0.00	0.00	0.00
ELECTROLYTES (Note 1)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulfur	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
TOTAL	27.36	100.00	100.00	100.00	100.00	100.00	100.00	100.00
V&L MIXTURE PROPERTIES								
Vapor Fraction	0.0	1.0	1.0	0.0	0.0	0.0	1.0	1.0
Temperature	110.0	120.0	110.0	277.2	100.1	110.0	400.0	110.0
Pressure	50.0	475.0	50.0	10.3	35.0	2,200.0	475.0	485.0
Mass Flow	497.3	384.5	3,599.0	1,399.7	674.7	105,147.0	66,905.3	3,983.5
Molecular Weight	18.07	2.33	17.16	32.07	18.02	43.86	10.64	10.64
Enthalpy	-3.357	0.047	-5.719	0.070	-4.619	-413.000	-82.765	-5.680
Density	51.787	0.182	0.182	281.730	61.215	44.457	0.558	0.862
Vapor Phase Properties								
MMSCFD		1.5	1.9				57.3	3.4
Std. Volume Flow		0.6	5.5				33.3	1.3
Flowing Volume		384.5	3,599.0				66,905.3	3,983.5
Mass Flow		0.182	0.182				0.558	0.862
Density		2.935	0.404				0.663	0.657
Heat Capacity		0.096	0.034				0.075	0.053
Conductivity		0.009	0.014				0.016	0.012
Viscosity		1.423	1.409				1.403	1.430
PrCo		1.011	1.000				1.011	1.009
Z-Factor		2.33	17.16				10.64	10.64
NW								
Liquid Phase Properties								
Flowing Volume	1.2			0.6	1.4	294.9		
Mass Flow	497.3			1,399.7	674.7	105,147.0		
Density	51.79			291.73	61.21	44.46		
Heat Capacity	1.07			0.20	1.08	0.69		
Conductivity	0.78					0.67		
Viscosity	0.82					0.07		
Surface Tension	69					0		
SOLID PROPERTIES								
Mass Flow								
Enthalpy								
MM Btu/hr								
VALUES								
TOTAL Mass Flow	497.3	384.5	3,599.0	1,399.7	674.7	105,147.0	66,905.3	3,983.5
TOTAL Enthalpy flow	-3.357	0.047	-5.719	0.070	-4.619	-413.000	-82.765	-5.680

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)
Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PPFD Stream Number	101	102	103	104	105	105	107	108	109	110	111
DESCRIPTION	Hot Raw Syngas from Gasification	Quenched Syngas to Venturi Mixer	Quench Bottoms Water to MP Flash Drum	Quench Spray Water to Quench Drum	Scrubbed Syngas to Shift Inlet KO Drum	Flash Gas to Ammonia Stripper	Flash Drum Bottoms to Cooler	Sellier Overflow to Gray Water Tank	Sellier Bottoms to Solids Filters	Filtrate to Gray Water Tank	Filter Cake to Offsite Disposal
V&L MIXTURE COMPONENTS											
H ₂	4,167.9	4,169.2	0.5	1.6	4,188.9	0.5	0.0	0.0	0.0	0.0	0.0
CO	126,967.0	126,950.0	8.9	33.7	126,923.0	45.11	8.9	0.0	0.0	0.0	0.0
CO ₂	3,039.0	3,016.5	3.1	11.2	3,186.0	1.13	3.8	0.0	0.0	0.0	0.0
CH ₄	1,054.4	1,054.4	0.0	0.0	1,054.4	0.0	0.0	0.0	0.0	0.0	0.0
H ₂ S	1,398.0	1,398.0	0.0	0.0	1,398.0	0.0	0.0	0.0	0.0	0.0	0.0
COS	171.2	171.2	0.0	0.0	171.2	0.0	0.0	0.0	0.0	0.0	0.0
NH ₃	1.7	1.7	0.0	0.0	1.7	0.0	0.0	0.0	0.0	0.0	0.0
H ₂ O	15.4	15.4	0.0	0.0	15.4	0.0	0.0	0.0	0.0	0.0	0.0
HCL	66.4	66.4	0.0	0.0	66.4	0.0	0.0	0.0	0.0	0.0	0.0
AR	819.3	819.3	0.0	0.0	819.3	0.0	0.0	0.0	0.0	0.0	0.0
N ₂	1,101.8	1,101.8	0.0	0.0	1,101.8	0.0	0.0	0.0	0.0	0.0	0.0
H ₂ O	828.2	828.2	0.0	0.0	828.2	0.0	0.0	0.0	0.0	0.0	0.0
ELECTROLYTES (Note 1)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Sulfur	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL	138,671.2	138,671.2	100.00	286,364.1	100.00	54,423.5	100.00	54,423.5	100.00	134,464.0	100.00
V&L MIXTURE PROPERTIES											
Vapor Fraction	1.0	1.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0	0.0	0.0
Temperature	2,732.0	405.7	408.0	401.8	401.2	320.3	320.3	140.2	140.5	140.6	140.5
Pressure	500.0	490.0	490.0	600.0	475.0	75.0	75.0	0.0	5.0	25.0	0.0
Mass Flow	138,671.0	286,364.0	54,423.5	201,844.0	281,358.0	5,474.7	48,948.8	33,801.8	15,147.0	13,464.0	1,683.0
Molecular Weight	20.28	19.04	18.08	18.02	19.07	18.03	18.09	18.09	18.09	18.10	18.02
Enthalpy	-92,917	-1,054,600	-351,380	-1,309,700	-1,023,800	-30,981	-320,370	-227,350	-101,870	-90,491	-11,580
Density	0.303	1.084	53.222	53.662	1.056	0.189	56.325	61.012	61.017	60.984	61.370
Vapor Phase Properties											
Std. Volume Flow	62.3	137.2			134.4	28					
Flowing Volume	127.1	73.5			74.0	7.6					
Mass Flow	138,671.0	286,364.0			281,358.0	5,474.7					
Density	0.303	1.084			1.056	0.189					
Heat Capacity	0.402	0.438			0.435	0.435					
Conductivity	0.119	0.031			0.031	0.017					
Viscosity	0.060	0.022			0.022	0.015					
Cp/Cv	1.323	1.410			1.408	1.348					
Z Factor	1.005	0.955			0.957	0.970					
MW	20.28	19.04			19.07	18.03					
Liquid Phase Properties											
Flowing Volume	gm		127.5	468.0		108.3		66.1	30.9	27.5	3.4
Mass Flow	54,423.5	201,844.0	54,423.5	201,844.0		48,948.8		33,801.8	15,147.0	13,464.0	1,683.0
Density	53.22	53.66	53.22	53.66		56.33		61.01	61.02	60.98	61.37
Heat Capacity	1.07	1.08	1.07	1.08		1.03		0.99	0.99	0.99	1.00
Conductivity	0.37	0.37	0.37	0.37		0.40		0.37	0.37	0.37	0.37
Viscosity	0.12	0.11	0.12	0.11		0.17		0.48	0.48	0.48	0.45
Surface Tension	36	37	36	37		47		66	66	66	66
SOLID PROPERTIES											
Mass Flow	1,761.3		1,683.0			1,683.0			1,683.0		1,683.0
Enthalpy	-5,926		-6,554			-6,536			-6,536		-6,536
V&L S											
TOTAL Mass Flow	140,432.3	286,864.0	56,106.5	201,844.0	281,358.0	5,474.7	50,531.8	33,891.8	16,830.0	13,464.0	3,363.0
TOTAL Enthalpy Flow	-98,843	-1,054,600	-357,634	-1,309,700	-1,023,800	-30,981	-328,652	-227,350	-108,408	-90,491	-17,918

Note 1: Electrolytes are the sum of the ionic species and dissolved solids in the water phase.

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFID Stream Number	112	113	114	115	201	202	203	204	205	206	207
DESCRIPTION	HP Gray Water to Gray Water Preheater	Blowdown Water to Offsite Treating	Makeup Water to Gray Water Tank	Caustic Addition to Quench Water	Bypass Syngas to Second Shift Reactor	Steam Addition for First Shift Reactor	Syngas/Steam to Shift Feed/Effluent Exchanger	Syngas/Steam to First Shift Reactor		Total Syngas to Second Shift Reactor	
V&L MIXTURE COMPONENTS											
H ₂	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	
CO ₂	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
CH ₄	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
H ₂ S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
COS	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
NH ₃	0.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
HCN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
HCL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
AR	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
N ₂	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
PRO	74,789.8	31,983.3	59,940.0	55.2	43,675.4	50.98	335.5	335.5		671.0	
ELECTROLYTES (Note 1)											
Water	137.0	295.0	0.0	138.8	20.00	0.0	0.1	0.0		0.0	
TOTAL	74,927.0	32,278.8	59,940.0	694.0	85,673.5	100.00	15,950.6	101,624.1		187,298.0	
V&L MIXTURE PROPERTIES											
Vapor Fraction	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	
Temperature	63.5	140.6	60.0	100.0	401.2	486.3	418.8	464.0		474.0	
Pressure	600.0	50.0	50.0	600.0	475.0	580.0	475.0	465.0		435.0	
Mass Flow	74,927.0	32,278.8	59,940.0	694.0	85,673.5	15,950.6	101,624.0	101,624.0		187,298.0	
Molecular Weight	18.03	16.10	16.02	18.36	19.00	18.02	18.90	18.90		18.97	
Enthalpy	-509,480	-217,050	-408,920	-4,475	-311,730	-89,897	-401,630	-399,410		-728,550	
Density	62.116	61.002	62.233	75.639	1.056	1.388	1.045	0.960		0.877	
Vapor Phase Properties											
Std. Volume Flow											
Flowing Volume											
Mass Flow											
Density											
Heat Capacity											
Conductivity											
Viscosity											
Cap/Cov											
Z-Factor											
MW											
Liquid Phase Properties											
Flowing Volume											
Mass Flow											
Density											
Heat Capacity											
Conductivity											
Viscosity											
Surface Tension											
SOLID PROPERTIES											
Mass Flow											
Enthalpy											
V&L &S											
TOTAL Mass Flow	74,927.0	32,278.8	59,940.0	694.0	85,673.5	15,950.6	101,624.0	101,624.0		187,298.0	
TOTAL Enthalpy flow	-509,480	-217,050	-408,920	-4,475	-311,730	-89,897	-401,630	-399,410		-728,550	

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PPD Stream Number	208	209	210	211	212	213	214	215	216	217	218
DESCRIPTION	Bypass Syngas to Shift Cooling	Shifted Syngas from Second MP Steam Generator	Shifted Syngas from Shift Condensate Heater	Shifted Syngas from Gray Water Heater	Shifted Syngas from Steam Generator	Shifted Syngas from LP Water Heater KO Drum	Shift Condensate from Water Heater KO Drum	Shifted Syngas from Airfan Cooler KO Drum	Shift Condensate from Airfan Cooler KO Drum	Shifted Syngas from Trim Cooler KO Drum	Shift Condensate from Trim Cooler KO Drum
V&L MIXTURE COMPONENTS											
H ₂	1637.9	7,372.2	3.94	2,020.3	3.03	9,008.5	4.14	9,007.9	0.00	9,007.9	0.00
CO	49,626.9	45.11	10,317.1	5.51	13,440.8	20.16	46,503.2	27.56	7.0	59,933.3	0.00
CO ₂	1,245.7	1.13	107,267.0	57.27	24,330.7	36.50	84,178.1	161.7	0.00	108,175.0	56.71
CH ₄	41.2	0.04	64.2	0.03	23.6	0.04	81.8	0.05	0.00	105.4	0.06
H ₂ S	847.0	0.50	909.3	0.49	326.5	0.49	1,129.5	0.67	7.2	1,441.2	0.00
COS	66.9	0.06	3.2	0.00	15.7	0.02	54.4	0.03	0.00	70.1	0.04
NH ₃	5.7	0.01	14.8	0.01	4.4	0.01	14.4	0.00	0.00	0.00	0.00
HCN	6.1	0.01	0.0	0.00	1.4	0.00	4.7	0.00	0.00	0.00	0.00
HCL	0.0	0.00	0.0	0.00	0.0	0.00	0.0	0.00	0.00	0.00	0.00
AR	320.4	0.29	499.0	0.27	183.7	0.28	635.7	0.38	0.1	819.2	0.45
N ₂	430.3	0.39	671.0	0.36	247.1	0.37	854.8	0.51	0.1	1,101.7	0.61
H ₂ O	56,092.4	50.98	60,180.0	32.13	26,068.6	39.10	90,192.3	16.84	79,633.1	34,279.0	99.25
ELECTROLYTES (Note 1)	0.0	0.00	0.0	0.00	0.7	0.00	6.8	0.00	0.00	39.5	0.00
Sulfur	0.0	0.00	0.0	0.00	0.0	0.00	0.0	0.00	0.00	0.00	0.00
TOTAL	110,011.0	100.00	187,297.7	100.00	66,663.4	100.00	230,645.1	100.00	79,852.3	181,137.2	100.00
V&L MIXTURE PROPERTIES											
Vapor Friction	1.0	1.0	1.0	0.945769000	0.8741	1.0	0.0	1.0	0.0	1.0	0.0
Temperature	401.2	550.0	410.0	356.4	345.0	299.9	354.0	160.0	180.0	100.0	100.0
Pressure	475.0	415.0	405.0	385.0	385.0	340.0	384.0	373.0	373.0	362.0	362.0
Mass Flow	110,011.0	187,298.0	187,298.0	66,663.4	230,645.0	217,456.0	79,852.3	182,992.0	34,464.2	181,138.0	1,854.5
Molecular Weight	19.07	19.07	19.07	19.07	19.07	19.39	18.04	19.66	18.07	19.68	18.09
Enthalpy	-400,280	-741,240	-753,300	-262,980	-924,780	-712,330	-526,380	-526,700	-231,810	-520,280	-12,571
Density	1.056	0.762	0.875	0.966	1.053	0.959	0.7221	1.151	60.762	1.243	61.501
Vapor Phase Properties											
Std. Volume Flow	52.5	89.9	89.9	30.2	96.6	102.1	84.8	84.8	70.7	83.8	3.8
Flowing Volume	28.0	86.2	86.2	19.2	60.7	63.0	44.2	44.2	34.464.2	40.5	1.854.5
Mass Flow	110,011.0	187,298.0	187,298.0	66,663.4	230,645.0	217,456.0	79,852.3	182,992.0	34,464.2	181,138.0	1,854.5
Density	1.056	0.762	0.875	0.966	1.053	0.959	0.7221	1.151	60.762	1.243	61.501
Heat Capacity	0.435	0.465	0.458	0.440	0.445	0.418	0.394	0.394	0.390	0.390	0.390
Viscosity	0.031	0.052	0.054	0.041	0.042	0.046	0.045	0.045	0.045	0.045	0.045
Conductivity	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001
CapCo	1.408	1.852	1.871	1.371	1.369	1.385	1.385	1.385	1.385	1.385	1.385
Z-Factor	0.957	0.987	0.973	0.973	0.973	0.973	0.973	0.973	0.973	0.973	0.973
MW	19.07	18.97	18.97	19.07	19.15	19.39	19.39	19.66	19.68	19.68	19.68
Liquid Phase Properties											
Flowing Volume	7.8	3,441.7	3,441.7	7.8	61.6	63.0	79,852.3	79,852.3	70.7	83.8	3.8
Mass Flow	110,011.0	187,298.0	187,298.0	66,663.4	230,645.0	217,456.0	79,852.3	182,992.0	34,464.2	181,138.0	1,854.5
Density	1.056	0.762	0.875	0.966	1.053	0.959	0.7221	1.151	60.762	1.243	61.501
Heat Capacity	0.435	0.465	0.458	0.440	0.445	0.418	0.394	0.394	0.390	0.390	0.390
Viscosity	0.031	0.052	0.054	0.041	0.042	0.046	0.045	0.045	0.045	0.045	0.045
Conductivity	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001
CapCo	1.408	1.852	1.871	1.371	1.369	1.385	1.385	1.385	1.385	1.385	1.385
Z-Factor	0.957	0.987	0.973	0.973	0.973	0.973	0.973	0.973	0.973	0.973	0.973
MW	19.07	18.97	18.97	19.07	19.15	19.39	19.39	19.66	19.68	19.68	19.68
SOLID PROPERTIES											
Mass Flow	110,011.0	187,298.0	187,298.0	66,663.4	230,645.0	217,456.0	79,852.3	182,992.0	34,464.2	181,138.0	1,854.5
Enthalpy	-400,280	-741,240	-753,300	-262,980	-924,780	-712,330	-526,380	-526,700	-231,810	-520,280	-12,571
V&L&S											
TOTAL Mass Flow	110,011.0	187,298.0	187,298.0	66,663.4	230,645.0	217,456.0	79,852.3	182,992.0	34,464.2	181,138.0	1,854.5
TOTAL Enthalpy flow	-400,280	-741,240	-753,300	-262,980	-924,780	-712,330	-526,380	-526,700	-231,810	-520,280	-12,571

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFD Stream Number	219	220	221	222	223	224	225	226	301	302	303
DESCRIPTION	Shifted Syngas from Mercury Removal Drum	Hot Gray Water to Syngas Scrubber	Hot Shift Condensate to Syngas Scrubber	Sour Gas to Ammonia Stripper Overhead Condenser	Shift Condensate to Ammonia Stripper	Sour Gas to Sulfur Recovery System	Striped Condensate to HP Shift Condensate Drum	Mixed Bypass & Shifted Syngas	Sweet Syngas to Syngas Compression	CO2 to CO2 Compression	Acid Gas to SRU
VAL MIXTURE COMPONENTS											
H2	9,007.9 lb/hr	0.0 lb/hr	1.5 lb/hr	0.6 lb/hr	0.0 lb/hr	1.2 lb/hr	0.50 lb/hr	9,010.0 lb/hr	8,993.3 lb/hr	0.2 lb/hr	2.1 lb/hr
CO	59,833.0 lb/hr	0.0 lb/hr	7.0 lb/hr	3.7 lb/hr	4.93 lb/hr	12.8 lb/hr	5.58 lb/hr	59,844.0 lb/hr	59,517.7 lb/hr	32.3 lb/hr	4.4 lb/hr
CO2	108,162.0 lb/hr	0.0 lb/hr	173.0 lb/hr	0.14 lb/hr	89.17 lb/hr	172.6 lb/hr	75.01 lb/hr	108,513.0 lb/hr	107,951.4 lb/hr	561.6 lb/hr	55.3 lb/hr
CH4	105.4 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.01 lb/hr	0.0 lb/hr	0.01 lb/hr	105.4 lb/hr	101.6 lb/hr	3.8 lb/hr	3.4 lb/hr
H2S	1,440.7 lb/hr	0.0 lb/hr	7.6 lb/hr	0.01 lb/hr	1.3 lb/hr	11.0 lb/hr	4.77 lb/hr	1,456.3 lb/hr	1,428.0 lb/hr	28.3 lb/hr	2.7 lb/hr
COS	70.1 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.03 lb/hr	0.0 lb/hr	0.02 lb/hr	70.1 lb/hr	68.0 lb/hr	2.1 lb/hr	2.1 lb/hr
NH3	0.0 lb/hr	0.0 lb/hr	9.7 lb/hr	0.0 lb/hr	0.00 lb/hr	7.3 lb/hr	0.00 lb/hr	20.5 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr
HCN	4.4 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.01 lb/hr	1.0 lb/hr	0.00 lb/hr	6.1 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr
HCL	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr
AR	819.2 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.00 lb/hr	819.4 lb/hr	819.2 lb/hr	0.2 lb/hr	0.2 lb/hr
N2	1,101.7 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.00 lb/hr	1,101.8 lb/hr	1,100.3 lb/hr	1.55 lb/hr	0.3 lb/hr
H2O	492.9 lb/hr	74,786.7 lb/hr	99.82 lb/hr	2.4 lb/hr	3.18 lb/hr	38,117.2 lb/hr	10.37 lb/hr	116,282.0 lb/hr	116,282.0 lb/hr	0.0 lb/hr	0.0 lb/hr
ELECTROLYTES (Note 1)	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr
Sulfur	0.0 lb/hr	136.9 lb/hr	0.0 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.00 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr	0.0 lb/hr
TOTAL	181,137.2 lb/hr	74,927.0 lb/hr	121,411.5 lb/hr	75.5 lb/hr	100.00 lb/hr	36,243.2 lb/hr	100.00 lb/hr	297,308.5 lb/hr	297,308.5 lb/hr	105,145.1 lb/hr	4,503.1 lb/hr
VAL MIXTURE PROPERTIES											
Vapor Fraction	1.0	0.0	0.0	1.0	0.0	1.0	0.0	1.0	1.0	1.0	1.0
Temperature	105.7	346.0	394.7	157.3	157.5	200.0	286.2	407.1	80.0	65.0	80.0
Pressure	342.0	580.0	580.0	100.0	100.0	45.0	400.0	405.0	300.0	40.0	45.0
Mass Flow	181,138.0	74,927.0	121,411.0	75.5	36,243.2	230.1	41,563.3	297,309.0	70,888.8	105,147.0	4,505.1
Molecular Weight	19.86	18.03	18.03	35.01	18.05	32.51	18.02	19.01	10.64	43.86	38.89
Enthalpy	-519,850	-490,400	-785,580	-0.278	-244,090	-0.825	-274,470	-1,153,600	-102,590	-404,050	-12,244
Density	1.165	55,804	53,883	0.400	60,922	0.277	57,499	0.884	0.576	0.436	0.421
Vapor Phase Properties											
Std. Volume Flow	83.8			0.0		0.1		142.4	60.7	21.8	1.1
Flowing Volume	43.2			0.1		0.2		93.4	34.2	67.0	3.0
Mass Flow	181,138.0			75.5		230.1		297,309.0	70,888.8	105,147.0	4,505.1
Density	1.166			0.400		0.277		0.884	0.576	0.436	0.421
Heat Capacity	0.353			0.253		0.275		0.450	0.652	0.205	0.221
Conductivity	0.036			0.019		0.018		0.050	0.050	0.011	0.011
Viscosity	0.014			0.017		0.017		0.018	0.011	0.014	0.014
CP/CV	1.405			1.310		1.304		1.377	1.428	1.318	1.328
Z-Factor	0.985			0.988		0.988		0.970	1.004	0.978	0.976
NW	19.86			35.01		32.51		19.01	10.64	43.86	38.89
Liquid Phase Properties											
Flowing Volume	167.4			74.2		0.1		93.4	34.2	67.0	3.0
Mass Flow	74,927.0			36,243.2		230.1		297,309.0	70,888.8	105,147.0	4,505.1
Density	55.80			60.92		57.50		0.884	0.576	0.436	0.421
Heat Capacity	1.04			1.00		1.03		0.450	0.652	0.205	0.221
Conductivity	0.36			0.35		0.40		0.050	0.050	0.011	0.011
Viscosity	0.14			0.39		0.17		1.377	1.428	1.318	1.328
Surface Tension	43			64		49		0.970	1.004	0.978	0.976
SOLID PROPERTIES											
Mass Flow											
Enthalpy											
MM Bu/hr											
VAL&S											
TOTAL Mass Flow	181,138.0	74,927.0	121,411.0	75.5	36,243.2	230.1	41,563.3	297,309.0	70,888.8	105,147.0	4,505.1
TOTAL Enthalpy flow	-519,850	-490,400	-785,580	-0.278	-244,090	-0.825	-274,470	-1,153,600	-102,590	-404,050	-12,244

EnviroRes Coal/Biomass Gasification Project - Syngas Treating Units

Heat & Material Balance (Mass Flows)

Case 2 - Syngas to Fischer-Tropsch - Revision 0 (Aug 28, 2009)

PFD Stream Number	304	401	402	501	502	601	701	702
DESCRIPTION	AGR Waste Water to Offsite Treating	Hydrogen Product	Fuel Gas to Offsite	Sulfur Product	SRU Waste Water to Offsite Treating	CO2 Product	Treated Syngas to Fischer-Tropsch	Sweet Syngas to Hydrogen Purification
V&L MIXTURE COMPONENTS								
H2	0.0	0.0	177.0	0.0	0.0	0.0	0.0	0.0
CO	0.0	0.0	3,344.5	0.0	0.0	0.0	0.0	0.0
CO2	0.0	0.0	19.7	0.55	0.0	0.0	0.0	0.0
CH4	0.0	0.0	0.3	0.07	5.4	0.15	0.0	0.0
H2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
COS	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HCN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HCL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
AR	0.0	0.0	33.6	873	12.5	0.35	0.0	0.0
WGO	0.0	0.0	21.9	570	39.9	1.11	0.0	0.0
ELECTROLYTES (Note 1)	492.9	99.11	0.0	0.0	0.0	0.0	0.0	0.0
Sulfur	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL	497.3	100.00	3,399.0	1,399.7	100.00	105,146.1	66,905.3	100.00
V&L MIXTURE PROPERTIES								
Vapor Fraction	0.0	1.0	1.0	0.0	0.0	0.0	1.0	1.0
Temperature	110.0	120.0	110.0	277.2	100.1	110.0	400.0	110.0
Pressure	50.0	475.0	50.0	10.3	35.0	2,200.0	475.0	485.0
Mass Flow	497.3	384.5	3,399.0	1,399.7	674.7	105,147.0	66,905.3	3,983.5
Molecular Weight	18.07	2.33	17.16	33.97	18.02	43.86	10.64	10.64
Enthalpy	-3,357	0.047	-5,719	0.070	-4,619	-413,000	-82,765	-5,690
Density	51.787	0.182	0.182	291.730	61.215	44.457	0.558	0.862
Vapor Phase Properties								
Std. Volume Flow	1.5	1.9	1.9					
Flowing Volume	0.6	5.5	5.5					
Mass Flow	384.5	3,399.0	3,399.0					
Density	0.182	0.182	0.182					
Heat Capacity	2.935	2.935	2.935					
Conductivity	0.088	0.088	0.088					
Viscosity	0.009	0.009	0.009					
Cp/Cv	1.423	1.009	1.009					
Z-Factor	1.011	1.009	1.009					
MW	2.33	17.16	17.16					
Liquid Phase Properties								
Flowing Volume	1.2			0.6	1.4	284.9		
Mass Flow	497.3			1,399.7	674.7	105,147.0		
Density	51.79			281.73	61.21	44.46		
Heat Capacity	1.07			0.20	1.06	0.69		
Conductivity	0.78					0.05		
Viscosity	0.92					0.07		
Surface Tension	89					0		
SOLID PROPERTIES								
Mass Flow								
Enthalpy								
V&L &S								
TOTAL Mass Flow	497.3	384.5	3,399.0	1,399.7	674.7	105,147.0	66,905.3	3,983.5
TOTAL Enthalpy flow	-3,357	0.047	-5,719	0.070	-4,619	-413,000	-82,765	-5,690

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendix 4

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC
FINAL REPORT - SYNGAS TREATMENT STUDY
FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendix 5

WorleyParsons

resources & energy

ICRC-Envires Case 1 CAPEX Estimate

Equipment Number	Equipment Name	Quantity	Equipment Cost (Each)	Total Equipment Cost	Factor	Bulk Material Cost	Installation Labor Cost	Total Installed Cost
Unit 21 - Syngas Quench & Scrubbing				5,414,862		1,723,796	3,447,598	10,586,256
E-2101 A/B	Flash Bottoms Coolers	2						
Y-2101	Venturi Mixer	1						
Y-2102 A/B	Solids Filters	2						
P-2101 A/B	Syngas Scrubber Pumps	2						
P-2102 A/B	Settler Bottoms Pumps	2						
P-2103 A/B	HP Gray Water Pumps	2						
P-2104 A/B	Blowdown Water Pumps	2						
P-2105 A/B	Filtrate Pumps	2						
V-2101	Quench Drum	1						
V-2102	Syngas Scrubber	1						
V-2103	MP Flash Drum	1						
V-2104	Filtrate Drum	1						
T-2101	Solids Settler	1						
T-2102	Gray Water Tank	1						
Unit 22 - Shift Reaction & Cooling				7,175,671		2,799,728	5,599,465	15,574,864
E-2201	Shift Startup Heater	1						
E-2202	Shift/Feed Effluent Exchanger	1						
E-2203	First MP Steam Generator	1						
E-2204	Second MP Steam Generator	1						
E-2205	Shift Condensate Heater	1						
E-2206	Gray Water Heater	1						
E-2207	LP Steam Generator	1						
E-2208	Gray Water Preheater	1						
E-2209	Utility Water Heater	1						
E-2210	Syngas Airfin Cooler	1						
E-2211	Syngas Trim Cooler	1						
E-2212	Syngas Heater	1						
E-2213	Ammonia Stripper Overhead Condenser	1						
E-2214	Ammonia Stripper Reboiler	1						
E-2215	First LP Steam Generator	1						
P-2201 A/B	HP Shift Condensate Pumps	2						
P-2202 A/B	Ammonia Stripper Feed Pumps	2						
P-2203 A/B	Ammonia Stripper Reflux Pumps	2						
P-2204 A/B	Ammonia Stripper Bottoms Pumps	2						
R-2201	First Shift Reactor	1						
CAT-2201	First Shift Reactor Catalyst	1						
R-2202	Second Shift Reactor	1						
CAT-2202	Second Shift Reactor Catalyst	1						
R-2203	Third Shift Reactor	1						
CAT-2203	Third Shift Reactor Catalyst	1						
V-2201	Shift Inlet KO Drum	1						
V-2202	HP Shift Condensate Drum	1						
V-2203	Water Heater KO Drum	1						
V-2204	Syngas Airfin Cooler KO Drum	1						
V-2205	Syngas Trim Cooler KO Drum	1						
V-2206	Mercury Removal Drum	1						
V-2206	Mercury Removal Bed	1						
V-2207	Ammonia Stripper Feed Drum	1						
V-2208	Ammonia Stripper	1						
V-2209	Ammonia Stripper Reflux Drum	1						
Unit 23 - Acid Gas Removal				37,594,790		0	0	37,594,790
Z-2301	Rectisol AGR Package	1						
Unit 24 - H2 Purification				6,958,000		896,499	1,793,001	9,647,500
Z-2401	PSA Package	1						
Z-2402	Fuel Gas Compressor Package	1						
Unit 25 - Sulfur Recovery				17,584,177		77,843	155,685	17,817,705
Z-2501	Claus Package	1						
Z-2502	Sulfur Degassing Package	1						
Z-2503	Tail Gas Treating Package	1						
Unit 26 - CO2 Compression				11,088,905		319,630	639,261	12,047,796
Z-2601	CO2 Compressor (Centrifugal)	1						
Z-2602	CO2 Thermal Oxidizer Package	1						
Miscellaneous BOP Costs				12,872,461		1,647,924	2,359,551	16,879,937
	Allowance for Electrical Equipment & Bunks	1						
	Allowance for Instrumentation & Control Equipment & Bunks	1						
	Allowance for Pipe Rack	1						
	Allowance for Site Prep, Roads & Gravel	1						
Engineering, Procurement, & Construction Management								14,420,000
	Engineering, Procurement & Construction Management	1						
Contingency								33,640,000
	Contingency	1						
TOTAL				98,688,866	1.7	7,465,421	13,994,561	168,208,847

WorleyParsons

resources & energy

ICRC-Envires Case 2 CAPEX Estimate

Equipment Number	Equipment Name	Quantity	Equipment Cost (Each)	Total Equipment Cost	Factor	Bulk Material Cost	Installation Labor Cost	Total Installed Cost
Unit 21 - Syngas Quench & Scrubbing				5,405,742		1,717,716	3,435,438	10,558,896
E-2151 A/B	Flash Bottoms Coolers	2						
Y-2151	Venturi Mixer	1						
Y-2152 A/B	Solids Filters	2						
P-2151 A/B	Syngas Scrubber Pumps	2						
P-2152 A/B	Settler Bottoms Pumps	2						
P-2153 A/B	HP Gray Water Pumps	2						
P-2154 A/B	Blowdown Water Pumps	2						
P-2155 A/B	Filtrate Pumps	2						
V-2151	Quench Drum	1						
V-2152	Syngas Scrubber	1						
V-2153	MP Flash Drum	1						
V-2154	Filtrate Drum	1						
T-2151	Solids Settler	1						
T-2152	Gray Water Tank	1						
Unit 22 - Shift Reaction & Cooling				5,033,619		2,224,105	4,448,216	11,705,939
E-2251	Shift Startup Heater	1						
E-2252	Shift/Feed Effluent Exchanger	1						
E-2253	First MP Steam Generator	1						
E-2254	Second MP Steam Generator	1						
E-2255	Shift Condensate Heater	1						
E-2256	Gray Water Heater	1						
E-2257	LP Steam Generator	1						
E-2258	Gray Water Preheater	1						
E-2259	Utility Water Heater	1						
E-2260	Syngas Airfin Cooler	1						
E-2261	Syngas Trim Cooler	1						
E-2262	Syngas Heater	1						
E-2263	Ammonia Stripper Overhead Condenser	1						
E-2264	Ammonia Stripper Reboiler	1						
P-2251 A/B	HP Shift Condensate Pumps	2						
P-2252 A/B	Ammonia Stripper Feed Pumps	2						
P-2253 A/B	Ammonia Stripper Reflux Pumps	2						
P-2254 A/B	Ammonia Stripper Bottoms Pumps	2						
R-2251	First Shift Reactor	1						
CAT-2251	First Shift Reactor Catalyst	1						
R-2252	Second Shift Reactor	1						
CAT-2252	Second Shift Reactor Catalyst	1						
V-2251	Shift Inlet KO Drum	1						
V-2252	HP Shift Condensate Drum	1						
V-2253	Water Heater KO Drum	1						
V-2254	Syngas Airfin Cooler KO Drum	1						
V-2255	Syngas Trim Cooler KO Drum	1						
V-2256	Mercury Removal Drum	1						
V-2256	Mercury Removal Bed	1						
V-2257	Ammonia Stripper Feed Drum	1						
V-2258	Ammonia Stripper	1						
V-2259	Ammonia Stripper Reflux Drum	1						
Unit 23 - Acid Gas Removal				33,354,079		0	0	33,354,079
Z-2351	Rectisol AGR Package	1						
Unit 24 - H2 Purification				3,220,000		363,333	726,667	4,310,000
Z-2451	PSA Package	1						
Z-2452	Fuel Gas Compressor Package	1						
Unit 25 - Sulfur Recovery				17,584,177		77,843	155,685	17,817,705
Z-2551	Claus Package	1						
Z-2552	Sulfur Degassing Package	1						
Z-2553	Tail Gas Treating Package	1						
Unit 26 - CO2 Compression				7,698,706		221,876	443,753	8,364,335
Z-2651	CO2 Compressor (Centrifugal)	1						
Z-2652	CO2 Thermal Oxidizer Package	1						
Unit 27 - Syngas Compression				1,927,666		690,306	1,380,615	3,998,587
E-2751	Sulfur Guard Heater	1						
Z-2752	PSA Feed Cooler	1						
V-2751	Sulfur Guard Vessel	1						
CAT-2751	Sulfur Guard Catalyst/Vessel	1						
Z-2751	Syngas Compressor Package	1						
Miscellaneous BOP Costs				11,133,598		1,569,577	2,202,856	14,906,031
	Allowance for Electrical Equipment & Bulks	1						
	Allowance for Instrumentation & Control Equipment & Bulks	1						
	Allowance for Pipe Rack	1						
	Allowance for Site Prep, Roads & Gravel	1						
Engineering, Procurement, & Construction Management								12,600,000
	Engineering, Procurement & Construction Management	1						
Contingency								29,400,000
	Contingency	1						
TOTAL				85,357,587	1.7	6,864,756	12,793,230	147,015,573

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendix 6

Operations Expense Summary

EXPENSE CATEGORIES	ANNUAL OPEX
Personnel Expenses	Expenses
Management/Finance/Office Staff (Note 1)	\$504,000
Technical/Engineering Staff (Note 1)	\$443,000
Maintenance/Warehouse Staff (Note 1)	\$60,000
Operations (Shift Staff)	\$3,106,000
Total Direct Hire	\$4,113,000
Contract Maintenance Labor (Note 2)	\$2,254,000
Total Personnel Expenses	\$6,367,000
Materials Expenses	
Raw Materials Expenses (Note 3)	\$0
Maintenance Material Expenses (Note 2)	\$1,110,000
Catalysts and Chemical Expenses	\$870,000
Utilities Expenses	\$8,239,000
Waste Disposal Expenses	\$907,000
Total Materials Expense	\$11,126,000
Miscellaneous Expenses	
Miscellaneous Supplies (Note 4)	\$111,000
Miscellaneous Contracts (Note 5)	\$445,000
Other (Note 6, 7)	\$0
Total Miscellaneous Expenses	\$556,000
Total OPERATING EXPENSES:	\$18,049,000
Notes:	
1. Syngas Treatment was allocated 33% of the cost for these personnel.	
2. Total Maintenance is 2% of Total Installed Cost, sub-divided into 67% labor and 33% materials.	
3. The cost of feedstock to the gasifier (i.e. coal, biomass, and oxygen) as well as cost for Raw Syngas from the gasifier have been excluded from the OPEX.	
4. Miscellaneous Supplies include office supplies, equipment, laboratory and operating supplies (non repair) allocated to Syngas Treatment.	
5. Contract analytical testing allocated to Syngas Treatment.	
6. Taxes, interest on loans, insurance, permits, corporate overhead have been excluded.	
7. Licensor fees have been included in the CAPEX.

Operations Expense Summary

EXPENSE CATEGORIES	ANNUAL OPEX
Personnel Expenses	Expenses
Management/Finance/Office Staff (Note 1)	\$504,000
Technical/Engineering Staff (Note 1)	\$443,000
Maintenance/Warehouse Staff (Note 1)	\$60,000
Operations (Shift Staff)	\$3,106,000
Total Direct Hire	\$4,113,000
Contract Maintenance Labor (Note 2)	\$1,970,000
Total Personnel Expenses	\$6,083,000
Materials Expenses	
Raw Materials Expenses (Note 3)	\$0
Maintenance Material Expenses (Note 2)	\$970,000
Catalysts and Chemical Expenses	\$634,000
Utilities Expenses	\$6,483,000
Waste Disposal Expenses	\$906,000
Total Materials Expense	\$8,993,000
Miscellaneous Expenses	
Miscellaneous Supplies (Note 4)	\$90,000
Miscellaneous Contracts (Note 5)	\$360,000
Other (Note 6, 7)	\$0
Total Miscellaneous Expenses	\$450,000
Total OPERATING EXPENSES:	\$15,526,000
Notes:	
1. Syngas Treatment was allocated 33% of the cost for these personnel.	
2. Total Maintenance is 2% of Total Installed Cost, sub-divided into 67% labor and 33% materials.	
3. The cost of feedstock to the gasifier (i.e. coal, biomass, and oxygen) as well as cost for Raw Syngas from the gasifier have been excluded from the OPEX.	
4. Miscellaneous Supplies include office supplies, equipment, laboratory and operating supplies (non repair) allocated to Syngas Treatment.	
5. Contract analytical testing allocated to Syngas Treatment.	
6. Taxes, interest on loans, insurance, permits, corporate overhead have been excluded.	
7. Licensors fees have been included in the CAPEX.

WorleyParsons

resources & energy

EcoNomics™

ENVIRES LLC

FINAL REPORT - SYNGAS TREATMENT STUDY

FOR ENVIRES COAL/BIOMASS GASIFICATION PROJECT

Appendix 7

WorleyParsons
resources & energy

**EnviRes Coal/Biomass Gasification Project
Syngas Treatment Study
Case 1 and 2
Utility Summary by Unit**

By: JCM
Check: JLH
App: JW
Date: 27-Oct-09
Rev: 0

Utility	Case 1	Case 2	Units
MP Steam			
Consumed			
Unit 21	0	0	lb/hr
Unit 22	26,210	15,951	lb/hr
Unit 23	0	0	lb/hr
Unit 24	0	0	lb/hr
Unit 25	0	0	lb/hr
Unit 26	0	0	lb/hr
Unit 27	NA	12,739	lb/hr
Total	26,210	28,690	lb/hr
Produced			
Unit 21	0	0	lb/hr
Unit 22	46,655	28,417	lb/hr
Unit 23	0	0	lb/hr
Unit 24	0	0	lb/hr
Unit 25	2,700	2,700	lb/hr
Unit 26	0	0	lb/hr
Unit 27	NA	0	lb/hr
Total	49,355	31,117	lb/hr
Net Exported (Imported)	23,145	2,427	lb/hr
LP Steam			
Consumed			
Unit 21	0	0	lb/hr
Unit 22	7,242	7,197	lb/hr
Unit 23	21,488	17,568	lb/hr
Unit 24	0	0	lb/hr
Unit 25	0	0	lb/hr
Unit 26	0	0	lb/hr
Unit 27	NA	0	lb/hr
Total	28,730	24,765	lb/hr
Produced			lb/hr
Unit 21	0	0	lb/hr
Unit 22	20,786	30,902	lb/hr
Unit 23	0	0	lb/hr
Unit 24	0	0	lb/hr
Unit 25	0	0	lb/hr
Unit 26	0	0	lb/hr
Unit 27	NA	0	lb/hr
Total	20,786	30,902	lb/hr
Net Exported (Imported)	(7,944)	6,137	lb/hr

WorleyParsons
resources & energy

**EnviRes Coal/Biomass Gasification Project
Syngas Treatment Study
Case 1 and 2
Utility Summary by Unit**

By: JCM
Check: JLH
App: JW
Date: 27-Oct-09
Rev: 0

Utility	Case 1	Case 2	Units
Electric Power (Operating)			
Unit 21	222	193	HP
Unit 22	301	266	HP
Unit 23	6,234	5,106	HP
Unit 24	777	194	HP
Unit 25	6	6	HP
Unit 26	9,443	5,150	HP
Unit 27	NA	2,211	HP
Total	16,983	13,125	HP
Total	12,664	9,787	kW
Cooling Water (Note 1)			
Unit 21	895	895	gpm
Unit 22	793	403	gpm
Unit 23	5,795	4,746	gpm
Unit 24	179	46	gpm
Unit 25	0	0	gpm
Unit 26	4,089	2,215	gpm
Unit 27	NA	585	gpm
Total	11,750	8,890	gpm
Boiler Feed Water			
Unit 21	0	0	lb/hr
Unit 22	67,441	59,319	lb/hr
Unit 23	1,443	1,182	lb/hr
Unit 24	0	0	lb/hr
Unit 25	2,700	2,700	lb/hr
Unit 26	0	0	lb/hr
Unit 27	NA	0	lb/hr
Total	71,584	63,201	lb/hr
Demin Water			
Unit 21	86,370	59,940	lb/hr
Unit 22	0	0	lb/hr
Unit 23	0	0	lb/hr
Unit 24	0	0	lb/hr
Unit 25	0	0	lb/hr
Unit 26	0	0	lb/hr
Unit 27	NA	0	lb/hr
Total	86,370	59,940	lb/hr
Condensate Return			
Unit 21	0	0	lb/hr
Unit 22	7,242	7,197	lb/hr
Unit 23	21,488	17,568	lb/hr
Unit 24	0	0	lb/hr
Unit 25	0	0	lb/hr
Unit 26	0	0	lb/hr
Unit 27	NA	12,739	lb/hr
Total	28,730	37,504	lb/hr

**EnviRes Coal/Biomass Gasification Project
Syngas Treatment Study
Case 1 and 2
Utility Summary by Unit**

By: JCM
Check: JLH
App: JW
Date: 27-Oct-09
Rev: 0

Utility	Case 1	Case 2	Units
Utility Water for cooling / heat Recovery (Note 2)			
Unit 21	0	0	gpm
Unit 22	400	400	gpm
Unit 23	0	0	gpm
Unit 24	0	0	gpm
Unit 25	0	0	gpm
Unit 26	0	0	gpm
Unit 27	NA	0	gpm
Total	400	400	gpm
Fuel Gas			
Consumed			
Unit 21	0.0	0.0	MM Btu/hr
Unit 22	0.0	0.0	MM Btu/hr
Unit 23	0.0	0.0	MM Btu/hr
Unit 24	0.0	0.0	MM Btu/hr
Unit 25	2.4	2.4	MM Btu/hr
Unit 26	0.0	0.0	MM Btu/hr
Unit 27	NA	0.0	MM Btu/hr
Total	2.4	2.4	MM Btu/hr
Produced			
Unit 21	0.0	0.0	MM Btu/hr
Unit 22	0.0	0.0	MM Btu/hr
Unit 23	0.0	0.0	MM Btu/hr
Unit 24	103.9	25.5	MM Btu/hr
Unit 25	0.0	0.0	MM Btu/hr
Unit 26	0.0	0.0	MM Btu/hr
Unit 27	NA	0.0	MM Btu/hr
Total	103.9	25.5	MM Btu/hr
Net Exported	101.5	23.1	MM Btu/hr
<p>Note 1: Cooling water flows are based on a 20°F temperature rise.</p> <p>Note 2: Case 1 has 22 MM Btu/hr and Case 2 has 36 MMBtu/hr of Low Level Heat @ 300°F available. For heat exchanger sizing this heat has been used to heat 400 gpm of Utility Water. This heat is available for other services when the Balance of Plant systems are more defined.</p>			

APPENDIX 6
Hydrogen Recovery from Coke Oven Gas

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

TECHNICAL STUDY DOCUMENT PRESSURE SWING ADSORPTION UNIT PSA

CUSTOMER:	AKER KVAERNER FOR ENVIRES
PROJECT:	SYNTHETIC AMMONIA FROM COKE OVEN GAS AND COAL PRESSURE SWING ADSORPTION UNIT.
PROJECT No.:	J10556
PROCESS ENGINEER:	MR C. GODWIN
SALES MANAGER:	MR C. ROPER
ISSUING OFFICE:	NATCO UK LIMITED

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

TABLE OF CONTENTS

1.0	SUMMARY	4
2.0	INTRODUCTION	4
3.0	COKE OVEN GAS EXPERIENCE	5
4.0	DESIGN BASIS	7
4.1	Feedstock	7
4.2	Product	8
4.3	Offgas	8
4.4	Utility consumption	9
4.5	Study Assumptions	9
4.6	Exceptions	9
5.0	SCOPE OF STUDY	10
6.0	PROCESS DESCRIPTION	11
6.1	PSA Unit	12
6.2	PSA Technology	14
6.3	Coke Oven Gas Compressor	15
6.4	Intermediate Storage Tank	15
7.0	GENERAL COMMENTS	16
8.0	SCOPE OF SUPPLY	17
8.1	PSA Unit	18
8.2	General	20
8.3	Budget cost	21
9.0	DATA SHEETS AND DRAWINGS	22
9.1	Material Balance	22

<i>NATCO</i>		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIREC	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

10.0 STUDY CONCLUSIONS.....23

APPENDICES

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

1.0 SUMMARY

The following study was commissioned by EnviRes/CTL Ashland and Aker Kvaerner Technical Services (Purchase order No. A0755501-R001-0001). This study has been carried out on the proposal to recover Hydrogen from a Coke Oven Gas (COG) stream using Pressure Swing Adsorption (PSA) technology.

The scope of the study is detailed in section 5.0 of this report.

2.0 INTRODUCTION

Natco UK Limited have been commissioned by EnviRes/CTL Ashland and Aker Kvaerner Technical services to carry out a feasibility and verification study on the recovery of Hydrogen from Coke Oven Gas (COG) at an existing AK STEEL process plant in Ashland Kentucky. The COG Conditioning and PSA plant are part of a wider process design being undertaken by Aker Kvaerner, in order to produce synthesised Ammonia from Coke Oven Gas (COG).

The study has been commissioned to assess the options for the PSA purification unit prior to further upstream processing for Ammonia synthesis and refining. The study has reviewed the compression of COG, Pre-purification, PSA purification, Hydrogen pre-treatment, and has reported on its findings.

The submission of Heat and Material balances, Process Flow Diagrams (PFD's), Process Instrument Diagrams (P&ID's), Plot Plan and preliminary Datasheets for vessels and rotating machines were agreed with EnviRes/CTL Ashland as the deliverables for this study.

Included with this design study is the budgetary cost for plant engineering, design and equipment supply for Pre-purification/COG conditioning, PSA Plant, and Hydrogen Pre-treatment.

Indicative prices for the COG compressor have also been identified.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

3.0 COKE OVEN GAS EXPERIENCE

For many years, Natco (as Howmar) licensed its PSA technology to a medium sized domestic Japanese engineering contractor, Cosmo Engineering Corporation (CEC).

They had a particular interest and specialty in supplying hydrogen generating plants to the local market, mainly based upon the steam reforming of methanol under license from Mitsubishi Gas Chemical for the crude hydrogen production (or front end) section.

Pressure swing adsorption was used to purify the hydrogen from the crude hydrogen and Natco technology was used exclusively for this purpose.

However, CEC also supplied 'stand alone' PSA's also using Natco technology and one such application was for the purification of coke oven gas for Nippon Sanso (translates as Japanese Oxygen) who are a domestic industrial gas supplier.

The plant was installed at a steel works, Kokan Centre, in the district of Hiroshima, Japan.

It had a product capacity of 1,000 Nm³/hr of pure hydrogen and operated at a pressure of 9 Bar(a).

The flowscheme was configured as follows

- The coke oven gas was compressed to operating pressure in a two stage oil injected screw compressor
- Two sacrificial activated carbon beds, one on line and one standby, principally to remove naphthalene with each sized for 6 months operation.
- PSA to purify the hydrogen
- De-oxo unit to remove oxygen from the product

The plant was successfully commissioned and performance tested on 28th - 29th December 1985, and the following performance was achieved.

J1303 PSA for Nippon Sanso/Kokan Sanso centre operation data.

1. Feedstock

	Design	Actual
H ₂	53.45	54.5
N ₂	3.76	2.80
CH ₄	28.51	29.30
O ₂	0.3	0.21
CmHn	5.4	3.74
CO ₂	2.19	2.49
CO	6.29	6.96
Water	Sat	Sat
Total	100 vol %	100 vol %
Temperature	40°C	46.2°C

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

Pressure (Inlet)	9.4 kg/cm ² (g)	9.0 kg/cm ² (g)
(Outlet)	-	8.9 kg/cm ² (g)
Feed Gas Flow Rate	2,711 Nm ³ /hr	2,702 Nm ³ /hr

Note: The flow rate, instruments values have been corrected for temperature and pressure.

2. Product Hydrogen

	Design	Actual
Flow rate	1,000 Nm ³ /h	1,032 Nm ³ /h
Hydrogen recovery	69%	70%
'RUN' cycle time	4min	5 min and 15 sec to 5 min and 40 sec
Repress Gas flow	1,175 Nm ³ /hr	810 Nm ³ /hr
Impurity in the Product Hydrogen (Guarantee)		
	Design	Actual
CH ₄	<1.0 ppmV	No Peak
CO ₂	<1.0 ppmV	No Peak
CO	<1.0ppmV	No Peak
O ₂	<500 ppmV	3.4 ppmV
N ₂ +Ar	<9.0 ppmV Expected	6.6 ppmV
NH ₃	-	0.0
Purity of Hydrogen product		
	Design	Actual
Vol %	99.95	99.999

Note that we are uncertain of the current situation with the plant but CEC have advised us that they believe it to be shut down.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

4.0 DESIGN BASIS

The Study is to examine the PSA technical feasibility and capacity using the following feed stock as defined by Aker Kvaerner Design Basis B-10 Rev A dated 15/04/07, amended by AK e-mail dated 25/09/2007 with attached Excel file 'COG Feed Analysis' and acknowledge/clarified by Natco UK Limited dated 26/09/07.

4.1 Feedstock

Description	Units	Design	Comment
Hydrogen	Volume %, dry	56.31	
Carbon Monoxide	Volume %, dry	5.06	
Carbon Dioxide	Volume %, dry	1.26	Some removal expected with desulfurization process
Oxygen	Volume %,dry	0.78	Max. excursion observed- Dan Howell 14SEP07
Nitrogen	Volume %,dry	6.81	
Methane	Volume %,dry	23.34	
Ethane	Volume %,dry	0.39	
Ethylene	Volume %,dry	1.85	
Propane	Volume %,dry	0.1	
Propylene	Volume %,dry	0.29	
Benzene	Volume %,dry	0.35	Coke plant light oil removal could be improved.
Toluene	Volume %,dry	0.16	
Xylene	Volume %,dry	0.00	
Hydrogen Sulfide	ppm, dry basis	5	New acid gas system will be specified to meet this level to avoid problems in the PSA
Hydrogen Cyanide	ppm, dry	5	New acid gas system will be specified to meet this level to avoid problems in the PSA
Ammonia	Volume %,dry	0.04	50% removal experienced in existing COG Desulfurization- Dan Howell 14SEP07
Sulfur Dioxide	Volume %,dry	0.004	Acid gas removal systems are not designed to remove SO ₂ . Only partial removal expected.
Organic Sulfur COS,CS ₂	Volume %,dry	0.01	Only partial removal of organic sulfur expected in Acid Gas removal. Wilson and Wells, pg 268 estimates 20 to 30 grains/100 ft ³ in raw gas and 50% removal in Light Oil Scrubbers
Water vapor content of COG	Volume	3.21	Saturated at 35 C. Variation depends on total pressure at B/L.
Naphthalene content of COG	Volume %,dry	0.01	Based on inspection of data. 70% removal of Naphthalene in Light Oil Scrubbers and saturation at 35 C. 14SEP07
Particulate content of COG	mg/Nm ³ ,dry	5	Typical particulate loading for gas cleaning systems
HCl	Volume %	0.0001	Assumed. Coke plant has ammonia water scrubbers
NO _x	Volume %	0.0001	Assumed based on ' Coal, Coke and Coal Chemicals',
Temperature of COG at Light Oil Scrubber	Degrees C	35	
COG production	Million scfd	38	Based on variation of approx. +/- 20% as seen in data from AK Steel Coke Plant (Nov.2006) analyzed by Dan Howell.
Pressure of Coke Oven Gas at inlet to PSA	Psig	400	

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

4.2 Product

PSA Outlet (H₂ Product);

Flow (100% Hydrogen)	mmscfd	17	
Pressure	psi(g)	375	
Temperature	°F	95 - 100	Approx
Hydrogen Purity			
Hydrogen	Vol %	99.9	Min (water saturated)
Design Turndown	%	30	

4.3 Offgas

Offgas Temperature	95 – 100 °F
Offgas Pressure (minimum)	10 psi(g) (Notes 1&2)

Notes:

1. Minimum pressure measured downstream of PSA offgas FCV.
2. In order that hydrogen recovery performance is achieved, it is important that this pressure is not exceeded.

4.3.1 Battery Limit Conditions of Feed, Product and Relief Header

Stream	Operating Conditions		Design Conditions	
	Temp (°F)	Pressure psi(g)	Temp (°F)	Pressure (psi(g))
Feed Gas to PSA	95-100	406	150	510
Product Hydrogen	95-100	375	150	460
Offgas exit PSA	100	10 (Notes 1&2)	150	70
Flare header at PSV outlet	Ambient	TBA	TBA	TBA

4.3.2 Nitrogen

Pressure		
Operating	Norm	100 psi(g) 600 psi(g)
Temperature		
Operating	Norm.	77-81 °F
Quality		
Content		99.5 mol% with max 20ppmv O ₂

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

4.3.3 Instrument Air

Pressure		
Operating	Min.	60 psi (g)
	Normal	80 psi (g)
	Design	100 psi (g)
Design for actuators		100 psi (g)
Temperature		
Norm		Ambient
Design		150 °F
Quality		
Dew Point		Dry, Oil Free and clean
		-20.0 °C

4.3.4 Cooling water

Supply pressure	60 psi (g)
Supply Temperature	85°F
Return temperature	110°F
Return Pressure	40 psi (g)

4.3.5 Power

Power	LV	120
	MV	460
	HV	4000

4.4 Utility consumption

Oil Filled Gas Screw Compressor requirements;

First Stage	kW	5000
Second stage	kW	6000
Cooling water (Heat exchanger)	lb/hr	153,572
Cooling water compressor	lb/hr	544,301
Instrument Air	ft ³	1800
Nitrogen (Purge)	ft ³	290,000

4.5 Study Assumptions

Ambient conditions	
Barometric Pressure	14.7 psia
Ambient temperature	24 to 100 °F

4.6 Exceptions

- The study except where specifically stated otherwise is restricted to a process design review only;
- Detailed instrument design and engineering;
- Detailed pipe design.
- All and any costs quoted are of 'indicative' accuracy only.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIREC	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

5.0 SCOPE OF STUDY

The scope of the study is limited to the following, which is detailed below;

- The study covers the pre- and post- purification of the COG in the PSA unit and advises on the most suitable equipment;
- Information on PSA technology, detailing the required equipment and operating conditions in order to maximise hydrogen production reliability and purity.
- Provide preliminary basic design information for ancillary equipment such as COG compressor and storage tank.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

6.0 PROCESS DESCRIPTION

The following process descriptions should be read in conjunction with Process Flow Diagrams (PFD's) J10556 – 1002/1003 and Process and Instrument Diagrams (P&ID's) J10556 – 1010/1011/1012/1013/1014. The process description is intended to give a detailed narrative of the system but also serves as a justification for the equipment selected thereby highlighting some of the difficulties in processing coke oven gas for hydrogen purification.

Coke oven gas is delivered to the PSA Unit at approximately 429psig and 100°F. At this stage the COG contains hydrocarbons liquids and sulphur compounds that would be poisons to the adsorbents in the PSA Unit and must therefore be removed / reduced in the pre-purification and conditioning.

Under flow control, at approximately 38mmscfd the COG is routed to the Feed Coalescer V-212. The purpose of this equipment is to prevent liquid carry over of hydrocarbons to the upstream units, the coalescer is under level control and all liquids flow to drain.

The liquid free gas then passes out of the Feed Coalescer and flows into Two (2) x 100% Pre-Purification Units (PPU) V-220A/B. The PPU vessels are used to remove Tar, Naphthalene, Bulk BTX, compressor oil and small amounts of sulphur compounds from the feedstock. The two PPU Vessels are operated in a led-lag configuration to provide frequent change out of the silica gel/activated carbon media without disturbing the flow of COG through the PSA plant.

Now that the COG has been cleaned and is in a suitable condition it is passed to the 5 bed Pressure Swing Adsorbers V-201A/B/C/D/E at approximately 406psig and 100°F.

The pressure swing adsorption unit is a series of adsorber vessels connected by a number of pipe manifolds. Flow to each of the adsorbers is controlled by switching valves. Continuous product flow is achieved by cycling adsorber operation between run, depressurisation, purge and repressurisation.

In these units high purity hydrogen is produced (99.9%) using a 5-vessel pressure swing adsorption (PSA) system. Each of the adsorber units selectively adsorbs the impurities from a hydrogen rich feed gas onto a fixed bed of adsorbents.

Further description of the PSA Unit and technology can be found in section 6.1 of this document.

Although the Hydrogen is now at a high purity level due to the composition of the COG feedstock it will still contain a large amount of oxygen. This will need to be removed before it can be used in ammonia synthesis. The Purified Hydrogen is therefore routed to One (1) x 100% Deoxo vessel V – 206 where it is reacted at (109°F) with the oxygen over a Palladium catalyst bed in the following stoichio-metric exothermic reaction;

The temperature is expected to rise by 245°F. The catalyst bed must be protected against excessive temperature rise. This is achieved by an over temperature shutdown system.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

The hydrogen product from the Deoxo unit is then cooled by the Hydrogen product cooler E-207 from 354°F to the battery limit of 100°F. As a result of this large amounts of water are condensed. The two phases are separated from each other in the vertical Product separator V-213.

The Pure Hydrogen (99.9%) product, saturated with water vapour is then delivered to the battery limits under pressure control, (375psig and 100F) for further processing and ammonia synthesis.

6.1 PSA Unit

The plant uses Natco UK Limited fixed bed Pressure Swing Adsorption System to produce a high purity product stream. The adsorption process operates on a repeated cycle with the basic steps of adsorption and regeneration. Although the process is a batch operation, continuous product and offgas flows are achieved by operating the adsorbers in a stepwise manner. No matter how complex the PSA sequence it can always be broken down into the following 5 fundamental steps:

- Adsorption
- Co-Current Depressurisation
- Counter-Current Depressurisation
- Purge
- Counter-Current Repressurisation

Impurities from the feed are passed into the two offgas drums V-203A/B for further use in the plants fuel gas system. The product pressure control maintains a back pressure in the system and spills over pressure into the vent header. The system offered is a 5 Absorber system

A brief description is given below of the operating cycle that each adsorber goes through.

6.1.1 Adsorption

Impure feed gas flows into the adsorber vessel currently online during the run step at pressure. As the gas is passed over the adsorbent beds, the adsorbents, solid granular materials, selectively adsorb the impurities, leaving a high purity hydrogen product. Feed flow continues through the bed until the bed is just short of being fully saturated with impurities, and the feed gas is then automatically switched to a clean adsorber. The loaded adsorber is taken off line for regeneration.

6.1.2 Regeneration

Regeneration is the process of desorbing the impurities, which were adsorbed during the adsorption step, followed by repressurisation to adsorption pressure so that the cycle can be repeated.

There are 3 basic steps to the regeneration process:

- i. Depressurisation,
- ii. Purging and,
- iii. Repressurisation.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

6.1.3 Depressurisation

At the end of the adsorption step, the adsorber vessel contains a significant amount of high purity hydrogen in the void volume of the adsorbent bed. Much of this hydrogen is saved by pressure equalisation with other vessels, after which the adsorber is blown down to low pressure in preparation for the next step.

6.1.4 Purging

The impurities remaining after blow down are removed by passing hydrogen from one adsorber at pressure in a counter current direction through the adsorber which has just been blown down.

6.1.5 Repressurisation

The adsorber is now ready to be repressurised to adsorption pressure ready for a new run step. This is achieved by a combination of pressure equalisation with other vessels and repressurisation with a slipstream of pure hydrogen.

The feed flowmeter is used to compute the PSA's repressurisation gas requirements and hence cycle time. As the feedrate to the plant is reduced so the cycle time is extended to maintain optimum efficiency.

6.1.6 Offgas Drum

Two offgas drums, a configuration selected to provide the required geometric volume in the most cost effective and practical way, complete with internal mixing device is included. With appropriate controls, a steady and continuous offgas flowrate can be maintained.

The offgas drum control system is so designed that the drum runs at the lowest average pressure consistent with requirements of battery limits (10 psig), thus enabling maximum PSA efficiency to be achieved.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

6.2 PSA Technology

The design of the Pressure Swing Adsorption System utilises technology developed by Natco UK Limited. The basic design concepts have been well proven in service, in excess of one hundred PSA units having been installed over the past twenty years, operating on a wide variety of hydrogen rich feedstocks.

6.2.1 Reliability

The reliability of the Pressure Swing Adsorption System depends upon the successful operation of the switching logic and the associated switch valves. These switch valves are required to open and close bubble tight at a relatively high frequency of approximately three complete cycles per hour. In order to ensure the tight shut off is maintained, the plug and seat arrangement incorporates a PTFE insert.

Prior to acceptance by Natco UK of valves suitable for this arduous duty the manufacturer is required to successfully carry out a "life" test, whereby a tight shut off is maintained over the test period, representing well in excess of a year's operation of the unit.

6.2.2 Operating Range and Efficiency

The efficiency of Pressure Swing Adsorption Units, in terms of ability to recover hydrogen from the feed stream, is very sensitive to the off-gas pressure specified for the system.

However, turndown to virtually zero product flow is possible whilst still maintaining specified purity, although efficiency will be lost as a consequence of the loss of accuracy of flow measurements at low flow rates.

The PSA is designed for the entire range of 30 – 100% of the product capacity.

6.2.3 Control

The control system controls the sequencing functions and automatic capacity control for a 5 bed, one vessel online PSA System. Timing is critical to the process and to this end the control system design should include consideration of the scan time of the system which is to be implemented.

All PID control loop process variables, Adsorber Vessel pressures and product / feed flow rates shall be available to the operator for historical trending to enable accurate tuning / commissioning and monitoring of the PSA. PID control loop set point adjustment, mode selection, PSA sequence status and Run / Output Inhibit / Auto / Manual mode selection should also be available to the operator.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

6.3 Coke Oven Gas Compressor

The COG will be discharged from the Ovens at approximately 20 psig. In order to maximise the recovery of the hydrogen present in the COG high pressure will be required on the inlet to the PSA system. (See general comments below). In addition to this large fluctuations in flow from the COG ovens (± 20) means that not only will the compressor have to cope with high pressure but varying flowrates.

Initial contact has been made with KOBELCO EDTI COMPRESSORS INC for the design of such a specialised compressor. From the information supplied 2 (Two) Screw type compressors will be required. The 1st compressor will be 5,347 HP oil-free screw compressor and the 2nd stage, 4,370 oil-injected screw compressor. The pressure outlet from the 2nd stage oil filled compressor will be 470psi(g) sufficient to send the COG to intermediate storage tank.

The indicative price of these 2 COG compressors Units is

\$5,000,000 (Five Million US Dollars)

Details of KOBELCO EDTI proposal can be found in Appendix A.

6.4 Intermediate Storage Tank

The PSA system does not lend itself well to significant pressure or flow fluctuation its preferred mode of operation is for steady state conditions. This provides the best opportunity to maximise hydrogen recovery and purity.

The PSA is designed for 38mmscfd maximum working capacity. The maximum flow fluctuations observed over a 10 min period from the COG oven is 38 mmscfd $\pm 20\%$. From Natco UK calculations a 10,000 cubic foot (ft³) with dimensions 13'7" i/d x 68'3" Tan/Tan should be sufficient to even the flow of gas prior to reaching the PSA unit.

During high flow a maximum operating pressure of 456.81psi(g) will be observed. This will therefore not exceed the design limitation of the compressor outlined above. The flow can then be delivered to the PSA unit under pressure control.

The calculations for the size of the Intermediate storage tank can be found in Appendix B.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

7.0 GENERAL COMMENTS

The statements below are intended to highlight decisions made during the execution of the design study. The points are of interest and may be useful in any future justification of the Ammonia synthesis project.

- A design pressure of 400psig has been selected for the operating pressure of the PSA Unit. This was selected as from Natco PSA calculation runs higher pressures (600-650psig) do not produce an increase in hydrogen recovery. The 400psig pressure is not excessively large as to make the design of the COG complicated and not feasible;
- From Natco UK experience liquids carried through to the PSA absorbents drastically reduced the purity and recovery of hydrogen from the PSA unit. The Feed Coalescer V-212 plays an important part in achieving this objective;
- The recovery of the hydrogen from the COG is approximately 78% producing 17mmsfd;
- The Pre-Purification Units (PPU) vessels are used to remove Tar Naphthalene Bulk BTX and compressor oil from the feedstock. Although Naphthalene does not pose a severe problem and design of the PSA can account for this Naphthalene crystallisation is an issue and must be removed prior to the COG entering the PSA adsorber beds.
- The PPU is also expected to removal small amounts of sulphur compounds and thereby further protecting the PSA absorber beds.
- The Pre-purification Units V-220 use 'fifty-fifty' wide pore silica gel and activated carbon as the media to remove hydrocarbons/sulphur compounds which is non-regenerative and life expectancy is approximately 6 months.
- The Deoxo unit is positioned after the PSA to prevent poisoning of the catalyst from sulphide compounds. Carbon Monoxide will similarly be at low levels (less than 10ppm) and will therefore not act as a Deoxo catalyst inhibitor. A feed temperature of 109⁰F is sufficient to catalyse reaction.
- The Deoxo Unit V-206 use a Palladium (Pd) based catalyst which has a guaranteed design life of 2 years;

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

8.0 SCOPE OF SUPPLY

The following schedule outlines the main items of equipment for supply of a COG PSA Unit. The scope of supply includes all necessary equipment in order to achieve maximum reliability purity and recovery.

Details of individual items of equipment and ancillaries are given below;

VESSELS	
Item	No Required
V – 212 Feed Coalescer	1-off
V – 220A/B Pre-Purification Units (PPU)	2-off
V – 201A/B/C/D/E PSA Adsorber Vessels	5-off
V – 203A/B Offgas Drum	2-off
V – 206 Deoxo Unit	1-off
V – 213 Product Separator	1-off

Vessels to be designed in accordance with ASME VIII Div1 and 2. Please refer to vessel datasheets in Appendix

Heat Exchanger	
Item	No Required
E – 207 Product Cooler	1-off

Heat exchangers to be designed in accordance with ASME VIII Div1 and TEMA class R.

PSA SKID	
Item	No Required
200 – V Complete skid mounted valve skid	1-off

The plant is complete within battery limits, supplied as a modular unit, prefabricated to the maximum extent thus minimising field installation time and labour. All vessels and major equipment will be installed off module.

Piping for all module-mounted equipment will be prefabricated and installed between items of equipment and from equipment to module edge where the termination will be with a standard flanged connection. Interconnecting pipework between module and off module equipment and pipework to battery limit will either be supplied prefabricated with cutting lengths for fit-up or will be site run.

All on-module electrical and instrument connections will be made in a junction box, with appropriate segregation of electrical circuits, mounted at the edge of the module.

Control of PSA sequencing will be by a new control system programmed to provide the necessary control required by the sequence logic diagrams.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

8.1 PSA Unit

8.1.1 Adsorbents

Natco UK Limited will supply the initial charge of all, adsorbents for the five adsorber vessels. Packed in steel or plastic airtight drums to minimise water pick up from the atmosphere.

Approximate quantities per Adsorber vessel	Alumina	5.268	Te
	Carbon	24.129	Te
	Mol Sieve	21.834	Te
Approximate quantities per PPU vessel	WP Silica gel	30.74	Te
	Carbon	19.08	Te
Approximate quantities per Deoxo vessel	Palladium	8.84	Te

8.1.2 Relief Valves

1 full sized safety valve's with inlet and outlet isolation valve (locked) will be supplied and installed on the skid for each adsorber and 2 full sized safety valves with inlet and outlet isolation valves (locked) for the offgas drum.

8.1.3 Vessel Isolation Valves

Isolation valves complete with spectacle blinds are provided for each set of adsorber automatic valves, such that a complete adsorber including instrumentation can be isolated. A hard piped nitrogen connection is included for each absorber.

8.1.4 Instrumentation

The instrumentation provided by Natco UK Limited will be engineered and designed in accordance with the following standards;

Instrumentation will be electronic, certified to North American standards designed for use within a Class 1 Div 2 Group 3 hazardous, Temperature Class T3.

Equipment supplied by Natco UK Limited will include:-

- Automatic switching valves for cycle sequence control, complete with solenoid valves certified or positioner as required by service.
- Interconnecting pipework, wiring and tubing for all items located on module;
- All other local on module instrumentation and devices necessary to make the unit automatic and safe during operation;
- All services such as control stations, instrument air supply and instrument signals will be piped or wired by Natco UK to module edge;
- Switching valve position detection by proximity switch is included indicating closed/not closed, wired to module edge junction box;
- Control System for PSA sequencing functions.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

8.1.5 Adsorber Logic System

Natco UK will supply the control system as described below. Built into the system as standard features are the following:-

- Automatic manual/manual step function - Enables system to run on automatic or the sequence to be stepped manually.
- Output Inhibit Function - Enables controller to continue to sequence but all outputs are isolated.
- Vessel State Indication - Outputs from the PLC indicate which vessel is on RUN, PURGE, HOLD, etc.

The control system shall have the capability to automatically close feed, product and purge valves and signal shutdown for the following conditions:

- Long cycle time
- Low product gas pressure
- Power failure
- Low instrument air pressure

8.1.6 Noise Control

A silencer is included in the offgas line to the offgas drum

8.1.7 Insulation and Heat Tracing

Insulation and heat tracing is supplied. No fireproofing has been included.

8.1.8 Painting

Off module piping, equipment and steelwork will be primed only. Module and on module piping, equipment and steelwork will be supplied grit blasted, primed and finish painted in the fabrication shop.

8.1.9 Lighting

Lighting is excluded from Natco UK scope of supply.

8.1.10 Exclusions

- Interconnecting pipework fittings and valves between skid and absorber vessels;
- Interconnecting cable, cable tray and insulation materials between new module and PSA module and control system/panel
- Analysis equipment
- All field installation work
- Any item not specifically mentioned above

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRES	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

8.2 General

The following check list provides a summary of the split in responsibility between Natco UK Limited and the Purchaser for supply and installation of equipment and provision of services.

EQUIPMENT/SERVICE PLANT 1	NATCO UK		PURCHASER	
	SUPPLY	INSTALL	SUPPLY	INSTALL
V – 212 Feed Coalescer	√			√
V – 220A/B Pre-Purification Units (PPU)	√			√
V – 201A/B/C/D/E PSA Adsorber Vessels	√			√
V – 203A/B Offgas Drum	√			√
V – 206 Deoxo Unit	√			√
V – 213 Product Separator	√			√
2 stage screw compressor			√	√
<i>Prefabricated Switching Valve Module</i>	√			√
PSA Adsorbents	√			√
PPU Activated carbon	√			√
Deoxo Unit Palladium catalyst	√			√
Pipework:	√	√		
<i>Pipework, Valves and Fittings On Module</i>	√			
<i>Pipework, Valves and Fittings Off Module</i>	√			√
Instrument and Electrics:	√			√
<i>Field Instrumentation</i>	√			
<i>PLC for Sequence Control</i>	√			
<i>Control Panel</i>	√			√
<i>On Module Interconnections</i>	√	√		
<i>Off Module Interconnections (ISBL)</i>			√	√
<i>Plant to Control Room Interconnections</i>			√	√
<i>Heat Tracing</i>	N/A			
<i>Lighting</i>			√	√
<i>Insulation</i>	Within BL			√
On Module Painting:	√	√		
<i>Primer</i>	√	√		
<i>Finish</i>	√	√		
Delivery – FOB	√			
Installation on Prepared Foundations				√
Foundations:	√			
<i>Foundation Loading Diagram</i>				

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIREC	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

EQUIPMENT/SERVICE PLANT 1	NATCO UK		PURCHASER	
	SUPPLY	INSTALL	SUPPLY	INSTALL
<i>Design</i>			√	
<i>Civil Work</i>			√	
<i>Anchor Bolts</i>			√	√
Operating & Maintenance Manuals	√			
Commissioning Services per Diem Rates	√			
Spare Parts: <i>Commissioning</i>	√			
<i>2 Years' Operation</i>			√	

8.3 Budget cost

Our budget estimate for the design supply of the complete PSA Plant Including pre-purification units, Deoxo unit and product cooler as detailed in PFD's J10556 – 1002/1003 and generally as described above is:

\$ 4,500,000 (Four Million Five Hundred Thousand US Dollars)

Please note this is a budget price and not an offer to sell.

[NATCO]		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRE	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

9.0 DATA SHEETS AND DRAWINGS

9.1 Material Balance

A material balance for the PSA plant has been generated and is presented together with the PFD's in Appendix C.

NATCO		Project:	Synthetic Ammonia from Coke Oven Gas and Coal Project
Client:	AKER KVAERNER FOR ENVIRE	Equipment:	Pressure Swing Adsorption Unit
Client Ref:	-	Natco Ref:	J10556

10.0 STUDY CONCLUSIONS

It is the conclusion of this study that with the Process design and equipment scheduled in section 8.0 of this report, operation with the defined Coke oven gas feedstock is feasible and practical. The recovery of pure hydrogen saturated with water, (99.9%) from coke oven gas will be 78% giving a total flow of 17mmscfd.

Pre-purification Units have been included in the design to prevent large amounts hydrocarbon and sulphur compounds poisoning the absorbents beds present in the PSA.

Due to the large quantities of oxygen present a Deoxo unit will need to be installed after the PSA to provide pure hydrogen (99.9%) at the battery limits of 375psig and 100°F.

APPENDIX A

OIL SCREW COMPRESSOR

301 N. Smith Ave. Corona, CA 92880-1742 Ph: 951-279-9400 • Fax: 951-279-9464 ISO 9001 REGISTERED	TITLE: COKE OVEN GAS COMPRESSOR Q-27150A-00 CUSTOMER: NATCO UK, LTD. CUSTOMER REF:	DATE: October • 2007
DESIGN CRITERIA	Inlet Press. P1 (kPa) Outlet Press P2 (kPa) Inlet Temp. T1 (deg C) Outlet Temp. T2 (degC) Gas M.W. (Avg.) Gas K Value Main/Aux/Control Voltage (VAC) Flow Rate at Inlet (M ³ /hr)	34.7 114.7 104.0 Uncooled 10.32 1.358 4000 / 460 / 120 45.40	
COMPRESSOR DATA	Compressor Model Compressor Stages Compressor Quantity Compressor Speed (rpm) Compressor Seal Compressor BHP (kw)	KS50LNZ One One 1770 Dry Gas, Non-Contact With Back Up. 4370	
SCOPE OF SUPPLY	Compressor Motor (kw) Motor Speed (rpm) Gear Box No. Of Shafts Oil Reservoir Lube Oil Piping (Before Oil Filter) Lube Oil Piping (After Oil Filter) Oil-Pumps Oil-Coolers Oil Filters After-Cooler Inlet Filters Control System Silencer Process Gas Piping / Instruments Noise Enclosure / Sound Jacket Vibration Probes Bearing Temperature Detector Vibration / Bearing Temperature Monitor Painting Shipping	KS50LNZ 5000, WP II 1770 AGMA Class 2,1770→4567 8' x 9' x 6' (C.S) Carbon Steel Stainless Steel Dual Dual - Shell & Tube - BEM with Transfer Valve Duplex N/A 2 x 42" Dia By customer (JB is provided) Included (Suction/ Discharge) Not Included Not Included Accelerometer(1) / Axial Probe(2) per Compressor Accelerometer(1) / Axial Probe(0) / Vibration Probe(0) per Gear Box Bearing RTD (0) per Gear Box Not Included Manufacturer's Standard Manufacturer's Standard for Domestic	
PRICING, DELIVERY DIMENSIONS,	COMPLETE SYSTEM (EX- FACTORY) BUDGETARY PRICING <u>DIMENSIONS</u> <u>DELIVERY</u>	See Attached	
ETC NOTES

301 N. Smith Ave.
Corona, CA 92880-1742
Ph: 951-279-9400 Fax : 951-279-9464
ISO 9001 REGISTERED

TITLE: COKE OVEN GAS COMPRESSOR
Q-27150B-00 CUSTOMER: NATCO UK, LTD.
CUSTOMER REF:

DATE: October 2, 2007

		CASE 1
DESIGN CRITERIA	Inlet Press P1 @ Comp Flange (psig)	110.7
	Outlet Press P2 @ Comp Flange (psig)	469.7
	Inlet Temp. T1 (EF)	120.0
	Outlet Temp. T2 (EF)	Uncooled
	Gas M.W. (Avg.)	10.32
	Gas K Value	1.358
	Main/Aux/Control Voltage (VAC)	4000 / 460 / 120
	Flow Rate at Inlet (mmscfd)	42.83
	Oil Carryover Limit (ppmw)	1.0
COMPRESSOR DATA	Compressor Model	KS50MEH
	Compressor Stages	One
	Compressor Quantity	One
	Compressor Speed (rpm)	1770
	Compressor Capacity (mmscfd)	42.83
	Compressor bhp (hp)	5347
	Disch Temp. (EF)	210
	Comp. Oil Flow/Temp. (gpm/EF)	546 / 120
SCOPE OF SUPPLY	Comp. Suction Knockout Drum	By Others
	Compressor	KS50MEH
	Motor (kW)	6,000 WP11 (Space Heater, Winding RTD's)
	Bulk Oil Separator (in)	84" Dia
	Coalescing Oil Separator(s) (in)	2 x 42" Dia, Duplex with switchover valving
	Charcoal Adsorber (in)	N/A
	Dust Filter (in)	N/A
	After-Cooler	N/A
	Knockout Drum, Disch. (in)	N/A
	Oil-Pumps	Single, PT, 3 Rotor Screw
	Oil-Coolers	Single, Shell & Tube, BEM Design
	Oil Filters	Duplex
	Control System	PLC (AB), NEMA 4X
	No. of Skids	Three
PRICING, DELIVERY DIMENSIONS, ETC. (Budgetary Basis)	COMPLETE SYSTEM (Ex-Factory) Net Each.	\$
	COMMISSIONING SPARES	\$
	LUBE OIL	\$
	INSULATION	\$
	NOISE JACKET	\$
	DIMENSIONS	See Attached
DELIVERY		
NOTES

OIL-INJECTED SCREW COMPRESSOR DATA SHEET

E7B0458 / Q-27150

SERVICE : Coke Oven Gas
MODEL : KS50LNZ-KS50MEH
NO. OF STAGE : 2

ITEM NO. :
TYPE OF DRIVER : ELECTRIC MOTOR
NO. REQUIRED : 1

ALL DATA ON PER ONE UNIT BASIS

		Design	Design		
		1st stage	2nd stage		
			Full Load		
GAS HANDLED					
FLOW RATE	lbm/h WET	54610	50873		
(14.7PSIA & 32 Deg.F)	Nm3/h WET	50630	48552		
(14.7PSIA & 32 Deg.F)	Nm3/h DRY	49069	47802		
(14.7PSIA & 60 Deg.F)	MMSCFD WET	45.36	43.50		
(14.7PSIA & 60 Deg.F)	MMSCFD DRY	43.96	42.83		
ATM PRESSURE		PSI	14.7	14.7	
SUCTION CONDITIONS					
PRESSURE AT TERMINAL POINT	PSIG	20	95		
TEMPERATURE	deg.F	104	120		
RELATIVE HUMIDITY	%	100	100		
MOLECULAR WEIGHT		10.96	10.32		
SPECIFIC HEAT RATIO (K1=Cp/Cv)		1.355	1.358		
COMPRESSIBILITY(Z1)		1.000	1.0010		
INLET VOLUME	ACFM	14470	4517		
DISCHARGE CONDITIONS					
PRESSURE AT OUTLET FLANGE	PSIG	100	470		
TEMPERATURE	deg.F	349	210		
TEMP. COOLER OUTLET	deg.F	-----	-----		
PRESSURE RATIO		3.304	4.419		
BHP AT COMP. INLET SHAFT	HP	4370	5347		
SEPARATE GEAR LOSS	HP	109	0		
TOTAL BHP(GEAR LOSS INCL.)	HP	4479	5347		
SPEED	rpm	4567	1770		
DRIVER RATED	HP	5000	5900		
DRIVER SPEED	rpm	1770	1770		
SLIDE VALVE VOLUME RATIO	%	NA	100.0		
(*1) FLOW RATE SHALL NOT BE LESS THAN 0% OF SPECIFIED FIGURE AT "Design"					
(*2) BHP SHALL NOT EXCEED 104% OF QUOTED FIGURE AT "Design"					

		Design	Design		
		1st stage	2nd stage		
GAS COMPOSITION(Mol. %)					
HYDROGEN		55.986	58.415		
CARBON MONOXIDE		5.028	5.246		
CARBON DIOXIDE		1.257	0.012		
OXYGEN		0.774	0.807		
NITROGEN		6.769	7.062		
METHANE		23.207	24.213		
ETHANE		0.387	0.404		
ETHYLENE		1.837	1.917		
PROPANE		0.097	0.101		
PROPYLENE		0.290	0.303		
BENZENE		0.345	0.360		
TOLUENE		0.155	0.161		
O-XYLENE		0.001	0.002		
P-XYLENE		0.001	0.002		
HYDROGEN SULFIDE		0.376	0.392		
HYDROGEN CYANIDE		0.242	0.252		
AMMONIA		0.073	0.076		
SULFUR DIOXIDE		0.037	0.038		
WATER		3.139	0.238		
TOTAL	%	100.00	100.00		
RELATIVE HUMIDITY	%	100.00	100.00		
AVERAGE MOL. WEIGHT		10.96	10.32

FIRST STAGE COMPRESSOR SYSTEM

**KOBELCO EDI
COMPRESSORS INC.**

301 N. SMITH AVE.
CORONA, CA 92880
PH.: (951) 279-9400
FAX.: (951) 279-9464

TITLE: MECHANICAL FLOW DIAGRAM
COKE OVEN GAS COMPRESSOR SYSTEM

DESIGN:	C.C.	DATE	SCALE:
DRAWN:	M.M.		NONE
CHECKED:			SHEET
APPROVED:			1 OF 2

PROJECT NO. Q-27150A

DRAWING NO.
Q-27150A-001

REVISION DATE :
CAD FILENAME: 27150A0011
LAST PLOT DATE:

REV.
A

**KOBELCO EDI
COMPRESSORS INC.**

301 N. SMITH AVE.
CORONA, CA 92880
PH.: (951) 279-9400
FAX.: (951) 279-9464

TITLE: MECHANICAL FLOW DIAGRAM
COKE OVEN GAS COMPRESSOR SYSTEM
(SEAL SECTION)

DESIGN:	C.C.	DATE	SCALE:
DRAWN:	MARCUS		NONE
CHECKED:			SHEET
APPROVED:			2 OF 2

PROJECT NO. Q-27150A

DRAWING NO.
Q-27150A-001

REVISION DATE :
CAD FILENAME: 27150A0012
LAST PLOT DATE:

REV.
A

SECOND STAGE COMPRESSOR SYSTEM

**KOBELCO EDI
COMPRESSORS INC.**

301 N. SMITH AVE.
CORONA, CA 92880
PH.: (951) 279-9400
FAX.: (951) 279-9464

TITLE: MECHANICAL FLOW DIAGRAM
COKE OVEN GAS COMPRESSOR SYSTEM

DESIGN:	M.G.	DATE	SCALE:
DRAWN:	MARCUS		NONE
CHECKED:			SHEET
APPROVED:			1 OF 1

PROJECT NO. Q-27150B

DRAWING NO.

Q-27150B-001

REV.
A

REVISION DATE :
CAD FILENAME: 27150B0011
LAST PLOT DATE:

FIRST STAGE COMPRESSOR SYSTEM

**KOBELCO EDTI
COMPRESSORS INC.**

301 N. SMITH AVE.
CORONA, CA 92880
PH.: (951) 279-9400
FAX.: (951) 279-9464

TITLE: MACHINERY ARRANGEMENT
COKE OVEN GAS COMPRESSOR SYSTEM

DESIGN:	C.C.	DATE	SCALE:
DRAWN:	MARCUS		NONE
CHECKED:			SHEET
APPROVED:			1 OF 1

PROJECT NO. Q-27150A

DRAWING NO. Q-27150A-011

REV.
A

REVISION DATE :
CAD FILENAME: 27150A0111
LAST PLOT DATE:

HEIGHT = 168' [14'-0"]
WEIGHT = 90,000#

HEIGHT = 156' [13'-0"]
WEIGHT = 85,000#

SECOND STAGE COMPRESSOR SYSTEM

**KOBELCO EDI
COMPRESSORS INC.**

301 N. SMITH AVE.
CORONA, CA 92880
PH.: (951) 279-9400
FAX.: (951) 279-9464

TITLE: MACHINERY ARRANGEMENT
COKE OVEN GAS COMPRESSOR SYSTEM

DESIGN: C.C.	DATE	SCALE:
DRAWN: MARCUS		NONE
CHECKED:		SHEET
APPROVED:		1 OF 1

PROJECT NO. Q-27150B

DRAWING NO.
Q-27150B-011

REV.
A

REVISION DATE :
CAD FILENAME: 27150B0111
LAST PLOT DATE:

APPENDIX B

STORAGE TANK CALCULATION

CONSIDER PSA FEED STORAGE

38.0 mScfd

Assumptions		
1. Average Feedgas Rate, Nm3/hr	42,415	4,174.66 Lbmols/hr
2. Maximum Feedgas Rate, Nm3/hr	50,897	5,009.59 Lbmols/hr
3. Temperature, F	100	
4. Minimum storage pressure, psig	415	
5. FCV DP Upstream of PSA, psi	15	
6. PSA Operating Pressure, psig	400	

	Time	Buffer Storage				Storage Flowrate, Lbmols/hr		Net Flow to Storage	Storage Pressure
	Minutes	L/D Ratio	Diameter (ft)	Length (ft)	Cubic Feet	Feed	Product	Lbmols	psig
CASE A									
Max Flow Cycle Start	0.0	5.0	6.3	31.7	1,000	5,009.59	4,174.66	0.00	415.00
Max Flow Cycle End	5.0	5.0	6.3	31.7	1,000	5,009.59	4,174.66	69.58	833.08
Min Flow Cycle End	10.0	5.0	6.3	31.7	1,000	3,339.72	4,174.66	-69.58	415.00
CASE B									
Max Flow Cycle Start	0.0	5.0	10.8	54.2	5,000	5,009.59	4,174.66	0.00	415.00
Max Flow Cycle End	5.0	5.0	10.8	54.2	5,000	5,009.59	4,174.66	69.58	498.62
Min Flow Cycle End	10.0	5.0	10.8	54.2	5,000	3,339.72	4,174.66	-69.58	415.00
CASE C									
Max Flow Cycle Start	0.0	5.0	13.7	68.3	10,000	5,009.59	4,174.66	0.00	415.00
Max Flow Cycle End	5.0	5.0	13.7	68.3	10,000	5,009.59	4,174.66	69.58	456.81
Min Flow Cycle End	10.0	5.0	13.7	68.3	10,000	3,339.72	4,174.66	-69.58	415.00

DESIGN BASIS

APPENDIX C

PRELIMINARY MATERIAL BALANCE

AND

PROCESS FLOW DIAGRAMS (PFD'S)

PRESSURE SWING ADSORPTION UNIT

MASS BALANCE

0		Issued For Design Study	CPG						
REV	DATE	DESCRIPTION	Originator.	Checked.	Mech.	Process	Instr.	Proj. Man.	
PROJECT TITLE:			SPECIFICATION NUMBER:						
PRESSURE SWING ADSORPTION UNIT			J10556 - 3401						
CLIENT:			JOB NO.:						
AKER KVAERNER FOR ENVIRES			J10556						
			SHEET 1 OF 2						

J10556 -3401 38MMSCFD PSA AKER KVAERNER FOR ENVIRES

Stream Number		1	2	3	4	5	6	7	8
Stream Name		H ₂ Feed COG	PSA Feed	PSA Product	Offgas	De-Oxo Product	Ex Cooler	H ₂ product	Water H ₂ O
Vapor Fraction		1.00	1.00	1.00	1.00	1.00	0.99	1.00	0.00
Temperature	F	100.00	99.95	108.95	90.95	354.47	100.00	99.88	99.88
Pressure	psig	413.00	406.00	401.00	10.00	396.00	391.00	375.00	375.00
Enthalpy	btu/hr	1955089.50	1991120.31	1047572.19	1030352.64	4266108.17	426900.14	905506.66	-478606.49
Entropy	btu/R/hr	-13785.34	-13602.02	-10591.44	6861.02	-5603.63	-11353.14	-10437.40	-763.37
Vapor Density	lb/ft3	0.75	0.74	0.15	0.07	0.11	0.14	0.13	
Liquid 1 Density	lb/ft3	53.86					62.93		62.93
Liquid 1 Specific Gravity	60F@STP	0.88					1.00		1.00
Vapor Cp	btu/lbmol/R	7.68	7.68	6.92	7.99	6.96	6.92	6.92	
Vapor Cv	btu/lbmol/R	5.55	5.56	4.91	5.99	4.96	4.91	4.91	
Liquid 1 Cp	btu/lbmol/R								17.97
Vapor Viscosity	cP	0.01	0.01	0.01	0.01	0.01	0.01	0.01	
Liquid 1 Viscosity	cP	0.57					0.68		0.68
Vapor Thermal Conductivity	btu/ft/hr/F	0.05	0.05	0.11	0.03	0.14	0.11	0.11	
Liquid 1 Thermal Conductivity	btu/ft/hr/F	0.08					0.36		0.36
Vapor Flowrate	mmscf/day@STP	38.08	38.09	17.38	20.71	17.24	17.00	17.00	
Liquid 1 Flowrate	gal/min@STP	0.50					0.96		0.95
Liquid 2 Flowrate	gal/min						0.96		0.95
Molecular Weight		10.64	10.62	2.26	17.63	2.28	2.28	2.05	18.01
Molar Flowrate	lbmol/hr	4184.27	4182.73	1908.87	2273.86	1893.40	1893.40	1866.99	26.41
Mass Flowrate	lb/hr	44534.91	44410.13	4311.95	40098.14	4311.85	4311.85	3836.30	475.60
Note: All Liquid 1 Phase calculations exclude Free Water									

J10556 -3401 38MMSCFD PSA AKER KVAERNER FOR ENVIRES

Stream Number		1	2	3	4	5	6	7	8
Stream Name		H ₂ Feed COG	PSA Feed	PSA Product	Offgas	De-Oxo Product	Ex Cooler	H ₂ product	Water H ₂ O
Molar Flowrate By Component									
HYDROGEN	lbmol/hr	2427.44	2427.44	1893.40	534.04	1862.47	1862.47	1862.46	0.01
CARBON MONOXIDE	lbmol/hr	218.13	218.13	0.00	218.13	0.00	0.00	0.00	0.00
CARBON DIOXIDE	lbmol/hr	54.32	54.32	0.00	54.32	0.00	0.00	0.00	0.00
OXYGEN	lbmol/hr	33.62	33.62	15.47	18.16	0.00	0.00	0.00	0.00
NITROGEN	lbmol/hr	293.57	293.57	0.00	293.57	0.00	0.00	0.00	0.00
METHANE	lbmol/hr	1006.15	1006.15	0.00	1006.15	0.00	0.00	0.00	0.00
ETHANE	lbmol/hr	16.81	16.81	0.00	16.81	0.00	0.00	0.00	0.00
ETHYLENE	lbmol/hr	79.75	79.75	0.00	79.75	0.00	0.00	0.00	0.00
PROPANE	lbmol/hr	4.31	4.31	0.00	4.31	0.00	0.00	0.00	0.00
PROPYLENE	lbmol/hr	12.50	12.50	0.00	12.50	0.00	0.00	0.00	0.00
BENZENE	lbmol/hr	15.09	15.09	0.00	15.09	0.00	0.00	0.00	0.00
TOLUENE	lbmol/hr	6.90	6.90	0.00	6.90	0.00	0.00	0.00	0.00
HYDROGEN SULFIDE	lbmol/hr	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HYDROGEN CYANIDE	lbmol/hr	0.02	0.01	0.00	0.01	0.00	0.00	0.00	0.00
AMMONIA	lbmol/hr	1.72	1.72	0.00	1.72	0.00	0.00	0.00	0.00
SULFUR DIOXIDE	lbmol/hr	1.72	0.86	0.00	0.86	0.00	0.00	0.00	0.00
CARBONYL SULFIDE	lbmol/hr	0.43	0.22	0.00	0.22	0.00	0.00	0.00	0.00
WATER	lbmol/hr	11.32	11.32	0.00	11.32	30.93	30.93	4.54	26.40
NAPHTHALENE	lbmol/hr	0.43	0.00	0.00	0.00	0.00	0.00	0.00	0.00
HYDROGEN CHLORID	lbmol/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NITRIC OXIDE	lbmol/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
XYLENE	lbmol/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	lbmol/hr	4184.27	4182.73	1908.87	2273.86	1893.40	1893.40	1866.99	26.41

J10556 -3401 38MMSCFD PSA AKER KVAERNER FOR ENVIRES

[illegible]

PRESSURE SWING ADSORPTION UNIT

FEED COALESCER DATA SHEET

EQUIPMENT TAG No.

V - 212

0		Issued for Design Study	CPG					
REV	DATE	DESCRIPTION	Originator.	Checked.	Mech.	Process	Instr.	Proj. Man.
PROJECT TITLE: PRESSURE SWING ADSORPTION UNIT			SPECIFICATION NUMBER: J10556-2205					
CLIENT: AKER KVAERNER FOR ENVIRE			JOB NO.: J10556					
			SHEET 1 OF 3					

Specification Sheet For		Specification Number:	
FEED COALESCER		J10556 - 22056	
Rev. Service : Coke Oven Gas		Sheet 2 of 3	
		Vendor to complete/confirm items marked *	
1			
2	TAG No.	V -212	
3	Fluid	Coke Oven Gas (Hydrogen rich Feed Gas)	
4	Quantity	1	
5	Body diameter	*	
6	Length	*	
7			
8	Operating Conditions		
9	Flowrate	lb/hr	44535
10	Inlet pressure	psig	429
11	Inlet temperature	°F	100.000
12	Allowable pressure drop	psig	0.15 (clean) 0.7(dirty) across filter
13	Molecular weight		10.640
14	density @ operating conditions vap/liq	lb/ft³	0.75 / 53.86
15	viscosity @ operating conditions vap/liq	cP	0.01 / 0.57
16			
17	Design Data		
18	Design Temperature	°F	150
19	Design Pressure	psig	460
20	Corrosion Allowance	in	0.125
21	Radiography		100%
22	Code		ASME VIII Div 1 (latest)
23	Stress relief		to code
24	Flange finish		3.2 - 3.6 µm
25			
26	Line size and specification		8" NB ANSI 300#
27			
28	Connections		
29	Inlet		8" RF WN 300#
30	Outlet		8" RF WN 300#
31	Vent		1" RF WN 300#
32	Drain		1" RF WN 300#
33	Instruments		1.5" RF WN 300#
34	Access Manway		20" RF WN 300#
35			
36	Instrumentation requirement		
37	Differential pressure Indicator		N/A
38	Differential pressure switch		N/A
39	Other		Level Gauge (2 off connections)
40			
41	Filtration efficiency required		99.9% removal of 5 microns and above
42			Liquid droplets and solid particles
43	Materials of construction		
44	Filter housing		ASTM A516 Gr 60
45	Filter Material		Glass Media
46	Gaskets		Non Asbestos
47	Bolting		ASTM A194-2H
48			
49	Coatings		
50	Internal		*
51	External		*
52			
53	closure / Access		Blind Flange with davit and lugs
54			
55			
56			
57			
58			
59			
60			
61			
62	Manufacturer		
63	Model No.		
64			

Specification Sheet For		Specification Number:
FEED COALESCER		J10556-2205
Rev. Service : Coke Oven Gas		Sheet 2 of 3
		Vendor to complete/confirm items marked *
1	<p>NOTES:</p> <ol style="list-style-type: none"> 1. All flanges to be raised face with 3.2 to 6.3µm finish suitable for spirally wound gaskets 2. Skirt to vessel attachment weld to be tested 100% MPI and ground smooth. 3. Vessel is in hydrogen service. Low hydrogen welding process and full penetration welds required. 4. fabricator shall supply design calculations for all items even if detailed by Natco UK. 5. Post weld heat treatment shall be carried out after complete vessel fabrication as required by code. 6. Supplier to consider operation when flowrate is increased to 120% of normal flow and 30% turndown of normal flow. 7. Size of Feed coalescer to be confirmed during detailed engineering. 	
2		
3		
4		
5		
6		
7		
8		
9		
10		
11		
12		
13		
14		
15		
16		
17		
18		
19		
20		
21		
22		
23		
24		
25		
26		
27		
28		
29		
30		
31		
32		
33		
34		
35		
36		
37		
38		
39		
40		
41		
42		
43		
44		
45		
46		
47		
48		
49		
50		
51		
52		
53		
54		
55		
56		
57		
58		
59		
60		
61		
62		
63		
64

PRESSURE SWING ADSORPTION UNIT

VESSEL DATA SHEET PRE-PURIFICATION ADSORBERS

EQUIPMENT TAG No.

V-220 A/B

0		Issued for Design Study	CPG					
REV	DATE	DESCRIPTION	Originator.	Checked.	Mech.	Process	Instr.	Proj. Man.
PROJECT TITLE: PRESSURE SWING ADSORPTION UNIT			SPECIFICATION NUMBER: J10556-2204					
CLIENT: AKER KVAERNER FOR ENVIRE			JOB NO.: J10556					
			SHEET 1 OF 3					

NATCO UK LIMITED

SPECIFICATION SHEET For				Specification Number:																																									
VESSEL DATA SHEET				J10556-2204																																									
				Sheet: 2 of 3																																									
Rev				Vendor to complete all information marked *																																									
SERVICE:																																													
2 DESCRIPTION																																													
3 Type VERTICAL PACKED BED																																													
4 Vessel Unit No. V-220																																													
5 Vessel Tag No. V-220 A/B																																													
6 Quantity 2 - OFF																																													
7 Vessel - Diameter 9' 9" ft I.D.																																													
8 - Length 36' ft Tan./Tan.																																													
9 Head Type 2:1 Semi Ellipsoidal																																													
10 DESIGN DATA																																													
11 Operating Pressure 413 psig																																													
12 Design Pressure 460 psig																																													
13 Design Temperature 150 °F / 40 °F																																													
14 Max. Allow. Press. * psig (cold & new)																																													
15 Corrosion Allowance 1.5 mm SHELL & HEADS																																													
16 Codes ASME VIII Div 1																																													
17 Fluid Hydrogen (Hydrogen Service)																																													
18 Radiography 100%																																													
19 Joint Efficiency 1																																													
20 Stress Relief To Code																																													
21 Flange Finish RF 3.2 to 6.3 µm																																													
22																																													
23																																													
24																																													
25 MATERIALS																																													
26 Shell ASTM A516 Gr 70																																													
27 Heads ASTM A516 Gr 70																																													
28 Nozzle Necks ASTM A350 LF2																																													
29 Nozzle Flanges ASTM A350 LF2																																													
30 Internal Parts See below																																													
31 External Studs/Nuts ASTM A193-B7/ ASTM A194-2H																																													
32 Internal Bolts/Nuts ASTM A193-B8/ASTM A194 Gr 4																																													
33 Clips (External) *																																													
34 Gaskets (External) Spiral Wound 3mm																																													
35 Skirt/Leg Support *																																													
36																																													
37																																													
38 INTERNALS																																													
39 Baffel Screens 316 SS																																													
40																																													
41																																													
42 ADSORBENT QUANTITIES																																													
43 The vessel shall be filled with adsorbent, quantities which																																													
44 will be finalised at the detailed engineering stage.																																													
45																																													
46 ACCESSORIES																																													
47 Item Ref. Details																																													
48 Manway N3 Via Adsorbent Fill nozzle/manway																																													
49 Platform Clips - No																																													
50 Ladder Clips - No																																													
51 Lifting Lugs 4 Yes																																													
52 Earth Boss 4 Yes																																													
53 Trace Heating - No																																													
54																																													
55 COATINGS Internal External																																													
56 Preparation Note 5 Note 6																																													
57 Finish Note 5 Note 6																																													
<table border="1" style="width: 100%; border-collapse: collapse;"> <thead> <tr> <th>Rev</th> <th>Ref</th> <th>Service</th> <th>Qty</th> <th>Size</th> <th>Rating & Type</th> </tr> </thead> <tbody> <tr> <td></td> <td>N1</td> <td>FEED</td> <td>1</td> <td>8"</td> <td>ANSI 300#RFLWN</td> </tr> <tr> <td></td> <td>N2</td> <td>PRODUCT</td> <td>1</td> <td>8"</td> <td>ANSI 300#RFLWN</td> </tr> <tr> <td></td> <td>N3</td> <td>ADSORBENT FILL</td> <td>1</td> <td>24"</td> <td>ANSI 300#RFLWN</td> </tr> <tr> <td></td> <td>N4</td> <td>SKIRT VENT</td> <td>1</td> <td>*</td> <td></td> </tr> <tr> <td></td> <td>N5</td> <td>SKIRT ACCESS</td> <td>1</td> <td>*</td> <td></td> </tr> <tr> <td></td> <td>N6</td> <td>PIPE SLEVE</td> <td>1</td> <td>*</td> <td></td> </tr> </tbody> </table>				Rev	Ref	Service	Qty	Size	Rating & Type		N1	FEED	1	8"	ANSI 300#RFLWN		N2	PRODUCT	1	8"	ANSI 300#RFLWN		N3	ADSORBENT FILL	1	24"	ANSI 300#RFLWN		N4	SKIRT VENT	1	*			N5	SKIRT ACCESS	1	*			N6	PIPE SLEVE	1	*	
Rev	Ref	Service	Qty	Size	Rating & Type																																								
	N1	FEED	1	8"	ANSI 300#RFLWN																																								
	N2	PRODUCT	1	8"	ANSI 300#RFLWN																																								
	N3	ADSORBENT FILL	1	24"	ANSI 300#RFLWN																																								
	N4	SKIRT VENT	1	*																																									
	N5	SKIRT ACCESS	1	*																																									
	N6	PIPE SLEVE	1	*																																									
OTHER ITEMS:																																													
Ref Description																																													
1. Vessel support details by vessel supplier.																																													
2. Deleted																																													
3. Deleted																																													
/bolts/gasket by vessel vendor.																																													
4. Vessel lugs: 2-off lifting lugs, 1-off tailing lug.																																													
5. Earthing lugs: 2-off.																																													
6. Adsorbent fill flange: 2-off lifting lugs.																																													
Approx. Weights - Empty * kg																																													
- Hydrotest * kg																																													

NATCO UK LIMITED

SPECIFICATION SHEET For VESSEL DATA SHEET		Specification Number: P10556-2203
SERVICE:		Sheet: 3 of 3
Rev	Vendor to complete all information marked *	

1	<p>NOTES:</p> <p>1 Vessel is in Hydrogen Service. Low hydrogen Welding process and full penetration welds required.</p> <p>2 Internal welds to be ground flush.</p> <p>3 No compensation pads allowed.</p> <p>4 All welds to have smooth finish</p> <p>5 <u>Internal Finish</u> To be recommended by Vendor.</p> <p>6 <u>External Finish</u> To be recommended by Vendor.</p>
2	
3	
4	
5	
6	
7	
8	
9	
10	
11	
12	
13	
14	
15	
16	
17	
18	
19	
20	
21	
22	
23	
24	
25	
26	
27	
28	
29	
30	
31	
32	
33	
34	
35	
36	
37	
38	
39	
40	
41	
42	
43	
44	
45	
46	
47	
48	
49	
50	
51	
52	
53	
54	
55	
56	
57

PRESSURE SWING ADSORPTION UNIT

VESSEL DATA SHEET

ADSORBERS

EQUIPMENT TAG No.

V-201 A/B/C/D/E

0		Issued for Design Study	CPG					
REV	DATE	DESCRIPTION	Originator.	Checked.	Mech.	Process	Instr.	Proj. Man.
PROJECT TITLE:			SPECIFICATION NUMBER:					
PRESSURE SWING ADSORPTION UNIT			J10556-2201					
CLIENT:			JOB NO.:					
AKER KVAERNER FOR ENVIRES			J10556					
			SHEET 1 OF 3					

SPECIFICATION SHEET For				Specification Number:					
VESSEL DATA SHEET				J10556-2201					
				Sheet: 2 of 3					
Rev	SERVICE:			Vendor to complete all information marked *					
1									
2	DESCRIPTION								
3	Type	VERTICAL PACKED BED							
4	Vessel Unit No.	V-201							
5	Vessel Tag No.	V-201 A/B/C/D/E							
6	Quantity	5 - OFF							
7	Vessel - Diameter	12' 5" ft I.D.							
8	- Length	24' 9" ft Tan./Tan.							
9	Head Type	2:1 Semi Ellipsoidal							
10	DESIGN DATA								
11	Operating Pressure	406 - 10 - 406 psig							
12		every 12.5 minutes							
13	Design No. of Cycles	841000 (20 Years)							
14	Cycle Time (Mins.)	12.5							
15	Design Pressure	460 psig							
16	Design Temperature	150 °F / 40 °F							
17	Max. Allow. Press.	* psig (cold & new)							
18	Corrosion Allowance	1.5 mm SHELL & HEADS							
19	Codes	ASME VIII Div 2							
20		-							
21		-							
22	Fluid	Hydrogen (Hydrogen Service)							
23	Radiography	100%							
24	Joint Efficiency	1							
25	Stress Relief	To Code							
26	Flange Finish	RF 3.2 to 6.3 µm							
27	MATERIALS								
28	Shell	ASTM A516 Gr 70							
29	Heads	ASTM A516 Gr 70							
30	Nozzle Necks	ASTM A350 LF2							
31	Nozzle Flanges	ASTM A350 LF2							
32	Internal Parts	See below							
33	External Studs/Nuts	ASTM A193-B7/ ASTM A194-2H							
34	Internal Bolts/Nuts	ASTM A193-B8/ASTM A194 Gr 4							
35	Clips (External)	*							
36	Gaskets (External)	Spiral Wound 3mm							
37	Skirt/Leg Support	*							
38	INTERNALS								
39	Inlet/Outlet Screens	316 SS							
40									
41									
42	ADSORBENT QUANTITIES			NOZZLE SCHEDULE					
43	The vessel shall be filled with adsorbent, quantities which will be finalised at the detailed engineering stage.			Rev	Ref	Service	Qty	Size	Rating & Type
44					N1	FEED	1	12"	ANSI 300#RFLWN
45					N2	PRODUCT	1	8"	ANSI 300#RFLWN
					N3	ADSORBENT FILL	1	24"	ANSI 300#RFLWN
					N4	SKIRT VENT	1	*	
					N5	SKIRT ACCESS	1	*	
					N6	PIPE SLEEVE	1	*	
46	ACCESSORIES			OTHER ITEMS:					
47	Item	Ref.	Details	Ref Description					
48	Manway	N3	Via Adsorbent Fill nozzle/manway	1. Vessel support details by vessel supplier.					
49	Platform Clips	-	No	2. Pipe supports, one clip & pad & one coupling/vessel.					
50	Ladder Clips	-	No	3. Screens will be free issued by Axsia internal flange /bolts/gasket by vessel vendor.					
51	Lifting Lugs	4	Yes	4. Vessel lugs: 2-off lifting lugs, 1-off tailing lug.					
52	Earth Boss	4	Yes	5. Earthing lugs: 2-off.					
53	Trace Heating	-	No	6. Adsorbent fill flange: 2-off lifting lugs.					
54									
55	COATINGS								
		Internal	External						
56	Preparation	Note 5	Note 6	Approx. Weights		- Empty	*	kg	
57	Finish	Note 5	Note 6			- Hydrotest	*	kg	

NATCO UK LIMITED

SPECIFICATION SHEET For		Specification Number:
VESSEL DATA SHEET		J10556 - 2201
		Sheet: 3 of 3
Rev	SERVICE:	Vendor to complete all information marked *
1	<p>NOTES:</p> <p>1 Vessel is in Hydrogen Service. Low hydrogen Welding process and full penetration welds required.</p> <p>2 Internal welds to be ground flush.</p> <p>3 No compensation pads allowed.</p> <p>4 All welds to have smooth finish</p> <p>5 <u>Internal Finish</u> To be recommended by Vendor.</p> <p>6 <u>External Finish</u> To be recommended by Vendor.</p>	
2		
3		
4		
5		
6		
7		
8		
9		
10		
11		
12		
13		
14		
15		
16		
17		
18		
19		
20		
21		
22		
23		
24		
25		
26		
27		
28		
29		
30		
31		
32		
33		
34		
35		
36		
37		
38		
39		
40		
41		
42		
43		
44		
45		
46		
47		
48		
49		
50		
51		
52		
53		
54		
55		
56		
57

PRESSURE SWING ADSORPTION UNIT

VESSEL DATA SHEET

OFFGAS DRUM

EQUIPMENT TAG No.

V-203A/B

0		Issued for Design Study	CPG					
REV	DATE	DESCRIPTION	Originator.	Checked.	Mech.	Process	Instr.	Proj. Man.
PROJECT TITLE:			SPECIFICATION NUMBER:					
PRESSURE SWING ADSORPTION UNIT			J10556-2202					
CLIENT:			JOB NO.:					
AKER KVAERNER FOR ENVIRE			J10556					
			SHEET 1 OF 3					

Specification Number:

J10556-2202

Sheet: 2 of 3

Rev	SERVICE:
-----	----------

Vendor to complete all information marked *

The diagram shows a vertical vessel with a hemispherical head and a hemispherical skirt. The vessel is supported by a base. Various nozzles and supports are indicated by numbered callouts:

- 1: Vessel support details by vessel supplier.
- 2: Pipe supports, one clip & pad & one coupling/vessel.
- 3: Vessel lugs; 2-off lifting lugs, 1-off tailing lug.
- 4: Earthing lugs, 2-off.
- N1: INLET
- N2: OUTLET
- N3: DRAIN
- N4: SKIRT VENT
- N5: SKIRT ACCESS
- N6: PIPE SLEEVE
- N7: MANWAY

The vessel is labeled "Tan" (Tanks) at the top and bottom. The nozzle schedule is as follows:

Rev	Ref	Service	Qty	Size	Rating & Type
	N1	INLET	1	12"	ANSI 150#RFWN
	N2	OUTLET	1	12"	ANSI 150#RFWN
	N3	DRAIN	1	2"	ANSI 150#RFWN
	N4	SKIRT VENT	1	*	
	N5	SKIRT ACCESS	1	*	
	N6	PIPE SLEEVE	1	*	
	N7	MANWAY	1	24"	c/w blind and davit

OTHER ITEMS:

Ref	Description
1.	Vessel support details by vessel supplier.
2.	Pipe supports, one clip & pad & one coupling/vessel.
3.	Vessel lugs; 2-off lifting lugs, 1-off tailing lug.
4.	Earthing lugs, 2-off.

Approx. Weights

	- Empty	- Hydrotest
	*	*

NATCO UK LIMITED

SPECIFICATION SHEET For		Specification Number:
VESSEL DATA SHEET		J10556 - 2202
Sheet: 3 of 3		
Rev	SERVICE:	Vendor to complete all information marked *
1		
2		
3		
4		
5		
6		
7		
8		
9		
10		
11		
12		
13		
14		
15		
16		
17		
18		
19		
20		
21		
22		
23		
24		
25		
26		
27		
28		
29		
30		
31		
32		
33		
34		
35		
36		
37		
38		
39		
40		
41		
42		
43		
44		
45		
46		
47		
48		
49		
50		
51		
52		
53		
54		
55		
56		
57		

NOTES:

- 1 Vessel is in Hydrogen Service. Low hydrogen Welding process and full penetration welds required.
- 2 Internal welds to be ground flush.
- 3 No compensation pads allowed.
- 4 All welds to have smooth finish
- 5 Internal Finish
To be recommended by Vendor.
- 6 External Finish
To be recommended by Vendor.

PRESSURE SWING ADSORPTION UNIT

VESSEL DATA SHEET

DEOXO UNIT

EQUIPMENT TAG No.

V-206

0		Issued for Design Study	CPG					
REV	DATE	DESCRIPTION	Originator.	Checked.	Mech.	Process	Instr.	Proj. Man.
PROJECT TITLE:			SPECIFICATION NUMBER:					
PRESSURE SWING ADSORPTION UNIT			J10556-2203					
CLIENT:			JOB NO.:					
AKER KVAERNER FOR ENVIRES			P10556					
			SHEET 1 OF 3					

J10556 -2203 Deoxo Vessel Rev 0

NATCO UK LIMITED

SPECIFICATION SHEET For VESSEL DATA SHEET		Specification Number: J10556-2203 Sheet: 3 of 3
Rev	SERVICE:	Vendor to complete all information marked *

1	
2	
3	
4	
5	
6	
7	
8	
9	
10	
11	
12	
13	
14	
15	
16	
17	
18	
19	
20	
21	
22	
23	
24	
25	
26	
27	
28	
29	
30	
31	
32	
33	
34	
35	
36	
37	
38	
39	
40	
41	
42	
43	
44	
45	
46	
47	
48	
49	
50	
51	
52	
53	
54	
55	
56	
57	

NOTES:

1 Vessel is in Hydrogen Service. Low hydrogen Welding process and full penetration welds required.

2 Internal welds to be ground flush.

3 No compensation pads allowed.

4 All welds to have smooth finish

5 Internal Finish
 To be recommended by Vendor.

6 External Finish
 To be recommended by Vendor.

7 **Catalyst Specification**

Catalyst	Type D2 (0.2% Pd on spheres
Catalyst dimensions	2.5mm spheres
Catalyst space velocity	20,000 hr ⁻¹
Catalyst bed volume	960 litres
Catalyst density	approx 0.9kg/litre

Normal catalyst guarantee is for 2 years continuous

Analysis: less than 5 ppmv O₂

NATCO UK LIMITED

PRESSURE SWING ADSORPTION UNIT

EQUIPMENT DATA SHEET SHELL AND TUBE HEAT EXCHANGER

EQUIPMENT TAG No.

E-207

0		Issued for Design Study	CPG					
REV	DATE	DESCRIPTION	Originator	Checked	Mech.	Process	Instr.	Proj. Man.
PROJECT TITLE: PRESSURE SWING ADSORPTION UNIT			SPECIFICATION NUMBER: J10556 - 3301					
CLIENT: AKER KVAERNER FOR ENVIRES			JOB NO.: J10556 - 3301					
			SHEET 1 OF 3					

Rev											
SPECIFICATION SHEET FOR:					SPECIFICATION NUMBER						
SHELL AND TUBE HEAT EXCHANGER					J10556 - 3301						
					Sht 2 of 3						
Rev.	ITEM NO.	E-207			Vendor to complete all items marked *						
1	Service:	Product Cooler			Orientation:		Horizontal				
2	Size: (ID/L)	1ft '5" / 19'ft .0"	Type:	BEU	Connected:		1	Parallel	1 Series		
3	Surf/Unit:	828 ft ² (Gross)	Shells/Unit:	1	Surf/Shell:		836 ft ² (Effective)				
4	PERFORMANCE OF ONE UNIT							Units Required:			
5	Fluid Allocation			SHELL SIDE		TUBE SIDE					
6	Fluid Name:			Service Water (H ₂ O)		Hydrogen (H ₂) Water (H ₂ O)					
7	Fluid Quantity TOTAL			lb/hr	153,572	4,312					
8	Vapor:			lb/hr	-	-	4,241	4,241			
9	Liquid:			lb/hr	-	-	-	-			
10	Steam:			lb/hr	-	-	-	-			
11	Water:			lb/hr	153,572	153,572	71	71			
12	Noncondensables:			lb/hr	-	-	-	-			
13	Operating Temperature, In/Out:			°F	85.0	110	354	100			
14	Density L/V:			lb/ft ³	61.96 (Avg)		- / 0.11	62.9 / 0.14			
15	Viscosity, L/V:			cP	0.612		- / 0.01	0.68 / 0.01			
16	Molecular Weight of Vapor:				-	-	2.2773	2.2773			
17	Molecular Weight of Noncondensables:				-	-	-	-			
18	Specific Heat Capacity, L/V:			Btu/lb. °F	1.03 (Avg)		/	/			
19	Thermal Conductivity, L/V:			Btu/hr.ft. °F	0.368 (Avg)		- / 0.14	0.36 / 0.110			
20	Latent Heat:			Btu/lb @ °F	/		/				
21	Inlet Pressure (Operating):			psig	75		396				
22	Velocity:			ft/s	*		*				
23	Pressure Drop, Allow/Calc:			psi	5	/ VTA *	5	/ VTA *			
24	Fouling Resistance, Min:			ft ² .°F/Btu	0.002		0.002				
25	Heat Exchanged:			3831717 / *	Btu/hr	MTD:	51.93 °F (Corrected)				
26	Transfer Rate Service: Dirty:			105.87 / *	Btu/hr.ft ² .°F	Clean:	89.05/* Btu/hr.ft ² .°F				
27	CONSTRUCTION OF ONE SHELL					SKETCH (Bundle/Nozzle Orientation)					
28				SHELL SIDE	TUBE SIDE						
29	Design/Test Pressure:	psig	270 /	Code	460 /	Code					
30	Design Temperature/(MDMT):	°F	580 /	*	580 /	*					
31	No. Passes per Shell:		1		4						
32	Corrosion Allowance:	inches	0.0625		0.0625						
33	Connections	In	NPS	4" 300#RF	6" 300#RF						
34	Size, Rating &	Out	NPS	4" 300#RF	6" 300#RF						
35	Facing:	Intermediate	NPS	*	*						
36	Tube No.:	*	OD.:	* in	Thk.:	* in	Length:	* in	Pitch:	* in	
37	Tube Type:	Plain		Tube Material:	Steel SA-214						
38	Shell Material:	Carbon Steel		I.D.:	* in	O.D.:	* in	Shell Cover Material:	Carbon Steel		
39	Channel or Bonnet Material:	Steel SA-214		Channel Cover Material:	*						
40	Stationary Tubesheet Material:	Steel SA-214		Floating Tubesheet Material:	*						
41	Floating Head Cover Material:	*		Impingement Protection:			Yes/No	Type:	N/A		
42	Baffles - Cross Material:	*	Type:	Segm.	%Cut	*	Spacing/c/c	* in	Inlet:	- in	
43	Long Baffles Material:	*		Seal Type:	Sealing Strip						
44	Tube Supports Material:	*		U-bend:	*	Type:	*				
45	Bypass Seal Arrangement:	*		Tube-Tubesheet Join:	*						
46	Expansion Joint:	Yes/No	Material:	*	Type:	*	Avg. Metal Temp.:	Shell:	* °F	Tube:	* °F
47	pV ² - Shell Entrance:	* lb/ft.s ²		pV ² - Bundle Entrance:	* lb/ft.s ²		pV ² - Bundle Exit:	* lb/ft.s ²			
48	Gaskets:	Shell Side:	Non-Asbestos		Tube Side:	Non-Asbestos		Floating Head:	*		
49	Code Requirements:	ASME, Sec VIII, Div 1 (latest)		Code Stamp:	Yes/No	Natl. Board Reg.:	Yes/No	TEMA Class	R		
50	Weight/Shell:	Shipping:	* lb		Filled with Water:	* lb		Bundle:	* lb		
51	Post Weld Heat Treatm	Shell:	Yes/No	Channel:	Yes/No	Floating Head:	Yes/No	U-Bends	Yes/No		

Rev	
SPECIFICATION SHEET FOR: <div style="text-align: center; font-weight: bold; font-size: 1.2em;">SHELL AND TUBE HEAT EXCHANGER</div>	SPECIFICATION NUMBER <div style="text-align: center; font-weight: bold; font-size: 1.2em;">J10556 - 3301</div>
Sheet 3 of 3	

Rev.	NOTES
1	
2	1 Required unit: 1 (One).
3	2 10% extra duty correlates to 10% additional surface area, preferably 10% (min) additional tube length
4	3 Design Basis: ASME Section VIII Div. 1 Latest Edition.
5	4 Vendor to design and provide lifting lugs for unit, channels, and channel covers
6	5 Flange face finish to be : 125 - 250 Ra.
7	6 All flange and threaded connections will be protected with an anti-rust material. (not painted).
8	7 Shop Hydrostatic Test Pressure: Per ASME Code Sec. VIII, Div 1. Vendor to submit hydrotest chart.
9	8 Radiographing: per ASME Code - minimum Spot Radiography.
10	9 Joint Efficiency: per ASME Code.
11	10 All components welded directly to the pressure boundary shall be of the same material as the pressure boundary
12	11 Nozzle Schedule by vendor.
13	12 The fabrication and acceptance of the vessel will be subject to inspection by NATCO and/or client representative
14	13 All attachment and reinforcement plates shall be continuous welding.
15	
16	
17	
18	
19	
20	
21	
22	
23	
24	
25	
26	
27	
28	
29	
30	
31	
32	
33	
34	
35	
36	
37	
38	
39	
40	
41	
42	
43	
44	
45	
46	
47	
48	
49	
50	
51

PRESSURE SWING ADSORPTION UNIT

PRODUCT SEPARATOR DATA SHEET

EQUIPMENT TAG No.

V - 213

0		Issued for Design Study	CPG						
REV	DATE	DESCRIPTION	Originator.	Checked.	Mech.	Process	Instr.	Proj. Man.	
PROJECT TITLE: PRESSURE SWING ADSORPTION UNIT			SPECIFICATION NUMBER: J10556-2206						
CLIENT: AKER KVAERNER FOR ENVIRES			JOB NO.: J10556						
			SHEET 1 OF 3						

NATCO UK LIMITED

SPECIFICATION SHEET For VESSEL DATA SHEET				Specification Number: J10556-2201																																																															
				Sheet: 2 of 3																																																															
Rev	SERVICE:			Vendor to complete all information marked *																																																															
1																																																																			
2	DESCRIPTION																																																																		
3	Type	VERTICAL																																																																	
4	Vessel Unit No.	V-213																																																																	
5	Vessel Tag No.	V-213																																																																	
6	Quantity	1 - OFF																																																																	
7	Vessel - Diameter	1' 4" ft I.D.																																																																	
8	- Length	6' 0" ft Tan./Tan.																																																																	
9	Head Type	2:1 Semi Ellipsoidal																																																																	
10	DESIGN DATA																																																																		
11	Operating Pressure	391 psig																																																																	
12	Operating Temperature	100°F																																																																	
13	Design Pressure	460 psig																																																																	
14	Design Temperature	150 °F / 40 °F																																																																	
15	Max. Allow. Press.	* psig (cold & new)																																																																	
16	Corrosion Allowance	3 mm SHELL & HEADS																																																																	
17	Codes	ASME VIII Div 1																																																																	
18	Fluid	Hydrogen (Hydrogen Service)																																																																	
19	Radiography	100%																																																																	
20	Joint Efficiency	1																																																																	
21	Stress Relief	To Code																																																																	
22	Flange Finish	RF 3.2 to 6.3 μm																																																																	
23	MATERIALS																																																																		
24																																																																			
25	Shell	ASTM A516 Gr 70																																																																	
26	Heads	ASTM A516 Gr 70																																																																	
27	Nozzle Necks	ASTM A350 LF2																																																																	
28	Nozzle Flanges	ASTM A350 LF2																																																																	
29	Internal Parts	See below																																																																	
30	External Studs/Nuts	ASTM A193-B7/ ASTM A194-2H																																																																	
31	Internal Bolts/Nuts	ASTM A193-B8/ASTM A194 Gr 4																																																																	
32	Clips (External)	*																																																																	
33	Gaskets (External)	Spiral Wound 3mm																																																																	
34	Skirt/Leg Support	*																																																																	
35																																																																			
36																																																																			
37																																																																			
38	INTERNALS																																																																		
39	TP vane Mist pad	316 SS																																																																	
40																																																																			
41																																																																			
42																																																																			
43																																																																			
44																																																																			
45																																																																			
46	ACCESSORIES																																																																		
47	Item	Ref.	Details																																																																
48	Manway	N3	Via Adsorbent Fill nozzle/manway																																																																
49	Platform Clips	-	No																																																																
50	Ladder Clips	-	No																																																																
51	Lifting Lugs	4	Yes																																																																
52	Earth Boss	4	Yes																																																																
53	Trace Heating	-	No																																																																
54																																																																			
55	COATINGS																																																																		
56	Preparation	Note 5	Note 6																																																																
57	Finish	Note 5	Note 6																																																																
				NOZZLE SCHEDULE <table border="1" style="width:100%; border-collapse: collapse;"> <thead> <tr> <th>Rev</th> <th>Ref</th> <th>Service</th> <th>Qty</th> <th>Size</th> <th>Rating & Type</th> </tr> </thead> <tbody> <tr> <td></td> <td>N1</td> <td>FEED</td> <td>1</td> <td>6"</td> <td>ANSI 300#RFLWN</td> </tr> <tr> <td></td> <td>N2</td> <td>HYDROGEN PRODUCT</td> <td>1</td> <td>6"</td> <td>ANSI 300#RFLWN</td> </tr> <tr> <td></td> <td>N3</td> <td>WATER OUTLET</td> <td>1</td> <td>2"</td> <td>ANSI 300#RFLWN</td> </tr> <tr> <td></td> <td>N4</td> <td>SKIRT VENT</td> <td>1</td> <td>*</td> <td></td> </tr> <tr> <td></td> <td>N5</td> <td>SKIRT ACCESS</td> <td>1</td> <td>*</td> <td></td> </tr> <tr> <td></td> <td>N6</td> <td>PIPE SLEEVE</td> <td>1</td> <td>*</td> <td></td> </tr> <tr> <td></td> <td>N7</td> <td>MANWAY</td> <td>1</td> <td>*</td> <td></td> </tr> <tr> <td></td> <td>N8</td> <td>LSH</td> <td>1</td> <td>2"</td> <td></td> </tr> <tr> <td></td> <td>N9</td> <td>LEVEL CONTROLLER</td> <td>1</td> <td>2"</td> <td></td> </tr> </tbody> </table>				Rev	Ref	Service	Qty	Size	Rating & Type		N1	FEED	1	6"	ANSI 300#RFLWN		N2	HYDROGEN PRODUCT	1	6"	ANSI 300#RFLWN		N3	WATER OUTLET	1	2"	ANSI 300#RFLWN		N4	SKIRT VENT	1	*			N5	SKIRT ACCESS	1	*			N6	PIPE SLEEVE	1	*			N7	MANWAY	1	*			N8	LSH	1	2"			N9	LEVEL CONTROLLER	1	2"	
Rev	Ref	Service	Qty	Size	Rating & Type																																																														
	N1	FEED	1	6"	ANSI 300#RFLWN																																																														
	N2	HYDROGEN PRODUCT	1	6"	ANSI 300#RFLWN																																																														
	N3	WATER OUTLET	1	2"	ANSI 300#RFLWN																																																														
	N4	SKIRT VENT	1	*																																																															
	N5	SKIRT ACCESS	1	*																																																															
	N6	PIPE SLEEVE	1	*																																																															
	N7	MANWAY	1	*																																																															
	N8	LSH	1	2"																																																															
	N9	LEVEL CONTROLLER	1	2"																																																															
				OTHER ITEMS: <table border="1" style="width:100%; border-collapse: collapse;"> <thead> <tr> <th>Ref</th> <th>Description</th> </tr> </thead> <tbody> <tr> <td>1.</td> <td>Vessel support details by vessel supplier.</td> </tr> <tr> <td>2.</td> <td>Deleted.</td> </tr> <tr> <td>3.</td> <td>Screens will be free issued by Axsia internal flange /bolts/gasket by vessel vendor.</td> </tr> <tr> <td>4.</td> <td>Vessel lugs: 2-off lifting lugs, 1-off tailing lug.</td> </tr> <tr> <td>5.</td> <td>Earthing lugs: 2-off.</td> </tr> </tbody> </table>				Ref	Description	1.	Vessel support details by vessel supplier.	2.	Deleted.	3.	Screens will be free issued by Axsia internal flange /bolts/gasket by vessel vendor.	4.	Vessel lugs: 2-off lifting lugs, 1-off tailing lug.	5.	Earthing lugs: 2-off.																																																
Ref	Description																																																																		
1.	Vessel support details by vessel supplier.																																																																		
2.	Deleted.																																																																		
3.	Screens will be free issued by Axsia internal flange /bolts/gasket by vessel vendor.																																																																		
4.	Vessel lugs: 2-off lifting lugs, 1-off tailing lug.																																																																		
5.	Earthing lugs: 2-off.																																																																		
				<table border="1" style="width:100%; border-collapse: collapse;"> <thead> <tr> <th colspan="2">Approx. Weights</th> <th></th> <th></th> </tr> </thead> <tbody> <tr> <td>- Empty</td> <td>*</td> <td></td> <td>kg</td> </tr> <tr> <td>- Hydrotest</td> <td>*</td> <td></td> <td>kg</td> </tr> </tbody> </table>				Approx. Weights				- Empty	*		kg	- Hydrotest	*		kg																																																
Approx. Weights																																																																			
- Empty	*		kg																																																																
- Hydrotest	*		kg																																																																

NATCO UK LIMITED

SPECIFICATION SHEET For VESSEL DATA SHEET		Specification Number: <div style="text-align: center; font-weight: bold;">J10556 - 2201</div> <div style="text-align: right; font-size: small;">Sheet: 3 of 3</div>
Rev	SERVICE:	Vendor to complete all information marked *

1	<div style="margin-bottom: 10px;">NOTES:</div> <div> <div style="display: flex; margin-bottom: 5px;"> <div style="width: 20px; text-align: center;">1</div> <div>Vessel is in Hydrogen Service. Low hydrogen Welding process and full penetration welds required.</div> </div> <div style="display: flex; margin-bottom: 5px;"> <div style="width: 20px; text-align: center;">2</div> <div>Internal welds to be ground flush.</div> </div> <div style="display: flex; margin-bottom: 5px;"> <div style="width: 20px; text-align: center;">3</div> <div>No compensation pads allowed.</div> </div> <div style="display: flex; margin-bottom: 5px;"> <div style="width: 20px; text-align: center;">4</div> <div>All welds to have smooth finish</div> </div> <div style="display: flex; margin-bottom: 5px;"> <div style="width: 20px; text-align: center;">5</div> <div> <u>Internal Finish</u> To be recommended by Vendor. </div> </div> <div style="display: flex; margin-bottom: 5px;"> <div style="width: 20px; text-align: center;">6</div> <div> <u>External Finish</u> To be recommended by Vendor. </div> </div> <div style="display: flex; margin-bottom: 5px;"> <div style="width: 20px; text-align: center;">7</div> <div> The TP vane mist eliminator will be supplied by NATCO It will be fabricated from 316L stainless steel. It should be manufactured in sections suitable for installation through the 24 " manway. </div> </div> </div>
2	
3	
4	
5	
6	
7	
8	
9	
10	
11	
12	
13	
14	
15	
16	
17	
18	
19	
20	
21	
22	
23	
24	
25	
26	
27	
28	
29	
30	
31	
32	
33	
34	
35	
36	
37	
38	
39	
40	
41	
42	
43	
44	
45	
46	
47	
48	
49	
50	
51	
52	
53	
54	
55	
56	
57	

1.0 DESIGN BASIS

The study is to examine the hydrogen purification unit capacity efficiency and using the following coke oven gas feedstock, as defined by Aker Kvaerner e-mail dated 14/09/2007.

1.1 Feedstock.

Description	Units	Design	Comment
Hydrogen	Volume %, dry	56.31	
Carbon Monoxide	Volume %, dry	5.06	
Carbon Dioxide	Volume %, dry	1.26	Some removal expected with desulfurization process
Oxygen	Volume %,dry	0.78	Max. excursion observed- Dan Howell 14SEP07
Nitrogen	Volume %,dry	6.81	
Methane	Volume %,dry	23.34	
Ethane	Volume %,dry	0.39	
Ethylene	Volume %,dry	1.85	
Propane	Volume %,dry	0.1	
Propylene	Volume %,dry	0.29	
Benzene	Volume %,dry	0.35	Coke plant light oil removal could be improved.
Toluene	Volume %,dry	0.16	
Xylene	Volume %,dry	0.00	
Hydrogen Sulfide	Ppm, dry basis	5	New acid gas system will be specified to meet this level to avoid problems in the PSA
Hydrogen Cyanide	Ppm, dry	5	New acid gas system will be specified to meet this level to avoid problems in the PSA
Ammonia	Volume %,dry	0.04	50% removal experienced in existing COG Desulfurization- Dan Howell 14SEP07
Sulfur Dioxide	Volume %,dry	0.004	Acid gas removal systems are not designed to remove SO ₂ . Only partial removal expected.
Organic Sulfur COS,CS ₂	Volume %,dry	0.01	Only partial removal of organic sulfur expected in Acid Gas removal. Wilson and Wells, pg 268 estimates 20 to 30 grains/100 ft ³ in raw gas and 50% removal in Light Oil Scrubbers

Description	Units	Design	Comment
Water vapor content of COG	Volume	3.21	Saturated at 35 C. Variation depends on total pressure at B/L.
Naphthalene content of COG	Volume %,dry	0.01	Based on inspection of data. 70% removal of Naph in Light Oil Scrubbers and saturation at 35 C. 14SEP07
Particulate content of COG	mg/Nm3,dry	5	Typical particulate loading for gas cleaning systems
HCl	Volume %	0.0001	Assumed. Coke plant has ammonia water scrubbers
NOx	Volume %	0.0001	Assumed based on 'Coal, Coke and Coal Chemicals',
Temperature of COG at Light Oil Scrubber	Degrees C		
COG production	Million scfd	45.4	Based on variation of approx. +- 20% as seen in data from AK Steel Coke Plant (Nov.2006) analyzed by Dan Howell.
Pressure of Coke Oven Gas at inlet to PSA	Psig	400	

1.2 Product.

PSA Outlet (H₂ Product);

Component	Unit	New Feed
Hydrogen	Vol%	99.9
Pressure	Psi (g)	375
Temperature	°C	35

1.3 Offgas.

Pressure	psi (g)	Average 10 Min 5 Design 10
Temperature	°C	35-40

1.4 Utilities.

1.4.1 Nitrogen

Pressure		
Operating	Norm	100 psi(g)
Temperature		
Operating	Norm.	77-81 °F
Quality		
Content		99.5 mol% with max 20ppmv O ₂

1.4.2 Instrument Air

Pressure		
Operating	Min.	90 psi (g)
	Normal	100 psi (g)
	Design	150 psi (g)
Design for actuators		100 psi (g)
Temperature		
Norm		Ambient
Design		150 °F
Quality		Dry, Oil Free and clean
Dew Point		-20.0 °C

1.4.3 Cooling water

Supply pressure	60 psi (g)
Supply Temperature	70°F
Return temperature	90°F
Return Pressure	40 psi (g)

1.4.4 Power

To be advised.

Aspen ICARUS

Project Cost Summary

Project Title: ENVIRES CTL

Project Name: AREA 1

Proj. Location: Ashland, KY

Estimate Date: 17-Dec-07

Scenario Name: COG CONDITIONING

Job No: A0757500

Prep. By: AKTS

Est. Class: Budget-KBASE

Currency: DOLLARS USD

Account	MH	Wage Rate	Labor Cost	Matl Cost	Total Cost	Percentages
(2) Equipment	2,815	51.87	145,991	7,739,300	7,885,291	74.4% of TDC
(3) Piping	8,696	52.47	456,272	769,903	1,226,175	11.6% of TDC
(4) Civil	2,958	47.44	140,349	55,968	196,317	1.9% of TDC
(5) Steel	1,786	50.44	90,062	83,353	173,415	1.6% of TDC
(6) Instruments	4,243	52.13	221,183	345,864	567,046	5.4% of TDC
(7) Electrical	267	50.67	13,538	5,541	19,079	0.2% of TDC
(8) Insulation	5,610	50.79	284,966	144,003	428,970	4.0% of TDC
(9) Paint	1,705	48.55	82,769	13,880	96,649	0.9% of TDC
(labor estimated at 90% productivity)						
Total Direct Field Costs	28,081		1,435,129	9,157,813	10,592,942	100.0% of TDC
	(TDMH)		(TDL)	(TDM)	(TDC)	
CONSUMABLES					86,108	6.0% of TDL
VENDOR REPRESENTATIVES					45,789	0.5% of TDM
EQUIPMENT RENTAL					131,000	9.1% of TDL
OFFICE FIELD STAFF					71,756	5.0% of TDL
CONTRACT FEE					215,269	15.0% of TDL
TEMP CONST FACILITIES					43,054	3.0% of TDL
Indirect Field Costs					592,976	3.0% of TDL
					(IFC)	
Total Field Costs	28,081				11,185,918	72.2% of TIC
	(TFMH)					
FREIGHT					457,891	5.0% of TDM
ESCALATION					317,788	3.0% of TDC
DETAIL ENGINEERING					635,576	6.0% of TDC
PROJ. MGT / PROCUREMENT / H.O. SUPPORT					529,647	5.0% of TDC
CONSTRUCTION MGT.					635,576	6.0% of TDC
START-UP & COMMISSIONING					317,788	3.0% of TDC
CONTINGENCY					1,408,019	10.0% of TIC
Total Non-Field Costs					4,302,286	27.8% of TIC
Estimated Total Installed Cost				(Rounded)	15,490,000	146.2% of TDC
					(TIC)

Aspen ICARUS**Project Cost Summary****Project Title:** ENVIRES CTL**Project Name:** AREA 3A**Proj. Location:** Ashland, KY**Estimate Date:** 17-Dec-07**Scenario Name:** COG SYNGAS PREP**Job No:** A0757500**Prep. By:** AKTS**Est. Class:** Budget-KBASE**Currency:** DOLLARS USD

Account	MH	Wage Rate	Labor Cost	Matl Cost	Total Cost	Percentages
(2) Equipment	2,519	51.74	130,313	4,129,400	4,259,713	88.5% of TDC
(3) Piping	1,754	52.50	92,099	115,842	207,942	4.3% of TDC
(4) Civil	462	47.26	21,830	7,200	29,030	0.6% of TDC
(5) Steel	347	50.43	17,493	16,901	34,394	0.7% of TDC
(6) Instruments	1,799	52.05	93,623	103,079	196,702	4.1% of TDC
(7) Electrical	105	50.77	5,344	1,720	7,064	0.1% of TDC
(8) Insulation	762	50.87	38,771	19,239	58,010	1.2% of TDC
(9) Paint	330	48.41	15,995	3,296	19,291	0.4% of TDC
(labor estimated at 90% productivity)						
Total Direct Field Costs	8,078		415,467	4,396,678	4,812,144	100.0% of TDC
	(TDMH)		(TDL)	(TDM)	(TDC)	
CONSUMABLES					24,928	6.0% of TDL
VENDOR REPRESENTATIVES					21,983	0.5% of TDM
EQUIPMENT RENTAL					50,600	12.2% of TDL
OFFICE FIELD STAFF					20,773	5.0% of TDL
CONTRACT FEE					62,320	15.0% of TDL
TEMP CONST FACILITIES					12,464	3.0% of TDL
Indirect Field Costs					193,069	3.0% of TDL
					(IFC)	
Total Field Costs	8,078				5,005,213	71.8% of TIC
	(TFMH)					
FREIGHT					219,834	5.0% of TDM
ESCALATION					144,364	3.0% of TDC
DETAIL ENGINEERING					288,729	6.0% of TDC
PROJ. MGT / PROCUREMENT / H.O. SUPPORT					240,607	5.0% of TDC
CONSTRUCTION MGT.					288,729	6.0% of TDC
START-UP & COMMISSIONING					144,364	3.0% of TDC
CONTINGENCY					633,184	10.0% of TIC
Total Non-Field Costs					1,959,811	28.1% of TIC
Estimated Total Installed Cost				(Rounded)	6,970,000	144.8% of TDC
					(TIC)	

APPENDIX 7
Comparison of Cobalt and Iron FT Catalyst Systems

**Overall Comparison of Iron and Cobalt based Fischer-Tropsch Catalyst
Systems for the Production of High-Hydrogen Content Coal-Derived
Liquid Transportation Fuels**

Type of Report: Technical
Reporting Period Start Date: March 30, 2006
Reporting Period End Date: March 30, 2011
Principal Authors:
Rafael Espinoza
Kym B. Arcuri
Daniel B. Montoto
Emerging Fuels Technology, Inc.
Date Report Issued: March 2011
DOE Award Number: DE-FC26-06NT42449

Submitted: March 30, 2011

VSE Corporation
41150 Technology Park Drive, Suite 101
Sterling Heights, MI 48313

DISCLAIMER

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

ABSTRACT

The purpose of this project is to evaluate and compare the effect of the intrinsic differences between cobalt (Co) and iron (Fe) catalysts for Fischer-Tropsch (FT) synthesis using coal-derived syngas and using a realistic commercial design for a Coal-to-Liquids (CTL) plant. The ultimate goal is to consider all the relevant factors, and to evaluate the net tradeoff between using cobalt or iron catalyst formulations for the production of high-hydrogen content liquid fuels from coal. This report covers Task 6 of the overall project conducted under US Department of Energy Cooperative Agreement DE-FC26-06NT42449, entitled “Production of High-Hydrogen Content coal-Derived Liquids.”

An additional objective, added to this project as a result of Action Item 1 from a NETL Peer Review in October 2010, is to demonstrate the tie-in between the syngas preparation system and the comparison of cobalt and iron FT catalyst systems. Therefore, the plant production capacity considered, in this report on the FT catalyst comparison is approximately 2,000 barrels per day (BPD) of FT liquid products, because this FT plant capacity corresponds to the capacity of the syngas preparation system modeled under Tasks 3 and 4 of this overall project. The syngas required is obtained from the co-gasification of 800 tons per day (t/d) of coal and 200 t/d of biomass (Reference 1). A feedstock mix such as this has many advantages compared to the use of coal as the only feedstock (Reference 2).

The results of the overall comparison show that the overwhelmingly decisive factor favoring cobalt over iron FT catalyst for this application is the much higher capital cost of the major equipment needed for the iron-based FT portion of the CTL plant, as compared to a cobalt-based FT plant with essentially the same liquid fuel production capacity. The higher capital cost of the iron-based FT plant is attributable primarily to the need for significantly larger FT reactors, and to the need for expensive internal recycle compressors. For cobalt, these compressors are not needed for the case where the CO₂ is not removed from the syngas after the shift reactor, but a smaller compressor (recycle ratio of 0.55) is needed only for the first stage for the case where the CO₂ is removed after the shift reactor.

Other factors are considered as well, but they are shown to have relatively little influence on the overall comparison of cobalt and iron, in view of the large difference in capital cost of the plant. For example, iron FT catalyst, while not cheap, costs less than cobalt FT catalyst. However, more than twice as much iron catalyst is needed for the same production of FT fuels. Furthermore, the productive lifetime of the cobalt catalyst is normally an order of magnitude longer than that of iron, meaning that effective catalyst cost per barrel of product is actually lower for cobalt.

Although iron FT catalyst is able to tolerate a somewhat higher sulfur level in the syngas, this does not necessarily translate into a significant saving in practice. The same expensive initial-stage syngas clean-up system, capable of reducing syngas sulfur level to about 200 parts per billion (ppb) is required for both catalyst types. Although cobalt's

required final guard-bed type system for additional sulfur removal (down to ~40 ppb) might not be absolutely required with iron, its added cost is much lower than the initial-stage clean-up system. Furthermore, even with iron FT catalyst, the final guard-bed system can function as a valuable “insurance policy” to protect the catalyst charge in the event of upstream upsets in the plant.

Iron FT catalyst is able to promote the water-gas-shift (WGS) reaction in addition to catalyzing FT synthesis. This means that syngas with a lower H_2/CO ratio can be fed to an iron catalyst FT reactor. However, both the total CO_2 produced by both catalysts per barrel of products is similar, and ultimately the total amount of hydrogen required to produce high-hydrogen content coal-derived finished liquid fuels, post upgrading, is also similar for both catalysts.

While for cobalt the CO_2 is produced during the shift reaction before the cobalt-catalyst FT reactor, for iron a similar total amount of CO_2 is produced; this includes both some amount of the initial water-gas shift (WGS) ahead of the iron-catalyst FT reactor, and additional WGS shift inside the FT reactor. For iron it is therefore convenient to first evaluate a range of H_2/CO ratios in the syngas feed in order to assess the tradeoff--as the H_2/CO ratio of the feed increases, the FT reactor size decreases, but there is lower overall FT fuel production for a given initial flow rate of raw (i.e., pre-shift) syngas.

In terms of CO_2 management, two scenarios have been evaluated. One scenario assumes that CO_2 from the initial water-gas shift is “left in” the synthesis gas through the FT synthesis process, and that the CO_2 could be removed subsequently from the tail gas of the FT process. The other scenario assumes that CO_2 is removed (using a Rectisol-based syngas preparation system, see Reference 1) from the synthesis gas prior to entering the FT process. The results obtained under these two scenarios constitute the project team’s response to Action Item 3 of the Peer Review referred to previously, namely to demonstrate the impact of syngas contaminant CO_2 on Reactor Design, the model FT processes, and system economics.

TABLE OF CONTENTS

DISCLAIMER	II
ABSTRACT.....	III
TABLE OF CONTENTS.....	V
ACRONYMS, ABBREVIATIONS AND NOMENCLATURES.....	VI
1. EXECUTIVE SUMMARY	2
2. OBJECTIVE.....	5
3.0 INTRINSIC DIFFERENCE BETWEEN COBALT AND IRON BASED FISCHER-TROPSCH CATALYSTS	6
4.0 METHODOLOGY FOR THE COMPARISONS	10
5.0 REGIONS WITH OPERATING CONDITIONS MORE SUITABLE FOR AN IRON AND FOR A COBALT BASED FISCHER-TROPSCH CATALYST	14
6.0 DESIGN OF A COMMERCIAL CTL PLANT USING COBALT AND IRON BASED FISCHER-TROPSCH CATALYSTS	17
6.1. Syngas composition used for both catalysts	17
6.2. Optimum Plant Design for the iron and cobalt catalysts without removal of the initial-shift CO ₂ from the syngas feed.....	18
6.3. Optimum Plant Design for the iron and cobalt catalysts with removal of the initial-shift CO ₂ from the syngas feed	30
6.3.2.Design data for the iron case	33
6.3.3.Design data for the cobalt case	38
7.0 SIMPLIFIED PFD FOR THE OPTIMUM IRON AND COBALT CASES	43
8.0 IRON VS. COBALT (MAJOR EQUIPMENT, GENERAL LAYOUT AND COSTING)	45
9.0 IRON CASE WITHOUT CO ₂ REMOVAL (MAJOR EQUIPMENT AND COSTING)	46
10.0 COBALT WITHOUT CO ₂ REMOVAL (MAJOR EQUIPMENT AND COSTING)	48
11.0 IRON WITHOUT CO ₂ REMOVAL (GENERAL LAYOUT)	50
12.0 COBALT WITHOUT CO ₂ REMOVAL (GENERAL LAYOUT).....	51
13.0 IRON WITH CO ₂ REMOVAL (MAJOR EQUIPMENT AND COSTING).....	52
14.0 COBALT WITH CO ₂ REMOVAL (MAJOR EQUIPMENT AND COSTING)...	53
15.0 IRON WITH CO ₂ REMOVAL (GENERAL LAYOUT)	55
16.0 COBALT WITH CO ₂ REMOVAL (GENERAL LAYOUT)	56
REFERENCES	57
APPENDIX.....	58

ACRONYMS, ABBREVIATIONS AND NOMENCLATURES

ASTM	American Society for Testing and Materials
atm	Atmosphere
BPD	Barrels per day
BTU	British Thermal Unit
°C	Degrees Celsius
cCO/ccCat/h	Conversion of CO/cc of Catalyst/hour
CH ₄	Methane
Cm	Centimeters
CO	Carbon monoxide
Co	Cobalt
CO ₂	Carbon dioxide
comp	Completed
conv	Conversion
CTL	Coal-to-liquids
CX	Carbon chain length (paraffin)
EFT	Emerging Fuels Technology
EST\$PRO	Gulf Publishing estimating software
°F	Degrees Fahrenheit
Fe	Iron
FT	Fischer-Tropsch
Ft ³	Cubic feet
GHSV	Gas hour space velocity
H ₂	Hydrogen
HFTL	Heavy Fischer-Tropsch Liquids
hrs	Hours
ID	Inside Diameter
LFTL	Light Fischer-Tropsch Liquids
LV	Linear Velocity
m	Meters
MFTL	Middle Fischer-Tropsch Liquids
min	Minutes
MMscfd	Million standard cubic feet per day
NETL	National Energy Technology Laboratory
nm ³ /h	Normal Cubic Meters Per Hour
OD	Outside Diameter
PFD	Process flow diagram
PH ₂ O	Water partial pressure
ppb	Parts per billion
psig	Pounds Per Square Inch Gauge
ROI	Return on investment
Rx	Reactor(s)
s	Seconds

S	Sulfur
scfd	Standard Cubic Feet Per Day
SCFH	Standard Cubic Feet Per Hour
t/d	tons per day
TEMA	Tubular Exchanger Manufacturers Association
tickn	Thickness
Tot	Total
US	United States
WGS	Water-gas-shift reaction
wt	Weight
ZnO	Zinc Oxide

1. EXECUTIVE SUMMARY

Before proceeding to perform a comparison of the effect of the intrinsic activities of the Cobalt and Iron based Fischer-Tropsch (FT) catalysts, it was necessary to ascertain whether some of the current claims in the literature about the production superiority of one metal over the other were a function of the operating conditions.

We found that the operating conditions where one metal (in this case, iron) had a higher volumetric productivity than the other (cobalt) were outside those suitable for practical application.

Once this had been demonstrated, we proceeded to design a commercial CTL plant based on the composition and amount of the raw synthesis gas described in the Milestone report for Tasks 3 and 4 of this project issued in February 2010 (Reference 1). This syngas is obtained by the high-temperature gasification (in molten iron) of a total of 1,000 t/d of feedstock, consisting of 80% coal and 20% biomass. The operating conditions selected for the comparison of cobalt and iron FT catalyst systems were the optimum ones for each specific catalyst.

In response to the NETL Peer Review Action Items from October 2010, and to fully demonstrate both the:

- Tie-in between the syngas preparation system and the FT catalyst system comparison, and
- The effects of syngas contaminant CO₂ on Reactor Design, the model FT processes, and system economics,

the evaluation was performed for two scenarios: with and without the removal of CO₂ after the shift reactor.

Comparison without removal of CO₂ after the shift reactor

For the first case study, the CO₂ was not removed from the syngas after the shift took place (despite the fact that the equipment configuration described in Reference 1 implies otherwise). Leaving the CO₂ in the syngas provides some benefit because the CO₂ behaves as an inert in the Fischer-Tropsch reactor. As an inert, it helps to lower the partial pressure of water and therefore allows for a higher per-pass conversion for both the iron and cobalt based catalysts. This could also benefit the syngas preparation section of the plant by postponing, or possibly eliminating, the CO₂ removal step, depending on the overall configuration and planned operating scenario of the plant with respect to CO₂ management. The excess water not consumed during the shift has to be removed because it would increase the deactivation of both catalysts.

It was found that for the same amount of raw syngas (described in Reference 1), and after performing water gas shift at a level suitable for each case, the results favored a cobalt based FT catalyst. This is shown in Table 1.1 below. The specific iron case is a compromise selected after a study of nine (9) iron cases, each with a different H₂/CO ratio in the syngas feed (after shift) to the first Fischer-Tropsch stage.

Table 1.1 shows the plant production for the case where naphtha and diesel are the only products. The plant could, however, also include hydrotreating equipment suitable for the production of jet

fuel. The carbon number range for the jet fuel overlaps with both the naphtha and diesel fractions. The wide cut jet fuels are comprised of highly isomerized hydrocarbons with an overall carbon range between C8 to C17 or C18, depending upon the particular specification. This approach will reduce the total amount of naphtha produced.

Table 1.1 Design of a CTL plant based on iron and cobalt FT catalysts (without CO ₂ removed after shift reactor)				
	Iron		Cobalt	
	1 st Stage	2 nd Stage	1 st Stage	2 nd Stage
FT reactor weight (tons)	635	254	240	158
Total plant FT reactor weight (tons)	889		398	
FT Reactor recycle, mol/mol	1.25	0.4	None	None
Catalyst Volume, ft ³	8200	2500	2380	1585
Total plant catalyst vol (ft ³)	10700		3965	
Plant production, BPD				
naphtha (C5-C9)	490	85	420	180
diesel (C10-C19)	1420	220	1085	465
Total per Stage	1910	305	1505	645
Total plant production (BPD)	2215		2150	

As described in more detail later in the report, the compromise iron case selected results in a lower production (intermediate) than the most productive case, but also with less tubes, catalyst and reactor weight than the most productive case.

Notice that for the same amount of raw syngas, a cobalt based CTL plant will produce less barrels per day, the reactor volume will be about 1/3, and the reactors will weigh less than half as much as for a CTL plant designed using an iron based catalyst. In addition, a cobalt based CTL plant will not need internal recycle compressors, if the CO₂ from the shift is left in the syngas. In the case of the iron catalyst, 82 % of the CO reacted in the first stage, 52 % to hydrocarbons and the rest to CO₂. Because of this, the feed to the second stage is even more diluted by the CO₂, affecting negatively the iron kinetics by considerably lowering the FT/WGS reaction ratio and therefore increasing the necessary FT reactor volume. Lower conversions in the first stage resulted in a lower overall plant capacity.

The ratio of FT reactor tons per bpd in the first iron stage is about 0.33 and in the second stage about 0.83, and the ft³ of catalyst per bpd in the first iron stage is about 4.3 and in the second stage about 8.2.

For cobalt, the ratio of FT reactor tons per bpd in the first stage is about 0.16 and in the second stage about 0.24, while the ft³ of catalyst per bpd in the first cobalt stage is about 1.6 and in the second stage about 2.5.

The total installation cost for the major equipment, piping and instrumentation (excluding civil/utilities and support equipment/structures) is shown in the cost analysis in sections (9) and (10) of this report.

Comparison with removal of CO₂ after the shift reactor

In addition to its beneficial effect of allowing higher per-pass conversion, the diluent CO₂ from the initial shift also increases the total mass and volume of syngas that must be processed, which increases the required FT reactor sizes. Removing the CO₂ after the shift has a cost, but it reduces the size of FT reactors needed for both catalyst types, as well as capturing this CO₂ for subsequent application or sequestration, depending upon the carbon management strategy being used by the plant. Note that this approach can be particularly effective with cobalt FT catalyst, because hardly any additional CO₂ is produced within the cobalt FT reactors.

In this case, the results again favor the cobalt catalyst in terms of FT reactor tonnage, catalyst volume and in the much lower recycle needs for the cobalt case. The total C₅+ production, however, favors the iron case by about 140 bpd for the same amount of raw syngas feed.

According to the results shown in Table 1.2, the reactor tonnage for the iron case is over 2 times that required for the cobalt case; the amount of catalyst is almost three times higher; and the recycle needs are again much higher for iron than for cobalt. For iron to be better than cobalt in this scenario, the difference in the investment cost for the two options would have to be justified by this difference in production.

The total installation cost for the major equipment, piping and instrumentation (excluding civil/utilities and support equipment/structures) is shown for this case in the cost analysis in sections (13) and (14) of this report.

The difference in capital investment for the iron and cobalt cases (excluding civil/utility and support equipment/structures) is almost \$24 million, or an investment of about \$170,000 per barrel produced per day for the extra 140 bpd. On a per-barrel-per-day basis, this incremental investment would be about an order of magnitude greater than the typical overall equipment cost for an FT plant. This means that in this scenario, where the CO₂ is removed after the shift reactor, the cobalt case is again more economic than the iron one.

Both catalysts will need to have extensive syngas cleaning. While the iron-based catalyst will need to have the sulfur (S) contaminants levels reduced to about 200 ppb, the cobalt catalyst will need an extra step, probably a ZnO bed, to lower it further from about 200 to about 40 ppb. The cost for lowering the contaminants in the coal derived syngas down to about 200 ppb is much more expensive than its further lowering down to 40 ppb.

Table 1.2 Design of a CTL plant based on iron and cobalt FT catalysts (with CO ₂ removed after the shift reactor)				
	Iron		Cobalt	
	1 st Stage	2 nd Stage	1 st Stage	2 nd Stage
FT reactor weight (tons)	404	205	235	45
Total plant FT reactor weight (tons)	609		280	
FT Reactor recycle, mol/mol	1.6	1.2	0.55	None
Catalyst Volume, ft ³	5321	2752	2337	456
Total plant cat vol (ft ³)	8073		2793	
Plant production, BPD				
naphtha (C5-C9)	466	125	505	93
diesel (C10-C19)	1354	336	1306	239
Total per Stage	1820	461	1811	332
Total plant production (BPD)	2281		2143	

Based on the technical section of this study, a supported cobalt catalyst is more effective than an iron catalyst for the conversion of coal to high-hydrogen content liquid transportation fuels. This conclusion is based on the larger volume of the FT reactors, FT reactor weight, catalyst volume, and increased plant complexity--including recycle compressors (e.g., the iron option as compared to cobalt).

2. OBJECTIVE

The objective of this project is to evaluate and compare the effect of the intrinsic differences between cobalt (Co) and iron (Fe) catalysts for Fischer-Tropsch (FT) synthesis using coal-derived syngas and using a realistic commercial design for a CTL plant. Two additional objectives were added to this project as a result of Action Items 1 and 3 from a NETL Peer Review of this project in October 2010:

- Demonstrate the tie-in between the syngas preparation system and the comparison of cobalt and iron FT catalyst systems, and
- Demonstrate the impact of syngas contaminant CO₂ on Reactor Design, the model FT processes, and system economics.

3.0 INTRINSIC DIFFERENCE BETWEEN COBALT AND IRON BASED FISCHER-TROPSCH CATALYSTS

The commercially practical iron and cobalt-based catalysts are very different in composition, properties and behavior. In this study, we will summarize those differences that have an impact on the configuration, cost and productivity of a CTL plant.

A summary of these differences is shown in Table 3.1 below.

Generally, a precipitated iron FT catalyst has a higher activity than a supported iron FT catalyst. The cobalt based catalyst, however, typically has a much higher activity than an iron based one. This difference in activity could be two or threefold in fixed-bed Fischer-Tropsch reactors appropriate for a plant capacity of nominal 2,000 barrels per day (BPD). (The plant production capacity considered in detail is approximately 2,000 BPD of FT liquid products, because this FT plant capacity corresponds to the capacity of the syngas preparation system modeled under Tasks 3 and 4 of this overall project.) Because of heat transfer considerations, however, the cobalt catalyst is not able to apply the whole of its activity in fixed-bed reactors. The iron catalyst, on the other hand, is more often limited by kinetics than by heat transfer.

One area where the iron catalyst has an advantage for producing liquid transportation fuels over a cobalt catalyst is methane (CH₄) selectivity. For iron catalysts, the CH₄ selectivity varies between 4 and 5 % while for cobalt catalysts this selectivity will vary between 6 to 10 %. This represents an advantage of between 2 to 5 % more of the most desired products via an iron-based FT catalyst.

This advantage, however, is lost due to the iron catalyst's high activity for the water-gas-shift reaction.

At the operating conditions typical of low temperature FT, the WGS equilibrium constant is close to 100. This means that there is no net reverse reaction and the H₂/CO ratio will keep increasing as the CO is converted to CO₂.

On the other hand, the average hydrogen content of the Fischer-Tropsch products is about CH_{2.14} and the Fischer-Tropsch reaction is :

This means that for the iron catalyst to be able to consume the CO in an optimum manner and avoid making more CO₂ than necessary, the H₂/CO ratio in the outlet should be as low as possible, without promoting deactivation or excessive oxygenates production. This is difficult to achieve because as the reactant's partial pressures decrease with conversion and more H₂O is present, the ratio of the FT/WGS rates decrease due to their respective kinetic behaviors. Therefore, the best H₂/CO ratio practically possible at the outlet of the iron second stage is about 2.5. This means that the iron catalyst is producing more CO₂ than the cobalt based catalyst, which has an H₂/CO ratio at the outlet of the second stage of about 1.3.

Because the best iron catalysts are prepared by the precipitation technique, they are mechanically weaker than the cobalt ones that are typically prepared by impregnating the metal precursor over

a strong metal oxide such as alumina or silica. This means that the diameter for the iron catalyst extrudates has to be larger than that for the cobalt catalyst (or it will break during loading or upon deactivation by oxidation). This means that the effectiveness factor for iron catalysts is lower than that for cobalt ones and therefore the volumetric productivity will decrease even more than that due solely to the lower intrinsic activity of iron as compared to cobalt.

The iron catalysts are more prone to oxidation than cobalt catalysts. It has been reported that cobalt catalysts can stand up to 5 to 5.5 atm of P_{H_2O} , as compared to less than 1.3 to 1.5 for iron catalysts.

This is a very important difference. It means that at the same operating pressure, and because water is a product of the FT reaction, the iron catalyst is able to accept a much lower per pass conversion than a cobalt catalyst. A plant using an iron catalyst will therefore need both more stages and more reactors to achieve a high CO conversion (say, above 90 %), or the use of expensive compressors for internal recycle. The water is condensed and separated from the recycle stream, therefore allowing a higher per stage conversion.

Again, because the iron catalyst is prepared by the precipitation of, for example, iron nitrate and then mixed with a structural promoter such as silica or alumina, its physical strength is also a function of the degree of deactivation or oxidation of the catalyst. As the catalyst loses its FT activity by oxidation of the iron carbide to magnetite (Fe_3O_4 , active for WGS) this is accompanied by sintering and loss of surface area. Because of this, the deactivation of precipitated iron catalysts by oxidation is irreversible. On the other hand, because the cobalt metal is supported on an oxide such as alumina or silica, the oxidation of the surface of the cobalt crystallites is reversible by a re-reduction process. The regenerability of the cobalt based FT catalyst, together with its slow deactivation, facilitates a 2 to 3 years catalyst life (in the absence of irreversible poisons such as sulfur-based compounds).

The Fischer-Tropsch reaction is a hydrogenation reaction. This means that catalysts with a strong hydrogenation activity (e.g., cobalt catalysts), will have a faster FT reaction rate and will produce more hydrogenated products than catalysts with a lower hydrogenating activity (e.g., iron catalysts).

This also means that the higher the hydrogenation activity the more methane is produced and explains why cobalt methane selectivity is higher than that for iron. Further, for catalysts with a high hydrogenation activity it is easier to hydrogenate the growing hydrocarbon chain on the catalyst surface to a paraffin therefore terminating its growth and producing shorter hydrocarbons. This explains why the iron catalyst's products are intrinsically "heavier" (longer carbon chain lengths) than those for cobalt catalysts.

Because of its lower hydrogenating activity, the iron catalyst has a higher selectivity towards olefin and oxygenated hydrocarbon products. The treatment of water produced by iron FT catalysts is then more expensive than that for cobalt catalysts, and if hydrogenation of some of the hydrocarbon products is needed to produce high-hydrogen content fuels, more hydrogen will be needed.

Both catalysts react in a similar manner to the poisons present in the syngas feed and will need to have extensive syngas cleaning ahead of the FT reactors. While the iron based catalyst will need to have the sulfur contaminants levels reduced to about 200 ppb, the cobalt catalyst will need an extra step, probably a ZnO bed to lower it further from about 200 to about 40 ppb. The cost for

lowering the contaminants in the coal-derived syngas down to about 200 ppb is much more expensive than its further lowering down to 40 ppb.

The reason for the higher maximum amount of contaminants in the iron case is not due to an intrinsic higher resistance to sulfur. There are two main reasons for this:

- The precipitated iron catalyst has more iron sites per cc of catalyst (or reactor unit volume) than the supported cobalt catalyst.
- The expected life time of the iron catalyst is measured in months while the cobalt is measured in years. At the same contamination levels, the cobalt catalyst will have more time (about 10 times more) than an iron catalyst to absorb contaminants.

The incremental syngas clean-up step for cobalt catalysts is shown in Figure 3.1. The lowering of sulfur contaminants from about 200 to about 40 ppb can be achieved by means of a guard bed such as the commercially available ZnO option.

Figure 3.1. Syngas cleaning for iron and cobalt Fischer-Tropsch catalysts

The iron catalyst is less expensive per pound of catalyst. Although we need about twice as much iron catalyst as cobalt catalyst on a mass basis, the initial catalyst cost for an iron catalyst is lower than that for cobalt. Because of the longer lifetime of the cobalt catalyst, the final cost of catalyst

per barrel of products is lower for cobalt than for iron, as based upon the experience of Emerging Fuels Technology, Inc. and other industry experts.

Table 3.1 Comparison between iron and cobalt Fischer-Tropsch catalysts		
Item	Iron	Cobalt
Activity	Lower	Higher
Methane Selectivity	Lower	Higher
Diesel Yield	Similar	Similar
Olefins/Oxygenates in products	Higher	Lower
Catalyst life	Lower (months)	Higher (years)
Regeneration	No	Yes
Catalyst mechanical strength	Lower	Higher
Contaminants tolerance	Higher	Lower
Reactor volume (for same production)	Larger	Smaller
Optimum H ₂ /CO in inlet	~ 1.5 – 1.8	~ 2.1
CO ₂ selectivity in FT reactor	Higher	Lower
Catalyst cost, \$/lb cat	Lower	Higher
Catalyst cost, \$/barrel	Higher	Lower

4.0 METHODOLOGY FOR THE COMPARISONS

It has been stated in the literature that there are certain operating regions where an iron catalyst has a higher productivity than a cobalt catalyst and vice versa. Before proceeding with a comparison of both catalysts, it is therefore necessary to evaluate those operating regions in order to include them in this comparison.

Once the regions are identified, they will be selected for the comparison. Or, if the statements in the literature prove to be erroneous or not applicable, then the operating conditions for each catalyst will be individually optimized.

The cobalt catalyst activity and selectivity has been taken from repeated experimental results at the Emerging Fuels Technology, Inc. (EFT) using fixed-bed micro reactors. For the iron catalyst, literature data (Sasol) was used to determine the catalyst activity and the relative activity ratio between the Fischer-Tropsch and the Water-Gas-Shift reactions.

The equations used for the Fischer-Tropsch and Water Gas Shift reactions are:

Iron Catalyst: The FT activity was calculated using the Anderson kinetic equation (Reference 3):

$$r_{-CO} = A e^{(-\Delta E/RT)} P_{H_2} P_{CO} / (P_{CO} + a P_{H_2O}) \quad \text{-----} \quad (1)$$

Where, r_{-CO} is the rate of disappearance of CO (FT rate) in terms of cc CO reacted/cc cat/h

A is the pre-exponential factor or the catalyst intrinsic activity

ΔE is the activation energy

R is the Universal gas constant, $\Delta E/R = -8000$

T is the reaction temperature, in °K

“a” is the desorption constant equal to about 3.0

P_{H_2} , P_{H_2O} and P_{CO} are the partial pressures of hydrogen, water and carbon monoxide, in atm.

Notice that because of Equation (1), the FT rate of iron catalysts is almost directly related to the concentration of the reactants. That is, the rate will diminish as conversion proceeds. This is shown in Figure 4.1. The data in this figure was taken from the calculations for the base case for an iron commercial CTL plant based on the syngas composition gas described in Reference 1, after an optimum degree of WGS upstream of the FT reactors.

For this initial case being considered, the CO_2 was not removed from the syngas after the shift took place (despite the fact that the equipment configuration described in Reference 1 implies otherwise). This was because the CO_2 behaves as an inert in the Fischer-Tropsch reactor. As an

inert, it helps to lower the partial pressure of water and therefore allows for a higher per-pass conversion for both the iron and cobalt based catalysts.

This figure shows that at the reactor inlet the rate for the FT reaction rate is about 4 times higher than that at the exit of the reactor. The “bump” in the curve is because of the axial temperature profile in the FT fixed-bed reactor, which reaches its maximum at about 1/3 from the inlet of the reactor.

Figure 4.1. Axial profile for the FT rate for an iron catalyst at typical commercial conditions

Iron Catalyst: The WGS activity was calculated using the equation recommended by G Botes (Reference 4):

$$-r_{\text{WGS}} = K e^{(-\Delta E/RT)} = \frac{[P_{\text{CO}} P_{\text{H}_2\text{O}} - 1/K_{\text{WGS}} P_{\text{H}_2} P_{\text{CO}_2}]}{[1 + 1.1 P_{\text{H}_2} + 6.3 P_{\text{H}_2\text{O}}/P_{\text{H}_2}^{0.5}]^2} \quad (2)$$

Where:

$\Delta E/R = -7,000$, and

K_{WGS} is the water gas shift equilibrium constant, function of temperature according to :

$$K_{\text{WGS}} = 0.0132 e^{(4578/T)} \quad (3)$$

Cobalt catalyst: The activity is calculated using the Yates and Satterfield equation (Ref 5), shown below

$$r_{-\text{CO}} = A e^{(-\Delta E/RT)} P_{\text{H}_2} P_{\text{CO}} / (1 + a P_{\text{CO}})^2 \quad (4)$$

Where,

$\Delta E/R = -11000$, and

$a = f(T)$ At typical operating conditions, this values varies from ~ 1.5 to ~ 2.5

Notice that in this case the equation could be simplified to :

$$r_{-CO} \approx A e^{(-\Delta E/RT)} P_{H_2} / P_{CO} \text{-----} (5)$$

This means that for the case of a typical cobalt FT catalyst, the rate is more related to the ratio of the reactants than to their absolute concentrations.

Figure 4.2. Axial profile for the FT rate for a cobalt catalyst at typical commercial conditions

This is reflected in the axial FT rate profile for a typical cobalt catalyst at commercial conditions. The data in this figure was taken from the calculations for the base case for a cobalt catalyst based commercial CTL plant based on the syngas composition gas described in Reference 1, after an optimum degree of WGS upstream of the FT reactors . Same as for the iron catalyst case, the “bump” in the curve is because of the axial temperature profile in the FT fixed-bed reactor, which reaches its maximum at about 1/3 from the top, or inlet, of the reactor

From Figures 4.1 and 4.2 it is evident that the reactor volume for a cobalt FT catalyst is better utilized than the reactor volume for an iron one.

Calculation of the CO conversions

To calculate the CO conversion at any given set of operating conditions, the above kinetic equations were inserted into a kinetic model consisting of 100 increments or cells. In each cell, the partial pressures of the reactants are calculated and the kinetic equations applied. The gas composition is then recalculated by subtracting the molecules that reacted and adding the ones formed. Once the mass balance is performed, this becomes the feed for the next cell.

In addition, a typical axial temperature profile, obtained from the EFT Fischer-Tropsch pilot plant has been included in the model to calculate the maximum rate at any axial point in the reactor based on the temperature and the reactant's partial pressures at any specific axial point.

When using this model to calculate the CO conversions, the following rules applied:

Limitations used for the design of the fixed-bed Fischer-Tropsch reactors

- The maximum gas linear velocity used was 30 – 33 cm/s. This is calculated at operating conditions in an empty tube. Higher linear velocities would increase the pressure drop above acceptable figures. The reactor height was 33 ft for cobalt and 39.37 ft (10 meters) for iron. These are typical values for commercial FT fixed-bed reactors.
- The minimum linear velocity used was about 20 cm/s. This is because the lower the linear velocity, the lower the axial overall heat transfer coefficient. At low linear velocities, the heat transfer may decrease to the point where a temperature runaway may occur.
- The maximum average Fischer-Tropsch productivity was about 200 – 215 cc CO converted per cc of catalyst per hour. Based on our experience, the maximum rate could be about 275 (at an axial position close to 1/3 from the reactor top, or inlet, at the center of the tube), which is still within the heat transfer capabilities of a 1" (inner diameter) ID tube, operating at a linear velocity above 20 cm/s.
- For the case of iron based catalysts, and because of their intrinsic lower activity, the ID of the tubes could be as large as 2 inches, which translates to less tubes and therefore lower construction costs (as compared to 1" ID tubes for the cobalt case).

5.0 REGIONS WITH OPERATING CONDITIONS MORE SUITABLE FOR AN IRON AND FOR A COBALT BASED FISCHER-TROPSCH CATALYST

It has been stated in the literature (Reference 6) that under some operating conditions the iron catalyst is more productive than the cobalt catalyst and vice versa. This is discussed by Professor Burt Davis in (Reference 7) and shown in Figure 5.1 below.

Figure 5.1. Productivity comparison between the iron catalyst (240 °C) and the cobalt catalyst (220 °C)

Should this be true, then this study should be oriented towards identifying these conditions and optimizing the plant design for each catalyst accordingly.

In order to perform this comparison, we had to identify the catalysts' intrinsic activity.

Intrinsic activity of the cobalt and iron catalysts used in this study

For the cobalt catalyst case, the activity used was that typical of a moderately active catalyst after about 500 hrs of continuous run time. This is an extruded catalyst with dimensions suitable for fixed-bed reactors

The literature information on the Sasol Arge reactors was used to calculate the relative catalyst activity for the precipitated iron catalyst using equations 1, 2 and 3 above. The comparison in this work is therefore based on Sasol's precipitated iron catalyst. The model used 2,052 tubes of 2" ID and 12 m height (Reference 8) and back calculated the activity necessary for the production quoted in Reference (8) of between 400 to 600 bpd. The water partial pressure at those conditions was assumed to be close to the upper limit.

Both catalysts' productivity was then calculated by means of the fixed-bed model described in section 4 of this report (*Calculation of the CO conversions*).

The pressure range for the comparison was between 260 and 600 psig (~ 18.7 and 41.8 atm) and the gas hour space velocity (GHSV) between 1.75 and 6.5.

Using any given set of conditions, the models were run at a constant temperature for each catalyst (that is, a constant temperature for the iron of 237 °C), which is a typical commercial temperature for Fe catalyst as well as to compare with example in Figure 5.1 and for the cobalt of 192 °C--typical optimal temperature for this specific cobalt catalyst--varying the space velocity for each pressure value until a productivity match was found, where the productivity of iron and cobalt were equal.

Because the iron catalyst typically needs recycle and the cobalt catalyst recycle is a function of the H_2+CO concentration, this process was performed with an iron typical recycle ratio and for the typical cobalt case, without recycle. The results are shown in Figure 5.2 below.

Figure 5.2. Productivity comparison between the iron catalyst (237 °C) and the cobalt catalyst (192 °C)

According to Figure 5.2, there are indeed operating regions (combinations of GHSV and pressure), where typical commercial iron catalysts will be more productive than typical commercial cobalt ones and vice versa.

The next step was to perform analyses of these two regions. The analyses are shown in tables 5.1 and 5.2 below.

Table 5.1 Fischer-Tropsch catalysts productivity : Cobalt vs iron, RR=0										
	Co	Fe	Co	Fe	Co	Fe	Co	Fe	Co	Fe
GHSV	6.5	6.5	5.6	5.6	3.1	3.1	2.45	2.45	2.0	2.0
P (psig)	280	280	300	300	400	400	500	500	600	600
% CO to HC	15.5	15.25	18.7	17.05	36.17	32.6	48.9	43.82	60.9	54.3
LV, cm/s	122.3	70	97	56	41.4	23.9	26.4	15.2	18	10.4
P _{H2O} , atm	0.89	0.69	1.17	0.88	1.62	2.14	6.3	3.6	10.7	5.4
Rate, cCO/ccCat/h	93.2	93	95.2	96	103.6	103	108.8	109	112.4	111

Table 5.2 Fischer-Tropsch catalysts productivity : Cobalt vs iron, Fe RR=1.2, Co RR=0								
	Co	Fe	Co	Fe	Co	Fe	Co	Fe
GHSV	5	5	3.1	3.1	2.1	2.1	1.75	1.75
P (psig)	260	260	300	300	400	400	500	500
% CO to HC	18.9	18.3	33.8	30.3	53.8	48.4	56.3	59.9
LV, cm/s	101	128	54.6	69	28	36	23	24
P _{H2O} , atm	1.0	0.38	2.29	0.7	5.8	1.47	7.9	2.3
Rate, cCO/ccCat/h	92	93	96.8	96	104.4	104	108	107

In these tables, a yellow highlight means that the value exceeds the fixed-bed reactors design limits or that the water partial pressure is too high for practical applications. The light tan colored highlight means that the values are either borderline or that the CO conversions are too low for practical applications.

The interpretation of this study is that there are combinations of space velocities and operating pressures where both catalysts have the same productivity. Therefore, at higher space velocities and higher pressures, iron may exceed the productivity of cobalt in spite of its lower intrinsic activity, but these regions are outside commercial application due to either high linear velocities, high partial pressure of water or too low volumetric productivity.

Based on these results, the comparison will now proceed by assigning each catalyst its optimum space velocity, reactants partial pressures and per pass conversion.

6.0 DESIGN OF A COMMERCIAL CTL PLANT USING COBALT AND IRON BASED FISCHER-TROPSCH CATALYSTS

6.1. Syngas composition used for both catalysts

The raw syngas composition used for this comparison is shown in Table 6.1, taken from Reference 1. This syngas is obtained by the high-temperature gasification, in molten iron, of a total of 1,000 t/d feedstock, consisting of 80% coal and 20% biomass.

Table 6.1 Syngas flow and composition					
Component	Vol %	m3/h	Mol Kg/h	lbmol/h	lb/h
H ₂	30.76	21130	952	2094	4188
CO	66.61	45758	2061	4535	126967
CH ₄	0.10	66	3	7	105
CO ₂	1.00	690	31	68	3009
H ₂ O	0.68	464	21	46	828
N ₂	0.58	397	18	39	1102
Ar	0.27	188	8	19	819
Totals	100.00	68693	3086	6807	137019

Each catalyst has its own optimum H₂/CO ratio. This ratio is typically a function of the syngas composition (e.g., concentration of H₂ plus CO in syngas), the desired per pass conversion (which will result in a minimum H₂/CO ratio in the outlet for cobalt or a higher one for iron), etc. This ratio plays an important role for the optimization of the plant design for both catalysts.

In order to increase the H₂/CO ratio, a water gas shift reactor is needed. After the shift reactor, the CO₂ formed during the shift reaction could be either allowed to remain in the syngas (as the CO₂ is practically an inert on both iron and cobalt catalysts) or removed in order to increase the concentrations of H₂ and CO in the syngas. The concentrations of H₂ and CO may affect the optimum plant design in different manners for the iron and for the cobalt based catalysts. Similarly, the decision to remove the initial-shift CO₂ from the syngas, or to “leave it in” for later removal from the FT tail gas, for example, must be considered in terms of the plant’s carbon management strategy, because iron produces a significant additional amount of CO₂ in the FT reactors after the initial shift, but cobalt does not. For these reasons, this study will include both options, that is, an optimum design for both Fischer-Tropsch catalysts with and without CO₂ removal after the shift reactor.

6.2. Optimum Plant Design for the iron and cobalt catalysts without removal of the initial-shift CO₂ from the syngas feed

Optimized Design for the iron catalyst

It is frequently mentioned in the literature that “coal syngas is for iron and natural gas syngas is for cobalt”. This is based on the iron catalyst activity for the WGS reaction, allowing feeding of a sub-stoichiometric H₂/CO. This has two effects. This first one is that the further the feed ratio is from the stoichiometric ratio (approximately 2.14) the more WGS is necessary to produce the needed hydrogen. The H₂/CO ratio for the raw syngas (i.e., before water gas shift) shown in Table 6.1 is about 0.46, well below the stoichiometric ratio.

At this point, it is not possible to predict which H₂/CO ratio in the fresh feed syngas would be optimum for an iron FT catalyst. Because of this, 9 different cases were selected for further study within the constraints of linear velocity, reaction rate and water partial pressure as described in Section 4.1 of this report. These cases for iron are shown in Table 6.2 below. Notice in this table, that the number of tubes and therefore the necessary reactor volume and ft³ of catalyst increase as the H₂/CO ratio in the syngas fresh feed decreases. On the other hand, the number of barrels per day produced increases.

The optimum H₂/CO ratio will therefore have to be calculated based on the additional cost versus the additional production as the H₂/CO ratio decreases.

Table 6.2. Optimization of the H ₂ /CO ratio on the FT section syngas fresh feed (After Shift)									
H ₂ /CO after shift	0.75	0.87	1.11	1.25	1.40	1.50	1.60	1.70	1.80
Production,bpd									
1 st Stage	1976	1959	1910	1878	1811	1739	1642	1581	1497
2 nd Stage	333	328	304	245	257	290	338	359	391
Plant Total	2308	2287	2214	2123	2068	2029	1980	1940	1889
Recycle Ratio									
1 st Stage	1.50	1.40	1.25	1.20	1.20	1.20	1.20	1.20	1.20
2 nd Stage	0.70	0.55	0.40	0.45	0.60	0.70	0.80	0.90	1.00
Number tubes (2" ID)									
1 st Stage	9170	9027	8935	8947	8433	7714	6789	6338	5700
2 nd Stage	3651	3499	2765	2810	3116	3322	4113	4369	4681
Total tubes	12821	12525	11700	11757	11549	11037	10902	10707	10381
ft ³ of catalyst									
1 st Stage	8413	8282	8197	8208	7737	7077	6228	5815	5229
2 nd Stage	3350	3210	2537	2578	2858	3048	3773	4008	4295
Total ft ³ cat per plant	11762	11491	10734	10786	10595	10125	10002	9823	9524
CO conversions									
Tot CO conv., %	93.72	94.46	93.57	93.88	93.76	93.10	93.11	92.87	92.25
Hydrocarbons and CO ₂ Selectivity (%), after shift									
CO to hydrocarbons	46.03	45.17	43.01	41.76	40.60	39.95	38.95	38.17	37.25
CO to CO ₂	53.97	54.83	56.99	58.24	59.40	60.05	61.05	61.83	62.75

Optimized design for the cobalt catalyst

The optimization for the plant design running with a cobalt catalyst is much easier than that for an iron catalyst. The reason is that the cobalt catalyst practically has no water gas shift activity and typically less than 1 % of the CO that reacts is converted to CO₂.

The optimization therefore has to consider the minimum H₂/CO ratio acceptable at the outlet of the second stage at the total CO conversion. The optimum design was also calculated using the limits described in Section 4.1 above. The results for cobalt are shown in Table 6.3 below. According to summary table 1.1, a cobalt based plant compares with an iron based one as:

- Cobalt makes about 2,150 bpd. This is about 65 bpd less than iron would produce using an H₂/CO after shift of 1.1.
- If the syngas is shifted only up to an H₂/CO ratio of 0.75, iron will produce about 160 bpd more than the cobalt catalyst.
- To produce 160 bpd more than cobalt, iron will need more than 490 additional tons of FT reactor than that for the cobalt case. This additional mass is more than the total weight of both cobalt reactors.
- Also, as shown in Table 6.2, in order to produce 160 bpd more than cobalt, iron will need 11,762 ft³ of catalyst, or about 7,500 ft³ more than that needed for the cobalt case. Notice that the cobalt needs only 3,965 ft³ of catalyst to produce about 2,150 bpd of C5+ products.

While a rigorous answer needs to quantify these differences in terms of capital investment and ROI, the differences are so large that it is evident that a cobalt catalyst based CTL plant has a much lower capital investment than an iron based CTL plant, as shown in the equipment cost summaries later in this report, and thus a higher return on investment.

Table 6.3. Optimized cobalt design	
H ₂ /CO after shift	1.83
Production,bpd	
1 st Stage	1506
2 nd Stage	644
Plant Total	2149
Recycle Ratio	
1 st Stage	0.00
2 nd Stage	0.00
Number tubes (1" ID)	
1 st Stage	12000
2 nd Stage	8000
Total tubes	20000
ft ³ of catalyst	
1 st Stage	2377
2 nd Stage	1584

Total ft ³ cat per plant	3961
CO conversions	
Tot CO conv., %	91.36
Hydrocarbons and CO ₂ Selectivity (%), after shift	
% CO to hydrocarbons	99.0
% CO to CO ₂	1.0

6.2.1. Design Data for the iron and cobalt optimized cases without removal of the initial-shift CO₂ from the syngas feed

6.2.2. Design data for the iron case with initial-shift CO₂ not removed

The data-column in Table 6.2 selected is the one of the H₂/CO in the fresh feed is equal to 1.11. This case was further refined, by adding more tubes in the second stage (increased from 2,765 to 3,696) in order to maintain the gas linear velocity inside the reactor below about 30 cm/s and therefore still within the range of a manageable pressure drop (which increases exponentially with gas linear velocity). These data are shown in more detail in the following series of tables below:

FIRST STAGE – IRON

SG Feed to first stage	
Flow (nm ³ /h)	82378
SCFD	7.38E+07
H ₂ /CO in	1.106
SYNGAS	Comp (%)
H ₂	42.64
CO	38.55
CH ₄	0.08
CO ₂	17.83
H ₂ O	0.18
N ₂	0.48
Ar	0.23
Total	100

OPERATING PARAMETERS		
Temperature	467.86	°F
Pressure	350	psig
ΔP	45	psi
Recycle	1.25	mol/mol

SELECTIVITIES	
ALPHA	0.94
CH4 SEL (Carbon %)	5
CO2 SEL (Carbon %)	36.82
USAGE RATIO	0.956

Note : Selectivity as carbon % means the

percentage of CO reacted that ended as

CH4, CO2, etc

CO Conversions (%)			
	To HC	To CO2	Tot CO conv
Per Pass	35.47	20.67	56.13
Total	51.83	30.2	82.03

PRODUCTIVITY	
Avg FT rate in cc CO/cc cat/h)	71

Partial Pressures (atma)			
	Inlet	Outlet	Kin Avg
H2	8.22	5.23	6.48
CO	6.21	2.91	4.65
H2O	0	1	0.74
CO2	9.72	11.77	10.78
Total P (atm)	24.81	21.81	

FIRST STAGE PRODUCTION		
Naphtha	BPD	489
Diesel	BPD	1421
Total		
BPD		1910

H2/CO	F Feed	Tot Feed	Rx outlet
	1.106	1.324	1.794

DRY TAIL GAS (TO NEXT STAGE)			
Flows			
	nm3/h	scfd	Mol %
H2	10237	9170326	24.48
CO	5706	5111528	13.65
CO2	24281	21752401	58.07
H2O	0	0	0
INERTS	585	524050	1.4
CH4	889	796851	2.13
C2-C4	116	104108	0.28
TOT	41814	3.75E+07	100
Equiv H2O nm3/h and scfd	36532	32727447	

SECOND STAGE - IRON

SG Feed after H2 removal	
Flow (nm3/h)	39565
SCFD	35441512
H2/CO in	1.4
SYNGAS	Comp (%)
H2	20.19
CO	14.42
CH4	2.25
CO2	61.37
H2O	0
N2	1.48
C2-C4	0.29
Totals	100

OPERATING PARAMETERS		
Temp.	467.86	°F
Pressure	291.2	psig
ΔP	45	psi
Recycle	1.2	mol/mol

SELECTIVITIES	
ALPHA	0.92
CH4 SEL (Carbon %)	6.5
CO2 SEL (Carbon %)	34.34
USAGE RATIO	0.689

H2 Removed	
nm3/h	-2251
SCFD	-2.02E+06

CO Conversions (%)			
	To HC	To CO2	Tot CO conv
Per Pass	33.2	17.37	50.57
Total	47	24.58	71.58

PRODUCTIVITY	
Avg FT rate in cc CO/cc cat/h)	28

Partial Pressures (atma)			
	Inlet	Outlet	Kin Avg
H2	3.23	2.29	2.67
CO	1.93	0.86	1.39
H2O	0	0.27	0.2
CO2	14.64	13.42	14.2
Total P (atm)	20.81	17.81	

SECOND STAGE PRODUCTION		
Naphtha	BPD	82
Diesel	BPD	221
Total BPD		304

H2/CO	F Feed	Tot Feed	Rx outlet
	1.4	1.672	2.678

DRY TAIL GAS (TO NEXT STAGE)			
Flows			
	nm3/h	scfd	Mol %
H2	5175	4635674	15.1
CO	1621	1452426	4.73
CO2	25684	23008815	74.95
H2O	0	0	0
INERTS	701	628154	2.05
CH4	1064	952971	3.1
C2-C4	24	21517	0.07
TOT	34269	3.07E+07	100
Equiv H2O nm3/h and scfd	10690	9577011	

REACTORS DESIGN – IRON CASE

FISCHER-TROPSCH REACTORS		
DATA	1st Stage	2nd Stage
Tubes ID, in	2.067	2.067
Tubes length, ft	39.37	39.37
Number tubes	8935	3696
Reactor OD, ft	29.89	19.38
Fresh Feed, SCFH	3.07E+06	1.48E+06
Total feed, SCFH	6.92E+06	3.25E+06
Recycle, SCFH	3.84E+06	1.77E+06
LV, cm/s	20.16	27.29
Tot feed/tube, cc/hr	9.22E+06	1.07E+07
Cat vol/tube, ft3	0.92	0.92
Tot Rx cat vol, ft3	8197	3391
Total Plant Catalyst, ft3	11588	
Indiv tube wt, lb	28.7	28.7
Total tubes wt, ton	144.75	59.88
Rx shell wt, ton	151.93	65.18
Tube sheets tickn, ft	1	1
Each tube sheet wt,ton	80.75	34.48
Each Rx head,ton	43.99	12.29
*Approx total Rx wt, ton	635	254

**Includes 20% addition for support structures*

6.2.3. Design data for the cobalt case with initial-shift CO₂ not removed

FIRST STAGE - COBALT

Feed : SG plus added H2	
Flow (nm3/h)	87215
SCFD	7.81E+07
H2/CO in	1.826
SYNGAS	Comp (%)
H2	45.82
CO	25.1
CH4	0.08
CO2	28.16
H2O	0.17
N2	0.46
C2-C4	0.22
Totals	100

OPERATING PARAMETERS		
Temperature	397.35	°F
Pressure	350	psig
ΔP	45	psi
Recycle	0	mol/mol

SELECTIVITIES	
ALPHA	0.92
CH4 SEL (Carbon %)	9
CO2 SEL (Carbon %)	1
USAGE RATIO	2.122

H2 Added	
nm3/h	0
SCFD	0.00E+00

CO Conversions (%)	
Per Pass	64
Total	64

PRODUCTIVITY	
Avg FT rate in cc CO/cc cat/h)	210

	Partial Pressures (atma)		
	Inlet	Outlet	Kin Avg
H2	11.37	3.75	7.83
CO	6.23	2.89	4.67
H2O	0	5.09	2.33
CO2	6.99	9.05	7.94
Total P (atm)	24.81	21.75	

FIRST STAGE PRODUCTION		
Naphtha	BPD	420
Diesel	BPD	1086
Total BPD		1506

H2/CO	F Feed	Tot Feed	Rx outlet
	1.826	1.826	1.299

DRY TAIL GAS (TO NEXT STAGE)			
Flows			
	nm3/h	scfd	Mol %
H2	10239	9172350	22.82
CO	7881	7059995	17.57
CO2	24697	22124358	55.05
H2O	0	0	0
INERTS	585	524050	1.3
CH4	1327	1189149	2.96
C2-C4	135	121365	0.3
TOT	44864	4.02E+07	100
Equiv H2O nm3/h and scfd	13880	12434449	

SECOND STAGE – COBALT

Feed : SG plus added H2

Flow (nm3/h)	44864
SCFD	40188617
H2/CO in	1.299
SYNGAS	Comp (%)
H2	22.82
CO	17.57
CH4	2.96
CO2	55.05
H2O	0
N2	1.3
C2-C4	0.3
Totals	100

OPERATING PARAMETERS

Temperature	400	°F
Pressure	290	psig
ΔP	45	psi
Recycle	0	mol/mol

SELECTIVITIES

ALPHA	0.9
CH4 SEL (Carbon %)	10
CO2 SEL (Carbon %)	1
USAGE RATIO	2.141

H2 Added

nm3/h	5030
SCFD	4.51E+06

CO Conversions (%)

Per Pass	76
Total	76

PRODUCTIVITY	
Avg FT rate in cc CO/cc cat/h)	134

	Partial Pressures (atma)		
	Inlet	Outlet	Kin Avg
H2	6.34	1.14	3.84
CO	3.27	0.88	2.13
H2O	0	2.73	1.32
CO2	10.26	11.51	10.99
Total P (atm)	20.73	17.67	

SECOND STAGE PRODUCTION		
Naphtha	BPD	180
Diesel	BPD	464
Total BPD		644

H2/CO	F Feed	Tot Feed	Rx outlet
	1.937	1.937	1.294

DRY TAIL GAS (TO NEXT STAGE)			
Flows			
	nm3/h	scfd	Mol %
H2	2448	2192802	7.69
CO	1891	1694442	5.94
CO2	24757	22178013	77.8
H2O	0	0	0
INERTS	720	645415	2.26
CH4	1926	1725704	6.05
C2-C4	78	69640	0.24
TOT	31820	2.85E+07	100
Equiv H2O nm3/h and scfd	5870	5258239	

REACTORS DESIGN – COBALT CASE

FISCHER-TROPSCH REACTORS DATA		
	1st Stage	2nd Stage
Tubes ID, in	1.049	1.049
Tubes length, ft	33	33
Number tubes	12000	8000
Reactor OD, ft	19.33	15.86
Fresh Feed, SCFH	3.26E+06	1.86E+06
Total feed, SCFH	3.26E+06	1.86E+06
Recycle, SCFH	0	0
LV, cm/s	25.46	26.21
Tot feed/tube, cc/hr	7.27E+06	6.24E+06
Cat vol/tube, ft3	0.2	0.2
Tot Rx cat vol, ft3	2377	1584
Total Plant Catalyst, ft3		3961
Indiv tube wt, lb	12.39	12.39
Total tubes wt, ton	162.95	108.64
Rx shell wt, ton	46.99	31.84
Tube sheets tickn, ft	1	1
Each tube sheet wt,ton	29.36	19.8
Each Rx head,ton	9.8	5.49
*Approx total Rx wt, ton	240	158

**Includes 20% addition for support structures*

6.3. Optimum Plant Design for the iron and cobalt catalysts with removal of the initial-shift CO₂ from the syngas feed

The raw syngas feed used in this case is the one shown in Table 6.1. The CO₂ formed during the shift to increase the H₂/CO ratio to various levels is removed and therefore the syngas feed is now richer in CO and H₂.

Optimized Design for the iron catalyst

At this point, it is not possible to predict which H₂/CO ratio in the fresh feed syngas would be optimum for an iron FT catalyst. Because of this, 10 different cases were selected for further study within the constraints of linear velocity, reaction rate and water partial pressure described in Section 4.1 of this report.

The syngas compositions for these 10 cases are shown in Table 6.4. Notice that the CO₂ formed during the shift as well as the extra water in the shift reactor have been removed from the shifted syngas feeds.

Table 6.4. Syngas feeds compositions and flows after shift and CO ₂ removal												
Component	Vol %	m3/h	m3/h	m3/h	m3/h	m3/h	m3/h	m3/h	m3/h	m3/h	m3/h	m3/h
H ₂ /CO ratio	Raw feed	0.75	0.87	1.00	1.11	1.25	1.40	1.50	1.60	1.70	1.80	
H ₂	30.76	21130	28681	31137	33451	35130	37159	39018	40131	41161	42114	42994
CO	66.61	45758	38207	35751	33437	31758	29729	27870	26757	25727	24774	23894
CH ₄	0.1	66	66	66	66	66	66	66	66	66	66	66
CO ₂	1	690	690	687	687	687	687	687	687	687	687	687
H ₂ O	0.68	464	464	464	464	464	464	464	464	464	464	464
N ₂	0.58	397	397	397	397	397	397	397	397	397	397	397
Ar	0.27	188	188	188	188	188	188	188	188	188	188	188
Totals	100	68693	68693	68690	68690	68690	68690	68690	68690	68690	68690	68690
CO shifted, m3/h			7551	10007	12321	14000	16029	17888	19001	20031	20984	21864
H ₂ formed, m3/h			7551	10007	12321	14000	16029	17888	19001	20031	20984	21864
CO ₂ removed, m3/h			7551	10011	12324	14003	16033	17891	19005	20034	20987	21867
H ₂ O used , lb/h			13350	17692	21781	24750	28338	31623	33592	35411	37096	38652

Fischer-Tropsch units were designed for each of the 10 cases shown in Table 6.4, following the same limitations on reactor operating conditions, tube diameter and

maximum water partial pressure as in the previous case (without removal of the CO₂ formed during the shift process).

The results are shown in Table 6.5 below. The total barrels produced only shows a marginal decrease as the H₂/CO ratio increases from about 0.75 to 1.1, while the required amount of catalyst decreases by more than 10 %. Because of these considerations, the H₂/CO ratio selected is again 1.11.

Table 6.5 Optimization of the H₂/CO ratio for iron on the FT section - With Removal of the CO₂ formed during shift										
H ₂ /CO after shift	0.75	0.87	1.00	1.11	1.25	1.40	1.50	1.60	1.70	1.80
Production,bpd										
1 st Stage	1861	1900	1812	1821	1684	1635	1643	1583	1527	1504
2 nd Stage	440	382	476	460	579	623	599	613	617	594
Plant Total	2301	2282	2289	2281	2263	2257	2242	2196	2144	2097
Recycle Ratio										
1 st Stage	1.00	1.00	1.50	1.60	1.80	2.05	2.20	2.20	2.20	2.20
2 nd Stage	1.00	1.00	1.00	1.20	1.20	1.20	1.40	1.40	1.40	1.50
Number tubes (2" ID)										
1 st Stage	6600	6600	5650	5800	6400	6800	7200	7200	7200	7200
2 nd Stage	3700	2900	3300	3000	3500	3365	3150	3192	3200	3100
Total tubes	10300	9500	8950	8800	9900	10165	10350	10392	10400	10300
ft ³ of catalyst										
1 st Stage	6055	6055	5184	5321	5872	6239	6606	6606	6606	6606
2 nd Stage	3395	2661	3028	2752	3211	3087	2890	2928	2936	2844
Total ft ³ cat per plant	9450	8716	8211	8073	9083	9326	9495	9534	9541	9450

Optimized design for the cobalt catalyst

As mentioned in the previous section, the optimization for the plant design running with a cobalt catalyst is much easier than that for an iron catalyst

The optimization in this case has to consider the minimum H₂/CO ratio acceptable at the outlet of the reactors (min 1.11) as well as the maximum water partial pressure (5.5 atm) and acceptable linear velocity (between 20 and 30 cm/s). The results are shown in Table 6.6 below. According to table 1.2, a cobalt based plant compares with an iron based one as follows:

- Cobalt makes about 2,140 bpd. This is lower than the amount that iron would produce using an H₂/CO after shift of 1.1 by about 140 bpd. Further attempts to optimize this cobalt case only succeeded in adding another 48 bpd by further optimization of the amount of CO shifted and the use of the minimum H₂/CO acceptable ratio of 1.11 at the exit of the cobalt reactors in the first and second stages. The optimized cobalt case would still make about 90 bpd less than the optimized iron case when the CO₂ is removed after the shift section.
- The iron catalyst would need more than twice as much tonnage for the FT reactors than cobalt.
- The first stage of the cobalt plant now needs a recycle ratio of 0.55 mol/mol with the initial-shift CO₂ removed from the syngas. This is because the syngas feed is now richer (no CO₂) and the water partial pressure has to be kept below 5.5 atm. For the same reason, the iron now needs a higher degree of recycle as compared with the case where no CO₂ was removed. The necessary recycle for the first stage is now 1.6 (vs 1.25, Table 6.2) and for the second stage is now 1.2 (vs 0.4, Table 6.2).
- The recycle ratios needed for the iron case are much higher than those needed for cobalt (1.6 for iron vs. 0.55 for cobalt in their respective first stage FT reactors).
- The volume of iron catalyst is now about 8,073 ft³ (an improvement over the 10,734 ft³ needed when CO₂ was not removed). The volume of catalyst for the cobalt case is about 2,793 ft³, much lower than that needed for iron and an improvement over the 3,965 ft³ needed for the previous case, without CO₂ removal.

Table 6.6. Optimized cobalt design	
Production,bpd	
1 st Stage	1811
2 nd Stage	332
Plant Total	2143
Recycle Ratio	
1 st Stage	0.55
2 nd Stage	0
Number tubes (1" ID)	
1 st Stage	11800
2 nd Stage	2300
Total tubes	14100
ft ³ of catalyst	
1 st Stage	2337
2 nd Stage	456
Total ft ³ cat per plant	2793
CO conversions	

Tot CO conv., %	93.49
HC and CO ₂ Selectivity (%)	
% CO to hydrocarbons	99
% CO to CO ₂	1

6.3.1. Design Data for the iron and cobalt optimized cases with removal of the initial-shift CO₂ from the syngas feed

6.3.2. Design data for the iron case

The data-column in Table 6.5 selected is the one of the H₂/CO in the fresh feed is equal to 1.11. The total reactor weight includes an additional 20%, which includes necessary structural components for transportation and mounting.

The design data for this case is shown in more detail in the series of tables below.

FIRST STAGE - IRON

Feed : SG after shift and CO₂ removal

Flow (nm ³ /h)	68690
SCFD	6.15E+07
H ₂ /CO in	1.106
SYNGAS	Comp (%)
H ₂	51.14
CO	46.23
CH ₄	0.10
CO ₂	1.00
H ₂ O	0.68
N ₂ + Ar	0.85
C ₂ -C ₄	0.00
Totals	100.00

OPERATING PARAMETERS

Temperature	462.00	°F
Pressure	350.00	psig
ΔP	44.1	psig
Recycle	1.60	mol/mol

SELECTIVITIES

Alpha Value	0.94
CH ₄ Sel (%)	5
CO ₂ Sel (%)	32.48
USAGE RATIO	1.090

CO Conversions (%)

	To HC	To CO ₂	Tot CO conv
Per Pass	23.18	11.15	34.33
Total	49.39	23.76	73.16

PRODUCTIVITY (FT + WGS)

Avg FT rate in cc CO/

cc cat/h 155

**Partial Pressures
(atma)**

	Inlet	Outlet	Kin Avg
H2	10.62	7.58	8.88
CO	9.40	6.58	8.09
H2O	0.00	1.28	0.87
CO2	3.73	5.09	4.34
Total P (atm)	24.81	21.81	

H2/CO	F Feed	Tot Feed	Rx outlet
	1.106	1.13	1.15

FIRST STAGE PRODUCTION

Naphtha		466
Diesel		1354
Total C5+	TOTAL	1821

DRY TAIL GAS (TO NEXT STAGE)**Flows**

	nm3/h	scfd	Mol %
H2	9812	8789141	34.90
CO	8525	7636974	30.32
CO2	8233	7374733	29.28
H2O	0	0	0.00
INERTS	585	524035	2.08
CH4	850	761726	3.02
C2-C4	111	99209	0.39
TOT	28116	2.52E+07	100.00

SECOND STAGE - IRON

**Feed : TG from stage 1 plus
2768 nm3/h of added
H₂**

Flow (nm3/h)	28116
SCFD	2.52E+07
H2/CO in	1.151
SYNGAS	Comp (%)
H2	34.90
CO	30.32
CH4	3.02
CO2	29.28
H2O	0.00
N2 + Ar	2.08
C2-C4	0.39
Totals	100.00

OPERATING PARAMETERS

Temperature	468	oF
Pressure	291.20	psig
ΔP	44.10	psi
Recycle	1.20	mol/mol

SELECTIVITIES

ALPHA	0.92
CH4 SEL (Carbon %)	6.50
CO2 SEL (Carbon %)	35.81
USAGE RATIO	0.629

CO Conversions (%)

	To HC	To CO2	Tot CO conv
Per Pass	31.88	17.78	49.66
Total	47.68	26.59	74.27

PRODUCTIVITY (FT + WGS)

Avg FT rate in cc CO/ cc cat/h	82
-----------------------------------	----

Partial Pressures (atma)

	Inlet	Outlet	Kin Avg
H2	5.61	4.03	4.64
CO	4.29	2.04	3.19
H2O	0.00	0.57	0.41
CO2	9.19	9.40	9.41
Total P (atm)	20.81	17.81	

H2/CO	F Feed	Tot Feed	Rx outlet
	1.151	1.31	1.98

SECOND STAGE PRODUCTION		
Naphtha	BPD	125
Diesel	BPD	336
Total BPD		460

DRY TAIL GAS (TO NEXT STAGE)			
	Flows		
	nm3/h	scfd	Mol %
H2	5830	5222710	12.64
CO	2194	1965038	4.96
CO2	10500	9405575	63.76
H2O	0	0	0.00
INERTS	585	524035	0.12
CH4	1115	998398	18.46
C2-C4	147	131822	0.05
TOT	20370	1.82E+07	100.00

TOTAL PLANT PRODUCTION

PLANT PRODUCTION		
Naphtha	BPD	591
Diesel	BPD	1690
Total BPD		2281

FISCHER-TROPSCH REACTORS DATA			
	1st Stage	2nd Stage	
Tubes ID, in	2.067	2.067	
Tubes length, ft	39.37	39.37	
Number tubes	5800	3000	

Reactor OD, ft	24.17	17.50
Fresh Feed, SCFH	2.56E+06	1.05E+06
Total feed, SCFH	6.67E+06	2.31E+06
Recycle, SCFH	4.10E+06	1.26E+06
LV, cm/s	29.92	23.89
Tot feed/tube, cc/hr	1.18E+07	9.37E+06
Cat vol/tube, ft3	0.92	0.92
Tot Rx cat vol, ft3	5321	2752
Total Plant Catalyst, ft3	8073	
Indiv tube wt, lb	29	29
Total tubes wt, ton	94	49
Rx shell wt, ton	100.21	53.5
Tube sheets tickn, ft	1.00	1.00
Each tube sheet wt,ton	53	28
Each Rx head,ton	23.51	9.13
*Approx total Rx wt, ton	404	205

**Includes 20% addition for support structures*

6.3.3. Design data for the cobalt case

The detailed design data for the cobalt case (Table 6.6) is shown in more detail in the tables below. The total reactor weight includes an additional 20%, which includes necessary structural components for transportation and mounting.

FIRST STAGE - COBALT

Feed : SG after shift and CO2 removal		OPERATING PARAMETERS		
Flow (nm3/h)	64359	Temperature	400.00	°F
SCFD	5.77E+07	Pressure	350.00	psig
H2/CO in	1.950	ΔP	45	psig
SYNGAS	Comp (%)	Recycle	0.55	mol/mol
H2	64.63	SELECTIVITIES		
CO	33.14	Alpha Value	0.92	
CH4	0.10	CH4 Sel (%)	9	
CO2	0.98	CO2 Sel (%)	1	
H2O	0.23			
N2	0.62			
Ar	0.29			
Totals	100.00			
CO Conversion (%)		PRODUCTIVITY		
Per Pass	50.88	Avg FT rate in cc CO/ cc cat/h		
Total	79.00	256		
Partial Pressures (atma)				
	Inlet	Outlet	Kin Avg	
H2	14.17	6.96	10.82	
CO	8.24	5.34	6.88	
H2O	0.00	5.47	2.47	
CO2	0.68	0.96	0.81	
Total P (atm)	24.81	21.75		

H2/CO	F Feed	Tot Feed	Rx outlet
	1.950	1.72	1.30

FIRST STAGE PRODUCTION		
Naphtha		505
Diesel		1306
Total C5+	TOTAL	1811

DRY TAIL GAS (TO NEXT STAGE)				
Flows				
	nm3/h	scfd	Mol %	
H2	5839	5230963	43.41	
CO	4479	4012721	33.30	
CO2	801	717964	5.96	
H2O	0	0	0.00	
INERTS	585	524038	4.35	
CH4	1583	1418142	11.77	
C2-C4	163	145969	1.21	
TOT	13451	1.20E+07	100.00	

SECOND STAGE - COBALT

Feed : TG from stage 1 plus 2768 nm3/h of added H ₂	
Flow (nm3/h)	16219
Flow (ft3/day)	1.37E+07
SCFD	1.45E+07
H ₂ /CO in	1.921
SYNGAS	Comp (%)
H ₂	53.07
CO	27.62
CH ₄	9.76
CO ₂	4.94
H ₂ O	0.00
N ₂ + Ar	3.61
C ₂ -C ₄	1.00
Totals	100.00

INPUTS		
Temperature	400.00	°F
Pressure	290.00	psig
ΔP	45	psig
Recycle	0.00	mol/mol

SELECTIVITIES	
Alpha Value	0.90
CH ₄ Sel (%)	10
CO ₂ Sel (%)	1

CO Conversion (%)	
Per Pass	69.00
Total	69.00

PRODUCTIVITY	
Avg FT rate in cc CO/ cc cat/h	241

	Partial Pressures (atma)		
	Inlet	Outlet	Kin Avg
H ₂	11.00	3.49	7.57
CO	5.72	2.43	4.22
H ₂ O	0.00	5.31	2.34
CO ₂	1.02	1.46	1.22
Total P (atm)	20.73	17.67	

H ₂ /CO	F Feed	Tot Feed	Rx outlet
	1.921	1.92	1.43

SECOND STAGE PRODUCTION	
Naphtha	93
Diesel	239
Total C5+	332

DRY TAIL GAS (TO NEXT STAGE)			
Flows			
	nm3/h	scfd	Mol %
H2	1991	1783396	29.59
CO	1389	1243943	20.64
CO2	832	745652	12.37
H2O	0	0	0.00
INERTS	585	524038	8.69
CH4	1892	1695019	28.12
C2-C4	40	35936	0.60
TOT	6729	6.03E+06	1.00E+02

TOTAL PLANT PRODUCTION		
PLANT PRODUCTION		
Naphtha	BPD	598
Diesel	BPD	1545
Total C5+		2143

FISCHER-TROPSCH REACTORS		
DATA		
	1st Stage	2nd Stage
Tubes ID, in	1.049	1.049
Tubes length, ft	33	33
Number tubes	11800	2300
Reactor OD, ft	19.18	8.70

Fresh Feed, SCFH	2.40E+06	6.05E+05
Total feed, SCFH	3.72E+06	6.05E+05
Recycle, SCFH	1.32E+06	
LV, cm/s	29.78	29.64
Tot feed/tube, cc/hr	5.45E+06	7.05E+06
	0	0
Cat vol/tube, ft3	0.20	0.20
Tot Rx cat vol, ft3	2337	456
Total Plant Catalyst, ft3		3961
Indiv tube wt, lb	12.40	12.40
Total tubes wt, ton	160	31
Rx shell wt, ton	46.24	9.86
Tube sheets tickn, ft	1.00	1.00
Each tube sheet wt,ton	29	6
Each Rx head,ton	9.56	0.92
*Approx total Rx wt, ton	235	45

**Includes 20% addition for support structures*

7.0 SIMPLIFIED PFD FOR THE OPTIMUM IRON AND COBALT CASES

The simplified PFDs are shown below. Notice that in order to satisfy all the limitations described in Section 4 the iron case must have two internal recycle compressors for both cases studied.

There are no internal recycle compressors for the cobalt case where the CO₂ has not been removed from the syngas. For the case where the CO₂ has been removed, recycle is necessary to maintain the partial pressure of water below 5.5 atm.

For H₂ plant management, the H₂/CO ratio after the shift for the cobalt catalyst is higher than the one needed as feed for the first stage. Since for cobalt the usage ratio is higher than the feed ratio, the H₂/CO ratio will decrease with conversion. Because of this, the hydrogen rich stream from the membrane after the shift reactor is added before the second cobalt FT Stage.

In the case of iron without CO₂ removal, and because its WGS activity will increase the H₂/CO ratio as conversion increases, it is necessary to add a hydrogen membrane between the first and second stages to lower the H₂/CO ratio in the feed to the second FT Stage. For the iron case with CO₂ removal, the H₂/CO only shows a small increase and therefore it is not necessary to add a membrane between the first and second stages

Both catalysts will have the same Fischer-Tropsch products, separation and upgrading equipment, although product upgrading for iron-based catalysts will consume more hydrogen due to the higher amount of olefins and oxygenates in their raw products.

First Case: Without CO₂ removal after Shift

Second Case: With CO₂ removal after Shift

8.0 IRON VS. COBALT (MAJOR EQUIPMENT, GENERAL LAYOUT AND COSTING)

This comparison employs a typical FT process configuration using a two-stage, fixed-bed reactor sequence with three stages of separation. As shown in the Process Flow Diagrams (PFDs) for each catalyst system, the FT products are collected as a HFTL (Heavy Fischer Tropsch Liquid), MFTL (middle Fischer Tropsch product) and LFTL (light Fischer Tropsch Liquid). The HFTL predominantly consists of the heavier (wax) (C20+) which must be hydroprocessed to diesel or jet fuel. The MFTL contains predominantly C10-C20 hydrocarbons (diesel range), which require isomerization to produce refined diesel products. The LFTL is predominantly less than C12 containing the naphtha (<C9). There is a significant overlap in carbon number between the collected liquids and distillation is required to secure the desired end points for the diesel and jet products. The equipment costs were derived from EST\$PRO version 3.7 for the exchangers, separators, and compressors.

The cost estimate is based on the modular cost of the major equipment followed by a multiplier representing the installed cost. The installed cost is an approximation covering the erection (set-up) as well as the tie-ins for piping as well as instrumentation costs. Compressors typically have a larger install factor due to the required foundation work, instrumentation and control systems. The install factors can vary with site location and how much pre-fabrication of instrumentations/controls is done in the shop as opposed to in the field. The sum of the factored costs is then multiplied by 1.1 in order to capture the piping and control systems, which are not part of individual units. This factor again is a gross approximation since the detailed process configuration has not yet been established. The cost estimates in this report are useful for a relative comparison. The Fe system in the comparison will carry additional costs over the Co system due to the added utilities associated with the larger cooling and heating duty, expressed in Btu/hour, of the syngas.

The reactors for both the Fe and Co cases were scaled up to a nominal 2,000 barrels per day from a 650 barrel per day “hard estimate” (e.g., costs are on a “as quoted” basis) obtained from a third party fabricator. The first stage Fe based FT reactor is difficult to cost estimate due to the large size. The FT fixed-bed reactor costs cannot be accurately estimated using the standard shell/tube/TEMA ASTM vessel estimation models due to their large size. Install factors of 2.8 to 4.0 are used for the process equipment recognizing that site location, construction methods and transport vectors can significantly affect this factor. These factors are ones generally accepted by the engineering and design community for cost estimating purposes.

Both catalyst systems used representative activity and product selectivity as described above. The products made with the iron catalyst FT system are more olefinic (>15%) and straight run diesel range hydrocarbons are not readily blendable due to stability and pour point issues. Hydrotreating is required for the production of either a blend stock or neat fuel (i.e. ASTM D975). The cobalt catalyst produces predominantly fewer olefins and oxygenates and the straight run diesel is compatible with blending with existing specified fuels. The C20+ products (HFTL) require hydrocracking to make blendable or neat fuels. The LFTL has marginal markets in solvent, blending or refinery feed stock

and in both catalyst cases, LFTL distillation is required to maximize the recovery of diesel and jet fuels.

The FT reactor is the predominant cost in the hydrocarbon synthesis section. (Note that the capital costs for FT reactors are for the reactors themselves, not including any FT catalyst, for either the cobalt or the iron case. Either type of catalyst charge will need to be changed periodically, so FT catalyst is considered here to be an operating cost rather than a capital cost.) For the 2,000 barrel per day throughput specified in this project, a tube/jacket heat exchanger was employed utilizing a conventional boiler system. The Fe system utilized 2" ID tubes requiring shell diameters on the order of at least 24 feet for the first stage. The larger sizes for the first Fe stages limits the number of fabrication shops and cost/practicality of transport (>400 tons). The cost estimates for the larger vessels are dependent upon both of these factors. The use of reactors with a 16' diameter or less allows more competitive bidding and somewhat reduces the complexities of transport.

9.0 IRON CASE WITHOUT CO₂ REMOVAL (MAJOR EQUIPMENT AND COSTING)

Vessels		Install Factor :		2.8	Installed Cost :	\$1,081,780
Equipment No.	Name	Diameter (ft)	Length (ft)	Volume (ft ³)	Cost (USD)	Description
20-V-100	#1 HFTL Separator	10	14	1361.3	\$109,196	Horizontal Separator
20-V-101	#1 MFTL Separator	8	14	837.8	\$65,729	3-Way Separator
20-V-102	#1 LFTL Separator	8	14	837.8	\$64,707	3-Way Separator
30-V-100	#2 HFTL Separator	8	12	737.2	\$48,979	Horizontal Separator
30-V-101	#2 MFTL Separator	8	14	837.8	\$50,134	3-Way Separator
30-V-102	#2 LFTL Separator	8	14	837.8	\$47,605	3-Way Separator

Reactors		Install Factor :		3.2	Installed Cost :	\$25,551,212
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description		
20-R-100	#1 FT Reactor	194,574,364	\$5,394,027	Fix-Bed Reactor		
30-R-100	#2 FT Reactor	26,433,183	\$2,590,727	Fix-Bed Reactor		

Heaters			Install Factor :	3.0	Installed Cost :	\$2,164,737
Equipment No.	Name	Duty btu/hr	Cost (USD)	Description		
10-E-100	FT #1 Preheater	63,511,556	\$524,726	Heat Exchanger		
10-E-101	FT #2 Preheater	27,708,507	\$196,853	Heat Exchanger		

Coolers			Install Factor :	2.5	Install Cost :	\$4,706,058
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description		
20-E-100	#1 MFTL Chiller	18,829,565	\$241,624	Heat Exchanger		
20-E-101	#1 LFTL Chiller	12,701,321	\$1,111,663	Heat Exchanger		
30-E-100	#2 MFTL Chiller	5,413,613	\$79,591	Heat Exchanger		
30-E-101	#2 LFTL Chiller	6,549,709	\$449,545	Heat Exchanger		

Compressors			Install Factor :	4.0	Total Cost :	\$7,687,652
Equipment No.	Name	Power (hp)	Cost (USD)	Description		
10-K-100	FT #1 Recycle Compressor	2100	\$1,114,820	Reciprocating Compressor		
10-K-101	FT #2 Recycle Compressor	1200	\$807,093	Reciprocating Compressor		

Air Fin Coolers			Install Factor :	2.5	Total Cost :	\$1,766,678
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description		
20-AC-100	FT #1 Air Fin	50,721,473	\$476,856	Air Cooler		
30-AC-100	FT #2 Air Fin	20,747,032	\$229,815	Air Cooler		

Total Install Cost : \$42,958,117

Total Install plus Piping & Instrumentation : \$51,549,740

(Estimate excludes civil/utilities and support equipment/structures.)

10.0 COBALT WITHOUT CO₂ REMOVAL (MAJOR EQUIPMENT AND COSTING)

Vessels			Install Factor :	2.8	Install Cost :	\$876,694
Equipment No.	Name	Diameter (ft)	Length (ft)	Volume (ft ³)	Cost (USD)	Description
20-V-100	#1 HFTL Separator	8	12	737.2	\$66,631	Horizontal Separator
20-V-101	#1 MFTL Separator	8	10	636.7	\$59,266	3-Way Separator
20-V-102	#1 LFTL Separator	6	10	339.3	\$36,487	3-Way Separator
30-V-100	#2 HFTL Separator	10	12	1204.3	\$67,901	Horizontal Separator
30-V-101	#2 MFTL Separator	8	10	636.7	\$43,385	3-Way Separator
30-V-102	#2 LFTL Separator	8	10	636.7	\$39,435	3-Way Separator

Reactors			Install Factor :	3.2	Install Cost :	\$13,586,924
Equipment No.	Name	Duty (BTU/hr)	Cost (USD)	Description		
20-R-100	#1 FT Reactor	92,471,602	\$2,471,927	Fixed-Bed Reactor		
30-R-100	#2 FT Reactor	38,715,593	\$1,773,987	Fixed-Bed Reactor		

Heaters			Install Factor :	3.0	Install Cost :	\$408,351.00
Equipment No.	Name	Duty (BTU/hr)	Cost (USD)	Description		
10-E-100	FT #1 Preheater	14,389,212	\$74,676	Heat Exchanger		
10-E-101	FT #2 Preheater	13,290,000	\$61,441	Heat Exchanger		

Coolers			Install Factor :	2.5	Install Cost :	\$2,074,013
Equipment No.	Name	Duty (BTU/hr)	Cost (USD)	Description		
20-E-100	#1 HFTL Chiller	4,974,353	\$31,680	Heat Exchanger		
20-E-101	#1 MFTL Chiller	7,884,903	\$285,134	Heat Exchanger		
20-E-102	#1 LFTL Chiller	2,251,396	\$141,598	Heat Exchanger		

30-E-100	#1 HFTL Chiller	3,433,833	\$25,327	Heat Exchanger
30-E-101	#2 MFTL Chiller	6,578,358	\$225,768	Heat Exchanger
30-E-102	#2 LFTL Chiller	1,870,493	\$120,098	Heat Exchanger

Air Fin Coolers		Install Factor : 2.5		Install Cost : \$1,202,438
Equipment No.	Name	Duty (BTU/hr)	Cost (USD)	Description
20-AC-100	FT #1 Air Fin	25,461,671	\$305,915	Air Cooler
30-AC-100	FT #2 Air Fin	10,168,776	\$175,060	Air Cooler

Total Installed Cost : \$18,148,419

Total Install plus Piping & Instrumentation : \$21,778,103

(Estimate excludes civil/utilities and support equipment/structures.)

11.0 IRON WITHOUT CO₂ REMOVAL (GENERAL LAYOUT)

(See Appendix for more Detail)

(See Appendix for more Detail)

12.0 COBALT WITHOUT CO₂ REMOVAL (GENERAL LAYOUT)

(See Appendix for more Detail)

(See Appendix for more Detail)

13.0 IRON WITH CO₂ REMOVAL (MAJOR EQUIPMENT AND COSTING)

Vessels		Install Factor :		2.8	Installed Cost :	\$1,143,293
Equipment No.	Name	Diameter (ft)	Length (ft)	Volume (ft ³)	Cost (USD)	Description
20-V-100	#1 HFTL Separator	12	16	6616.2	\$123,296	Horizontal Separator
20-V-101	#1 MFTL Separator	14	16	3181.4	\$70,497	3-Way Separator
20-V-102	#1 LFTL Separator	12	16	2261.9	\$65,584	3-Way Separator
30-V-100	#2 HFTL Separator	12	8	1357.2	\$49,645	Horizontal Separator
30-V-101	#2 MFTL Separator	10	8	890.1	\$50,927	3-Way Separator
30-V-102	#2 LFTL Separator	10	8	890.1	\$48,370	3-Way Separator

Reactors		Install Factor :		3.2	Installed Cost :	\$19,012,461
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description		
20-R-100	#1 FT Reactor	119,000,000	\$3,754,813	Fix-Bed Reactor		
30-R-100	#2 FT Reactor	34,170,000	\$2,186,581	Fix-Bed Reactor		

Heaters		Install Factor :		3.0	Installed Cost :	\$1,345,662
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description		
10-E-100	FT #1 Preheater	45,860,000	\$297,250	Heat Exchanger		
10-E-101	FT #2 Preheater	20,660,000	\$151,304	Heat Exchanger		

Coolers		Install Factor :		2.5	Install Cost :	\$3,504,463
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description		
20-E-100	#1 MFTL Chiller	15,480,000	\$197,393	Heat Exchanger		
20-E-101	#1 LFTL Chiller	10,140,000	\$803,594	Heat Exchanger		
30-E-100	#2 MFTL Chiller	5,170,000	\$77,237	Heat Exchanger		
30-E-101	#2 LFTL Chiller	5,104,000	\$323,561	Heat Exchanger		

Compressors			Install Factor :	4.0	Total Cost :	\$7,419,440
Equipment No.	Name	Power (hp)	Cost (USD)	Description		
10-K-100	FT #1 Recycle Compressor	1,995	\$1,082,271	Reciprocating Compressor		
10-K-101	FT #2 Recycle Compressor	1,116	\$772,589	Reciprocating Compressor		

Air Fin Coolers			Install Factor :	2.5	Total Cost :	\$1,390,195
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description		
20-AC-100	FT #1 Air Fin	38,140,000	\$361,865	Air Cooler		
30-AC-100	FT #2 Air Fin	15,470,000	\$194,213	Air Cooler		

Total Install Cost : \$33,815,513

Total Install plus Piping & Instrumentation : \$40,578,616

(Estimate excludes civil/utilities and support equipment/structures.)

14.0 COBALT WITH CO₂ REMOVAL (MAJOR EQUIPMENT AND COSTING)

Vessels			Install Factor :	2.8	Install Cost :	\$890,092
Equipment No.	Name	Diameter (ft)	Length (ft)	Volume (ft ³)	Cost (USD)	Description
20-V-100	#1 HFTL Separator	10	16	1518.4	\$65,161	Horizontal Separator
20-V-101	#1 MFTL Separator	8	16	938.3	\$57,883	3-Way Separator
20-V-102	#1 LFTL Separator	8	8	536.2	\$35,925	3-Way Separator
30-V-100	#2 HFTL Separator	6	8	282.7	\$62,404	Horizontal Separator
30-V-101	#2 MFTL Separator	4	8	117.3	\$42,984	3-Way Separator
30-V-102	#2 LFTL Separator	4	8	117.3	\$53,533	3-Way Separator

Reactors			Install Factor :	3.2	Install Cost : \$9,896,601
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description	
20-R-100	#1 FT Reactor	116,700,000	\$2,437,947	Fixed-Bed Reactor	
30-R-100	#2 FT Reactor	21,780,000	\$654,741	Fixed-Bed Reactor	

Heaters			Install Factor :	3.0	Install Cost : \$253,806
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description	
10-E-100	FT #1 Preheater	6,908,000	\$56,165	Heat Exchanger	
10-E-101	FT #2 Preheater	3,892,000	\$28,437	Heat Exchanger	

Coolers			Install Factor :	2.5	Install Cost : \$1,611,990
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description	
20-E-100	#1 HFTL Chiller	5,441,000	\$32,973	Heat Exchanger	
20-E-101	#1 MFTL Chiller	8,975,000	\$335,592	Heat Exchanger	
20-E-102	#1 LFTL Chiller	2,369,000	\$145,185	Heat Exchanger	
30-E-100	#1 HFTL Chiller	1,009,000	\$14,329	Heat Exchanger	
30-E-101	#2 MFTL Chiller	1,874,000	\$76,621	Heat Exchanger	
30-E-102	#2 LFTL Chiller	426,000	\$40,096	Heat Exchanger	

Air Fin Coolers			Install Factor :	2.5	Install Cost : \$1,195,150
Equipment No.	Name	Duty (btu/hr)	Cost (USD)	Description	
20-AC-100	FT #1 Air Fin	31,440,000	\$370,923	Air Cooler	
30-AC-100	FT #2 Air Fin	5,578,000	\$107,137	Air Cooler	

Compressors			Install Factor :	4.0	Install Cost : \$2,565,220
Equipment No.	Name	Power (hp)	Cost (USD)	Description	
10-K-100	FT #1 Recycle Compressor	810	\$641,305	Reciprocating Compressor	

Total Installed Cost : \$13,847,642
Total Install plus Piping & Instrumentation : \$16,617,170

(Estimate excludes civil/utilities and support equipment/structures.)

15.0 IRON WITH CO₂ REMOVAL (GENERAL LAYOUT)

(See Appendix for more Detail)

(See Appendix for more Detail)

16.0 COBALT WITH CO₂ REMOVAL (GENERAL LAYOUT)

(See Appendix for more Detail)

(See Appendix for more Detail)

References

1. Malone, Donald P., and Holcombe, Thomas C., "Design and Performance Simulation of a Clean-up and Water-Gas-Shift System for Syngas Derived from Coal and Biomass," Project DE-FC26-06NT42449 Milestone Report for Tasks 3 and 4, Feb 2010.
2. Tarka, Thomas J., P.E., et. al., "Affordable, Low-Carbon Diesel Fuel from Domestic Coal and Biomass," US Department of Energy, National Energy Technology Laboratory Report DOE/NETL-2009/1349, January 14, 2009.
3. R.B. Anderson, The Fischer-Tropsch Synthesis, Academic Press 1984.
4. Frederick Gideon Botes, Kinetic and Selectivity Modeling of the Iron-Based Low-Temperature Fischer-Tropsch Synthesis, Chapter 5: Water-Gas-Shift Kinetics in the Fe-LTFT Synthesis. Doctoral Thesis, Technical University of Eindhoven, 2008
5. I.C. Yates and C.N. Satterfield, Energy and Fuels, 5, p. 168, 1991.
6. Van Berge, P. J.; Everson, R. C. Cobalt as an alternative Fischer-Tropsch catalyst to iron for the production of middle distillates. Stud. Surf. Sci. Catal. 1997, 107 (Natural Gas Conversion IV), p.207-212.
7. Burtron H. Davis, Fischer-Tropsch Synthesis: Comparison of Performances of Iron and Cobalt Catalysts, Ind. Eng. Chem. Res. 2007, 46, p.8938-8945.
8. Fischer-Tropsch (FT) synthesis, 6:791, 827; 12:431 of hydrocarbons, 10:532, 621; 13:768.
Kirk-Othmer Encyclopedia of Chemical Technology, John Wiley & Sons, Inc. Fifth Edition, March 2007.

Appendix

Section 11.0 Detail - PFD Iron without CO₂ removal (General Layout)

Section 12.0 Detail - PFD Cobalt without CO₂ removal (General Layout)

Section 15.0 Detail - PFD Iron with CO₂ removal (General Layout)

Section 16.0 Detail - PFD Cobalt with CO₂ removal (General Layout)

APPENDIX 8
CTL Plant Site Evaluation & Feasibility Analysis

FINAL PROJECT REPORT

March 14, 2011
Purchase Order No. VH56760

TASK 9

Prepared for

2550 Huntington Avenue
Alexandria, VA 22303-1499

41150 Technology Park Drive, Suite 101
Sterling Heights, MI 48314

By

Smith Management Group
1405 Mercer Road
Lexington KY 40511

ACKNOWLEDGEMENT

This material is based upon work supported by the U.S. Department of Energy's National Energy Technology Laboratory (U.S. DOE-NETL) under cooperative agreement DE-FC26-06NT42449.

DISCLAIMER

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

ABSTRACT

Development of a robust coal to liquid (CTL) fuel industry is critically important in the move towards energy independence. The United States has large deposits of coal which have historically been mined at reasonable costs. In addition, the world-wide move toward control and reduction of greenhouse gases is driving the use of coal for energy production toward newer technologies which are better able to capture and store or sequester CO₂.

Any industrial development must comply with many layers of environmental and governmental regulation and control. The CTL industry will be one of the more heavily regulated industries, due to its potential emissions, especially of greenhouse gases. CTL projects will be further delayed by protests and legal action from groups ranging from NIMBYs (not in my back yard) to those interested in stopping the use of coal.

Development of a road map and analysis of the permitting requirements applicable to the development of a commercial-scale CTL facility should help prospective developers make good decisions about schedules, costs and the economic feasibility of their proposed projects. Understanding the likely time and cost requirements of planning, siting and permitting a CTL facility is critical in any developer's decision matrix.

Due to the constant changes in the status of regulatory and legislative developments in this field, with limited exception a cut-off date of September 30, 2010 has been applied to enable this report to be completed. This decision means that portions of this report should be updated in order to remain viable.

TABLE OF CONTENTS

1.0	OBJECTIVE.....	1
2.0	INTRODUCTION	2
3.0	EXECUTIVE SUMMARY	4
4.0	METHODS USED TO GATHER AND ANALYZE INFORMATION	6
5.0	GENERAL PROCESS DESCRIPTION	8
6.0	RESOURCE REQUIREMENTS AND PROCESS OUTPUTS	13
6.1	Process Resources and Inputs	13
6.2	Process Outputs	14
6.2.1	Solid By-Products	15
6.2.2	Gaseous By-Products and Air Emissions	15
6.2.2.1	SO ₂ Emissions.....	17
6.2.2.2	Particulate Emissions	17
6.2.2.3	NO _x Emissions	18
6.2.2.4	CO Emissions.....	19
6.2.2.5	Lead Emissions	19
6.2.2.6	Hazardous Air Pollutants	20
6.2.2.7	Mercury.....	21
6.2.2.8	Greenhouse Gas Emissions	21
6.2.3	Liquid By-Products and Wastewater Discharges.....	22
7.0	GENERAL SITE REQUIREMENTS.....	24
8.0	ENVIRONMENTAL AUTHORIZATION REQUIREMENTS.....	26
8.1	Site Planning and Development.....	26
8.1.1	National Environmental Policy Act (NEPA)	27
8.1.1.1	Influence on Plant Siting, Design and Operation	28
8.1.2	Local Zoning Board or Planning Commission	29
8.1.2.1	Influence on Plant Siting, Design and Operation	29
8.2	Air Emissions	29
8.2.1	Kentucky Division for Air Quality (DAQ).....	32
8.2.2	Permitting	32
8.2.2.1	PSD Requirements.....	33
8.2.2.2	Nonattainment Area Requirements	34
8.2.2.3	Federal Class I Areas	35
8.2.2.4	NSPS and NESHAPS Requirements.....	35
8.2.2.5	GHG Emissions Requirements.....	37
8.2.2.6	Risk Management Plan.....	38
8.2.2.7	Influence on Plant Siting, Design and Operation	38
8.3	Waste Management.....	41

8.3.1	Kentucky Division of Waste Management (DWM)	42
8.3.2	Special Waste Permit	43
8.3.2.1	Influence on Plant Siting, Design and Operation	44
8.3.3	Hazardous Waste Management	46
8.3.3.1	Influence on Plant Siting, Design and Operation	47
8.3.4	Brownfield Developments	48
8.3.4.1	Influence on Plant Siting, Design and Operation	48
8.4	Water Supply and Discharges.....	49
8.4.1	U.S. Army Corps of Engineers (USACE)	50
8.4.1.1	Influence on Plant Siting, Design and Operation	51
8.4.2	Kentucky Division of Water (DOW).....	52
8.4.3	Stream and Floodplain Construction	52
8.4.3.1	Influence on Plant Siting, Design and Operation	53
8.4.4	Water Quality Certification	54
8.4.4.1	Influence on Plant Siting, Design and Operation	55
8.4.5	Water Withdrawal Permit.....	56
8.4.5.1	Influence on Plant Siting, Design and Operation	56
8.4.6	Potable Water	57
8.4.6.1	Influence on Plant Siting, Design and Operation	57
8.4.7	Kentucky Pollution Discharge Elimination System (KPDES)	58
8.4.7.1	Construction Activity Stormwater Discharges	58
8.4.7.1.1	Influence on Plant Siting, Design and Operation.....	59
8.4.7.2	Process Wastewater and Stormwater Discharges.....	60
8.4.7.2.1	Influence on Plant Siting, Design and Operation.....	62
8.4.7.3	Sanitary Wastewater Discharges.....	63
8.4.7.3.1	Influence on Plant Siting, Design and Operation.....	63
8.5	Electric Service, Generation and Transmission	64
8.5.1	Electric Service	64
8.5.2	Generation and Transmission.....	65
8.5.3	Influence on Plant Siting, Design and Operation	66
8.5.4	Kentucky Siting Board	67
8.5.4.1	Influence on Plant Siting, Design and Operation	69
9.0	REGULATORY AND PERMITTING TIME FRAMES - IMPACTS UPON THE ECONOMICS OF A PROJECT	71
9.1	Permitting Delays for Fossil Fuel Projects.....	72
9.2	Coal Fired Power Plants and Electricity Supply	78
9.3	Coal Gasification.....	80
10.0	POTENTIAL FOR GEOLOGIC SEQUESTRATION	82
11.0	LIABILITY, LEGISLATION, PARTNERSHIPS	84
11.1	Current Status of Carbon Dioxide Liability	86
11.2	Development of Legislation - Carbon Capture and Sequestration	89
11.2.1	States (Enacted).....	89
11.2.2	States (Proposed)	96

11.3 Selected International Legislative Developments	102
11.3.1 Summary - European Community	102
11.3.2 Summary - United Kingdom (UK)	107
11.3.3 Summary - Germany	108
11.3.4 Summary - Norway.....	108
11.3.5 Summary - Australia.....	110
11.4 Partnerships.....	114
11.5 Recent Developments in Kentucky	118
12.0 RECOMMENDATIONS	120

FIGURES

Figure 1	Process Flow Diagram
Figure 2	Conceptual Landfill Diagram

ATTACHMENTS

Appendix 1	Acronyms, Abbreviations and Nomenclature
Appendix 2	References and Sources
Appendix 3	Permit List
Appendix 4	Energy Contacts
Appendix 5	Permit Map
Appendix 6	Analysis of Carbon Capture and Sequestration Potential
Appendix 7	Kentucky Workgroup for Legal Issues of Carbon Sequestration
Appendix 8	Draft Report of the West Virginia CCS Workgroup
Appendix 9	House Bill 259, Senate Bill 50, 2011 Kentucky Legislative Session

1.0 OBJECTIVE

Work was performed to provide a generalized, public report on plant siting and permitting issues. The objective is to provide the coal-to-liquid (CTL) project developer tools to gain a better perspective of the information required for permit submittals, and guidance on timeframes for the preparation of information, regulatory review and approval.

The report is designed to identify and provide information on environmental permitting requirements and timeframes for designing, building and operating a commercial scale CTL plant, including those relevant to land use and zoning, coal handling and storage, transportation infrastructure requirements, control of water runoff and discharge, emissions to air and water, impact to jurisdictional waters and wetlands, among others.

Work was also performed to identify the impact of the choice of plant site on the permitting requirements for a CTL plant; and the impact of permitting requirements on the design, layout and operation of such a plant. Discussion of the economic impact of environmental compliance on the feasibility of building and operating a commercial plant has been included.

Strategies to manage carbon dioxide were examined, utilizing a specific site in South Shore, Kentucky as a test case. Carbon sequestration was explored in more general terms as well and the current, albeit ever-changing, status of legislation and regulation on carbon management was explored.

2.0 INTRODUCTION

This report describes work performed and tools created to inform the development of a coal-to-liquid gasification industry. Development issues were examined in a number of ways with a goal of providing practical tools to aid in decision and execution of a project.

Aspects of this report were prepared using a specific site located in Greenup County, Kentucky. This method was used to illustrate the processes required in project development to fully analyze required aspects. For example, an analysis of the potential for geologic sequestration at or near the site was conducted. While this analysis is necessarily site specific, it illustrates the information available and the process required to make an initial determination of potential success. Likewise, while the Permit Handbook has been prepared to include Kentucky permitting references, each state will have parallel agencies and requirements.

The report is organized to provide analysis and introduction to several attached tools and reports. Information may be found as follows.

- Section 5 explains the coal-to-liquid process and describes the major equipment and processes.
- Section 6 describes the resources needed or inputs to the process, and the outputs expected, including product, emissions and wastes.
- Section 7 describes site attributes that are either required or beneficial for a successful coal-to-liquids project.
- Section 8 introduces the listing of major environmental permits and authorizations included in the attachments and a chart identifying appropriate contacts within each state environmental, energy or economic development agencies for development of a coal-to-liquids facility. This section also introduces the Permit Handbook which has been developed to provide a manual to developers to guide permit preparation and scheduling. Finally, this section

examines in detail the influences and impacts that each permitting initiative may have on choice of site and design and development of the project.

- Section 9 provides a discussion of the economic impact of environmental compliance on the viability of developing and operating a coal-to-liquids facility.
- Section 10 introduces the analysis of geologic sequestration potential which is found in the attachments.
- Section 11 describes the current status of legislation and regulation applicable to geologic sequestration. Understandably, information in this section changes rapidly with developments at the U.S. Environmental Protection Agency (EPA) and under state law. This section also introduces a report of a policy Workgroup which evaluated legal issues of geologic sequestration as applied to Kentucky.
- Section 12 provides a conclusion with a summary of lessons learned and issues identified in the report development process.

3.0 EXECUTIVE SUMMARY

This project has evaluated permitting issues and plant siting for development of a coal-to-liquids plant. The research has used a development site in Greenup County, Kentucky as a model. This work has provided perspective on the information, cost and time required for preparation of applications and required information, regulatory review and approval.

Any developer preparing to design, site and construct a coal-to-liquid facility should be prepared for a long, protracted effort with regard to developing a comprehensive permit package. Permits and authorizations will be required from dozens of state and federal agencies with applications that are among the most complex possible. Public input will be extensive and likely adversarial for the most part. Opportunities for delay will be found within every regulatory program.

Legal requirements for reduction, limitation, or capture and sequestration of carbon dioxide are still in development. While no federal legislation has been passed mandating the control of such emissions, U.S. EPA has adopted requirements to inventory and report greenhouse gas emissions (GHG) and now is requiring emission reductions for the first few categories of facilities. Litigation over EPA actions may reverse or encourage those developments.

The information provided in this report and its attachments reflects this flux and the changes that have been recognized over the past couple of years. The necessary conclusion is that regulatory requirements, while to some extent still in development, are substantial, time consuming and costly to any coal-to-liquid project. Siting aspects and technology design will equally drive the permitting requirements. Delays inherent in the permitting process and allowed through public input and litigation will add substantially to the time and cost of permit acquisition.

With limited exception, an overall decision has been made to apply a date of September 30, 2010 as a cutoff date for regulation or statutory changes. This cutoff date has been necessary to allow completion of the project.

4.0 METHODS USED TO GATHER AND ANALYZE INFORMATION

The breadth of information required for this project indicated that a range of methods would be required. Methods included traditional research and literature review, interviews with industry and agency personnel and partnership with specific agencies and individuals with special knowledge of the subject matter. In addition, as a part of the analysis of the status of legislation, regulation and the existing law as it relates to coal-to-liquid industry development and to geologic sequestration, substantial efforts were used to engage with a Workgroup examining Kentucky's current status and potential approach to the state aspects of that discussion.

Resources reviewed and relied upon are listed in the attachments to this report. These resources include numerous articles, research reports and analyses. This information was particularly useful in developing the primary functions and process of a facility and as a source for the required or likely inputs and outputs of the process.

The chart identifying appropriate contacts within each state was developed by first reviewing information posted on the websites for each state's energy, environmental and economic development agencies. Preliminary information was confirmed through telephone and email contacts with each likely responsible agency. The results of this inquiry were as expected. Those states with coal production or with significant electric generation from coal have identified coal-to-liquids as an important initiative. Those states with little interface with coal generally have few resources available.

The permit listing and Permit Book were developed through painstaking analysis of existing regulatory programs. Contact was made with each agency with responsibility for the permitting or approval programs in order to confirm requirements and analyze timeframes for review and issuance. The completed Permit Handbook was reviewed by the Department for Compliance Assistance and the Department for Energy Development and Independence of the Kentucky Energy and Environment Cabinet for accuracy and completeness.

The analysis of the potential for geologic sequestration was prepared in conjunction with the Kentucky Geologic Survey, a research organization deeply involved with sequestration research and testing. This analysis relied on publicly available information from prior drilling in the area as well as traditional geologic research methods.

Finally, information about the legal issues associated with geologic sequestration and the status of legislation and regulation associated with sequestration and coal-to-liquids development was obtained through repeated review of both state and federal legislative filings and monitoring of EPA rule making and policy changes. While there is a significant federal interest in sequestration, many of the legal issues associated with the development and commercialization of that technology are property matters and therefore subject to state law. Numerous states are taking legislative action or performing studies to identify their preferred approach to the new issues presented by sequestration. As a part of the research for this project, SMG participated in a policy Workgroup which reviewed and analyzed the legal issues and through participation by a wide range of stakeholders, and developed a report identifying a recommended route. The information developed through this participation was supplemented by discussions with a similar Workgroup in West Virginia.

Individuals and organizations relied upon for analysis and assistance have been identified in the attachments.

5.0 GENERAL PROCESS DESCRIPTION

To assess the ramifications of the various regulatory programs upon the development of a coal-to-liquids (CTL) facility, it is helpful to begin with a general understanding of the CTL process. Presented below is a brief summary of the typical components of the process.

Coal can be used directly or indirectly to produce liquid fuels. Direct use (or direct liquefaction) involves the hydrogenation of coal under very high pressures to develop a liquid product. Direct liquefaction is less prominent, since the end products are highly aromatic and therefore difficult to use as high-quality synthetic fuel products.¹ Direct liquefaction has been evaluated in China¹, South Africa, Germany, and to a lesser extent in the United States.²

Indirect liquefaction requires an initial step of transforming coal to a synthesis gas (“syngas”) that is subsequently cleaned and converted to liquid fuels or several other products (e.g., diesel, gasoline, electrical power, methanol and ammonia). Due to a variety of reasons, including process flexibility, indirect liquefaction is currently the favored process for CTL facilities. One of the most commonly used methods for converting syngas to liquid fuels is the Fischer-Tropsch (FT) process. For purposes of discussion throughout this document, a conceptual CTL facility using indirect liquefaction with a FT process is considered.

The main unit processes in a CTL facility include:

¹ Radtke, K.R., M. Heinritz-Adrian and C. Marsico, New Wave of Coal-to-Liquids, an Opportunity to Decrease Dependence on Oil and Gas Imports and An Appropriate Approach to a Partial Revival of Domestic Coal Industries. VGB Powertech, Volume 86, pages 78-84, May 2006.

² Bowen, B., M. Irwin. Coal Gasification & Fischer-Tropsch, CCTR Basic Facts File #1. Indiana Center for Coal Technology Research and The Energy Center at Discovery Park, Purdue University, July 2006.

- gasification,
- gas cleaning or conditioning,
- gas conversion to liquid (such as by Fischer-Tropsch (FT) synthesis), and
- product refining.

Gasification is the process of converting carbon-based materials (e.g., coal) into gaseous products using high temperatures and elevated pressure in the presence of steam, oxygen and sometimes a catalyst. Carbon monoxide and hydrogen are the primary usable products that are the basis of synthesis gas (or “syngas”). The process enables the separation of pollutants and byproducts to generate “clean” syngas, which can then be converted to liquid transportation fuels, chemicals, or synthetic natural gas.

For the production of Synthetic Natural Gas (SNG), syngas is converted to methane in a methanation reactor and the product gas is adjusted to meet natural gas pipeline specifications.³

Regardless of the size, overall configuration, and feedstock, CTL plants generally have the same process units. These are portrayed in **Figure 1** and are briefly described below. This discussion was obtained from (CAER, 2007).³

• **Coal Preparation.** The coal is crushed and ground to a pulverized size distribution just prior to combination with water to form a slurry feed. Alternatively, pulverized coal may be dried and pneumatically conveyed to the gasifier with nitrogen or carbon dioxide.

• **Coal Slurry.** Coal is ground and combined with water to create a coal-slurry. The coal-slurry is pumped at high pressure into single-stage, slurry-feed gasifiers. The solids content of the coal-slurry is typically 65-70 percent, by weight.

³ Technologies for Producing Transportation Fuels, Chemicals, Synthetic Natural Gas and Electricity from the Gasification of Kentucky Coal. A Report in Response to HB 299, Series 3(1), (2) and (6). University of Kentucky Center for Applied Energy Research, July 2007.

- **Air Separation Unit.** Oxygen for coal gasification is provided by an air separation unit (ASU). A conventional cryogenic ASU produces 95 percent purity oxygen for coal gasification and nitrogen for inert gas uses. Higher purity oxygen (99.5 percent) is necessary to meet the SNG pipeline specifications.

- **Gasification.** Coal slurry is fed to the gasifier with oxygen from the ASU. Gasification takes place at temperatures as high as 2,600 °F and pressures of about 450 psia. The carbon conversion in a gasifier can be as high as 98 percent. In a quench-type system, the hot syngas leaves the bottom of the gasifier along with liquid ash and enters a water quench chamber. The quench removes hydrogen chloride and particulate matter before further processing of the syngas.

- **Gas Cooling, Raw Water Gas Shift, Carbonyl Sulfide Hydrolysis and Mercury Removal.** As described in more detail under Tasks 3 and 4 of this overall project, the synthesis gas stream leaving the gasifier quench section may be split, and a portion of the stream sent to a raw water gas shift reactor to adjust the hydrogen to carbon monoxide molar ratio to that required for indirect liquefaction using Fischer-Tropsch (FT) reactors. The other portion of the synthesis gas may be sent to a carbonyl sulfide hydrolysis unit where the COS is hydrolyzed to hydrogen sulfide. The two streams would then be combined and cooled in gas coolers before being sent to activated carbon filtration for removal of mercury.

- **Acid Gas Removal.** After mercury removal, the raw synthesis gas is sent to an Acid Gas Removal (AGR) unit. The AGR unit removes hydrogen sulfide and can remove carbon dioxide. The acid gas produced by this selective absorption is suitable for feeding to a Claus-type unit for acid gas treatment (AGT) and recovery of elemental sulfur.

- **Hydrogen Recovery.** A portion of the clean synthesis gas leaving the AGR unit is sent to a hydrogen recovery unit where sufficient hydrogen is separated and purified for

use in the FT upgrading section of the plant. This hydrogen is required for hydrotreating and hydrocracking.

- **Sulfur Polishing.** Depending on operating conditions, the synthesis gas exiting the AGR unit may still contain about 0.1 to 2 ppmv H₂S. This quantity of H₂S is still likely to be too high for the sulfur sensitive metal-based catalysts in the Fischer-Tropsch synthesis process. To remove this residual H₂S, zinc oxide polishing reactors are used. The zinc oxide reacts with the hydrogen sulfide to form solid zinc sulfide. The product gas leaving the polishing reactor contains less than 0.03 ppmv H₂S. (Task 2 of this overall project is a comprehensive experimental study of the effects of contaminants, such as H₂S and many others as well, in the syngas on cobalt FT catalysts.)

- **Fischer-Tropsch Synthesis.** The clean synthesis gas is heated (to a temperature that can be as high as 400°F) and fed to the FT reactors. In slurry-type reactors, the syngas bubbles up through the reactors that are filled with liquid hydrocarbons in which are suspended fine metal (e.g., iron or cobalt) catalyst particles. In smaller-scale FT plants, fixed beds of iron or cobalt FT catalyst may be used. (Task 6 of this overall project is an in-depth comparison of iron and cobalt FT catalyst systems.) Synthesis gas conversions of about 75-80 percent per pass can be obtained. Volatile overhead product swept from the reactors is separated in hot and cold separators to recover liquid hydrocarbons. The effluent from the FT reactors is cooled to recover the portion constituting liquid fuels and the unconverted synthesis gas is recycled back to the FT reactors to increase the conversion to fuels.

Heavy product that is non-volatile under reaction conditions is removed from the reactor and must be separated from the catalyst in the slurry-type system. The raw FT products consisting of crude naphtha, crude middle distillate and crude wax are sent directly to product upgrading.

- **FT Product Upgrading.** As described in more detail under Task 7 of this overall project, the raw FT products need to be upgraded to produce naphtha and high quality

jet and diesel fuels. The raw naphtha and middle distillate may be sent to a hydrotreating unit to saturate any olefins that may have been produced in the FT process. The wax material is sent to a hydrocracker where the wax is converted into hydrocarbon gases, naphtha and diesel fuel.

- **Recycle Loop.** The FT tail gas containing light hydrocarbon gases, unconverted hydrogen and carbon monoxide, some nitrogen, and carbon dioxide may be split into two streams, with one stream recycled back to the FT unit to increase liquids yield and the other stream sent to the power generation block.

- **Power Generation Block.** FT tail gas not recycled to the FT reactors can be sent to a gas combustion turbine where electric power is generated. The hot effluent gas from the gas turbine is sent to a heat recovery steam generator (HRSG). High temperature flue gas exiting the gas turbine is sent to the HRSG to recover the large quantity of thermal energy as steam for the steam turbines. The steam turbine consists of a high pressure section (~1,800 psig, 1,050 °F), an intermediate pressure section (~ 400 psig, 1,050 °F) and a low pressure section. All three sections are connected mechanically to an electric power generator by a common shaft. Power generated is used on site or can be transferred to the electric grid.

6.0 RESOURCE REQUIREMENTS AND PROCESS OUTPUTS

6.1 Process Resources and Inputs

The primary resources required for a CTL facility and associated process units described in Section 2.0 of this report are relatively few:

- Inexpensive and abundant coal resources
- Water resources for coal slurry, steam production and cooling
- Concentrated oxygen (usually generated on site)
- Electrical power (can be generated on site)

An abundant and inexpensive supply of coal is obviously the most critical input for coal gasification. CTL plants require approximately 0.5 ton of coal per barrel of liquid product.^{3,4} Commercial grade facilities producing 10,000 to 50,000 barrels per day will therefore require somewhere between 5,000 and 25,000 tons of coal per day of operation (minimum of 1 million tons per year up to 7.5 million annual tons). Coal is most economically transported by either railroad or river barge, but coal can also be delivered by truck with affordable logistics, especially in the case of a smaller-scale FT plant designed to produce only a few thousand barrels of liquid fuel products per day. Therefore, it is critical for a potential CTL facility to have dependable access to at least one of these transportation options.

The exothermal processes at CTL facilities produce significant amounts of heat that is typically removed with large amounts of water for cooling. Although air cooling can be used for some processes, typically water evaporative cooling towers are more feasible and commonly used. These cooling systems and process requirements require a consistent supply of water in the range of 1500 to 2500 gallons per minute.^{3, 4}

⁴ Baseline Technical and Economic Assessment of a Commercial Scale Fischer-Tropsch Liquids Facility. Final Report prepared for the U.S. Department of Energy, National Energy Technology Laboratory, DOE/NETL-2007/1260, April 2007.

Consequently, these types of facilities are located near dependable water resources (rivers and lakes) that can supply cooling water consistently throughout the year.

A slurry-fed gasification process requires oxygen at 95% purity that is typically produced on site using a cryogenic air separation unit (ASU). The cryogenic process uses ambient air as a feedstock. The ambient air is compressed, purified and separated into the required components and purity levels. Liquid products are sent to storage and the gaseous products are re-warmed by the feed streams and compressed to the required final product pressure(s). The exact compressor, expander and column configurations are determined by the product purities, flows and pressures required. All configurations have common characteristics that impact the performance of process control systems.⁵

The facility needs electrical power to operate equipment and can generate electricity on site. With proper design, heat generated from the process can be used to create steam for electric power generation. If electricity is generated on site, unneeded power can be exported if there is access to the regional electric transmission grid.

The FT process is a catalytic conversion that uses iron or cobalt as catalysts. Although the iron catalyst is cheaper than cobalt and less sensitive to variations in quality of the syngas feed, iron catalysts are not effective for as long as cobalt.¹ (Task 6 of this overall project is an in-depth comparison of cobalt and iron FT catalyst systems.)

6.2 Process Outputs

The primary product of coal gasification is syngas, which is cleaned and converted to liquid fuels via the FT process. In addition to these primary products there are a variety of other useable by-products as well as materials that must be treated and appropriately managed to minimize impact to the environment. Summarized below are typical by-products, waste streams, discharges and air emissions.

⁵ Vinson, D.R. Air Separation Control Technology, Computers and Chemical Engineering, Volume 30, pages 1436-1446, May 2006.

6.2.1 Solid By-Products

The high temperature in a slagging-type gasifier converts the inorganic materials left behind by gasification and fuses them into a glassy material, generally referred to as slag. Slag is generally the largest quantity solid waste stream for a CTL facility. It is a black, glassy material with the consistency of coarse sand. Slag is essentially chemically inert and has been marketed for use as an aggregate in asphalt roads, structural fill in construction applications, roofing granules, and blasting grit. Where there is no commercial market, the slag must be disposed in a solid waste landfill. In lower-temperature gasification processes, the non-combustible (ash) solid material originally in the coal will be in the form of a bottom-ash rather than fused slag.

The other large-volume byproduct produced from syngas conditioning is sulfur, which can be extracted in solid or liquid form. In either form, sale of sulfur can be made to help offset plant operating costs.

6.2.2 Gaseous By-Products and Air Emissions

When coal-derived syngas is intended to be used in a chemical process, such as in FT synthesis, high-intensity syngas clean-up is required to protect the FT catalyst and prolong its life to an acceptable interval. Contaminants, including CO₂, are removed from the syngas, but they are captured and contained within the syngas clean-up processes rather than being exhausted into the air.

Potential gaseous pollutants, such as sulfur and nitrogen compounds, form species that can be readily extracted and sold as commercial products. Hydrogen sulfide (H₂S) and carbonyl sulfide (COS), once hydrolyzed, are removed by dissolution in an organic solvent and converted to valuable commercial by-products, such as elemental sulfur or

sulfuric acid. Nitrogen is converted to NH_3 , as well as some cyanide and thiocyanate, which are readily removed via water scrubbing.⁶

The coal gasification process also has the potential to produce hydrogen by extracting it from the syngas stream. However, hydrogen as a product may be limited since a certain ratio of hydrogen to carbon monoxide must be maintained for converting synthetic gas to liquid fuels.

When coal-derived syngas is intended simply to be combusted, to generate electric power for example, a less intense, and less effective, syngas clean-up process will likely be employed. High-intensity syngas clean-up requires cooling the syngas to a low temperature. While this is necessary and acceptable in FT synthesis, cooling the syngas during clean-up before combustion would directly reduce its ability to generate electricity most efficiently.

An FT plant could use a portion of its coal-derived syngas for combustion and electric power generation. The portion of the syngas to be combusted would likely not receive the same high-intensity clean-up as the syngas fed to the FT reactors. A general discussion of the level of known air pollutants that can be generated from the gasification process, when followed by a relatively low intensity clean-up of the syngas before its combustion for power generation, is presented below. This discussion is largely derived from a publication entitled *Major Environmental Aspects of Gasification-Based Power Generation Technologies*.⁷ The following discussion is effectively a comparison between coal gasification and conventional coal combustion as approaches for generating electric power, rather than a focused discussion of the (very low) atmospheric emissions from a plant producing FT liquid fuels from coal.

⁶ Ratafia-Brown, J., L. Manfredo, J. Hoffmann, M. Ramezan and G. Stiegel. An Environmental Assessment of IGCC Power Systems. Presented at the Nineteenth Annual Pittsburgh Coal Conference, September 2002.

⁷ Major Environmental Aspects of Gasification-Based Power Generation Technologies. Final Report prepared for the U.S. Department of Energy, National Energy Technology Laboratory, December 2002.

6.2.2.1 SO₂ Emissions

During coal gasification, most of the sulfur in the coal matrix is released and converted to hydrogen sulfide (H₂S), as well as a small amount of carbonyl sulfide (COS), due to the reduced oxygen environment. These H₂S, COS and particulate contaminants are mostly removed from the syngas prior to combustion or other forms of fuel conversion. Acid gas removal equipment extracts 95-99% of the H₂S and COS from the fuel gas and converts it to a salable sulfur or sulfuric acid (H₂S O₄) byproduct. The small amount of residual sulfur that remains in the syngas is converted to SO₂ in the combustion turbine and released to the atmosphere in the primary stack gas or in the secondary stack gas from the sulfur recovery equipment.⁷

For comparison, the Polk and Wabash River IGCC plants achieve emissions below 0.15 lb. SO₂/MMBtu heat input or greater than 97% sulfur reduction. This is almost an order of magnitude lower than Federal limits on SO₂ emissions from utility plants burning solid fuels and also less than the Federal SO₂ emission limits required for stationary gas turbines.⁶

6.2.2.2 Particulate Emissions

Particulate control in gasification processes is highly efficient, as gasifiers operate at high pressure and generate a significantly smaller gas volume than conventional coal combustion. Not only does the gasification process provide an inherent capability to remove most ash as slag or bottom ash, but the fly ash produced is concentrated in the smaller gas volume, which further assists its cost effective collection.

Both the Polk and Wabash River plants use a wet scrubber to efficiently capture fine particulates that are entrained in the syngas. Additional particulate removal occurs in the gas cooling operations and in the acid gas removal systems. As a result, very low particulate emission levels are achieved. The Wabash plant reported emissions of less than 0.012 lb./MMBtu heat input (0.088 lb./MWh output), while the Polk plant typically

emits less than 0.015 lb./MMBtu.⁷ These emissions are significantly less than the current Federal NSPS requirement of 0.03 lb./MMBtu heat input.⁶

6.2.2.3 NO_x Emissions

The term “NO_x” refers to the sum of the nitric oxide (NO) and nitrogen dioxide (NO₂) emissions from a combustion source. While most of the NO_x produced during the combustion of syngas is in the form of NO, it is subsequently oxidized to NO₂ in the atmosphere. NO_x is formed in fossil fuel combustion systems by two primary mechanisms; “fuel NO” is formed via the oxidation of chemically-bound nitrogen in the fuel, and “thermal NO” is formed via the dissociation of molecular nitrogen and oxygen to their atomic forms (at high temperatures) and subsequent recombination into oxides of nitrogen. Unlike natural gas, coal contains chemically-bound nitrogen that forms most of the NO_x emissions when it is fired in a typical excess-oxygen environment, such as a utility boiler.

The gasification process differs significantly from conventional coal combustion plants with respect to the impact of chemically-bound nitrogen in solid fuels, like coal. Because it operates with a deficiency of oxygen, gasification converts most of the fuel nitrogen into harmless nitrogen gas (N₂). While a small portion is converted to ammonia (NH₃), as well as small amounts of hydrogen cyanide (HCN) and thiocyanate, these water-soluble species are removed during fuel gas cooling and cleaning and are usually converted to nitrogen in the sulfur recovery process. The fuel gas produced is virtually free of fuel-bound nitrogen.⁷

NO_x formation is primarily the result of thermal NO produced at the high temperatures in the combustion turbine. By maintaining a low fuel-air ratio (lean combustion) and adding a diluent (e.g., nitrogen from the air separation unit or steam), the flame temperature can be lowered to reduce the potential for NO_x formation. IGCC NO_x emissions of less than 0.1 lb./MMBtu are quite low relative to the emissions of a PC plant with low-NO_x burners, approximately 0.4 lb./MM Btu for a tangentially- fired boiler.⁶

6.2.2.4 CO Emissions

CO emissions are typically the result of incomplete combustion, but can also result from fugitive emissions. In an IGCC system, sources are typically the gas turbine, sulfur recovery unit tail gas incinerator, the flare system and equipment leaks. Total CO emissions from the Wabash IGCC plant for 1998 were 0.30 lb./MWh. The original Wabash coal-fired plant emitted CO at an annual average rate of 0.64 lb./MWh.⁷

6.2.2.5 Lead Emissions

Lead, a semi-volatile metal, is released from coal during combustion or gasification and partially volatilizes and becomes enriched on fly ash particles of decreasing particle size. Both bench-scale testing and thermodynamic equilibrium models indicate that the most likely chemical forms of lead in gasifier product gas will be Pb, PbS, PbCl₂ and PbCl. Key variables that influence the formation of these lead species are the lead species present in the coal, coal pretreatment, gasifier temperature profile, oxygen partial pressure and reaction time. Most, but not all, of the lead species should be removed in the plant's particulate and acid gas cleanup systems.⁷

Any residual lead in the syngas will be discharged from the combustion turbine as Pb, PbCl₂, or PbO. Trace metal mass balance results for LGTI's IGCC plant showed about one-third of the lead in the coal ended up in the gasifier slag and less than 5% as air emissions. The remaining lead was assumed to be removed in the particulate and acid gas cleanup systems and discharged with solid and liquid waste streams. The form of lead discharged in the solid waste or through the plant's waste water treatment facility is unknown at this time. Turbine stack emissions showed an average lead content of 1.6 $\mu\text{g}/\text{Nm}^3$, with 62% in the particulate phase and 38% in the vapor phase. A total average air emission factor for lead at the LGTI plant was calculated to be 2.9 lb./10¹² Btu of heat input. Thus, trace amounts of lead contained in coal can be efficiently removed in an IGCC plant with minimal discharge to the atmosphere.⁷

6.2.2.6 Hazardous Air Pollutants

Potential trace substance emissions from coal- fueled power plants include ionic species, trace elements, and trace organic compounds. These trace substances can be emitted in flue gas, aqueous discharges, and solid effluents. Ionic species of environmental concern in the effluent streams of coal-fueled power plants include sulfate, nitrogen-containing ions (e.g., nitrate, ammonium), chloride, fluoride, phosphate and cyanide. The ionic forms of these species in stack gases are present only in the aerosol phase. Chloride and fluoride, however, can exist as acids and, thus, may appear in the gas phase as well. Stack emissions of all ionic species are reduced to very low levels via particulate and acid gas control equipment.⁶

Release of trace organic compounds is also an environmental concern, since some of these compounds, such as dioxins, furans, and formaldehyde, can adversely affect the environment or human health. While there isn't much corroborating data available on trace organic releases from gasification systems, detailed test results from the LGTI plant indicate extremely low levels of trace organic emissions, in line with emissions expected from conventional coal-fired plants. Furthermore, results from both LGTI and a Shell coal gasification pilot plant corroborate that dioxins and furans are not present at the detection limit of 1 part per billion by volume in the synthesis gas, nor were there any precursors at the same detection level. Due to the effects of dilution and combustion, the concentration of dioxins and furans in the HRSG stack gas should be less than one part per trillion by volume. Additionally, formaldehyde emissions from a syngas-fired combustion turbine appear to be more than an order-of-magnitude lower than from a natural gas-fired combustion turbine.⁶

Coal contains most of the naturally occurring chemical elements, including potentially toxic trace metals. Specific elements and their concentrations are dependent upon the type of coal and its geological origins. These trace species may be released during gasification. Most either remain with the slag/bottom ash or are removed from the

syngas in downstream process equipment. The trace metals of greatest environmental concern are reported to be arsenic, boron, cadmium, mercury, and selenium. While in-situ measurement of these species has proven to be quite difficult in the reducing atmosphere of gasification system, computer-based thermodynamic equilibrium studies have indicated that these metals are highly volatile and may be hard to control.⁶ Of these, mercury has received the most attention from regulators.

6.2.2.7 Mercury

“Mercury is a particular problem in gasification systems, since it primarily remains in the vapor phase due to its low boiling point (357°C or 180°F). Its partitioning and speciation may vary between different gasification systems, but should be broadly similar. The likely chemical forms are elemental mercury (Hg), oxidized mercury (HgCl₂), and mercuric sulfide (HgS)”⁶. These mercury species may remain in the gaseous phase, be adsorbed onto particulates, or be removed in the liquid scrubbers. Mercury can be removed from the syngas via use of activated carbon beds.⁶

6.2.2.8 Greenhouse Gas Emissions

The largest contributor to greenhouse gas (GHG) emissions is the production of CO₂ from the carbon contained in the fuel fed to the gasifier. The production of other GHG emissions, such as N₂O and NH₃, are small compared with CO₂. Although CO₂ emissions are higher than natural gas-fired plants, gasification’s improved efficiency reduces CO₂ emissions relative to conventional coal-fired plants. On average, IGCC plants produce CO₂ at a rate of about 1.85 lb./kWh, while pulverized coal-fired plants yield about 2 lb./kWh.

Gasification has two major advantages that can be exploited to capture CO₂ more efficiently. The syngas has a high CO₂ concentration, which can be further increased by converting CO to CO₂ prior to combustion (while simultaneously producing more hydrogen), and IGCC gasifiers typically operate under relatively high pressure (~400

psig in the Wabash plant). This makes recovery of the CO₂ from the syngas much easier than capture from flue gas. A recent study of one design concept concluded that 75% of the CO₂ could be captured from an IGCC plant with only a 4 percent loss in efficiency, but this does not account for transport of the CO₂ to a utilization or sequestration site and further processing.⁶

6.2.3 Liquid By-Products and Wastewater Discharges

Some beneficial byproducts of FT synthesis include: liquid sulfur extracted from the syngas; paraffinic and olefinic waxes that can be used for various products including high quality lubricants; ammonia; alcohols; ketones and aldehydes in modest quantities.

Coal gasification plants have two principal sources of water effluents. The first is wastewater from the steam cycle that includes blowdown from the boiler feed water purification system and the cooling tower. The amount of wastewater effluent depends upon the hardness and other qualities of the raw water and process quality needs. These effluents typically contain salts and minerals that have been concentrated from the raw feed water.

The second aqueous effluent is process water blowdown. Gasification processes typically purify and recycle most of the raw process streams, with some blowdown to remove concentrated salts and minerals. This blowdown stream is typically high in dissolved solids and gases along with the various ionic species washed from the syngas, such as sulfide, chloride, ammonium, and cyanide.⁶

The blowdown streams are recycled through a variety of ancillary processes including the coal feed preparation area, the scrubber after entrained solids have been removed, a zero discharge water treatment/recirculating system; or an on-site wastewater treatment system.⁶

An additional wastewater stream is stormwater runoff in contact with plant operations, the coal stockpile area and slag handling and disposal areas.