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Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup, and Oxygen Separation Equipment

Task 2: Gas Cleanup Design and Cost Estimates – Wood Feedstock

Nexant Inc.
San Francisco, California

Subcontract Report
NREL/SR-510-39945
May 2006

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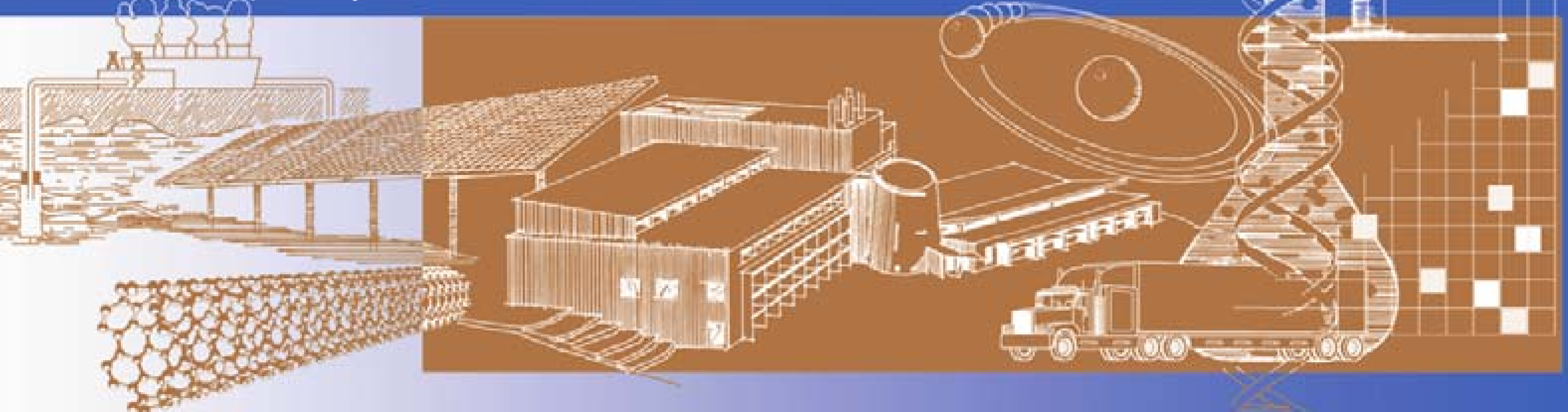

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Executive Summary

As part of Task 2, Gas Cleanup and Cost Estimates, the team investigated the appropriate process scheme for treatment of wood derived syngas for use in the synthesis of liquid fuels. Two different 2,000 metric tonne per day gasification schemes, a low-pressure, indirect system using the BCL gasifier, and a high-pressure, direct system using GTI gasification technology, were evaluated. Initial syngas conditions from each of the gasifiers was provided to the team by NREL. Nexant was the prime contractor and principal investigator during this task; technical assistance was provided by both GTI and Emery Energy.

The first task explored the different process options available for the removal of the main process impurities, including particulates, sulfur, carbon dioxide, tar, ammonia, and metals. From this list, selection of commercial technologies appropriate for syngas clean-up was made based on the criteria of cost and the ability to meet the final specifications. Preliminary flow schemes were established and presented to NREL; after discussion and modification, final designs, including unit sizes, energy use, capital and operating costs, and labor requirements, were developed. Finally, Nexant performed an analysis to determine how changes in syngas flowrates and compositions would impact the designs, for future reference as the plant size changes.

The technologies chosen for both cases did not differ considerably. Each case possesses the following pieces of equipment:

- Cyclones for particulate removal
- Tar cracking for the removal of heavy and light hydrocarbons. Steam is injected in varying amounts into the tar cracker to set the appropriate hydrogen to carbon monoxide ratio.
- Syngas cooling, necessary for downstream sulfur treatment, and a water quench/venturi scrubber for ammonia and trace contaminant removal
- Amine treatment for sulfur and carbon dioxide removal
- Zinc oxide beds for additional sulfur removal down to the low levels required for fuels synthesis
- Liquid phase oxidation of acid gas for sulfur recovery

The low-pressure gasifier case required the use of a process gas compressor to raise the gas pressure to the level appropriate for downstream treatment and product synthesis. Information was also provided for the level of clean syngas compression necessary to prepare both cases for methanol synthesis.

The results of the analysis for both cases can be seen in Table A below, with information on the capital and operating costs:

TABLE A SYNGAS CLEAN-UP CASE SUMMARY

	Low-Pressure BCL Gasifier	High-Pressure GTI Gasifier
Wood Feedrate (MTPD)	2,000	2,000
Syngas Rate (lb/hr)	316,369	418,416
Total Installed Cost ($MM)	109.4	76.5
Power Required (MW)	18.5	(5.2)
Net Steam Required (lb/hr)	44,000	114,000
Water Required (GPM)	37,806	25,454
Natural Gas (MMSCFD)	7	8
Catalysts and Chemicals ($/day)	1,931	1,457

The bulk of the cost difference between the two cases is due to the process gas compressor required in the low-pressure case. The two cases use similar equipment for all other steps of the process; although the cases had different gas flowrates and compositions, the equipment impact is small relative to that of the process gas compressor. While these results imply that direct gasification is preferred, this study did not take into account other differences in the two process schemes, such as the potential need for an oxygen plant in the high-pressure to chemicals case.

The team also compared the clean-up system design and costs versus the design developed by NREL for a recent biomass to hydrogen study. The cost for the clean-up section of the biomass to chemicals designs is more expensive due to three main reasons: more equipment necessary in the chemical production designs, the increase in steel prices from 2002 to 2005, and different engineering assumptions made in the chemicals production case. The main engineering difference is the cost assumed for the process gas compressor in the low pressure case; a larger compressor and selection of a different design type increases the installed cost by $25MM versus the NREL design. In addition, gas clean-up cost assumptions made by NREL from previous studies likely underestimated the cost of the tar cracker and heat exchange equipment.

This study updates previous NREL investigations by providing the most up-to-date information for appropriate technologies and their respective costs. Future studies should focus on the following areas to further define suitable technologies and confirm costs:

- ***Alternatives for Tar Removal:*** Further study and analysis should be performed to validate the methods used by the team. In addition, alternative tar removal technology should be considered, including cracking within the gasifier.
- ***Process Integration, Gasification Systems and Biorefinery:*** Integration of the clean-up section with the other parts of the gasification plant will provide a better picture of the overall plant costs.
- ***Alternate CO_2/Sulfur Removal Steps:*** A cost comparison of amine versus physical solvents would provide additional data to confirm the appropriate use of amine in this design Advanced technologies for acid gas removal, such as warm gas clean-up, should also be considered.

- ***Other Impurities in the Syngas:*** If it is deemed that the level of items such as metals and halides entering the scrubber will not adversely impact the FT or methanol catalysts, this step could be removed.

Introduction and Methodology

This study provides designs and costs for cleaning wood derived syngas in preparation for feed to liquid fuel synthesis units. Two different starting conditions, one with syngas derived from a low-pressure, indirect gasifier, and one from a high-pressure, direct gasifier, were evaluated. The goal was to provide NREL with a complete design package, including process flow diagrams, equipment specification sheets, mass and energy balances, capital and operating costs, and labor requirements, that can be used to evaluate the feasibility of biomass to chemicals technologies. The study also addressed how the designs would be impacted by changing flowrates and syngas compositions, so that the designs could be adapted to other process conditions.

The work was divided into three main task areas. The first Subtask (2.1) presented a list of possible gas clean-up technologies, with recommendations provided for the most suitable ones for additional analysis. The results of this study can be seen in Appendix D. Next, preliminary process flow diagrams were developed, along with an initial material balance (Subtasks 2.2.1 and 2.2.2). This was reviewed with NREL, and modifications made before the final design work began. The final phase consisted of performing equipment sizing, development of costs, and scaling analysis (Subtasks 2.2.3 through 2.2.7).

A variety of resources were used throughout the project to produce the final designs. In gathering the initial technology data, previous team studies, literature reviews, vendor information, and NREL input were all used to establish the items for consideration. Vendors and R&D facilities were especially helpful in providing data for novel technologies, such as tar cracking and liquid phase sulfur oxidation. Team members involved in biomass gasification, GTI and Emery Energy, provided valuable insight on reliability and feasibility issues.

HYSYS was used for modeling the overall process, with vendor input for specialty equipment. Design and performance of the amine system, LO-CAT™ unit, tar cracker, and process gas compressor were provided by vendors and estimated through other modeling work. All other process equipment was sized by the HYSYS program. Since the basis for the tar cracker, the NREL TCPDU, is not commercial, data from NREL was used, along with assumptions for bed fluidization needs and heat transfer requirements to produce a size estimate. Greater detail for the assumptions made can be found in Section 2.

Costing was performed in a similar fashion as design, with commercially available software, ICARUS, used for much of the equipment sized using HYSYS. All cost estimates use a second quarter 2005 basis. Quotes were obtained from vendors for unique and capitally intensive items, such as the process gas compressor, cyclones, ZnO beds, and LO-CAT™ unit. Industry derived cost curves were used for the amine system and as a check on other process items. Operating costs were developed from vendor supplied information and the energy balance. Finally, labor requirements are derived from a scale-up of a detailed study by Emery Energy specific to biomass gasification. For all results, comparisons were made throughout the study to results from previously developed NREL reports.

1.1 INTRODUCTION

The initial task for the Nexant team was to identify and evaluate all commercially available technology for clean-up of wood derived syngas. The technology list, with information on operating size ranges and conditions, materials of construction, and cleanup parameters, can be seen in Appendix D. After a review of technology options with NREL, flow schemes were developed for both the high and low pressure cases. The result of this analysis and justification for the technologies chosen is detailed in this section.

The compositions of the syngas from the gasifiers and the cleanup requirements are listed in Tables 1-1 and 1-2 below[1]. Each case being evaluated assumed a wood feedrate of 2,000 metric tonnes per day (MTPD).

TABLE 1-1 SYNGAS COMPOSITIONS AND OPERATING PARAMETERS

	Syngas from BCL Gasifier	Syngas from GTI Gasifier
Temperature, °F	1,598°F (870°C)	1,598°F (870°C)
Pressure	33 psia (1.6 bar)	460 psia (32 bar)
Steam/bone dry feed	0.4 lb/lb	0.76 kg/kg
Compositions	Mol% (wet)	Mol% (wet)
H_2	12.91	13.10
CO_2	6.93	19.40
CO	22.84	8.10
H_2O	45.87	50.70
CH_4	8.32	7.80
C_2H_2	0.22	---
C_2H_4	2.35	0.10
C_2H_6	0.16	0.20
C_6H_6	0.07	0.30
Tar ($C_{10}H_8$)	0.13	0.10
NH_3	0.18	0.10
H_2S	0.04	0.04
Gas Yield	0.04 lbmol of dry gas/lb bone dry feed	0.05 lbmol of dry gas/lb bone dry feed
Char Yield	0.22 lb/lb bone dry feed	0.0514 lb/lb bone dry feed
H_2:CO molar ratio	0.57	1.62

1 Information provided by Pamela Spath, NREL.

The gas pressure assumed from the BCL gasifier, 33 psia, is higher than initially evaluated during this project. Preliminary investigations were performed using a syngas pressure of 23 psia. Raising the pressure by 10 psia allows for a simpler and more reliable design, by allowing a water wash upstream of the compression stage.

TABLE 1-2 GAS CLEANUP REQUIREMENTS

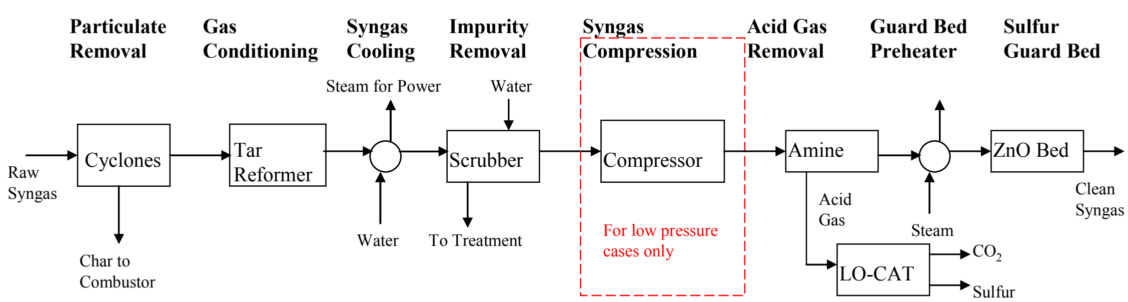
Process	Contaminants	Level	Source/Comment
Fischer-Tropsch Synthesis	Sulfur	0.2 ppm 1 ppmv 60 ppb	Dry, 1981 Boerrigter, et al, 2002 Turk, et al, 2001
	Halides	10 ppb	Boerrigter, et al, 2002
	Nitrogen	10 ppmv NH3 0.2 ppmv NOx 10 ppb HCN	Turk, et al, 2001
Methanol Synthesis	Sulfur (not COS)	<0.5 ppmv (<0.1 ppmv preferred)	Kung, 1992
	Halides	0.001 ppmv	Twigg and Spencer 2001
	Fe and Ni	0.005 ppmv	Kung, 1992

The main impurities in the syngas exiting the gasifier that must be removed are char, tars, hydrocarbons, sulfur, and CO_2. In addition, trace contaminants such as ammonia, metals, halides, and alkali species were of sufficient concern that equipment was added to remove them as well. Finally, the syngas must also be adjusted to obtain the appropriate H_2/CO ratio.

1.2 PROCESS DESCRIPTION AND RATIONALE

A schematic for the process design developed for both cases can be seen in Figure 1-1. Both the low and high pressure cases used very similar processes for syngas clean-up: particulate removal with cyclones, tar reforming, cooling and water scrubbing, acid gas removal with amine, and sulfur polishing. The main difference between the cases is the inclusion of a compression step in the low-pressure case. A detailed description of each design is addressed in this section.

FIGURE 1-1 GENERAL SYNGAS CLEAN-UP PROCESS FLOW



1.2.1 Low-Pressure Syngas Process Description

Particulate Removal

The syngas exiting the gasifier contains impurities that must be removed in order to meet the specifications required for methanol or FT synthesis. Cyclones are used as the initial step in the gas cleanup process to remove the bulk of the char entrained in the syngas stream. This technology is standard in industry due to its low cost and high level of performance for removing particulates. Syngas from the low-pressure gasifier is sent through four parallel cyclones operating at 1598°F and 33 psia.

Tar Reforming

Syngas is fed to a tar reformer to remove tars, light hydrocarbons, and ammonia before any additional gas treating or cooling. Reforming must occur prior to cooling the syngas to prevent tar condensation and deposition on downstream equipment. The tar reformer was modeled using NREL's "goal design" reactor conversion for the Thermochemical Pilot Development Unit (TCPDU). Table 1-3 shows the assumed reactor conversion rate as provided by NREL. In the tar reformer, tars (mono and polyaromatic compounds) and light hydrocarbons such as methane, ethylene, and ethane are converted to H_2 and CO. Ammonia is converted to N_2 and H_2. Since the reactor effluent contains about 1.3 mol% CH_4, and 0.2 mol% of other hydrocarbons, additional downstream steam reforming was deemed not necessary. This conclusion was confirmed by NREL[2].

TABLE 1-3 TAR REFORMER PERFORMANCE

Compound	% Conversion
Methane (CH_4)	80
Ethane (C_2H_6)	99
Ethylene (C_2H_4)	90
Tars (C10+)	99.9
Benzene (C_6H_6)	99
Ammonia (NH_3)	90

Syngas exiting the tar reformer enters another cyclone to separate both entrained reforming catalyst and any residual char. The solids are then sent to a catalyst regenerator. The catalyst is sent to a regenerator vessel, where char and residual carbon is combusted. The hot, regenerated catalyst is then recycled back to the reactor vessel, acting as the heat source for the reforming reactions.

Syngas Cooling

The remaining gas treatment steps require the syngas to be at a much lower temperature. Therefore, the gas is cooled in three stages from 1598°F to 225°F prior to scrubbing. The heat

[2] Nexant team discussion with Pamela Spath, April 2005.

recovered from the process is used for steam generation throughout the system. The process design has been optimized as much as possible to use this steam, reducing the plant utility load. Integration was limited to the needs of the clean-up section; broader heat integration with the overall thermochemical platform or biomass refinery may lead to additional efficiency gains.

Scrubbing and Quench

The syngas is sent to the Syngas Venturi Scrubber, C-200, to remove any remaining ammonia, particulates, metals, halides, or alkali remaining in the system. The water circulation rate to the scrubber is adjusted such that the exiting syngas is quenched to the appropriate temperature for feed to the first stage of the compressor.

Compression

Any residual condensate in the syngas exiting the scrubber is removed in the Syngas Compressor KO Drum, V-300. The cooled syngas stream is compressed to 445 psia using a 4-stage centrifugal compressor with interstage cooling. The compressor is modeled assuming a horizontally split centrifugal design, with a polytropic efficiency of 78% and 110°F intercoolers. After discussion with compressor vendors[3] and internal analysis by Nexant, it was determined that this type of compressor is appropriate for this gas flowrate, pressure ratio, and reliability requirements. While an integrally geared compressor was considered due to its lower cost, this type of compressor was not recommended due to the high flowrate and reliability required. The discharge pressure is designed such that the compressed gas is at the operating pressure range for FT synthesis.

Sulfur Removal

Originally, the scheme developed was use of LO-CAT™ and ZnO polishing for H_2S removal, followed by amine for CO_2 removal. After discussions with NREL, this was modified so that amine was used for both H_2S and CO_2 removal. The ZnO beds remained in the design as a guard/polishing step after the amine unit, while the LO-CAT™ unit is now used to remove H_2S from the acid gas stream. The benefit of this design is reduced load on both the LO-CAT™ and ZnO units; the flow going to the LO-CAT™ unit in this case is now only the acid gas stream instead of the entire syngas stream, and the inlet H_2S concentration at the ZnO bed is expected to be lower. This should increase the lifespan of the ZnO catalyst.

The syngas exiting the gasifier contains ~400 ppmv of H_2S. An amine unit with a high circulation rate can reduce the syngas sulfur concentration to below 10 ppmv, with a target of 2-3 ppmv. Due to the high amount of CO_2 removal required, it is this component that drives the circulation rate and unit size, not H_2S. The ZnO beds are used as a polishing step to reduce the sulfur concentration to the < 0.1 ppmv level required for methanol and FT synthesis. The gas exiting the amine absorber is heated to the operating temperature of the ZnO beds, 750°F.

For the low-pressure case, DEA was selected, while MDEA is used for the high-pressure case. This selection is based on design simulation runs by matching the desired CO_2 and H_2S removal

3 Consultation made with both Elliott Compressor and GE.

requirements to the selectivity of the amine solvents. Attempts were also made to choose solvents that minimized net energy requirements.

Water-Gas Shift and CO_2 Removal

FT synthesis requires a H_2/CO ratio of 2:1, and methanol synthesis requires the following stoichiometric ratio of H_2, CO, and CO_2:

$$(H_2 - CO_2) / (CO + CO_2) = 2$$

The syngas stream exiting the ZnO beds has a H_2/CO ratio of 1.7 and a stoichiometric ratio of 0.89, which are inadequate for FT or methanol synthesis. A combination of water injection into the tar cracker, followed by CO_2 removal in the amine unit, has been selected to adjust these ratios. In methanol synthesis, H_2 will react preferentially with CO_2 over CO to form methanol. This results in a significantly lowered methanol yield, greatly impacting the process efficiency. In FT synthesis, CO_2 acts as a diluent; however, for a design in which the off-gas from the FT reactor is recycled back to the reactor to improve conversion, removal of CO_2 is necessary to prevent CO_2 buildup in the reactor.

The initial designs for the low pressure system incorporated a shift reactor instead of water injection to assist in obtaining the necessary composition ratios. Further analysis and review with NREL led to the determination that a shift reactor was unnecessary, and that steam injection into the tar cracker is sufficient to perform the required shift. Elimination of this unit operation helps to reduce the overall system cost.

CO_2 removal can be achieved through different processes such as chemical (amine) or physical (Selexol or Rectisol) absorption, as outlined in Appendix D. The syngas stream entering the CO_2 removal unit is at about 420 psia and 110°F. Since physical absorption process is best suited for high pressure (>700 psia) and low temperature systems, an amine system was selected to remove CO_2 from the syngas. In addition to the syngas already possessing the appropriate operating conditions for chemical absorption, an amine system is also likely to be less expensive than the Selexol or Rectisol system. A side-by-side cost analysis from vendors would be necessary to confirm the optimal design. Approximately 98% of the CO_2 in the syngas stream must be removed in order to meet the stoichiometric ratio requirement for methanol synthesis.

The treated syngas exits the amine absorber at approximately 110°F and 440 psia. The treated syngas is sent to either the methanol or FT reactor. For methanol synthesis, the treated gas is compressed and heated to the operating conditions of the methanol reactor, about 1160 psia and 460°F. For FT synthesis, the treated gas is heated to 350°F.

1.2.2 High-Pressure Syngas Process Description

The cleanup process scheme for the syngas from the high-pressure gasifier is similar to that of the syngas from the low-pressure gasifier with the exception of the syngas compression step, differences in the heat balances, and process unit size variations due to different syngas compositions and conditions. Information about these differences is presented below.

Similar to the low-pressure case, high-pressure syngas is sent through a series of cyclones to remove the bulk of the char entrained in the syngas stream. The syngas is then sent to the tar reformer for removal of tars, methane, other light hydrocarbons, and ammonia. Steam is added to the syngas entering the tar reformer so that the shift reaction that occurs in the reformer can yield the required H_2/CO ratio for methanol or FT synthesis. Due to a more appropriate synthesis ratio in the raw syngas stream, less steam is required relative to the low-pressure case. The reformer effluent is then sent to the water scrubbing unit for removal of residual char, alkali, metals, halides, and ammonia.

Following the water scrubbing unit, the syngas is sent to an amine unit where MDEA is used for the removal of both H_2S and CO_2. As in the low-pressure case, a LO-CAT^{TM} unit is used for sulfur recovery, while ZnO beds are used for reducing the syngas sulfur content to below < 0.1 ppmv H_2S. Rationale for process selection of the sulfur and CO_2 removal units is similar to that of the low-pressure syngas case, although MDEA was used instead of DEA in the amine system. The treated syngas is sent to either the methanol or FT reactor. For methanol synthesis, the treated gas requires compression and pre-heating to 1160 psia and 460°F prior to entering the methanol reactor. For FT synthesis, the treated gas requires pre-heating to 350°F.

1.3 DISCUSSION

1.3.1 Technologies Not Chosen

As presented in Appendix D, a list of technologies was provided for performing the various gas cleanup tasks required. From this list, specific technologies have been selected for each of the designs presented here. Below is a list of the technologies that were not chosen, and the rationale behind those decisions.

Particulate Removal

Ceramic and Metal Candle Filters: Candle filters could be used in place of cyclones for char and catalyst separation from the syngas stream. Little commercial experience exists in operating these types of filters at the temperatures (1500°F+) that the cyclones operate under. At this temperature, only ceramic filters could be considered. A recent study performed by Nexant for the DOE's National Energy Technology Laboratory[4] examined replacing a third stage cyclone with a ceramic candle filter. The cost of this high temperature filter, even assuming an "nth plant design", did not justify the change. Because of the limited commercial experience and high cost, these options were eliminated.

Baghouse Filters: As with candle filters, baghouse filters are not appropriate for high temperature applications. Therefore, they cannot replace the cyclones as an effective solids removal option.

Electrostatic Precipitators: Since dry ESPs can only operate up to ~750°F and wet ESPs up to ~200°F, this option cannot replace cyclones for solids removal. In addition, the high cost and waste streams produced make them unattractive relative to other filtration options.

4 "Gasification Alternatives for Industrial Applications: Subtask 3.3—Alternate Design for the Eastern Coal Case, DOE Contract DE-AC26-99FT40342, April 2005.

Tar and Hydrocarbon Removal

Wet Scrubbing: Due to the relatively low content of tar in the syngas stream and the non-power application being considered, wet scrubbing could be considered a viable option for tar removal. However, inclusion of a wet scrubber may make a steam reformer necessary to remove hydrocarbons from the system. In addition, wet scrubbing for tar removal creates considerable waste removal and treatment issues and lowers process efficiencies. A detailed analysis comparing the current configuration with a wet scrubber/steam reformer would be of interest to confirm these assumptions.

Hydrocarbon Reforming (SMR/POx/ATR): Due to the low content of hydrocarbons exiting the tar cracker, it was determined that this step was unnecessary. Both FT and methanol synthesis reactors should be able to handle the quantity of hydrocarbons without severely impacting performance.

Other Technologies: During the course of the design work for the current configuration, other alternatives, such as injection of cracking catalyst directly into the gasifier and changes in gasifier operation, were identified. Limited empirical data for these technology options make them impractical for design use at this time.

Sulfur Removal

LO-CAT™: The initial designs for sulfur removal from the syngas stream used the LO-CAT™ technology due to the low net syngas sulfur content. Redesigns of the combined sulfur and CO_2 removal system demonstrated that using LO-CAT™ for sulfur recovery and amine for sulfur and CO_2 removal was more economic.

Physical Solvents: As can be seen in Appendix D, physical solvents (Rectisol/Selexol processes, for example) typically operate at low temperatures and high pressures. Changes in the stream pressure leaving the scrubber/quench may be required prior to entering a physical solvent unit for optimum performance, whereas the current process conditions are more appropriate for feed to an amine system. In addition, previous Nexant studies have determined little to no cost benefit in implementing a physical solvent system over other treatment methods for systems of this nature. A more in-depth analysis would be required to confirm the cost difference between physical absorbents and an amine/ZnO treatment system.

COS Hydrolysis: Due to the limited COS expected to be produced from a biomass gasification system, this removal step was omitted.

2.1 INTRODUCTION AND METHODOLOGY

Design and cost estimates were obtained using three major sources:

- HYSYS and ICARUS were used to obtain design and cost estimates for generic equipment such as vessels, pumps, compressors, and heat exchangers. The design basis was agreed upon after the submission of the design information outlined in Section 1.
- Vendor quotes were obtained for unique and specialized equipment such as cyclones, ZnO catalyst/reactors, LO-CAT™ sulfur absorption, and compressors. Some items, such as compressors and blowers, were estimated both by HYSYS/ICARUS and through vendor quotes in order to validate the results.
- The amine unit performance and energy requirements were estimated using commercially available software that is specific for amine unit modeling. Once performance requirements were obtained, an industry developed cost curve was used for estimating installed cost.

An updated set of PFDs can be seen in Appendices A and B. The design and cost estimates for the high-pressure and low-pressure cases are presented in the Equipment List and Data Sheets, which can be seen in Appendix C. The Equipment List groups process equipment by the following categories: reactors, cyclones, vessels, heat exchangers, compressors, pumps, turbines, and packaged units (the amine and LO-CAT™ units). Shown in the Equipment List are the following items:

- Unit size and weight
- Design duty (exchangers)
- Design temperature and pressure
- Power usage
- Materials of construction
- Price (uninstalled) on both a Q2 2004 and Q2 2005 basis
- Source for cost estimate
- Comments and notes

An installation factor of 2.57 was applied to all base equipment costs, with the exception of the process gas compressor, to arrive at the total installed cost. The installation factor was derived based upon previous experience and vendor estimates. An installation factor of 2.47 was used for the compressor based on previous detailed compressor cost analysis. The total installed cost for the low-pressure case is $109MM, while the installed cost for the high-pressure case is $76MM. The difference is largely due to the process gas compressor used in the low-pressure case.

2.2 KEY DESIGN ASSUMPTIONS

A complete description of the process and rationale for choosing the technologies in this deliverable can be seen in Section 1. Each case assumed a feedrate of 2,000 MTPD. Issues encountered when performing the unit designs are outlined below.

2.2.1 Sulfur and CO_2 Removal

As mentioned in Section 1, DEA was selected for the low-pressure case, while MDEA is used for the high-pressure case. This selection is based on design simulation runs by matching the desired CO_2 and H_2S removal requirements to the selectivity of the amine solvents. The level of CO_2 removal is the major driving force in determining the amine system size and cost; without the need for CO_2 removal, the unit cost decreases significantly.

2.2.2 Tar Reforming

Design and cost estimation of the tar reformer/regenerator presented a challenge to the team. Because no commercial data exists on design or cost for the performance outlined by the "goal" TCPDU case, a number of assumptions have been made:

- Reaction temperatures equal to the inlet gas temperature (1598 and 1576°F). These temperatures are derived from conversations with NREL. Recent experimental studies at Iowa State University on catalytic tar destruction have demonstrated successful operation at ~1350 to 1550°F [5]. Sensitivity cases were run at 1472 and 1200°F; the results show that heat duty is strongly impacted by the reaction temperature. Since the catalyst is the heat carrier in the reaction, the reaction temperature will greatly impact natural gas use and catalyst circulation rates. Minimizing these factors will trade-off with catalyst activity as the reaction temperature is lowered. This may be an area for future optimization and testing at the TCPDU.
- Low pressure operation for the regenerator to cut down on combustion air blower costs. This design is assuming the use of a pressurized rotary lock to increase recycle catalyst pressure. There is the risk that a rotary lock may be inadequate for this service due to the high catalyst circulation rates leading to premature erosion. If this is the case, either a lockhopper system or pressurized regenerator vessel would need to be included, significantly adding to the cost.
- Catalyst recycle rate based entirely off of thermodynamic requirements. Because of the endothermic reforming reactions, the regenerated catalyst must carry the heat necessary to maintain reactor temperature.
- Catalyst heat capacity of 0.25 Btu/lb/°F
- Plug flow within the reactor, with a Gas Hour Space Velocity (GHSV) of 2000/hr, to establish the basis for the bed volume and catalyst inventory. The calculated cracker

5 Zhang, R., Brown, R., Suby, A., Cummer, K., "Catalytic Destruction of Tar in Biomass Derived Producer Gas", Energy Conversion and Management, Vol. 45, pp. 995-1014, 2004.

bed length was multiplied by a factor of four to account for deviations from ideal plug flow.

- Bed diameter calculated by first estimating the minimum and maximum bed fluidization velocities, then an average of these estimates taken. Fluidization velocities calculated from catalyst and syngas properties.

Both ASPEN and HYSYS were used to model these systems, with all necessary thermodynamic and kinetic assumptions included. The results from both simulations came out very close to one another with a very high heat duty (~150 to 170 MMBTU/hr) and catalyst circulation rate (~24,000 to 29,000 MTPD) in each case. While the cost of the actual vessels are not very high ($1.3MM to $1.5MM), the catalyst load is substantial, and costs could be high based on what assumptions are made for catalyst losses and system maintenance requirements. Since the catalyst is regenerated in the process, minimizing losses is key to reducing operating costs.

2.2.3 Cyclones

A number of assumptions were made for the particle size distribution, efficiency, and outlet particle loading. Since no explicit direction was given by NREL, assumptions using experimental data from small-scale gasifiers was assumed and given to vendors for sizing (99%+ particulate removal and an average particle size of 50 μm).

2.2.4 Heat Integration

The process heating and cooling needs were evaluated and heat integration performed to maximize heat recovery. The process design includes a steam cycle that recovers the majority of the process heat by generating steam. For hot process streams that could not be integrated in the steam cycle, cooling water was used to provide cooling duty. A steam turbine is included in the design to generate power from the excess process steam.

2.2.5 Methanol Compressor

It was assumed that a clean syngas pressure of 1160 psia was required for methanol synthesis. Therefore, a compression system with interstage cooling has been included in the design.

2.3 OPERATING COSTS AND UTILITY REQUIREMENTS

Catalyst and chemical needs, along with utility requirements, can be seen in Tables 2-1 through 2-3. The units with the highest operating cost are the amine system and the tar cracker. Steam cost contributes the largest cost component for the amine unit. A portion of the steam required for the amine unit is extracted from the steam turbine, and the remainder is assumed to be imported. About 44,000 lb/hr of steam is imported for the low-pressure case, and 113,500 lb/hr for the high-pressure case. Imports may be unnecessary if excess steam from elsewhere in the gasification unit is available.

The other major source of operating cost is the catalyst requirement for the tar cracker. The tar cracker specifics were determined by estimating the minimum fluidization velocity, required space velocity, and the required heat duty demanded of the regenerated catalyst. The total amount of catalyst is equal to the settled bed volume of the two fluidized beds, plus an additional 10% for transfer line inventory. Due to the very high heat load and quantity of gas to be handled, the initial catalyst loading is substantial: ~300 tonnes in the HP case, and ~830 tonnes in the LP case.

The remaining catalyst and chemicals cost are in-line with the assumptions made by NREL; in fact, some of the costs used by NREL in the biomass to hydrogen report are used here either for consistency, or because little other information exists. For example, it is unknown what the cost will be of tar cracker catalyst that can perform as expected in the NREL "goal" design.

Nexant has not made assumptions for the total yearly operating cost at this time; this cost could vary considerably based on the assumptions made for plant performance and the assumptions for catalyst, chemicals, and power costs. An estimate for operating cost should be performed for an entire integrated gasification unit or biorefinery, instead of the clean-up unit as a stand-alone facility. Suggestions for proper estimation and reducing operating costs include:

- An availability of 85 to 90% would be appropriate for this design
- Both low and high pressure designs would likely require steam imports. This could come from purchases or excess steam production elsewhere in the gasification plant
- A 0.01% per day catalyst loss in the tar cracker, as assumed by NREL in the "goal" hydrogen design, is appropriate for initial cyclone operation, but will likely degrade over time. Typical catalyst assumptions and make-up rates for similar technologies range from 0.01% to 0.1%.

  If a loss rate of 0.01% is assumed, and costs for the ZnO beds are amortized over the year, the daily catalyst and chemical cost is $1931/day for the low-pressure case, and $1457/day for the high pressure case. This takes into account tar cracker losses, ZnO bed replacement, and LO-CAT™ requirements. This is shown in Table 2-1 below.

TABLE 2-1 CATALYST AND CHEMICAL REQUIREMENTS

Variable	Amount Required	Cost	Notes
Tar Reformer Catalyst	Low- Pressure Case: 1,820,000 lbs High-Pressure Case: 662,000 lbs	Price: $4.67/lb (NREL H_2 Report)	No commercial catalyst is currently available for this operation. Assuming a GHSV of 2000/hr, and a catalyst volume equal to the settled bed volume of the two fluidized beds plus 10% for transfer lines.
ZnO Catalyst	Low-Pressure Case: 777 cubic feet High-Pressure Case: 707 cubic feet	Price: $355/cubic foot (Johnson Matthey).	Initial fill then replaced every year. Catalyst inventory based on H_2S removal capacity from 2 ppmv to 0.1 ppmv.
Sulfur Recovery Chemicals	Low-Pressure Case: 1.7 Tonnes/Day of Sulfur Removal High-Pressure Case: 2.4 Tonnes/Day of Sulfur Removal	Price: $191/tonne sulfur removed (GTP Quote)	Assumes price for all LO-CAT™ chemicals required. Does not include utility requirements.

Steam, water, natural gas, and combustion air requirements are similar between both the high and low pressure cases. The main difference is in the power and cooling requirements. This is mostly due to the syngas compressor; the large energy and interstage cooling duty required adds considerably more to the utility requirements. Some of the cooling duty is recaptured in the steam system.

High-pressure case utility requirements can be seen in Table 2-2 below.

TABLE 2-2 HIGH-PRESSURE CASE UTILITY REQUIREMENTS

Item No	Item Name	Load BHP		Elect. Power	Steam M Pounds per Hour					Water, GPM		Cooling MMBTU/HR	Nat. Gas	Combustion Air
		Norm.	Max (3).	KW	445 psig	85 psig	5 psig	___ psig	Cond.	Proc.	C.W. circ. (2).	Water	MMSCFD	MMSCFD
H-200	Quench Water Recirculation Cooler										2,232	22.3		
H-302	Lean Solvent Cooler										13,487	135.0		
H-303	Amine Stripper Reboiler					243.9			244					
H-305	Acid Gas Condenser										8,520	85.3		
H-400A	K-400 Interstage Cooler										1,046	10.5		
H-401	MeOH Reactor Preheater					17.61			17.6					
H-501	Blowdown Cooler										84	0.8		
K-100	Combustion Air Blower	1,022		762										
K-320	Flue Gas Blower	207		154							2	0.02		
K-400	MeOH Compressor - 2 Stages	8,388		6,257							84	0.8		
P-201	Quench Water Recirculation Pump	3		2										
P-300	Lean Solvent Pump	1,474		1,100										
P-500	Condensate Make-up Water Pump	1		1										
P-501	Deaerator Feed Pump	8		6										
P-502	Boiler Feed Water Pump	710		530										
R-xxx	Gasifier				139.6									
R-100	Tar Reformer				26									
R-101	Catalyst Regenerator												7.8	84.4
	LO-CAT unit	1,004		749			0.9			2,500				
M-501	Extraction Steam Turbine/Generator	(19,721)		(14,712)	(165.6)	(148.0)								
	TOTAL	**(6,903)**		**(5,150)**	**0**	**114**	**1**		**262**	**2,500**	**25,454**	**255**	**8**	**84**

NOTES: 1. All Figures shown above represent normal utility usage requirements except:
() indicates normal utility make
* indicates intermittent usage or make, not included in totals
2. CWS temperature is 80 F and CWR temperature is 100 F. Makeup water to cooling tower is not shown
3. Utility consumption for max. load conditions is not shown.

Low-pressure case utility requirements can be seen in Table 2-3.

TABLE 2-3 LOW-PRESSURE CASE UTILITY REQUIREMENTS

Item No	Item Name	Load BHP		Elect. Power	Steam M Pounds per Hour					Water, GPM		Cooling MMBTU/HR	Nat. Gas	Combustion Air
		Norm.	Max (3).	KW	85 psig	35 psig	5 psig	___ psig	Cond.	Proc.	C.W. circ. (2).	Water	MMSCFD	MMSCFD
H-200	Quench Water Recirculation Cooler										2,213	22.2		
H-300A	1st Stage intercooler										12,188	122.0		
H-300B	2nd Stage intercooler										3,276	32.8		
H-300C	3rd Stage intercooler										2,766	27.7		
H-300D	Post compressor cooler										1,819	18.2		
H-402	Lean Solvent Cooler										11,388	114.0		
H-403	Amine Stripper Reboiler					150.6			151					
H-405	Acid Gas Condenser										2,900	29.0		
H-500A	K-500 Interstage Cooler										1,105	11.1		
H-501	MeOH Reactor Preheater				18.8				18.8					
H-601	Blowdown Cooler										61	0.6		
K-100	Combustion Air Blower	910		679										
K-300	Syngas Compressor - 4 Stages	38,786		28,934										
K-420	Flue Gas Blower	347		259							3	0.03		
K-500	MeOH Compressor - 2 Stages	8,717		6,503							87	0.9		
P-201	Quench Water Recirculation Pump	20		15										
P-400	Lean Solvent Pump	802		599										
P-600	Condensate Make-up Water Pump	1		1										
P-601	Deaerator Feed Pump	7		5										
P-602	Boiler Feed Water Pump	570		425										
R-xxx	Gasifier					73.47								
R-100	Tar Reformer					53								
R-101	Catalyst Regenerator												7.0	74.8
	LO-CAT unit	639.9		477			0.56			1,800				
M-601	Extraction Steam Turbine/Generator	(26,019)		(19,410)	(18.8)	(232.9)								
	TOTAL	**24,781**		**18,486**	**0**	**44**	**1**		**169**	**1,800**	**37,806**	**378**	**7**	**75**

NOTES:
1. All Figures shown above represent normal utility usage requirements except:
   () indicates normal utility make
   * indicates intermittent usage or make, not included in totals
2. CWS temperature is 80 F and CWR temperature is 100 F. Makeup water to cooling tower is not shown
3. Utility consumption for max. load conditions is not shown.

2.4 DIFFERENCES WITH NREL BIOMASS TO HYDROGEN DESIGN

In general, the cost of the clean-up section of the biomass to chemicals designs is more expensive than for the NREL Biomass to Hydrogen design[6]. There are three main reasons for this: more equipment necessary in the chemicals designs, the increase in steel prices from 2002 to 2005, and different engineering assumptions made in the chemicals case. Information on each reason will be elaborated upon below.

2.4.1 Added Equipment to Chemicals Design

The two major unit operations that are new to this design versus the hydrogen cases are the amine unit and the syngas compressor for methanol synthesis. In the hydrogen cases, a LO-CAT™ unit and ZnO bed was used for H_2S removal, while the PSA removed carbon dioxide. The chemicals cases also use the LO-CAT™ and ZnO units, but instead of a PSA, an amine unit is used for the bulk H_2S and CO_2 removal. The cost for the amine units is driven largely by the need for CO_2 removal; due to the low H_2S content in the syngas, the cost of the amine unit would be roughly half as much if CO_2 removal was not required. The LO-CAT™ unit is used in this case for clean-up of the acid gas stream from the amine unit instead of bulk H_2S removal. Because of the CO_2 content and different operating requirements versus the hydrogen case, the quote provided by GTP is roughly double the price used in the hydrogen case.

6 Spath, P.; Aden, A.; Eggeman, T.; Ringer, M.; Wallace, B.; Jechura, J. (2005). Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier. 161 pp.; NREL Report No. TP-510-37408.

In order to compress the clean syngas up to methanol synthesis pressure, a ~8,000 HP compressor is required. This unit was not necessary in the hydrogen case, adding to the overall cost. Taking into account a $12MM credit by not using the PSA, the LP cost increases by ~$8.5MM, while the HP cost increases by ~$18.5MM due specifically to the extra equipment needed.

2.4.2 Increase in Steel Price

NREL used 2002 as the cost basis for the biomass to hydrogen designs, while Nexant is using Q2 2005. The increase in steel price between 2002 and 2005 has been significant, impacting the prices quoted in the Nexant design. The Q2 2005 basis for hot-rolled steel is ~$400 to $450/ton, up from ~$250 to $300/ton in 2002[7]. Steel prices have been very volatile in the last 3 years due to strong worldwide demand, a sharp rise in energy prices, consolidation in the US steel market, and a weak US dollar.

Because of this basis difference, the 2002 NREL basis would need to be escalated not only for inflation but also for steel price in order to put it on the same basis as this study. It is difficult to place a blanket escalation factor on the design due to the impacts that steel price has on different pieces of equipment; for example, this may make up much of the difference in price in equipment like vessels and exchangers, but have less of an impact on compressor prices. Each unit should be evaluated independently to determine the impact that steel price has on overall unit cost.

2.4.3 Engineering Assumptions

A side-by-side comparison of all the major process units was performed for the HP and LP cases versus the NREL hydrogen design. A few differences were noticed that are outlined below. A direct comparison cannot be performed on units that were lumped into the "Gas Cleanup" section of the NREL design and not explicitly sized. While the major differences are outlined here, only a brief attempt at determining the cost difference has been made.

Reactors and Columns

ZnO Beds: While the size of the ZnO beds in this design is smaller than the hydrogen case, the installed cost is roughly double. This is likely due to the difference in steel price.

Tar Reformer/Regenerator: In the hydrogen design, this is included in the "Cleanup" costs, so no explicit design information is available. The NREL assumption for "Cleanup" took the average of a number of different studies; however, only one of these studies, Weyerhaeuser (2000), had a tar cracker. The "Cleanup" section for the Weyerhaeuser study was ~$9MM greater than the other designs, implying that the majority of the cost may be due to the tar cracker cost. The NREL "Cleanup" assumption may be low since the hydrogen design has a tar cracker, yet only one of the studies used to obtain the "Cleanup" cost also has a tar cracker.

7 For more information, see the Bureau of Labor Statistics "Producer Price Series", along with Lazaroff, Leon, "Steel Regains Some Luster", Detroit Free Press, 25 July 2005

Cyclones

Since these were part of the "Cleanup" average, no explicit design numbers were provided as part of the hydrogen study. Design quotes from vendors are used for this part of the plant in the chemicals design.

Vessels

The Nexant estimate is higher than the hydrogen design due to 1) the venturi and quench being included as part of the "Cleanup" estimate, 2) larger vessel sizes for the steam system than what was assumed in the hydrogen design, and 3) steel prices. Depending on the price assumed for the venturi /quench in the hydrogen design, the Nexant estimate appears to be ~$3MM greater than the hydrogen case.

Heat Exchangers

A number of differences exist between the hydrogen and chemicals designs, making the installed cost for exchangers in the chemical production case ~$4MM to $6MM higher than in the hydrogen case:

- There is a large cost discrepancy between the exchangers downstream of the tar reformer. The Nexant designs are larger and considerably more expensive; Nexant assumed refractory lining, while it is unclear if this assumption is made in the hydrogen design.
- The Nexant design has a number of exchangers not included in the hydrogen design: amine precoolers (HP case), methanol compressor coolers (both cases), and ZnO coolers (both cases).
- A few of the exchangers in the hydrogen design are included in the "Cleanup" section, so it is difficult to make a direction comparison.

Compressors and Blowers

As mentioned earlier, the syngas compressor for methanol synthesis adds ~$7MM to the installed cost relative to the hydrogen case. This compressor was not necessary in the NREL hydrogen design.

There is a major difference between the NREL and Nexant assumptions for the syngas compressor in the LP case. While NREL shows an installed cost of ~$12MM for a 30,000 HP compressor, Nexant estimates that a ~38,000 HP compressor is required at an installed cost of ~$37MM ($15MM for the equipment alone). The equipment cost comes directly from Elliott Compressor; checks on the validity of the estimate using cost curves, ICARUS, and other vendors show that this is within the +/- 30% estimate desired by the study. The NREL study assumed that an integrally geared compressor type would be appropriate, while this report uses a horizontally split centrifugal compressor recommended by vendors. Analysis using cost estimating software shows that this assumption is the main reason for the cost difference.

Pumps

Both Nexant and NREL designs are in agreement in regards to the pumps.

Steam Turbine

The Nexant estimate is slightly higher than the NREL estimate, ~$12MM installed versus $10MM. This difference is likely due to steel prices.

The other difference that should be pointed out between the hydrogen and chemicals cases is the assumption made for the installation factor. NREL used a 2.47 installation factor, which is derived from literature sources. Nexant used 2.57 in both the HP and LP cases, except on the process gas compressor, where 2.47 is used. These numbers are derived independently from previous experience and vendor engineering estimates. While the factors are very similar to one another, this difference can make a 4% difference ($2MM) on an equipment cost of $20MM.

2.5 CHANGING FLOWS, CONDITIONS, AND COMPOSITIONS

Per the scope of work outlined by NREL as part of this project, Nexant has been asked to provide input on how the design estimates will be adjusted if the syngas flowrates or compositions vary. Information for both the high and low-pressure cases, along with the scaling factors appropriate for each major piece of process equipment, are outlined below.

2.5.1 Flowrate Impacts

In general the limits on process equipment sizes are usually the result of manufacturing restraints, transportation limits, and maintenance restrictions. For this evaluation, it was assumed that the throughput would be increased by 50% and the equipment size or capacity would increase accordingly. The affects of this change are discussed below with respect to both the low- and high-pressure cases.

Low-Pressure Syngas Design Cases

For the Low-Pressure Syngas Design Cases some of the equipment has already reached size limitations that required multiple trains or parallel equipment. Thus, increasing the capacity by 50% will require more parallel equipment and a more complex and expensive piping manifold. Examples include:

- Gasifier Cyclones (4 required for the base capacity)
- Tar Reformer SG Cooler/Steam Generator (2 required)
- Tar Reformer SG Cooler/BFW Preheater (2 required)
- Compressor Interstage Cooling - 1st stage (2 required)
- Syngas Venturi Scrubber/Quench Tower (2 required)

Thus, for a 50% increase in capacity, the design would require 6 gasifier cyclones, 3 of each major heat exchanger, and 3 venturi scrubbers.

Other items, such as the 1st Stage KO Drum, may require either a parallel unit or field construction due to equipment size and weight limitations during transportation. While the limits for ground transportation vary from state to state, typically, codes limit standard transport sizes to ~14 feet in width and height, 53 feet long and 80,000 pounds. Locating this facility in Iowa will mean that most equipment will be transported to the site either by rail or truck. Access to the Mississippi or Missouri Rivers may allow larger vessels to be used. For the 1st Stage KO Drum, the inside diameter would increase to about 16 feet (from a 13 foot diameter) at a capacity 50% greater than the base case. However, when considering transportation by road, auxiliary equipment such as nozzles and flanges must be taken into consideration. This item would be well beyond most road transportation limits in the U.S. To manage this limitation, options are either transportation by rail or barge, parallel pieces of equipment, or field fabrication.

Other equipment may exceed the maximum recommended size for a single train, and would require a second, parallel unit. This includes items such as the Syngas Compressor and the shell and tube heat exchanger for the Flue Gas Cooler/Steam Superheater service. In the latter case, the size of the heat exchanger is actually a maintenance issue. The diameter of the tube bundle of these units is larger than a normal bundle puller could handle (maximum limit is about 6-7 feet diameter). It then becomes an economic question of bringing in special maintenance equipment during turnarounds or using smaller, parallel process equipment.

High-Pressure Syngas Design Cases

For the High-Pressure Syngas Design Cases, most of the equipment is smaller than the corresponding equipment for the Low-Pressure Syngas Design Cases as a result of the high pressure operation. Only a few items, when scaled by +50%, would require a parallel unit. Two major exchangers, the Tar Reformer SG Cooler/Steam Generator and Flue Gas Cooler/Steam Superheater, were discussed above. Another area is equipment within the LO-CAT™ unit. These include the Inlet Gas KO Drum and the LO-CAT™ Oxidizer Vessel. The former would require a vessel with an inside diameter of over 17 feet and the latter would required an inside diameter of about 16 feet. As noted previously, the outside diameter (including nozzles and flanges) would be well beyond most road transportation limits in the U.S. Vendors for process items of this nature can provide input for the appropriate process configuration for this service.

Appropriate vessel sizing for the amine system is also of concern in this design. The amine system contains two relatively large columns – the scrubber and the regenerator. Considering a 50% increase in capacity, the column diameters will increase by about 20 to 25%. In particular, the regeneration column may exceed the transportation size limitations and thus, require parallel trains or field fabrication.

General Information

A plant that is 50% larger will require more plot area not only due to the larger equipment and storage, but due to offsite considerations. For example, the flare will have to be designed for a load that is 50% larger. This will require either a taller flare or moving the flare further away from the main process units. A higher flare may meet with height restrictions. Thus, the area that is restricted around the flare may increase.

Estimating the Capital Investment Cost

In most cases the capital cost for a capacity increase or decrease of 50% can be estimated using exponential methods. That is, the new capital cost can be estimated by using capacity ratio exponents based on published correlations and the following formula:

$$C_2 = C_1 (q_2/q_1)^n$$

where C stands for cost, q for flowrate, and where the value of the exponent n depends on the type of equipment. In reviewing the literature for the various exponents, some discrepancies in published factors are apparent due to variation in definition, scope and size. Technology has also advanced over time, making it less expensive to produce larger machinery now than in years past. In addition, new regulations dictate expenditures for environmental control and safety not included in earlier equipment. In the table that follows, the most recent literature information is listed. Traditionally, when a specific value is not known, an exponent value of 0.6 is often used for equipment and a value of 0.7 for chemical process plants (usually expressed in terms of annual production capacity). Table 2-4 gives typical values of n for most of the equipment included in these designs.[8,9,10,11,12]

TABLE 2-4 EXAMPLES OF TYPICAL EXPONENTS FOR EQUIPMENT COST VERSUS CAPACITY

Equipment	Size Range	Units	Exponent**
Reactor – fixed beds	N/A		0.65-0.70
Column (including internals)	300-30,000	Feed rate, million lb/yr	0.62
Cyclone	20-8,000	Cubic feet/m	0.64
Vessel – vertical	100-20,000	US gallons	0.30
Vessel – horizontal	100-80,000	US gallons	0.62
Heat exchanger (S&T)	20-20,000	Square feet	0.59
Venturi scrubber	N/A		0.60
Compressor – centrifugal*	200-30,000	hp	0.62
Blower*	0.5 - 150	Thousand standard cubic feet per minute	0.60
Pump*	0.5-40 40-400	hp	0.30 0.67
Turbine Pressure discharge Vacuum discharge	 20-5,000 200-8,000	hp	0.81
Motor	10-25	hp	0.56

8 Perry, Robert H., and Green Don W., Perry's Chemical Engineers' Handbook, 7th edition, page 9-69.
9 Walas, Stanley M., "Chemical Process Equipment – Selection and Design," Butterworths, page 665
10 Blank, L. T. and A. J. Tarquin, "Engineering Economy," McGraw-Hill
11 Peters, Max S. and Timmerhaus, Klaus D., "Plant Design and Economics for Chemical Engineers," McGraw-Hill, page 170
12 Remer, Donald S. and Chai, Lawrence H., "Design Cost Factors for Scaling-up Engineering Equipment," *Chemical Engineering Progress*, August 1990, pp 77-82

Equipment	Size Range	Units	Exponent**
	25-200		0.77
Package unit	N/A		0.75
Other	N/A		0.6 – 0.7

* excluding driver

** this estimating method gives only the purchase price of the equipment; additional installation cost for labor, foundations and construction expenses will make the final cost higher.

2.5.2 Composition Impacts

The major units that will be impacted by a large change in syngas composition are the tar reformer and the venturi scrubber. Due to the relatively low concentration of sulfur in the syngas stream, +/-50% fluctuations in the H_2S content should not impact how the sulfur removal system is designed. Significant changes in the inlet H_2/CO ratio may also require modifications of the design in order to establish the appropriate downstream composition.

The obvious change that will influence the design of the tar reformer is the amount of hydrocarbons in the syngas from the gasifier. Currently, the design is assuming that a separate reformer is not necessary, with the tar reformer converting most hydrocarbons exiting the gasifier. If either the hydrocarbon yield increases or the tar reformer conversion is lower than planned, a separate reformer for light hydrocarbons should be considered. The amount and type of hydrocarbons will affect the operating conditions which will in turn affect the water gas shift reaction. A change in the H_2/CO ratio may require divorcing the shift reaction from the tar reformer (i.e., a separate shift reactor instead of just adding steam to the tar reformer).

A 50% increase in particulates may require different/larger cyclones or a redesign of the venturi scrubber in order to handle the larger load. This is largely controlled by the gasifier operation; reliable performance data should be established prior to deciding upon a particulate removal scheme. Higher particulate loading than planned can significantly hurt overall plant performance.

A 50% increase in H_2S will not affect the sulfur recovery processes. LO-CAT^{TM} can handle between 150 lbs to 20 tonnes of sulfur per day, and concentrations between 100 ppm and about 10% H_2S. Even at 50 percent more H_2S, the concentration still remains within the operating limits for LO-CAT^{TM}. In addition, the solvent circulation rate in the amine unit can be increased to remove additional H_2S if the sulfur concentration is higher than expected.

2.6 FOLLOW-UP AND AREAS FOR FURTHER STUDY

The analysis performed sets the base case for the clean-up section of two different biomass-to-chemicals designs. After in-depth analysis of these cases, the team has identified a number of areas for further study:

- ***Alternatives for Tar Removal:*** A number of assumptions have been made for sizing and costing of this unit. Greater study and analysis, both in the laboratory and through simulations, should be performed to determine if the methods used are valid. In addition, alternative tar removal technology should be considered, including:

  - Introduction of tar cracking catalyst into the gasifier. Typically, this has not been done due to concerns with deactivation and erosion.
  - Gasifier operation to reduce hydrocarbon yields.
  - Using a water wash for tars, followed by a standard reformer for hydrocarbons. While this increases the cost of quenching and wastewater handling, the cost tradeoff may be economic.

- ***Process Integration, Gasification Systems and Biorefinery:*** Integration of the clean-up section with the other parts of the gasification plant will provide a better picture of the overall plant costs. In addition, use of this thermochemical platform has been considered for future application into an integrated "biorefinery". This base case could be used for a determination of the process requirements and offerings that a thermochemical platform could provide.
- ***Alternate CO_2/Sulfur Removal Steps:*** Based on the design information provided and past studies that have been examined, the steps incorporated for CO_2 and sulfur removal has been determined to be appropriate at this stage. A cost comparison of amine versus physical solvents and new technologies for acid gas removal would provide additional data to confirm the appropriate use of amine in this design.

  New technology is currently being explored to remove sulfur without having to cool to 110°F or below. Since none of this technology is currently commercial, it has not been evaluated for use in this design. If available however, warm sulfur clean-up may increase efficiency in this design, by reducing the amount of reheat necessary prior to entering the shift reactor.
- ***Other Impurities in the Syngas:*** For the low pressure case, a scrubber has been included to remove residual ammonia, and any metals, halides, or alkali remaining in the system. If it is deemed that the level of these impurities entering the scrubber will not adversely impact the FT or methanol catalysts, this step could be removed.

Section 3 Labor Requirements

3.1 SUMMARY

The labor projections for the 2000 MTPD biomass gasification plant are based on a combination of 1) models developed from Emery Energy's 70MWe Gasification Plant design completed under prior DOE contracts, 2) additional "adders" for the scale and complexity (chemical plant nature / hydrogen production) of the 2000 MTPD plant being considered, and 3) previous experience of Nexant and other team members. The high pressure, oxygen-blown, 2000 MTPD plant requires labor skills with slightly greater operating experience than power-only facilities, and thus commands a premium for these skills.

The labor rates derived from Emery's 70 MWe Biomass IGCC (1200 MTPD plant) case were ~$1,650,000 per year (not including subcontracted services) versus the $2,274,720 projected for the labor costs for the 2000 MTPD biomass to chemicals design. This difference of roughly $625,000 represents the higher level of experience needed for the larger plant, greater materials handling rates, and increased labor for plant maintenance. A discussion of the reasons for this difference, along with differences between the recent NREL Biomass to Hydrogen report, is contained below. Some of the main differences with the NREL Hydrogen report include different job descriptions, the use of a back-up shift crew, utilization of contract labor, and lower assumptions for overhead costs.

3.2 LABOR REQUIREMENTS

The following labor categories and positions will be required for the 2000 MTPD biomass plant.

- ***General Plant Manager:*** Responsible for all personnel and plant decisions, including new employee hiring, operator training, fuel contracts, maintenance contracts, general equipment purchases, external communications, and operating schedules. Engineering degree required, with 10+ years of chemical plant operating experience. Salary of $100,000/yr.
- ***Administrative Assistant/Company Controller:*** Support the general plant manager, manages personnel records, completes company payroll, manages time accounting records, manages company benefits, employee investment accounts, and insurance enrollments. Accountant degree required with 5+ years of experience. Salary of $45,000/yr.
- ***Secretary/Receptionist:*** Supports the General Plant Manager and Company Controller. Receives visitors, answers phone, and attends to office administrative duties. Salary/Wages of $25,000/yr.
- ***Laboratory Manager:*** Oversees all laboratory equipment and laboratory technicians. Responsible for product quality; testing performed both on finished product and intermediate streams (via on-line equipment and sample draws). Works straight days, with some overtime possible. Salary/Wages of $50,000/yr.
- ***Laboratory Technician:*** Responsible for sample gathering, analytical equipment maintenance, and laboratory testing. Works straight days, with some overtime

possible. Shift operating crew can assist with some sample gathering as necessary; contract equipment technicians can assist with analytical equipment repair as necessary. Salary/Wages of $35,000/yr.

- ***Shift Operating Crew:*** The plant will be operated by a four-member crew shift each week, with responsibilities defined below:
- ***Shift Superintendent.*** The shift superintendent is the chief operator who mans the control station and simultaneously directs the activities of the shift crew. The shift superintendent is a degreed engineer who understands the plant, understands the technical and physical operations, and makes key operating decisions. The shift superintendent ensures compliance with plant quality, safety, industrial hygiene, and environmental requirements. 5-10 years of chemical plant operating experience is preferred for this position. Salary of $75,000/yr.
- ***Support Operator.*** The support operator aids the shift superintendent with plant operation. The support operator is also tasked with bulk material handling such as feedstock receipts/inspection/weigh-in and ash weigh-out/disposal shipments. The support operator attends to feed and ash sampling/characterization, waste water disposal sampling, and provides general plant support in relief of the shift superintendent. The support operator is also tasked with monitoring plant emissions rates, including daily/weekly calibration of effluent gas monitors. The support operator verifies that plant operating records and daily logs are correct. This position coordinates fuel characterizations and waste water analyses. A novice degreed engineer or experienced technician is sufficient for this position. Salary of $45,000/yr
- ***Millwright.*** The shift millwright conducts hourly and daily equipment inspections, safety rounds, completes scheduled equipment process maintenance, supports equipment maintenance and equipment replacements, contracts and supervises crafts such as pipe fitters, electricians, welders, and special instrument technicians when such functions exceed the millwright's capabilities. The millwright preferably has an associate degree in mechanical, industrial, or design engineering technology with 5-10 years experience. Salary of $60,000.
- ***Millwright Assistant/Yard Labor.*** Supports millwright and accompanies millwright and contracted crafts, particularly during dangerous work activities, such as confined space entries and working from heights. The millwright assistant supports tool setup, job errands, and plant cleanup. Salary of $35,000.

Shifts run for 12 hours with two crews per day. Crews report to work 30 minutes prior to the shift turnover to perform receive shift operating instructions and to pass information on critical operations and maintenance. Each crew member is allotted 30 minutes for a meal break. Thus, each shift extends 12.5 hours, with 0.5 hours meal break, or 12 hours of labor. Crews operate on a 4 days on / 4 days off rotation. This requires 84 hours on average per crew member for any two-week pay period.

Five complete shift teams are engaged. The fifth crew provides coverage for individual vacations, sick leave, and holidays. The fifth crew also fills in for continuing training and for new hire training. The fifth crew also supports ongoing maintenance and periodic outage/turnaround planning. In addition, the fifth crew supports updates to control system programming, data collection, and instruments. The millwright assistant on the fifth crew supports plant cleanup and janitorial activities. The fifth crew works 40-hour straight days when not substituting for members of the four-crew rotation.

Table 3-1 summarizes the plant operating labor by category, salary, and total cost.

TABLE 3-1 LABOR COSTS

Position	Number	Base Salary or Hourly Rate	Annual Overtime and Holiday Hours	Overtime Rate	Total Annual Cost
General Plant Manager	1	$100,000	N/A	N/A	$100,000
Company Controller	1	$45,000	N/A	N/A	$45,000
Secretary/ Receptionist	1	$25,000	None	N/A	$25,000
Laboratory Manager	1	$50,000	240	$30	$57,200
Laboratory Technician	2	$35,000	240	$22.50	$80,800
Shift Superintendent	5	$75,000	680	$45	$405,600
Support Operator	5	$45,000	680	$25	$242,000
Millwright	5	$60,000	680	$32.50	$322,100
Millwright Assistant	5	$15.00/hr	560	$22.50	$144,000
Total Base Salaries and Wages					$1,421,700
General Overhead and Benefits (60% of total salaries)					$853,020
Total Base Wages and Benefits					**$2,274,720**
Subcontracted Crafts					
Welder	$80/hr	1200			$96,000
Electrician	$75/hr	640			$48,000
Pipe Fitter	$65/hr	600			$39,000
Insulator/Painter	$60/hr	400			$24,000
Carpenter	$55/hr	400			$22,000
Instrument Technician	$90/hr	400			$36,000
Total Subcontracted Labor					**$265,000**
Total Labor and Benefits (Operating Labor Cost)					**$2,539,720**

3.3 DIFFERENCES WITH EMERY ENERGY 70 MWE CASE

Both the complexity and size of this facility increases the labor costs over what Emery Energy has assumed for their 70 MWe biomass gasification facility. The size of the unit (1200 MTPD vs. 2000 MTPD) slightly increases the number of shift workers and contract hours required, but does not increase the plant management or engineering requirements. This represents an economy-of-scale advantage enjoyed by larger gasification facilities; while the total labor requirement is greater than the 1200 MTPD facility, the marginal amount of labor required decreases as plant size increases.

This design contains additional equipment than what is assumed in Emery Energy's 70 MWe facility design. While this design does not contain a gas turbine, steam turbine, or HRSG, additional equipment includes enhanced sulfur removal (an amine system and ZnO beds), chemicals synthesis equipment, and tar cracking. It is this increase in complexity, rather than the increase in size, that adds the majority of the increase in labor costs.

3.4 DIFFERENCES WITH NREL BIOMASS TO HYDROGEN CASE

In the 2005 study, NREL made assumptions for the labor requirements necessary for a 2000 TPD wood gasification to hydrogen plant. The size being considered in this design is exactly the same, and the complexity is roughly the same as the NREL case. The only main difference is the inclusion of chemicals synthesis equipment, which takes the place of the PSA and related equipment required for hydrogen production.

The labor requirements developed for the chemicals synthesis cases are lower by almost $1.5MM due to the assumptions made by the Nexant team. The main differences are highlighted below:

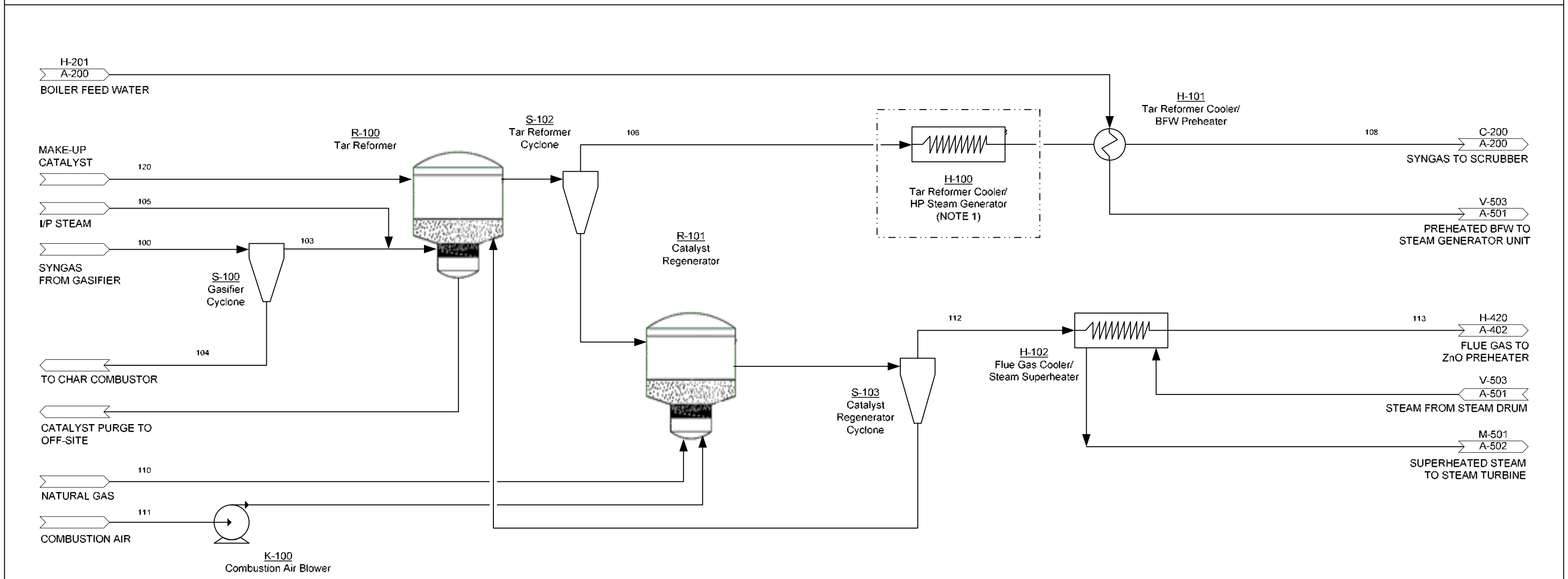
- ***Salary Assumptions:*** In general, slightly higher salaries are assumed in the chemicals synthesis design for employees such as the plant manager, engineers, and operators. Higher salaries may be necessary to attract workers to facilities employing complicated and novel technologies.
- ***Administrative Assistants:*** Instead of the three assistants assumed by NREL, this design assumes only two: the company controller/administrative assistant and the main receptionist. The main difference is that the truck handling work performed by the assistant in the NREL design will now be split amongst the millwrights and assistants.
- ***Work Assignments for Shift Workers:*** As mentioned in the job descriptions, it is assumed that support operators will assist with yard issues, feedstock delivery, and field work, while the superintendent will largely be responsible for control issues. This reduces the need for yard employees and operators whose sole job is to man control boards. The five crews effectively allow for additional personnel capable of supporting offloading and weighing of the biomass feedstock.
- ***Subcontract Labor:*** In order to reduce the need for full-time staff for part-time work, a number of specific skills, such as welders, electricians, and carpenters, will be

contracted out. This reduces the overall labor costs and overhead. No subcontract labor was assumed in the NREL hydrogen case.

- ***Overhead:*** The labor estimate made in this case has roughly half as much full-time staff by utilizing more contract labor and changing the job description of day and shift employees. This is one reason that the estimate for overhead expenses (60%) is less than the biomass to hydrogen case (95%). In addition, the assumption has been made that a small firm will own and operate this facility. In general, overhead has been found to be less in smaller firms than in large multinationals; this assumption could be revised based on the ownership basis. This assumption for the overhead rate has been confirmed by Emery Energy, and is consistent with other small gasification companies that have limited facilities and indirect labor costs.
- ***Overtime Assumptions:*** The NREL hydrogen case assumed straight salaries for all employees, with no overtime. The chemicals case assumes ~2500 hours of overtime per year, roughly split over the 4 main shift worker categories. Allowing overtime reduces the number of full-time employees required, and decreases overall labor costs versus the NREL hydrogen case.
- ***Back-Up Shift Crew:*** Unlike the NREL hydrogen design, the back-up fifth shift team would be available to cover a number of different duties during the day shift, decreasing the need for specialty workers in each area.

Appendix A High-Pressure Syngas Design Case PFDs

S-100	S-102	S-103	R-100	R-101	H-100	H-101	H-102	K-100
Gasifier Cyclone	**Tar Reformer Cyclone**	**Catalyst Regenerator Cyclone**	**Tar Reformer**	**Catalyst Regenerator**	**Tar Reformer Cooler/ HP Steam Generator**	**Tar Reformer Cooler/ BFW Preheater**	**Flue GasCooler/ Steam Superheater**	**Combustion Air Blower**
5' ID x 25' T/T	5'-6" ID x 27'-6" T/T	5' ID x 25' T/T	20' ID x 20' T/T	22' ID x 70' T/T	Duty: 203.7 MMBTU/hr	Duty: 50.8 MMBTU/hr	Duty: 83.7 MMBTU/hr	61910 ACFM 3 BHP



	Unit	100	101	103	104	105	106	108	110	111	112	113
Mass Flow	lb/hr	418,416	409,164	408,980	9,436	26,000	434,982	434,982	14,650	265,200	279,847	279,847
Temperature	F	1,576	1,576	1,576	1,576	757	1,576	370	60	90	1,798	839
Pressure	psia	460.0	460.0	459.9	459.9	460.0	456.9	446.9	14.7	14.7	14.7	14.5
Vapour Fraction		0.96	1.00	1.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
MW		21.5	21.9	21.9	12.0	18.0	18.7	18.7	17.1	28.6	27.6	27.6
Heat Flow	MMBtu/hr	-1,423.6	-1,437.5	-1,437.7	14.1	-141.8	-1,419.7	-1,674.3	-28.3	-29.5	-218.0	-301.7
Components												
Hydrogen	lb/hr	4,935	4,935	4,935	-	-	14,182	14,182	-	-	0	0
CO2	lb/hr	159,234	159,234	159,234	-	-	161,857	161,857	188	131	39,529	39,529
CO	lb/hr	42,314	42,314	42,314	-	-	90,773	90,773	-	-	0	0
H2O	lb/hr	170,183	170,183	170,183	-	26,000	162,870	162,870	-	5,175	36,433	36,433
Methane (CH4)	lb/hr	23,334	23,334	23,334	-	-	4,667	4,667	12,995	-	0	0
Acetylene (C2H2)	lb/hr	-	-	-	-	-	0	0	-	-	0	0
Ethylene (C2H4)	lb/hr	523	523	523	-	-	52	52	-	-	0	0
Ethane (C2H6)	lb/hr	1,122	1,122	1,122	-	-	11	11	800	-	0	0
Propane	lb/hr	-	-	-	-	-	0	0	183	-	0	0
Isobutane	lb/hr	-	-	-	-	-	0	0	50	-	0	0
n-Butane	lb/hr	-	-	-	-	-	0	0	50	-	0	0
Pentane	lb/hr	-	-	-	-	-	0	0	122	-	0	0
Benzene (C6H6)	lb/hr	4,370	4,370	4,370	-	-	44	44	-	-	0	0
Tar (C10H8)	lb/hr	2,390	2,390	2,390	-	-	2	2	-	-	0	0
Ammonia (NH3)	lb/hr	318	318	318	-	-	32	32	-	-	0	0
H2S	lb/hr	254	254	254	-	-	254	254	-	-	0	0
Char	lb/hr	9,440	188	4	9,436	-	4	4	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	0	0	-	60,170	3,898	3,898
Argon	lb/hr	-	-	-	-	-	0	0	-	3,349	3,349	3,349
Nitrogen	lb/hr	-	-	-	-	-	235	235	263	196,376	196,638	196,638
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-

NOTES:
1. SEE SECTION A-501 FOR DETAIL OF STEAM GENERATOR UNIT

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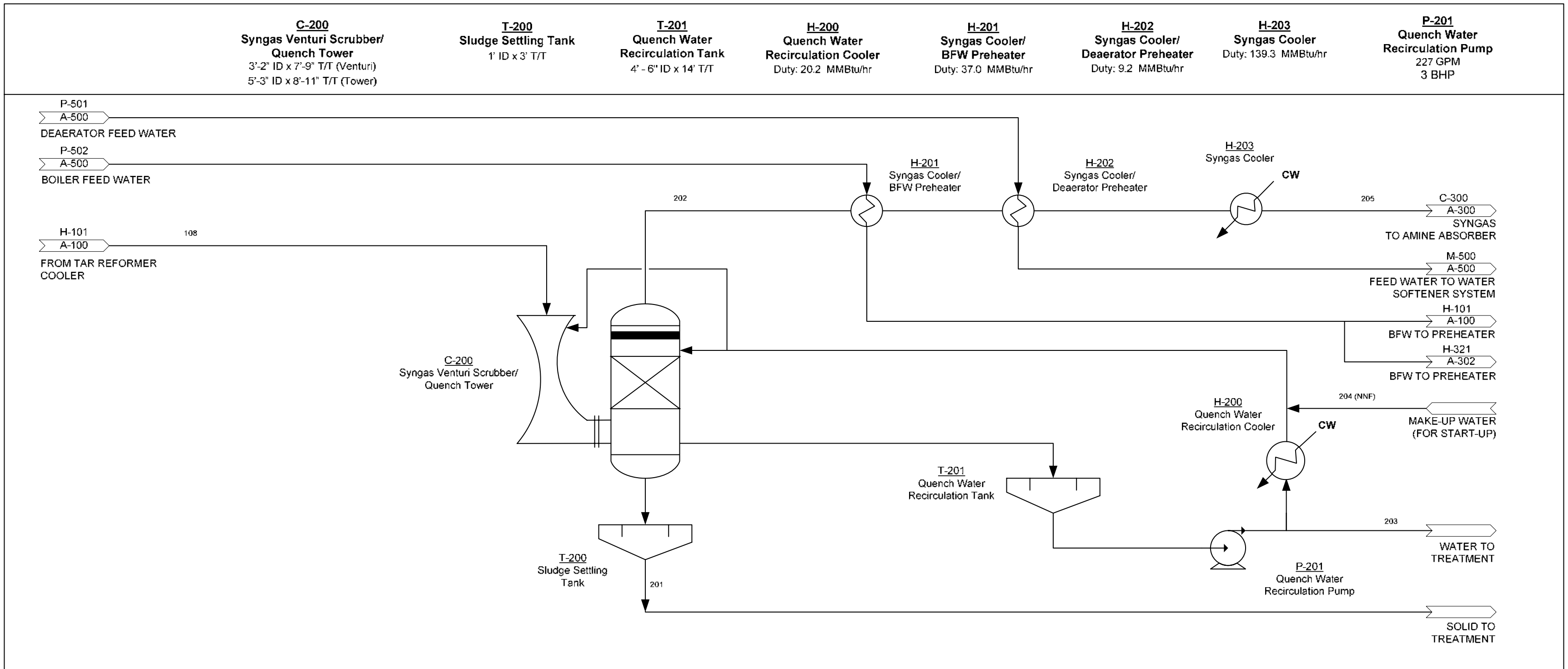
NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-100: TAR REFORMING

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov, 2005



	Unit	108	201	202	203	205
Mass Flow	lb/hr	434,982	504	413,207	21,271	413,207
Temperature	F	370	364	356	364	110
Pressure	psia	446.9	444.5	444.5	456.0	429.5
Vapour Fraction		1.00	0.00	1.00	0.00	0.65
MW		18.7	18.0	18.7	18.0	18.7
Heat Flow	MMBtu/hr	-1,674.3	-3.2	-1,554.2	-138.2	-1,739.6
Components						
Hydrogen	lb/hr	14,182	0	14,179	1	14,179
CO2	lb/hr	161,857	1	161,688	36	161,688
CO	lb/hr	90,773	0	90,758	3	90,758
H2O	lb/hr	162,870	499	141,289	21,230	141,289
Methane (CH4)	lb/hr	4,667	0	4,667	0	4,667
Acetylene (C2H2)	lb/hr	0	0	0	0	0
Ethylene (C2H4)	lb/hr	52	0	52	0	52
Ethane (C2H6)	lb/hr	11	0	11	0	11
Propane	lb/hr	0	0	0	0	0
Isobutane	lb/hr	0	0	0	0	0
n-Butane	lb/hr	0	0	0	0	0
Pentane	lb/hr	0	0	0	0	0
Benzene (C6H6)	lb/hr	44	0	44	0	44
Tar (C10H8)	lb/hr	2	0	2	0	2
Ammonia (NH3)	lb/hr	32	0	29	1	29
H2S	lb/hr	254	0	253	0	253
Char	lb/hr	4	4	0	-	0
Oxygen	lb/hr	0	0	0	0	0
Argon	lb/hr	0	0	0	0	0
Nitrogen	lb/hr	235	0	235	0	235
Sulfur (solid)	lb/hr	-	-	-	-	-

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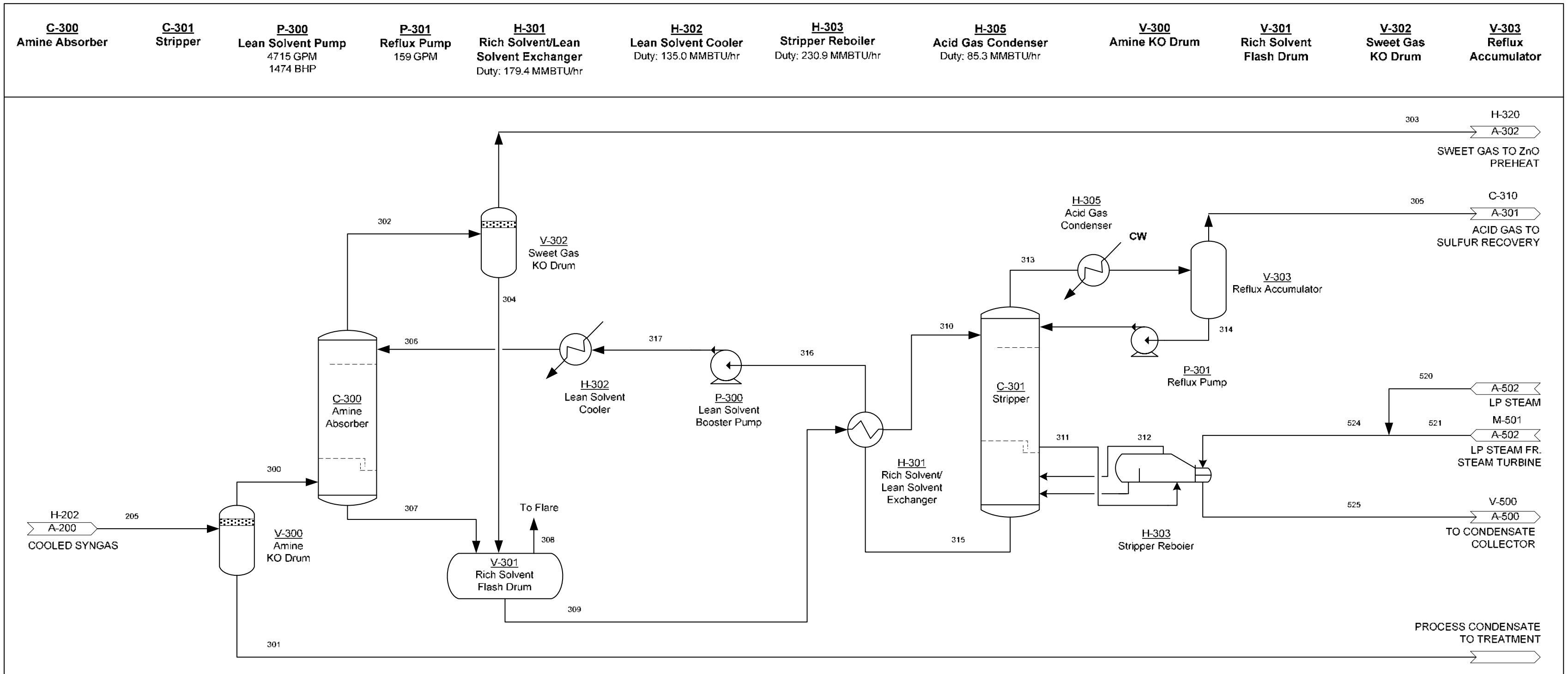
NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-200: SYNGAS SCURBBER AND QUENCH

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov. 2005



	Unit	205	300	301	302	303	305	306	307	308	309	310	311	312	313	314	315	316	317	520	521	524	525
Mass Flow	lb/hr	413,207	272,078	141,129	118,505	118,505	161,485	2,384,401	2,537,954	1,685	2,536,269	2,536,269	2,587,826	213,042	240,048	78,564	2,374,785	2,374,785	2,374,785	113,499	130,439	243,938	243,924
Temperature	F	110	110	110	110	110	130	110	152	152	152	200	245	258	203	130	258	176	177	338	472	410	310
Pressure	psia	429.5	429.5	429.5	422.0	422.0	19.6	432.0	445.0	30.0	30.0	25.0	25.0	30.0	21.4	16.4	30.0	30.0	432.0	100.0	100.0	100.0	90.0
Vapour Fraction		0.65	1.00	0.00	1.00	1.00	1.00	0.00	0.00	1.00	0.00	0.05	0.00	1.00	1.00	0.00	0.00	0.00	0.00	1.00	1.00	1.00	0.00
MW		18.7	19.0	18.1	11.0	11.0	40.4	31.3	31.9	28.9	31.9	31.9	29.7	18.3	28.7	18.0	31.4	31.4	31.4	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-1,739.6		-953.9																-640.4	-727.6	-1,368.1	-1,601.6
Components																							
Hydrogen	lb/hr	14,179	14,178	0	14,149	14,149	1	-	30	29	1	1	-	-	1	0	-	-	-	-	-	-	-
CO2	lb/hr	161,688	160,911	778	8,184	8,184	151,360	1,796	154,523	1,367	153,156	153,156	6,432	4,637	151,419	59	1,796	1,796	1,796	-	-	-	-
CO	lb/hr	90,758	90,755	3	90,592	90,592	5	-	166	161	5	5	-	-	5	0	-	-	-	-	-	-	-
H2O	lb/hr	141,289	970	140,319	581	581	9,868	1,189,367	1,189,733	114	1,189,619	1,189,619	1,387,605	207,854	88,372	78,504	1,179,751	1,179,751	1,179,751	113,499	130,439	243,938	243,924
Methane (CH4)	lb/hr	4,667	4,666	0	4,655	4,655	0	-	11	11	0	0	-	-	0	0	-	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	52	52	0	52	52	0	-	0	0	0	0	-	-	0	0	-	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	11	11	0	11	11	0	-	0	0	0	0	-	-	0	0	-	-	-	-	-	-	-
Propane	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Isobutane	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
n-Butane	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentane	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	44	44	0	40	40	2	-	3	1	2	2	-	-	2	0	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	2	2	0	2	2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	29	4	25	4	4	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
H2S	lb/hr	253	249	4	1	1	248	3	252	1	251	251	3	1	248	0	3	3	3	-	-	-	-
Char	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Oxygen	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Argon	lb/hr	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	lb/hr	235	235	0	235	235	0	-	0	0	0	0	-	-	0	0	-	-	-	-	-	-	-
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MDEA	lb/hr	-	-	-	-	-	-	1,193,235	1,193,235	0	1,193,235	1,193,235	1,193,786	551	0	0	1,193,235	1,193,235	1,193,235	-	-	-	-

No | DATE | REVISIONS

SCALE | DESIGNED | DRAWN

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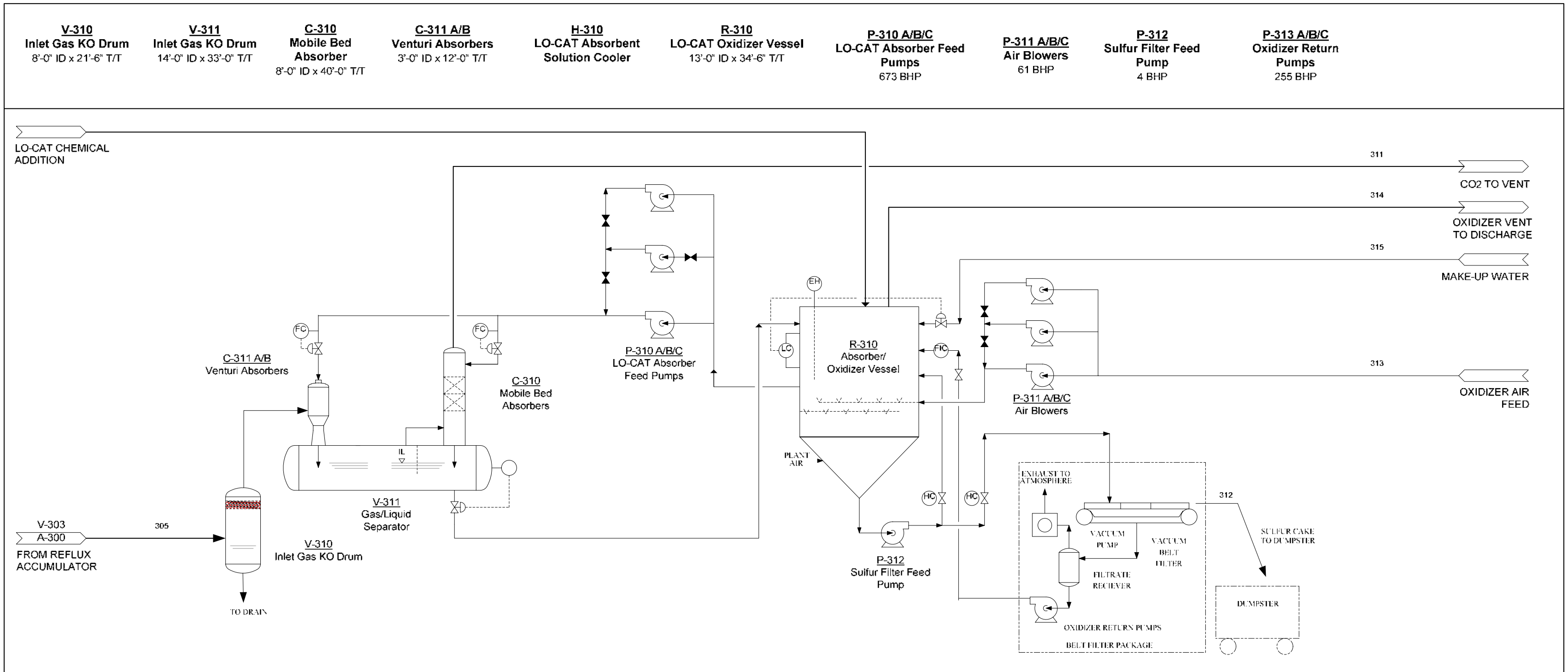
NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-300: AMINE SYSTEM

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov. 2005



	Unit	305	331	332	333	334	335
Mass Flow	lb/hr	160,969	156,507	375	4,923	9,400	381
Temperature	F	130	132	132	70	122	70
Pressure	psia	19.7	18.7	14.7	18.0	14.8	64.7
Vapour Fraction		1.00	1.00	0.00	1.00	1.00	0.00
MW		40.4	42.0	25.5	28.8	33.2	18.0
Heat Flow	MMBtu/hr						
Components							
Hydrogen	lb/hr	1	1	-	-	-	-
CO2	lb/hr	150,873	146,625	3	-	4,301	-
CO	lb/hr	5	5	-	-	-	-
H2O	lb/hr	9,841	9,811	137	-	387	381
Methane (CH4)	lb/hr	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-
Propane	lb/hr	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-
H2S	lb/hr	248	2	-	-	-	-
Char	lb/hr						
Oxygen	lb/hr	-	-	-	1,136	988	-
Argon	lb/hr						
Nitrogen	lb/hr	-	63	-	3,787	3,724	-
Sulfur (solid)	lb/hr	-	-	223	-	-	-
KHCO3	lb/hr	-	-	5	-	-	-
K2S2O3	lb/hr	-	-	7	-	-	-
FE	lb/hr	-	-	0	-	-	-
KOH	lb/hr	-	-	-	-	-	-

Nexant San Francisco

NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-301: SULFUR RECOVERY

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov, 2005

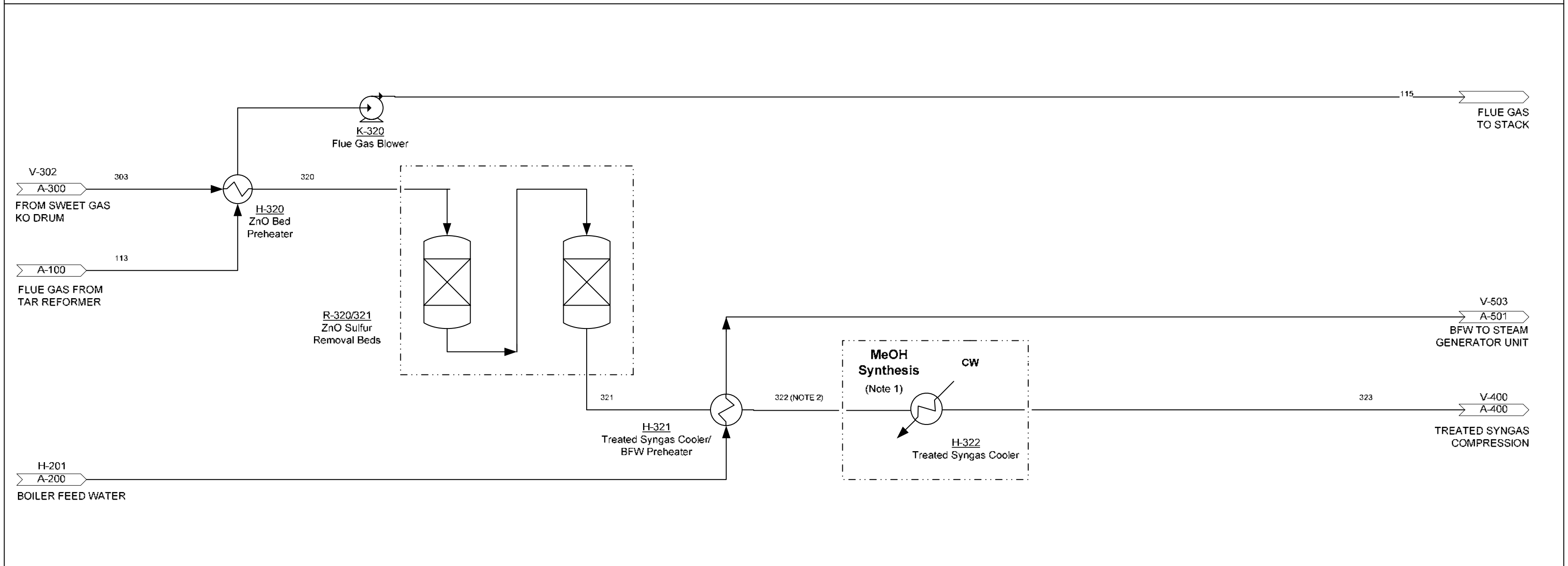
R-320/321
ZnO Sulfur Removal Beds
4'-3" ID x 8' T/T

H-320
ZnO Bed Preheater
Duty: 49.7 MMBTU/hr

H-321
Treated Syngas Cooler/ BFW Preheater
Duty: 29.9 MMBTU/hr

H-322
Treated Syngas Cooler
Duty: 19.9 MMBTU/hr

K-320
Flue Gas Blower
85400 ACFM
207 BHP



	Unit	113	115	303	320	321	323
Mass Flow	lb/hr	279,847	279,847	118,506	118,506	118,505	118,505
Temperature	F	839	221	110	750	750	110
Pressure	psia	14.5	14.7	422.0	417.0	412.0	402.0
Vapour Fraction		1.00	1.00	1.00	1.00	1.00	1.00
MW		27.6	27.6	11.0	11.0	11.0	11.0
Heat Flow	MMBtu/hr	-301.7	-350.8	-195.7	-146.0	-146.0	-195.7
Components							
Hydrogen	lb/hr	0	0	14,149	14,149	14,149	14,149
CO2	lb/hr	39,529	39,529	8,184	8,184	8,184	8,184
CO	lb/hr	0	0	90,592	90,592	90,592	90,592
H2O	lb/hr	36,433	36,433	581	581	581	581
Methane (CH4)	lb/hr	0	0	4,655	4,655	4,655	4,655
Acetylene (C2H2)	lb/hr	0	0	-	-	-	-
Ethylene (C2H4)	lb/hr	0	0	52	52	52	52
Ethane (C2H6)	lb/hr	0	0	11	11	11	11
Propane	lb/hr	0	0	-	-	-	-
Isobutane	lb/hr	0	0	-	-	-	-
n-Butane	lb/hr	0	0	-	-	-	-
Pentane	lb/hr	0	0	-	-	-	-
Benzene (C6H6)	lb/hr	0	0	40	40	40	40
Tar (C10H8)	lb/hr	0	0	2	2	2	2
Ammonia (NH3)	lb/hr	0	0	4	4	4	4
H2S	lb/hr	0	0	1	1	0	0
Char	lb/hr	-	-	-	-	-	-
Oxygen	lb/hr	3,898	3,898	-	-	-	-
Argon	lb/hr	3,349	3,349	-	-	-	-
Nitrogen	lb/hr	196,638	196,638	235	235	235	235
Sulfur (solid)	lb/hr	-	-	-	-	-	-

NOTES:
1. FOR FT SYNTHESIS, EXCHANGER H-322 IS NOT REQUIRED

2. FOR FT SYNTHESIS, STREAM 322 IS COOLED TO 350F PRIOR TO ENTERING THE FT REACTOR

No DATE REVISIONS
SCALE DESIGNED DRAWN

Nexant
San Francisco

NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-302: SULFUR RECOVERY PROCESS

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov. 2005

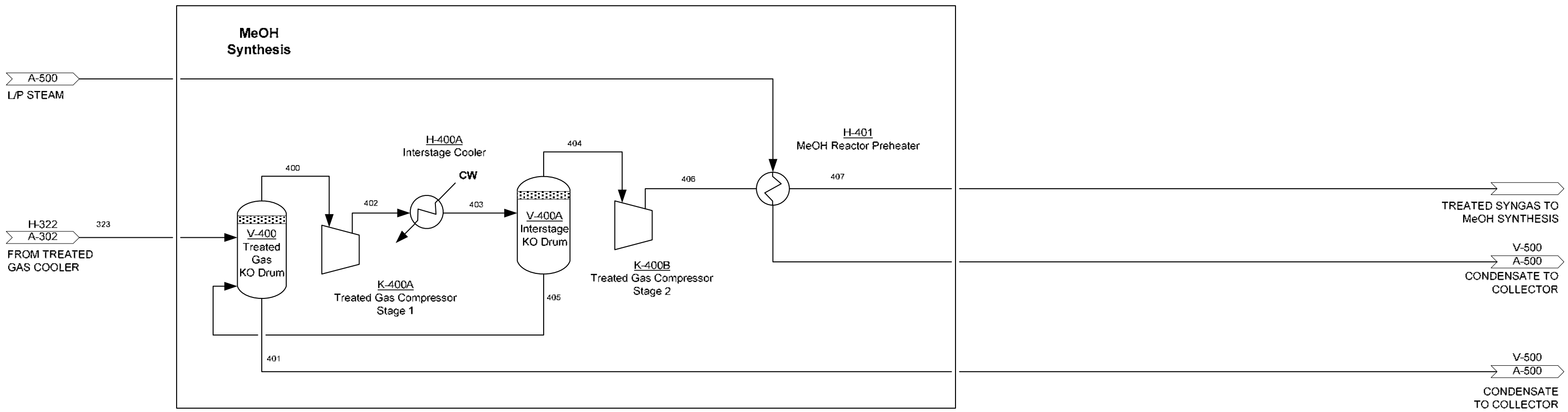
K-400 A/B
Treated Gas Compressor
2746 ACFM
8388 BHP

V-400
Treated Gas KO Drum
5' ID x 15' T/T

V-400A
Interstage KO Drum
5' ID x 15' T/T

H-400A
Interstage Cooler
Duty: 10.5 MMBTU/hr

H-401
MeOH Reactor Preheater
Duty: 17.1 MMBTU/hr



	Unit	323	400	401	403	406	407	518	526
Mass Flow	lb/hr	118,505	118,505	-	118,505	118,505	118,505	17,614	17,614
Temperature	F	110	110	110	200	240	460	472	324
Pressure	psia	402.0	402.0	402.0	995.0	1,165.0	1,160.0	100.0	95.0
Vapour Fraction		1.00	1.00	0.00	1.00	1.00	1.00	1.00	0.00
MW		11.0	11.0	18.0	11.0	11.0	11.0	18.0	18.0
Heat Flow	MMBtu/hr	-195.7	-195.7	0.0	-189.0	-185.9	-168.8	-98.3	-115.4
Components									
Hydrogen	lb/hr	14,149	14,149	-	14,149	14,149	14,149	-	-
CO2	lb/hr	8,184	8,184	-	8,184	8,184	8,184	-	-
CO	lb/hr	90,592	90,592	-	90,592	90,592	90,592	-	-
H2O	lb/hr	581	581	-	581	581	581	17,614	17,614
Methane (CH4)	lb/hr	4,655	4,655	-	4,655	4,655	4,655	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	52	52	-	52	52	52	-	-
Ethane (C2H6)	lb/hr	11	11	-	11	11	11	-	-
Propane	lb/hr	-	-	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	40	40	-	40	40	40	-	-
Tar (C10H8)	lb/hr	2	2	-	2	2	2	-	-
Ammonia (NH3)	lb/hr	4	4	-	4	4	4	-	-
H2S	lb/hr	0	0	-	0	0	0	-	-
Char	lb/hr	-	-	-	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-	-	-	-
Nitrogen	lb/hr	235	235	-	235	235	235	-	-
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-

Nexant
San Francisco

NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-400: MeOH SYNTHESIS COMPRESSION & PREHEAT

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov, 2005

P-500 Condensate Make-up Water Pump 147 GPM 1 BHP

P-501 Deaerator Feed Pump 674 GPM 8 BHP

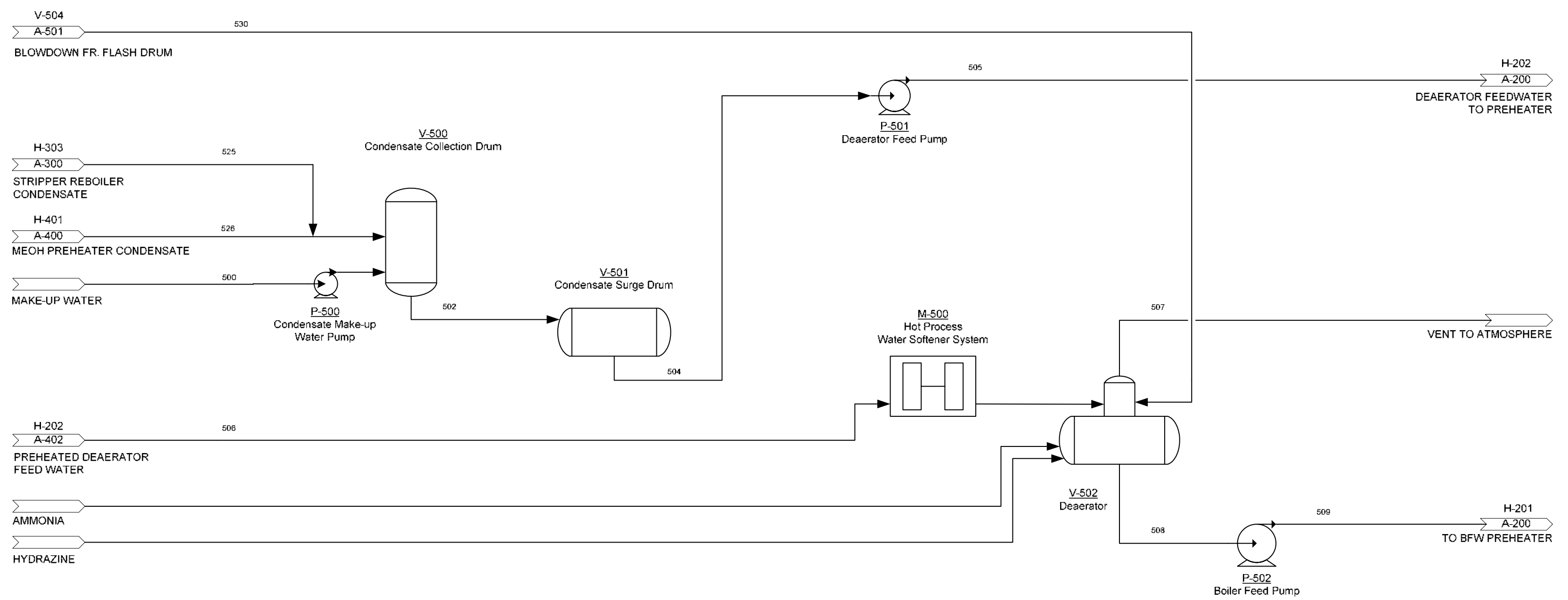
P-502 Boiler Feed Pump 684 GPM 710 BHP

V-500 Condensate Collection Drum 5' ID x 13' T/T

V-501 Condensate Surge Drum 6' ID x 17' T/T

V-502 Deaerator 2' ID x 7' T/T (Top) 6' ID x 17' T/T (Bottom)

M-500 Hot Process Water Softener System



	Unit	500	502	504	505	506	507	508	509	525	526	530
Mass Flow	lb/hr	74,631	319,878	319,878	319,878	319,878	2,180	320,040	320,040	243,924	17,614	2,154
Temperature	F	60	212	212	212	239	239	240	242	310	324	298
Pressure	psia	14.7	14.7	14.7	30.0	25.0	20.0	25.0	1,295.0	90.0	95.0	65.0
Vapour Fraction		0.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00	0.00	1.00
MW		18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-509.6	-2,134.0	-2,134.0	-2,134.0	-2,124.7	-12.4	-2,125.6	-2,123.9	-1,601.6	-115.4	-12.2
Components												
Hydrogen	lb/hr	-	0	0	0	0	0	0	0	-	-	-
CO2	lb/hr	-	0	0	0	0	0	0	0	-	-	-
CO	lb/hr	-	0	0	0	0	0	0	0	-	-	-
H2O	lb/hr	74,631	319,878	319,878	319,878	319,878	2,180	320,040	320,040	243,924	17,614	2,154
Methane (CH4)	lb/hr	-	0	0	0	0	0	0	0	-	-	-
Acetylene (C2H2)	lb/hr	-	0	0	0	0	0	0	0	-	-	-
Ethylene (C2H4)	lb/hr	-	0	0	0	0	0	0	0	-	-	-
Ethane (C2H6)	lb/hr	-	0	0	0	0	0	0	0	-	-	-
Propane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	0	0	0	0	0	0	0	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	-	-	-	-
H2S	lb/hr	-	0	0	0	0	0	0	0	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	lb/hr	-	0	0	0	0	0	0	0	-	-	-
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-

No DATE REVISIONS

SCALE DESIGNED DRAWN

Nexant San Francisco

NREL BIOMASS GASIFICATION HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM A-500: STEAM SYSTEM AND POWER GENERATION

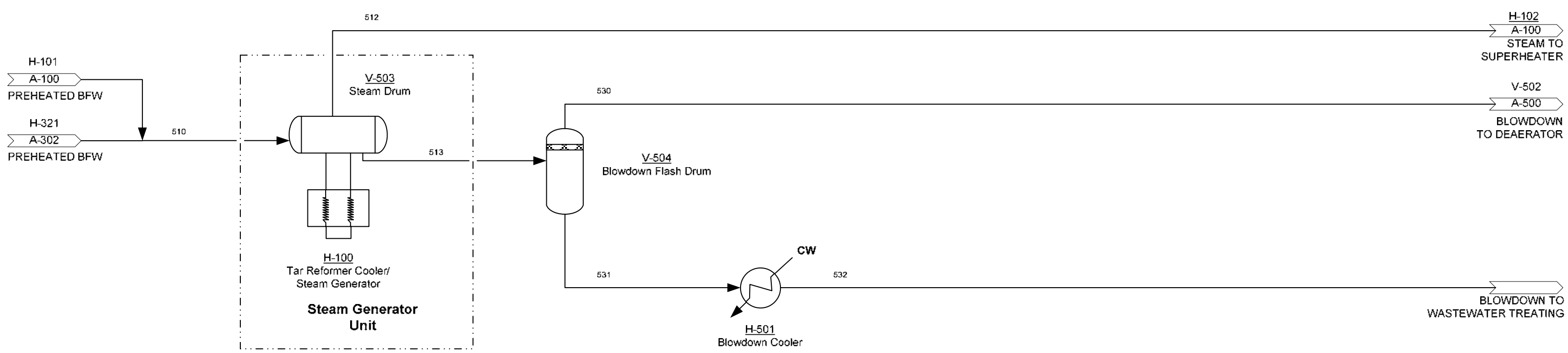
JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov, 2005

V-503
Steam Drum
9' ID x 27' T/T

V-504
Blowdown Flash Drum
2' ID x 6' T/T

H-501
Blowdown Cooler
Duty: 0.8 MMBTU/hr



	Unit	510	512	513	530	531	532
Mass Flow	lb/hr	320,040	313,639	6,401	2,154	4,247	4,247
Temperature	F	555	575	555	298	298	110
Pressure	psia	1,280.0	1,270.0	1,280.0	65.0	65.0	60.0
Vapour Fraction		0.00	1.00	0.00	1.00	0.00	0.00
MW		18.0	18.0	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-2,006.4	-1,762.6	-40.1	-12.2	-27.9	-28.8
Components							
Hydrogen	lb/hr	-	-	-	-	-	-
CO2	lb/hr	-	-	-	-	-	-
CO	lb/hr	-	-	-	-	-	-
H2O	lb/hr	320,040	313,639	6,401	2,154	4,247	4,247
Methane (CH4)	lb/hr	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-
Propane	lb/hr	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-
H2S	lb/hr	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-	-
Nitrogen	lb/hr	-	-	-	-	-	-
Sulfur (solid)	lb/hr	-	-	-	-	-	-

No DATE REVISIONS

SCALE DESIGNED DRAWN

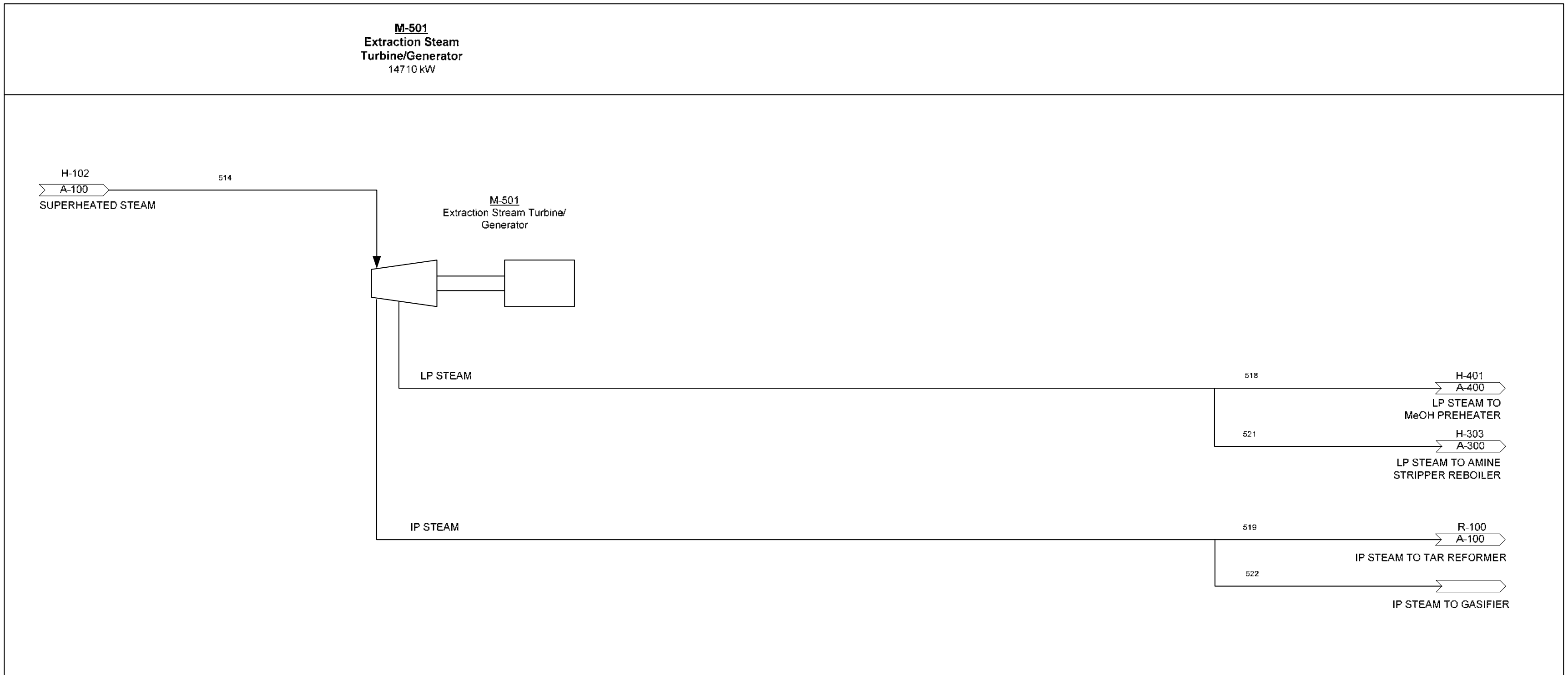
Nexant
San Francisco

NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-501: STEAM SYSTEM AND POWER GENERATION

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov, 2005



	Unit	514	518	519	521	522
Mass Flow	lb/hr	313,639	17,614	26,000	130,439	139,586
Temperature	F	1,000	472	758	472	758
Pressure	psia	1,260.0	100.0	460.0	100.0	460.0
Vapour Fraction		1.00	1.00	1.00	1.00	1.00
MW		18.0	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-1,678.9	-98.3	-141.8	-727.6	-761.4
Components						
Hydrogen	lb/hr	-	-	-	-	-
CO2	lb/hr	-	-	-	-	-
CO	lb/hr	-	-	-	-	-
H2O	lb/hr	313,639	17,614	26,000	130,439	139,586
Methane (CH4)	lb/hr	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-
Propane	lb/hr	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-
H2S	lb/hr	-	-	-	-	-
Char	lb/hr	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-
Nitrogen	lb/hr	-	-	-	-	-
Sulfur (solid)	lb/hr	-	-	-	-	-

519

Nexant
San Francisco

NREL BIOMASS GASIFICATION
HIGH-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-502: STEAM SYSTEM AND POWER GENERATION

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 2

15 Nov. 2005

Appendix B Low-Pressure Syngas Design Case PFDs

S-100 A/B/C/D Gasifier Cyclones 7' ID x 35' T/T

S-102 Tar Reformer Cyclone 5' ID x 25' T/T

S-103 Catalyst Regenerator Cyclone 4' ID x 17' T/T

R-100 Tar Reformer 29' ID x 89' T/T

R-101 Catalyst Regenerator 23' ID x 72' T/T

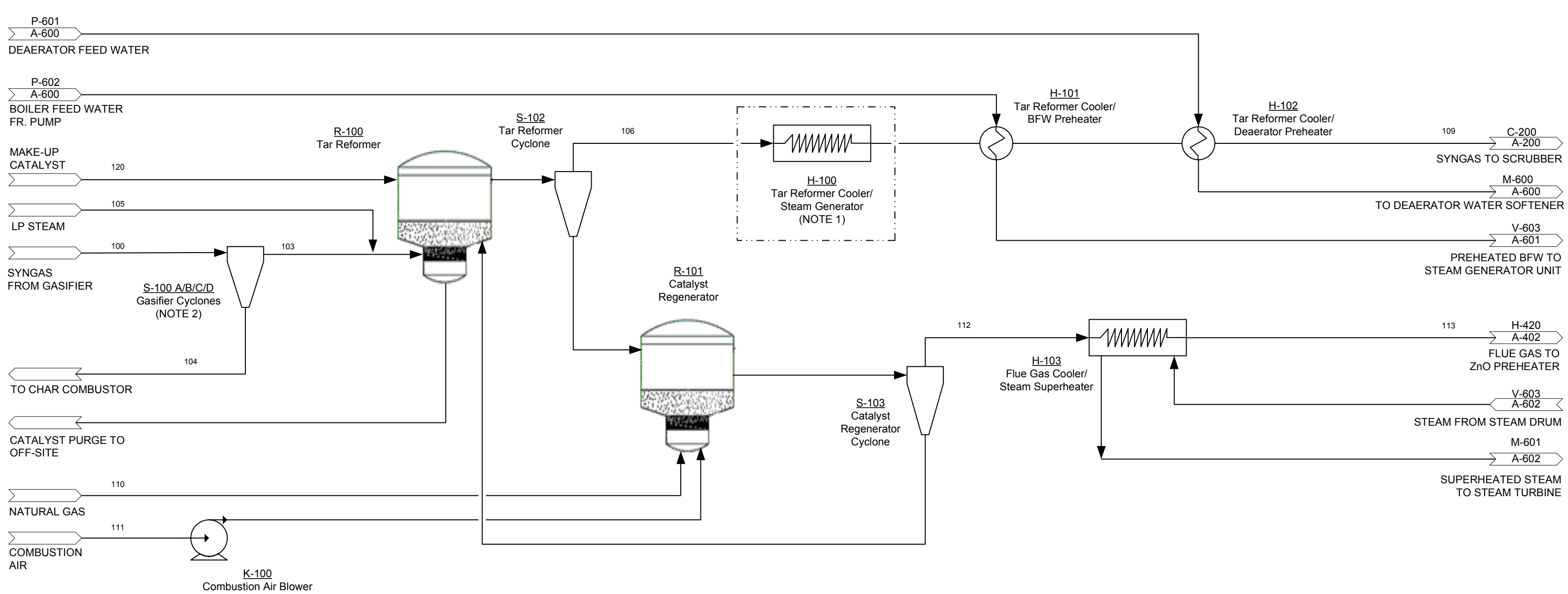
H-100 Tar Reformer Cooler/ Steam Generator Duty: 167 MMBTU/hr

H-101 Tar Reformer Cooler/ BFW Preheater Duty: 50.6 MMBTU/hr

H-102 Tar Reformer Cooler/ Deaerator Preheater Duty: 11.3 MMBTU/hr

H-103 Flue GasCooler/ Steam Superheater Duty: 67.3 MMBTU/hr

K-100 Combustion Air Blower 54910 ACFM 3 BHP



	Unit	100	103	104	105	106	109	110	111	112	113
Mass Flow	lb/hr	316,369	275,978	40,391	53,000	328,979	328,979	13,190	235,200	248,388	248,388
Temperature	F	1,598	1,598	1,598	376	1,598	225	60	90	1,798	935
Pressure	psia	33.0	32.9	32.9	50.0	29.9	18.2	14.7	14.7	14.7	14.5
Vapour Fraction		0.80	1.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
MW		18.7	20.3	12.0	18.0	16.7	16.7	17.1	28.6	27.6	27.6
Heat Flow	MMBtu/hr	-722.8	-784.2	61.4	-297.9	-936.3	-1,165.3	-25.5	-26.1	-197.5	-264.8
Components											
Hydrogen	lb/hr	3,538	3,538	-	-	15,262	15,262	-	-	0	0
CO2	lb/hr	41,379	41,379	-	-	103,295	103,295	169	116	35,588	35,588
CO	lb/hr	86,796	86,796	-	-	97,976	97,976	-	-	0	0
H2O	lb/hr	112,176	112,176	-	53,000	107,295	107,295	-	4,589	32,732	32,732
Methane (CH4)	lb/hr	18,106	18,106	-	-	3,621	3,621	11,700	-	0	0
Acetylene (C2H2)	lb/hr	777	777	-	-	78	78	-	-	0	0
Ethylene (C2H4)	lb/hr	8,944	8,944	-	-	894	894	-	-	0	0
Ethane (C2H6)	lb/hr	653	653	-	-	7	7	720	-	0	0
Propane	lb/hr	-	-	-	-	0	0	165	-	0	0
Isobutane	lb/hr	-	-	-	-	0	0	45	-	0	0
n-Butane	lb/hr	-	-	-	-	0	0	45	-	0	0
Pentane	lb/hr	-	-	-	-	0	0	110	-	0	0
Benzene (C6H6)	lb/hr	742	742	-	-	7	7	-	-	0	0
Tar (C10H8)	lb/hr	2,260	2,260	-	-	2	2	-	-	0	0
Ammonia (NH3)	lb/hr	416	416	-	-	42	42	-	-	0	0
H2S	lb/hr	175	175	-	-	175	175	-	-	0	0
Char	lb/hr	40,407	16	40,391	-	16	16	-	-	-	-
Oxygen	lb/hr	-	-	-	-	0	0	-	53,363	2,699	2,699
Argon	lb/hr	-	-	-	-	0	0	-	2,970	2,970	2,970
Nitrogen	lb/hr	-	-	-	-	308	308	237	174,161	174,398	174,398
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-	-	-

NOTES:
1. SEE SECTION A-601 FOR DETAIL OF STEAM GENERATOR UNIT
2. FOUR UNITS OPERATE IN PARALLEL

No.	DATE	REVISIONS	BY	CHKD	DESIGN SUPV	ENGR.	PROJ. ENGR.	APPRV.

SCALE: DESIGNED: DRAWN:

Nexant
San Francisco

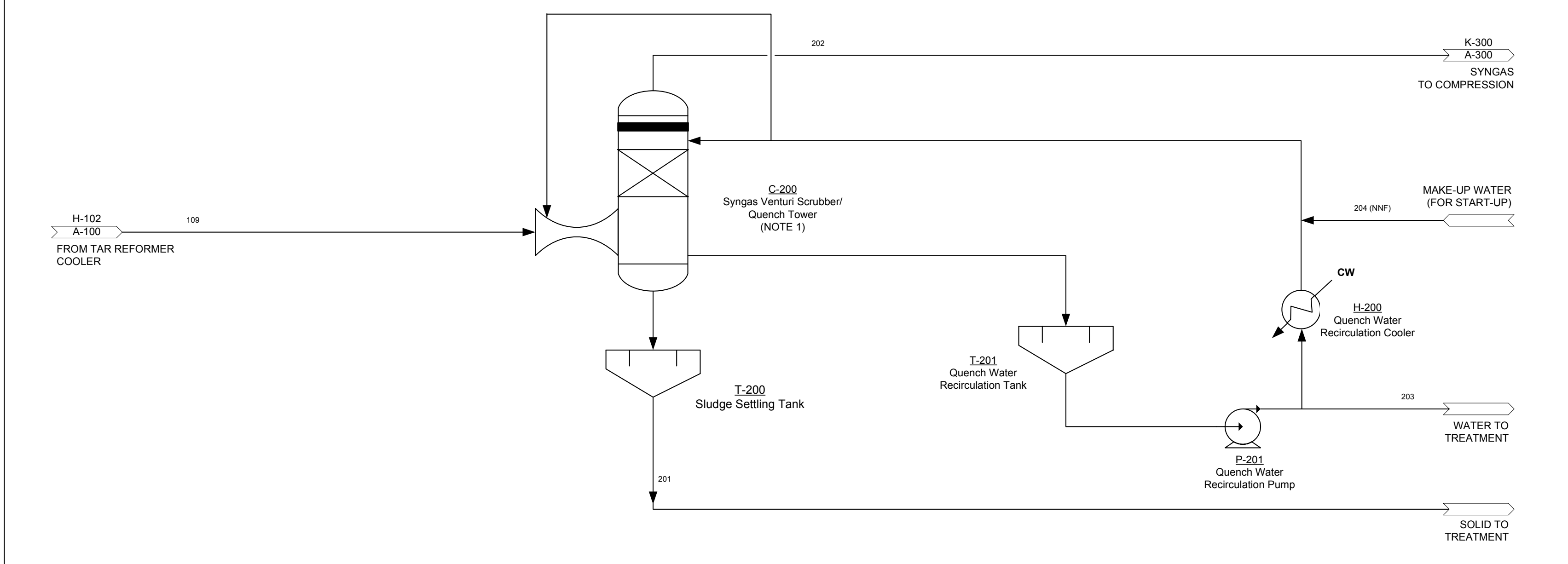
NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-100: TAR REFORMING

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005

C-200	T-200	T-201	H-200	P-201
Syngas Venturi Scrubber/ Quench Tower	**Sludge Settling Tank**	**Quench Water Recirculation Tank**	**Quench Water Recirculation Cooler**	**Quench Water Recirculation Pump**
14' ID x 29' T/T (Quench Tower)	1' ID x 3' T/T	9' ID x 27' T/T	Duty: 21.8 MMBtu/hr	844 GPM 7 BHP



	Unit	109	201	202	203
Mass Flow	lb/hr	328,979	516	317,371	11,092
Temperature	F	225	163	157	163
Pressure	psia	18.2	15.9	15.9	26.0
Vapour Fraction		1.00	0.00	1.00	0.00
MW		16.7	17.7	16.7	18.0
Heat Flow	MMBtu/hr	-1,165.3	-3.4	-1,109.3	-74.6
Components					
Hydrogen	lb/hr	15,262	0	15,262	0
CO2	lb/hr	103,295	0	103,268	1
CO	lb/hr	97,976	0	97,976	0
H2O	lb/hr	107,295	500	95,740	11,091
Methane (CH4)	lb/hr	3,621	0	3,621	0
Acetylene (C2H2)	lb/hr	78	0	78	0
Ethylene (C2H4)	lb/hr	894	0	894	0
Ethane (C2H6)	lb/hr	7	0	7	0
Propane	lb/hr	0	0	0	0
Isobutane	lb/hr	0	0	0	0
n-Butane	lb/hr	0	0	0	0
Pentane	lb/hr	0	0	0	0
Benzene (C6H6)	lb/hr	7	0	7	0
Tar (C10H8)	lb/hr	2	0	2	0
Ammonia (NH3)	lb/hr	42	0	33	0
H2S	lb/hr	175	0	175	0
Char	lb/hr	16	16	0	-
Oxygen	lb/hr	0	0	0	0
Argon	lb/hr	0	0	0	0
Nitrogen	lb/hr	308	0	308	0
Sulfur (solid)	lb/hr	-	-	-	-

NOTES:
1. TWO UNITS OPERATE IN PARALLEL

No.	DATE	REVISIONS	BY	CHKD	DESIGN SUPV	ENGR.	PROJ. ENGR.	APPRV

SCALE: DESIGNED: DRAWN:

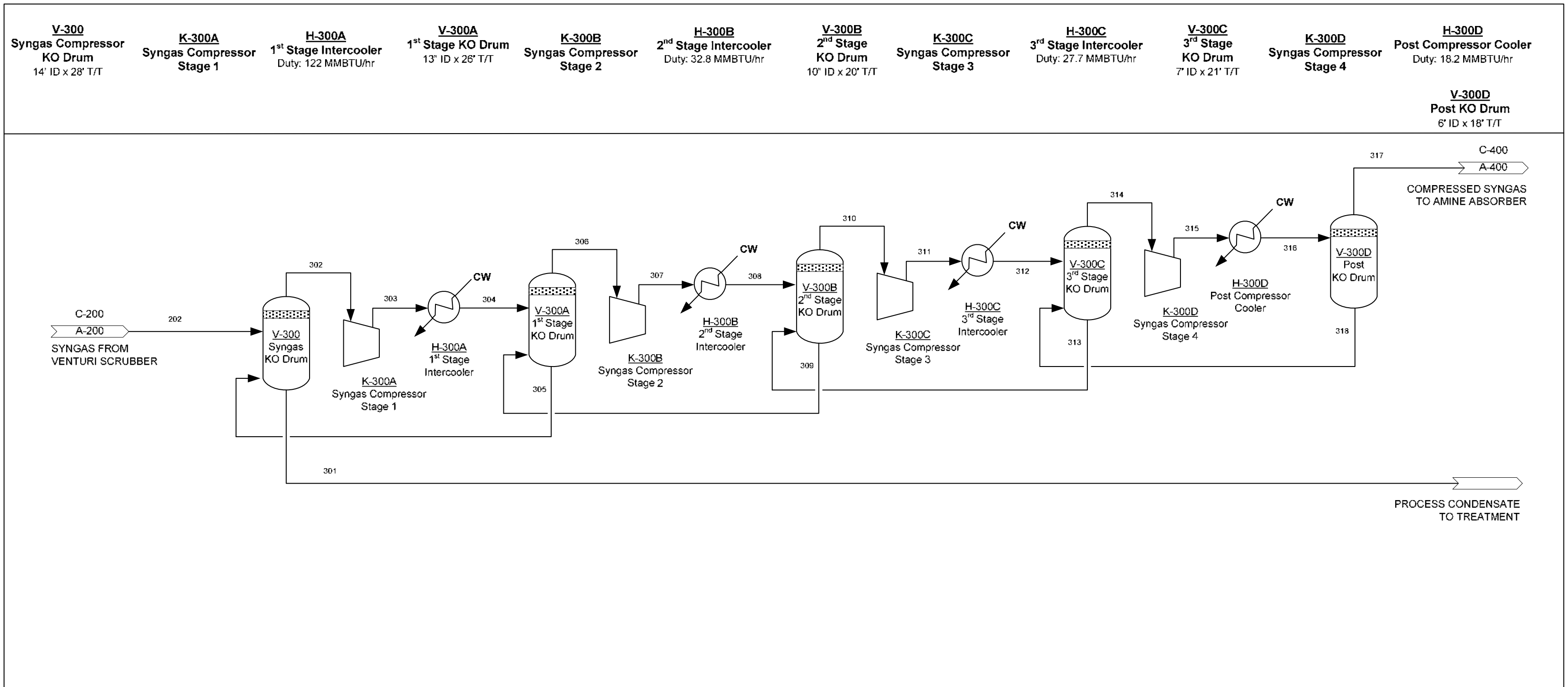
Nexant
San Francisco

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-200: SYNGAS SCURBBER AND QUENCH

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005



	Unit	301	302	304	306	308	310	312	314	316	317
Mass Flow	lb/hr	94,902	317,371	317,371	232,617	232,617	225,773	225,773	223,220	223,220	222,469
Temperature	F	110	157	110	110	110	110	110	110	110	110
Pressure	psia	15.9	15.9	30.0	30.0	79.0	79.0	215.0	215.0	445.0	445.0
Vapour Fraction		0.00	1.00	0.75	1.00	0.97	1.00	0.99	1.00	1.00	1.00
MW		18.0	16.7	16.7	16.3	16.3	16.2	16.2	16.2	16.2	16.2
Heat Flow	MMBtu/hr	-643.0	-1,109.3	-1,204.1	-629.8	-637.0	-590.7	-593.6	-576.3	-577.5	-572.4
Components											
Hydrogen	lb/hr	0	15,262	15,262	15,262	15,262	15,262	15,262	15,262	15,262	15,262
CO2	lb/hr	36	103,268	103,268	103,245	103,245	103,240	103,240	103,235	103,235	103,232
CO	lb/hr	0	97,976	97,976	97,976	97,976	97,976	97,976	97,976	97,976	97,976
H2O	lb/hr	94,854	95,740	95,740	11,017	11,017	4,180	4,180	1,633	1,633	886
Methane (CH4)	lb/hr	0	3,621	3,621	3,621	3,621	3,621	3,621	3,621	3,621	3,621
Acetylene (C2H2)	lb/hr	0	78	78	78	78	78	78	78	78	78
Ethylene (C2H4)	lb/hr	0	894	894	894	894	894	894	894	894	894
Ethane (C2H6)	lb/hr	0	7	7	7	7	7	7	7	7	7
Propane	lb/hr	-	0	0	-	-	-	-	-	-	-
Isobutane	lb/hr	-	0	0	-	-	-	-	-	-	-
n-Butane	lb/hr	-	0	0	-	-	-	-	-	-	-
Pentane	lb/hr	-	0	0	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	0	7	7	7	7	7	7	7	7	7
Tar (C10H8)	lb/hr	0	2	2	2	2	2	2	2	2	2
Ammonia (NH3)	lb/hr	12	33	33	26	26	24	24	23	23	22
H2S	lb/hr	0	175	175	175	175	175	175	175	175	175
Char	lb/hr	-	-	-	-	-	-	-	-	-	-
Oxygen	lb/hr	-	0	0	-	-	-	-	-	-	-
Argon	lb/hr	-	0	0	-	-	-	-	-	-	-
Nitrogen	lb/hr	0	308	308	308	308	308	308	308	308	308
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-	-	-

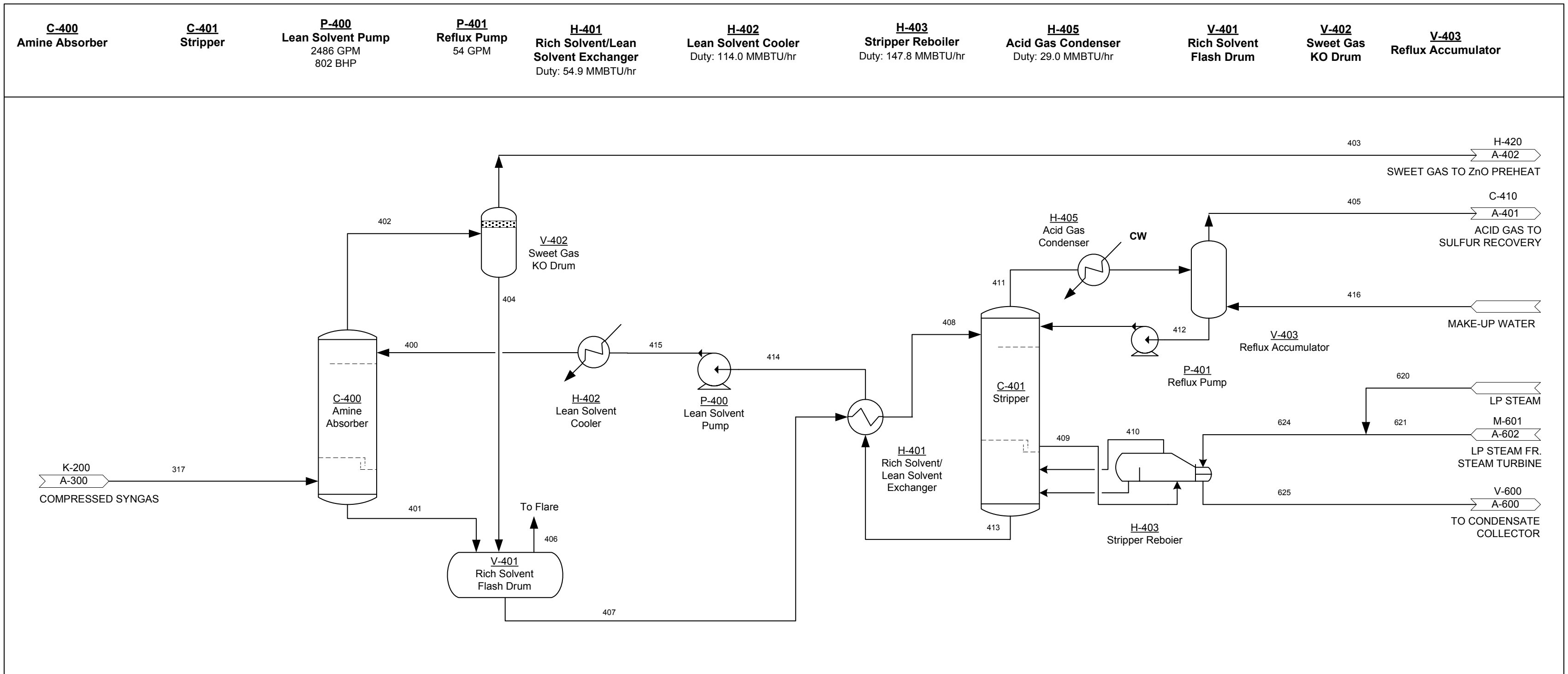
Nexant San Francisco

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-300: SYNGAS COMPRESSION

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

13 Dec, 2005



	Unit	317	400	401	402	403	405	406	407	408	409	410	411	412	413	414	415	620	621	624	625
Mass Flow	lb/hr	222,470	1,244,605	1,339,954	127,107	127,107	94,477	6,788	1,333,166	1,333,166	1,379,976	140,473	120,942	26,465	1,239,503	1,239,503	1,239,503	44,137	106,459	150,596	150,596
Temperature	F	110	110	160	115	115	130	157	157	200	241	253	193	130	253	209	209	292	376	352	267
Pressure	psia	445.0	445.0	445.0	440.4	440.4	19.6	50.0	50.0	45.0	25.0	30.0	21.6	19.6	30.0	30.0	445.0	50.0	50.0	50.0	40.0
Vapour Fraction		1.00	0.00	0.00	1.00	1.00	1.00	1.00	0.00	0.00	0.00	1.00	1.00	0.00	0.00	0.00	0.00	1.00	1.00	1.00	0.00
MW		16.2	24.2	24.9	11.0	11.0	40.5	37.3	24.9	24.9	23.5	18.6	31.8	18.0	24.2	24.2	24.2	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-572.4			-206.2	-206.2												-249.8	-598.3	-848.1	-995.8
Components																					
Hydrogen	lb/hr	15,262	-	22	15,240	15,240	-	22	0	0	-	-	-	-	-	-	-	-	0	0	0
Carbon Dioxide	lb/hr	103,233	11,519	106,499	8,260	8,260	88,628	6,356	100,143	100,143	18,366	6,847	88,663	35	11,519	11,519	11,519	-	0	0	0
Carbon Monoxide	lb/hr	97,976	-	122	97,854	97,854	-	121	1	1	-	-	-	-	-	-	-	-	0	0	0
Water	lb/hr	886	859,599	859,638	825	825	5,681	268	859,370	859,370	987,990	133,553	32,111	26,430	854,437	854,437	854,437	44,137	106,459	150,596	150,596
Methane	lb/hr	3,621	-	6	3,615	3,615	-	6	0	0	-	-	-	-	-	-	-	-	0	0	0
Acetylene	lb/hr	78	-	-	78	78	-	-	-	-	-	-	-	-	-	-	-	-	0	0	0
Ethylene	lb/hr	894	-	4	890	890	-	4	0	0	-	-	-	-	-	-	-	-	0	0	0
Ethane	lb/hr	7	-	0	7	7	-	0	0	0	-	-	-	-	-	-	-	-	0	0	0
Propane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Benzene	lb/hr	7	-	0	7	7	-	0	0	0	-	-	-	-	-	-	-	-	0	0	0
Tar	lb/hr	2	-	-	2	2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia	lb/hr	22	-	-	22	22	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrogen Sulfide	lb/hr	175	0	175	0	0	168	11	164	164	0	0	168	0	0	0	0	-	0	0	0
Char	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Argon	lb/hr																	-	-	-	-
Nitrogen	lb/hr	308	-	0	308	308	-	0	0	0	-	-	-	-	-	-	-	-	0	0	0
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DEA	lb/hr	-	373,487	373,487	0	0	-	0	373,487	373,487	373,620	73	0	0	373,547	373,547	373,547	-	-	-	-

No.	DATE	REVISIONS	BY	CHKD	DESIGN SUPV	ENGR.	PROJ. ENGR.	APPRV.

SCALE: DESIGNED: DRAWN:

Nexant

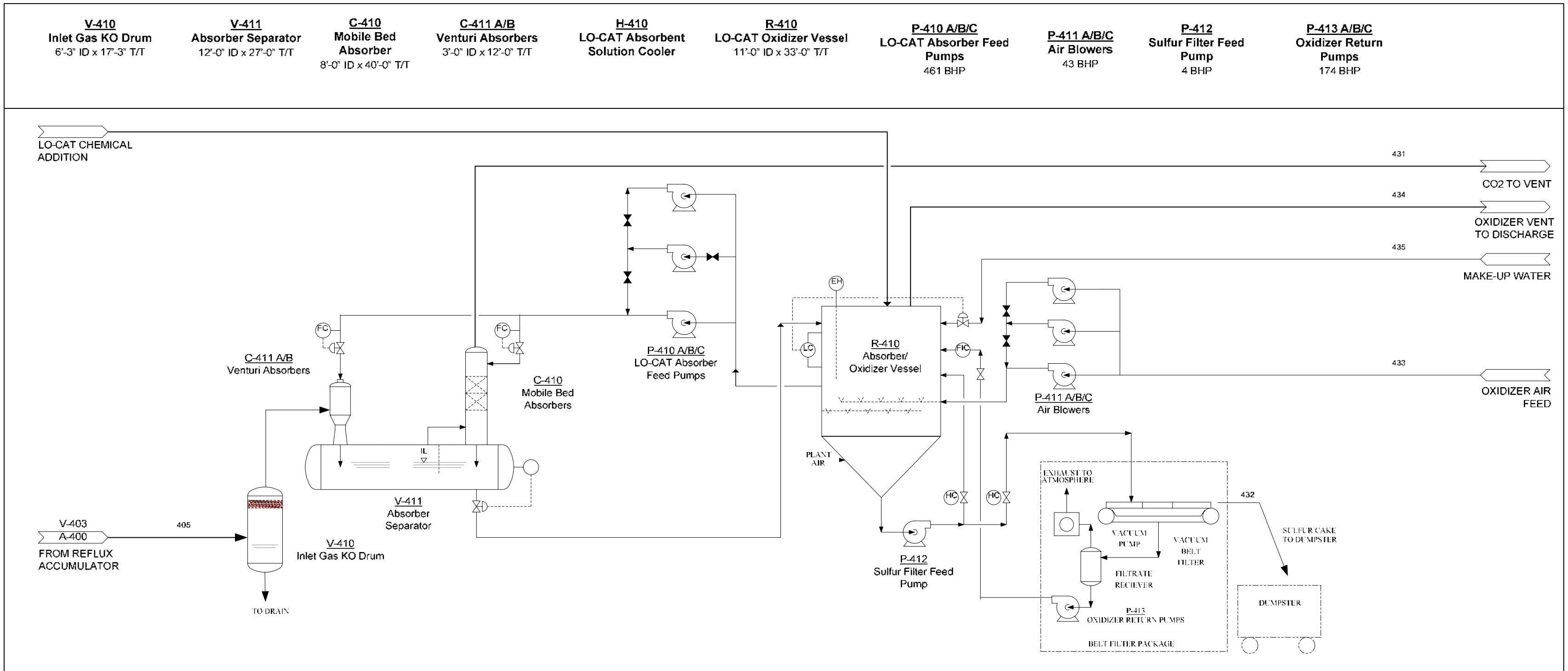
San Francisco

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-400: AMINE SYSTEM

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005



	Unit	405	431	432	433	434	435
Mass Flow	lb/hr	94,477	92,150	262	3,479	6,574	270
Temperature	F	130	132	132	70	132	70
Pressure	psia	19.7	18.7	14.7	17.9	14.8	64.7
Vapour Fraction		1.00	1.00	0.00	1.00	1.00	0.00
MW		40.5	40.4	92.9	28.8	33.1	18.0
Heat Flow	MMBtu/hr						
Components							
Hydrogen	lb/hr	-	-	-	-	-	-
CO2	lb/hr	88,628	86,444		-	2,969	-
CO	lb/hr	-	-	-	-	-	-
H2O	lb/hr	5,681	5,661	97	-	274	270
Methane (CH4)	lb/hr	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-
Propane	lb/hr	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-
H2S	lb/hr	168	2	-	-	-	-
Char	lb/hr						
Oxygen	lb/hr	-	-	-	803	699	-
Argon	lb/hr						
Nitrogen	lb/hr	-	43	-	2,676	2,633	-
Sulfur (solid)	lb/hr	-	-	157	-	-	-
KHCO3	lb/hr	-	-	4	-	-	-
K2S2O3	lb/hr	-	-	5	-	-	-
FE	lb/hr	-	-	0	-	-	-
KOH	lb/hr	-	-	-	-	-	-

No.	DATE	REVISIONS			
SCALE	DESIGNED	DRAWN			

Nexant
San Francisco

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-401: SULFUR RECOVERY

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005

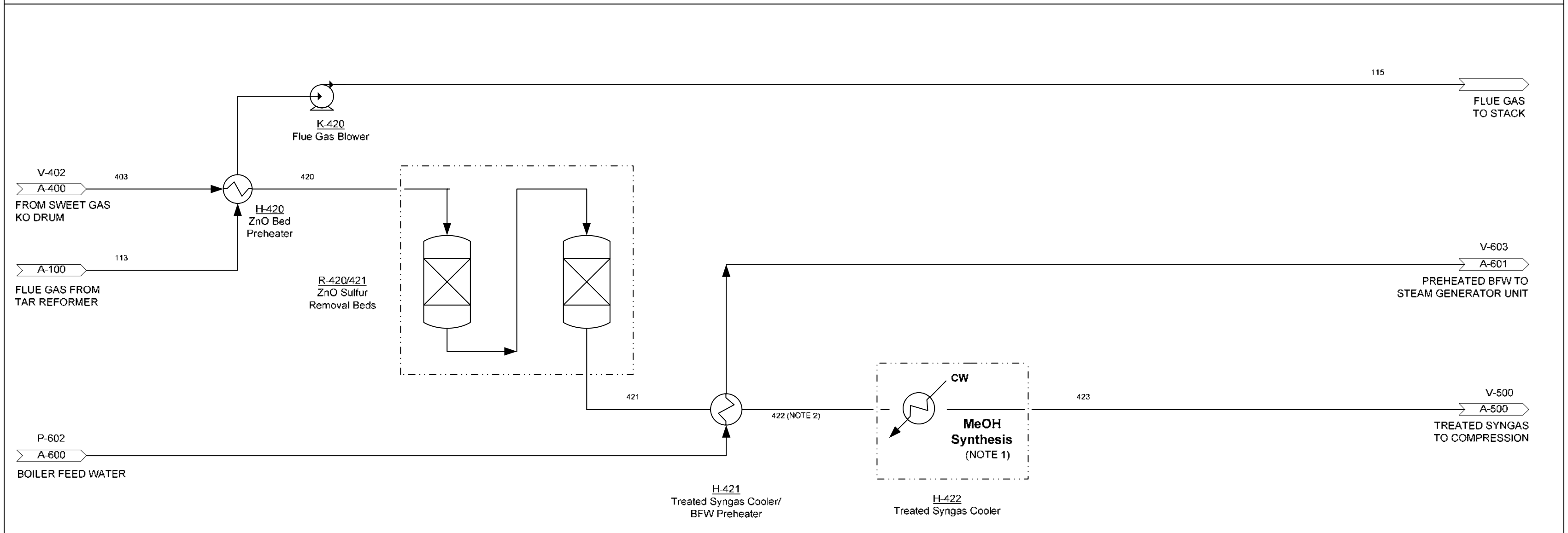
R-420/421 ZnO Sulfur Removal Beds
4'-6" ID x 8' T/T

H-420 ZnO Bed Preheater
Duty: 52.9 MMBTU/hr

H-421 Treated Syngas Cooler/ BFW Preheater
Duty: 40.6 MMBTU/hr

H-422 Treated Syngas Cooler
Duty: 11.9 MMBTU/hr

K-420 Flue Gas Blower
73100 ACFM
177 BHP



	Unit	113	115	403	420	421	423
Mass Flow	lb/hr	248,388	248,388	127,099	127,099	127,098	127,098
Temperature	F	935	196	115	750	750	110
Pressure	psia	14.5	14.7	440.4	435.4	425.0	415.0
Vapour Fraction		1.00	1.00	1.00	1.00	1.00	1.00
MW		27.6	27.6	11.0	11.0	11.0	11.0
Heat Flow	MMBtu/hr	-264.8	-317.2	-206.1	-153.2	-153.2	-206.5
Components							
Hydrogen	lb/hr	0	0	15,240	15,240	15,240	15,240
CO2	lb/hr	35,588	35,588	8,260	8,260	8,260	8,260
CO	lb/hr	0	0	97,853	97,853	97,853	97,853
H2O	lb/hr	32,732	32,732	817	817	817	817
Methane (CH4)	lb/hr	0	0	3,615	3,615	3,615	3,615
Acetylene (C2H2)	lb/hr	0	0	78	78	78	78
Ethylene (C2H4)	lb/hr	0	0	890	890	890	890
Ethane (C2H6)	lb/hr	0	0	7	7	7	7
Propane	lb/hr	0	0	-	-	-	-
Isobutane	lb/hr	0	0	-	-	-	-
n-Butane	lb/hr	0	0	-	-	-	-
Pentane	lb/hr	0	0	-	-	-	-
Benzene (C6H6)	lb/hr	0	0	7	7	7	7
Tar (C10H8)	lb/hr	0	0	2	2	2	2
Ammonia (NH3)	lb/hr	0	0	22	22	22	22
H2S	lb/hr	0	0	1	1	0	0
Char	lb/hr	-	-	-	-	-	-
Oxygen	lb/hr	2,699	2,699	-	-	-	-
Argon	lb/hr	2,970	2,970	-	-	-	-
Nitrogen	lb/hr	174,398	174,398	308	308	308	308
Sulfur (solid)	lb/hr	-	-	-	-	-	-

NOTES:
1. FOR FT SYNTHESIS, EXCHANGER H-422 IS NOT REQUIRED

2. FOR FT SYNTHESIS, STREAM 422 IS COOLED TO 350F PRIOR TO ENTERING THE FT REACTOR

No	DATE	REVISIONS

SCALE | DESIGNED | DRAWN

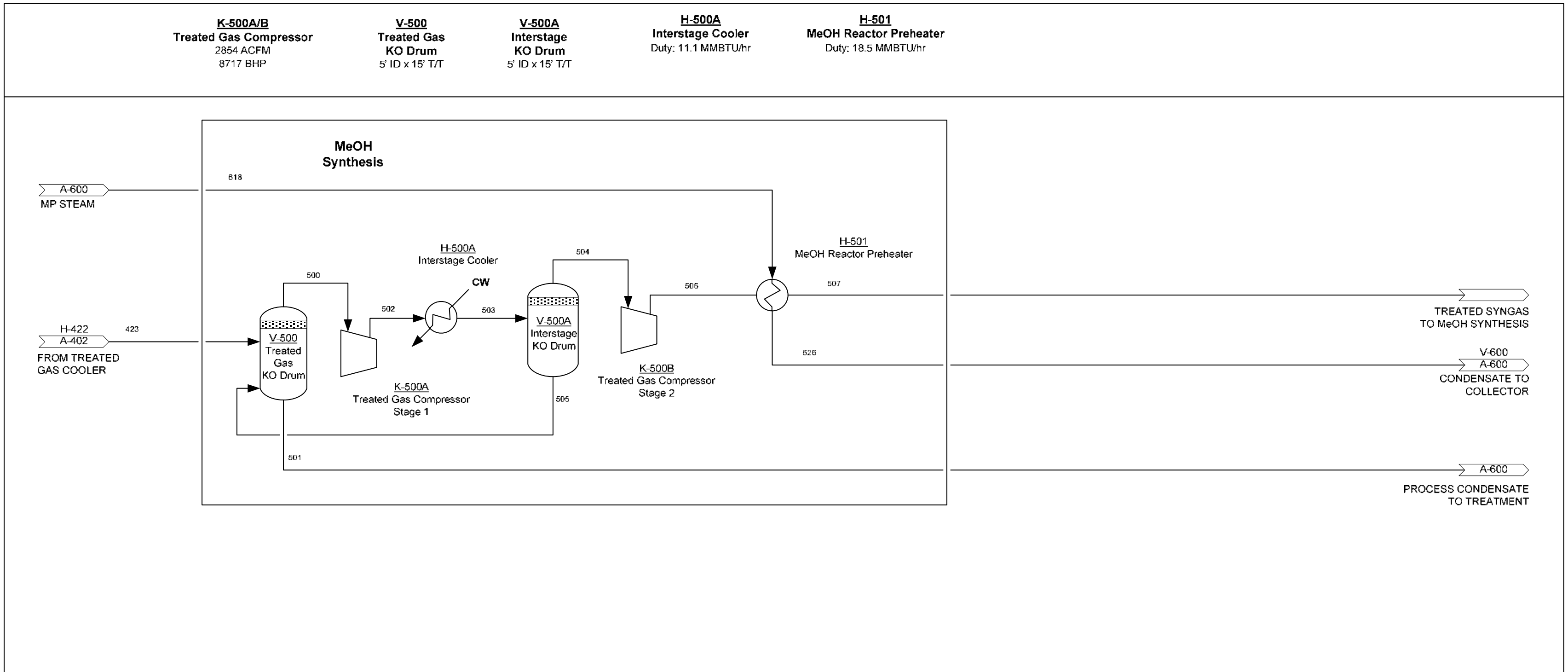
Nexant
San Francisco

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-402: SULFUR RECOVERY PROCESS

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005



	Unit	423	500	501	503	506	507	618	626
Mass Flow	lb/hr	127,098	127,035	63	127,035	127,035	127,035	18,832	18,832
Temperature	F	110	110	110	200	239	460	487	324
Pressure	psia	415.0	415.0	415.0	995.0	1,160.0	1,155.0	100.0	95.0
Vapour Fraction		1.00	1.00	0.00	1.00	1.00	1.00	1.00	0.00
MW		11.0	11.0	18.0	11.0	11.0	11.0	18.0	18.0
Heat Flow	MMBtu/hr	-206.5	-206.1	-0.4	-198.9	-195.7	-177.3	-104.9	-123.4
Components									
Hydrogen	lb/hr	15,240	15,240	0	15,240	15,240	15,240	0	0
CO2	lb/hr	8,260	8,260	0	8,260	8,260	8,260	0	0
CO	lb/hr	97,853	97,853	0	97,853	97,853	97,853	0	0
H2O	lb/hr	817	754	63	754	754	754	18,832	18,832
Methane (CH4)	lb/hr	3,615	3,615	0	3,615	3,615	3,615	0	0
Acetylene (C2H2)	lb/hr	78	78	0	78	78	78	0	0
Ethylene (C2H4)	lb/hr	890	890	0	890	890	890	0	0
Ethane (C2H6)	lb/hr	7	7	0	7	7	7	0	0
Propane	lb/hr	-	-	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	7	7	0	7	7	7	0	0
Tar (C10H8)	lb/hr	2	2	0	2	2	2	-	-
Ammonia (NH3)	lb/hr	22	22	0	22	22	22	-	-
H2S	lb/hr	0	0	0	0	0	0	0	0
Char	lb/hr	-	-	-	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-	-	-	-
Nitrogen	lb/hr	308	308	0	308	308	308	0	0
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-

Nexant
San Francisco

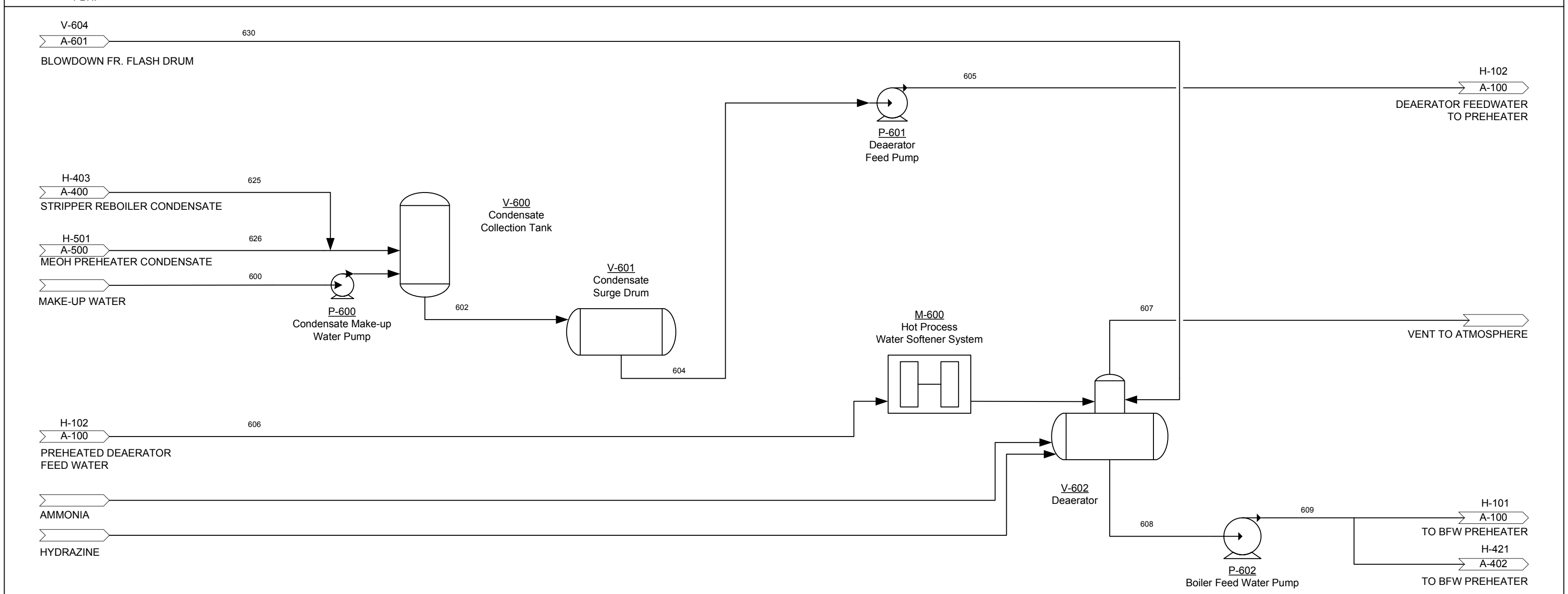
NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-500: MeOH SYNTHESIS COMPRESSION & PREHEAT

JOB No. NREL Contract ACO-5-44027 | DRAWING No. | REV 1

15 Nov, 2005

P-600 Condensate Make-up Water Pump, 173 GPM, 1 BHP
P-601 Deaerator Feed Pump, 537 GPM, 7 BHP
P-602 Boiler Feed Pump, 548 GPM, 570 BHP
V-600 Condensate Collection Tank, 5' ID x 12' T/T
V-601 Condensate Surge Drum, 5'-6" ID x 16'-6" T/T
V-602 Deaerator, 2' ID x 7' T/T (Top), 5' - 6" ID x 16' - 6" T/T (Bottom)
M-600 Hot Process Water Softener System



	Unit	600	602	604	605	606	607	608	609	625	626	630
Mass Flow	lb/hr	87,404	256,818	256,818	256,818	256,818	1,579	256,896	256,896	150,588	18,826	1,648
Temperature	F	60	195	195	195	237	237	238	240	257	315	298
Pressure	psia	14.7	14.7	14.7	30.0	25.0	20.0	24.0	1,295.0	40.0	95.0	65.0
Vapour Fraction		0.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00	0.00	1.00
MW		18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-596.9	-1,717.8	-1,717.8	-1,717.8	-1,706.4	-9.0	-1,706.8	-1,705.5	-997.4	-123.5	-9.3
Components												
Hydrogen	lb/hr	-	0	0	0	0	0	0	0	-	0	0
CO2	lb/hr	-	0	0	0	0	0	0	0	-	0	0
CO	lb/hr	-	0	0	0	0	0	0	0	-	0	0
H2O	lb/hr	87,404	256,818	256,818	256,818	256,818	1,579	256,896	256,896	150,588	18,826	1,648
Methane (CH4)	lb/hr	-	0	0	0	0	0	0	0	-	0	0
Acetylene (C2H2)	lb/hr	-	0	0	0	0	0	0	0	-	0	0
Ethylene (C2H4)	lb/hr	-	0	0	0	0	0	0	0	-	0	0
Ethane (C2H6)	lb/hr	-	0	0	0	0	0	0	0	-	0	0
Propane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	0	0	0	0	0	0	0	-	0	0
Tar (C10H8)	lb/hr	-	0	0	0	0	-	-	-	-	0	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	-	-	-	-
H2S	lb/hr	-	0	0	0	0	0	0	0	-	0	0
Char	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	lb/hr	-	0	0	0	0	0	0	0	-	0	0
Sulfur (solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-

No.	DATE	REVISIONS	BY	CHKD	DESIGN SUPV	ENGR	PROJ. ENGR	APPR

SCALE: DESIGNED: DRAWN:

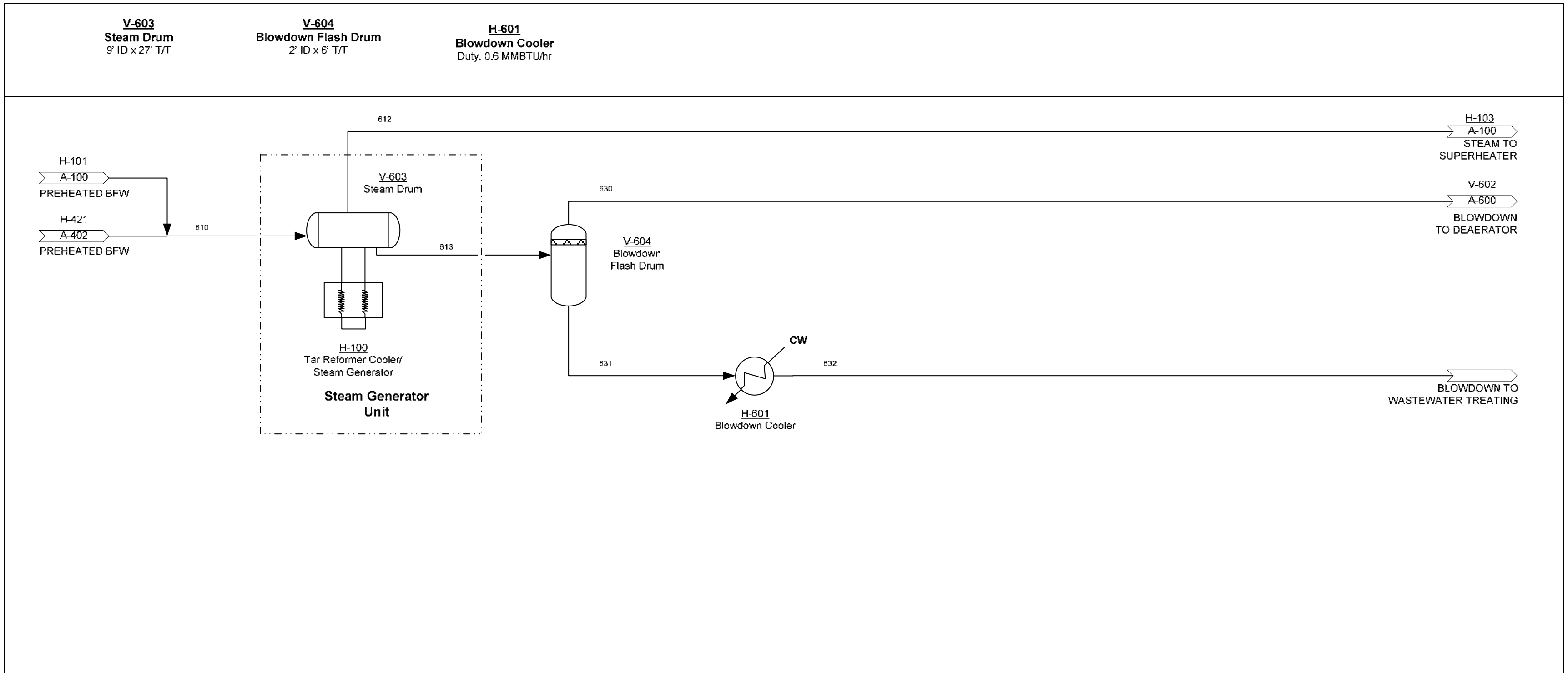
Nexant — *San Francisco*

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-600: STEAM SYSTEM AND POWER GENERATION

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005



	Unit	610	612	613	630	631	632
Mass Flow	lb/hr	256,896	251,758	5,138	1,648	3,490	3,490
Temperature	F	547	575	547	298	298	115
Pressure	psia	1,285.0	1,275.0	1,285.0	65.0	65.0	60.0
Vapour Fraction		0.00	1.00	0.00	1.00	0.00	0.00
MW		18.0	18.0	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-1,614.3	-1,415.0	-32.3	-9.3	-23.0	-23.6
Components							
Hydrogen	lb/hr	0	0	0	0	0	0
CO2	lb/hr	0	0	0	0	0	0
CO	lb/hr	0	0	0	0	0	0
H2O	lb/hr	256,896	251,758	5,138	1,648	3,490	3,490
Methane (CH4)	lb/hr	0	0	0	0	0	0
Acetylene (C2H2)	lb/hr	0	0	0	0	0	0
Ethylene (C2H4)	lb/hr	0	0	0	0	0	0
Ethane (C2H6)	lb/hr	0	0	0	0	0	0
Propane	lb/hr	-	-	-	-	-	-
Isobutane	lb/hr	-	-	-	-	-	-
n-Butane	lb/hr	-	-	-	-	-	-
Pentane	lb/hr	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	0	0	0	0	0	0
Tar (C10H8)	lb/hr	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-
H2S	lb/hr	0	0	0	0	0	0
Char	lb/hr	-	-	-	-	-	-
Oxygen	lb/hr	-	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-	-
Nitrogen	lb/hr	0	0	0	0	0	0
Sulfur (solid)	lb/hr	-	-	-	-	-	-

Nexant
San Francisco

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-601: STEAM SYSTEM AND POWER GENERATION

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005

M-601
Extraction Steam Turbine/Generator
19410 kW

	Unit	614	618	619	621	622	
Mass Flow	lb/hr	251,758	18,832	53,000	106,459	73,467	
Temperature	F	1,000	487	376	376	376	
Pressure	psia	1,265.0	100.0	50.0	50.0	50.0	
Vapour Fraction		1.00	1.00	1.00	1.00	1.00	
MW		18.0	18.0	18.0	18.0	18.0	
Heat Flow	MMBtu/hr	-1,347.7	-104.9	-297.9	-598.3	-412.9	
Components							
Hydrogen	lb/hr	0	0	0	0	0	
CO2	lb/hr	0	0	0	0	0	
CO	lb/hr	0	0	0	0	0	
H2O	lb/hr	251,758	18,832	53,000	106,459	73,467	
Methane (CH4)	lb/hr	0	0	0	0	0	
Acetylene (C2H2)	lb/hr	0	0	0	0	0	
Ethylene (C2H4)	lb/hr	0	0	0	0	0	
Ethane (C2H6)	lb/hr	0	0	0	0	0	
Propane	lb/hr	-	-	-	-	-	
Isobutane	lb/hr	-	-	-	-	-	
n-Butane	lb/hr	-	-	-	-	-	
Pentane	lb/hr	-	-	-	-	-	
Benzene (C6H6)	lb/hr	0	0	0	0	0	
Tar (C10H8)	lb/hr	-	-	-	-	-	
Ammonia (NH3)	lb/hr	-	-	-	-	-	
H2S	lb/hr	0	0	0	0	0	
Char	lb/hr	-	-	-	-	-	
Oxygen	lb/hr	-	-	-	-	-	
Argon	lb/hr	-	-	-	-	-	
Nitrogen	lb/hr	0	0	0	0	0	
Sulfur (solid)	lb/hr	-	-	-	-	-	

No.	DATE	REVISIONS	BY	CHK'D	DESIGN SUPV	ENGR.	PROJ. ENGR.	APPR.

SCALE: DESIGNED: DRAWN:

Nexant
San Francisco

NREL BIOMASS GASIFICATION
LOW-PRESSURE SYNGAS CASE

PROCESS FLOW DIAGRAM
A-602: STEAM SYSTEM AND POWER GENERATION

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 1

15 Nov, 2005

Appendix C Equipment Lists and Data Sheets

The following two appendices show the equipment lists for the high-pressure and low-pressure syngas design cases, along with detailed data sheets for some of the major pieces of equipment. No specific detail was developed for the tar cracking equipment due to the preliminary nature of its design. In addition, no additional information beyond what is presented in the equipment list was produced for vessels and pumps. Detailed equipment sheets are only shown for exchangers, cyclones, and compressors, where additional design data was developed.

HIGH PRESSURE SYNGAS DESIGN CASE

									Design		Operating								
Item No	Description	Type	Quantity Per Train	Size	Weight	Head	Design Duty		P	T	P	T	Power Useage	Materials	Price, total (unistalled)	Price Escalated, total (uninstalled)	Total Installed Cost	Quote Source	Comments
					lbs	PSI			PSIG	ºF	PSIG	ºF	(No.) HP		Q2 2004 Cost Index (US $)	Q2 2005 Cost Index (US $)	(US $)		
Reactors																			
R-100	Tar Reformer	Fluidized Bed		20' ID x 20' T/T					490	1675	445	1576		Refractory lined CS		$950,942		GTI	662,000 lbs catalyst req'd
R-101	Catalyst Regenerator			22' ID x 70' T/T					20	1950	5	1850		Refractory lined CS		$329,616		GTI	
R-320	ZnO Beds	Vertical	1	4' - 3" ID x 8' T/T	43,856		2 ppmv H2S inlet		445	850	402	750		CS		$219,280		Johnson Matthey	707 ft^3 total catalyst volume req'd
R-321	ZnO Beds	Vertical	1	4' - 3" ID x 8' T/T	43,856		2 ppmv H2S inlet		445	850	402	750		CS		$219,280		Johnson Matthey	
Total																**$1,719,118**			
Cyclones																			
S-100	Gasifier Cyclone	Cyclone	1	5' ID x 25' T/T			3304 lb/hr dust loading		490	650	445	1576		CS w/ 4" refractory lining		$355,000		Fisher Kosterman	Refractory lining will bring the shell temperature to 590F.
S-102	Tar Reformer Cyclone	Cyclone	1	5' - 6" ID x 27' - 6" T/T			1128 lb/hr dust loading		490	650	442	1576		CS w/ 4" refractory lining		$410,000		Fisher Kosterman	Refractory lining will bring the shell temperature to 590F.
S-103	Catalyst Regenerator Cyclone	Cyclone	1	5' ID x 25' T/T			1128 lb/hr dust loading		490	650	442	1576		CS w/ 4" refractory lining		$265,000		Fisher Kosterman	Refractory lining will bring the shell temperature to 590F.
Total																**$1,030,000**			
Columns, Vessels & Tanks																			
C-200	Syngas Venturi Scrubber & Quench Tower	Vertical	1	3' - 2" ID x 7' - 9" T-T (Venturi); 5' - 3" ID x 8' - 11" T-T (Quench Tower)					485	420	432	370		CS		$316,000		EPA Cost Curve	
V-400	Treated Gas KO Drum	Vertical	1	5' ID x 15' T-T	31700				422	160	382	110		CS	$31,700	$37,580		ICARUS	
V-400A	Interstage KO Drum	Vertical	1	5' ID x 15' T-T	29300				1030	250	980	200		CS	$57,800	$68,522		ICARUS	
V-500	Condensate Collection Drum	Vertical	1	5' ID x 13' T-T	4170				15	265	0	212		CS	$14,745	$17,480		ICARUS	
V-501	Condensate Surge Drum	Horizontal	1	6' ID x 17' T-T	6300				15	145	0	94		CS	$22,195	$26,312		ICARUS	
V-502	Deaerator	Horizontal	1	6' ID x 17' T-T; 2' ID x 7' T-T	7900				25	290	10	240		CS	$31,350	$37,165		ICARUS	
V-503	Steam Drum	Horizontal	1	9' ID x 27' T-T	139300				1335	625	1265	575		SA 302B	$764,205	$1,018,227		ICARUS	
V-504	Blowdown Flash Drum	Vertical	1	2' ID x 6' T-T	1300				65	350	50	298		CS	$8,200	$9,721		ICARUS	
T-200	Sludge Settling Tank	Horizontal	1	1' ID x 3' T/T	300				475	415	430	364		CS	$4,800	$5,690		ICARUS	
T-201	Quench Water Recirculation Tank	Horizontal	1	4' - 6" ID x 14' T/T	3600				475	360	430	311		CS	$14,460	$17,142		ICARUS	
Total																**$1,553,840**			
Heat Exchangers																			
				5' - 7" ID x 12' T-T				T	1335	625	1270	575		CS					
H-100	Tar Reformer SG Cooler/Steam Generator	Shell & Tube	2	Surface area: 5206 SQFT			203.7 MMBTU/hr	S	485	1675	442	1576		CS - refractory	$1,465,600	$1,664,628		ICARUS	Refractory Lined
				7' - 6" ID x 20' T-T				T	1335	600	1270	551		CS					
H-101	Tar Reformer SG Cooler/BFW Preheater	Shell & Tube	1	Surface area: 23969 SQFT			50.84 MMBTU/hr	S	485	675	437	624		CS	$513,500	$583,233		ICARUS	
				8' - 4" ID x 14' T-T				T	1335	1100	1255	1000		316S					
H-102	Flue Gas Cooler/Steam Superheater	Shell & Tube	1	Surface area: 8915 SQFT			83.65 MMBTU/hr	S	15	1900	0	1798		CS - refractory	$1,598,750	$1,815,860		ICARUS	Refractory Lined
				3' - 6" ID x 10' T-T				T	485	415	441	364		CS					
H-200	Quench Water Recirculation	Shell & Tube	1	Surface area: 2867 SQFT			22.34 MMBTU/hr	S	20	150	5	100		CS	$80,000	$90,864		ICARUS	
				4' - 8" ID x 14' T-T				T	1335	400	1280	349		CS					
H-201	Amine Precooler/BFW Preheat	Shell & Tube	1	Surface area: 7511 SQFT			36.99 MMBTU/hr	S	470	410	427	356		CS	$260,300	$295,649		ICARUS	
				3' - 4" ID x 6' T-T				T	30	300	15	239		CS					
H-202	Amine Precooler/Deaerator FW Preheat	Shell & Tube	1	Surface area: 585 SQFT			9.24 MMBTU/hr	S	465	400	422	338		CS	$16,260	$18,468		ICARUS	
				8' ID x 8' T-T				T	65	150	50	100		CS					
H-203	Amine Precooler	Shell & Tube	1	Surface area: 11541 SQFT			139.3 MMBTU/hr	S	460	350	432	305		CS	$309,600	$351,644		ICARUS	
				8' ID x 8' T-T				T	450	800	407	750		CS					
H-320	ZnO Preheater	Shell & Tube	1	Surface area: 19400 SQFT			49.69 MMBTU/hr	S	15	910	0	839		CS	$288,000	$327,110		ICARUS	
				5' ID x 16' T-T				T	1335	615	1270	565		CS					
H-321	ZnO SG Cooler/BFW Preheater	Shell & Tube	1	Surface area: 5440 SQFT			29.85 MMBTU/hr	S	440	800	397	750		CS	$192,600	$218,755		ICARUS	
				3' ID x 8' T-T				T	65	150	50	100		CS					
H-322	Post ZnO Syngas Cooler	Shell & Tube	1	Surface area: 1620 SQFT			19.91 MMBTU/hr	S	435	420	393	370		CS	$56,100	$63,718		ICARUS	
				1' - 11" ID x 6' T-T				T	1035	390	985	338		CS					
H-400A	MeOH Compressor Interstage Cooler	Shell & Tube	1	Surface area: 476 SQFT			10.47 MMBTU/hr	S	65	150	100	50		CS	$32,200	$36,573		ICARUS	
				6' ID x 18' T-T				T	1210	515	1150	460		CS					
H-401	MeOH Syngas Preheat	Shell & Tube	1	Surface area: 16212 SQFT			17.14 MMBTU/hr	S	100	525	85	472		CS	$355,140	$403,368		ICARUS	
				1' - 3" ID x 4' T-T				T	65	150	50	100		CS					
H-501	Blowdown Cooler	Shell & Tube	1	Surface area: 130 SQFT			0.84 MMBTU/hr	S	65	350	50	298		CS	$19,100	$21,694		ICARUS	
Total																**$5,891,565**			
Compressors & Blowers																			
K-100	Combustion Air Blower	Blower	2	61910 ACFM		5					0	90	1800	CS		$274,305		Chicago Blower Corp./ ICARUS	Used ICARUS to cost motor. 2 - 100% blowers
K-320	Flue Gas Blower	Blower	2	85400 ACFM		0.4					0	214	207	CS		$233,875		Scaled fr. Chicago Blower	2 - 100% blowers
K-400	MeOH Compressor - 2 Stages	Centrifugal	1	2746 ACFM	74,500	758					387	110	8388	CS	$2,133,200	$2,522,936		ICARUS	
Total																**$3,031,115**			
Pumps																			
P-201	Quench Water Recirculation	Centrifugal	2	282 GPM	420	14			475	360	430	311	3	CS	$10,600	$11,021		ICARUS	2 - 100% pumps
P-500	Condensate Make-up Water Pump	Centrifugal	2	147 GPM	440	5			20	110	0	60	1.3	CS	$5,400	$5,614		ICARUS	2 - 100% pumps
P-501	Deaerator Feed Pump	Centrifugal	2	674 GPM	680	15			30	150	0	98	8	CS	$17,200	$17,883		ICARUS	2 - 100% pumps
P-502	Boiler Feed Water Pump	Centrifugal	2	684 GPM	9,000	1,270			1345	290	11	240	710	CS	$325,000	$337,903		ICARUS	2 - 100% pumps
Total																**$372,421**			
Steam Turbine																			
M-501	Steam Turbine	Steam Turbine	1		172,900	-1,160					1245	1000	(14710 kW)	CS	$4,534,500	$5,362,953		ICARUS	
Total																**$5,362,953**			
Package Units																			
A-300	Amine Unit																**$22,413,600**	GRI Cost Curve	
A-301	LO-CAT Unit															**$3,998,550**	**$5,348,550**	Gas Technology Products	
TOTAL EQUIPMENT COST, (excld. Package units)																**$18,961,012**	**$48,729,802**		Installation factor of 2.57 used
TOTAL INSTALLED COST																	**$76,491,952**		

LOW PRESSURE SYNGAS DESIGN CASE

									Design		Operating								
Item No	Description	Type	Quantity	Size, each	Weight	Head	Design Duty, total		P	T	P	T	Power Useage	Materials	Price, total (unistalled)	Price Escalated, total (uninstalled)	Total Installed Cost	Quote Source	Comments
					lbs	PSI			PSIG	ºF	PSIG	ºF	(No.) HP		Q2 2004 Cost Index (US $)	Q2 2005 Cost Index (US $)	(US $)		
Reactors																			
R-100	Tar Reformer	Fluidized Bed	1	29' ID x 89' T/T					30	1700	15	1598		Refractory lined CS		$921,786		GTI	1,820,000 lbs catalyst req'd
R-101	Catalyst Regenerator		1	23' ID x 72' T/T					30	1700	15	1598		Refractory lined CS		$545,886		GTI	
R-420	ZnO Beds	Vertical	1	4' - 6" ID x 8' T/T	44,522		2 ppmv H2S inlet		455	850	415	750		CS		$222,610		Johnson Matthey	777 ft^3 total catalyst volume req'd
R-421	ZnO Beds	Vertical	1	4' - 6" ID x 8' T/T	44,522		2 ppmv H2S inlet		455	850	415	750		CS		$222,610		Johnson Matthey	
Total																**$1,912,892**			
Cyclones																			
S-100 A/B/C/D	Gasifier Cyclone	Cyclone	4	7' ID x 35' T/T			14,142 lb/hr dust loading		33	650 (see comments)	18	1598		CS w/ 4" refractory lining		$1,225,000		Fisher Kosterman	Refractory lining will bring the shell temperature to 590F.
S-102	Tar Reformer Cyclone	Cyclone	1	5' ID x 25' T/T			1,000 lb/hr dust loading		33	650 (see comments)	15	1598		CS w/ 4" refractory lining		$370,000		Fisher Kosterman	Refractory lining will bring the shell temperature to 590F.
S-103	Catalyst Regenerator Cyclone	Cyclone	1	4' ID x 17' T/T			1,000 lb/hr dust loading		33	650 (see comments)	15	1598		CS w/ 4" refractory lining		$250,000		Fisher Kosterman	Refractory lining will bring the shell temperature to 590F.
Total																**$1,845,000**			
Columns, Vessels & Tanks																			
C-200	Syngas Venturi Scrubber & Quench Tower	Vertical	2	14' ID x 29' T/T					19	275	4	225		CS		$340,000		Croll Reynolds	
V-300	Syngas KO Drum	Vertical	2	14' ID x 28' T/T	31,500				16	210	1	157		CS	$306,800	$363,711		ICARUS	
V-300A	1st Stage KO Drum	Vertical	1	13' ID x 26' T/T	25,500				30	160	15	110		CS	$73,400	$87,016		ICARUS	
V-300B	2nd Stage KO Drum	Vertical	1	10' ID x 20' T/T	24,700				79	160	64	110		CS	$54,300	$64,373		ICARUS	
V-300C	3rd Stage KO Drum	Vertical	1	7' ID x 21' T/T	21,900				220	160	200	110		CS	$41,800	$49,554		ICARUS	
V-300D	Post KO Drum	Vertical	1	6' ID x 18' T/T	23,600				475	160	430	110		CS	$45,400	$53,822		ICARUS	
V-500	Treated Gas KO Drum	Vertical	1	5' ID x 15' T/T	14,900				440	160	400	110		CS	$31,800	$37,699		ICARUS	
V-500A	Interstage KO Drum	Vertical	1	5' ID x 15' T/T	29,300				1,030	250	980	200		CS	$57,800	$68,522		ICARUS	
V-600	Condensate Collection Tank	Vertical	1	5' ID x 12' T/T	3,990				15	245	0	195		CS	$14,100	$16,716		ICARUS	
V-601	Condensate Surge Drum	Horizontal	1	5' - 6" ID x 16' - 6" T/T	5,483				15	245	0	195		CS	$19,320	$22,904		ICARUS	
V-602	Deaerator	Vertical	1	6' ID x 18' T/T; 2' ID x 6' T/T	7,800				25	290	10	237		CS	$35,700	$42,322		ICARUS	
V-603	Steam Drum	Horizontal	1	9' ID x 27' T/T	139,300				1335	625	1270	575		SA 302B	$764,205	$1,018,227		ICARUS	
V-604	Blowdown Flash Drum	Vertical	1	2' ID x 6' T/T	1,200				65	350	50	298		CS	$7,500	$8,891		ICARUS	
T-200	Sludge Settling Tank	Horizontal	1	1' ID x 3' T/T	300				16	180	1	128		CS	$4,000	$4,742		ICARUS	
T-201	Quench Water Recirculation Tank	Horizontal	1	9' ID x 27' T/T	15,300				16	180	1	128		CS	$60,700	$71,960		ICARUS	
Total																**$2,250,458**			
Heat Exchangers																			
				6' ID x 14' T/T				T	1335	625	1270	575		CS					
H-100	Tar Reformer SG Cooler/Steam Generator	Shell & Tube	2	Surface area: 5354 SQFT			167 MMBTU/hr	S	30	1700	15	1598		CS - refractory	$989,400	$1,129,202		ICARUS	Refractory Lined
				4' - 9" ID x 14' T/T				T	1335	600	1,280	542		CS					
H-101	Tar Reformer SG Cooler/BFW Preheater	Shell & Tube	2	Surface area: 6667 SQFT			50.61 MMBTU/hr	S	20	675	12	624		CS	$682,550	$775,240		ICARUS	
				6' - 3" ID x 14' T/T				T	30	280	15	227		CS					
H-102	Tar Reformer Cooler/Deaerator FW Preheat	Shell & Tube	1	Surface area: 5621 SQFT			11.34 MMBTU/hr	S	20	350	9	300		CS	$104,600	$118,805		ICARUS	
				7' - 6" ID x 14' T/T				T	1335	1100	985	1275		316S					
H-103	Flue Gas Cooler/Steam Superheater	Shell & Tube	1	Surface area: 5770 SQFT			67.26 MMBTU/hr	S	15	1900	0	1798		CS - refractory	$1,016,858	$1,154,947		ICARUS	Refractory Lined
				5' - 11" ID x 10' T/T				T	30	150	5	100		CS					
H-200	Quench Water Recirculation Cooler	Shell & Tube	1	Surface area: 9232 SQFT			22.2 MMBTU/hr	S	30	215	11	161		CS	$203,800	$231,476		ICARUS	
				6' - 10" ID x 12' T/T				T	35	400	20	344		CS					
H-300A	Compressor Interstage Cooling	Shell & Tube	2	Surface area: 14235 SQFT			122 MMBTU/hr	S	65	150	50	100		CS	$802,600	$911,593		ICARUS	
				3' - 11" ID x 10' T/T				T	65	150	50	100		CS					
H-300B	Compressor Interstage Cooling	Shell & Tube	1	Surface area: 3435 SQFT			32.79 MMBTU/hr	S	85	400	69	350		CS	$72,300	$82,118		ICARUS	
				4' - 3" ID x 10' T/T				T	230	400	205	349		CS					
H-300C	Compressor Interstage Cooling	Shell & Tube	1	Surface area: 4368 SQFT			27.69 MMBTU/hr	S	65	150	50	100		CS	$95,000	$107,901		ICARUS	
				3' - 6" ID x 10' T/T				T	485	330	435	277		CS					
H-300D	Compressor Interstage Cooling	Shell & Tube	1	Surface area: 2934 SQFT			18.21 MMBTU/hr	S	65	150	50	100		CS	$74,900	$85,071		ICARUS	
				7' - 6" ID x 8' T/T				T	465	800	420	750		CS					
H-420	ZnO Preheater	Shell & Tube	1	Surface area: 14480 SQFT			52.90 MMBTU/hr	S	15	990	0	945		CS	$289,300	$328,587		ICARUS	
				5' - 4" ID x 12' T/T				T	1335	600	1,280	542		CS					
H-421	ZnO Syngas Cooler/BFW Preheat	Shell & Tube	1	Surface area: 6915 SQFT			40.57 MMBTU/hr	S	455	800	410	750		CS	$244,300	$277,476		ICARUS	
				2' - 6" ID x 8' T/T				T	65	150	50	100		CS					
H-422	ZnO Syngas Cooler	Shell & Tube	1	Surface area: 1190 SQFT			11.86 MMBTU/hr	S	450	315	405	265		CS	$41,210	$46,806		ICARUS	
				2' - 6" ID x 8' T/T				T	1,035	385	985	333		CS					
H-500A	MeOH Compressor Interstage Cooling	Shell & Tube	1	Surface area: 511 SQFT			11.06 MMBTU/hr	S	65	150	50	100		CS	$33,800	$38,390		ICARUS	
				6' ID x 14' T/T				T	1,261	515	1,145	460		CS					
H-501	MeOH Syngas Preheat	Shell & Tube	1	Surface area: 12712 SQFT			18.45 MMBTU/hr	S	100	540	85	487		CS	$278,500	$316,320		ICARUS	
				1' ID x 4' T/T				T	65	150	50	100		CS					
H-601	Blowdown Cooler	Shell & Tube	1	Surface area: 89 SQFT			0.609 MMBTU/hr	S	65	350	50	298		CS	$18,400	$20,899		ICARUS	
Total																**$5,624,833**			
Compressors																			
K-100	Combustion Air Blower	Blower	2	54910 ACFM		3					0	90	1600	CS		$256,425		Chicago Blower Corp./ ICARUS	Used ICARUS to cost motor. 2 - 100% blowers
K-420	Flue Gas Blower	Blower	2	73100 ACFM		0.4					0	176	177	CS		$202,375		Scaled fr. Chicago Blower	2 - 100% blowers
K-300	Syngas Compressor- 4 stages	Centrifugal	1	131800 ACFM	333,100	434					1	157	38,786	CS		$15,000,000	$37,050,000	Elliott	2.47 installation factor
K-500	MeOH Compressor- 2 stages	Centrifugal	1	2854 ACFM	31,100	745					399	115	8,717	CS		$2,369,000		Ariel Corp.	
Total																**$17,827,800**			
Pumps																			
P-201	Quench Recirculation Pump	Centrifugal	2	2423 GPM	800	10.12			26	211	1.18	160.9	20	CS	$91,000	$94,613		ICARUS	2 - 100% pumps
P-600	Condensate Make-up Water Pump	Centrifugal	2	172 GPM	440	5			20	110	0	60	1	CS	$6,320	$6,571		ICARUS	2 - 100% pumps
P-601	Deaerator Feed Pump	Centrifugal	2	537 GPM	810	15.3			30	160	15	108.8	7	CS	$17,400	$18,091		ICARUS	2 - 100% pumps
P-602	Boiler Feedwater Pump	Centrifugal	2	548 GPM	8,900	1275			1350	278	20	227.9	570	CS	$316,800	$329,377		ICARUS	2 - 100% pumps
Total																**$448,651**			
Steam Turbine																			
M-601	Steam Turbine- 2 extraction stages	Steam Turbine	1		221,200	-1200					1,250	1,000	(19410 kW)	CS	$5,459,900	$6,457,424		ICARUS	
Total																**$6,457,424**			
Package Units																			
A-400	Amine Unit																**$12,452,000**	GRI Cost Curve	
A-401	LO-CAT Unit															**$3,733,550**	**$5,003,550**	Gas Technology Products	
TOTAL EQUIPMENT COST, (excld. Package units)																**$36,367,057**	**$91,963,336**		Installation factor of 2.57 used on all equipment except syngas compressor
TOTAL INSTALLED COST																	**$109,418,886**		

DATA SHEETS, HIGH PRESSURE DESIGN

Heat Exchanger Specification sheet

					Job No.		
Customer	NREL				Ref No.	HP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Tar Reformer SG Cooler/HP Steam Generator				Item No	H-100	
Size 67x 144		Type	BEM - HORZ	Connected in	2 Parallel		1 Series
Surf/Unit (Eff)	10411 ft²	Shells/Unit	2	Surface/Shell (Effective)		5206 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas fr Tar Reformer	Preheated BFW
Total Fluid Entering	lb/hr	435,000	313,900
Vapor		435,000	0
Liquid		0	313,900
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	313,900
Liquid Density (In/Out)	lb/ft³	0.000/0.000	46.162/45.419
Liquid Viscosity	cP	0.000	0.091
Liquid Specific Heat	Btu/lb-F	0.000	1.644
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.320
Vapor Mol. Weight (In/Out)		18.66/18.66	0.0/18.02
Vapor Viscosity	cP	0.0285	0.0200
Vapor Specific Heat	Btu/lb-F	0.492	0.774
Vapor Thermal Conductivity	Btu/hr-ft-F	0.067	0.025
Temperature (In/Out)	°F	1,576.0/624.0	556.0/575.0
Operating Pressure	psi(Abs)	457.000	1,285.000
Velocity	ft/sec	43.504	8.337
Pressure Drop (Allow/Calc)	psi	5.000/3.245	5.000/1.133
Fouling resistance	hr-ft²-F/Btu	0.001000	0.005000

Heat Exchanged	203,700,000 Btu/hr		mtd (corr)	307.674 °F
Transfer Rate, Service		63.6	Clean	128.2 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	500/	1,360/	
Design Temp.	°F	1675	600	
No. Passes per Shell		1	6	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-19.0	6.0	
Size &	Out	1-17.0	12.0	
Rating	Intermediate	0	0	

Tube No	1912	OD 1.000 in	Thk 0.065	Length 12.00 ft		Pitch 1.25000 / 30.0°
Tube Type		PLAIN	Material			
Shell		I.D 67.00 OD in	Shell Cover		INT	
Channel or Bonnet			Channel Cover			
Tubesheet-Stationary			Tubesheet-Floating			
Floating Head Cover			Impingement Protection		YES	
Baffles Cross		Type VERT- SEG	%Cut 19.1 (Area)		Spacing-cc	29.1
Baffles-Long			Seal Type			
Supports-Tube		U-Bend		Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint			
Expansion Joint			Type			
Rho-V2 Inlet Nozzle		2,412	Bundle Entrance	1,266	Bundle Exit	2,915
Gasket-Shellside			Tubeside		Floating Head	
Code Requirement		ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell			Filled with Water		Bundle	

Heat Exchanger Specification sheet							
					Job No.		
Customer	NREL				Ref No.	HP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Tar Reformer SG Cooler/BFW Preheater				Item No	H-101	
Size 90x 240		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	23969 ft²	Shells/Unit	1	Surface/Shell (Effective)		23969 ft²	

PERFORMANCE OF ONE UNIT			
Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas fr Tar Reformer	BFW
Total Fluid Entering	lb/hr	435,000	208,600
Vapor		435,000	0
Liquid		0	208,600
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	55.214/46.316
Liquid Viscosity	cP	0.000	0.116
Liquid Specific Heat	Btu/lb-F	0.000	1.368
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.358
Vapor Mol. Weight (In/Out)		18.66/18.66	0.0/0.0
Vapor Viscosity	cP	0.0199	0.0000
Vapor Specific Heat	Btu/lb-F	0.461	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.044	0.000
Temperature (In/Out)	°F	624.0/370.0	349.0/551.0
Operating Pressure	psi(Abs)	452.000	1,285.000
Velocity	ft/sec	33.096	-
Pressure Drop (Allow/Calc)	psi	10.000/8.600	5.000/0.359
Fouling resistance	hr-ft²-F/Btu	0.001000	0.005000

Heat Exchanged	50,840,000 Btu/hr	mtd (corr)	41.736 °F
Transfer Rate, Service	50.8	Clean	86.9 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL				
		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	500/	1,350/	
Design Temp.	°F	675	600	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-19.0	6.0	
Size &	Out	1-19.0	6.0	
Rating	Intermediate	0	0	

Tube No	6830	OD 0.750 in	Thk 0.065	Length 20.00 ft		Pitch 1.00000 / 30.0°
Tube Type		PLAIN	Material			
Shell		I.D 90.00 OD in	Shell Cover		INT	
Channel or Bonnet			Channel Cover			
Tubesheet-Stationary			Tubesheet-Floating			
Floating Head Cover			Impingement Protection		YES	
Baffles Cross		Type VERT- SEG	%Cut 13.8 (Area)		Spacing-cc	24.1
Baffles-Long			Seal Type			
Supports-Tube		U-Bend		Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint			
Expansion Joint			Type			
Rho-V2 Inlet Nozzle	5,194	Bundle Entrance		1,440	Bundle Exit	4,997
Gasket-Shellside		Tubeside			Floating Head	
Code Requirement	ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell		Filled with Water			Bundle	
Remarks:						

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	HP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	Flue Gas Cooler/Steam Superheater		Item No	H-102	
Size 100x 168	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	8915 ft²	Shells/Unit 1	Surface/Shell (Effective)	8915 ft²	

PERFORMANCE OF ONE UNIT			
Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Flue Gas fr. Tar Regen	Superheated Steam
Total Fluid Entering	lb/hr	280,200	313,900
Vapor		280,200	313,900
Liquid		0	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	0.000/0.000
Liquid Viscosity	cP	0.000	0.000
Liquid Specific Heat	Btu/lb-F	0.000	0.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.000
Vapor Mol. Weight (In/Out)		27.56/27.56	18.02/18.02
Vapor Viscosity	cP	0.0399	0.0254
Vapor Specific Heat	Btu/lb-F	0.314	0.676
Vapor Thermal Conductivity	Btu/hr-ft-F	0.039	0.036
Temperature (In/Out)	°F	1,798.0/839.0	575.0/1,000.0
Operating Pressure	psi(Abs)	14.700	1,270.000
Velocity	ft/sec	211.463	4.576
Pressure Drop (Allow/Calc)	psi	2.000/1.798	5.000/0.484
Fouling resistance	hr-ft²-F/Btu	0.001000	0.005000

Heat Exchanged	83,650,000 Btu/hr	mtd (corr)	482.751 °F
Transfer Rate, Service	19.4	Clean	22.7 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL				
		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	30/	1,350/	
Design Temp.	°F	1900	1100	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-61.0	15.0	
Size &	Out	1-55.0	15.0	
Rating	Intermediate	0	0	

Tube No	3900	OD 0.750 in	Thk 0.065	Length 14.00 ft	Pitch 1.25000 / 45.0°
Tube Type	PLAIN		Material		
Shell	I.D 100.00 OD in		Shell Cover	INT	
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 40.7 (Area)	Spacing-cc	69.9
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	880	Bundle Entrance	3,144	Bundle Exit	1,037
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet

				Job No.		
Customer	NREL			Ref No.	HP Syngas Csae	
Address				Proposal No.		
Plant Location				Date		Rev. 0
Service of Unit	Quench Water Recirculation			Item No	H-200	
Size 42x 120		Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	2867 ft²	Shells/Unit	1	Surface/Shell (Effective)	2867 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Cooling Water	Quench Water
Total Fluid Entering	lb/hr	1,117,000	105,700
Vapor		0	0
Liquid		1,117,000	105,700
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	61.436/61.060	57.041/61.765
Liquid Viscosity	cP	0.510	0.301
Liquid Specific Heat	Btu/lb-F	1.005	1.017
Liquid Thermal Conductivity	Btu/hr-ft-F	1.122	0.381
Vapor Mol. Weight (In/Out)		0.0/0.0	0.0/0.0
Vapor Viscosity	cP	0.0000	0.0000
Vapor Specific Heat	Btu/lb-F	0.000	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.000
Temperature (In/Out)	°F	80.0/100.0	311.0/110.0
Operating Pressure	psi(Abs)	20.000	456.000
Velocity	ft/sec	3.475	-
Pressure Drop (Allow/Calc)	psi	5.000/3.632	5.000/0.424
Fouling resistance	hr-ft²-F/Btu	0.002000	0.001000

Heat Exchanged	22,340,000 Btu/hr	mtd (corr)	92.789 °F
Transfer Rate, Service	84.0	Clean	115.0 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	35/	500/	
Design Temp.	°F	150	415	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-12.0	4.0	
Size &	Out	1-12.0	4.0	
Rating	Intermediate	0	0	

Tube No	1558	OD 0.750 in	Thk 0.065	Length 10.00 ft		Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material			
Shell	I.D 42.00 OD in		Shell Cover		INT	
Channel or Bonnet			Channel Cover			
Tubesheet-Stationary			Tubesheet-Floating			
Floating Head Cover			Impingement Protection		YES	
Baffles Cross	Type VERT- SEG		%Cut 22.8 (Area)		Spacing-cc	24.0
Baffles-Long			Seal Type			
Supports-Tube		U-Bend		Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint			
Expansion Joint			Type			
Rho-V2 Inlet Nozzle	2,540	Bundle Entrance		1,308	Bundle Exit	3,750
Gasket-Shellside		Tubeside			Floating Head	
Code Requirement	ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell		Filled with Water			Bundle	
Remarks:						

Heat Exchanger Specification sheet					
			Job No.		
Customer	NREL		Ref No.	HP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev.
Service of Unit	Amine Precooler/BFW Preheat		Item No	H-201	
Size 56x 168	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	7511 ft²	Shells/Unit 1	Surface/Shell (Effective)	7511 ft²	

PERFORMANCE OF ONE UNIT			
Fluid Allocation		Shellside	Tubeside
Fluid Name		BFW	Syngas to Amine Absorber
Total Fluid Entering	lb/hr	320,300	414,200
Vapor		0	414,200
Liquid		320,300	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	40,260
Liquid Density (In/Out)	lb/ft³	58.527/55.201	0.000/56.407
Liquid Viscosity	cP	0.188	0.150
Liquid Specific Heat	Btu/lb-F	1.086	1.037
Liquid Thermal Conductivity	Btu/hr-ft-F	0.393	0.404
Vapor Mol. Weight (In/Out)		0.0/0.0	18.69/18.69
Vapor Viscosity	cP	0.0000	0.0176
Vapor Specific Heat	Btu/lb-F	0.000	0.467
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.040
Temperature (In/Out)	°F	242.0/349.0	356.0/338.0
Operating Pressure	psi(Abs)	1,295.000	442.000
Velocity	ft/sec	0.893	18.179
Pressure Drop (Allow/Calc)	psi	5.000/0.697	5.000/0.635
Fouling resistance	hr-ft²-F/Btu	0.002000	0.001000

Heat Exchanged	36,985,000 Btu/hr	mtd (corr)	34.052 °F
Transfer Rate, Service	144.6	Clean	300.2 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL				
		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	1,425/	1,360/	
Design Temp.	°F	410	400	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-8.0	23.0	
Size &	Out	1-8.0	23.0	
Rating	Intermediate	0	0	

Tube No	3030	OD 0.750 in	Thk 0.065	Length 14.00 ft	Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 56.00 OD in		Shell Cover	INT	
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 10.0 (Area)	Spacing-cc	11.1
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	1,110	Bundle Entrance	167	Bundle Exit	1,585
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	HP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	Amine Precooler/Deaerator FW Preheat		Item No	H-202	
Size 40x 72	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	585 ft^2	Shells/Unit 1	Surface/Shell (Effective)	585 ft^2	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas to Amine Absorber	Deaerator Feed Water
Total Fluid Entering	lb/hr	414,200	320,000
Vapor		373,940	0
Liquid		40,260	320,000
Steam			
Noncondensable			
Fluid Vaporized or Condensed		9,444	0
Liquid Density (In/Out)	lb/ft^3	55.290/55.492	59.180/58.595
Liquid Viscosity	cP	0.092	0.262
Liquid Specific Heat	Btu/lb-F	1.111	1.020
Liquid Thermal Conductivity	Btu/hr-ft-F	0.395	0.385
Vapor Mol. Weight (In/Out)		18.96/18.9436	0.0/0.0
Vapor Viscosity	cP	0.0179	0.0000
Vapor Specific Heat	Btu/lb-F	0.445	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.041	0.000
Temperature (In/Out)	°F	338.0/332.0	212.0/239.4
Operating Pressure	psi(Abs)	437.000	30.000
Velocity	ft/sec	25.051	-
Pressure Drop (Allow/Calc)	psi	5.000/1.075	5.000/0.287
Fouling resistance	hr-ft^2-F/Btu	0.001000	0.002000

Heat Exchanged	9,238,000 Btu/hr	mtd (corr)	108.950 °F
Transfer Rate, Service	145.0	Clean	322.3 Btu/hr-ft^2-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	480/	45/	
Design Temp.	°F	400	300	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-23.0	8.0	
Size &	Out	1-19.0	8.0	
Rating	Intermediate	0	0	

Tube No	550	OD 0.750 in	Thk 0.065	Length 6.00 ft	Pitch 1.25000 / 45.0°
Tube Type	PLAIN		Material		
Shell	I.D 40.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 49.0 (Area)	Spacing-cc	38.9
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	1,486	Bundle Entrance	2,490	Bundle Exit	2,529
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	

Heat Exchanger Specification sheet							
					Job No.		
Customer	NREL				Ref No.	HP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Amine Precooler				Item No	H-203	
Size 96x 96		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	11541 ft²	Shells/Unit	1	Surface/Shell (Effective)		11541 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		Shellside	Tubeside
Fluid Name		Syngas to Amine Absorber	Cooling Water
Total Fluid Entering	lb/hr	414,200	6,965,000
Vapor		364,537	0
Liquid		49,663	6,965,000
Steam			
Noncondensable			
Fluid Vaporized or Condensed		97,296	0
Liquid Density (In/Out)	lb/ft³	55.608/62.120	62.000/61.573
Liquid Viscosity	cP	0.211	0.627
Liquid Specific Heat	Btu/lb-F	1.063	1.001
Liquid Thermal Conductivity	Btu/hr-ft-F	0.384	0.365
Vapor Mol. Weight (In/Out)		18.8591/18.96	0.0/0.0
Vapor Viscosity	cP	0.0168	0.0000
Vapor Specific Heat	Btu/lb-F	0.424	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.041	0.000
Temperature (In/Out)	°F	332.0/110.0	80.0/100.0
Operating Pressure	psi(Abs)	432.000	65.000
Velocity	ft/sec	13.546	-
Pressure Drop (Allow/Calc)	psi	5.000/1.874	5.000/0.592
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000

Heat Exchanged	139,300,000 Btu/hr		mtd (corr)	98.751 °F
Transfer Rate, Service		122.2	Clean	210.0 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	475/	80/	
Design Temp.	°F	350	150	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-23.0	31.0	
Size &	Out	1-17.0	31.0	
Rating	Intermediate	0	0	

Tube No	8842	OD 0.750 in	Thk 0.065	Length 8.00 ft		Pitch 0.93750 / 30.0°
Tube Type		PLAIN	Material			
Shell		I.D 96.00 OD in	Shell Cover		INT	
Channel or Bonnet			Channel Cover			
Tubesheet-Stationary			Tubesheet-Floating			
Floating Head Cover			Impingement Protection		YES	
Baffles Cross		Type VERT- SEG	%Cut 18.6 (Area)		Spacing-cc	39.8
Baffles-Long			Seal Type			
Supports-Tube		U-Bend		Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint			
Expansion Joint			Type			
Rho-V2 Inlet Nozzle	1,463	Bundle Entrance		1,418	Bundle Exit	3,610
Gasket-Shellside		Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water			Bundle	
Remarks:						

Heat Exchanger Specification sheet

				Job No.		
Customer	NREL			Ref No.	HP Syngas Case	
Address				Proposal No.		
Plant Location				Date		Rev. 0
Service of Unit	ZnO Preheater			Item No	H-320	
Size 96x 96		Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	19400 ft²	Shells/Unit	1	Surface/Shell (Effective)	19400 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		Shellside	Tubeside
Fluid Name		Flue Gas fr. Tar Regen	Sweet Syngas
Total Fluid Entering	lb/hr	280,200	118,500
Vapor		280,200	118,500
Liquid		0	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	0.000/0.000
Liquid Viscosity	cP	0.000	0.000
Liquid Specific Heat	Btu/lb-F	0.000	0.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.000
Vapor Mol. Weight (In/Out)		27.56/27.56	10.99/10.99
Vapor Viscosity	cP	0.0157	0.0182
Vapor Specific Heat	Btu/lb-F	0.312	0.659
Vapor Thermal Conductivity	Btu/hr-ft-F	0.012	0.076
Temperature (In/Out)	°F	839.0/214.0	100.0/750.0
Operating Pressure	psi(Abs)	14.500	422.000
Velocity	ft/sec	64.628	2.701
Pressure Drop (Allow/Calc)	psi	2.000/1.675	5.000/0.488
Fouling resistance	hr-ft²-F/Btu	0.002000	0.002000

Heat Exchanged	49,960,000 Btu/hr	mtd (corr)	96.31 °F
Transfer Rate, Service	26.55	Clean	Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	30/	465/	
Design Temp.	°F	910	800	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections Size & Rating	In	1-53.0	12.0	
	Out	1-47.0	15.0	
	Intermediate	0	0	

Tube No	14190	OD 0.750 in	Thk 0.065	Length 8.00 ft		Pitch 1.25000 / 45.0°
Tube Type		PLAIN	Material			
Shell		I.D 163.00 OD in	Shell Cover		INT	
Channel or Bonnet			Channel Cover			
Tubesheet-Stationary			Tubesheet-Floating			
Floating Head Cover			Impingement Protection	YES		
Baffles Cross		Type VERT- SEG	%Cut 36.0 (Area)	Spacing-cc	65.0	
Baffles-Long			Seal Type			
Supports-Tube		U-Bend		Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint			
Expansion Joint			Type			
Rho-V2 Inlet Nozzle	900	Bundle Entrance	673	Bundle Exit	651	
Gasket-Shellside		Tubeside		Floating Head		
Code Requirement		ASME Section 8, Divsion 1		TEMA Class	R	
Weight/Shell		Filled with Water		Bundle		
Remarks:						

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	HP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	ZnO SG Cooler/BFW Preheater		Item No	H-321	
Size 60x 192	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	5440 ft²	Shells/Unit 1	Surface/Shell (Effective)	5440 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		Shellside	Tubeside
Fluid Name		Syngas fr ZnO Beds	BFW
Total Fluid Entering	lb/hr	118,500	111,600
Vapor		118,500	0
Liquid		0	111,600
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	54.688/45.460
Liquid Viscosity	cP	0.000	0.115
Liquid Specific Heat	Btu/lb-F	0.000	1.429
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.352
Vapor Mol. Weight (In/Out)		10.99/10.99	0.0/0.0
Vapor Viscosity	cP	0.0203	0.0000
Vapor Specific Heat	Btu/lb-F	0.663	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.086	0.000
Temperature (In/Out)	°F	750.0/370.0	349.0/565.0
Operating Pressure	psi(Abs)	412.000	1,285.000
Velocity	ft/sec	30.448	-
Pressure Drop (Allow/Calc)	psi	5.000/3.935	5.000/0.407
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000

Heat Exchanged	29,850,000 Btu/hr	mtd (corr)	75.373 °F
Transfer Rate, Service	72.8	Clean	99.1 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	455/	1,350/	
Design Temp.	°F	800	615	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-15.0	4.0	
Size &	Out	1-13.0	6.0	
Rating	Intermediate	0	0	

Tube No	1902	OD 0.750 in	Thk 0.065	Length 16.00 ft	Pitch 1.25000 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 60.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 14.0 (Area)	Spacing-cc	14.5
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	2,063	Bundle Entrance	272	Bundle Exit	2,203
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet							
					Job No.		
Customer	NREL				Ref No.	HP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Post ZnO Syngas Cooler				Item No	H-322	
Size 36x 96		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	1620 ft²	Shells/Unit	1	Surface/Shell (Effective)		1620 ft²	

PERFORMANCE OF ONE UNIT			
Fluid Allocation		Shellside	Tubeside
Fluid Name		Syngas fr. ZnO Beds	Cooling Water
Total Fluid Entering	lb/hr	118,500	995,500
Vapor		118,500	0
Liquid		0	995,500
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	62.000/62.000
Liquid Viscosity	cP	0.000	0.762
Liquid Specific Heat	Btu/lb-F	0.000	1.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.363
Vapor Mol. Weight (In/Out)		10.99/10.99	0.0/0.0
Vapor Viscosity	cP	0.0148	0.0000
Vapor Specific Heat	Btu/lb-F	0.647	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.065	0.000
Temperature (In/Out)	°F	370.0/110.0	80.0/100.0
Operating Pressure	psi(Abs)	407.000	65.000
Velocity	ft/sec	47.403	-
Pressure Drop (Allow/Calc)	psi	5.000/3.747	5.000/0.585
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000

Heat Exchanged	19,910,000 Btu/hr	mtd (corr)	109.229 °F
Transfer Rate, Service	112.6	Clean	183.2 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL				
		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	450/	80/	
Design Temp.	°F	420	150	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-13.0	12.0	
Size &	Out	1-12.0	12.0	
Rating	Intermediate	0	0	

Tube No	1102	OD 0.750 in	Thk 0.065	Length 8.00 ft		Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material			
Shell	I.D 36.00 OD in		Shell Cover		INT	
Channel or Bonnet			Channel Cover			
Tubesheet-Stationary			Tubesheet-Floating			
Floating Head Cover			Impingement Protection		YES	
Baffles Cross	Type VERT- SEG		%Cut 24.3 (Area)		Spacing-cc	24.0
Baffles-Long			Seal Type			
Supports-Tube		U-Bend		Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint			
Expansion Joint			Type			
Rho-V2 Inlet Nozzle	2,539	Bundle Entrance		1,981	Bundle Exit	3,675
Gasket-Shellside		Tubeside			Floating Head	
Code Requirement	ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell		Filled with Water			Bundle	
Remarks:						

Heat Exchanger Specification sheet						
				Job No.		
Customer	NREL			Ref No.	HP Syngas Case	
Address				Proposal No.		
Plant Location				Date		Rev. 0
Service of Unit	MeOH Compressor Interstage Cooler			Item No	H-400A	
Size 23x 72		Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	476 ft²	Shells/Unit	1	Surface/Shell (Effective)	476 ft²	

PERFORMANCE OF ONE UNIT			
Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Cooling water	Syngas
Total Fluid Entering	lb/hr	537,000	118,500
Vapor		0	118,500
Liquid		537,000	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	62.000/62.000	0.000/0.000
Liquid Viscosity	cP	0.762	0.000
Liquid Specific Heat	Btu/lb-F	1.000	0.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.363	0.000
Vapor Mol. Weight (In/Out)		0.0/0.0	10.99/10.99
Vapor Viscosity	cP	0.0000	0.0155
Vapor Specific Heat	Btu/lb-F	0.000	0.655
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.068
Temperature (In/Out)	°F	80.0/100.0	338.0/200.0
Operating Pressure	psi(Abs)	65.000	1,000.000
Velocity	ft/sec	4.236	25.340
Pressure Drop (Allow/Calc)	psi	5.000/2.578	5.000/0.675
Fouling resistance	hr-ft²-F/Btu	0.002000	0.001000

Heat Exchanged	10,470,000 Btu/hr	mtd (corr)	172.318 °F
Transfer Rate, Service	127.7	Clean	216.4 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL				
		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	80/	1,050/	
Design Temp.	°F	150	390	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-10.0	12.0	
Size &	Out	1-10.0	10.0	
Rating	Intermediate	0	0	

Tube No	442	OD 0.750 in	Thk 0.065	Length 6.00 ft	Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 23.25 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 23.5 (Area)	Spacing-cc	16.3
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	1,206	Bundle Entrance	1,316	Bundle Exit	1,940
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet

				Job No.			
Customer	NREL			Ref No.	HP Syngas Case		
Address				Proposal No.			
Plant Location				Date		Rev. 0	
Service of Unit	MeOH Syngas Preheat			Item No	H-401		
Size 72x 216		Type	BEM - HORZ	Connected in	1 Parallel	1 Series	
Surf/Unit (Eff)	16212 ft²	Shells/Unit	1	Surface/Shell (Effective)	16212 ft²		

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Steam	Syngas to MeOH Rxn
Total Fluid Entering	lb/hr	17,610	118,500
Vapor		17,610	118,500
Liquid		0	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		17,610	0
Liquid Density (In/Out)	lb/ft³	0.000/54.780	0.000/0.000
Liquid Viscosity	cP	0.128	0.000
Liquid Specific Heat	Btu/lb-F	1.157	0.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.393	0.000
Vapor Mol. Weight (In/Out)		18.02/18.02	10.99/10.99
Vapor Viscosity	cP	0.0161	0.0170
Vapor Specific Heat	Btu/lb-F	0.483	0.660
Vapor Thermal Conductivity	Btu/hr-ft-F	0.020	0.074
Temperature (In/Out)	°F	471.7/324.0	240.0/460.0
Operating Pressure	psi(Abs)	100.000	1,165.000
Velocity	ft/sec	4.482	2.118
Pressure Drop (Allow/Calc)	psi	5.000/0.586	5.000/0.430
Fouling resistance	hr-ft²-F/Btu	0.005000	0.001000
Heat Exchanged	17,140,000 Btu/hr	mtd (corr)	45.146 °F
Transfer Rate, Service	23.4	Clean	27.4 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	130/	1,225/	
Design Temp.	°F	545	515	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections Size & Rating	In	1-8.0	10.0	
	Out	1-2.0	12.0	
	Intermediate	0	0	

Tube No	5044	OD 0.750 in	Thk 0.065	Length 18.00 ft	Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 72.00 OD in		Shell Cover	INT	
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	NO	
Baffles Cross	Type VERT- SEG		%Cut 10.2 (Area)	Spacing-cc	14.3
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	1,057	Bundle Entrance	1,398	Bundle Exit	1,158
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet								
					Job No.			
Customer	NREL				Ref No.	HP Syngas Case		
Address					Proposal No.			
Plant Location					Date		Rev. 0	
Service of Unit	Blowdown Cooler				Item No	H-501		
Size 15x 48		Type	BEM - HORZ	Connected in	1 Parallel		1 Series	
Surf/Unit (Eff)	130 ft²	Shells/Unit	1	Surface/Shell (Effective)		130 ft²		

PERFORMANCE OF ONE UNIT			
Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Blowdown	Cooling water
Total Fluid Entering	lb/hr	3,987	41,985
Vapor		0	0
Liquid		3,987	41,985
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	56.607/62.000	62.000/62.000
Liquid Viscosity	cP	0.311	0.762
Liquid Specific Heat	Btu/lb-F	1.059	1.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.382	0.363
Vapor Mol. Weight (In/Out)		0.0/0.0	0.0/0.0
Vapor Viscosity	cP	0.0000	0.0000
Vapor Specific Heat	Btu/lb-F	0.000	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.000
Temperature (In/Out)	°F	298.0/110.0	80.0/100.0
Operating Pressure	psi(Abs)	65.000	65.000
Velocity	ft/sec	0.143	0.528
Pressure Drop (Allow/Calc)	psi	5.000/0.154	5.000/0.206
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000
Heat Exchanged 839,700 Btu/hr		mtd (corr) 89.027 °F	
Transfer Rate, Service 72.7		Clean 97.5 Btu/hr-ft²-F	

CONSTRUCTION OF ONE SHELL				
		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	80/	80/	
Design Temp.	°F	350	150	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-1.0	3.0	
Size &	Out	1-1.0	3.0	
Rating	Intermediate	0	0	

Tube No	170	OD 0.750 in		Thk 0.065	Length 4.00 ft		Pitch 0.93750 / 30.0°
Tube Type		PLAIN		Material			
Shell		I.D 15.25 OD in		Shell Cover		INT	
Channel or Bonnet				Channel Cover			
Tubesheet-Stationary				Tubesheet-Floating			
Floating Head Cover				Impingement Protection		YES	
Baffles Cross		Type VERT- SEG		%Cut 8.6 (Area)		Spacing-cc	3.0
Baffles-Long				Seal Type			
Supports-Tube			U-Bend		Type		
Bypass Seal Arrangement				Tube-Tubesheet Joint			
Expansion Joint				Type			
Rho-V2 Inlet Nozzle		728	Bundle Entrance		9	Bundle Exit	423
Gasket-Shellside			Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell			Filled with Water			Bundle	
Remarks:							

COMPRESSOR NUMBER			K-100			
SERVICE			Combustion Air			
GAS HANDLED			Air			
NORMAL FLOW	SCFM		58,597			
NORMAL FLOW	LB/HR		265,200			
DESIGN FLOW	SCFM					
MOL WT.			28.63			
C_p/C_v	Value		1.4			
	@ F / PSIA		90 / 14.7			
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		14.7			
COMPR. FACTOR @ SUCTION			0.999			
FLOW AT SUCTION	ACFM		61,910			
ORIGIN	PSIA					
TEMPERATURE	F		90			
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		20			
DISCH. TEMPERATURE	F	(2)	157			
COMPR. FACTOR @ DISCH.			0.999			
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			1.36			
EFFICIENCY		(2)	0.75			
BHP		(2)	1800			
COMPRESSOR TYPE						
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
	H_2O		3.1			
	O_2		20.3			
	Ar		0.9			
	N_2		75.7			

(1) INCLUDES ALLOWANCE FOR SUCTION OR DISCHARGE SNUBBER
(2) VALUE TABULATED IS ESTIMATED AND MUST BE VERIFIED BY FINAL MECHANICAL DESIGN

NO	DATE	REVISIONS	PROC	PROJ.	CLIENT	
		NREL BIOMASS GASIFICATION: High Pressure Syngas Case (GTI Gasifier)				JOB NO: NREL Contract ACO-5-44027; DRAWING NO; REV

COMPRESSOR NUMBER			K-320			
SERVICE			Flue Gas Blower			
GAS HANDLED			Flue Gas			
NORMAL FLOW	SCFM		64,194			
NORMAL FLOW	LB/HR		279,800			
DESIGN FLOW	SCFM					
MOL WT.			27.58			
C_p/C_v	Value		1.365			
	@ F / PSIA		202.5 / 14.3			
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		14.3			
COMPR. FACTOR @ SUCTION			0.9985			
FLOW AT SUCTION	ACFM		85,400			
ORIGIN	PSIA					
TEMPERATURE	F		214			
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		14.7			
DISCH. TEMPERATURE	F	(2)	221			
COMPR. FACTOR @ DISCH.			0.9985			
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			1.03			
EFFICIENCY		(2)	0.75			
BHP		(2)	207			
COMPRESSOR TYPE						
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
	CO_2		14.33			
	H_2O		10.93			
	O_2		1.03			
	Ar		0.73			
	N_2		72.98			

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NO	DATE	REVISIONS	PROC	PROJ.	CLIENT	
		NREL BIOMASS GASIFICATION: High Pressure Syngas Case (GTI Gasifier)				JOB NO: NREL Contract ACO-5-44027
						DRAWING NO / REV

COMPRESSOR NUMBER			K-400A	K-400B		
SERVICE			MeOH Comp-1	MeOH Comp-2		
GAS HANDLED			Treated Syngas	Treated Syngas		
NORMAL FLOW	SCFM		68,247	68,247		
NORMAL FLOW	LB/HR		118,500	118,500		
DESIGN FLOW	SCFM					
MOL WT.			10.99	10.99		
C_p/C_v	Value		1.418	1.423		
	@ F / PSIA		110 / 402	200 / 995		
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		402	995		
COMPR. FACTOR @ SUCTION			1.006	1.021		
FLOW AT SUCTION	ACFM		1,567	1,306		
ORIGIN	PSIA					
TEMPERATURE	F		110	200		
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		1,000	1,165		
DISCH. TEMPERATURE	F	(2)	334.8	240.4		
COMPR. FACTOR @ DISCH.			1.022	1.026		
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			2.49	1.17		
EFFICIENCY		(2)	0.75	0.75		
BHP		(2)	7,102	1,286		
COMPRESSOR TYPE						
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
		H_2	65.10	65.1		
		CO_2	1.50	1.5		
		CO	30.08	30.08		
		H_20	0.27	0.27		
		CH_4	2.70	2.7		
		C_2H_2				
		C_2H_4	0.02	0.02		
		C_2H_6	0.00003	0.00003		
		Benzene (C_6H_6)	0.00005	0.00005		
		Tar ($C_{10}H_8$)	0.000001	0.000001		
		NH_3	0.00002	0.00002		
		N_2	0.08	0.08		

(1) INCLUDES ALLOWANCE FOR SUCTION OR DISCHARGE SNUBBER
(2) VALUE TABULATED IS ESTIMATED AND MUST BE VERIFIED BY FINAL MECHANICAL DESIGN

NO	DATE	REVISIONS	PROC	PROJ.	CLIENT		
		NREL BIOMASS GASIFICATION: High Pressure Syngas Case (GTI Gasifier)				JOB NO	NREL Contract ACO-5-44027
						DRAWING NO	REV

COMPRESSOR NUMBER			M-501A	M-501B		
SERVICE			Steam Turbine - Extraction Stage 1	Steam Turbine - Extraction Stage 2		
GAS HANDLED			Steam	Steam		
NORMAL FLOW	SCFM		110,138	51,979		
NORMAL FLOW	LB/HR		313,600	148,100		
DESIGN FLOW	SCFM					
MOL WT.			18.02	18.02		
C_p/C_v	Value		1.384	1.353		
	@ F / PSIA		1000 / 1260	758 / 460		
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		1260	460		
COMPR. FACTOR @ SUCTION			0.9334	0.9521		
FLOW AT SUCTION	ACFM		3,369	3,709		
ORIGIN	PSIA					
TEMPERATURE	F		1000	758		
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		460	100		
DISCH. TEMPERATURE	F	(2)	758	472		
COMPR. FACTOR @ DISCH.			0.9521	0.974		
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			-	-		
EFFICIENCY		(2)	0.75	0.75		
kW Generated		(2)	9,341	5,371		
Turbine TYPE			Steam	Steam		
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
		H_2				
		CO_2				
		CO				
		H_20	100%	100%		
		CH_4				
		C_2H_2				
		C_2H_4				
		C_2H_6				
		Benzene (C_6H_6)				
		Tar ($C_{10}H_8$)				
		NH_3				
		N_2				

(1) INCLUDES ALLOWANCE FOR SUCTION OR DISCHARGE SNUBBER
(2) VALUE TABULATED IS ESTIMATED AND MUST BE VERIFIED BY FINAL MECHANICAL DESIGN

NO	DATE	REVISIONS	PROC	PROJ.	CLIENT	
		NREL BIOMASS GASIFICATION: High Pressure Syngas Case (GTI Gasifier)				JOB NO: NREL Contract ACO-5-44027; DRAWING NO; REV

Cyclone Specification Sheet

Site Location		Date			Rev.

SERVICE OF HIGH PRESSURE UNIT S-100

Inlet Conditions	Flow	Viscosity	Density	Molecular Weight (Ave.)	Particle Size (mm) (Stokes' MMD)	Volumetric Flowrate	Temperature
	lb/h	lb/ft-sec	lb/ft3	lb/mole		acfm	°F
Gas	418,416.00	2.54E-05	0.47800	21.5		14,589.00	1,576
Particulate	9,440.00		62.40		60		

Gas Inlet Pressure (psia)	460.00
Gas Discharge Pressure (psig)	455.57
Pressure Drop, Max Allow. (" .WC.)	120.00
Design/Test Pressure Psig	460.00
Design Particulate Cutpoint	50
Design Separation Efficiency at Cutpoint (%)	98

Emery Design Calculations Summary for S-100 (for Reference Only)

Mechanical Sizing		Inside Diam (in)	Uninsulated Outside Diam (in)		ID (in)	OD (in)	Thickness (in)	Designation		Overall Height (ft)
Connections Size & Rating	In	32	42	Upper Shell	58	60	1	ASME VIII		25
	Out	24	34	Inner Tube	24	26	1			
	Bottom			Cone			1	ASME VIII		
				Refractory	50		4			

Component Data					Cyclone Body Materials of Construction					
	Design Temperature (°F)	Solids Removal Flowrate (CFM)	Differential Design Pressure (psig)	Type	Upper Section		Lower Conical section		Nozzles	
Rotary Air Lock	1598				Inner Wall	Outer Shell	Inner Wall	Outer Shell	Inner Wall	Outer Shell
Level Indicator	1598				Cercast™	MS	Cercast™	MS	Cercast™	MS
					Inner Tube					
					MS					

Vendor/Supplier Specifications and Price Quote

Fisher-Klosterman, Inc
Ryan Bruner, Sales Manager
P.O. Box 11190
Lousville, KY
Ph: 502-572-4000 ext 213
Email: rab@fkinc.com

(Refer to Vendor Communications and Data Sheets)

Recommendation: Replace S-100 and S-101 with one (1) cyclone only:

One (1) cyclone (XQ120-30M) with the following features:

Design, fabricated, tested, and stamped as an ASME vessel		Interior surfaces to be lined with 4" of Vesuvius Cercast 3300 castable refractory	
1-1/4" plate carbon steel construction		All welding per FKI Class 3 preocedures with 100% penetration	
Dust receiver section with flanged discharge		Exterior to be sandblasted and painted with high temperature aluminum paint	
Inlet transition to 24"∅ gas inlet flange		Design pressure (psig)	460
30"∅ verticle gas outlet flange		Design Temperature (F)	650
Approximate Overall Dimensions:	5 ft∅ x 25 ft tall		

Gas Conditions at Inlet:		Particulate Conditions at Inlet:	
Volume per cylone (acfm)	14,589	Specific Gravity	1.000
Density (lbm/ft3)	0.478	Dust Loading (Grains/acf)	31.3
Viscosity (lbm/ft-sec)	2.54E-05		

Inlet Velocity (ft/sec)	68.39
No load pres. drop (in.W.C.)	106.35
Full load pres. Drop (in. W.C.)	85.46

Fraction Efficiencies: Stokes Equiv. % Efficiency

Dia.(microns)	Weight %
2.5	6.11
3	15.75
3.5	21.47
4	27.4
4.5	33.3
5	39.04
6	44.49
6.5	49.6
7.5	58.71
8.5	66.32
9.5	72.57
12	83.53
16	89.99
23	95.08
33	97.84

Price (2005 U.S.$) **$ 355,000.00**

Remarks: Inlet and outlet manifolding is not included in Fisher-Klosterman quote for these four cylones. Estimated cost of splitter and collection is $25,000.
Refer to supplier data sheet for Vesuvius CERCAST™ 3300 Castable refractory.

Cyclone Specification Sheet

Site Location		Date			Rev.

SERVICE OF HIGH PRESSURE UNIT S-102

Inlet Conditions	Flow	Viscosity	Density	Molecular Weight (Ave.)	Particle Size (mm) (Stokes' MMD)	Volumetric Flowrate	Temperature
	lb/h	lb/ft-sec	lb/ft3	lb/mole		acfm	°F
Gas	434,982.00		0.38390	27.6		18,883.00	1,576
Particulate	9,440.00		62.40		60		
Gas Inlet Pressure (psia)	460.00						
Gas Discharge Pressure (psig)	455.57						
Pressure Drop, Max Allow. (" .WC.)	120.00						
Design/Test Pressure Psig	460.00						
Design Particulate Cutpoint	50						
Design Separation Efficiency at Cutpoint (%)	98						

Emery Design Calculations Summary for S-102 (for Reference Only)

Mechanical Sizing		Inside Diam (in)	Uninsulated Outside Diam (in)		ID (in)	OD (in)	Thickness (in)	Designation	Height (In)	Height (ft)
Connections Size & Rating	In	32.0769	42.10	Upper Shell			1	ASME VIII	160	13.4
	Out			Inner Tube	32.10		4			
	Bottom			Cone			1	ASME VIII		
				Refractory			4			

Component Data					Cyclone Body Materials of Construction					
	Design Temperature (°F)	Solids Removal Flowrate (CFM)	Differential Design Pressure (psig)	Type	Upper Section		Lower Conical section		Nozzles	
Rotary Air Lock	1598	20.4	15		Inner Wall	Outer Shell	Inner Wall	Outer Shell	Inner Wall	Outer Shell
Level Indicator	1598				Cercast™	MS	Cercast™	MS	Cercast™	MS
					Inner Tube					
					MS					

Vendor/Supplier Specifications and Price Quote

Fisher-Klosterman, Inc (Refer to Vendor Communications and Data Sheets)
Ryan Bruner, Sales Manager
P.O. Box 11190
Lousville, KY
Ph: 502-572-4000 ext 213
Email: rab@fkinc.com

Recommendation:

One (1) cyclone (XQ120-30M) with the following features:

Design, fabricated, tested, and stamped as an ASME vessel		Interior surfaces to be lined with 4" of Vesuvius Cercast 3300 castable refractory	
1-1/4" plate carbon steel construction		All welding per FKI Class 3 preocedures with 100% penetration	
Dust receiver section with flanged discharge		Exterior to be sandblasted and painted with high temperature aluminum paint	
Inlet transition to 24"∅ gas inlet flange		Design pressure (psig)	460
30"∅ verticle gas outlet flange		Design Temperature (F)	650
Approximate Overall Dimensions:	5-1/2 ft∅ x 27 1/2 ft tall		

Gas Conditions at Inlet:		Particulate Conditions at Inlet:	
Volume per cylone (acfm)	18,883	Specific Gravity	1.000
Density (lbm/ft3)	0.3839	Dust Loading (Grains/acf)	6.97
Viscosity (lbm/ft-sec)	2.78E-05		

Inlet Velocity (ft/sec)	69.94
No load pres. drop (in.W.C.)	83.63
Full load pres. Drop (in. W.C.)	72.52

Fraction Efficiencies: Stokes Equiv. % Efficiency

Dia.(microns)	Weight %
3	7.64
4	16.37
4.5	21.33
5	26.43
5.5	31.53
6	36.52
7	45.88
8	54.2
9	61.4
10	67.53
11	72.7
14	83.68
18	89.34
25	94.31
35	97.29
80	99.72

Price (2005 U.S.$) | **$ 410,000.00**

Remarks: Inlet and outlet manifolding is not included in Fisher-Klosterman quote for these four cylones. Estimated cost of splitter and collection is $25,000.
Refer to supplier data sheet for Vesuvius CERCAST^{TM} 3300 Castable refractory.

Cyclone Specification Sheet

Site Location		Date			Rev.

SERVICE OF HIGH PRESSURE UNIT S-103

Inlet Conditions	Flow	Specific Heat	Density	Molecular Weight (Ave.)	Particle Size (mm) (Stokes' MMD)	Volumetric Flowrate	Temperature
	lb/h	BTU/lb°F	lb/ft3	lb/mole		acfm	°F
Gas	434,982.00		0.41421	20.14507		16,835.82	1576.0
Particulate	9,440.00		33.00		60		
Gas Inlet Pressure (psia)	460.00						
Gas Discharge Pressure (psig)	455.57						
Pressure Drop, Max Allow. (" .WC.)	120.00						
Design/Test Pressure Psig	460.00						
Design Particulate Cutpoint	50						
Design Separation Efficiency at Cutpoint (%)	98						

Emery Design Calculations Summary for S-103 (for Reference Only)

Mechanical Sizing		Inside Diam (in)	Uninsulated Outside Diam (in)		ID (in)	OD (in)	Thickness (in)	Designation	Height (In)	Height (ft)
Connections Size & Rating	In	32.0769	42.10	Upper Shell			1	ASME VIII	160	13.4
	Out			Inner Tube	32.10		4			
	Bottom			Cone			1	ASME VIII		
				Refractory			4			

Component Data	Design Temperature (°F)	Solids Removal Flowrate (CFM)	Differential Design Pressure (psig)	Type
Rotary Air Lock	1598	20.4	15	
Level Indicator	1598			

Cyclone Body Materials of Construction

Upper Section		Lower Conical section		Nozzles	
Inner Wall	Outer Shell	Inner Wall	Outer Shell	Inner Wall	Outer Shell
Cercast™	MS	Cercast™	MS	Cercast™	MS
Inner Tube					
MS					

Vendor/Supplier Specifications and Price Quote

Fisher-Klosterman, Inc
Ryan Bruner, Sales Manager
P.O. Box 11190
Lousville, KY
Ph: 502-572-4000 ext 213
Email: rab@fkinc.com

(Refer to Vendor Communications and Data Sheets)

Recommendation:

One (1) cyclone (XQ120-30M) with the following features:

- Design, fabricated, tested, and stamped as an ASME vessel
- 1-1/4" plate carbon steel construction
- Dust receiver section with flanged discharge
- Inlet transition to 24"∅ gas inlet flange
- 30"∅ verticle gas outlet flange
- Approximate Overall Dimensions: 4 ft∅ x 18 ft tall
- Interior surfaces to be lined with 4" of Vesuvius Cercast 3300 castable refractory
- All welding per FKI Class 3 preocedures with 100% penetration
- Exterior to be sandblasted and painted with high temperature aluminum paint
- Design pressure (psig): 460
- Design Temperature (F): 650

Gas Conditions at Inlet:	
Volume per cylone (acfm)	8,223
Density (lbm/ft3)	0.5679
Viscosity (lbm/ft-sec)	2.87E-05
Inlet Velocity (ft/sec)	68.53
No load pres. drop (in.W.C.)	103.76
Full load pres. Drop (in. W.C.)	85.86

Particulate Conditions at Inlet:	
Specific Gravity	1.000
Dust Loading (Grains/acf)	16

Fraction Efficiencies: Stokes Equiv. % Efficiency

Dia.(microns)	Weight %
2.5	6.71
3.5	15.89
4	21.16
4.5	26.55
5	31.91
5.5	37.11
6	42.08
7	51.14
8	58.96
9	65.59
10	71.14
13	82.8
17	89.12
24	94.37
34	97.41
89	99.83

Price (2005 U.S.$) | $ 265,000.00

Remarks: Inlet and outlet manifolding is not included in Fisher-Klosterman quote for these four cylones. Estimated cost of splitter and collection is $25,000.
Refer to supplier data sheet for Vesuvius CERCAST™ 3300 Castable refractory.

DATA SHEETS, LOW PRESSURE DESIGN

Heat Exchanger Specification sheet

					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Tar Reformer SG Cooling/Steam Generator				Item No	H-100 Tar Ref Cooler	
Size 72x 168		Type	BEM - HORZ	Connected in	2 Parallel		1 Series
Surf/Unit (Eff)	10708 ft^2	Shells/Unit	2	Surface/Shell (Effective)		5354 ft^2	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas fr Tar Reformer	Preheated BFW
Total Fluid Entering	lb/hr	329,000	251,800
Vapor		329,000	0
Liquid		0	251,800
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	251,800
Liquid Density (In/Out)	lb/ft^3	0.000/0.000	46.533/45.419
Liquid Viscosity	cP	0.000	0.092
Liquid Specific Heat	Btu/lb-F	0.000	1.636
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.321
Vapor Mol. Weight (In/Out)		16.74/16.74	0.0/18.02
Vapor Viscosity	cP	0.0280	0.0200
Vapor Specific Heat	Btu/lb-F	0.520	0.774
Vapor Thermal Conductivity	Btu/hr-ft-F	0.078	0.025
Temperature (In/Out)	°F	1,598.0/624.0	546.5/575.0
Operating Pressure	psi(Abs)	29.900	1,285.000
Velocity	ft/sec	280.241	7.682
Pressure Drop (Allow/Calc)	psi	5.000/3.920	5.000/0.977
Fouling resistance	hr-ft^2-F/Btu	0.001000	0.005000

Heat Exchanged	167,000,000	Btu/hr	mtd (corr)	318.656 °F	
Transfer Rate, Service		48.9	Clean	80.8 Btu/hr-ft^2-F	

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	45/	1,350/	
Design Temp.	°F	1700	625	
No. Passes per Shell		1	6	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-33.0	6.0	
Size &	Out	1-29.0	10.0	
Rating	Intermediate	0	0	

Tube No	1664	OD 1.000 in		Thk 0.065	Length 14.00 ft		Pitch 1.25000 / 30.0°
Tube Type		PLAIN		Material			
Shell		I.D 72.00 OD in		Shell Cover		INT	
Channel or Bonnet				Channel Cover			
Tubesheet-Stationary				Tubesheet-Floating			
Floating Head Cover				Impingement Protection		YES	
Baffles Cross		Type VERT- SEG		%Cut 34.7 (Area)		Spacing-cc	73.7
Baffles-Long				Seal Type			
Supports-Tube			U-Bend		Type		
Bypass Seal Arrangement				Tube-Tubesheet Joint			
Expansion Joint				Type			
Rho-V2 Inlet Nozzle		2,611	Bundle Entrance		3,399	Bundle Exit	4,375
Gasket-Shellside			Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell			Filled with Water			Bundle	
Remarks:							

Heat Exchanger Specification sheet

				Job No.		
Customer	NREL			Ref No.	LP Syngas Case	
Address				Proposal No.		
Plant Location				Date		Rev. 0
Service of Unit	Tar Reformer SG Cooling/BFW Preheat			Item No	H-101 Tar Ref Cooler	
Size 57x 168	Type	BEM - HORZ	Connected in	2 Parallel		1 Series
Surf/Unit (Eff)	13334 ft²	Shells/Unit	2	Surface/Shell (Effective)	6667 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas fr Tar Reformer	BFW
Total Fluid Entering	lb/hr	329,000	142,594
Vapor		329,000	0
Liquid		0	142,594
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	58.509/46.533
Liquid Viscosity	cP	0.000	0.139
Liquid Specific Heat	Btu/lb-F	0.000	1.340
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.359
Vapor Mol. Weight (In/Out)		16.74/16.74	0.0/0.0
Vapor Viscosity	cP	0.0189	0.0000
Vapor Specific Heat	Btu/lb-F	0.475	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.049	0.000
Temperature (In/Out)	°F	624.0/300.0	240.0/546.5
Operating Pressure	psi(Abs)	26.900	1,295.000
Velocity	ft/sec	234.572	-
Pressure Drop (Allow/Calc)	psi	5.000/4.568	5.000/0.513
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000

Heat Exchanged	50,610,000 Btu/hr	mtd (corr)	68.377 °F
Transfer Rate, Service	55.5	Clean	68.5 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	35/	1,360/	
Design Temp.	°F	675	600	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-29.0	3.0	
Size &	Out	1-29.0	4.0	
Rating	Intermediate	0	0	

Tube No	2688	OD 0.750 in	Thk 0.065	Length 14.00 ft	Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 57.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 37.1 (Area)	Spacing-cc	75.5
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	2,563	Bundle Entrance	2,914	Bundle Exit	3,741
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	LP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	Tar Reformer SG Cooler/Deaerator FW Preheat		Item No	H-102 Tar Ref Cooler	
Size 75x 168	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	5621 ft^2	Shells/Unit	1	Surface/Shell (Effective)	5621 ft^2

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas fr H-101	Deaerator Feed Water
Total Fluid Entering	lb/hr	329,000	257,000
Vapor		329,000	0
Liquid		0	257,000
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft^3	0.000/0.000	59.592/58.402
Liquid Viscosity	cP	0.000	0.273
Liquid Specific Heat	Btu/lb-F	0.000	1.054
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.392
Vapor Mol. Weight (In/Out)		16.74/16.74	0.0/0.0
Vapor Viscosity	cP	0.0156	0.0000
Vapor Specific Heat	Btu/lb-F	0.461	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.040	0.000
Temperature (In/Out)	°F	300.0/225.0	195.0/237.0
Operating Pressure	psi(Abs)	23.880	30.000
Velocity	ft/sec	168.427	-
Pressure Drop (Allow/Calc)	psi	5.000/2.790	5.000/0.489
Fouling resistance	hr-ft^2-F/Btu	0.001000	0.002000

Heat Exchanged	11,340,000 Btu/hr	mtd (corr)	44.478 °F
Transfer Rate, Service	45.4	Clean	58.8 Btu/hr-ft^2-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	35/	45/	
Design Temp.	°F	350	280	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-41.0	6.0	
Size &	Out	1-42.0	6.0	
Rating	Intermediate	0	0	

Tube No	2096	OD 0.750 in	Thk 0.065	Length 14.00 ft	Pitch 1.25000 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 75.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 41.0 (Area)	Spacing-cc	81.9
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	2,021	Bundle Entrance	1,271	Bundle Exit	2,155
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet							
					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Flue Gas Cooler/Steam Superheater				Item No	H-103	
Size 90x 168		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	5770 ft²	Shells/Unit	1	Surface/Shell (Effective)		5770 ft²	
PERFORMANCE OF ONE UNIT							
Fluid Allocation			**Shellside**			**Tubeside**	
Fluid Name			Flue Gas fr. Tar Regen			Superheated Steam	
Total Fluid Entering		lb/hr	248,400			251,800	
Vapor			248,400			251,800	
Liquid			0			0	
Steam							
Noncondensable							
Fluid Vaporized or Condensed			0			0	
Liquid Density (In/Out)		lb/ft³	0.000/0.000			0.000/0.000	
Liquid Viscosity		cP	0.000			0.000	
Liquid Specific Heat		Btu/lb-F	0.000			0.000	
Liquid Thermal Conductivity		Btu/hr-ft-F	0.000			0.000	
Vapor Mol. Weight (In/Out)			27.57/27.57			18.02/18.02	
Vapor Viscosity		cP	0.0405			0.0254	
Vapor Specific Heat		Btu/lb-F	0.313			0.678	
Vapor Thermal Conductivity		Btu/hr-ft-F	0.040			0.036	
Temperature (In/Out)		°F	1,798.0/935.0			575.0/1,000.0	
Operating Pressure		psi(Abs)	14.700			1,275.000	
Velocity		ft/sec	215.255			5.762	
Pressure Drop (Allow/Calc)		psi	2.000/1.727			5.000/0.629	
Fouling resistance		hr-ft²-F/Btu	0.001000			0.005000	
Heat Exchanged	67,260,000 Btu/hr			mtd (corr)	550.248 °F		
Transfer Rate, Service		21.2		Clean	25.0 Btu/hr-ft²-F		
CONSTRUCTION OF ONE SHELL							
		Shellside	**Tubeside**			**Sketch**	
Design/Test Pres.	psi	30/	1,350/				
Design Temp.	°F	1900	1100				
No. Passes per Shell		1	1				
Corrosion Allow.	in	0.0625	0.0625				
Connections Size & Rating	In	1-57.0	12.0				
	Out	1-53.0	15.0				
	Intermediate	0	0				
Tube No	2475	OD 0.750 in		Thk 0.065	Length 14.00 ft		Pitch 1.25000 / 45.0°
Tube Type		PLAIN		Material			
Shell		I.D 90.00 OD in		Shell Cover		INT	
Channel or Bonnet				Channel Cover			
Tubesheet-Stationary				Tubesheet-Floating			
Floating Head Cover				Impingement Protection		YES	
Baffles Cross		Type VERT- SEG		%Cut 38.4 (Area)		Spacing-cc	71.2
Baffles-Long				Seal Type			
Supports-Tube			U-Bend		Type		
Bypass Seal Arrangement				Tube-Tubesheet Joint			
Expansion Joint				Type			
Rho-V2 Inlet Nozzle		906	Bundle Entrance		2,230	Bundle Exit	1,030
Gasket-Shellside			Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell			Filled with Water			Bundle	
Remarks:							

Heat Exchanger Specification sheet

				Job No.		
Customer	NREL			Ref No.	LP Syngas Case	
Address				Proposal No.		
Plant Location				Date		Rev. 0
Service of Unit	NREL Biomass			Item No	H-200 Quech Water Cooler	
Size 71x 120	Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	9232 ft²	Shells/Unit	1	Surface/Shell (Effective)	9232 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Quench Water	Cooling Water
Total Fluid Entering	lb/hr	1,189,000	1,107,500
Vapor		0	0
Liquid		1,189,000	1,107,500
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	61.342/61.765	62.470/62.000
Liquid Viscosity	cP	0.578	0.744
Liquid Specific Heat	Btu/lb-F	1.003	0.998
Liquid Thermal Conductivity	Btu/hr-ft-F	0.366	0.361
Vapor Mol. Weight (In/Out)		0.0/0.0	0.0/0.0
Vapor Viscosity	cP	0.0000	0.0000
Vapor Specific Heat	Btu/lb-F	0.000	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.000
Temperature (In/Out)	°F	128.0/110.0	80.0/100.0
Operating Pressure	psi(Abs)	26.000	20.000
Velocity	ft/sec	1.924	-
Pressure Drop (Allow/Calc)	psi	5.000/1.722	5.000/0.549
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000

Heat Exchanged	22,150,000 Btu/hr	mtd (corr)	28.989 °F
Transfer Rate, Service	82.8	Clean	115.4 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	45/	45/	
Design Temp.	°F	215	150	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-13.0	12.0	
Size &	Out	1-13.0	12.0	
Rating	Intermediate	0	0	

Tube No	4860	OD 0.750 in	Thk 0.065	Length 10.00 ft	Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 71.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 9.2 (Area)	Spacing-cc	14.1
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	2,093	Bundle Entrance	913	Bundle Exit	2,867
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet							
					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Compressor Interstage Cooling				Item No	H-300A	
Size 82x 144		Type	BEM - HORZ	Connected in	2 Parallel		1 Series
Surf/Unit (Eff)	28471 ft²	Shells/Unit	2	Surface/Shell (Effective)		14235 ft²	

PERFORMANCE OF ONE UNIT			
Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Cooling water	1st Stage Syngas
Total Fluid Entering	lb/hr	6,100,000	317,400
Vapor		0	317,400
Liquid		6,100,000	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	85,698
Liquid Density (In/Out)	lb/ft³	62.000/62.000	0.000/62.020
Liquid Viscosity	cP	0.762	0.432
Liquid Specific Heat	Btu/lb-F	1.000	1.035
Liquid Thermal Conductivity	Btu/hr-ft-F	0.363	0.380
Vapor Mol. Weight (In/Out)		0.0/0.0	16.7/16.7
Vapor Viscosity	cP	0.0000	0.0157
Vapor Specific Heat	Btu/lb-F	0.000	0.460
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.043
Temperature (In/Out)	°F	80.0/100.0	344.0/110.0
Operating Pressure	psi(Abs)	65.000	35.000
Velocity	ft/sec	3.977	39.521
Pressure Drop (Allow/Calc)	psi	5.000/4.889	5.000/0.642
Fouling resistance	hr-ft²-F/Btu	0.002000	0.001000

Heat Exchanged	122,000,000 Btu/hr		mtd (corr)	80.189 °F
Transfer Rate, Service		53.4	Clean	64.5 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL				
		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	80/	50/	
Design Temp.	°F	150	400	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-23.0	25.0	
Size &	Out	1-23.0	23.0	
Rating	Intermediate	0	0	

Tube No	6298	OD 0.750 in		Thk 0.065	Length 12.00 ft	Pitch 0.93750 / 30.0°
Tube Type		PLAIN		Material		
Shell		I.D 82.00 OD in		Shell Cover		INT
Channel or Bonnet				Channel Cover		
Tubesheet-Stationary				Tubesheet-Floating		
Floating Head Cover				Impingement Protection		YES
Baffles Cross		Type VERT- SEG		%Cut 12.4 (Area)		Spacing-cc 24.0
Baffles-Long				Seal Type		
Supports-Tube			U-Bend		Type	
Bypass Seal Arrangement				Tube-Tubesheet Joint		
Expansion Joint				Type		
Rho-V2 Inlet Nozzle		1,391	Bundle Entrance		1,525	Bundle Exit 2,034
Gasket-Shellside			Tubeside			Floating Head
Code Requirement		ASME Section 8, Divsion 1				TEMA Class R
Weight/Shell			Filled with Water			Bundle
Remarks:						

Heat Exchanger Specification sheet

					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Compressor Interstage Cooling				Item No	H-300B	
Size 47x 120		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	3435 ft²	Shells/Unit	1	Surface/Shell (Effective)		3435 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		Shellside	Tubeside
Fluid Name		2nd Stage Syngas	Cooling water
Total Fluid Entering	lb/hr	232,600	1,639,500
Vapor		232,600	0
Liquid		0	1,639,500
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	62.000/62.000
Liquid Viscosity	cP	0.000	0.762
Liquid Specific Heat	Btu/lb-F	0.000	1.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.363
Vapor Mol. Weight (In/Out)		16.26/16.26	0.0/0.0
Vapor Viscosity	cP	0.0162	0.0000
Vapor Specific Heat	Btu/lb-F	0.470	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.050	0.000
Temperature (In/Out)	°F	350.0/110.0	80.0/100.0
Operating Pressure	psi(Abs)	84.000	65.000
Velocity	ft/sec	119.731	1.938
Pressure Drop (Allow/Calc)	psi	5.000/3.994	5.000/0.664
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000

Heat Exchanged	32,790,000 Btu/hr		mtd (corr)	103.761 °F
Transfer Rate, Service		92.0	Clean	134.2 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	100/	80/	
Design Temp.	°F	400	150	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-25.0	15.0	
Size &	Out	1-23.0	15.0	
Rating	Intermediate	0	0	

Tube No	1808	OD 0.750 in		Thk 0.065	Length 10.00 ft		Pitch 0.93750 / 30.0°
Tube Type		PLAIN		Material			
Shell		I.D 47.00 OD in		Shell Cover		INT	
Channel or Bonnet				Channel Cover			
Tubesheet-Stationary				Tubesheet-Floating			
Floating Head Cover				Impingement Protection		YES	
Baffles Cross		Type VERT- SEG		%Cut 37.2 (Area)		Spacing-cc	58.0
Baffles-Long				Seal Type			
Supports-Tube			U-Bend		Type		
Bypass Seal Arrangement				Tube-Tubesheet Joint			
Expansion Joint				Type			
Rho-V2 Inlet Nozzle		2,286	Bundle Entrance		3,535	Bundle Exit	3,995
Gasket-Shellside			Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell			Filled with Water			Bundle	
Remarks:							

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	LP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	Compressor Interstage Cooling		Item No	H-300C	
Size 51x 120	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	4368 ft²	Shells/Unit 1	Surface/Shell (Effective)	4368 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Cooling water	3rd Stage Syngas
Total Fluid Entering	lb/hr	1,384,500	225,800
Vapor		0	225,800
Liquid		1,384,500	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	2,710
Liquid Density (In/Out)	lb/ft³	62.000/62.000	0.000/62.250
Liquid Viscosity	cP	0.762	0.558
Liquid Specific Heat	Btu/lb-F	1.000	1.038
Liquid Thermal Conductivity	Btu/hr-ft-F	0.363	0.368
Vapor Mol. Weight (In/Out)		0.0/0.0	16.21/16.21
Vapor Viscosity	cP	0.0000	0.0164
Vapor Specific Heat	Btu/lb-F	0.000	0.468
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.051
Temperature (In/Out)	°F	80.0/100.0	349.0/110.0
Operating Pressure	psi(Abs)	65.000	220.000
Velocity	ft/sec	3.395	26.531
Pressure Drop (Allow/Calc)	psi	5.000/3.694	5.000/0.747
Fouling resistance	hr-ft²-F/Btu	0.002000	0.001000

Heat Exchanged	27,690,000 Btu/hr	mtd (corr)	92.157 °F
Transfer Rate, Service	68.8	Clean	88.3 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	80/	245/	
Design Temp.	°F	150	400	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-15.0	19.0	
Size &	Out	1-15.0	17.0	
Rating	Intermediate	0	0	

Tube No	2350	OD 0.750 in	Thk 0.065	Length 10.00 ft	Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 51.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 18.9 (Area)	Spacing-cc	24.0
Baffles-Long			Seal Type		
Supports-Tube		U-Bend	Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	1,584	Bundle Entrance	1,413	Bundle Exit	2,336
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet

					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Compressor Interstage Cooling				Item No	H-300D	
Size 42x 120		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	2934 ft²	Shells/Unit	1	Surface/Shell (Effective)		2934 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		Shellside	Tubeside
Fluid Name		Cooling water	4th Stage Syngas
Total Fluid Entering	lb/hr	910,500	223,200
Vapor		0	223,200
Liquid		910,500	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	670
Liquid Density (In/Out)	lb/ft³	62.000/62.000	0.000/62.210
Liquid Viscosity	cP	0.762	0.580
Liquid Specific Heat	Btu/lb-F	1.000	1.036
Liquid Thermal Conductivity	Btu/hr-ft-F	0.363	0.367
Vapor Mol. Weight (In/Out)		0.0/0.0	16.2/16.2
Vapor Viscosity	cP	0.0000	0.0160
Vapor Specific Heat	Btu/lb-F	0.000	0.470
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.049
Temperature (In/Out)	°F	80.0/100.0	277.0/110.0
Operating Pressure	psi(Abs)	65.000	450.000
Velocity	ft/sec	3.281	17.909
Pressure Drop (Allow/Calc)	psi	5.000/3.891	5.000/0.750
Fouling resistance	hr-ft²-F/Btu	0.002000	0.001000

Heat Exchanged	18,210,000 Btu/hr		mtd (corr)	79.340 °F
Transfer Rate, Service		78.2	Clean	104.5 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	80/	500/	
Design Temp.	°F	150	330	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-12.0	15.0	
Size &	Out	1-12.0	15.0	
Rating	Intermediate	0	0	

Tube No	1594	OD 0.750 in		Thk 0.065	Length 10.00 ft		Pitch 0.93750 / 30.0°
Tube Type		PLAIN		Material			
Shell		I.D 42.00 OD in		Shell Cover		INT	
Channel or Bonnet				Channel Cover			
Tubesheet-Stationary				Tubesheet-Floating			
Floating Head Cover				Impingement Protection		YES	
Baffles Cross		Type VERT- SEG		%Cut 19.2 (Area)		Spacing-cc	19.2
Baffles-Long				Seal Type			
Supports-Tube			U-Bend		Type		
Bypass Seal Arrangement				Tube-Tubesheet Joint			
Expansion Joint				Type			
Rho-V2 Inlet Nozzle		1,673	Bundle Entrance		1,289	Bundle Exit	2,454
Gasket-Shellside			Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell			Filled with Water			Bundle	
Remarks:							

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	LP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	ZnO Preheater		Item No	H-420 ZnO Preheater	
Size 90x 96	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	14480 ft²	Shells/Unit	1 Surface/Shell (Effective)	14480 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Flue Gas fr. Tar Regen	Sweet Syngas
Total Fluid Entering	lb/hr	248,400	127,000
Vapor		248,400	127,000
Liquid		0	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	0.000/0.000
Liquid Viscosity	cP	0.000	0.000
Liquid Specific Heat	Btu/lb-F	0.000	0.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.000
Vapor Mol. Weight (In/Out)		27.57/27.57	10.99/10.99
Vapor Viscosity	cP	0.0256	0.0182
Vapor Specific Heat	Btu/lb-F	0.286	0.659
Vapor Thermal Conductivity	Btu/hr-ft-F	0.024	0.076
Temperature (In/Out)	°F	935.0/190.0	114.0/750.0
Operating Pressure	psi(Abs)	14.500	440.000
Velocity	ft/sec	-	-
Pressure Drop (Allow/Calc)	psi	1.000/-	5.000/0.287
Fouling resistance	hr-ft²-F/Btu	0.001000	0.001000

Heat Exchanged	52,900,000 Btu/hr	mtd (corr)	122.52.15 °F
Transfer Rate, Service	29.82	Clean	Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	30/	480/	
Design Temp.	°F	990	800	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-35.0	10.0	
Size &	Out	1-31.0	12.0	
Rating	Intermediate	0	0	

Tube No	12160	OD 0.750 in	Thk 0.065	Length 8.00 ft	Pitch 0.9375 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 96.00 OD in		Shell Cover	INT	
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 49.0 (Area)	Spacing-cc	50
Baffles-Long			Seal Type		
Supports-Tube		U-Bend		Type	
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle		Bundle Entrance		Bundle Exit	
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	LP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	ZnO Syngas Cooler/BFW Preheat		Item No	H-421	
Size 64x 144	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	6915 ft^2	Shells/Unit 1	Surface/Shell (Effective)	6915 ft^2	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas fr ZnO Bed	BFW
Total Fluid Entering	lb/hr	127,000	114,306
Vapor		127,000	0
Liquid		0	114,306
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft^3	0.000/0.000	58.509/46.533
Liquid Viscosity	cP	0.000	0.139
Liquid Specific Heat	Btu/lb-F	0.000	1.340
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.359
Vapor Mol. Weight (In/Out)		10.99/10.99	0.0/0.0
Vapor Viscosity	cP	0.0196	0.0000
Vapor Specific Heat	Btu/lb-F	0.660	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.082	0.000
Temperature (In/Out)	°F	750.0/265.0	240.0/546.5
Operating Pressure	psi(Abs)	425.000	1,295.000
Velocity	ft/sec	27.606	-
Pressure Drop (Allow/Calc)	psi	5.000/2.034	5.000/0.399
Fouling resistance	hr-ft^2-F/Btu	0.001000	0.002000

Heat Exchanged	40,570,000 Btu/hr	mtd (corr)	85.130 °F
Transfer Rate, Service	68.9	Clean	90.2 Btu/hr-ft^2-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	470/	1,360/	
Design Temp.	°F	800	600	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-15.0	4.0	
Size &	Out	1-13.0	6.0	
Rating	Intermediate	0	0	

Tube No	3364	OD 0.750 in	Thk 0.065	Length 12.00 ft	Pitch 1.00000 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 64.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	YES	
Baffles Cross	Type VERT- SEG		%Cut 18.6 (Area)	Spacing-cc	24.0
Baffles-Long			Seal Type		
Supports-Tube		U-Bend	Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	2,297	Bundle Entrance	611	Bundle Exit	3,020
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet							
					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	ZnO Syngas Cooler				Item No	H-422	
Size 30x 96		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	1190 ft²	Shells/Unit	1	Surface/Shell (Effective)		1190 ft²	

PERFORMANCE OF ONE UNIT			
Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Syngas fr H-421	Cooling Water
Total Fluid Entering	lb/hr	127,000	593,000
Vapor		127,000	0
Liquid		0	593,000
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	0.000/0.000	62.850/62.283
Liquid Viscosity	cP	0.000	0.734
Liquid Specific Heat	Btu/lb-F	0.000	1.027
Liquid Thermal Conductivity	Btu/hr-ft-F	0.000	0.363
Vapor Mol. Weight (In/Out)		10.99/10.99	0.0/0.0
Vapor Viscosity	cP	0.0140	0.0000
Vapor Specific Heat	Btu/lb-F	0.645	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.062	0.000
Temperature (In/Out)	°F	265.0/120.0	80.0/100.0
Operating Pressure	psi(Abs)	420.000	65.000
Velocity	ft/sec	54.190	1.566
Pressure Drop (Allow/Calc)	psi	5.000/4.440	5.000/0.420
Fouling resistance	hr-ft²-F/Btu	0.001000	0.002000

Heat Exchanged	11,860,000 Btu/hr	mtd (corr)	88.210 °F
Transfer Rate, Service	113.0	Clean	184.6 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL				
		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	465/	80/	
Design Temp.	°F	315	150	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-13.0	10.0	
Size &	Out	1-12.0	10.0	
Rating	Intermediate	0	0	

Tube No	802	OD 0.750 in	Thk 0.065	Length 8.00 ft		Pitch 0.93750 / 30.0°
Tube Type		PLAIN	Material			
Shell		I.D 30.00 OD in	Shell Cover		INT	
Channel or Bonnet			Channel Cover			
Tubesheet-Stationary			Tubesheet-Floating			
Floating Head Cover			Impingement Protection	NO		
Baffles Cross		Type VERT- SEG	%Cut 32.3 (Area)	Spacing-cc	24.0	
Baffles-Long			Seal Type			
Supports-Tube		U-Bend		Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint			
Expansion Joint			Type			
Rho-V2 Inlet Nozzle	2,469	Bundle Entrance	3,979	Bundle Exit	4,341	
Gasket-Shellside		Tubeside		Floating Head		
Code Requirement		ASME Section 8, Divsion 1		TEMA Class	R	
Weight/Shell		Filled with Water		Bundle		
Remarks:						

Heat Exchanger Specification sheet

					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	MeOH Compressor Interstage Cooling				Item No	H-500A	
Size 24x 72		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	511 ft²	Shells/Unit	1	Surface/Shell (Effective)		511 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		Shellside	Tubeside
Fluid Name		Cooling water	Syngas
Total Fluid Entering	lb/hr	553,000	127,000
Vapor		0	127,000
Liquid		553,000	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft³	62.000/62.000	0.000/0.000
Liquid Viscosity	cP	0.762	0.000
Liquid Specific Heat	Btu/lb-F	1.000	0.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.363	0.000
Vapor Mol. Weight (In/Out)		0.0/0.0	10.99/10.99
Vapor Viscosity	cP	0.0000	0.0155
Vapor Specific Heat	Btu/lb-F	0.000	0.655
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.068
Temperature (In/Out)	°F	80.0/100.0	333.0/200.0
Operating Pressure	psi(Abs)	65.000	1,000.000
Velocity	ft/sec	4.182	25.131
Pressure Drop (Allow/Calc)	psi	5.000/2.552	5.000/0.721
Fouling resistance	hr-ft²-F/Btu	0.002000	0.001000

Heat Exchanged	11,060,000 Btu/hr	mtd (corr)	170.297 °F
Transfer Rate, Service	127.1	Clean	215.4 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	80/	1,050/	
Design Temp.	°F	150	385	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-10.0	12.0	
Size &	Out	1-10.0	10.0	
Rating	Intermediate	0	0	

Tube No	476	OD 0.750 in		Thk 0.065	Length 6.00 ft		Pitch 0.93750 / 30.0°
Tube Type		PLAIN		Material			
Shell		I.D 24.00 OD in		Shell Cover		INT	
Channel or Bonnet				Channel Cover			
Tubesheet-Stationary				Tubesheet-Floating			
Floating Head Cover				Impingement Protection		YES	
Baffles Cross		Type VERT- SEG		%Cut 23.4 (Area)		Spacing-cc	16.3
Baffles-Long				Seal Type			
Supports-Tube			U-Bend		Type		
Bypass Seal Arrangement				Tube-Tubesheet Joint			
Expansion Joint				Type			
Rho-V2 Inlet Nozzle		1,279	Bundle Entrance		1,349	Bundle Exit	2,039
Gasket-Shellside			Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell			Filled with Water			Bundle	
Remarks:							

Heat Exchanger Specification sheet

			Job No.		
Customer	NREL		Ref No.	LP Syngas Case	
Address			Proposal No.		
Plant Location			Date		Rev. 0
Service of Unit	MeOH Syngas Preheat		Item No	H-501	
Size 73x 168	Type	BEM - HORZ	Connected in	1 Parallel	1 Series
Surf/Unit (Eff)	12712 ft²	Shells/Unit 1	Surface/Shell (Effective)	12712 ft²	

PERFORMANCE OF ONE UNIT

Fluid Allocation		Shellside	Tubeside
Fluid Name		Steam	Syngas to MeOH Rxn
Total Fluid Entering	lb/hr	18,830	127,000
Vapor		18,830	127,000
Liquid		0	0
Steam			
Noncondensable			
Fluid Vaporized or Condensed		18,830	0
Liquid Density (In/Out)	lb/ft³	0.000/52.387	0.000/0.000
Liquid Viscosity	cP	0.148	0.000
Liquid Specific Heat	Btu/lb-F	1.120	0.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.404	0.000
Vapor Mol. Weight (In/Out)		18.02/18.02	10.99/10.99
Vapor Viscosity	cP	0.0161	0.0170
Vapor Specific Heat	Btu/lb-F	0.492	0.659
Vapor Thermal Conductivity	Btu/hr-ft-F	0.020	0.074
Temperature (In/Out)	°F	487.0/324.0	239.0/460.0
Operating Pressure	psi(Abs)	100.000	1,160.000
Velocity	ft/sec	4.726	2.192
Pressure Drop (Allow/Calc)	psi	5.000/0.548	5.000/0.492
Fouling resistance	hr-ft²-F/Btu	0.005000	0.001000

Heat Exchanged	18,450,000 Btu/hr	mtd (corr)	60.365 °F
Transfer Rate, Service	24.0	Clean	28.3 Btu/hr-ft²-F

CONSTRUCTION OF ONE SHELL

		Shellside	Tubeside	Sketch
Design/Test Pres.	psi	100/	1,220/	
Design Temp.	°F	540	515	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections Size & Rating	In	1-8.0	10.0	
	Out	1-2.0	12.0	
	Intermediate	0	0	

Tube No	5242	OD 0.750 in	Thk 0.065	Length 14.00 ft	Pitch 0.93750 / 30.0°
Tube Type	PLAIN		Material		
Shell	I.D 73.00 OD in		Shell Cover		INT
Channel or Bonnet			Channel Cover		
Tubesheet-Stationary			Tubesheet-Floating		
Floating Head Cover			Impingement Protection	NO	
Baffles Cross	Type VERT- SEG		%Cut 10.4 (Area)	Spacing-cc	14.5
Baffles-Long			Seal Type		
Supports-Tube		U-Bend	Type		
Bypass Seal Arrangement			Tube-Tubesheet Joint		
Expansion Joint			Type		
Rho-V2 Inlet Nozzle	1,228	Bundle Entrance	1,623	Bundle Exit	1,384
Gasket-Shellside		Tubeside		Floating Head	
Code Requirement	ASME Section 8, Divsion 1			TEMA Class	R
Weight/Shell		Filled with Water		Bundle	
Remarks:					

Heat Exchanger Specification sheet

					Job No.		
Customer	NREL				Ref No.	LP Syngas Case	
Address					Proposal No.		
Plant Location					Date		Rev. 0
Service of Unit	Blowdown Cooler				Item No	H-601	
Size 12x 48		Type	BEM - HORZ	Connected in	1 Parallel		1 Series
Surf/Unit (Eff)	89 ft^2	Shells/Unit	1	Surface/Shell (Effective)		89 ft^2	

PERFORMANCE OF ONE UNIT

Fluid Allocation		**Shellside**	**Tubeside**
Fluid Name		Blowdown	Cooling water
Total Fluid Entering	lb/hr	3,164	30,465
Vapor		0	0
Liquid		3,164	30,465
Steam			
Noncondensable			
Fluid Vaporized or Condensed		0	0
Liquid Density (In/Out)	lb/ft^3	56.607/62.000	62.000/62.000
Liquid Viscosity	cP	0.311	0.762
Liquid Specific Heat	Btu/lb-F	1.059	1.000
Liquid Thermal Conductivity	Btu/hr-ft-F	0.382	0.363
Vapor Mol. Weight (In/Out)		0.0/0.0	0.0/0.0
Vapor Viscosity	cP	0.0000	0.0000
Vapor Specific Heat	Btu/lb-F	0.000	0.000
Vapor Thermal Conductivity	Btu/hr-ft-F	0.000	0.000
Temperature (In/Out)	°F	298.0/110.0	80.0/100.0
Operating Pressure	psi(Abs)	65.000	65.000
Velocity	ft/sec	0.170	0.561
Pressure Drop (Allow/Calc)	psi	5.000/0.111	5.000/0.536
Fouling resistance	hr-ft^2-F/Btu	0.001000	0.002000

Heat Exchanged	609,300 Btu/hr	mtd (corr)	89.027 °F
Transfer Rate, Service	76.9	Clean	104.4 Btu/hr-ft^2-F

CONSTRUCTION OF ONE SHELL

		Shellside	**Tubeside**	**Sketch**
Design/Test Pres.	psi	80/	80/	
Design Temp.	°F	350	150	
No. Passes per Shell		1	1	
Corrosion Allow.	in	0.0625	0.0625	
Connections	In	1-1.0	2.0	
Size &	Out	1-1.0	2.0	
Rating	Intermediate	0	0	

Tube No	116	OD 0.750 in		Thk 0.065	Length 4.00 ft		Pitch 0.93750 / 30.0°
Tube Type		PLAIN		Material			
Shell		I.D 12.00 OD in		Shell Cover		INT	
Channel or Bonnet				Channel Cover			
Tubesheet-Stationary				Tubesheet-Floating			
Floating Head Cover				Impingement Protection		YES	
Baffles Cross		Type VERT- SEG		%Cut 10.1 (Area)		Spacing-cc	2.3
Baffles-Long				Seal Type			
Supports-Tube			U-Bend		Type		
Bypass Seal Arrangement				Tube-Tubesheet Joint			
Expansion Joint				Type			
Rho-V2 Inlet Nozzle		459	Bundle Entrance		10	Bundle Exit	268
Gasket-Shellside			Tubeside			Floating Head	
Code Requirement		ASME Section 8, Divsion 1				TEMA Class	R
Weight/Shell			Filled with Water			Bundle	
Remarks:							

COMPRESSOR NUMBER			K-100			
SERVICE			Combustion Air			
GAS HANDLED			Air			
NORMAL FLOW	SCFM		51,965			
NORMAL FLOW	LB/HR		235,200			
DESIGN FLOW	SCFM					
MOL WT.			28.63			
C_p/C_v	Value		1.4			
	@ F / PSIA		90 / 14.7			
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		14.7			
COMPR. FACTOR @ SUCTION			0.999			
FLOW AT SUCTION	ACFM		54,910			
ORIGIN	PSIA					
TEMPERATURE	F		90			
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		20			
DISCH. TEMPERATURE	F	(2)	157			
COMPR. FACTOR @ DISCH.			0.999			
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			1.36			
EFFICIENCY		(2)	0.75			
BHP		(2)	1600			
COMPRESSOR TYPE						
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
	H_2O		3.1			
	O_2		20.3			
	Ar		0.9			
	N_2		75.7			

(1) INCLUDES ALLOWANCE FOR SUCTION OR DISCHARGE SNUBBER
(2) VALUE TABULATED IS ESTIMATED AND MUST BE VERIFIED BY FINAL MECHANICAL DESIGN

NO	DATE	REVISIONS	PROC	PROJ.	CLIENT

NREL BIOMASS GASIFICATION: Low Pressure Syngas Case (BLC Gasifier)	JOB NO	NREL Contract ACO-5-44027
	DRAWING NO	REV

COMPRESSOR NUMBER			K-300A - Stage 1	K-300B - Stage 2	K-300C - Stage 3	K-300D - Stage 4
SERVICE			Syngas Compressor Stage 1	Syngas Compressor Stage 2	Syngas Compressor Stage 3	Syngas Compressor Stage 4
GAS HANDLED			Syngas	Syngas	Syngas	Syngas
NORMAL FLOW	SCFM		120,208	90,448	88,044	87,158
NORMAL FLOW	LB/HR		317,371	232,617	225,773	223,220
DESIGN FLOW	SCFM					
MOL WT.			16.7	16.26	16.21	16.2
C_p/C_v	Value		1.36	1.374	1.379	1.39
	@ F / PSIA		157 / 15.88	110 / 30	110 / 79	110 / 215
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		15.88	30	79	215
COMPR. FACTOR @ SUCTION			0.9979	0.999	0.9985	0.9972
FLOW AT SUCTION	ACFM		131,756	48,531	17,936	6,513
ORIGIN	PSIA					
TEMPERATURE	F		157.1	110	110	110
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		35	84	220	450
DISCH. TEMPERATURE	F	(2)	344.2	349.6	349.1	277
COMPR. FACTOR @ DISCH.			0.9982	1.001	1.003	1.005
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			2.204	2.8	2.78	2.093
EFFICIENCY		(2)	0.75	0.75	0.75	0.75
BHP		(2)	11,248	10,251	10,251	7,036
COMPRESSOR TYPE						
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
		H_2	39.79	52.87	54.32	54.88
		CO_2	12.36	16.42	16.87	17.04
		CO	18.42	24.48	25.15	25.41
		H_2O	27.97	4.28	1.67	0.66
		CH_4	1.19	1.58	1.62	1.64
		C_2H_2	0.02	0.02	0.02	0.02
		C_2H_4	0.02	0.22	0.23	0.23
		C_2H_6	0	0.00001	0.00002	0.00002
		Benzene (C_6H_6)	0	0.000006	0.000006	0.000007
		Tar ($C_{10}H_8$)	0	0.00001	0.000001	0.000001
		Ammonia (NH_3)	0.01	0.01	0.01	0.01
		H_2S	0.03	0.04	0.04	0.04
		N_2	0.06	0.08	0.08	0.08

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NO	DATE	REVISIONS	PROC	PROJ.	CLIENT

JOB NO	NREL Contract ACO-5-44027
DRAWING NO	REV

NREL BIOMASS GASIFICATION: Low Pressure Syngas Case (BCL Gasifier)

COMPRESSOR NUMBER			K-420			
SERVICE			Flue Gas Blower			
GAS HANDLED			Flue Gas			
NORMAL FLOW	SCFM		56,988			
NORMAL FLOW	LB/HR		248,400			
DESIGN FLOW	SCFM					
MOL WT.			27.57			
C_p/C_v	Value		1.367			
	@ F / PSIA		176 / 14.3			
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		14.3			
COMPR. FACTOR @ SUCTION			0.9982			
FLOW AT SUCTION	ACFM		71,490			
ORIGIN	PSIA					
TEMPERATURE	F		175.8			
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		14.7			
DISCH. TEMPERATURE	F	(2)	182			
COMPR. FACTOR @ DISCH.			0.9982			
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			1.028			
EFFICIENCY		(2)	0.75			
BHP		(2)	177			
COMPRESSOR TYPE						
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
	CO_2		0.03			
	H_2O		3.1			
	O_2		20.29			
	Ar		0.91			
	N_2		75.67			

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NO	DATE	REVISIONS	PROC	PROJ.	CLIENT

NREL BIOMASS GASIFICATION: Low Pressure Syngas Case (BCL Gasifier)

JOB NO	NREL Contract ACO-5-44027	
DRAWING NO		REV

				K-500A	K-500B		
COMPRESSOR NUMBER				K-500A	K-500B		
SERVICE				MeOH Compressor Stage 1	MeOH Compressor Stage 2		
GAS HANDLED				Treated Syngas	Treated Syngas		
NORMAL FLOW		SCFM		73,055	73,055		
NORMAL FLOW		LB/HR		127,035	127,035		
DESIGN FLOW		SCFM					
MOL WT.				10.99	10.99		
C_p/C_v		Value		1.418	1.424		
		@ F / PSIA		115 / 415	200 / 995		
SUCTION CONDITIONS							
	SUCTION PRESSURE	PSIA		415	995		
	COMPR. FACTOR @ SUCTION			1.006	1.021		
	FLOW AT SUCTION	ACFM		2,881	1,400		
	ORIGIN	PSIA					
	TEMPERATURE	F		110	200		
	LINE LOSS	PSI	(2)				
	OTHER LOSSES	PSI	(1, 2)				
	CONTINGENCY	PSI					
DISCHARGE CONDITIONS							
	DISCH. PRESSURE	PSIA		1,000	1,160		
	DISCH. TEMPERATURE	F	(2)	326	239.3		
	COMPR. FACTOR @ DISCH.			1.023	1.026		
	DELIVERY	PSIA					
	LINE LOSS	PSI	(2)				
	EXCHANGER LOSS	PSI	(2)				
	HEATER LOSS	PSI	(2)				
	CONTROL VALVE LOSS	PSI	(2)				
	OTHER LOSSES	PSI	(2)				
	CONTINGENCY	PSI	(2)				
	TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO				2.41	1.17		
EFFICIENCY			(2)	0.75	0.75		
BHP			(2)	7,377	1,340		
COMPRESSOR TYPE							
DRIVER TYPE							
GAS COMPOSITION: Vol. %							
			H_2	65.45	65.45		
			CO_2	1.63	1.63		
			CO	30.3	30.3		
			H_20	0.26	0.26		
			CH_4	1.96	1.96		
			C_2H_2	0.03	0.03		
			C_2H_4	0.28	0.28		
			C_2H_6	0.00002	0.00002		
			Benzene (C_6H_6)	0.000008	0.000008		
			Tar ($C_{10}H_8$)	0.000001	0.000001		
			NH_3	0.01	0.01		
			N_2	0.095	0.095		

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NO	DATE	REVISIONS	PROC	PROJ.	CLIENT		
		NREL BIOMASS GASIFICATION: Low Pressure Syngas Case (BCL Gasifier)				JOB NO	NREL Contract ACO-5-44027
						DRAWING NO	REV

COMPRESSOR NUMBER			M-601A	M-601B		
SERVICE			Steam Turbine - Extraction Stage 1	Steam Turbine - Extraction Stage 2		
GAS HANDLED			Steam	Steam		
NORMAL FLOW	SCFM		88,402	81,815		
NORMAL FLOW	LB/HR		251,800	232,900		
DESIGN FLOW	SCFM					
MOL WT.			18.02	18.02		
C_p/C_v	Value		1.384	1.336		
	@ F / PSIA		1000 / 1265	564.8 / 165		
SUCTION CONDITIONS						
SUCTION PRESSURE	PSIA		1265	100		
COMPR. FACTOR @ SUCTION			0.9332	0.977		
FLOW AT SUCTION	ACFM		2,691	21,390		
ORIGIN	PSIA					
TEMPERATURE	F		1000	487		
LINE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(1, 2)				
CONTINGENCY	PSI					
DISCHARGE CONDITIONS						
DISCH. PRESSURE	PSIA		100	50		
DISCH. TEMPERATURE	F	(2)	487	376		
COMPR. FACTOR @ DISCH.			0.977	0.9833		
DELIVERY	PSIA					
LINE LOSS	PSI	(2)				
EXCHANGER LOSS	PSI	(2)				
HEATER LOSS	PSI	(2)				
CONTROL VALVE LOSS	PSI	(2)				
OTHER LOSSES	PSI	(2)				
CONTINGENCY	PSI	(2)				
TOTAL LOSSES	PSI	(2)				
COMPRESSION RATIO			-	-		
EFFICIENCY		(2)	0.75	0.75		
kW Generated		(2)	16,067	3,343		
Turbine TYPE			Steam	Steam		
DRIVER TYPE						
GAS COMPOSITION: Vol. %						
		H_2				
		CO_2				
		CO				
		H_20	100%	100%		
		CH_4				
		C_2H_2				
		C_2H_4				
		C_2H_6				
		Benzene (C_6H_6)				
		Tar ($C_{10}H_8$)				
		NH_3				
		N_2				

(1) INCLUDES ALLOWANCE FOR SUCTION OR DISCHARGE SNUBBER
(2) VALUE TABULATED IS ESTIMATED AND MUST BE VERIFIED BY FINAL MECHANICAL DESIGN

NO	DATE	REVISIONS	PROC	PROJ.	CLIENT	
		NREL BIOMASS GASIFICATION: Low Pressure Syngas Case (BCL Gasifier)				JOB NO: NREL Contract ACO-5-44027
						DRAWING NO / REV

Cyclone Specification Sheet

Site Location	(Note: Four (4) parallel cyclones)	Date			Rev.

SERVICE OF LOW PRESSURE UNIT S-100 and S-101

Inlet Conditions	Flow	Viscosity	Density	Molecular Weight (Ave.)	Particle Size (mm) (Stokes' MMD)	Volumetric Flowrate	Temperature
	lb/h	lb/ft-sec	lb/ft3	lb/mole		acfm	°F
Gas (Split into four parallel flows)	316,369.00	2.35 x 10-5	0.03500	18.7		150,652.00	1,598
Particulate	40,407.00		62.40		60		
Gas Inlet Pressure (psia)	33.00						
Gas Discharge Pressure (psig)	32.64						
Pressure Drop, Max Allow. (" .WC.)	10.48						
Design/Test Pressure Psig	33.00						
Design Particulate Cutpoint	50						
Design Separation Efficiency at Cutpoint (%)	98						

Emery Design Calculations Summary for S-100 (for Reference Only)

Mechanical Sizing		Inside Diam (in)	Uninsulated Outside Diam (in)		ID (in)	OD (in)	Thickness (in)	Designation		Overall Height (ft)
Connections Size & Rating	In	48	58	Upper Shell	82	84	1	ASME VIII		35
	Out	36	46	Inner Tube	36					
	Bottom	TBD		Cone			1	ASME VIII		
				Refractory	74		4			

Component Data					Cyclone Body Materials of Construction					
	Design Temperature (°F)	Solids Removal Flowrate (CFM)	Differential Design Pressure (psig)	Type	Upper Section		Lower Conical section		Nozzles	
Rotary Air Lock	1598				Inner Wall	Outer Shell	Inner Wall	Outer Shell	Inner Wall	Outer Shell
Level Indicator	1598				Cercast™	MS	Cercast™	MS	Cercast™	MS
					Inner Tube					
					MS					

Vendor/Supplier Specifications and Price Quote

Fisher-Klosterman, Inc (Refer to Vendor Communications and Data Sheets)
Ryan Bruner, Sales Manager
P.O. Box 11190
Lousville, Ky
Ph: 502-572-4000 ext 213
Email: rab@fkinc.com

Recommendation: Replace S-100 and S-101 with 4 cyclones operated in parallel using split air flow:

Four (4) XQ120-48M cyclone assemblies each with the following Features:

Design, fabricated, tested, and stamped as an ASME vessel		Interior surfaces to be lined with 4" of Vesuvius Cercast 3300 castable refractory	
3/8" plate carbon steel construction		All welding per FKI Class 3 preocedures with 100% penetration	
Dust receiver section with flanged discharge		Exterior to be sandblasted and painted with high temperature aluminum paint	
40"∅ gas inlet flange		Design pressure (psig)	33
48"∅ verticle gas outlet flange		Design Temperature (F)	650
Approximate Overall Dimensions:	7 ft∅ x 35 ft tall		

Gas Conditions at Inlet:		Particulate Conditions at Inlet:	
Volume per cylone (acfm)	37,663	Specific Gravity	1.000
Density (lbm/ft3)	0.035	Dust Loading (Grains/acf)	31.3
Viscosity (lbm/ft-sec)	2.53E-05		

		Fraction Efficiencies: Stokes Equiv. % Efficiency	
Inlet Velocity (ft/sec)	78.46	Dia.(microns)	Weight %
No load pres. drop (in.W.C.)	12.6	3	7.37
Full load pres. Drop (in. W.C.)	10.02	3.5	16.3
		4	21.44
		4.5	26.75
		5	32.07
		5.5	37.27
		6.5	42.27
		7.5	51.48
		8.5	59.48
		9.5	66.29
		10.5	71.99
		13	82.36
		17	89.12
		24	94.36
		34	97.39
		89	99.83

Price (2005 U.S.$) **$ 1,225,000.00**

Remarks: Inlet and outlet manifolding is not included in Fisher-Klosterman quote for these four cylones. Estimated cost of splitter and collection is $25,000.
Refer to supplier data sheet for Vesuvius CERCAST™ 3300 Castable refractory.

Cyclone Specification Sheet

Site Location							Date			Rev.
SERVICE OF LOW PRESSURE UNIT S-102										
Inlet Conditions				Flow	Viscosity	Density	Molecular Weight (Ave.)	Particle Size (mm) (Stokes' MMD)	Volumetric Flowrate	Temperature
				lb/h	BTU/lb°F	lb/ft3	lb/mole		acfm	°F
Gas				328,979.00	2.78E-05	0.34470	16.7		150,612.01	1,598
Particulate				40,407.00		62.40		60		
Gas Inlet Pressure (psia)				33.00						
Gas Discharge Pressure (psig)				32.64						
Pressure Drop, Max Allow. (" .WC.)				10.00						
Design/Test Pressure Psig				33.00						
Design Particulate Cutpoint				50						
Design Separation Efficiency at Cutpoint (%)				98						
Emery Design Calculations Summary for S-102 (for Reference Only)										
Mechanical Sizing		Inside Diam (in)	Uninsulated Outside Diam (in)		ID (in)	OD (in)	Thickness (in)	Designation		Overall Height (ft)
Connections Size & Rating	In	34	44	Upper Shell	58	60	1	ASME VIII		25
	Out	26	36	Inner Tube	34					
	Bottom			Cone			1	ASME VIII		
				Refractory	50		4			
Component Data					Cyclone Body Materials of Construction					
	Design Temperature (°F)	Solids Removal Flowrate (CFM)	Differential Design Pressure (psig)	Type	Upper Section		Lower Conical section		Nozzles	
Rotary Air Lock	1598	20.4	15		Inner Wall	Outer Shell	Inner Wall	Outer Shell	Inner Wall	Outer Shell
Level Indicator	1598				Cercast™	MS	Cercast™	MS	Cercast™	MS
					Inner Tube					
					MS					
Vendor/Supplier Specifications and Price Quote										
Fisher-Klosterman, Inc						(Refer to Vendor Communications and Data Sheets)				
Ryan Bruner, Sales Manager										
P.O. Box 11190										
Lousville, Ky										
Ph: 502-572-4000 ext 213										
Email: rab@fkinc.com										
Recommendation: Quote Pending										
Four (4) XQ120-48M cyclone assemblies each with the following Features:										
Design, fabricated, tested, and stamped as an ASME vessel					Interior surfaces to be lined with 4" of Vesuvius Cercast 3300 castable refractory					
3/8" plate carbon steel construction					All welding per FKI Class 3 preocedures with 100% penetration					
Dust receiver section with flanged discharge					Exterior to be sandblasted and painted with high temperature aluminum paint					
40"∅ gas inlet flange					Design pressure (psig)		460			
48"∅ verticle gas outlet flange					Design Temperature (F)		650			
Approximate Overall Dimensions:			5 ft∅ x 25 ft tall							
Gas Conditions at Inlet:				Particulate Conditions at Inlet:						
Volume per cylone (acfm)		15,906		Specific Gravity		1.000				
Density (lbm/ft3)		0.3447		Dust Loading (Grains/acf)		7.33				
Viscosity (lbm/ft-sec)		2.78E-05								
Inlet Velocity (ft/sec)		70.11		Fraction Efficiencies: Stokes Equiv. % Efficiency						
No load pres. drop (in.W.C.)		73.64		Dia.(microns)	Weight %					
Full load pres. Drop (in. W.C.)		63.69		2.5	4.91					
				3.5	12.88					
				4.5	22.89					
				5	28.13					
				5.5	33.31					
				6	38.34					
				7	47.7					
				8	55.93					
				9	63					
				10	68.97					
				11	73.98					
				13	81.64					
				17	88.65					
				24	94.08					
				34	97.25					
				74	99.67					
Price (2005 U.S.$)		**$ 370,000.00**								

Remarks: Inlet and outlet manifolding is not included in Fisher-Klosterman quote for these four cylones. Estimated cost of splitter and collection is $25,000.
Refer to supplier data sheet for Vesuvius CERCAST^{TM} 3300 Castable refractory.

Cyclone Specification Sheet

Site Location		Date			Rev.

SERVICE OF LOW PRESSURE UNIT S-103

Inlet Conditions	Flow	Voscosity	Density	Molecular Weight (Ave.)	Particle Size (mm) (Stokes' MMD)	Volumetric Flowrate	Temperature
	lb/h	lb/ft-sec	lb/ft3	lb/mole		acfm	°F
Gas	248,368.00	2.87E-05	0.03501	27.6		7,289.00	1,798
Particulate	40,407.00		1.00		60		
Gas Inlet Pressure (psia)	33.00						
Gas Discharge Pressure (psig)	32.64						
Pressure Drop, Max Allow. (" .WC.)	10.00						
Design/Test Pressure Psig	33.00						
Design Particulate Cutpoint	50						
Design Separation Efficiency at Cutpoint (%)	98						

Emery Design Calculations Summary for S-103 (for Reference Only)

Mechanical Sizing		Inside Diam (in)	Uninsulated Outside Diam (in)		ID (in)	OD (in)	Thickness (in)	Designation		Overall Height (ft)
Connections Size & Rating	In	26	36	Upper Shell	46	48	1	ASME VIII		25
	Out	18	28	Inner Tube	18		4			
	Bottom			Cone			1	ASME VIII		
				Refractory	38		4			

Component Data	Design Temperature (°F)	Solids Removal Flowrate (CFM)	Differential Design Pressure (psig)	Type	Cyclone Body Materials of Construction: Upper Section		Lower Conical section		Nozzles	
Rotary Air Lock	938	20.4	15		Inner Wall	Outer Shell	Inner Wall	Outer Shell	Inner Wall	Outer Shell
Level Indicator	938				Cercast™	MS	Cercast™	MS	Cercast™	MS
					Inner Tube					
					MS					

Vendor/Supplier Specifications and Price Quote

Fisher-Klosterman, Inc
Ryan Bruner, Sales Manager
P.O. Box 11190
Lousville, Ky
Ph: 502-572-4000 ext 213
Email: rab@fkinc.com

(Refer to Vendor Communications and Data Sheets)

Recommendation: Quote Pending

Four (4) XQ120-48M cyclone assemblies each with the following Features:

Design, fabricated, tested, and stamped as an ASME vessel	Interior surfaces to be lined with 4" of Vesuvius Cercast 3300 castable refractory
3/8" plate carbon steel construction	All welding per FKI Class 3 preocedures with 100% penetration
Dust receiver section with flanged discharge	Exterior to be sandblasted and painted with high temperature aluminum paint
40"∅ gas inlet flange	Design pressure (psig): 460
48"∅ verticle gas outlet flange	Design Temperature (F): 650
Approximate Overall Dimensions: 4 ft∅ x 17 ft tall	

Gas Conditions at Inlet:		Particulate Conditions at Inlet:	
Volume per cylone (acfm)	7,289	Specific Gravity	1.000
Density (lbm/ft3)	0.5679	Dust Loading (Grains/acf)	16
Viscosity (lbm/ft-sec)	2.87E-05		
Inlet Velocity (ft/sec)	72.29	Fraction Efficiencies: Stokes Equiv. % Efficiency	
No load pres. drop (in.W.C.)	120.63	Dia.(microns)	Weight %
Full load pres. Drop (in. W.C.)	99.82	2.5	8.46
		3	13.57
		3.5	19.29
		4	25.27
		4.5	31.27
		5	37.1
		5.5	42.64
		6	47.84
		7	57.08
		8	64.8
		9	71.14
		10	76.31
		12	83.89
		16	90.07
		21	94.11
		31	97.52
		101	99.93

Price (2005 U.S.$) $ 250,000.00

Remarks: Inlet and outlet manifolding is not included in Fisher-Klosterman quote for these four cylones. Estimated cost of splitter and collection is $25,000. Refer to supplier data sheet for Vesuvius CERCAST™ 3300 Castable refractory.

D.1 INTRODUCTION

The first task undertaken by the team was to examine commercial technologies that are suitable for synthesis gas cleanup for biomass gasification. Currently, there are various types of technologies available dependent upon the specific cleanup requirements. For example, the clean-up required for syngas that will ultimately be fed to a reciprocating engine is much less than for syngas used in chemical synthesis. This study examined all technologies that could be required for syngas that will be used for Fischer-Tropsch (FT) liquids and alcohol synthesis.

The gas cleanup configuration for a system is generally determined by the composition of the syngas exiting the gasifier, the cleanup requirements for the intended use of the syngas, and economic considerations. Technologies such as cyclone separators, barrier filters, and electrostatic precipitators are routinely used for solid particulate removal. Catalytic tar crackers are employed to destroy tars and nitrogen contaminants. Wet scrubbers are used to remove a number of contaminants such particulates, alkali species, halides, soluble gases, and condensable liquids. Acid gas removal technologies encompass a large selection of processes including amine-based, physical solvent, liquid phase oxidation, and catalytic absorbent. Each section focuses on the operating size ranges and conditions, materials of construction, and cleanup parameters for each technology considered.

D.2 PARTICULATE REMOVAL TECHNOLOGIES

D.2.1 INTRODUCTION

During the gasification process, the mineral matter contained in the biomass feedstock will form inorganic ash, and the unconverted biomass will form char. These particulates are entrained in the syngas stream as it exits the gasifier. The concentration of particulates produced is often influenced by the gasifier design. These particulates can present emissions problem and can cause abrasion to downstream equipment. Therefore, the particulates concentration must be reduced using various technologies discussed in the following paragraphs.

Cyclone Separators

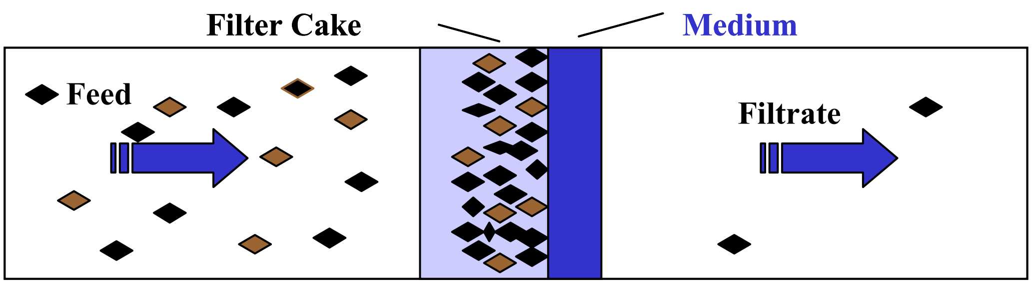
Cyclones use centrifugal forces to separate the bulk of large size particulates from a gas stream. In gasification systems, cyclones are normally used as the first step in the gas cleanup process. They are relatively inexpensive to manufacture and easy to operate which translate to low capital and maintenance costs. In general, 90-98% of particulates 10 μm or larger in diameter can be removed, but the removal efficiency decreases significantly for smaller particulates[13]. The removal efficiency also decreases as the operating temperatures increases. Cyclones are capable of handling operating temperatures up to 2000°F and can be designed to operate at pressures normally encountered in gasifiers. Cyclones are usually made from carbon steel and are refractory lined to withstand high temperature environments. A flow range from 300 to 13,000 CFM is typical for cyclones. This flow range is within the parameter of the syngas flow rate specified by NREL for this project.

13 Donaldson Co., Inc. "Cyclone Dust Collectors," July 2003, <http://www.donaldson.com/en/industrialair/literature/000984.pdf

D.2.2 BARRIER FILTERS

Barrier filters remove particulates by capturing the particulates on the filter surfaces as the gas stream passes through the filter medium. The particulates accumulated on the filter surfaces form a cake, which can be dislodged by initiating a blowback flow. The blowback gas flows in the reverse direction of normal process flow and dislodges the filter cake, which is then removed from the system. The operating principle of barrier filters is illustrated in Figure D-1. Barrier filters include high-temperature filters, such as ceramic and metal candle filters, and low-temperature filters, such as baghouse filters.

FIGURE D-1 PRINCIPLE OF BARRIER FILTERS



Ceramic Candle Filters

Ceramic filters are designed to remove particulate matter from gas streams at elevated temperatures. Ceramic filters can be designed for any flow requirement and can remove 90% of particulates larger than 0.3 µm[14]. In theory, the ceramic filter elements, normally made of aluminosilicate or silicon carbide powder with a sodium aluminosilicate binder, have exceptional physical and thermal properties, and should be able to withstand high temperature operations of up to 1800°F. However, commercial operations using ceramic filters at this temperature range have not been successful due to the susceptibility of the filter elements to cracking. Advances in composite filter element materials that have resistance to crack propagation at high temperatures are being developed and tested[15]. At temperatures below 850°F, ceramic filters have demonstrated satisfactory operational reliability.

In operations where tars are formed in the gasifier, ceramic filters should be operated at temperatures above the dew point of the tars (usually about 700-750°F) to avoid tar condensation. Condensed tar accumulates on filter surfaces and leads to plugging which will reduce the lifetime of the filter and impact process flowrates.

Metal Candle Filters

Metal filters are used in high temperature cleanup systems to remove particulate matter and can achieve filtration level as low as 1 µm. They can be designed to meet any flow requirement and can operate over a wide range of temperatures depending on the material of construction. Metal

[14] Pall Corp., "Syngas Filter Proposal," 26 January 2005, office communication
[15] Jay E. Lane, Jean-Francois LeCostaouec, "Ceramic Composite Hot Gas Filter Development," <http://www.netl.doe.gov/publications/proceedings/98/98ps/pspb-5.pdf

filters made from stainless steel can be used in cleanup systems for temperatures below 650°F while Inconel or alloy HR filters are suitable for operating temperatures up to 1100°F. At even higher temperatures, Fercalloy can withstand temperatures up to 1800°F[16], although commercial operation at this temperature has not been demonstrated. Commercial operation of metal filters operating at a maximum temperature of 915°F has been successful at a few gasification facilities in Europe[17].

Some operational considerations for metal filters are the corrosion rate and tar deposition on filter elements. Under similar stream compositions and conditions, the corrosion rate of metal filter elements is ten times that of the surrounding piping; thus, a regular maintenance schedule is essential to ensure operational reliability. Additionally, in operations where filter elements are subjected to frequent cleaning cycles due to tar deposition, the lifetime of the filter will be reduced. Therefore, it is recommended that the filter be operated at a temperature above the dew point of the tars in the syngas stream to avoid tar condensation and deposition.

Baghouse Filters

Baghouse filters are made of a woven fabric or felted (non-woven) material to remove particulate matter from an air or gas stream and can remove particulates down to 2.5 µm[18]. For woven fabric filters, the removal efficiency increases as the thickness of filter cake increases; thus, the removal efficiency of these systems is constantly changing. Felted filter systems have a constant removal efficiency that does not depend on the thickness of the filter cake[19]. Baghouse filters are modular in design and thus can accommodate a wide flow range from 1,500 to 150,000 CFM. The air-to-cloth ratio, or ratio of the volumetric flow to cloth area, sets the size of a baghouse unit. The bag fabric can be made from various materials including polyester, acrylic, NOMEX, Teflon, Ryton, and fiberglass[20]. The operating temperature range of an application influences the selection of bag material. For example, materials such as polyester or acrylic are suitable for applications with operating temperatures below 300°F, while NOMEX, Teflon, Ryton, or fiberglass is recommended for temperatures up to 500°F. Due to the temperature limits of the filter fabric, baghouse filters are only used in the low-temperature cleanup systems. They are often used downstream of the cyclones so that the particulate loading on the filters can be reduced.

Disadvantages of baghouse filters include the need for periodic bag replacement that can result in high maintenance costs and the potential for bag fire or explosion. A spark detection and extinguishment system, along with bag grounding strips, are recommended safety measures to mitigate the fire potential. Additionally, the performance of the filter fabrics degrades drastically with tar deposition on the fabric surface, so fabric surface treatments such as Teflon coating and pre-coating with limestone or other compatible filter aids is recommended. Such pre-coats can

16 Mott Corp., "Fiber Metal. The High-Flow, Low-Pressure Drop Alternative," June 2003, http://www.mottcorp.com/resource/pdf/PSFIBERfinal.pdf

17 Mike Wilson, Mott Corp., "Fercalloy Metal Filters," 2 February 2005, Vendor input

18 Donaldson Co., Inc. "Dalamatic Dust Collectors," December 2002, http://www.donaldson.com/en/industrialair/literature/000983.pdf

19 EPA, "Air Pollution Technology Fact Sheet-Fabric Filter – Pulse-Jet Cleaned Type," http://www.macrotek.net/pdf/FS_Pulse_Clean_Dust_Collector.pdf

20 Ducon, "Baghouse Filter," 2003, http://www.ducon.com/bag-house-filter.php

also be used to adsorb mercury and other contaminants.. Industry experience suggests that either ceramic or metal filters should be used in place of baghouse filters in high temperature operations.

D.2.3 ELECTROSTATIC PRECIPITATORS (ESPs)

ESPs are commonly used in large power plants to control fly ash emissions. ESPs consist of discharge electrodes centered between positively grounded collection plates. As the gas stream laden with particulates passes through the ESP, the discharge electrodes provide a negative charge to the particulates. The positively grounded collection plates act as a magnet for the negatively charged particulates, which collect on the plates. The collected particulates are transported into the collection hopper by the rapper or vibrator system.

ESPs are classified as either wet or dry processes. In wet ESPs, a water quench is applied either intermittently or continuously to the collection plates. The purpose of the water quench is to prevent possible fires that have occasionally resulted from the use of dry ESPs. The wastewater from wet ESPs must be treated prior to disposal.

For dry ESPs, the removal efficiency decreases for particulates with a high electrical resistivity since these particulates can introduce positive ions into the gas space resulting in reduced attraction of the negatively charged particulates to the collection plates. Particulates with a high resistivity are commonly produced from combustion of low-sulfur coals. Flow ranges of 10,000 – 300,000 CFM are typical for dry ESPs. Dry ESPs operate in the pressure range from vacuum conditions up to 150 psi and can operate at temperatures up to 750°F[21].

Wet ESPs can achieve 99.9% removal of sub-micron particulates down to 0.01 μm. Particulate resistivity does not affect removal efficiency of wet ESPs since the humid operating environment often reduces the resistivity of particulates. These systems are generally designed for gas flow range from 1,000 to 100,000 CFM. Gas streams with particulate sizes larger than 2 μm or with an exceptionally high particulate loading should be pretreated to reduce the load on the ESP. Wet ESPs operate in the pressure range from vacuum conditions up to 150 psi, with operating temperatures limited to 170-190°F[22,23].

The type of ESP selected for an application is largely influenced by the operating parameter and the type of particulates to be removed. However, the use of ESPs is limited in gasification systems due to the significant capital costs compared to other systems. Additionally, the removal efficiency of ESPs is sensitive to fluctuations in process conditions, such as changes in temperatures and pressures, gas compositions, and particulate loading. Therefore, ESPs are not suitable for biomass gasification applications that have highly variable syngas compositions from different feedstocks.

21 Gerry Graham, "Controlling Stack Emissions in the Wood Products Industry," http://www.ppcesp.com/ppcart.html
22 Ducon, "Wet & Dry Electrostatic Precipitators," 2003, http://www.ducon.com/wet-dry-precipitators.php (24 January 2005)
23 EPA, "Air Pollution Technology Fact Sheet-Wet Electrostatic Precipitator (ESP)-Wire-Pipe Type," http://www.p2pays.org/ref/10/09890.pdf (25 January 2005)

D.3 TAR REMOVAL TECHNOLOGIES

D.3.1 INTRODUCTION

Following NREL guidelines for the purpose of this project, tar is defined as C10+ hydrocarbons. Tar in syngas products can cause serious operational problems when the syngas stream cools below the dew point of the tars (usually about 700-750°F) and tar deposition occurs on downstream equipment and piping. Thus, tar removal is critical when there is tar present in the syngas. Tar can be removed either by physical or chemical processes. The most common physical process involves cooling the syngas stream to condense the tar into fine droplets and removing these droplets by wet scrubbing. Chemical process involves catalytic steam reforming of tars to lighter gases.

D.3.2 WET SCRUBBERS

Wet scrubbing is generally used to remove water-soluble contaminants from the syngas by absorption into a solvent. Tar components are water-soluble can be removed by this method. Additionally, wet scrubbing is also often used to remove a number of other contaminants such as particulates, alkali species, halides, soluble gases, and condensable liquids. In wet scrubbing, water is a common solvent choice. Wet scrubbers with the venturi design are frequently used in gas cleanup applications to achieve sub-micron particulate removal requirements. As the gas stream enters the venturi scrubber, the scrubbing liquid is sprayed into the gas stream. The two streams are thoroughly mixed by the turbulence in the venturi throat section where fine particles are impacted and agglomerate into liquid droplets. The liquid droplets are separated from the gas stream in a separator unit consisting of a cyclone separator or a mist eliminator.

Venturi scrubbers can achieve 99.9% removal efficiency of sub-micron particulates. Flow range for a single-throat venturi is 500-100,000 SCFM. Flows above this range require either multiple venturi scrubbers in series or a multiple-throat venturi[24]. Venturi scrubbers with a quench section can accommodate high temperature gas streams up to 450°F, and they can operate over a wide range of pressures[25].

The standard material of construction for venturi scrubbers is carbon steel. For corrosive or high temperature applications, stainless steel or special alloys such as FRP (fiberglass reinforced plastic) and Inconel are used.

The disadvantages of scrubbers include high pressure drop, the need to treat the wastewater effluent prior to disposal, and the loss of sensible heat of the syngas due to quenching. In power generation applications, the loss of sensible heat reduces the energy content of the gas and thus is undesirable; however, it is less of a concern in biomass refinery applications. Nevertheless, sensible heat loss will result in reduced overall system efficiency.

24 EPA, "Air Pollution Technology Fact Sheet-Venturi Scrubber" <http://www.macrotek.net/pdf/FS_Venturi_Scrubber.pdf

25 Envitech, Inc., "Venturi Scrubber," <http://www.envitechinc.com/scrubber.zhtml

D.3.3 CATALYTIC TAR REFORMING

Catalytic reforming of biomass tars is a developing technology for tar removal from syngas streams. The concept of this technology is to reform tar in a fluidized reactor bed, or tar cracker, into lighter gases using a proprietary catalyst. In addition to tar, light hydrocarbons (C1 to C5), benzene, and ammonia are also removed. A few large-scale biomass gasification facilities, such as Carbona in Denmark and the FERCO gasifier in Vermont, have demonstrated a novel catalyst in their tar crackers since commercial catalysts are too friable for this application[26]. The FERCO tar cracker removed 90% of the tar in the syngas stream using a novel catalyst known as DN34[27]. In both of these processes, a wet scrubber was used downstream of the tar cracker to remove residual tars and impurities.

A tar cracker known as the Reverse Flow Tar Cracking (RFTC) reactor developed by BTG uses the steam reforming process with a commercial nickel catalyst[28]. The nickel catalyst is very sensitive to sulfur impurities; therefore, a syngas stream containing sulfur contaminants has to be desulfurized prior to entering the RFTC reactor. Due to the cooling requirement for the desulfurization process, the syngas is fed to the reactor at a temperature from 660 -1200°F and is heated to the reaction temperature of 1650 -1740°F in the reactor entrance section. The heated gas passes through a bed of nickel catalyst where tar, light hydrocarbons, and ammonia are removed by steam reforming. The main reactions of the RFTC reactor are:

$C_nH_m + nH_2O \Leftrightarrow nCO + (\frac{1}{2} m+n)H_2$ Hydrocarbon reforming

$2NH_3 \Leftrightarrow N_2 + 3H_2$ Reverse ammonia synthesis

$CO + H_2O \Leftrightarrow CO_2 + H_2$ Water-gas shift

A small amount of the syngas is combusted to counterbalance the endothermic tar reforming reactions:

$$H_2 + \frac{1}{2} O_2 \rightarrow H_2O$$

$$CO + \frac{1}{2} O_2 \rightarrow CO_2$$

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

The typical conversion for the RFTC reactor is as follows:

Components	Conversion
Benzene	82
Napthalene	99
Phenol	96
Total Aromatic	94
Total Phenols	98
Total Tar	96
Ammonia	99

26 Don J. Stevens, "Hot Gas Conditioning: Recent Progress with Larger-Scale Biomass Gasification Systems," prepared by Pacific Northwest National Laboratory for NREL, August, 2001

27 Mark A. Paisley, Mike J. Welch, "Biomass Gasification Combined Cycle Opportunities Using the Future Energy *SilvaGas* Gasifier Coupled to Alstrom's Industrial Gas Turbines," ASME Turbo Expo Land, Sea, and Air, Georgia World Congress Center, June 16-19, 2003

28 BTG Biomass Technology Group, "Tar & Tar Removal," 22 March 2004, http://www.btgworld.com/technologies/tar-removal.html

The partial oxidation reaction (POx) was also investigated as a possible process for tar and hydrocarbons removal. In this process, the syngas enters the POx reactor and mixes with oxygen that is at about 300°F. Partial oxidation and reforming reactions occur in a combustion zone where tar, methane, light hydrocarbons, and benzene are converted to CO and H_2. The reformed gas exits the reactor at about 2500°F.

The main disadvantage of POx is a reduction of the product gas heating value. In order to achieve destruction of the tars and oils, a high temperature reactor is required. While it is possible to crack the tars and oils at moderate temperatures, it is very difficult to selectively react methane. However at high temperatures oxidation of CO and H_2 also occur. As a result, the gas composition will be shifted toward a lower H_2:CO ratio.

In order to improve the efficiency of POx, a catalyst can be used to lower the temperature, and hence also the amount of oxidizer required to destroy the tars and oils. A catalytic auto-reformer technology may provide a solution to biomass tar and oil elimination. Such an application would only apply to a particulate-free gas since any particulate in the gas could shortly blind the catalytic reactor. As shown in Table D-1 below, an auto-thermal reformer is essentially a hybrid between POx and steam reforming.

TABLE D-1 COMPARISON OF SYNGAS REFORMING PROCESS TECHNOLOGY

Gas Reforming Process	Typical H_2/CO ratio	Comments
Tar Cracking/Reforming	wide range	Developing technology. Operating information not widely available.
Steam (Methane) Reforming SR or SMR	3-4	Dominant technology for industrial H_2 production Typically high efficiency
Partial Oxidation (POx)	1.7-1.8	Used in refining to upgrade heavy liquid fuels Low efficiency May generate coke or soot
Auto-thermal Reforming (ATR)	2.4-4	Hybrid of POx and SR

D.4 ACID GAS REMOVAL TECHNOLOGIES

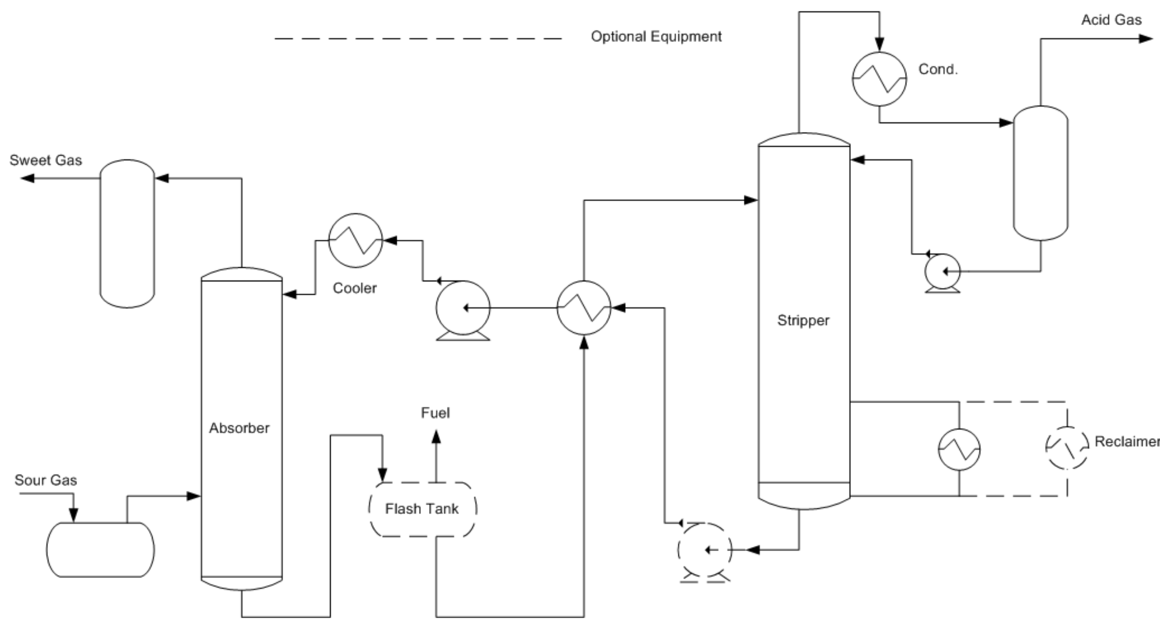
D.4.1 INTRODUCTION

Sulfur contaminants such as H_2S, COS, CO_2, mercaptans, and HCN poison catalysts used in liquid fuel synthesis. Therefore, the syntheses of methanol and FT liquids from syngas require that the sulfur be removed from the syngas to a residual level of 0.10 ppm or less. The syngas considered for this study contains approximately 400 ppmv of H_2S; therefore, acid gas removal is critical in the gas cleanup process. Acid gas removal technologies can be categorized as amine-based, physical solvent, liquid phase oxidation, or catalytic absorbent processes. The type of technology selected is largely influenced by the system operating conditions, the sulfur level in the syngas stream, and the desired purity of the treated syngas. Brief descriptions to explain the overall process for each system are given in the following paragraphs.

D.4.2 AMINE-BASED SYSTEM

Amine processes are proven technologies for the removal of H_2S and CO_2 from gas streams by absorption. Amine systems generally consist of an absorber, a stripper column, a flash separator, and heat exchangers. This is a low-temperature process in which the gas to be treated usually enters the absorber at approximately 110°F. In the absorber, acid gases are removed from the gas stream by chemical reactions with the amine solution. The sweet gas stream exits at the top of the absorber. Regeneration of the rich amine is accomplished through the flash separator to remove absorbed hydrocarbons followed by a stripper column to remove the H_2S and CO_2 from the amine solution. The lean amine solution is cooled and returned to the absorber. The stripped acid gas stream is cooled to recover water and then sent to a sulfur recovery unit. A typical amine system is shown in Figure D-2.

FIGURE D-2 TYPICAL AMINE SYSTEM FLOW DIAGRAM



Amine systems normally operate in the low to medium pressure range of 70-360 psi, although higher pressures can be accommodated with a specific amine solvent. However, in applications where the partial pressure of acid gases is high, the economy of an amine system declines in comparison to other systems. Amine systems can be designed to meet specific flow range and sulfur removal requirements. A sulfur removal level as low as 1 ppm can be achieved but at the expense of operating cost due to the large solvent circulation rate required[29].

There are a variety of amine solutions available. Each offers distinct advantages based on the specific treating condition. Commercially available amine solutions include[30]:

29 Input from GTI, "Gas Cleanup Technologies Discussion," 3 February 2005, office communication
30 GPSA

MEA – Monoethanolamine removes both H_2S and CO_2 from gas streams and is generally used in low-pressure systems and in operations requiring stringent sulfur removal.

DGA – Diglycolamine is used when there is a need for COS and mercaptan removal in addition to H_2S. DGA can hydrolyze COS to H_2S; thus, a COS hydrolysis unit is not needed in the cleanup system.

DEA - Diethanolamine is used in medium- to high-pressure systems (above 500 psi) and is suitable for gas stream with a high ratio of H_2S to CO_2.

MDEA - Methyldiethanolamine has a higher affinity for H_2S than CO_2. MDEA is used when there is a low ratio of H_2S to CO_2 in the gas stream so that the H_2S can be concentrated in the acid gas effluent. If a Claus plant is used for sulfur recovery, a relatively high concentration of H_2S (>15%) in the acid gas effluent is required for optimal Claus operation.

After prolonged use, MEA, DGA, and MDEA solutions accumulate impurities that reduce the H_2S removal efficiency of the solutions. A reclaim unit is needed to remove the impurities in order to improve system efficiency.

One major operating concern for amine systems is corrosion. In water, H_2S dissociates to form a weak acid while CO_2 forms carbonic acid. These acids attack and corrode metal. Therefore, equipment in the amine systems may be clad with stainless steel to improve equipment life.

D.4.3 PHYSICAL SOLVENT SYSTEM

This acid gas removal technology uses an organic solvent to remove acid gases from gas streams by physical absorption without chemical reaction. The driving force of this process is the high solubility of acid gases in the organic solvent. In most cases, solubility increases as the temperature decreases and the pressure increases. Thus, physical absorption is a low-temperature, high-pressure process, with high partial pressure of acid gases required for the economy and efficiency of this process. The temperature of the solvent should be as low as possible while the temperature of the gas to be treated usually enters the absorber at about 100°F. Physical solvent systems normally operate at pressures above 150 psi[31].

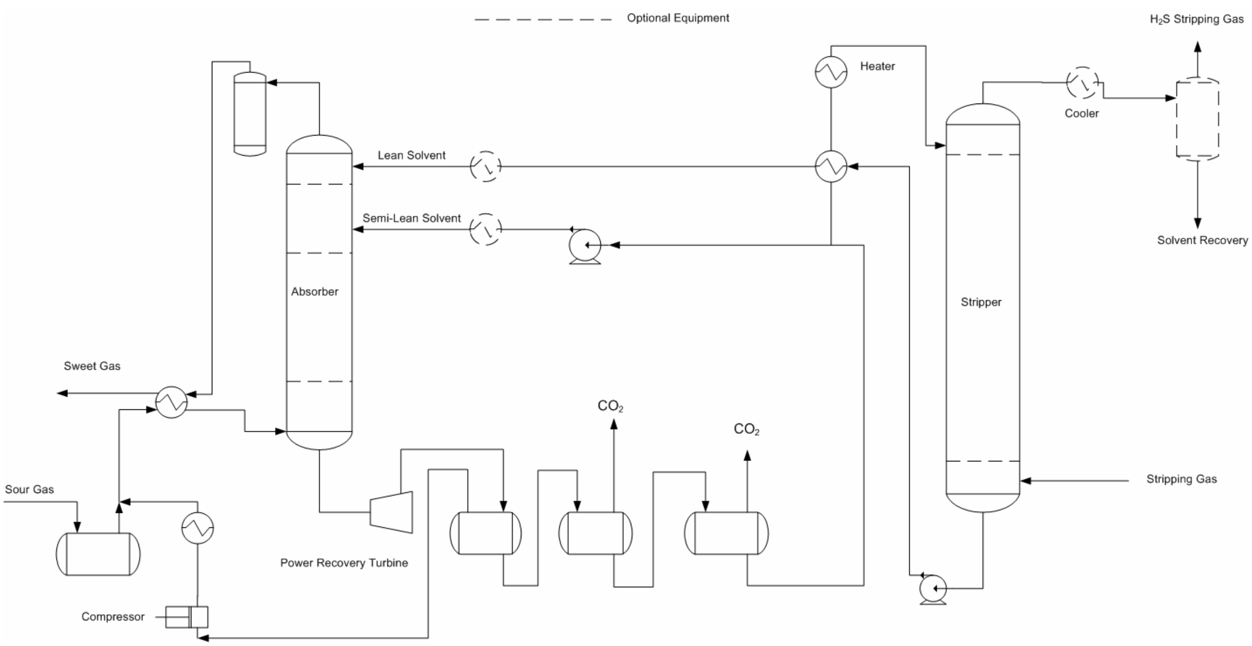
In general, physical solvent systems consist of an absorber, a stripper column, a series of flash separators, and heat exchangers. In the absorber, acid gases in the syngas stream are absorbed into the solvent solution. The sweet syngas stream exits the top of the absorber. Regeneration of the rich solvent stream is accomplished through a series of flash separators at reduced pressures to remove absorbed hydrocarbons followed by the stripper column to remove the acid gases from the solvent. The lean solvent solution is cooled and returned to the absorber. The stripped acid gas stream is cooled to recover water and then sent to a sulfur recovery unit. A typical physical solvent system is shown in Figure D-3.

[31] Gerhard Ranke, "Advantages of the Rectisol-Wash Process in Selective H_2S Removal from Gas Mixtures," 1973, office communication, 30 January 2005

The two common physical systems are Rectisol and Selexol. The Rectisol process, which uses methanol at temperatures < 32°F, can achieve a sulfur removal level as low as 0.1 ppm. The Selexol process, which uses mixtures of dimethyl ethers of polyethylene glycol, can achieve a sulfur removal level of 1ppm[32].

Selection of material of construction depends on the solvent used. For example, stainless steel is required for much of the Rectisol process equipment, contributing to a significant capital cost. In the Selexol process, carbon steel is the standard material of construction, except for those areas with high severity where stainless steel will be used.

FIGURE D-3 TYPICAL PHYSICAL SOLVENT SYSTEM FLOW DIAGRAM



D.4.4 LIQUID PHASE OXIDATION PROCESS -- LO-CAT™

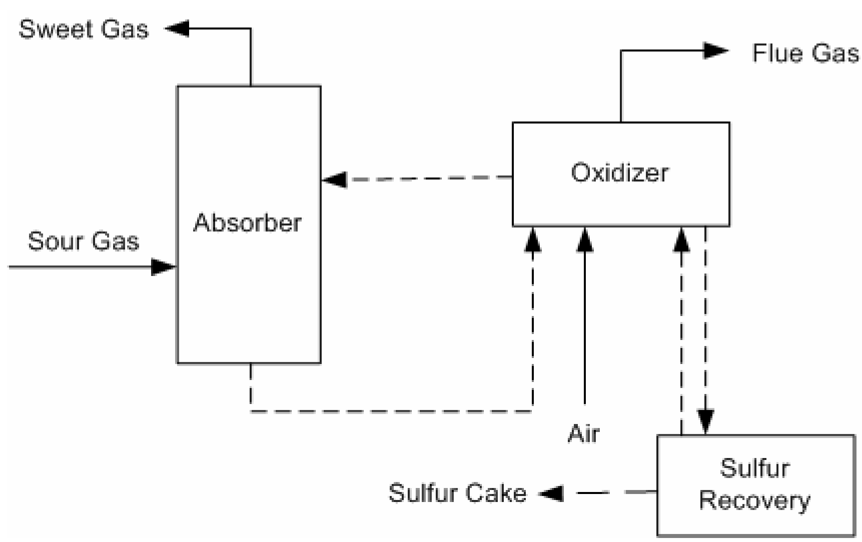
LO-CAT™ is an oxidation process that uses iron catalyst held in a chelating agent to oxidize H_2S to elemental sulfur. H_2S is the only acid gas being removed in this process but a high CO_2 concentration in the feedgas requires caustic for pH adjustment. A LO-CAT™ process consists of 3 sections that include an absorber, an oxidizer for catalyst regeneration, and a sulfur handling unit. Figure D-4 illustrates a typical LO-CAT™ unit. When the gas stream comes in contact with the LO-CAT™ solution in the absorber, H_2S in the gas stream is converted to elemental sulfur. The spent catalyst along with the elemental sulfur exit the absorber, then enter the oxidizer where the spent catalyst is regenerated by contact with oxygen in air, and the elemental sulfur is concentrated into a sulfur slurry. The sulfur slurry moves to the sulfur handling unit where it is washed to recover any entrained catalyst. The sulfur recovered from a LO-CAT™

32 D.J. Kubek, E. Polla, F.P. Wilcher, “Purification and Recovery for Gasification,” Gasification Technologies Conference, October 1996, San Francisco, CA.

process contains a small amount of entrained residual catalyst and is considered low-value sulfur that is suitable for agricultural purposes but is undesirable as a chemical feedstock.

The LO-CAT™ process is suitable for small-scale applications that require less than 20 TPD of sulfur recovery capacity, making the LO-CAT™ a candidate process for this study, which has less than 5 TPD of sulfur recovery. This process can achieve 99.9%+ of H_2S removal efficiency[33]. This process can operate over a wide range of pressures from atmospheric up to 600 psi, but most are low-pressure applications in amine acid gas service. The operating temperature is normally maintained at about 110°F since high temperatures degrade the LO-CAT™ solution that can affect removal efficiency. Advantages of this process include the ability to treat a wide range of gas compositions, a significant turndown flexibility, and less capital costs in comparison to the Claus process with the associated tail gas treating unit.

FIGURE D-4 TYPICAL LO-CAT™ SYSTEM FLOW DIAGRAM



Since LO-CAT™ only removes H_2S, a COS hydrolysis unit upstream of the LO-CAT™ is needed to hydrolyze any COS in the gas stream to H_2S. Other acid gases, such as HCN and mercaptans, would have to be removed by wet scrubbing.

The standard material used for LO-CAT™ systems is stainless steel. Under certain conditions where there is build-up of chloride ions from the feed gas, FRP (fiberglass reinforced plastic) material is used to provide added stability for the stainless steel components[34].

D.4.5 CATALYTIC ABSORBENT—ZnO

ZnO is often used as a polishing step for sulfur removal in gas streams where the sulfur level is below 20 ppmv. In a traditional purification system, illustrated in Figure D-5, ZnO is used in

33 Douglas L. Heguy, Gary J. Nagl, "The State of Iron Redox Sulfur Plant Technology New Developments to an Established Technology," <http://www.gtpmerichem.com/support/technical_papers/state_of_iron_redox.html> (25 January 2005)

34 GTP-Merichem, "FAQ's About Sulfur Removal and Recovery Using the LO-CAT System," <http://www.gtp-merichem.com/support/faq.html> (25 January 2005)

conjunction with hydrogenation catalysts based on cobalt, molybdenum and nickel. This system involves the hydrogenation of sulfur compounds such as mercaptans to H_2S, and halides such as chlorides to HCl. These compounds are then reacted with the ZnO absorbent where H_2S is converted to zinc sulfide, and HCl forms a stable chloride. Additionally, ZnO also removes COS by hydrolysis to form H_2S which is then adsorbed to form zinc sulfide. The general reactions are summarized below[35]:

Hydrogenation reactions:

$RSH + H_2 \rightarrow RH + H_2S$
$RCl + H_2 \rightarrow RH + HCl$

Reaction with ZnO:

$ZnO + H_2S \Leftrightarrow ZnS + H_2O$
$ZnO + COS \Leftrightarrow ZnS + H_2O$

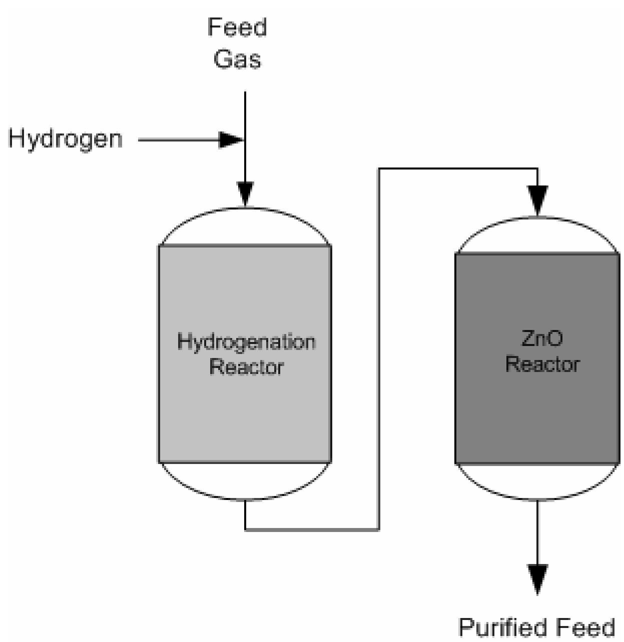


FIGURE D-5 TRADITIONAL ZNO PURIFICATION SYSTEM

A sulfur removal below 50 ppb is attainable with ZnO[36]. Since the sulfur specifications for alcohols and FT liquids are 0.10 ppm or less, ZnO will be used to achieve these requirements. However, a hydrogenation reactor will not likely be required since the syngas stream given by NREL does not contain halogens or any other sulfur compounds other than H_2S.

35 Johnson Matthey Group, "Purification Catalysts and Absorbents for Hydrogen Production," available at http://www.jmcatalysts.com (25 January 2005)
36 Johnson Matthey Group, "Absorbent for Sulphur Polishing," available at http://www.jmcatalysts.com (25 January 2005)

ZnO is active over a wide range of temperatures from ambient to 750°F; however, operating temperatures range between 660°F and 750°F are normally used to maximize absorption efficiency. Operating pressure limits are not a concern for the use of ZnO absorbent. The ZnO reactor is normally constructed from carbon steel clad with stainless steel to prevent corrosion caused by acid gases.

One drawback of this process is the significant operating costs contributed by frequent replacement and disposal of ZnO absorbent since it cannot be regenerated.

D.4.6 COS HYDROLYSIS

COS can be removed simultaneously with H_2S and other acid gases in some of the acid gas removal processes described above. In chemical absorption processes, the degree of COS removal is dependent upon the reactivity of the solvent solution with COS. For example, DGA can remove virtually all of the COS whereas MDEA has little reactivity with COS. In physical absorption processes, the solubility of COS in the physical solvent and the COS partial pressure determine the level of removal. A COS level of 0.1 ppm is attainable with the Rectisol process while the Selexol process can achieve 10 ppm COS[37]. In the ZnO process, approximately 80% of the COS can be removed by hydrolysis.

When COS cannot be effectively removed by the conventional acid gas removal processes, a COS hydrolysis reactor is required and is placed upstream of the acid gas removal unit. COS removal is accomplished by hydrolysis of COS on a catalyst to form H_2S which is sent to the downstream acid gas removal unit. Activated alumina catalysts are often used in these applications. COS removal to 0.1 ppm or below can be achieved[38]. COS hydrolysis reactors can operate over a wide range of pressures with temperatures in the range of 100°F – 450°F. The COS hydrolysis reactor is normally constructed from carbon steel clad with stainless steel to prevent corrosion caused by acid gases.

D.4.7 SULFUR RECOVERY UNIT (SRU)

In the sulfur recovery unit, the acid gas stream from the amine or physical solvent unit is recovered to elemental sulfur. In operations where the sulfur recovery is more than 20 TPD, a Claus SRU is generally an economical approach. However, since the amount of sulfur in the syngas for this study is small (< 5 TPD), a Claus operation would not be a cost-effective solution. For a low sulfur recovery capacity, a LO-CAT SRU would be a more suitable process.

D.5 AMMONIA, ALKALI, AND OTHER CONTAMINANTS

D.5.1 AMMONIA REMOVAL

Two methods for removing ammonia include catalytic tar reforming and wet scrubbing. Tar cracker catalysts have been demonstrated to be effective at reducing ammonia in the syngas stream by conversion to N_2 and H_2. A tar cracker can be used to remove ammonia followed by

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37 Robert Chu, Senior Design Engineer, Nexant, "COS Removal," office communication, 17 February 2005
38 United Catalysts Inc., "UCI COS Hydrolysis Catalysts," 22 June 1992, and office communication, 17 February 2005

gas cooling and a wet scrubber to remove residual ammonia. This cleanup configuration should achieve complete removal of ammonia.

D.5.2 ALKALI REMOVAL

Alkali removal is normally accomplished by cooling the syngas stream below 1100°F to allow condensation of alkali species followed by barrier filtration or wet scrubbing. Corrosion potential should be taken into consideration when using metal or ceramic candle filters due to possible reactions between the alkali and filter materials at high temperatures. Several demonstration facilities had used barrier filters to removal alkali along with other impurities. For example, ceramic filters were used at the Lahti facility in Finland and Varnamo in Sweden[39],[40]. The Varnamo facility experienced breakage of the ceramic filter elements and replaced them with sintered metal filters, which operated successfully. Baghouse filters were used in Lahti's low-pressure gasification system and the FERCO facility in Vermont.

Alkali can easily be removed by wet scrubbing, thus it is often the preferred method for alkali removal. Descriptions of operating and cleanup parameters for barrier filters and wet scrubbing are given earlier in this Appendix.

D.5.3 REMOVAL OF OTHER CONTAMINANTS

Contaminants such as halides or metals (i.e. nickel or iron) are not typical, but may exist in syngas produced from biomass gasification. If present, these impurities can be removed by wet scrubbing or purification by hydrogenation and ZnO absorption.

[39] OPET Finland, OPET Report 4 "Review of Finnish Biomass Gasification Technologies," May 2002
[40] Krister Stahl, et al. "Biomass IGCC at Varnamo, Sweden-Past and Future," GCEP Energy Workshop, 27 April 2004, Stanford University, CA.

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As part of Task 2, Gas Cleanup and Cost Estimates, Nexant investigated the appropriate process scheme for treatment of wood-derived syngas for use in the synthesis of liquid fuels. Two different 2,000 metric tonne per day gasification schemes, a low-pressure, indirect system using the gasifier, and a high-pressure, direct system using gasification technology were evaluated. Initial syngas conditions from each of the gasifiers was provided to the team by the National Renewable Energy Laboraoty. Nexant was the prime contractor and principal investigator during this task; technical assistance was provided by both GTI and Emery Energy.

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