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 – Black Liquor Gasification

Nexant Inc.
San Francisco, California

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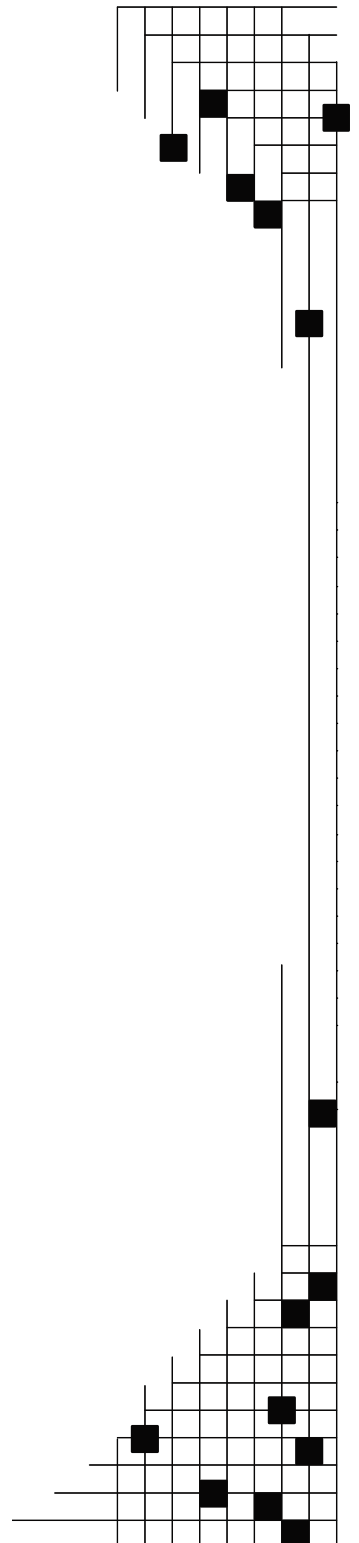
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Executive Summary

As part of Task 2, Gas Cleanup and Cost Estimates, the team investigated the appropriate process scheme for removal of acid gases from black liquor derived syngas for use in both power and liquid fuels synthesis. Two different 3,200 metric tonne per day gasification schemes, both low-temperature/low-pressure (1100°F, 40 psi) and high-temperature/high-pressure (1800°F, 500 psi) were used for syngas production. Initial syngas conditions from each of the gasifiers was provided to the team by NREL and Princeton University. Nexant was the prime contractor and principal investigator during this task; technical assistance was provided by both GTI and Emery Energy.

The initial task was to evaluate the different technologies suitable for acid gas removal from the different syngas streams, with the end-use (power or chemicals) defining the amount of treatment required. After these preliminary investigations, process models were developed for each of the cases that incorporated equipment deemed suitable for sulfur and carbon dioxide removal. Final designs were developed that include unit sizes, energy use, capital and operating costs, and labor requirements. An analysis of the impact that syngas flowrate and composition changes have on the designs was then performed as a sensitivity to the base cases. Finally, Nexant evaluated the possible ways to reintroduce recovered sulfur back into the front-end of the pulping process.

The technologies chosen differed considerably based on the end-use of the syngas. For power applications, the target was to removal sulfur to roughly 100 ppm, while minimizing the amount of carbon dioxide removed. It was determined that COS removal beds, followed by single absorber amine systems, are suitable to meet this specification. A standard Claus and Tail Gas Treating unit is included for recovery of elemental sulfur. For the chemicals synthesis cases, a multi-column physical solvent system (Genosorb, similar in style to Selexol) was selected for deep sulfur removal and to provide suitable H_2S concentration in the acid gas for feed to a standard Claus unit. Zinc oxide beds are incorporated downstream of the Genosorb units as a polishing/guard step. While amine units with zinc oxide beds were originally evaluated for the chemical synthesis designs, their inability to selectively concentrate the H_2S in the acid gas stream made them unsuitable. The results of the analysis for both cases can be seen in Table ES-1 below, with information on the capital cost and utility requirements:

Table ES-1 Syngas Clean-Up Case Summary

	Low-Temp to Power	High-Temp to Power	Low-Temp to Chemicals	High-Temp to Chemicals
Black Liquor Feedrate (MTPD)	3,200	3,200	3,200	3,200
Syngas Rate (Mlb/hr)	196.9	203.1	196.9	203.1
Total Installed Cost ($MM)	36.7	27.8	50.2	55.2
Power Required (MW)	1.6	1.0	5.7	5.7
Net Steam Required (Mlb/hr)	47	32	75	42

Table ES-1 shows how the chemical synthesis cases require significantly more capital and power during operation. The chemicals synthesis cases are more expensive than the power cases due to the need to remove CO_2 with physical solvents, the inclusion of ZnO beds, and the use of a shift reactor in the high temperature case. The need for the shift reactor creates additional carbon dioxide, placing a greater load on the Genosorb system. The pressure in both chemicals cases is lower than what is typically observed in physical solvent systems, leading to higher solvent circulation requirements and cost. The use of ZnO and shift reactors requires additional heat exchangers to be added to the design, further increasing the cost. Finally, the Genosorb system requires solvent refrigeration which increases the system power load.

The syngas from the low temperature gasifier has roughly double the mass of H_2S as the high temperature case, increasing the cost of the acid gas removal and Claus sections of the low temperature plants relative to the high temperature cases. This is the main reason for the cost difference in the power only cases.

Based on the pulping process used, these designs could be incorporated into a pulp and paper facility for recovery of some of the process sulfur. This is especially true of polysulfide processes, where recovered elemental sulfur can be reacted with sodium sulfide to create polysulfide chemicals suitable for pulping uses. The use of other pulping processes may limit the ability to use recovered sulfur until new technologies are developed.

Future studies should focus on the following areas to further define suitable technologies and confirm costs:

- Integration of the sulfur removal section with the other parts of the gasification plant to obtain a better understanding of overall plant costs.

- Alternate methods for CO_2 removal in the chemicals cases. A cost comparison of different physical solvents and other technologies for acid gas removal would provide additional data to confirm the appropriate physical solvent to use in the syngas to chemicals design. Raising the system pressure would decrease the cost of the physical solvent system.

Introduction and Methodology

This study provides designs and costs for sulfur removal from black liquor derived syngas for both liquid fuels synthesis and power generation. Two different syngas compositions, derived from both low (1100°F) and high (1800°F) temperature gasification conditions, were evaluated for a 3200 metric tonne per day plant. The different gasification conditions and end-uses created four distinct cases with different conditioning requirements. The goal was to provide NREL with design packages that included process flow diagrams, equipment requirements, mass and energy balances, capital and operating costs, and labor requirements for inclusion into a larger black liquor conversion study. The study also addressed how the designs would be impacted by changing flowrates and syngas compositions.

The work was divided into two main task areas. The first Subtasks (2.4.1 and 2.4.2) discussed the technologies suitable for sulfur removal for each case, along with major areas of uncertainty that would be studied in greater detail during the design. This rationale for technology selection has been updated and is shown in Section 1. Subtasks 2.4.3 through 2.4.7 consisted of equipment sizing, development of costs, and scaling analysis. This was an iterative process, for some of the technologies initially selected for use were later rejected after performing process modeling and receiving information from vendors. The final design and cost information, reflecting technologies appropriate for each case, is shown in Sections 2 and 3. While the issue of how recovered sulfur can be recycled for use in the pulping process was not quantitatively addressed, process options are discussed in Section 4.

A variety of resources were used throughout the project to produce the final designs. In gathering the initial technology data, previous team studies (especially the wood gasification cases addressed in Subtask 2.2 of this overall study), data on black liquor gasification from Dr. Eric Larson at Princeton University, vendor information, and NREL input were all used to establish the items for consideration. Vendors were especially helpful in providing data for packaged units and unique technologies, such as Ortloff Engineers (Claus sulfur recovery), Uhde/GTI (physical solvent systems), Sud-Chemie (COS beds), and Johnson Matthey (ZnO beds).

HYSYS was used for modeling the overall process. Design and performance for many of the units is based off of previous studies and vendor supplied information. 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 the packaged units and unique technologies mentioned above. Industry derived cost curves were used for the acid gas removal systems and as a check on other process items. Operating costs were developed from vendor supplied information and the system energy balance. Finally, labor requirements are derived from a scale-up of a detailed study put together by Emery Energy specific to biomass gasification. Comparisons are made to the labor estimate produced for the wood gasification case in Subtask 2.2.

1.1   INTRODUCTION

The initial goal for the Nexant team was to evaluate the different sulfur cleanup processes suitable for syngas produced from a 3200 metric tonne per day black liquor gasification process. Two design cases have been explored: meeting the sulfur specifications for power generation (~100 ppm) and chemicals production (~0.1 ppm). The rationale for the technology selection is discussed in this section.

The syngas compositions and sulfur cleanup requirements are listed in Tables 1-1 and 1-2 below. Table 1-1 lists the syngas compositions for the high-temperature gasification case after syngas cooling and heat recovery. For the low-temperature gasification case, the syngas composition listed is after syngas cooling and compression.

Table 1-1 Syngas Compositions and Operating Parameters[1]

	High-Temperature Gasifier Syngas	Low-Temperature Gasifier Syngas
Gasifier Temperature	1,832°F (1000°C)	1,112°F (600°C)
Gasifier Pressure	508 psia (35 bar)	39 psia (2.7 bar)
Syngas Flowrate (lb/hr)	203,133	196,870
Syngas temperature to the sulfur removal unit, °F	252	200
Syngas pressure to the sulfur removal unit, psia	487	332
Compositions	Mol%	Mol%
H_2	36.30	55.68
CO	34.44	21.35
CO_2	18.15	13.80
CH_4	1.88	3.13
H_2O	6.32	3.85
COS	0.07	0.01
H_2S	1.63	2.17
NH_3	0.005	-
Ar	0.87	-
N_2	0.32	-
H_2:CO molar ratio	1.05	2.61

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[1] Information from Dr. Eric Larson, Princeton University

Table 1-2 Sulfur Specifications for Gas Cleanup

Process	Contaminants	Level	Source/Comment
Power Generation	Sulfur	100 ppmv	Typically between 100 and 200 ppm for coal gasification facilities
Methanol Synthesis	Sulfur (not COS)	<0.5 ppmv (<0.1 ppmv preferred)	Kung, 1992

1.2   PROCESS DESCRIPTION AND RATIONALE

1.2.1   Sulfur Removal for Low-Temperature Design Case

Sulfur Removal for Power Generation

The syngas exiting the low-temperature gasifier contains sulfur compounds that must be removed prior to power generation or chemicals production. For the power generation cases, the selected sulfur removal process is a COS hydrolysis reactor followed by an amine unit. The syngas, at 200°F and 337 psia, enters the COS hydrolysis reactor where COS is converted to H_2S. The syngas exiting the COS hydrolysis unit is cooled to the operating temperature of the amine unit, approximately at 110°F. The amine unit is designed to selectively remove H_2S to a concentration of 100 ppmv; after an analysis of possible amine solvents, it was determined that MDEA is appropriate for this application. Selective removal of H_2S without CO_2 removal is desirable in power applications since CO_2 in the syngas increases the volumetric flow to the turbine and boosts power output.

In addition to MDEA, DGA was also evaluated as a possible solvent for the amine unit. Since DGA has the ability to remove COS along with H_2S, its use would allow the elimination of the upstream COS hydrolysis unit, reducing overall capital costs. While simulations run with DGA and vendor input from Huntsman showed that DGA could successfully remove ~90% of the COS, it removes too much CO_2 to be appropriate for a power application.

The acid gas from the amine system is sent to the sulfur recovery unit (SRU). The SRU is a Claus plant with a sulfur recovery capacity of approximately 100 TPD. Although this sulfur recovery rate is low for a typical Claus plant, recovery of pure sulfur is desired for recycle back into the pulping process.

Sulfur Removal for Chemicals Production

For the chemicals production case, the sulfur removal requirement is assumed to be what is required for methanol synthesis. The sulfur removal unit consists of a multi-absorber physical solvent system (Clarient/Uhde's Genosorb system, similar to that of Selexol), followed by ZnO reactor beds as a polishing step. The Genosorb unit is designed to remove the bulk of the H_2S to about 1 ppmv and ~70% of the CO_2. CO_2 removal is necessary for methanol synthesis due to the required stoichiometric ratio of H_2, CO, and CO_2 :

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

Initial investigations into sulfur removal for chemical production focused on chemical solvents, namely DEA. Although DEA can remove sufficient amounts of both H_2S and CO_2, the H_2S content in the resulting acid gas stream is too low for feed to a Claus SRU (a minimum concentration of 40% H_2S in the acid gas effluent is desired for normal Claus operation). Multiple simulations were attempted, including different amine circulation rates and two absorbers, in an attempt to increase the H_2S content. None of these efforts were successful, forcing the team to explore other options for sulfur removal.

The Genosorb process is frequently marketed in Europe as an alternative to the Selexol process. Both Genosorb and Selexol physical solvent processes are based on the use of polyethylene glycol ether solvent mixtures. The Selexol process solvent is a mixture of dimethyl ethers of polyethylene glycol, and has the formulation $CH_3O(CH_2CH_2O)_nCH_3$ where n is between 3 and 9. There are other process suppliers using solvents similar to the Selexol process, such as Clariant GmbH of Germany, which offers a family of dialkyl ethers of polyethylene glycol. The Clariant solvents, under the Genosorb name, include dimethyl and dibutyl ethers of polyethylene glycols.

Treatment in the Genosorb unit removes the H_2S and COS content to roughly 1 ppm. Following the Genosorb unit, the sweet syngas enters the ZnO reactor where residual COS and H_2S are removed at 750°F. The ZnO reactor is used as a polishing step to reduce the sulfur concentration to < 0.1 ppmv.

1.2.2   Sulfur Removal for High-Temperature Design Case

Sulfur Removal for Power Generation

For the power generation case, the sulfur removal process for the syngas from the high-temperature gasifier is similar to that of the syngas from the low-temperature gasifier. The sulfur removal unit consists of a COS hydrolysis reactor followed by an MDEA amine unit. The syngas enters the COS hydrolysis reactor at 252°F and 487 psia. Similar to the low-temperature design case, DGA was also evaluated as a possible solvent. As with the low-temperature case, DGA removes too much CO_2 than what is appropriate for use in power generation.

Sulfur Removal for Chemicals Production

For the chemicals production case, the sulfur removal process is also similar to that of the low-temperature gasifier case, with the exception of the shift reaction step prior to the Genosorb unit. The syngas exiting the high-temperature gasifier has a H_2/CO ratio of 1.05; thus, a portion of the syngas is sent to the shift reactor to adjust the H_2/CO ratio. The shifted syngas is mixed with the unshifted portion, then is cooled before entering the Genosorb unit. The configuration of the Genosorb unit is similar to that of the low-temperature syngas case. A greater amount of CO_2 removal is required in this case due to the CO_2 produced in the shift reactor. The sweet syngas exiting the amine unit then enters the ZnO reactor where the residual COS and H_2S compounds are removed to meet the sulfur requirement of < 0.1 ppmv. As in the low temperature case, a chemical solvent system (amine) could not be configured to remove the necessary acid gases and have an H_2S concentration high enough for feed to a Claus SRU.

1.3   DISCUSSION

1.3.1 Technologies Not Chosen

Physical solvent systems such as Selexol and Rectisol were considered but not originally selected for this process. Only when it was determined that an amine system could not effectively concentrate the H_2S at a high enough level for feed to a Claus SRU was a physical solvent system selected. Generally, physical solvents are considered when there is high system pressure (> 700 psi), or high acid gas partial pressure (> 50 psi).   The syngas in this study has adequate acid gas partial pressure to consider physical solvent systems; however, a chemical solvent system (amine) was originally chosen instead on the basis of comparable sulfur removal capability and cost. The relatively low syngas pressure available in each case decreases the physical solvent removal efficiency and increases solvent circulation requirements. This leads to higher overall capital and operating costs.

A Rectisol system can remove virtually all the acid gas components such as H_2S, CO_2, COS, HCN, and mercaptans to the level required for chemicals production without the need for separate COS hydrolysis and ZnO reactor units. However, the capital investment for a Rectisol system can be significant because of the complexity, power requirements, and use of stainless steel in parts of the design. Since the syngas in this study contains only COS and H_2S, the configurations established in both the power and chemicals designs should be sufficient to meet the sulfur removal requirements.

2.1   INTRODUCTION AND METHODOLOGY

After deciding on the appropriate technology for sulfur removal and recovery, the next step was to size process equipment for sulfur removal from black liquor derived syngas, and obtain cost estimates for individual pieces of equipment. 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 original process selection deliverable (Subtasks 2.4.1 and 2.4.2). Further refinement was made after vendor input and analysis of the process model results.

- Vendor quotes were obtained for unique and specialized equipment such as ZnO catalyst/reactors, Claus sulfur recovery, and all catalysts. Performance information for the physical solvent system, Genosorb, was obtained from Uhde. Cost for the system was estimated from Uhde supplied information, previous physical solvent studies, and industry developed cost estimates for amines versus physical solvents for IGCC applications.

- The performance and energy requirements for the amine units were estimated using commercially available software that is specific for amine unit modeling (ProTreat and HYSYS). In addition, input was provided by vendors for estimating COS removal from the amine systems and for assistance in simulation development. Once performance specifics were obtained, an industry developed cost curve was used for estimating installed cost.

A complete set of PFDs can be seen in Appendix A through D. The design and cost estimates for the high-pressure and low-pressure cases are presented in the Equipment List, which can be seen in Appendix E. The Equipment List groups process equipment by the following categories: reactors, cyclones, vessels, heat exchangers, compressors, pumps, turbines, and packaged units (the amine and sulfur recovery 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 the equipment cost to arrive at the total installed cost. The installation factor was derived based upon previous experience and Bechtel estimates. The total installed cost for each of the cases can be seen in Table 2-1.

Table 2-1 Black Liquor Syngas Sulfur Removal Costs (Installed)

Case	Capital Cost ($MM)
High–Temperature Gasifier to Power	27.8
Low-Temperature Gasifier to Power	36.7
High-Temperature Gasifier to Chemicals	55.2
Low-Temperature Gasifier to Chemicals	50.2

As can be seen in the equipment list (Appendix E), a major source of plant cost is due to the Claus sulfur recovery and Tail Gas Treating units. The cost of these units represents 60 to 70% of the overall plant costs in both power cases. The designs represent a standard Claus and Tail Gas Recovery system; however, the sulfur recovery configuration that is used in a black liquor gasification plant may vary based on the pulping process and technology choice for sulfur recovery. A more detailed explanation of the sulfur recycle options is presented in Section 4.

The chemicals synthesis cases are more expensive than the power cases due to the need to remove CO_2, the inclusion of ZnO beds, and the use of a shift reactor in the high temperature case. The use of ZnO and shift reactors also requires additional heat exchangers to be added to the design, further increasing the cost. In addition, the chemicals cases required the use of more expensive physical solvents for sulfur and CO_2 removal. Although COS beds were no longer necessary due to the ability of the Genosorb system to adequately remove COS, this reduction in cost did not offset the overall expense of the Genosorb unit. Lower solvent temperatures, inclusion of an additional stripping column, higher solvent flow rates, and relatively low system pressure for a physical solvent system all increases the cost of the physical solvent acid gas removal unit when compared to the amine system used in the power cases.

The high-temperature to chemicals case is more expensive than the low-temperature case due to the greater acid gas removal requirement. The shift reactor in the high-temperature case significantly increases the CO_2 content, which, in general, increases the cost of the acid gas removal system. However, the lower pressure of the low-temperature system raises the removal cost on a dollar/ton basis due to higher solvent circulation requirements. Future studies should look at increasing system pressure to lower the physical solvent system cost.

The syngas from the low temperature gasifier has roughly double the mass of H_2S as the high temperature cases. This increase in sulfur content not only increases the cost of the acid gas removal unit, but also significantly increases the cost of the Claus and Tail Gas Treating sections of the plant. The higher sulfur content increases the overall cost in the low-temperature cases by ~$8MM due to the increased cost of the Claus and Tail Gas Treating units.

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. Additional design specific assumptions are outlined below.

Amine System: For the power cases, an amine needed to be used that was selective for H_2S removal and that minimized CO_2 removal. MDEA is appropriate and relatively cost effective in this service. The simulations performed with MDEA show that a single absorber system can work effectively, provided the amine circulation rate and column size are large enough.

Genosorb: A multi-column process was determined to be appropriate for both removing the necessary acid gas and concentrating the H_2S adequately for feed to the Claus unit. The system consists of separate H_2S and CO_2 absorbers in order to remove the appropriate amount of acid gases specified in the design basis. This design is also necessary for flashing off CO_2 captured in the H_2S absorber. Multiple stages of solvent flash may be required for solvent regeneration. In addition, a CO_2 enrichment column may be necessary in certain instances, such as the low temperature case, to pre-load the solvent to prevent too much CO_2 removal from the syngas stream. A general multi-column process flow is shown in the PFD section of this report.

Heat Integration: Detailed heat integration was not practical for this design, since sulfur treating is only a small portion of a large black liquor gasification facility. However, areas have been identified in the PFDs where heat can be recovered via steam generation.

COS Removal Beds: In both power cases, COS hydrolysis beds were required to remove sufficient COS from the inlet syngas stream. Two different types of COS removal catalysts are used. Based on vendor recommendations, a platinum based catalyst has been selected for the high temperature gasification cases, while a chromium catalyst is used in the low temperature cases. Platinum catalysts are best suited for high pressure and streams that contain contaminants such as ammonia.

Sulfur Recovery: Per the scope agreed upon with NREL, a cost has been estimated for sulfur recovery and tail gas treating, with a generic PFD included with each design. The cost estimate is derived from previous vendor quotes for a Claus and Tail Gas Treating unit, scaled to meet the sulfur design requirements. The cost estimate has been confirmed with other literature sources. The actual configuration for sulfur recovery and reuse in the pulping process may vary.

2.3   OPERATING COSTS AND UTILITY REQUIREMENTS

Operating costs and utility requirements can be seen in Tables 2-2 through 2-6. The main sources of costs are catalysts for the ZnO, shift, and COS beds, and steam/power for the acid gas removal and sulfur recovery units. With the exception of the platinum catalyst required in the high temperature power case, catalyst costs are relatively low. Power and steam needs for the acid gas removal and recovery units make up the majority of the operating costs. The physical solvent system requires significantly more power to operate than the amine unit due to the chiller required to maintain low solvent temperatures. This power requirement greatly raises the operating cost of the chemicals cases when compared to the power-only designs.

Nexant has not made assumptions for the yearly operating cost at this time; this cost could vary considerably based on the assumptions made for plant performance and for steam, chemicals, and power costs. For example, the overall gasification plant may have excess power and steam that can be used by the sulfur treating units at low cost. Until this section of the plant is incorporated into a complete black liquor gasification unit, it is not possible to estimate the overall operating costs.

Table 2-2 Catalyst and Chemicals Requirements

Variable	Amount Required	Cost	Notes
COS Removal Catalyst	Low-Temperature to Power Case: 2470 cubic feet High-Temperature to Power Case: 2400 cubic feet	Price: LT to Power: $575/cubic foot (Cr based) HT to Power: $1587/cubic foot (Pt based)	Information provided by Sud-Chemie. Initial fill, then replaced every 5 years. HT case process conditions require the more robust catalyst.
ZnO Catalyst	Low-Temperature to Chemicals Case: 823 cubic feet High-Temperature to Chemicals Case: 818 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.
Sour Shift Catalyst	High-Temperature to Chemicals Case: 150 cubic feet	Price: $575/cubic foot (Sud-Chemie)	Initial fill then replaced every one to two years.
Acid Gas Removal Chemicals	Low-Temperature to Power Case: 337,200 lb/hr lean MDEA High-Temperature to Power Case: 297,500 lb/hr lean MDEA Low-Temperature to Chemicals Case: 1,160,000 lb/hr lean Genosorb High-Temperature to Chemicals Case: 668,000 lb/hr lean Genosorb	Price: MDEA = ~$0.50/lb Genosorb = ~$1.30/lb	Make-up requirements may vary. Chemical cost may vary.

Table 2-3 High Temperature Syngas To Power Case Utility Requirements

Item No	Item Name	Load BHP		Elect. Power	Steam M Pounds per Hour				Water, GPM		Cooling MMBTU/HR	Fuel Gas	
		Norm.	Max (3).	KW	385 psig	35 psig	5 psig	Cond.	Proc.	C.W. circ. (2).	Water	MMBtu/hr (LHV)	
H-100	Syngas Cooler/Steam Generator						(7.0)						
H-101	Amine Precooler									2,009	20.1		
H-103	Lean Solvent Cooler									1,674	16.8		
H-104	Amine Stripper Reboiler					39.9							
H-105	Acid Gas Condenser									1,672	16.7		
P-101	Lean Solvent Pump	230		172									
A-120	Claus Plant	1,111		829	3.5	(5.4)		(21)				29.7	
	TOTAL	1,341		1,000	4	35	(7)	(21)		5,356	54	30	

NOTES: 1. All Figures shown above represent normal utility usage requirements except:
 () indicates normal utility generated
 * 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.

Table 2-4 Low Temperature Syngas To Power Case Utility Requirements

Item No	Item Name	Load BHP		Elect. Power	Steam M Pounds per Hour				Water, GPM		Cooling MMBTU/HR	Fuel Gas	
		Norm.	Max (3).	KW	385 psig	35 psig	5 psig	Cond.	Proc.	C.W. circ. (2).	Water	MMBtu/hr (LHV)	
H-100	Syngas Cooler/Steam Generator						(8.6)						
H-101	Amine Precooler									2,080	20.8		
H-103	Lean Solvent Cooler									2,269	22.7		
H-104	Amine Stripper Reboiler					58.9		(59)					
H-105	Acid Gas Condenser									2,967	29.7		
P-101	Lean Solvent Pump	200		149									
A-120	Claus Plant	1,908		1423	6.0	(9.2)		(36)				50.9	
	TOTAL	2,108		1,572	6	50	(9)	(95)		7,316	73	51	

NOTES: 1. All Figures shown above represent normal utility usage requirements except:
 () indicates normal utility generated
 * 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.

Table 2-5 High Temperature Syngas To Chemicals Case Utility Requirements

Item No	Item Name	Load BHP		Elect. Power	Steam M Pounds per Hour					Water, GPM		Cooling MMBTU/HR	Fuel Gas	Combustion Air
		Norm.	Max (3).	KW	485 psig	385 psig	35 to 80 psig	5 psig	Cond.	Proc.	C.W. circ. (2).	Water	MMBtu/hr (LHV)	MMSCFD
H-110	LT Shift Cooler/ LP Steam Generator							(78.0)						
H-111	Syngas Cooler/ LP Steam Generator	15		11				(31.1)						
H-112	Solvent Precooler										3,095	31.0		
	Genosorb Unit	6,568		4,900			29.0				1,809	18.1		
H-118	ZnO Beds Preheater						29.9							
R-110	LT Shift Reactor				93.7									
A-120	Claus Plant	1,117		833		3.5	(5.4)		(21)				29.8	
	TOTAL	7,700		5,744	94	4	(25)	(31)	(21)		4,904	49		

NOTES: 1. All Figures shown above represent normal utility usage requirements except:
 () indicates normal utility generated
 * 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.

Table 2-6 Low Temperature Syngas To Chemicals Case Utility Requirements

Item No	Item Name	Load BHP		Elect. Power	Steam M Pounds per Hour				Water, GPM		Cooling MMBTU/HR	Fuel Gas	
		Norm.	Max (3).	KW	385 psig	35 or 80 psig	5 psig	Cond.	Proc.	C.W. circ. (2).	Water	MMBtu/hr (LHV)	
H-100	Syngas Cooler/Steam Generator						(8.6)						
H-111	Solvent Precooler									1,558	15.6		
H-117	ZnO Beds Preheater					45.0							
	Genosorb Unit	5764		4300		42.6				2,683	26.9		
A-120	Claus Plant	1,915		1429	6.1	(9.2)		(36)				51.1	
	TOTAL	**7,679**		**5,729**	**6**	**78**	**(9)**	**(36)**		**4,242**	**42**		

NOTES: 1. All Figures shown above represent normal utility usage requirements except:
         () indicates normal utility generated
         * 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.

A few differences exist in the utility requirements between the high and low temperature cases, as well as the power and chemicals cases. In general, the H_2S yield out of the low-temperature gasifier is higher, leading to an increased load on the acid gas removal and sulfur recovery units. A greater net amount of power and steam is therefore necessary in the low temperature cases. The need to remove CO_2 in the chemicals cases changes the type of solvent used, leading to a significantly greater power load. Solvent chilling requirements increase both the power and cooling water supply required in the chemicals cases.

2.4   CHANGING FLOWS, CONDITIONS, AND COMPOSITIONS

As was presented in the wood gasification cases[2], an analysis was performed to estimate the impact that changes in flowrate and composition would have on the black liquor sulfur removal designs. Information for all four design cases are outlined below; the scaling factors appropriate for each piece of equipment are listed in Table 2-7.

2.4.1   Flowrate Impacts

In general, the limits on process equipment sizes are usually the result of manufacturing guidelines, transportation limits, and maintenance restrictions. An analysis of how the designs would change if the throughput was increased by 50% was performed. The effects of this change are discussed below with respect to all four black liquor gasification cases. A discussion of changes due to lower flows (i.e., a minus 50% change in flow) is not included.   Designs of this nature would only make the plant equipment smaller and piping simpler (fewer parallel units).

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[2]   Nexant, "Task 2 Final Report, Gas Cleanup Design and Cost Estimates: Wood Feedstock", as part of Subcontract ACO-5-44027, December 2005

High-Temperature Gasification Cases

Power Production

For the high-temperature gasifier for power production design case, the COS hydrolysis reactor will have to have an additional parallel train or field fabrication performed due to equipment size and weight limitations during transportation. For this reactor, the drum inside diameter would increase from 14 feet to about 17 feet for a flowrate 50% greater than the base case. The outside diameter (including nozzles and flanges) would be well beyond most road transportation limits in the U.S. However, transportation by rail or barge may still be feasible depending on the exact plant location.

Another piece of equipment that may require a parallel train is the syngas cooler/LP steam generator (H-100). In this case, the size limitation of the heat exchanger is 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 for the tube bundle). It then becomes an economic question of bringing in special maintenance equipment during turnarounds, or using smaller, parallel, process equipment.

It is expected that the amine absorber and stripper required for this design will have already exceeded the maximum transport size, and thus require parallel units or field fabrication. Increasing the capacity by 50% may require additional parallel equipment and a more complex pumping manifold.

Chemicals Production

The syngas cooler has the same issue as the cooler in the power case, with a base case tube bundle that would be over 7 feet.   Also, the Genosorb unit, as with the power case, may already require a parallel unit or field fabrication in the base design. Expanding the capacity by 50% will require more parallel equipment and a more complex pumping manifold.

The chemicals case requires additional pieces of equipment that were also evaluated. The LT shift reactor is small enough in the base case that multiple trains would not be required in a scaled-up design. The LT shift cooler/LP steam generator would face the same tube bundle restriction that the syngas cooler does, since the base case exchanger has a diameter over 7 feet.

Low-Temperature Gasification Cases

Power Production

For the low-temperature gasifier for power production design case, the issues raised in the high-temperature gasifier for power production discussion above concerning the COS hydrolysis reactor and amine unit are applicable for this design as well. Units in parallel of field fabrication for these sections will be required.

Chemicals Production

For the chemicals production case, a larger ZnO bed preheater would exceed the general maintenance limits for pulling the tube bundle, and may require a parallel unit. An analysis will need to be made to determine if a parallel bed is worth the additional investment. And as with the high-temperature to chemicals case, it is likely that the Genosorb unit would require a parallel unit or field fabrication.

General Information

Other considerations must be made to take into account the impact that an increase in flowrate will have on the overall plant. A plant that is 50% larger will require more plot area due not only to the larger equipment and storage requirements, but due to offsite considerations. The utility requirements will need to be scaled accordingly. In addition, 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 operational units.

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 2-7 gives typical values of *n* for most of the equipment included in these designs.

Table 2-7 Examples of Typical Exponents for Equipment Cost Versus Capacity[3,4,5,6,7]

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
Equipment	**Size Range**	**Units**	**Exponent****
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 25-200	hp	0.56 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.

2.4.2   Composition Impacts

The designs presented here are most sensitive to changes in CO and COS. It does not appear that changes significant enough in CO_2, H_2S, or methane concentrations could occur that would change the designs significantly. The expected rates of fluctuation in these components could be handled by the current designs.

Increasing the CO concentration will have a significant influence on the COS hydrolysis reactor for the low-temperature to power case. At a higher CO concentration, a different catalyst (platinum on alumina) would be recommended by the vendor to provide a greater efficiency and appropriate performance. The platinum based catalyst will be more expensive, but a smaller reactor will be necessary due to the higher space velocity necessary. Changes in the H_2/CO ratio

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

could also impact the chemicals cases, and may require different conditions to assure the appropriate ratio for synthesis is maintained.

An increase in the COS concentration could be significant since the COS reactors are already at the limit for most forms of transport (14 feet). An elevated amount of COS entering the reactor would require either more frequent catalyst changeouts, a change in the way the reactor is fabricated, or a parallel train.

2.5 FOLLOW-UP AND AREAS FOR FURTHER STUDY

The analysis performed in this section of the project sets the base case for the sulfur removal section of four different black liquor gasification designs. After in-depth analysis of these cases, the team has identified a number of areas for further study:

- The original vendor quotes obtained for COS removal used a platinum catalyst for both power designs. Only through in-depth discussion with the vendor was it determined that a chromium catalyst could be substituted for some process conditions. A greater investigation of the assumed process conditions and discussions with other vendors will help to confirm if the more expensive platinum catalyst is necessary.

- Integration of the sulfur removal section with the other parts of the gasification plant to get a better picture of overall plant costs.

  - An analysis should also be performed for how the result of this analysis fits with previous black liquor gasification studies, such as the Larson papers[8]. Previous studies have spent little time on this topic; integrating the detailed results from this study will provide more insight and better accuracy for the actual facility costs.

- Alternate methods for CO_2 removal in the chemicals cases. A cost comparison of different physical solvents and other technologies for acid gas removal would provide additional data to confirm the appropriate physical solvent to use in the syngas to chemicals design. Use of amine has already been ruled out due to the inability to concentrate H_2S in the feed stream to the Claus unit.

- Appropriate assumptions for operating costs, including availability, catalyst life in each major reactor, and a determination of if steam and power could be exported from elsewhere in the gasification plant.

- An investigation into the ability to raise the syngas pressure in the chemicals cases to lower the cost of the acid gas removal system. While raising system pressure will either require different gasification design specifications or downstream compression, these changes may be justified based on the decrease in unit costs of the physical solvent acid gas removal system.

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[8] Larson E.D., S. Consonni, and R.E. Katofsky (2003). "A Cost-Benefit Assessment of Biomass Gasification Power Generation in the Pulp and Paper Industry," Princeton University, October 3, Page 96.

3.1   SUMMARY

The work completed in the previously referenced wood gasification designs acts as a base case to estimate the labor requirements for the four black liquor derived syngas cases. Because this design task focused on only a small portion of the overall black liquor conversion facility, a number of assumptions had to be made for the rest of the plant. Differences with the wood gasification cases were identified and will be highlighted below.

The two major items investigated relative to the wood gasification cases were the plant size and the level of complexity. For labor estimating purposes, the black liquor plant size is considered roughly similar to the wood gasification cases. While a greater amount of mass is being fed to the gasifier (3200 MTPD of black liquor, versus 2000 MTPD of wood), the amount of syngas entering the sulfur removal block is actually lower (~200 Mlb/hr versus 227 to 272 Mlb/hr). Based on these numbers, the plant size for labor purposes is considered the same.

All the black liquor designs are also estimated to have roughly the same complexity as the wood gasification cases. While the black liquor gasification plants have additional sulfur treatment requirements (COS beds, Claus units, and two absorber acid gas treatment units for the chemicals cases), some units that exist in the wood gasification cases, such as feedstock handling and the LO-CAT unit, are no longer required. In addition, the cases producing power would require a gas turbine and HRSG, but would no longer need units specific to chemicals synthesis, such as additional syngas compressors, reactors, and certain heat exchange equipment. For these reasons, the complexity between the power and chemicals cases is also considered similar.

In summary, the labor estimate for the chemicals cases is estimated to be the same as for the wood gasification cases, $2,539,720/year. For the power cases, it is not expected that a dedicated on-site laboratory would be required. Eliminating the laboratory manager and technician needs will cut the labor requirements by $220,800/year, to $2,318,920/year total labor costs.

3.2   LABOR REQUIREMENTS

The following labor categories and positions will be required for the 3200 MTPD black liquor gasification plant. It is assumed that this facility is connected to, but independent from, the pulp and paper plant. Having a dedicated feed source to the gasification plant will eliminate the need for feedstock handling and storage facilities. If the plant is independent, a different management team and overhead rate is necessary than what may be used in the pulping plant. If both the pulping and gasification facility was owned and operated by the same entity, it is possible that management, overhead, and technical labor costs could be reduced. Since black liquor gasification is not a widely accepted commercial technology that will likely be owned and operated by a group distinct from the existing pulp and paper facility, this synergy is not assumed at this time. Plant optimization studies should keep this potential cost reduction in mind.

The information below was largely presented in the wood gasification final report. Slight modifications have been made for the black liquor cases.

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. Necessary in chemicals case ONLY.**

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. Necessary in chemicals case ONLY.**

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 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, holidays, training, ongoing maintenance and periodic outage/turnaround planning. In addition, the fifth crew supports updates to control system programming, data collection, and instruments. 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 Black Liquor Plant 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** (**Chemicals Only)	1	$50,000	240	$30	$57,200
Laboratory Technician** (**Chemicals Only)	2	$35,000	240	$22.50	$80,800
Shift Superintendent	5	$75,000	680	$45	$405,600

Position	Number	Base Salary or Hourly Rate	Annual Overtime and Holiday Hours	Overtime Rate	Total Annual Cost
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—Chemicals)					**$2,539,720**
Total Labor and Benefits (Operating Labor Cost—Power)					**$2,318,920**

3.3   DIFFERENCES WITH WOOD GASIFICATION CASES

Two major items were considered when estimating the labor differences between the wood and black liquor gasification cases: size and complexity impacts. Note that a number of assumptions have been made in both sets of cases since only a portion of the plant was designed by the Nexant team. The labor estimates provided here are for the entire gasification facility, assuming a stand-alone plant with a dedicated feed source. After these designs are integrated into a complete facility, these assumptions should be checked for their validity.

3.3.1   Size

The wood gasification final report demonstrated that a larger gasification facility will only require an incrementally larger labor force as the size increases. The size related differences between Emery Energy's 70 MWe facility (which the labor estimate was originally based on) and the wood gasification cases were relatively minor. In the black liquor gasification cases, the total feed to the gasifier is larger relative to the wood gasification cases. However, gasifying a liquid feedstock is very different than a solid one. The total amount of syngas exiting the gasifier in the black liquor cases varies considerably depending on the type of gasification scheme used. The difference is largely due to the amount of entrained water assumed in the high-temperature case.

In order to perform a more direct comparison, Nexant looked closely at the water-free syngas flowrate entering the sulfur removal units. The black liquor syngas flowrate, ~200 Mlb/hr, is 10 to 35% lower than in the wood gasification cases. Since the main reason for the high syngas flowrate exiting the gasifier in the high-temperature case is only due to the amount of entrained water, this should not add considerably to the plant labor requirements. While it is possible that this lower water-free flow rate may lead to slightly lower labor requirements, it is likely that the amount of labor will be similar. Therefore, it was assumed that the size of the facility being considered in the black liquor gasification case is the same as the wood gasification case for labor requirements.

3.3.2   Complexity

The unit operation requirements in the black liquor and wood gasification cases were compared against each other to determine the complexity impacts. The major differences in the black liquor relative to the wood gasification designs can be seen in Table 3-2.

Table 3-2 Complexity Differences Between Black Liquor and Wood Gasification Cases

	All BL Cases	BL Power Cases	BL Chemicals Cases
Units Added	Claus S Recovery and Tail Gas Treatment	Gas Turbine HRSG COS Removal Beds	Shift Reactor (HT Case Only) Genosorb Unit
Units Removed	LO-CAT Major Feed Handling Equipment	Chemicals Synthesis Equipment Methanol Compressor	Amine Unit

The shift and COS removal beds do not add a significant amount of complexity to the black liquor designs, but the Claus, Tail Gas Treating, and Genosorb units do. This is balanced by removing the LO-CAT sulfur recovery unit, much of the feedstock handling equipment, and the amine system in the chemicals cases. Because of the assumption that a dedicated black liquor stream will be available for the gasifier, much of the weighing, crushing, and drying equipment necessary for the wood gasification cases has been removed. The net impact is that in general, the black liquor cases do not add to the design complexity.

The introduction of a gas turbine and HRSG adds significantly to the complexity of a power production case. However, once all the equipment necessary for chemicals synthesis, such as a dedicated syngas compressor, multiple reactors, preheat exchangers, recycle streams, and sampling stations, are removed, the net impact is estimated to be minor. In fact, the laboratory supervisor and technicians deemed necessary in the chemicals cases can be eliminated from the power only case. Any sampling of the feedstock is assumed to be done at the pulping facility.

As part of the project scope, Nexant was asked to consider the possible ways to reintroduce recovered sulfur back to the front-end of the pulping process. Being able to turn sulfur recovered from the syngas back into the chemicals used in the pulping process will increase process efficiency, eliminate waste disposal, and improve the environmental footprint. While Nexant did not evaluate the process economic or overall plant impacts, a number of potential recovery options, tied to different pulping processes, are outlined here for future study.

4.1   INTRODUCTION

The pulp and paper industry currently uses Tomlinson boilers to produce energy from the black liquor produced in the Kraft pulping process. A number of researchers and professionals advocate replacing Tomlinson boilers in the pulp and paper industry with gasifiers in order to more efficiently create energy from the black liquor produced in the Kraft pulping process.

While initial studies have shown that black liquor gasification has potential, there are some concerns with the technology. One concern involves the loss of sulfur that is present as Na_2S in both the white liquor and the black liquor as H_2S in the gasifier. A low temperature gasifier and a high temperature gasifier have been shown to lose 44% and 16% of the Na_2S present in the black liquor to the syngas, respectively[9]. This H_2S can be captured and converted to elemental sulfur, but the method in which this sulfur should be recycled is not established. The following summarizes possible methods for capturing and recycling the sulfur.

4.2   PULPING METHODS

The following information comes from "A Cost-Benefit Assessment of BLGCC Technology" from Eric Larson et al10. Of the four alternative ways of pulping, each requires a different way to recycle sulfur:

- *Polysulfide Pulping:* Polysulfide can be produced by capturing H_2S from the scrubber, converting it to elemental sulfur in a Claus plant, and then remixing the sulfur with white liquor. Subsequent use of polysulfide in the digester results in pulp yield increases on the order of 1–2%[11].

- *Split Sulfide Pulping:* Green liquor from the gas scrubber will have a high sulfide content and a low alkali content. Impregnation of wood chips with this stream will satisfy one of the well-known pulping rules-of-thumb, which recommends a high concentration of free hydrosulfide ions in the initial phase of the cook. The resulting pulp should have less residual lignin that is difficult to dissolve, and the strength will be enhanced by about 5%.[12]

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9 Eric Larson. "A Cost Benefit Assessment of Biomass Gasification Power Generation in the Pulp and Paper Industry." October 8, 2003. S6

10 Eric Larson et al. "A Cost-Benefit Assessment of BLGCC Technology." TAPPI Journal. Vol. 83 No. 6.

11 Jarneel, H., TAPPI 1994 Pulping Conference Proceedings, TAPPI PRESS,Atlanta, 1994, p. 781.; Prasad, D. Y., Jameel, H., and Gratzl, J., "Extended delignification of hardwood with anthraquinone and polysulfide," TAPPI 1995 Pulping Conference Proceedings,TAPPI PRESS, Atlanta, 1995.

12 Teder, A. and Tormund, D., Tappi J. 72(5): 205(1990).

- *Alkaline Sulfite and Anthraquinone (ASAQ) Pulping:* ASAQ pulping has been reported to give substantially higher yields than Kraft processes[13]. Pulping processes based on alkaline sulfite and anthraquinone also have much lower potential for producing odor than Kraft processes have. The problem with implementation of ASAQ pulping has been the inability to regenerate sodium sulfite with the Tomlinson recovery cycle. Recovery of H_2S from the gasifier makes it possible to regenerate part of the sulfite.

- *Hydrogen Sulfide Impregnation:* Researchers at STFI have shown that impregnation of Scandinavian softwood chips with green liquor in the presence of 0.3-MPa hydrogen sulfide results in a 5% yield increase[14].

Sulfur recycle options for different pulping methods, along with the benefits and drawbacks, are shown in Table 4-1.

Table 4-1 Sulfur Recycle Options for Various Pulping Methods

Method	Benefits	Drawbacks
Polysulfide Pulping Sulfur Recycle Options		
Option #1: Capture H_2S with conventional sulfur removal technology (amine, Selexol, or Rectisol) and convert to elemental S with a Claus Unit. Dissolve elemental sulfur in Na_2S at moderate temperature (<100ºF). This creates polysulfide (Na_2-S_3-S) that can be used to digest feedstock for polysulfide pulping[15].	▪ Polysulfide pulping has a higher digester yield than conventional pulping for a fixed pulp production. ▪ Pulp yield increases are on the order of 1-2%[16]. ▪ Conventional sulfur removal technologies are well established. ▪ CO2 co-absorption can be avoided when H2S is converted to elemental sulfur. Avoiding this decreases the causticizing load compared to Options #2 and #3[17].	▪ Increased causticizing load due to increased amount of Na_2CO_3 leaving the gasifier in the condensed phase. ▪ Decreases the alkali content of white-liquor.
Option #2: Capture H_2S with green liquor as scrubbing medium. Then, convert the green liquor to white liquor with a causticizing cycle[18].	▪ No need for sulfur removal technology or Claus. ▪ Polysulfide pulping has a higher digester yield than conventional pulping for a fixed pulp production.	▪ Not as established a process as Selexol, Rectisol, or amine. Experimental testing needs to be performed to determine if this process is appropriate. ▪ Increased causticizing load due to CO_2 in the syngas being co-absorbed in the green liquor. The co-absorbed CO_2 will form additional Na_2CO_3 in the liquor that must be eventually converted back to

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[13] Ingruber, O.V., Kocurek, M.J., and Wong, A., Pulp and Paper Manufacture, Vol. 4: Sulfite Science and Technology, Joint Textbook Committee of the Paper Industry, TAPPI and CPPA,Atlanta, Montreal, 1985, p. 46.

[14] Olm, L., Tormund, D., and Gidnert, E. B., TAPPI 1998 Breaking the Pulp Yield Barrier Symposium Proceedings, TAPPI PRESS, Atlanta, 1998, p. 69.

[15] ibid. S5

[16] Eric Larson et al. "A Cost-Benefit Assessment of BLGCC Technology." TAPPI Journal. Vol. 83 No. 6.

[17] Causticizing load: required lime kiln capacity and lime kiln fuel consumption per unit of black liquor solids.

[18] Larson (2003) S5-6

Method	Benefits	Drawbacks
		NaOH. ▪ Decreases the alkali content of the white-liquor.
Option #3: Capture H_2S with white liquor and convert completely to white liquor with causticizing cycle[19].	▪ No need for sulfur removal technology or Claus. ▪ Polysulfide pulping has a higher digester yield than conventional pulping for a fixed pulp production.	▪ As with Option #2, not as established a process as Selexol, Rectisol, or amine. Experimental testing needs to be performed to determine if this process is appropriate. ▪ Increased causticizing load due to CO_2 in the syngas being co-absorbed in the white liquor. The co-absorbed CO_2 will form additional Na_2CO_3 in the liquor that must be eventually converted back to NaOH. ▪ Decreases alkali content of white liquor.
Split Sulfide Pulping Sulfur Recycle Options		
In this process, the sulfided white liquor streams are split during the pulping process. To recycle the sulfur, H_2S can be captured by any method, then recycled to 2 different white liquor streams that differ by their sulfur content[20,21].	▪ Split sulfide pulping has a higher digester yield than conventional pulping for a fixed pulp production. It can increase digest yields by about 2%, compared to ~7% for polysufide pumping. ▪ Could use conventional technology for H_2S removal.	▪ Unproven method. ▪ Would require a fairly pure H_2S stream.
Other Sulfur Recycle Options		
Direct causticization with TiO_2 in gasifier and removal of sulfur with a regenerative calcium based process[22].	▪ 4.5% increase in overall energy efficiency compared to conventional recovery ▪ Elimination of lime cycle (Annual savings of 100,000 barrels of oil for a 1,000 TPD mill) ▪ Increased carbon conversion in gasifier ▪ Conversion of sulfate to H_2S ▪ Increased tar conversion	▪ Large amount of titanate solids handling ▪ Disintegration of titanate particles during leaching ▪ Agglomeration of bed particles ▪ May not be appropriate for all gasifier configurations

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[19] Ibid. S5-6

[20] Adriaan van Heiningen. "Causticizing Aspects and Sulfur Recovery in Black Liquor Gasification." Colloqium on Black Liquor Combustion and Gasification." Black Liquor Gasification Colloquium Presentations 2003.

[21] Jean Ranard. Hasan Jameel. "Pulping Options with Black Liquor Gasification." Black Liquor Gasification Colloquium Presentations 2003.

[22] Ibid.

Appendix A   High-Temperature Gasifier to Power Design Case PFDs

The Claus Sulfur recovery unit shown here is for general information purpose only, and was not explicitly sized as part of the project scope. A cost estimate for this unit and a Tail Gas Treating process is provided in the equipment list.

R-100
COS Hydrolysis
14' ID x 15'-9" T/T

H-100
Syngas Cooler/
LP Steam Generator
Duty: 6.8 MMBTU/hr

H-101
Amine Precooler
Duty: 20.1 MMBTU/hr

100 (NOTE 1)

COOLED SYNGAS FR. HIGH-
TEMPERATURE BLACK LIQUOR
GASIFIER

R-100
COS Hydrolysis

201

LP STEAM

CW

101   102   103

V-101
A-101
SYNGAS TO AMINE KO DRUM

H-100
Syngas Cooler/
LP Steam Generator

H-101
Amine Precooler

200

BOILER FEED WATER
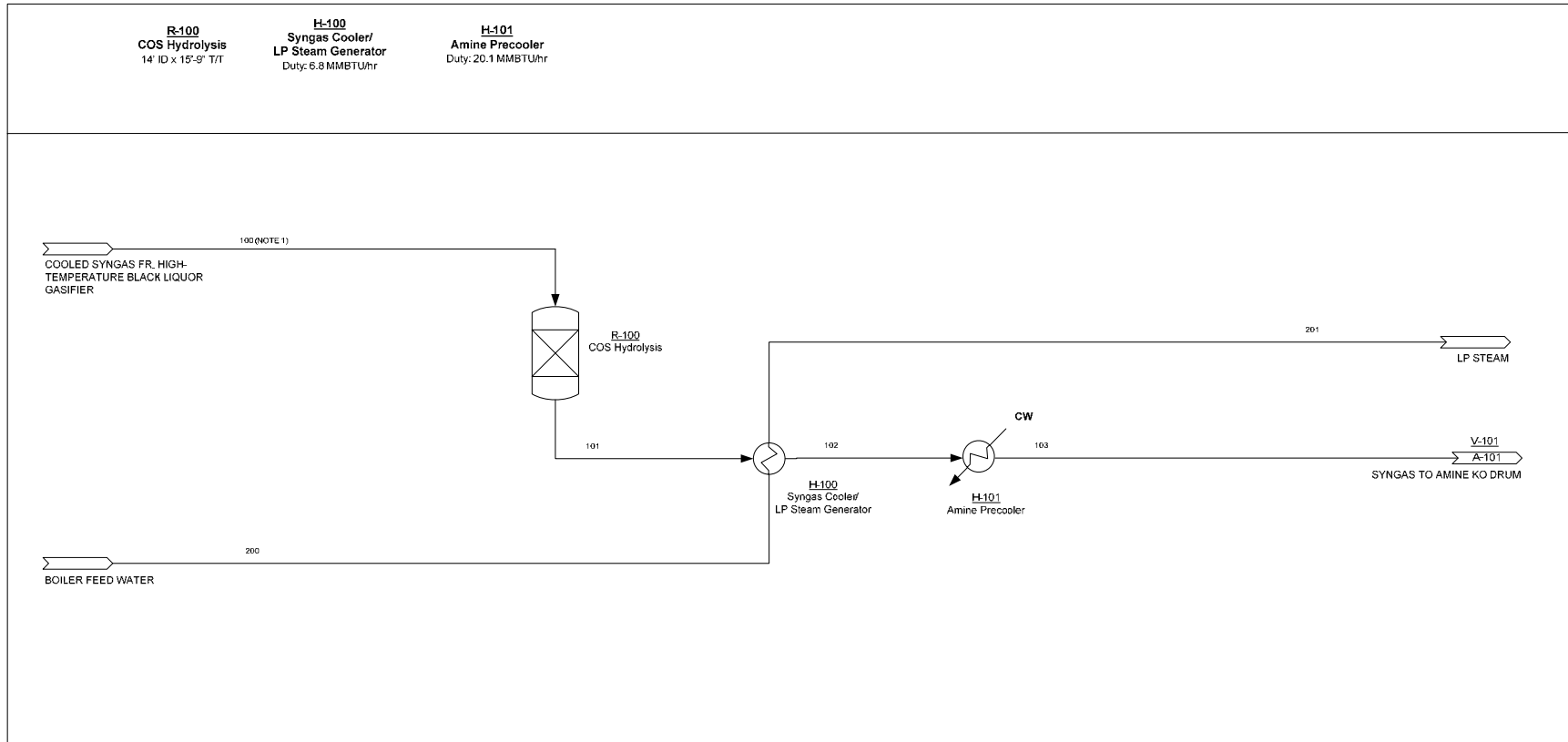
	Unit	100	101	102	103	200	201
Mass Flow	lb/hr	203,133	203,133	203,133	203,133	6,963	6,963
Temperature	F	325	326	240	110	220	230
Pressure	psia	487.3	482.3	477.3	472.3	25.0	20.0
Vapour Fraction		1.00	1.00	1.00	0.94	0.00	1.00
Molecular Weight		20.8	20.8	20.8	20.8	18.0	18.0
Heat Flow	MMBtu/hr	-513.0	-513.0	-519.8	-539.9	-46.4	-39.6
Components							
Hydrogen	lb/hr	7,133	7,133	7,133	7,133		
CO	lb/hr	94,013	94,013	94,013	94,013		
CO2	lb/hr	77,825	78,119	78,119	78,119		
Methane (CH4)	lb/hr	2,962	2,962	2,962	2,962		
H2O	lb/hr	11,094	10,973	10,973	10,973	6,963	6,963
COS	lb/hr	412	11	11	11		
H2S	lb/hr	5,416	5,644	5,644	5,644		
Ammonia (NH3)	lb/hr	8	8	8	8		
Argon	lb/hr	3,396	3,396	3,396	3,396		
Nitrogen	lb/hr	874	874	874	874		

NOTES:
1. THE OPERATING CONDITIONS OF STREAM 100 IS
AT DOWNSTREAM OF THE HEAT RECOVERY UNIT.

No.	DATE	REVISIONS				
SCALE		DESIGNED		DRAWN		

Nexant
San Francisco

NREL BLACK LIQUOR GASIFICATION
HIGH-TEMPERATURE SYNGAS
FOR POWER GENERATION

PROCESS FLOW DIAGRAM
A-100: COS HYDROLYSIS

JOB No.	DRAWING No.	REV
NREL Contract
ACO-5-44027		REV 0

28 Aug, 2005

Equipment list (top of diagram):

| C-101 Amine Absorber | C-102 Stripper | P-101 Lean Solvent Pump 650 GPM 230 BHP | P-102 Reflux Pump 31 GPM | H-102 Rich Solvent/Lean Solvent Exchanger Duty: 27.6 MMBTU/hr | H-103 Lean Solvent Cooler Duty: 16.8 MMBTU/hr | H-104 Stripper Reboiler Duty: 36.9 MMBTU/hr | H-105 Acid Gas Condenser Duty: 16.7 MMBTU/hr | V-101 Amine KO Drum | V-102 Sweet Gas KO Drum | V-103 Rich Solvent Flash Drum | V-104 Reflux Accumulator |

	Unit	103	104	105	106	107	108	110	111	112	113	114	115	116	117	118	119	120	121	122	123
Mass Flow	lb/hr	203,133	192,683	10,450	161,146	161,146	27,998	337,030	368,552	5,760	362,792	362,792	369,416	34,622	43,435	15,437	334,794	334,794	334,794	39,850	39,850
Temperature	F	110	110	110	124	124	130	110	171	156	156	200	245	258	206	130	258	168	169	298	298
Pressure	psia	472.3	472.3	472.3	468.4	468.4	17.1	472.3	472.3	30.0	30.0	25.0	25.0	30.0	22.1	17.1	30.0	30.0	472.3	65.0	60.0
Vapour Fraction		0.94	1.00	0.00	1.00	1.00	1.00	0.00	0.00	1.00	0.00	0.06	0.00	1.00	1.00	0.00	0.00	0.00	0.00	1.00	0.00
Molecular Weight		20.8	21.0	18.1	19.1	19.1	38.5	31.3	32.0	39.3	31.9	31.9	29.5	18.3	27.4	18.0	31.5	31.5	31.5	18.0	
Heat Flow	MMBtu/hr	-539.9		-70.6																	
Components																					
Hydrogen	lb/hr	7,133	7,133	0	7,129	7,129	0	-	4	4	0	0	-	-	0	0	-	-	-		
CO	lb/hr	94,013	94,012	0	93,976	93,976	0	-	39	39	0	0	-	-	0	0	-	-	-		
CO2	lb/hr	78,119	78,073	47	52,003	52,003	21,145	237	26,311	4,929	21,382	21,382	945	708	21,154	9	237	237	237		
Methane (CH4)	lb/hr	2,962	2,962	0	2,960	2,960	0	-	2	2	0	0	-	-	0	0	-	-	-		
H2O	lb/hr	10,973	586	10,388	762	762	1,720	168,176	167,974	314	167,660	167,660	199,757	33,817	17,140	15,421	165,940	165,940	165,940	39,850	39,850
COS	lb/hr	11	11	0	11	11	0	-	0	0	0	0	-	-	0	0	-	-	-		
H2S	lb/hr	5,644	5,633	11	32	32	5,133	14	5,617	470	5,146	5,146	20	6	5,140	7	14	14	14		
Ammonia (NH3)	lb/hr	8	4	3	4	4	-	-	-	-	-	-	-	-	-	-	-	-	-		
Argon	lb/hr	3,396	3,396	0	3,394	3,394	0	-	2	2	0	0	-	-	0	0	-	-	-		
Nitrogen	lb/hr	874	874	0	874	874	0	-	0	0	0	0	-	-	0	0	-	-	-		
MDEA	lb/hr	-	-	-	0	0	-	168,604	168,604	0	168,604	168,604	168,694	90	0	0	168,604	168,604	168,604		

NREL BLACK LIQUOR GASIFICATION
HIGH-TEMPERATURE SYNGAS
FOR POWER GENERATION

PROCESS FLOW DIAGRAM
A-101: AMINE SYSTEM

Nexant
San Francisco

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 0

29 Aug. 2005

Task 2: Gas Cleanup Design and Cost Estimates
Black Liquor Sulfur Removal Design and Cost Estimates
United States Department of Energy/National Renewable Energy Laboratory

A-3

R-120	H-120A	H-121A	R-121A	R-121B	H-120B	H-121B
Reaction Furnace	Sulfur Condenser	Reheater	Catalytic Converter	Catalytic Converter	Sulfur Condenser	Reheater

ACID GAS FR.
AMINE UNIT

AIR

Reaction
Furnace

R-120
Reaction Furnace

BFW

H-120A
Sulfur Condenser

STM

H-121A
Reheater

R-121A
Catalytic Converter

CW

H-120B
Sulfur Condenser

STM

H-121B
Reheater

R-121B
Catalytic Converter

CW

H-120C
Sulfur Condenser

Tail
Gas

Spent
Catalyst

Sulfur

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

SCALE: DESIGNED: DRAWN:

Nexant

San Francisco

**NREL BLACK LIQUOR GASIFICATION
HIGH-TEMPERATURE SYNGAS**

**PROCESS FLOW DIAGRAM
A-120: SULFUR RECOVERY**

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

28 Aug, 2005

Task 2: Gas Cleanup Design and Cost Estimates
Black Liquor Sulfur Removal Design and Cost Estimates
United States Department of Energy/National Renewable Energy Laboratory

A-4

Appendix B Low-Temperature Gasifier to Power Design Case PFDs

The Claus Sulfur recovery unit shown here is for general information purpose only, and was not explicitly sized as part of the project scope. A cost estimate for this unit and a Tail Gas Treating process is provided in the equipment list.

R-100
COS Hydrolysis
13' ID x 18'-6' T/T

H-100
Syngas Air Cooler
Duty: 8.4 MMBTU/hr

H-101
Amine Precooler
Duty: 20.8 MMBTU/hr

100 (NOTE 1)

COOLED SYNGAS FR. LOW-
TEMPERATURE BLACK LIQUOR GASIFIER

R-100
COS Hydrolysis

201

LP STEAM

101

102

103

CW

V-101
A-101
SYNGAS TO AMINE KO DRUM

H-100
Syngas Cooler/
LP Steam Generator

H-101
Amine Precooler

200

BOILER FEED WATER
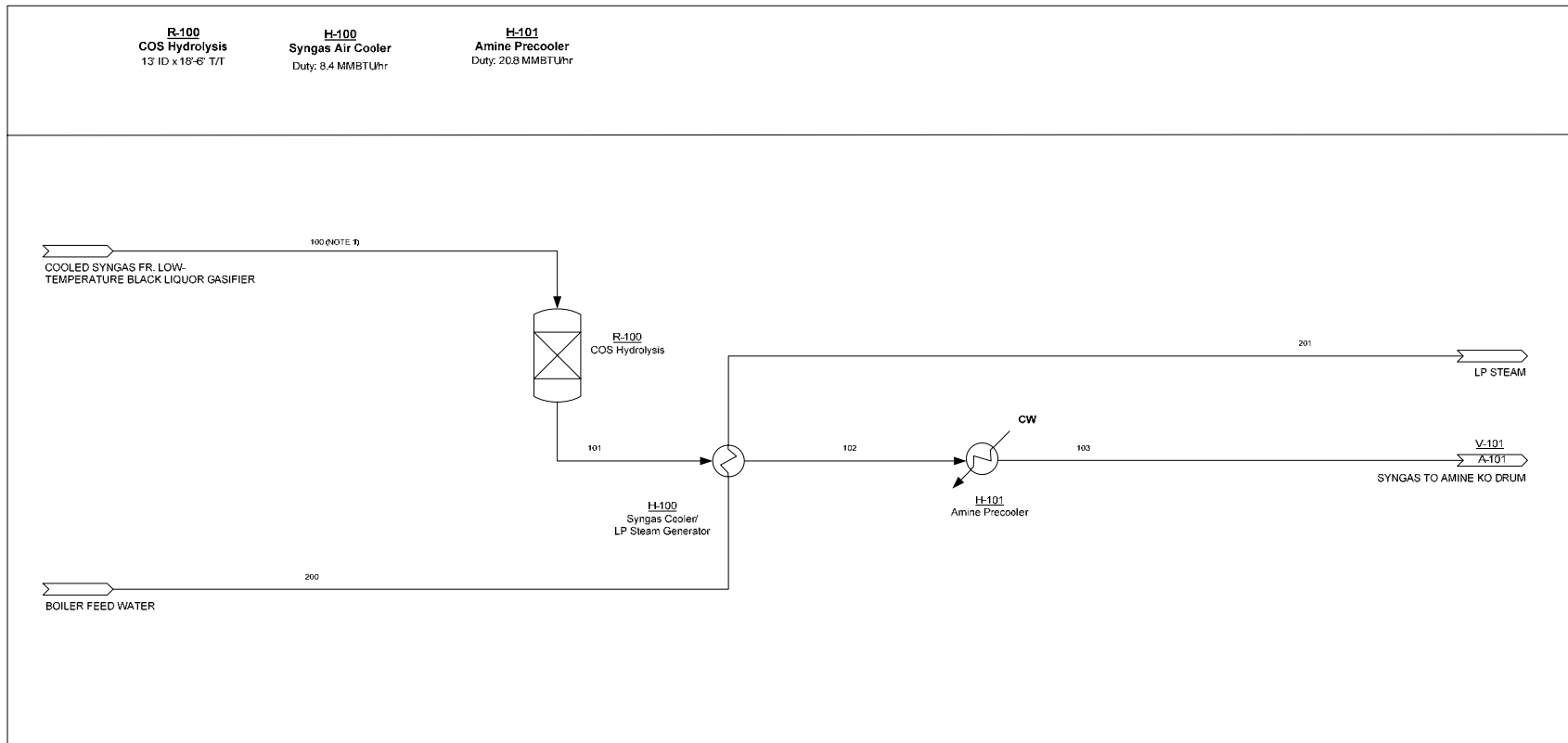
	Unit	100	101	102	103	200	201
Mass Flow	lb/hr	196,870	196,870	196,868	196,868	8,636	8,636
Temperature	F	325	325	240	110	220	230
Pressure	psia	336.5	331.5	326.5	321.5	25.0	20.0
Vapour Fraction		1.00	1.00	1.00	0.97	0.00	1.00
Molecular Weight		15.1	15.1	15.1	15.1	18.0	18.0
Heat Flow	MMBtu/hr	-480.5	-480.5	-488.9	-509.7	-57.5	-49.1
Components							
Hydrogen	lb/hr	14,617	14,617	14,617	14,617	-	-
CO	lb/hr	77,854	77,853	77,853	77,853	-	-
CO2	lb/hr	79,095	79,161	79,161	79,161	-	-
Methane (CH4)	lb/hr	6,547	6,547	6,547	6,547	-	-
H2O	lb/hr	9,027	9,003	9,003	9,003	8,636	8,636
COS	lb/hr	105	15	15	15	-	-
H2S	lb/hr	9,626	9,672	9,672	9,672	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-	-
Nitrogen	lb/hr	-	-	-	-	-	-

NOTES:
1. THE OPERATING CONDITIONS OF STREAM 100 IS
AT DOWNSTREAM OF THE HEAT RECOVERY UNIT.

NO	DATE	REVISIONS				
SCALE		DESIGNED		DRAWN		

Nexant
San Francisco

NREL BLACK LIQUOR GASIFICATION
LOW-TEMPERATURE SYNGAS
FOR POWER GENERATION

PROCESS FLOW DIAGRAM
A-100: COS HYDROLYSIS

JOB No.	DRAWING No.	REV
NREL Contract
ACO-5-44027		REV0

28 Aug, 2005

C-101 Amine Absorber

C-102 Stripper

P-101 Lean Solvent Pump 860 GPM 200 BHP

P-102 Reflux Pump 55 GPM

H-102 Rich Solvent/Lean Solvent Exchanger Duty: 30.2 MMBTU/hr

H-103 Lean Solvent Cooler Duty: 22.7 MMBTU/hr

H-104 Stripper Reboiler Duty: 54.9 MMBTU/hr

H-105 Acid Gas Condenser Duty: 29.7 MMBTU/hr

V-101 Amine KO Drum

V-102 Sweet Gas KO Drum

V-103 Rich Solvent Flash Drum

V-104 Reflux Accumulator

	Unit	103	104	105	106	107	108	110	111	112	113	114	115	116	117	118	119	120	121	122	123
Mass Flow	lb/hr	196,868	188,887	7,981	154,832	154,832	33,820	433,309	467,276	1,824	465,452	465,452	492,512	60,881	61,275	27,456	431,632	431,632	431,632	58,888	58,888
Temperature	F	110	110	110	118	118	130	110	160	157	157	200	253	248	220	130	248	171	172	281	281
Pressure	psia	321.5	321.5	321.5	323.3	323.3	25.0	326.5	326.5	30.0	30.0	30.0	28.0	25.0	25.0	25.0	25.0	25.0	326.5	50.0	50.0
Vapour Fraction		0.97	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	0.00
Molecular Weight		15.1	15.0	18.0	13.2	13.2	38.6	31.2	31.7	37.1	31.7	31.7	28.7	18.1	25.6	18.1	31.3	31.3	31.3	18.0	
Heat Flow	MMBtu/hr	-509.7		-54.0																	
Components																					
Hydrogen	lb/hr	14,617	14,617	0	14,612	14,612	0	-	5	5	0	0	-	-	0	0	-	-	-		
CO	lb/hr	77,853	77,853	0	77,834	77,834	0	-	22	22	0	0	-	-	0	0	-	-	-		
CO2	lb/hr	79,161	79,143	19	54,648	54,648	23,045	196	24,699	1,457	23,242	23,242	674	477	23,067	22	196	196	196		
Methane (CH4)	lb/hr	6,547	6,546	0	6,543	6,543	0	-	2	2	0	0	-	-	0	0	-	-	-		
H2O	lb/hr	9,003	1,049	7,954	1,155	1,155	1,430	217,783	217,645	108	217,537	217,537	276,369	60,263	28,797	27,368	216,106	216,106	216,106	58,888	58,888
COS	lb/hr	15	15	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		
H2S	lb/hr	9,672	9,665	8	40	40	9,344	7	9,582	231	9,351	9,351	10	3	9,373	29	7	7	7		
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		
Argon	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		
Nitrogen	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-		
MDEA	lb/hr	-	-	-	0	0	-	215,322	215,322	0	215,322	215,322	215,459	137	38	38	215,322	215,322	215,322		

San Francisco

NREL BLACK LIQUOR GASIFICATION LOW-TEMPERATURE SYNGAS FOR POWER GENERATION

PROCESS FLOW DIAGRAM A-101: AMINE SYSTEM

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 0

29 Dec, 2005

Task 2: Gas Cleanup Design and Cost Estimates
Black Liquor Sulfur Removal Design and Cost Estimates
United States Department of Energy/National Renewable Energy Laboratory

B-3

R-120	H-120A	H-121A	R-121A	R-121B	H-120B	H-121B
Reaction Furnace	Sulfur Condenser	Reheater	Catalytic Converter	Catalytic Converter	Sulfur Condenser	Reheater

ACID GAS FR.
AMINE UNIT

AIR

Reaction
Furnace

R-120
Reaction Furnace

H-120A
Sulfur Condenser

BFW

H-121A
Reheater

STM

H-120B
Sulfur Condenser

CW

H-121B
Reheater

STM

H-120C
Sulfur Condenser

CW

Tail
Gas

R-121A
Catalytic Converter

R-121B
Catalytic Converter

Spent
Catalyst

Sulfur

No   DATE   REVISIONS

SCALE   DESIGNED   DRAWN

Nexant

San Francisco

**NREL BLACK LIQUOR GASIFICATION
LOW-TEMPERATURE SYNGAS**

**PROCESS FLOW DIAGRAM
A-120: SULFUR RECOVERY**

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

28 Aug. 2005

Appendix C   High-Temperature Gasifier to Chemicals Design Case PFDs

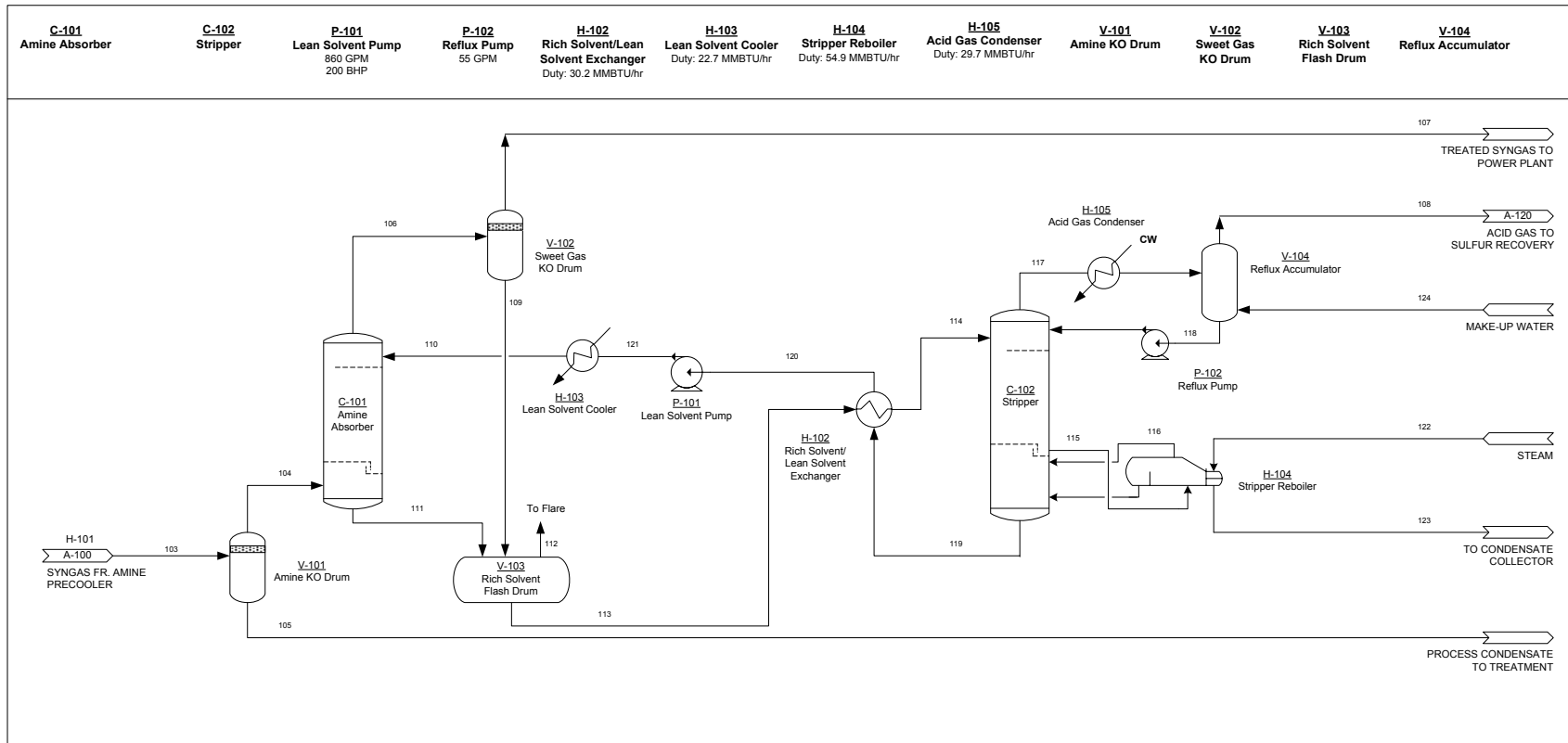The Claus Sulfur recovery unit shown here is for general information purpose only, and was not explicitly sized as part of the project scope. A cost estimate for this unit and a Tail Gas Treating process is provided in the equipment list.
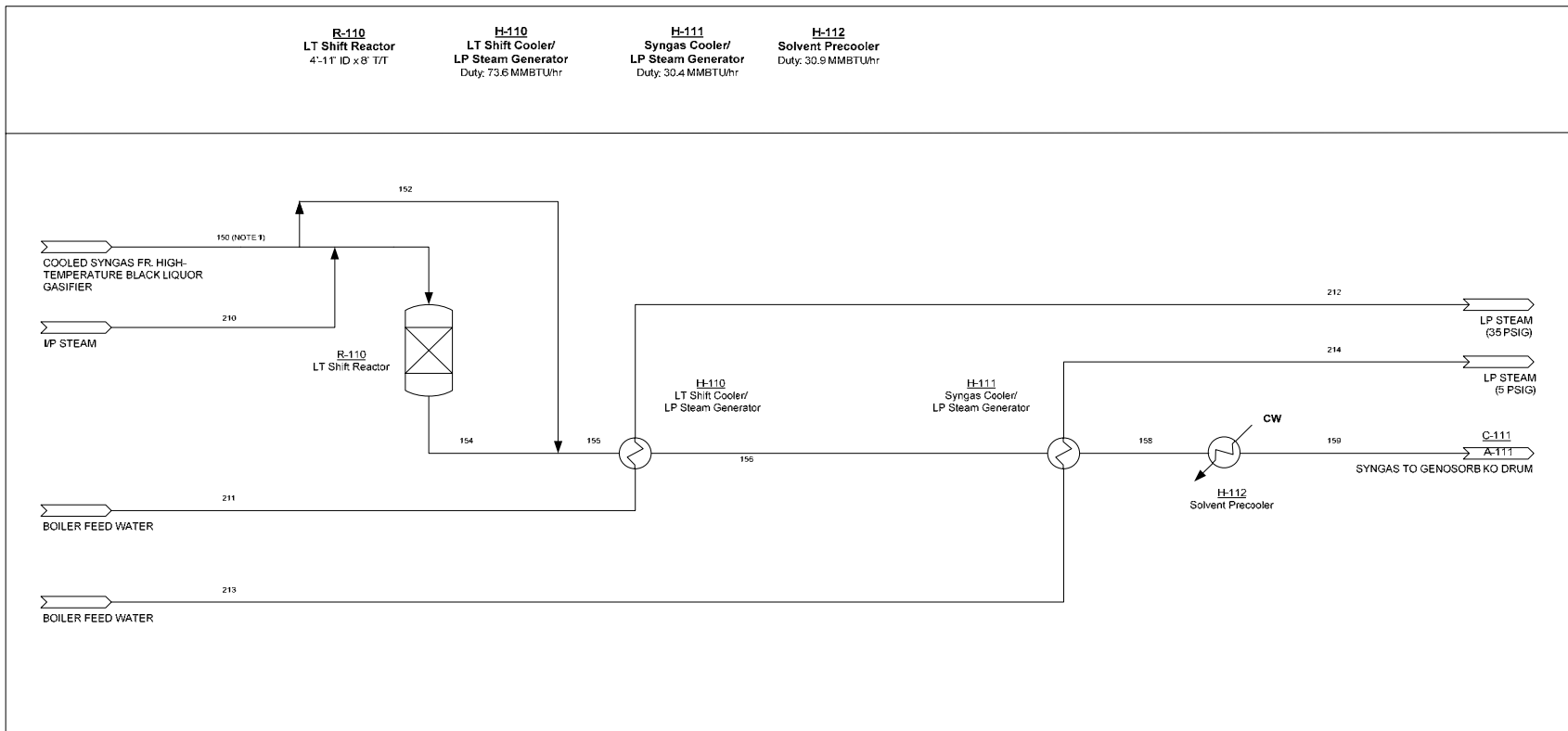
	Unit	150	152	153	154	155	156	158	159	210	211	212	213	214
Mass Flow	lb/hr	203,133	119,848	176,963	176,963	296,811	296,811	296,811	296,811	93,679	78,007	78,007	31,122	31,122
Temperature	F	252	252	460	718	551	300	240	120	608	272	282	220	230
Pressure	psia	487.3	487.3	487.3	482.3	482.3	477.3	472.3	467.3	500.0	55.0	50.0	25.0	20.0
Vapour Fraction		1.00	1.00	1.00	1.00	1.00	0.83	0.74	0.70	1.00	0.00	1.00	0.00	1.00
Molecular Weight		20.8	20.8	19.2	19.2	19.9	19.9	19.9	19.9	18.0	18.0	18.0	18.0	18.0
Heat Flow	MMBtu/hr	-518.7	-306.0	-731.3	-731.3	-1,037.3	-1,110.9	-1,141.4	-1,172.2	-518.7	-515.4	-441.8	-207.4	-176.9
Components														
Hydrogen	lb/hr	7,133	4,208	2,924	5,481	9,689	9,689	9,689	9,689	-	-	-	-	-
CO	lb/hr	94,013	55,467	38,545	3,026	58,494	58,494	58,494	58,494	-	-	-	-	-
CO2	lb/hr	77,825	45,917	31,908	87,714	133,631	133,631	133,631	133,631	-	-	-	-	-
Methane (CH4)	lb/hr	2,962	1,748	1,215	1,215	2,962	2,962	2,962	2,962	-	-	-	-	-
H2O	lb/hr	11,094	6,545	98,227	75,383	81,928	81,928	81,928	81,928	93,679	78,007	78,007	31,122	31,122
COS	lb/hr	412	243	169	169	412	412	412	412	-	-	-	-	-
H2S	lb/hr	5,416	3,196	2,221	2,221	5,416	5,416	5,416	5,416	-	-	-	-	-
Ammonia (NH3)	lb/hr	8	5	3	3	8	8	8	8	-	-	-	-	-
Argon	lb/hr	3,396	2,004	1,392	1,392	3,396	3,396	3,396	3,396	-	-	-	-	-
Nitrogen	lb/hr	874	516	358	358	874	874	874	874	-	-	-	-	-

NOTES:
1. THE OPERATING CONDITIONS OF STREAM 110 IS AT DOWNSTREAM OF THE HEAT RECOVERY UNIT.

Nexant
San Francisco

NREL BLACK LIQUOR GASIFICATION
HIGH-TEMPERATURE SYNGAS
FOR CHEMICALS PRODUCTION

PROCESS FLOW DIAGRAM
A-110: LT SHIFT & COS HYDROLYSIS

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 0

16 Dec, 2005

Task 2: Gas Cleanup Design and Cost Estimates
Black Liquor Sulfur Removal Design and Cost Estimates
United States Department of Energy/National Renewable Energy Laboratory

C-2

	C-111 H2S Absorber	C-112 CO2 Absorber	C-113 Stripper	P-111 Lean Solvent Pump	P-112 Loaded Solvent Pump	P-113 Semi-Lean Solvent Pump	P-114 Reflux Pump	H-113 Rich/Lean Solvent Exchanger	H-114 Loaded Solvent Cooler	H-115 Lean Solvent Cooler	H-116 Acid Gas Condenser	H-117 Stripper Reboiler	V-111 KO Drum	V-112 Rich Solvent Flash Drum	V-113 Flash Drum	V-114 Reflux Accumulator	V-115 Sweet Gas KO Drum

NOTES:

1. MATERIAL BALANCE PROVIDED BY UDHE BASED ON PROPRIETARY SIMULATION TOOLS. DESIGN PRESENTED IS A GENERIC TWO ABSORBER SYSTEM, AND MAY REQUIRE MODIFICATIONS BASED ON INLET COMPOSITION AND PRODUCT REQUIREMENTS.

2. MULTIPLE FLASH STAGES MAY BE NECESSARY. DEPENDING ON COMPOSITION, ADDITIONAL TREATMENT MAY BE REQUIRED

	Unit	159	160	161	164	165	166	168	178	182
Mass Flow	lb/hr	296,811	80,413	216,398	82,013	15,698	520,811	62,508	484,834	55,865
Temperature	F	120	120	120	102	113		43		33
Pressure	psia	462.3	462.3	462.3	454.0	23.2		72.5		16.0
Vapour Fraction		1.00	0.00	1.00	1.00	1.00		1.00		1.00
Molecular Weight		19.9	18.0	20.7	11.2	38.6		43.8		43.8
Components										
Hydrogen	lb/hr	9,689	-	9,689	9,610	-	-	53	-	1
CO	lb/hr	58,494	-	58,494	56,462	0	-	934	-	41
CO2	lb/hr	133,631	-	133,631	8,893	9,369	-	61,173	-	55,645
Methane (CH4)	lb/hr	2,962	-	2,962	2,664	1	-	254	-	43
H2O	lb/hr	81,928	80,413	1,515	18	447	-	27	-	84
COS	lb/hr	412	-	412	1	303	-	57	-	51
H2S	lb/hr	5,416	-	5,416	0	5,416	-	-	-	-
Solvent	lb/hr	-	-	-	-	-	520,811	-	484,834	-
Argon	lb/hr	3,396	-	3,396	3,388	-	-	7	-	-
Nitrogen	lb/hr	874	-	874	874	-	-	-	-	-

Task 2: Gas Cleanup Design and Cost Estimates
Black Liquor Sulfur Removal Design and Cost Estimates
United States Department of Energy/National Renewable Energy Laboratory

C-3

R-112/113
ZnO Sulfur
Removal Beds
7'-2' ID x 10' T/T

H-118
ZnO Bed Preheater
Duty: 35.1 MMBTU/hr

216
CONDENSATE TO
COLLECTION DRUM

V-115
A-111   164   190
FROM SWEET GAS
KO DRUM

H-118
ZnO Bed
Preheater

215

SUPERHEAT
STEAM

R-112/113
ZnO Sulfur
Removal Beds

191
TREATED SYNGAS TO
CHEMICALS PRODUCTION UNIT

	Unit	164	190	191	215	216
Mass Flow	lb/hr	82,013	82,013	82,012	29,748	29,748
Temperature	F	102	750	750	775	281
Pressure	psia	454.0	454.0	450.0	50.0	50.0
Vapor Fraction		1.00	1.00	1.00	1.00	0.00
Molecular Weight		11.2	11.2	11.2	18.0	18.0
Heat Flow	MMBtu/hr	-142.1	-107.3	-107.3	-161.4	-196.3
Components						
Hydrogen	lb/hr	9,610	9,610	9,610	-	-
CO	lb/hr	56,462	56,462	56,462	-	-
CO2	lb/hr	8,893	8,893	8,893	-	-
Methane (CH4)	lb/hr	2,664	2,664	2,664	-	-
H2O	lb/hr	18	18	18	29,748	29,748
COS	lb/hr	1	1	-	-	-
H2S	lb/hr	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-
Argon	lb/hr	3,388	3,388	3,388	-	-
Nitrogen	lb/hr	874	874	874	-	-

No	DATE	REVISIONS					
SCALE		DESIGNED		DRAWN			

Nexant
San Francisco

NREL BLACK LIQUOR GASIFICATION
HIGH-TEMPERATURE SYNGAS
FOR CHEMICALS PRODUCTION

PROCESS FLOW DIAGRAM
A-112: ZnO SULFUR REMOVAL

JOB No.	DRAWING No.	REV
NREL Contract
ACO-5-44027		REV 0

29 Dec, 2005

R-120	H-120A	H-121A	R-121A	R-121B	H-120B	H-121B
Reaction Furnace	Sulfur Condenser	Reheater	Catalytic Converter	Catalytic Converter	Sulfur Condenser	Reheater

NREL BLACK LIQUOR GASIFICATION
HIGH-TEMPERATURE SYNGAS

PROCESS FLOW DIAGRAM
A-120: SULFUR RECOVERY

NREL Contract
ACO-5-44027

REV 1

28 Aug. 2005

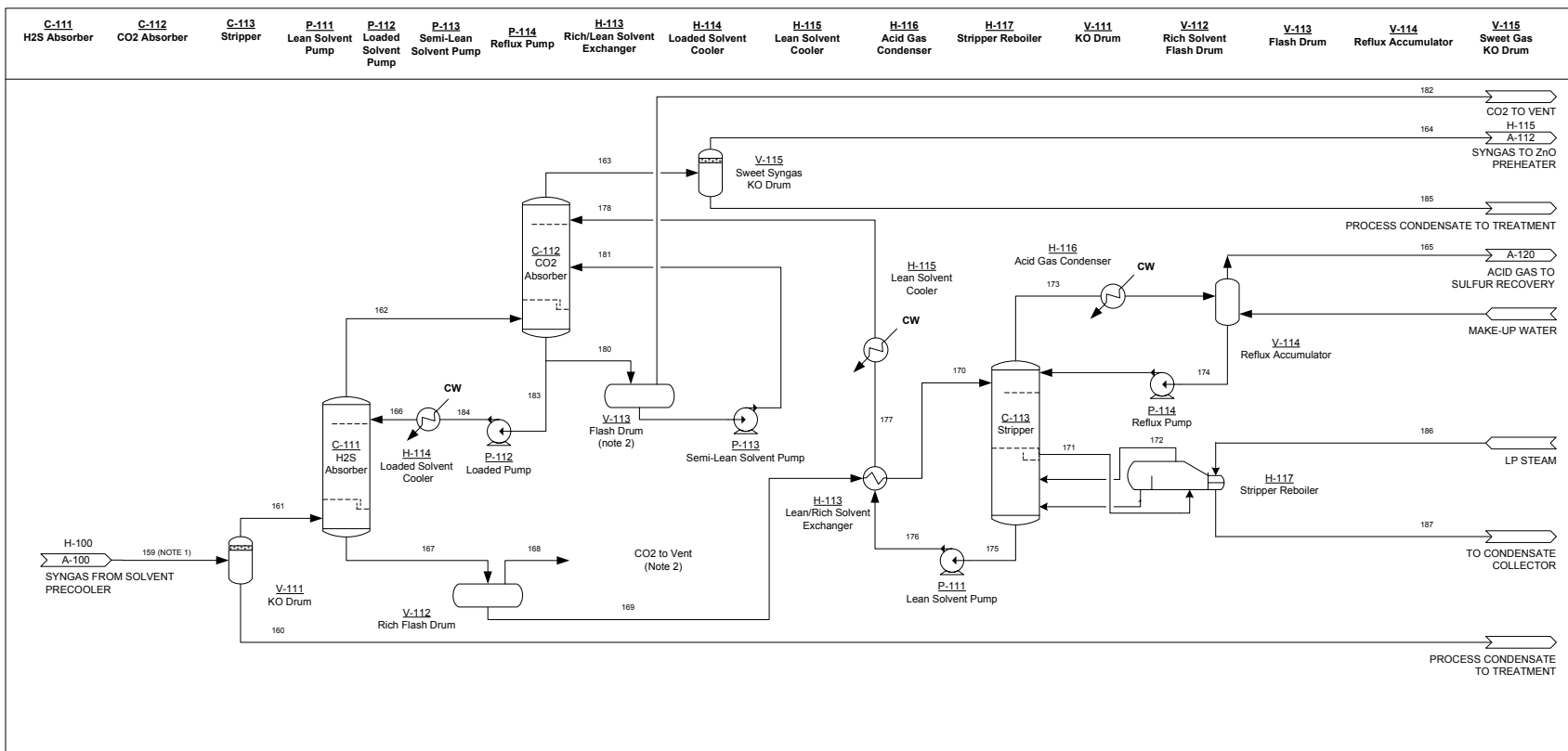# Appendix D   Low-Temperature Gasifier to Chemicals Design Case PFDs

The Claus Sulfur recovery unit shown here is for general information purpose only, and was not explicitly sized as part of the project scope. A cost estimate for this unit and a Tail Gas Treating process is provided in the equipment list.
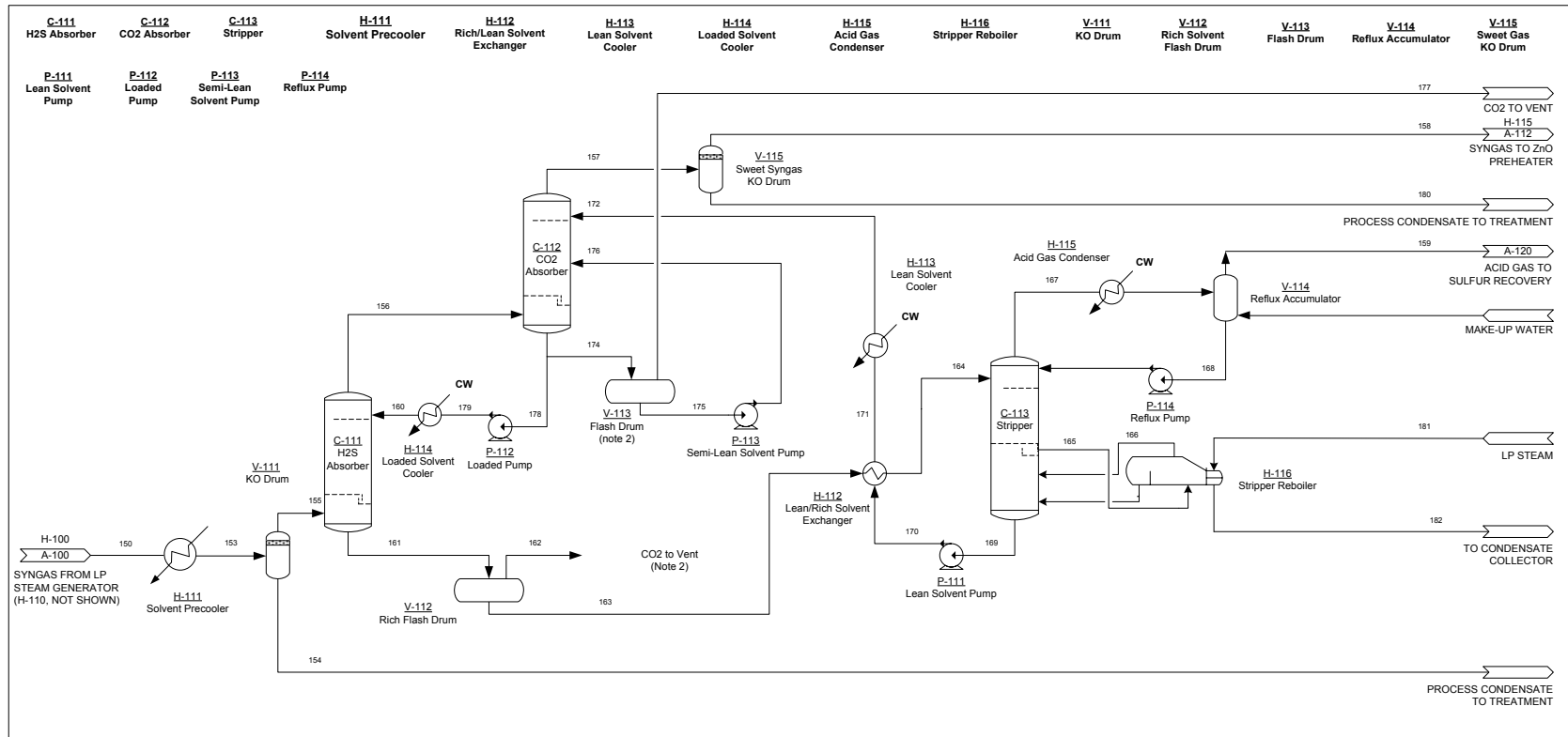
	Unit	153	154	155	158	159	160	162	172	177
Mass Flow	lb/hr	196,953	7,634	189,319	121,560	21,240	831,898	26,133	993,691	19,709
Temperature	F	120	120	120	98	113		43		33
Pressure	psia	326.5	326.5	326.5	318.5	23.2		72.5		16.0
Vapour Fraction		1.00	0.00	1.00	1.00	1.00		1.00		1.00
Molecular Weight		15.1	18.0	15.0	11.1	37.4		40.0		43.0
Components										
Hydrogen	lb/hr	14,617	-	14,617	14,539	-	-	74	-	5
CO	lb/hr	77,853	-	77,853	76,738	1	-	980	-	134
CO2	lb/hr	79,152	-	79,512	24,146	10,831	-	24,759	-	19,413
Methane (CH4)	lb/hr	6,547	-	6,547	6,103	5	-	308	-	129
H2O	lb/hr	9,003	7,634	1,369	33	623	-	9	-	28
COS	lb/hr	105	-	105	0	104	-	1	-	-
H2S	lb/hr	9,676	-	9,676	0	9,675	-	-	-	-
Solvent	lb/hr	-	-	-	-	-	831,898	-	993,691	-
Argon	lb/hr	-	-	-	-	-	-	-	-	-
Nitrogen	lb/hr	-	-	-	-	-	-	-	-	-

NOTES:

1. MATERIAL BALANCE PROVIDED BY UDHE BASED ON PROPRIETARY SIMULATION TOOLS. DESIGN PRESENTED IS A GENERIC TWO ABSORBER SYSTEM, AND MAY REQUIRE MODIFICATIONS BASED ON INLET COMPOSITION AND PRODUCT REQUIREMENTS

2. MULTIPLE FLASH STAGES MAY BE NECESSARY. DEPENDING ON COMPOSITION, ADDITIONAL TREATMENT MAY BE REQUIRED

Nexant

San Francisco

NREL BLACK LIQUOR GASIFICATION LOW-TEMPERATURE SYNGAS FOR CHEMICALS PRODUCTION

**PROCESS FLOW DIAGRAM
A-110: AMINE SYSTEM**

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 0

21 Dec, 2005

Task 2: Gas Cleanup Design and Cost Estimates
Black Liquor Sulfur Removal Design and Cost Estimates
United States Department of Energy/National Renewable Energy Laboratory

D-2

R-111/112
ZnO Sulfur
Removal Beds
7'-3" ID x 10' T/T

H-117
ZnO Bed Preheater
Duty: 52.7 MMBTU/hr

213
CONDENSATE TO
COLLECTION DRUM

V-115
A-111
158
FROM SWEET GAS
KO DRUM

190

H-117
ZnO Bed
Preheater

212
LP STEAM

R-111/112
ZnO Sulfur
Removal Beds

191

TREATED SYNGAS TO
CHEMICALS PRODUCTION UNIT

	Unit	158	190	191	212	213
Mass Flow	lb/hr	121,560	121,560	121,560	44,977	44,977
Temperature	F	98	750	750	775	281
Pressure	psia	318.5	318.0	313.0	50.0	50.0
Vapour Fraction		1.00	1.00	1.00	1.00	0.00
Molecular Weight		11.1	11.1	11.1	18.0	18.0
Components						
Hydrogen	lb/hr	14,538	14,538	14,538	-	-
CO	lb/hr	76,738	76,738	76,738	-	-
CO2	lb/hr	24,146	24,146	24,146	-	-
Methane (CH4)	lb/hr	6,103	6,103	6,103	-	-
H2O	lb/hr	33	33	33	44,977	44,977
COS	lb/hr	-	-	-	-	-
H2S	lb/hr	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-
Argon	lb/hr	-	-	-	-	-
Nitrogen	lb/hr	-	-	-	-	-

No	DATE	REVISIONS				
SCALE		DESIGNED		DRAWN		

Nexant
San Francisco

NREL BLACK LIQUOR GASIFICATION
LOW-TEMPERATURE SYNGAS
FOR CHEMICALS PRODUCTION

PROCESS FLOW DIAGRAM
A-111: ZnO SULFUR REMOVAL

JOB No.	DRAWING No.	REV
NREL Contract ACO-5-44027		REV 0

21 Dec, 2005

Task 2: Gas Cleanup Design and Cost Estimates
Black Liquor Sulfur Removal Design and Cost Estimates
United States Department of Energy/National Renewable Energy Laboratory

D-3

R-120	H-120A	H-121A	R-121A	R-121B	H-120B	H-121B
Reaction Furnace	Sulfur Condenser	Reheater	Catalytic Converter	Catalytic Converter	Sulfur Condenser	Reheater

ACID GAS FR.
AMINE UNIT

AIR

H-120A
Sulfur Condenser

H-121A
Reheater

H-120B
Sulfur Condenser

H-121B
Reheater

H-120C
Sulfur Condenser

Reaction
Furnace

R-120
Reaction Furnace

BFW   STM   CW   STM   CW

Tail
Gas

R-121A
Catalytic Converter

R-121B
Catalytic Converter

Spent
Catalyst

Sulfur

Nexant
San Francisco

**NREL BLACK LIQUOR GASIFICATION
LOW-TEMPERATURE SYNGAS**

**PROCESS FLOW DIAGRAM
A-120: SULFUR RECOVERY**

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

28 Aug. 2005

The equipment lists for all four black liquor gasification cases are shown below, with the two high temperature and two low temperature cases placed together on the same sheet.

High Temperature Syngas to Power and Chemicals Cases

Item No	Description	Type	Quantity Per Train	Size	Weight	Head	Design Duty	Design P	Design T	Operating P	Operating T	Power Useage	Materials	Price, total (unistalled) Q2 2004 Cost Index (US $)	Price Escalated, total (uninstalled) Q2 2005 Cost Index (US $)	Total Installed Cost (US $)	Quote Source	Comments
					lbs	PSI		PSIG	°F	PSIG	°F	(No.) HP						
Power Generation Case																		
Reactors																		
R-100	COS Hydrolysis	Vertical	1	14' ID x 15' - 9" T/T	132,230			520	375	472	325		CS	$315,598	$374,141		Sud-Chemie	2400 ft³ catalyst required
Total															$374,141			
Heat Exchangers																		
H-100	Syngas Cooler/ LP Steam Generator	Shell & Tube	1	6' ID x 12' T-T Surface area: 7907 SQFT			6.8 MMBTU/hr	T 25 / S 515	280 / 375	10 / 467	230 / 326		CS / CS	$165,900	$188,429		ICARUS	
H-101	Amine Pre-cooler	Shell & Tube	1	4' - 1" ID x 8' T-T Surface area: 3120 SQFT			20.1 MMBTU/hr	T 65 / S 510	150 / 290	50 / 462	100 / 240		CS / CS	$87,300	$99,155		ICARUS	
Total															$287,585			
Packaged Units																		
A-101	Amine Unit															$8,500,000	GRI Cost Curve	
A-120	Claus Plant with Tail Gas Treating															$17,622,000	Scaled from Ortloff quote	63.3 STPD sulfur production
TOTAL EQUIPMENT COST, (excld. Package units)														$661,726	$1,700,636			
TOTAL INSTALLED COST																$27,822,636		
Chemicals Production Case																		
Reactors																		
R-110	LT Shift Reactor	Vertical	1	4' - 11" ID x 8' T/T	46,100			515	770	467	718		CS		$230,500		Sud-Chemie	151 ft³ catalyst required
R-112	ZnO Sulfur Removal Beds	Vertical	1	7' - 2" ID x 10' T/T	66,800			480	800	437	750		CS		$333,400		Johnson Matthey	818 ft³ total catalyst volume
R-113	ZnO Sulfur Removal Beds	Vertical	1	7' - 2" ID x 10' T/T	66,800			480	800	437	750		CS		$333,400		Johnson Matthey	required
Total															$897,300			
Heat Exchangers																		
H-110	LT Shift Cooler/ LP Steam Generator	Shell & Tube	1	7' - 8" ID x 20' T-T Surface area: 29185 SQFT			73.59 MMBTU/hr	T 55 / S 515	350 / 600	40 / 467	282 / 551		CS / CS	$899,900	$1,022,106		ICARUS	
H-111	Syngas Cooler/ LP Steam Generator	Shell & Tube	2	7' - 3" ID x 20' T-T Surface area: 20268 SQFT			30.4 MMBTU/hr	T 25 / S 510	280 / 350	10 / 462	230 / 300		CS / CS	$578,400	$656,947		ICARUS	
H-112	Solvent Pre-cooler	Shell & Tube	1	4' - 9" ID x 8' T-T Surface area: 4270 SQFT			30.9 MMBTU/hr	T 65 / S 500	150 / 290	50 / 452	100 / 240		CS / CS	$113,100	$128,459		ICARUS	
H-118	ZnO Bed Preheater	Shell & Tube	1	4' - 6" ID x 20' T-T Surface area: 10672 SQFT			35.1 MMBTU/hr	T 490 / S 50	800 / 825	442 / 35	750 / 775		CS / CS	$147,800	$167,871		ICARUS	
Total															$1,975,383			
Packaged Units																		
A-111	Genosorb Unit															$30,125,000	Scaled estimates from previous physical solvent studies and IGCC data	Physical solvent necessary for CO2 removal and to concentrate the H2S for Claus Feed
A-120	Claus Plant with Tail Gas Treating															$17,677,000	Scaled from Ortloff quote	63.6 STPD sulfur production
TOTAL EQUIPMENT COST, (excld. Package units)														$2,872,683	$7,382,796			
TOTAL INSTALLED COST																$55,184,796		

Low Temperature Syngas to Chemicals Design Case

Item No	Description	Type	Quantity Per Train	Size	Weight	Head	Design Duty	Design P	Design T	Operating P	Operating T	Power Useage	Materials	Price, total (unistalled)	Price Escalated, total (uninstalled)	Total Installed Cost	Quote Source	Comments
					lbs	PSI		PSIG	°F	PSIG	°F	(No.) HP		Q2 2004 Cost Index (US $)	Q2 2005 Cost Index (US $)	(US $)		
Power Generation Case																		
Reactors																		
R-100	COS Hydrolysis	Vertical	1	13' ID x 18' - 6" T/T	130,200			355	375	322	325		CS	$310,767	$368,414		Sud-Chemie	2470 ft³ catalyst required
Total															$368,414			
Heat Exchangers																		
H-100	Syngas Cooler/ LP Steam Generator	Shell & Tube	1	4' - 7" ID x 12' T-T Surface area: 4433 SQFT			8.4 MMBTU/hr	T 25 S 350	280 375	10 316	230 325		CS CS	$93,000	$105,629		ICARUS	
H-101	Amine Pre-cooler	Shell & Tube	1	4' ID x 8' T-T			20.8 MMBTU/hr	T 65 S 345	150 290	50 312	100 240		CS CS	$73,600	$83,595		ICARUS	
Total															$189,224			
Packaged Units																		
A-101	Amine Unit															$10,200,000	GRI Cost Curve	
A-120	Claus Plant with Tail Gas Treating															$25,036,000	Scaled from Ortloff quote	109 STPD sulfur production
TOTAL EQUIPMENT COST, (excld. Package units)														$557,639	$1,433,131			
TOTAL INSTALLED COST																$36,669,131		
Chemicals Production Case																		
Reactors																		
R-111	ZnO Sulfur Removal Beds	Vertical	1	7' - 3" ID x 10' T/T	66,900			330	800	297	750		CS		$334,600			
R-112	ZnO Sulfur Removal Beds	Vertical	1	7' - 3" ID x 10' T/T	66,900			330	800	297	750		CS		$334,600			
Total															$669,200			
Heat Exchangers																		
H-110	Syngas Cooler/ LP Steam Generator	Shell & Tube	1	4' - 7" ID x 12' T-T Surface area: 4433 SQFT			8.4 MMBTU/hr	T 25 S 350	280 375	10 316	230 325		CS CS	$93,000	$105,629		ICARUS	
H-111	Solvent Pre-cooler	Shell & Tube	1	4' - 6" ID x 8' T-T Surface area: 3285 SQFT			15.6 MMBTU/hr	T 65 S 355	150 250	50 322	100 200		CS CS	$91,900	$104,380		ICARUS	
H-117	ZnO Bed Preheater	Shell & Tube	1	6' - 2" ID x 18' T-T Surface area: 17750 SQFT			52.7 MMBTU/hr	T 335 S 50	800 825	302 35	750 775		CS CS	$245,800	$279,180		ICARUS	
Total															$383,560			
Packaged Units																		
A-111	Genosorb Unit															$22,400,000	Scaled estimates from previous physical solvent studies and IGCC data	Physical solvent necessary for CO2 removal and to concentrate the H2S for Claus Feed
A-120	Claus Plant with Tail Gas Treating															$25,093,000	Scaled from Ortloff quote	109 STPD sulfur production
TOTAL EQUIPMENT COST, (excld. Package units)														$1,052,760	$2,705,592			
TOTAL INSTALLED COST																$50,198,592		

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		5b. GRANT NUMBER
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13. SUPPLEMENTARY NOTES NREL Technical Monitor: Kelly Ibsen		
14. ABSTRACT *(Maximum 200 Words)* As part of Task 2, Gas Cleanup and Cost Estimates, Nexant investigated the appropriate process scheme for removal of acid gases from black liquor-derived syngas for use in both power and liquid fuels synthesis. Two 3,200 metric tonne per day gasification schemes, both low-temperature/low-pressure (1100°F, 40 psi) and high-temperature/high-pressure (1800°F, 500 psi) were used for syngas production. Initial syngas conditions from each of the gasifiers was provided to the team by the National Renewable Energy Laboratory and Princeton University. Nexant was the prime contractor and principal investigator during this task; technical assistance was provided by both GTI and Emery Energy.		
15. SUBJECT TERMS biomass; Nexant; small modular; synthesis gas; syngas; fluidized catalytic cracking; steam methane reforming; natural gas liquids expanders; scaling; black liquor; National Renewable Energy Laboratory; NREL		

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