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Alliance Research Center

High-Solids Black Liquor Firing in Pulp and Paper Industry Kraft Recovery Boilers

Phase 1a: Final Report
Low-Temperature Black Liquor Gasifier Evaluation

PREPARED FOR:

The United States Department of Energy
UNDER CONTRACT NO. DE-FC36-94G010002, AMENDMENT A002

SUBMITTED BY:

The Babcock & Wilcox Company
RESEARCH & DEVELOPMENT DIVISION
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June 1997

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June 1997

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1.0

EXECUTIVE SUMMARY & OVERVIEW

1.1 Program Objectives and Background

This project, conducted under The United States Department of Energy (DOE) Cooperative Agreement DE-FC36-94GO10002/A002, was part of a multiple-phase effort to develop technologies that improve the energy efficiency and economics of chemical process recovery in the pulp and paper industry. The approach taken was to consider two major alternatives in two phases. Phase I, conducted previously, considered means to improve pulp mill recovery boilers using high-solids advanced combustion of black liquor; while this project, Phase Ia, considered means to recover kraft pulping mill process chemicals by low-temperature black liquor gasification. The principal steps previously proposed in this program were:

- o Evaluate these two technologies, high-solids advanced combustion and gasification, and then select a path forward using the more promising of these two options for future work.
- o Design and construct a pilot-scale unit based on the selected technology, and using that unit, develop the precompetitive data necessary to make commercialization attractive.
- o Develop and deploy a first-of-a-kind (FOAK) commercial unit in a kraft pulp mill.

Phase I, which evaluated the high-solids advanced combustion option, was concluded in 1995. Results of that project phase were reported previously [1]. This report describes the work

conducted in Phase Ia. The work is described in Sections 1 through 4 and six appendices provide additional detail. The six appendices address areas of:

- Appendix A Bench-scale gasifier design.
- Appendix B Bench-scale gasifier testing.
- Appendix C Characteristics of black liquor used for testing.
- Appendix D Pilot-scale gasifier (PSG) testing and modeling.
- Appendix E Pilot-scale gasifier design.
- Appendix F Market and Economic evaluation.

1.2 Recommended Technology

At this time gasification is recommended as the technology to pursue for the longer term. Ongoing commercial activities are advancing high-solids firing of black liquor in process recovery boilers. Other DoE-sponsored projects that support advances in numerical modeling and basic understanding of black liquor combustion would aid this process. This combination of these efforts supports the work needed in the near term to improve black liquor combustion.

Gasification offers the potential advantages of improved energy use, increased electrical power generation via combined cycles, and reduced emissions over direct combustion. However, large-scale units offering these advantages will not be available in the near term. Additional development of the low-temperature gasification processes described in this report is needed, to demonstrate its performance, assure that the process is economically viable, and then complete development to support commercial applications.

1.3 Phase Ia Objectives

Phase Ia evaluated black liquor gasification as an alternative to combustion. The objectives to support this goal were:

- o Conduct bench-scale tests of a low-temperature, partial-combustion gasification process. This was the majority of the Phase Ia activity, since it involved the design, construction, installation and testing of a bench-scale gasifier.
- o Prepare a preliminary design and cost estimate of the pilot-scale gasifier that might be developed in the future.
- o Outline a test program to evaluate gasification at the pilot scale.
- o Prepare an economic and market analysis of gasification, and solicit pulp and paper industry support for subsequent phases.

1.4 Phase Ia Key Results

The primary technical results of this phase were derived from the design, fabrication, installation and startup of the bench-scale test unit and its subsequent testing. The work showed that the bench-scale unit can be used to obtain useful qualitative and quantitative results. Phase Ia test results along with the recommended future bench-scale work will demonstrate the technical feasibility of low-temperature, partial-combustion black liquor gasification. At that point, a decision may be made of whether to continue with additional development at the pilot scale.

Results of the Phase Ia tests show that low-temperature black liquor gasification to form a low-heating-value gas is possible. However, further characterization of tar formation and char conversion is needed. Two key results regarding tar formation were: (1) the quantity of tar formed compared with the quantity of liquor processed was small and (2) the tars only condense from the gas stream in the lower-temperature gas processing equipment downstream of the gasifier particulate collector (cyclone). The test results were limited by equipment problems; consequently, determination of the optimal range of parameters (e.g., temperature, fuel/air ratio, fuel/steam ratio) for gasification in this system remain to be adequately investigated.

These important results have led us to an improved approach for this commercialization of the B&W gasification process. The recommended future work at the bench-scale includes novel gasifier design modifications that will cost effectively address char conversion and tar utilization.

As part of the work in Phase Ia, a plan was also developed for pilot-scale testing of our current gasifier design. The plan includes numerical modeling of the gasifier to compliment the experimental work.

The pilot-scale gasifier design was advanced to the next level of detail. Note that this work was also based on our current gasifier design. This task included preliminary sizing of the major components and development of sufficient design detail to obtain supplier cost estimates to fabricate these components. A preliminary evaluation of materials of construction was completed for all major components and piping. This work also simplified the design and identified other potential improvements.

The bench-scale work to date suggests that the gasifier design will need to be revised to achieve the desired char and tar conversion. This would affect the pilot-scale gasifier design and test plan. However, the design presented in this report and the overall test program approach provide a strong starting point for that future effort.

The market potential and economics of using the B&W low-temperature black liquor gasifier for incremental recovery capacity were compared with conventional technology. The market evaluation revealed good potential for incremental capacity, low-temperature gasifiers that can provide the path to commercialization of larger units in combined cycle applications in the longer term. During the near term, small recovery boilers and recovery boiler upgrades are expected to remain competitive with the currently-available air-blown, high-temperature gasifiers.

The economic evaluation compared three options for adding liquor processing capacity of approximately 500,000 lbs. of dry solids per day:

- (1) A small recovery boiler producing high-pressure steam.
- (2) A near-atmospheric pressure gasifier producing a low-heating-value gas.
- (3) A pressurized gasifier coupled with a gas turbine producing electricity and low-pressure process steam.

This analysis considered capital cost, one-time-fixed cost related to loss of production during installation, annual-fixed operating and maintenance cost, pilot-fuel cost, and the value of the energy outputs produced (high-pressure steam, low-heating-value gas, electricity, and low-pressure process steam). Sensitivity of the results to fuel prices, capital cost, operating and maintenance cost, and gasifier performance was also investigated.

The economic analysis demonstrated that the low-temperature gasifier is competitive under the conditions considered and has merit for incremental capacity increases. The gasifier would be even more effective in the combined-cycle application.

1.5 Recommended Direction for Future Work

Future work should include additional bench-scale tests, and then consider development at the pilot scale if the technical and economic viability of the process is justified. The recommended additional bench-scale tests are intended to evaluate gasifier design improvements to eliminate the concerns related to tar formation and char conversion in a low-temperature design.

Company-funded studies to evaluate char and tar conversion improvement produced promising results. Additional work has been proposed as a cooperative agreement under DOE's Agenda 2020 program. We have identified specific modifications to the bench-scale test equipment. These modifications will permit testing to address these areas of interest, make the unit easier to operate, and produce the additional test data to resolve char and tar conversion performance issues.

Following the additional bench-scale work, an assessment is recommended to evaluate the performance of the gasifier with these process changes. Assuming the results are positive, the impact of the design changes and the performance differences on the commercial design for incremental liquor processing capacity would be evaluated. Since this would affect the cost of the equipment and the quantity and value of its product streams, the economic evaluation would need to be revised.

After these additional evaluations are completed, a decision of whether or not to continue with development of this low-temperature gasification process would be made.

Assuming the decision is made to continue with development at the pilot-scale, the bench-scale and evaluation work described would set the stage for that development. The pilot-scale gasifier development and test program were outlined in this project. Adjustments would be needed to address issues from the current work and those that will undoubtedly arise from the additional bench-scale work. However, the same major stages of design development are envisioned. These major stages would be applied to the design as it would evolve based on the bench-scale test data developed in the recommended work. The pilot-scale design would be tested in several steps to evaluate the performance of the advanced gasifier design that would evolve out of Phase Ia and the recommended future work.

The pilot-scale work would include development and validation of numerical modeling tools to simulate the key gasifier processes. The results of the pilot-scale work and the efforts leading to it would set the stage for commercialization of the technology. That process, which would begin with the FOAK commercial unit, would occur outside the scope of this program.

2.0

BACKGROUND

2.1 Gasification and the Recovery Process

Black liquor is produced by chemical pulping of wood or other fiber sources via the kraft processes (Note that wood is the dominating fiber source worldwide and is nearly the singular source in U. S. kraft mills). The resulting black liquor is roughly one-fourth to one-third water, with the balance being organic and inorganic solids. These solids are approximately equally divided between complex organic compounds (lignin and other materials extracted from the fiber source), and inorganic chemicals (primarily sodium salts) used in the pulping process. Today, black liquor is primarily burned in chemical recovery boilers to produce steam and to convert the inorganic chemicals used in the pulping process to a form suitable for recovery and reuse. Recovery boilers are the most expensive capital equipment items in a typical pulp mill. This large investment, coupled with growing needs to be more energy self-sufficient, sustains a strong industry interest in more efficiently recovering energy from this process. To this end, the pulp and paper industry is interested in increasing the chemical recovery process efficiency either by improving recovery boiler performance or implementing alternative technologies.

Black liquor gasification is one alternative approach to conventional technology for chemical recovery in the kraft pulping process. Industry interest in gasification technology has increased due to its potential to generate a product gas that may be burned in a gas turbine and thus produce electricity more efficiently. Most current gasifier processes involve liberation of significant amounts of sulfur in varying forms from the liquor being processed to the product gas. Gas

cleanup processes recover sulfur from the gas by scrubbing with a caustic process stream produced from the sodium recovered as solids or molten material from the gasifier. This cleans the gas for subsequent use, while recovering the sulfur. However, this process generates some sodium bicarbonate, which requires more recaustizing lime than typically used in subsequent chemical processing associated with conventional chemical recovery technology [2].

T. M. Grace and W. M. Timmer reviewed several gasification processes under development [3]. Of those processes reviewed, all but one involve partial combustion. That one exception uses a steam reforming process. Other processes are proposed or being developed, such as the plasma-based process being investigated by Idaho National Engineering Laboratory (INEL) [4]. Black liquor gasification processes may be conveniently grouped as high-temperature and low-temperature processes. This grouping is based on the operating temperature of the gasifier relative to the melting point of the inorganic fractions in the liquor. High-temperature processes operate above the melting point and produce a molten product, while low-temperature processes operate below the melting point and produce a solid inorganic product.

The current gasifier designs target applications that will provide incremental chemical recovery capacity increases, with the product gas being combusted in place of purchased fuel, most probably in a power boiler. In the longer term, pressurized gasifiers could be developed and the gas supplied to a gas turbine. This development direction is anticipated, and at least one supplier is continuing with an active and published research program in that direction [2]. However, it should be noted that laboratory data suggest that increasing pressure in black liquor gasification processes may result in a reduction in the gasification rate of black liquor char [5].

Four major trends will influence the development and selection of recovery processes in the next two decades:

- (1) More stringent regulation of emissions,
- (2) Changes in mill energy mix, that is electricity generated versus steam raised,
- (3) Aging recovery equipment, and

(4) Need for incremental recovery capacity.

These trends and how gasification technology responds to each are described in Table 2.1.

Table 2.1

Black Liquor Gasification Responses to Major Trends in Pulp Mill Recovery Needs.

Trends	Black Liquor Gasification Response
<p>Emissions - The cluster rules that will regulate total mill discharge, will increase the load on the recovery plant. Also, more stringent control of air borne emissions including particulate, total reduced sulfur, and oxides of nitrogen can be expected over time [6,7].</p>	<p>Gasification requires low-temperature gas cleanup to complete the chemical recovery. As a result the emissions levels will be low. Typically, these will be similar to conventional technology with some emissions lower and others higher.</p>
<p>Energy Mix - Electrical loads compared to steam loads will increase for both pulp and integrated mills. The increased electrical loads will arise from: the installation of additional environmental control equipment, the increased use of mechanical pulps, and the integration of recycled fiber production with virgin pulp production [8,9]. Note that significant amounts of capital investment in recent years has been for the areas of environmental improvements, and installation of recycled fiber processing [7].</p>	<p>Gasification has an advantage of offering more electricity, i.e., a more favorable energy mix, if the gas is consumed in a combined cycle. Gasification is expected to start with smaller capacity units that provide incremental recovery capacity and produce fuel gas. The fuel gas would be consumed in a power boiler or other combustion application in the mill, so no advantage of energy mix would be gained. However, this may provide an economic benefit in mill-specific instances.</p>
<p>Age of Equipment - There is a significant base of installed pulp production equipment in North America. Between 2000 and 2020 a significant segment of this capacity will require replacement or upgrade. [10, 11]</p>	<p>After its initial deployment, gasification may be favored for smaller capacity replacements, where the economies of scale associated with a larger recovery boiler are not achievable. Gasification could develop an advantage in the longer term (10 - 15 years) if larger capacity units or modular designs are used in combined cycles.</p>
<p>Incremental Recovery Capacity - A shift in bleach processes to displace elemental chlorine will increase the quantity of solids in the recovery stream due to changes in the pulping operation needed for these bleach plant changes [8, 10]. The recovery system is typically the most expensive single capital investment in pulp mills. Due to economies of scale, recovery installations are typically in large increments. However, site specific needs for smaller increments of recovery capacity increases could be an advantage within North America.</p>	<p>Gasification may offer a cost advantage over new recovery boiler capacity for smaller capacity increments if the existing boiler cannot be upgraded to provide a sufficient capacity increase. However, gasification technology will need to be proven to have the same level of performance and reliability to achieve widespread industry acceptance.</p>

Gasification provides promising responses to each trends as shown in the table. However, the choice of gasification over conventional technology is not absolute due to site-specific advantages that conventional technology may offer. Also, while conventional technology continues to undergo incremental improvements and changes, gasification is only now being commercialized. Gasification offers the potential long term advantage of incorporating combined-cycle technology to increase power generation efficiency and reduce cost, but this promise requires development beyond the current gasifier designs. Conventional technology offers a proven record of performance, and a large installed base that will undoubtable remain viable for some time.

2.2 The B&W Gasifier

The B&W Gasifier design, as conceived before the Phase Ia testing, is shown schematically in Figure 2-1. Major components are the bubbling fluid bed gasifier, fluidizing gas preheater, bed drain dissolving tank, particulate separator, heat recovery boiler and gas cooling and cleanup system. The system and these components are further described in this section.

Black liquor is sprayed into the gasifier unit above the bubbling fluidized bed. This permits the use of a range of black liquor solids from a low of 65% to 80%+ solids. Note that 65% solids may be more typical in incremental retrofit applications, while 80% or more that may be present in mills that have a high-solids concentrator in the evaporator system. Increasing the solids level will improve both gasifier and recovery boiler performance. Thus, a trend toward higher solids in new mills or as part of mill upgrades is occurring today and is likely to continue.

The bubbling fluidized bed gasifier operates in the low-temperature range (below the melting point of the inorganics in the liquor) to produce a solid product by gasifying the char carbon and leaving a sodium carbonate residue.

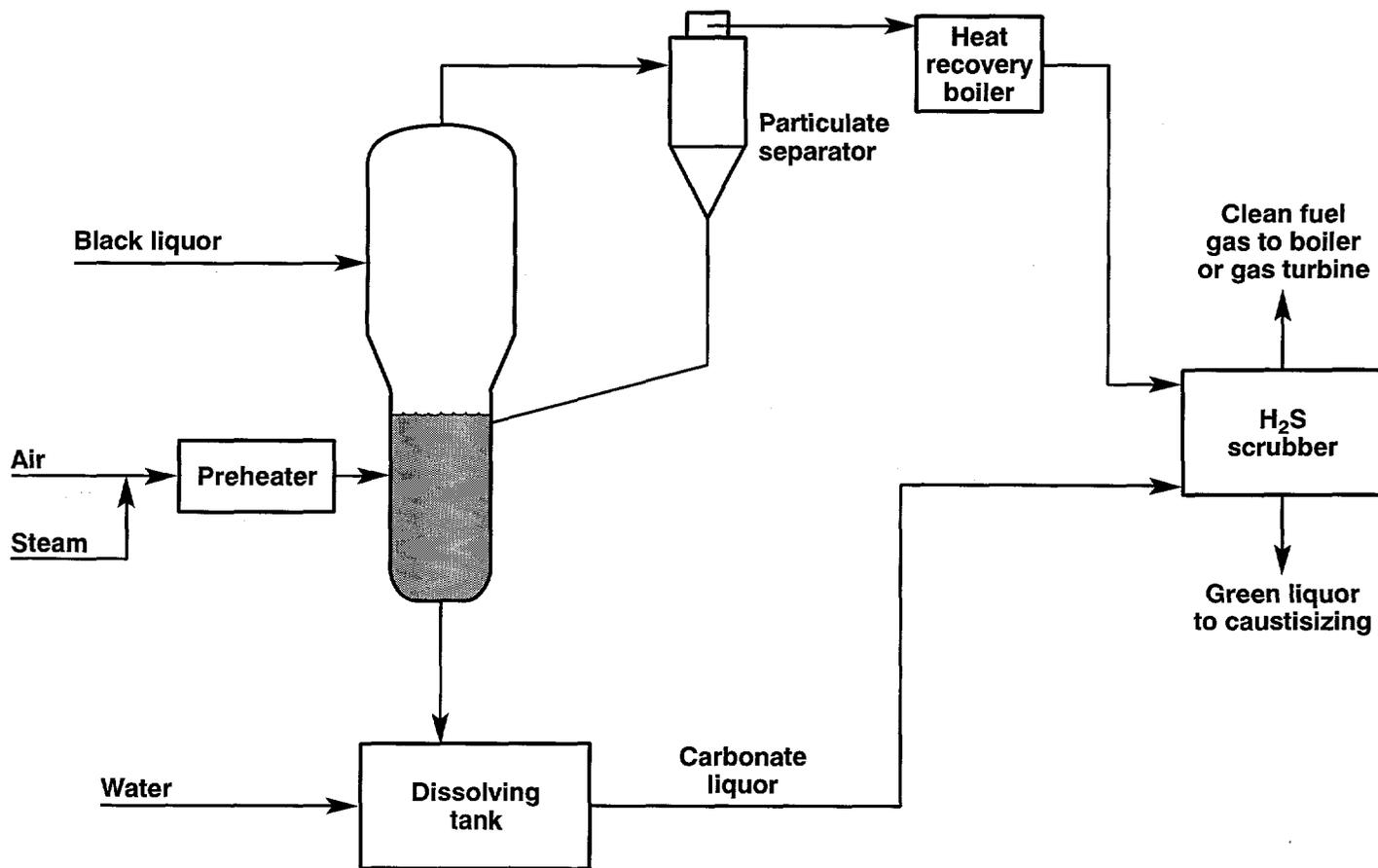


Figure 2-1. B&W Gasifier Design Concept.

The bed is supplied with a preheated mixture of air and steam, and operates as a partial oxidizing reactor. A bubbling fluidized bed also provides long solids residence time and relatively long gas phase residence time to assure that heterogenous reactions are as complete as feasible. Partial oxidation provides the heat needed for gasification. The solids produced by the gasifier are dissolved in water in the dissolving tank to produce a carbonate liquor that is in-turn used to scrub the product fuel gas.

At the bed temperatures used, the sulfur is liberated as the liquor is reacting and forms primarily hydrogen sulfide, which is captured in the scrubber. Green liquor is the scrubber product, which may contain some sodium bicarbonate, that is suitable for recaustization. The product gas is cooled before scrubbing to generate steam. Steam may be delivered at mill steam header conditions, or may be consumed in the gasifier. Low-temperature gasification provides an inherent thermal advantage over high-temperature gasification in that less energy is expended to heat the reactants to gasification conditions. Also, less energy is lost in cooling the gas to the conditions needed for low-temperature gas cleanup. The analysis presented by Grace and Timmer shows this with the comparisons of the energy conversion efficiency of low-temperature and high-temperature processes [3].

3.0

PHASE Ia RESULTS

3.1 Bench-Scale Gasifier Testing

Bench-Scale Test Objectives - A bench-scale gasifier was constructed and tested to evaluate some the practical issues of gasification. These issued included:

- o Achieving production of a clean fuel gas at the low temperature conditions.
- o Achieving adequate carbon conversion and limiting production of tars that may foul downstream equipment.
- o Examining gas composition to support future scrubber design.
- o Performing with adequate sulfate reduction.
- o Operating acceptably with higher chlorine and potassium liquors without bed fusion.

The test program was limited by equipment and operating difficulties, but was able partially to address the first three of these objectives. The results related to tar formation, and the location and temperature range for the condensation and collection of tars were very beneficial.

Bench-Scale Gasifier Test Apparatus - The bench-scale test apparatus partially simulated the B&W gasifier design features and permitted evaluation of practical performance issues. The bench-scale equipment includes simulation of the fluidized bed, the particulate removal equipment, waste heat boiler, and the gas cooling equipment. Spraying of the liquor above the bed for freeboard drying was not addressed by the bench-scale work. Gas cleanup evaluation was limited to determining gas composition to support future scrubber design.

A simplified schematic of the bench-scale black liquor gasifier is shown in Figure 3-1. The key test item is the fluidized bed reactor (on the left of the figure) in which gasification occurs. The balance of the equipment performed the following functions:

- o Supplying the reactants (black liquor solids or black liquor, and heated air, nitrogen, and steam) to the gasifier.
- o Removing gas, solids, or heat from the gasifier.
- o Cooling the gas downstream of the gasifier, and removes condensing tar and water.

Within the gasifier, heated air, nitrogen, and steam fluidize a bed of solids and react with the black liquor supplied. The bed solids are initially an inert material (silica sand). This material is displaced by the reacting black liquor solids and the resulting product solids that are primarily sodium carbonate and unreacted char carbon. Black liquor solids react with the steam and air mixture to produce a fuel gas, while the sulfur reacts primarily to form hydrogen sulfide, but also other reduced sulfur gas species.

Black liquor solids or black liquors (at 50% to 55% solids) are supplied to the reactor. The fluidizing air/nitrogen and steam are metered to the system and heated electrically to temperatures over the range of 900 to 1200°F. (Note that equipment limitations prevented exceeding 1000°F in operation.)

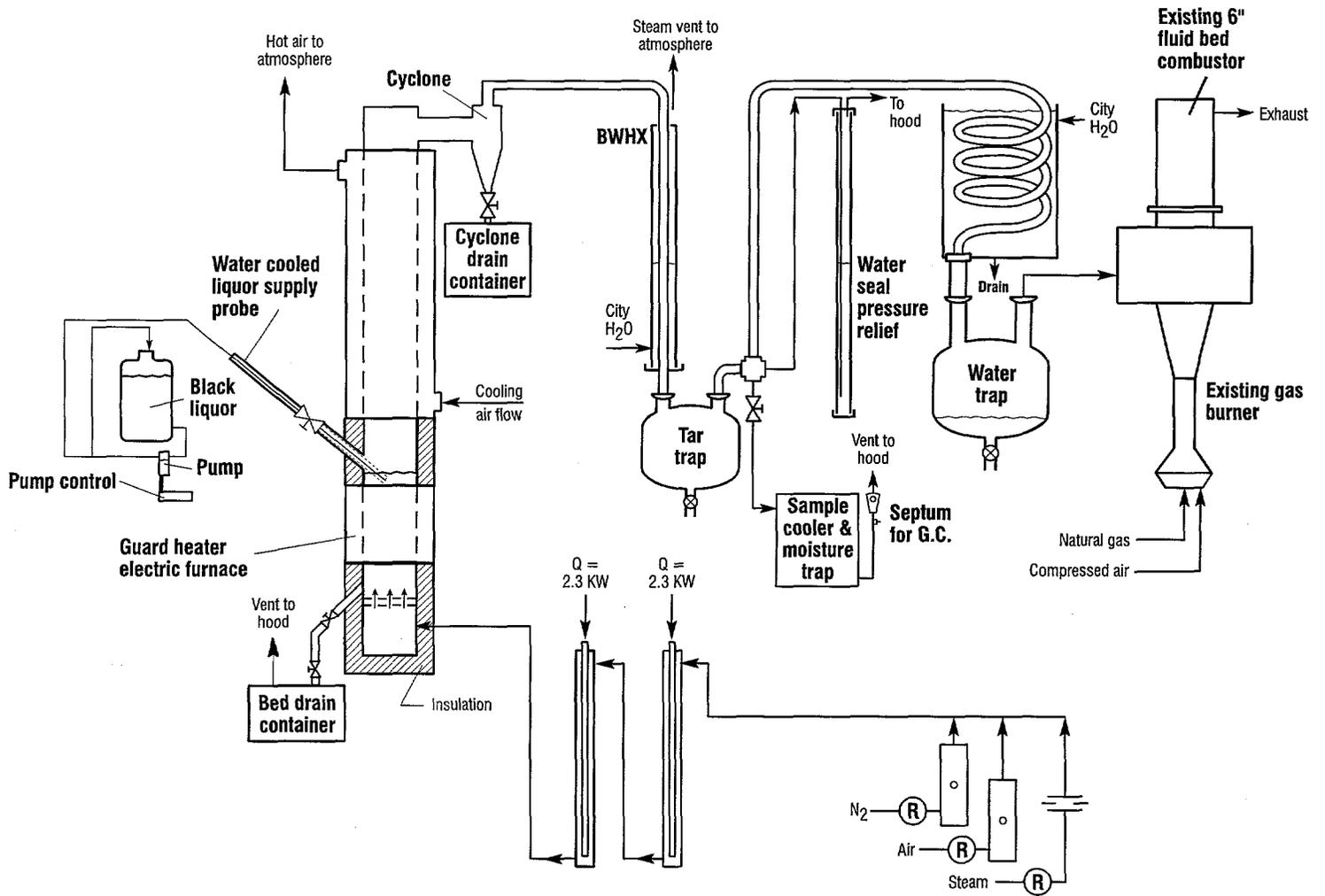


Figure 3-1. Schematic of the Bench-Scale Black Liquor Gasifier.

Product gas leaving the bed is cooled to approximately 500°F in the freeboard region of the gasifier. A cyclone removes particulate from the gas. The gas is cooled (to approximately 180 to 250°F) in a boiling water heat exchanger (BWHX) that simulates the heat recovery boiler. Any tars and water that condense in the BWHX are collected in a tar trap. Samples of the gas for analysis by gas chromatography are taken immediately downstream of the tar trap. The sample stream is cooled using an ice bath to minimize moisture in the samples.

The product gas is cooled to approximately 100°F in the condenser to simulate cooling for gas cleanup. Water and any low boiling point tars that condense are collected in the water trap. Since no gas cleanup or scrubbing is done to remove sulfur, the gas is thermally oxidized and discharged.

The bench-scale equipment before installing the insulation is pictured in Figure 3-2. The black liquor feed system, which was installed during the test program, is pictured in Figure 3-3. Additional details, which include photos of specific components, are included in Appendix A.

Bench-Scale Gasifier Test Results - Five tests, excluding the equipment shakedown and heater evaluation tests, conducted at the bench-scale provided data to address the objectives stated previously. These tests and key results of each are described in this section. More detailed test results are provided in Appendix B and analysis results for the liquor tested are provided in Appendix C.

The key results of the tests were:

- 1) Gasification was demonstrated under the partial-oxidation conditions, albeit intermittently, and for a short period in transition from combustion.
- 2) Tar formation was limited and the tars formed were shown to condense in the low-temperature equipment only -- thus tars appear manageable.

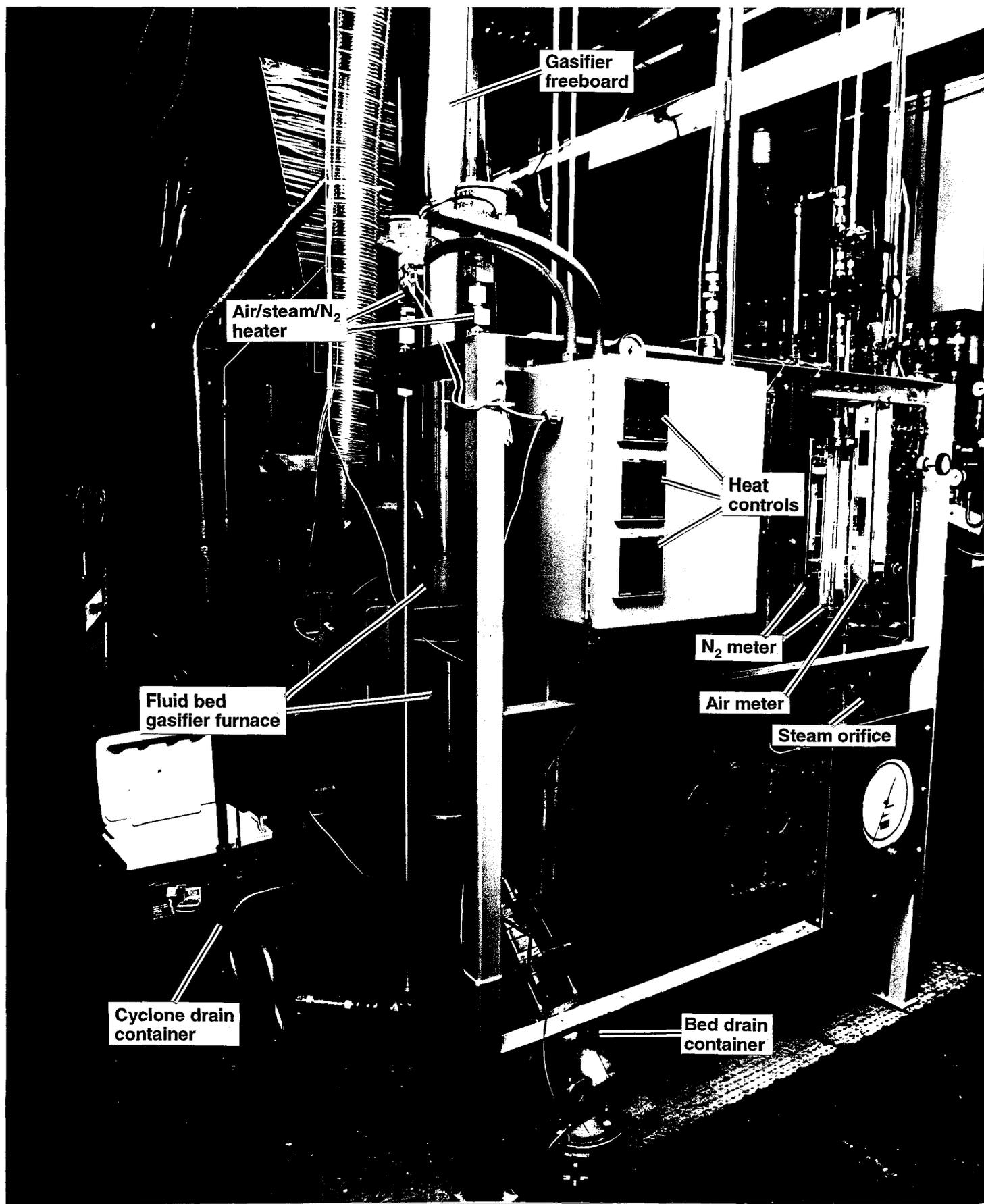


Figure 3-2. Bench-Scale Black Liquor Gasifier before Insulating.

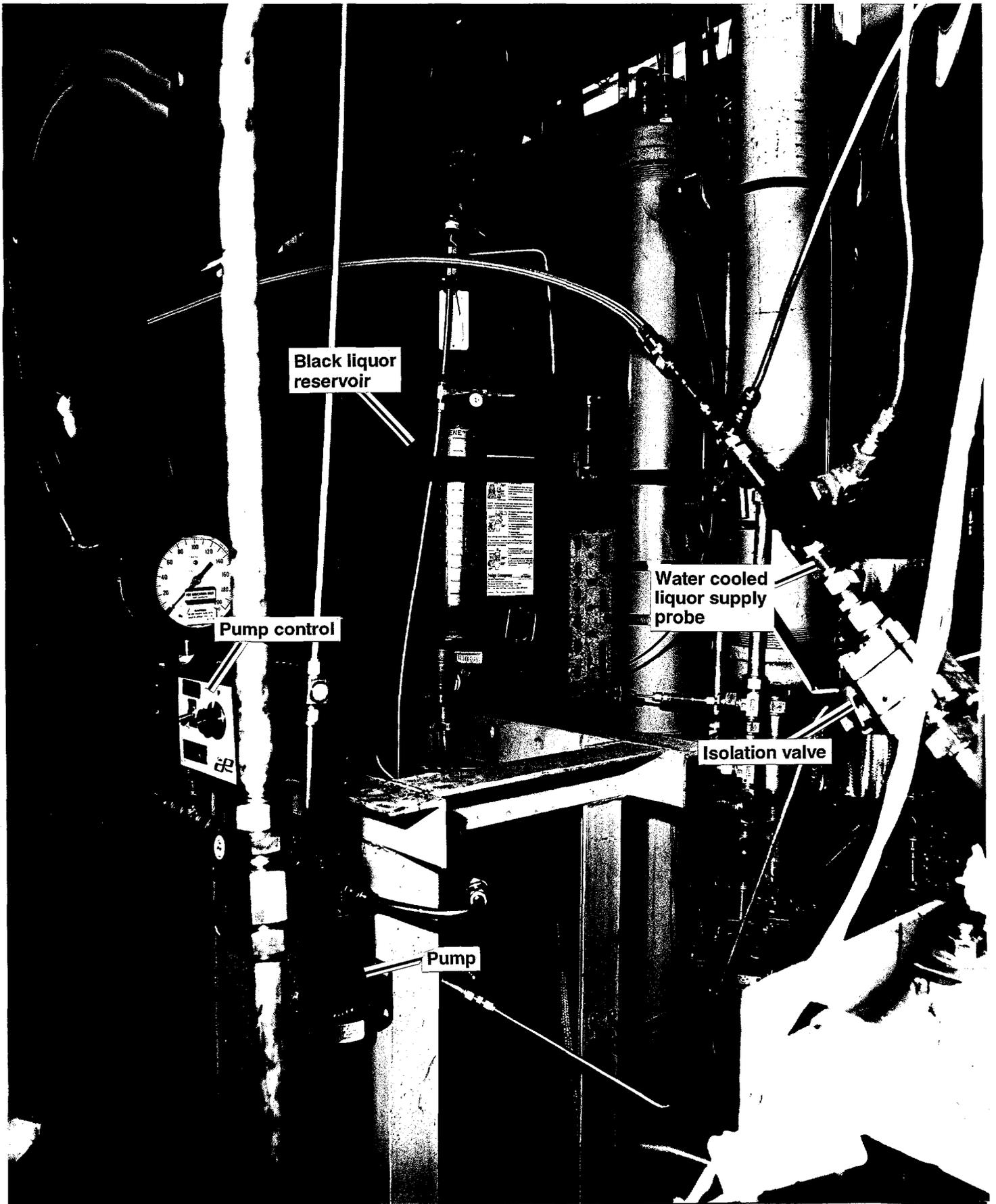


Figure 3-3. Bench-Scale Gasifier Black Liquor Feed System.

- 3) Char conversion was low, but the low temperatures of the test contributed to this. Higher temperatures should improve conversion. (Laboratory data show that char conversion rates climb an order of magnitude if the temperature can be raised to 1250°F) [5].
- 4) The bed material was drainable when fluidized.
- 5) Modifications to the test equipment to achieve a higher inlet temperature and eliminate cooler surfaces in the lower part of the reactor should make it possible to achieve sustained operation.

Shakedown testing was conducted to checkout the test hardware without liquor feed. These tests evaluated performance and operation of all supporting equipment that supplies and heats the reactants (air, steam, nitrogen) other than black liquor, the gas cooling equipment, and the solids feeder (operated with sand). Based on the results of these tests, some minor adjustments were made to ensure operation of these supporting components during the gasification runs. The maximum inlet temperature achievable in the fluidizing gas was 1000°F.

Test 1 [1st black liquor solids test - combustion]- This test supplied the reactor with black liquor solids and air at an air / fuel ratio consistent with combustion. Combustion with liquor solids was achieved, but the dry solids feed plugged where the liquor solids were introduced into the gasifier. Getting steady feed and moving to gasification conditions in this test was not possible. Operation under combustion conditions resulted in bed agglomeration when the temperature reached approximately 1370°F. Limited amounts of tar formed and deposited in the tar trap. Some minor amount of tar was observed mixed with water in the condensate trap as well. However, it was noted that the quantity of tars formed was small compared with the amount of solids reacted and did not foul the cyclone or the BWHX heat transfer surface. Because of the problems with solids feed, modifications to improve the feed system with increased motive nitrogen were made before Test 2.

Test 2 [2nd black liquor solids test - combustion and intermittent gasification]- Using the modified solids feed system, another attempt to achieve gasification conditions was made. This test intermittently produced gasification. The black liquor solids feed continued to plug where the material entered the gasifier. While gas chromatography (GC) results did not show gasification due to the intermittent nature of the gasification conditions, evaluation with an on-line oxygen analyzer showed that zero oxygen conditions existed periodically. Also, hydrogen sulfide and methane were detected by the GC analysis. These two results show that gasification conditions occurred, albeit intermittently, due to the intermittent feed of the black liquor solids. Unburned char collected in the cyclone catch, and relatively small amounts of tar collected in the tar trap and the condensate trap.

Test 3 [1st black liquor test - combustion and transitional gasification]- Due to the difficulties feeding black liquor solids and processing the solids to the form needed for the gasifier, a black liquor (liquid) feed system was installed. The system can supply sufficient black liquor to reach full gasifier capacity. This test run demonstrated that 55% solids liquor feed was successful and gasification conditions were achieved. The test was initiated under combustion conditions and then moved to gasification conditions by increasing the liquor flow. Sustained gasification conditions were achieved for approximately one-half hour of the test. The test was concluded when bed temperature excursions and high oxygen levels occurred due to the bed losing fluidization. This was due to the accumulation of unreacted char in the bed and operating at a lower bed velocity than design. However, under the gasification and pyrolysis conditions achieved, the quantity of tar was limited compared with the quantity of liquor consumed. Also, the tars were collected in the low temperature parts of the equipment. Unburned char was collected in the bed and in the cyclone catch. Most of the char in the bed was near the size of the sand used as the starting material. However, larger char agglomerates with mean diameters of one-half to one also formed. The agglomerates may have contributed to the loss of fluidization by settling on the air distributor plate producing flow maldistribution and channeling.

Test 4 [2nd black liquor test - pyrolysis] - This test attempted to return to the gasification conditions achieved previously, evaluate the potential to drain bed material from the unit, and operate the boiling water heat exchanger (BWHX) at more realistic minimum temperatures to evaluate tar deposition. Sustained gasification conditions that produced a quality product gas were not achieved. Failure to achieve gasification conditions was limited by the temperature reached in the reactor when attempting to move directly to gasification conditions from a lower starting temperature. Thus, the test was conducted essentially under pyrolysis conditions with air and steam. However, GC results do show that a significant quantity of reduced sulfur compounds was released, and that a limited amount of methane and other combustible species were formed. No significant amount of hydrogen or carbon monoxide was detected.

The bed was successfully drained during operation. The bed drain material is a mixture of char, residual sand from the starting material, and very limited amounts of consolidated inorganic material (agglomerated inorganic particles that can be detected visually). It is free flowing and composed of small granules that appear suitable for further handling and processing. Operating the BWHX above 220°F assured that no water condensate formed in the tar trap. Observation of the tar trap showed little tar accumulation. The tar observed was limited to a translucent coating of the inside surface of the one-liter tar trap vessel and a small accumulation of liquid tar (less than 20 ml) in the bottom of the tar trap. Tars were also observed in the condensate collected in the water trap downstream of the condenser, which reduced the gas temperature to approximately 80 to 100°F. This shows that operating heat transfer surface above 250 to 300°F, typically a lower limit for practical heat recovery boiler or economizer surface, may be sufficient to eliminate concerns of tar fouling.

The test also produced a significant quantity of unburned char in the bed and deposited a lower density char along the lower part of the freeboard. This material was removed after the test.

Heater Performance Test - Before conducting an additional test with black liquor, a test was conducted to establish the maximum bed temperature obtainable using external heating with a guard heater furnace. The results were used to decide if a final test at higher bed temperatures could be conducted without equipment modifications.

Test 5 - [3rd black liquor test - rapid defluidization] - This test attempted overcome the low temperature limitations of Test 4 by intensively heating the fluidized bed to approximately 1200°F using the guard heating furnace. The test failed to produce gasification; however, it did further characterize that the failure mechanism is due to limited inlet temperature and lower temperature surfaces located below the guard heater. These conditions cause the lower bed region to be quickly quenched by the significant amount of moisture in the liquor, which causes agglomerates to form on top of the distributor plate. The resulting channels cause a loss of fluidization, and impede heat transfer and reactions in the bed. Additional bench-scale tests that include equipment modifications are recommended to address this deficiency.

3.2 Pilot-Scale Gasifier (PSG)

The pilot-scale gasifier or PSG is the bridging step in black liquor gasification technology development between the bench-scale work and commercialization. It offers an opportunity to evaluate essentially all elements of the design at approximately 1/ 20th of the capacity of the projected FOAK commercial unit. The testing and modeling outline and the pilot-scale design, as prepared in Phase Ia are described in this section.

This activity began before completion of the bench-scale work and therefore was not refined to include the results of that work. Therefore, the pilot-scale design will need to be updated in the future. However, the task to develop the test / modeling outline, and advance the PSG design were valuable because a significant part of the results remains valid for the future work and provides a starting point for the future.

PSG Test / Modeling Plan Outline - This plan consists of three major elements that include testing and numerical modeling. The test and numerical modeling plans were integrated to assure that the two complimentary work areas are coordinated from the outset. The three elements of the outline follow.

- 1) PSG Model Development and Design Support - This element accomplishes the tasks of originating a numerical model to simulate the gasifier and to develop the experimental liquor and nozzle data needed to support the design of a PSG. Tasks described in the outline include:
 - Develop gasifier numerical model.
 - Screen mill liquors and spray nozzles.
 - Liquor and spray nozzle characterization.
 - Model liquor spray patterns.

- 2) PSG Commissioning and Performance Characterization - This element includes evaluating all systems as construction is completed, beginning operation of the PSG, and developing the ability to operate the unit in steady-state conditions for reasonable periods. Once steady operation is routinely achieved, data can be collected and compared with the numerical model. The model predictive capability can be evaluated and the model modified to the extent necessary, to give good comparison the PSG results. This includes the general tasks:
 - Shakedown tests.
 - Baseline characterization tests.
 - Model baseline operation.

- 3) PSG Performance Optimization and Commercialization Support - Using the model as a predictive tool, gasifier operating conditions and hardware modifications to optimize performance will be evaluated. Based on the model results, a course forward will be selected and followed. Tests to show extended operation of the PSG may then be undertaken, model improvements can be made based on the test results, and the PSG's sensitivity to process changes can be evaluated. Tasks to accomplish these objectives are:

- Gasifier performance modeling study.
- Gasifier performance optimization study.
- Extended operation tests.
- Model steady-state optimal performance.
- Test sensitivity to process changes.

Objectives of the PSG test / modeling approach are to evaluate performance at the pilot scale, and produce a tuned numerical model that can be used to predict gasifier performance.

The ultimate goal of this approach is to provide an adequate predictive capability that reduces the technical risk for designing the FOAK commercial unit. If that risk can be adequately managed, then investment in a FOAK incremental-capacity production unit will be commercially feasible.

The PSG approach will require changes from the design for which this test approach was outlined. However, the major elements of the test remain applicable, as does the underlying principal of integrated testing with numerical modeling to enhance the development process.

PSG Design - The Pilot Scale Gasifier, shown schematically in Figure 3-4, is designed to simulate the major features of the B&W gasifier design. It addresses key design issues and was used to solicit budgetary cost estimates from potential suppliers of major components. The PSG preliminary design work assumed that the equipment would be installed at an operating kraft mill located within the southeastern United States. Major component and subsystem functions are described in this section.

Black Liquor Feed System - Black liquor is received from the mill via pipeline and supplied through this system to the gasifier vessel. Mill selection will need to consider the specific liquor conditions available at the candidate mills. This includes the availability of black liquor at 65% solids or higher concentration, and at adequate temperature to maintain the liquor at a manageable viscosity for pumping. The system can lower liquor solids by dilution and can heat

the liquor as well. Liquor delivery into the gasifier vessel is by a retractable spray nozzle assembly. All components required for this system are commercially available. Instrumentation and controls would be similar to that for commercial systems.

Air / Steam Heater - The air/steam heater supplies the fluidizing gas, which consists of air and steam, to the gasifier. The system is fired with natural gas and it can recirculate part of the gas burner exhaust to the air inlet. This recirculation feature allows the heater to supply vitiated air with steam to control the oxygen level in the fluidizing gas. This flexibility is particularly important for startup and reduced-load operation. The heater is supplied by variable speed fans to provide the operational range needed for a test and development facility.

Gasifier Vessel - The main focus of the test program would be the performance of the fluidized bed gasifier. The gasifier vessel is sized to operate at approximately 1000 lbs/hr of black liquor (dry solids basis) flow, which is approximately 1/20th of the throughput of a small commercial application. The vessel consists of three sections, the air/steam plenum, the bubbling bed, and the freeboard region. The fluidizing air / steam is supplied to the plenum at the bottom of the vessel. It flows through a bubble cap distributor plate to the bubbling fluid bed. The bed is formed initially of sand, which is displaced over time by solids from the reacting liquor. Pyrolysis of liquor and steam gasification of char in the bed forms the low-heating-value product gas. The combustibles in the product gas are primarily hydrogen, carbon monoxide, and methane. Minor amounts of higher molecular weight organic compounds or tars may also be present. Sulfur is liberated primarily as hydrogen sulfide, but may also form other reduced sulfur species. The balance of the product gas is steam, carbon dioxide, and nitrogen, while the reacted solids will contain primarily sodium carbonate and carbon. Small amounts of other sodium salts will also be present. Potassium salts will be present in proportion to potassium in the liquor. The gas exits the vessel to the cyclone, carrying small amounts of solids as particulate while the bulk of the solids drains from the bed to the dissolving tank. The vessel includes penetrations for instrumentation and view ports. It also has a flanged connection at the distributor plate to provide access for inspection, and cleaning if necessary.

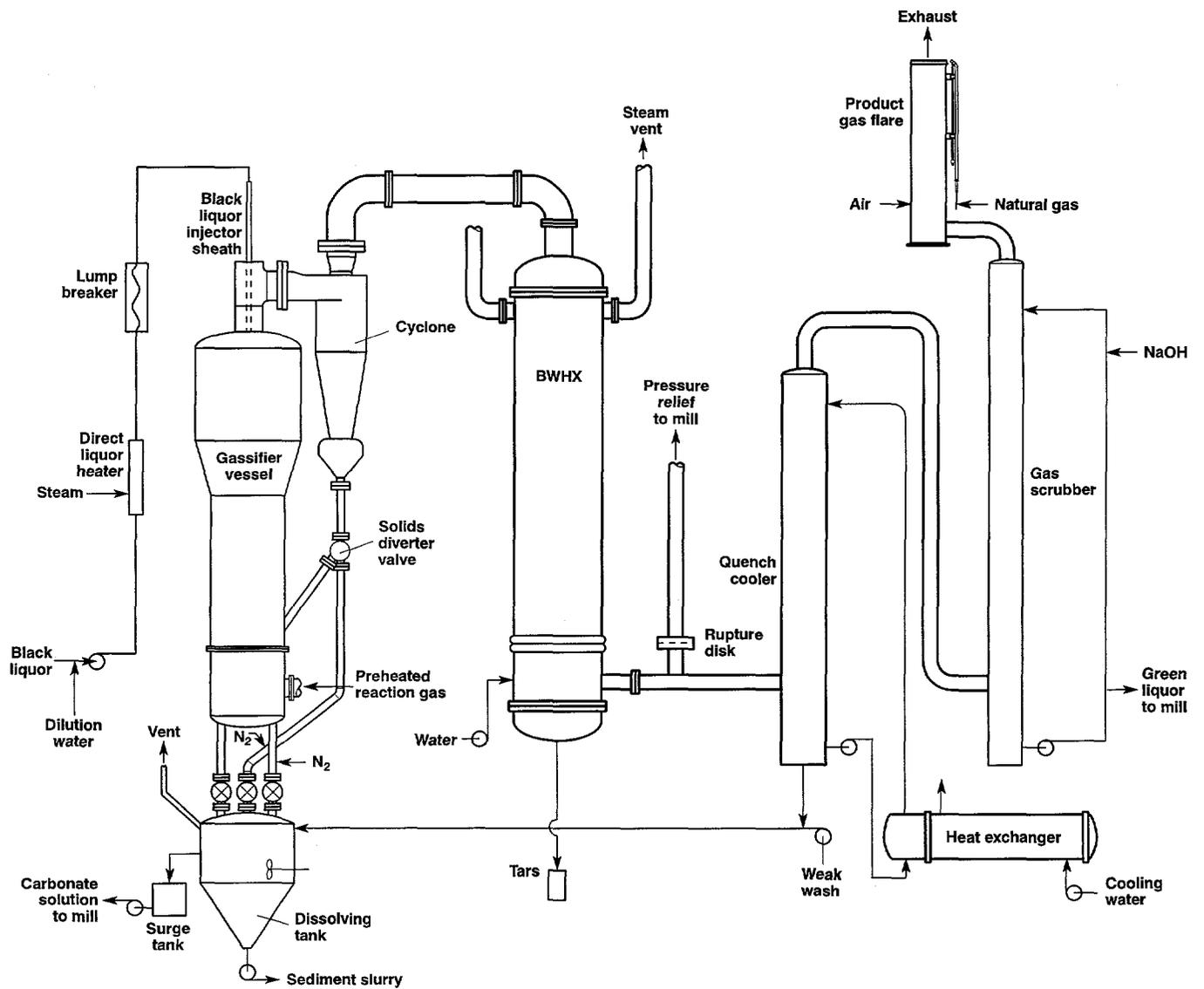


Figure 3-4. Pilot-Scale Gasifier Schematic.

Cyclone and Crossover Piping - The cyclone separates the majority of the particulate material from the product gas. The collected particulate is either recycled to the bubbling bed, or processed in the dissolving tank with the solids drained from the bubbling bed. The cleaned gas flows through the crossover piping to the boiling water heat exchanger (BWHX) for cooling prior to gas scrubbing.

Boiling Water Heat Exchanger (BWHX) - The BWHX cools the gas flowing from the cyclone (from approximately 1000 to 250°F) for processing. This shell and tube heat exchanger simulates the waste heat boiler of the commercial unit with a lower cost construction and greater flexibility of design. The gas flows on the tube side. Water boiling on the shell side cools the gas. Atmospheric pressure venting of the steam keeps the fabrication cost of this component low, and the ability to control the level of the boiling water provides variable heat transfer surface. This makes the BWHX more flexible for varying test conditions. Tars that condense on the long, vertical tubes will drain into the lower head. This design facilitates tar collection and removal, and provides access for cleaning, if it becomes necessary.

Dissolving Tank - Solids that drain from the gasifier and the cyclone dissolve in a mixture of weak wash from the mill and foul condensate from the quench cooler section of the product gas scrubber to form a caustic, carbonate solution. The product carbonate solution is returned to the mill, or may be diverted to the gas scrubber for evaluation as a scrubber solution. The dissolving tank includes provision for agitation, and removal of sedimentary solids (sand or other insoluble material) from the dissolving process.

Product Gas Scrubber - After the gas is cooled in the BWHX it flows to the product gas scrubber, which consists of a quench cooler stage and a scrubber stage. In the quench cooler stage, condensate is used to cool the gas from the BWHX outlet temperature to temperatures that are more optimal for scrubbing with caustic solution. The condensate is recirculated and cooled in a heat exchanger. The excess condensate is blown down continuously. Weak wash is combined with the blowdown and the mixture used in the dissolving tank. The scrubber stage

may be either a packed tower or a venturi scrubber. It uses sodium hydroxide solution from the mill to generate green liquor that is returned to the mill in routine operation. It also includes provision for using carbonate solution from the dissolving tank to permit its evaluation as a scrubbing solution as part of the test and development program.

Product Gas Flare - Since this is a small pilot, using the product gas as an energy source in the mill is not economically attractive. The gas will be adequately cleaned by the scrubber to permit it to be flared as a safe and economical means of disposal.

Refer to Appendix D for the full PSG test / modeling program outline, and to Appendix E for PSG process flow diagram, heat and material balance, and functional specifications.

3.3 Market and Economic Evaluation

Basis for Evaluation - A market and economic evaluation of alternative technologies for incremental recovery capacity compared gasification options with conventional technology. This work was conducted prior to the bench-scale testing, and was based on the gasifier design at that time. Because the economic analysis relies on best estimates of gasifier performance predictions for that design, the results therefore may be optimistic. However, the sensitivity analysis, which included gasifier (gas conversion) performance, demonstrates a clear advantage for gasification, particularly the combined cycle option that would be developed in the longer term.

The evaluation investigated the impact of capital costs, fuel costs, and operating and maintenance costs on financial performance. The three financial indicators were investigated:

- 1) Payback of the capital and sunk cost for the investment.
- 2) Internal rate of return for the investment.
- 3) Discounted cash flow of the investment.

These three indicators, respectively, measure of the relative risk of the investments, compare the performance of the alternative investments, and compare the impact of the investment on the value of the assets achieved by the alternative approaches. The three options evaluated were considered using three major assumptions.

- 1) The need for chemical recovery capacity is the primary justification for incremental capacity additions and the options considered are simply alternative means to achieve that end.
- 2) The energy products from the processes, which may include: high-pressure steam, low-heating-value gas, electricity, or low-pressure process steam, can be consumed within the mill or sold.
- 3) The values for the energy products other than process steam are set by market values as projected by DOE/EIA [12]. The value for process steam was taken from prior work on Phase I of this project [1]

Approach - The economic evaluation considered an incremental-capacity installation to process approximately 490,000 lbs. black liquor (dry solids) per day [based on a liquor higher heating value of 6020 Btu/lb dry solids] via different recovery approaches. This compared three options:

- Conventional recovery boiler technology (i.e., a small recovery boiler) producing high-pressure steam.
- A near-atmospheric pressure gasifier producing a low-heating-value gas.
- A pressurized gasifier / gas turbine combined cycle producing electricity and low-pressure process steam.

The financial performance of the two gasifier options was evaluated using the conventional recovery boiler as the baseline. The sensitivity analysis of the financial performance considered the impact of variations in the value of the energy outputs produced (high-pressure steam, low-heating-value gas, electricity and low-pressure process steam), capital cost, annual fixed operating and maintenance cost, and gasifier performance.

The future market potential was also reviewed and a scenario for commercialization of low-temperature gasification that is consistent with the development approach described in Section 4. was determined to be reasonable.

Results - Table 3.1 provides a qualitative summary of the financial performance of the two gasifier options compared with conventional technology. Here “Competitive” is option has equal or superior financial performance to the analogous small recovery boiler case.

Additional details concerning the market projections, the scenario for commercial development of the black liquor gasifier, and the economic evaluation of black liquor gasifier options versus conventional technology for incremental capacity addition are provided in Appendix F.

Table 3.1. Relative Financial Performance of Small Recovery Boiler, Black Liquor Gasifier (BLG) Producing Gas, and Black Liquor Gasifier / Gas Turbine Combined Cycle (BLG/GTCC) for the Sensitivity Variables Considered.

Sensitivity Variable	Technology Competing with Small Recovery Boiler	
	BLG Producing Gas	BLG/GTCC
Fuel Price Scenario (<i>Basis for the value of the energy output</i>)		
- High	Competitive	Competitive
- Mean	Competitive	Competitive
- Low	Competitive	Competitive
Capital & OTF Cost		
- High (Base Case +15%)	Not Competitive	Competitive
- Mean (Base Case)	Competitive	Competitive
- Low (Base Case -15%)	Competitive	Competitive
Annual Fixed O&M Cost		
- High (3% of TPC)	Competitive	Competitive
- Mean (1.5% of TPC)	Competitive	Competitive
- Low (1% of TPC)	Competitive	Competitive
Gasifier Performance		
- High (Base Energy Output + 7%)	Competitive	Competitive
- Mean (Base Energy Output)	Competitive	Competitive
- Low (Base Energy Output -7%)	Competitive	Competitive

TPC - Total Plant Cost - Allowances for A&E, owner engineering and management costs, project contingency, pre-production and startup costs, and inventory - refer to "Additions to Capital Cost," in Appendix F.

OTF - One-Time-Fixed Cost - Cost incurred by production loss during installation - refer to "Fixed Cost" in Appendix F.

Industry Impact

Prior to initiating Phase Ia, B&W conducted market interviews with 25 pulp and paper industry executives and technical personnel regarding the future of black liquor gasification. Forest products companies represented in these interviews own approximately 45% of the U. S. recovery boilers. The response was characterized as a uniform interest in gasification. Confidence that gasification would become a factor in the future ranged from "certain" to "probably not realistic any time soon."

Our internal studies show that over the next two decades a significant number of recovery boilers will be replaced and a significant quantity of new recovery capacity will be added. The total is estimated to be 12 million lbs. dry solids / day liquor processing capacity installation on average within the next two decades. This will include a mix of incremental capacity, replacement units, and new units.

Black liquor gasification is beginning to play a role in the incremental capacity market with the first commercial system being installed in North America and scheduled to begin operation within the year. The systems described for development in this report could enter a similar role in 2005. These low-temperature systems could offer higher thermal efficiency over the current technology.

4.0

FUTURE DEVELOPMENT PLANS AND MEANS

The first step forward for this program is to select which technology to pursue, high-solids advanced combustion or gasification. That would be followed by the steps necessary to advance that technology. The steps in that path forward are listed briefly here and then described in more detail.

Decision Point:

Technology Selection - Recommendation: Pursue gasification technology in this program over high-solids advanced combustion technology.

Agenda 2020 project:

Additional Bench-Scale Work - Modify the bench-scale equipment to evaluate the changes in the gasifier design concept that resulted from Phase Ia. (This project has been proposed to DOE.)

Decision Point:

Continue Development? Based on the additional bench-scale data, determine if the technology remains viable.

Future Project:

Evaluate the Impact on the Incremental Capacity Gasifier Design and Economics.

- Considering the results of the Phase Ia and the additional bench-scale tests, determine the impact on the gasifier performance and economics. Update pilot-scale design concept, development cost estimate, and the program schedule.

Decision Point:

Continue gasification development? Based on the evaluation results, determine if continuing with further development remains justified.

Future Major Program:

Pilot-Scale Program - Design, build and test the gasifier at the pilot scale.

(Assuming the decision is to continue is justified.)

FOAK Commercial Demonstration

Using the pilot-scale testing and modeling results, pursue the first commercial demonstration unit to provide incremental liquor processing capacity. *(Note that this activity is commercial and is beyond the scope of the DOE co-funded work scope.)*

Schedule and Cost

The estimate costs and schedule for these activities are provided in Table 4.1 and Figure 4-1.

The Agenda 2020 project that was proposed to DOE consisted of additional bench-scale experimental work to address the impact of improving char and tar conversion using a novel

gasifier design. The proposed work included consideration of the effects of high-chlorine liquor on the design.

The future project and future major program are budgetary estimates. Cost shares and final pricing would vary based upon the scope and specific business arrangements pursued.

Table 4.1. Estimated Costs for Agenda 2020 and Recommended Future Work.

<i>Description</i>	<i>Agenda 2020 Project^{1,2}</i>	<i>Future Project^{1,3}</i>	<i>Future Major Program^{1,3}</i>
Direct Cost (Labor + material)	\$242K	\$150K	\$6630K
Indirect Cost (Overhead + COM)	\$58K	\$36K	\$1590K
Total Cost	\$300K	\$186K	\$8220K

- 1) All costs show at time of performance assuming 5% escalation per year.
- 2) Proposed cost
- 3) Budgetary cost

<i>Project / Program</i>	<i>1997</i>	<i>1998</i>	<i>1999</i>	<i>2000</i>	<i>2001</i>	<i>2002</i>	<i>2003</i>
Agenda 2020	x-----x						
Future Program		x-----x					
Future Major Program			x-----	-----x			
Commercialization				x---	-----	-----	-----x

Figure 4-1. Schedule for Agenda 2020 and Recommended Future Work.

4.1 Technology Selection -- High-Solids Advanced Combustion or Gasification

Further investigation of gasification at the bench-scale is recommended, but larger-scale gasification research program is not recommended at this time. Other DOE-sponsored work in black liquor combustion that addresses the broader range of combustion and fireside deposition issues for conventional recovery technology will also be valuable for advancing high-solids firing technology.

High-solids advanced combustion is an extension of current recovery boiler technology. In general, the trend has been toward increasing solids levels in black liquor that is being fired in recovery boilers in recent years [13, 14, 15]. The solids levels are being increased incrementally, thus this technology is evolving through commercial activity. The gains possible by increasing the solids level will continue to diminish as the incremental process improvements are made. It is recommended that improvements in high-solids advanced combustion be limited to other precompetitive programs to improve fundamental understanding of black liquor combustion, advance the numerical modeling of those processes, and develop supporting data to validate those models. B&W's role in these areas is being addressed in other DOE programs.

Black liquor gasification is an emerging field that includes some limited commercial product offerings. Further pursuit of the precompetitive development of this technology remains sufficiently promising to merit additional investment. It continues to offer potential opportunity for increased self-generation of power, reduced emissions, and cost effective, incremental liquor processing capacity.

4.2 Additional Bench-Scale Work and Impact of Results on Commercial Potential

The results of Phase Ia, while encouraging, are insufficient to justify moving forward with anything other than additional investigation at the bench scale. The results also suggest that modifying the process design to improve tar and char conversion during gasification is prudent.

Agenda 2020 Proposed Project - The bench-scale tests showed that a limited amount of tar forms during gasification and that it condenses at relatively low temperatures. Therefore, tars may not be a fouling problem for the heat exchange equipment, but may be collected primarily in foul condensate from the gas cooling section of the gas cleanup system. Also, a significant amount of char, which is primarily carbon, will be generated in the fluid bed and will be available in the bed drain and cyclone catch byproduct streams. Our approach to handling these byproduct streams is a novel modification of the gasifier design. This requires additional testing

at the bench-scale to evaluate. It is recommended to first conduct additional tests at the bench-scale to evaluate the impact of improved char and tar gasifier performance. This will include characterizing the product gas, char, and tar streams to determine their quantities and composition using the novel gasifier design approach that resulted from Phase Ia. Based on the results of the additional bench-scale tests (Agenda 2020 project), a decision of whether or not to continue can be made.

Future Project - The changes in gasifier design will affect its performance. Based on the results of the current and additional bench-scale tests, the performance of a commercial scale, incremental capacity unit can be estimated. The existing process simulation models of the gasifier can then be updated. Then, relative economic performance of the gasifier can also be evaluated.

If the process and economics evaluations are favorable, consideration can be given to moving to the pilot scale. This would require several tasks:

- o Updating the pilot scale design, and the test / numerical modeling outline sufficiently to support cost estimating.
- o Updating cost estimates and schedules.
- o Reviewing the results with potential industry cosponsors and identifying interested participants including candidate host sites.

4.3 Decision Point - Continue Gasification Development

Based on the performance estimates for a commercial gasifier and the projected economics for that gasifier the potential benefits of continuing development of this technology could be quantified.

Other factors to consider at this point will include:

- o The level of continued industry interest in the process.
- o Additional investment required to develop the technology.
- o Potential environmental benefits, and safety and operational benefits.
- o The relative benefit of this technology compared with other competing technologies.

Based on this total picture, a decision of whether or not to begin a major pilot-scale program can be made.

4.4 Pilot Scale Gasifier (PSG)

If the decision is made to continue, then design, fabrication, installation and testing of a PSG would be the next steps toward a commercial product.

PSG Design - The major tasks involved in the design begin with test and numerical model planning and then progress through the design stages.

Update the Test and Numerical Modeling Outline - This outline would be updated to assure that the pilot-scale design simulates the appropriate elements of the commercial design.

PSG Preliminary Design and Review - The PSG design would be updated based on the results of the bench-scale tests and participation from potential host sites. The initial PSG numerical modeling and tests noted in Section 3.2 would be coordinated to support the preliminary design effort. Typical preliminary design outputs include the first release of

process and instrumentation diagrams, general arrangement drawings and process flow diagrams, and preliminary functional and component specifications. The preliminary design review would identify the major open issues that must be addressed in the balance of the design and assure that approaches for closing those issues are available.

Consideration of host site issues would be included in this stage. This stage would also update budgetary bids by suppliers and update the estimate to complete the design, fabrication, and construction of the equipment.

Final Design and Review - The design would be completed and all open items from the design review would be addressed and closed. The design outputs would include design drawings, functional specifications, and equipment specifications. This task would include extensive interaction with the host site and external suppliers.

Procurement, Fabrication, and Construction -Tasks include:

Procuring and fabricating components.

Controlling engineering changes and coordinating these with the host mill, and internal or external supplier organizations.

Conducting on-site component inspections prior to shipment and coordinating rework, repairs, or modifications.

Receipt and control of material at the site.

Site preparation and construction.

Testing and Numerical Modeling - After construction is completed, the test program outlined in Section 3.2 will be conducted. This includes two major parts.

PSG Commissioning and Performance.

PSG Performance Optimization and Commercialization Support.

The type of test and numerical modeling activities that would be conducted noted in Section 3.2. Additional detail for specific tests and modeling objectives for these major areas is given in the PSG Testing and Modeling Outline, provided in Appendix D.

First-of-a-Kind Commercial Demonstration - This technology would be introduced to the industry as a means to add incremental liquor processing capacity to existing mills with minimal downtime. The program described is to reduce the technical risk associated with the technology to an acceptable level for a first commercial unit.

The target capacity for the first incremental capacity units is 490,000 lbs. dry solids / day. However, the needs for capacity are site-specific. So the gasifier capacity may vary from this target.

4.5 Industry Interest and Support for Future Work

No established or developing alternative process can match the superior pulp quality, insensitivity to wood species, low operating costs, and proven recovery technology of the kraft process. Minor modifications to the current recovery process will satisfy the next generation of kraft pulp mill improvements.

Black liquor gasification (BLG) is currently foreseen as the replacement of the Tomlinson recovery boiler. Market penetration will be slow due to the capital demands of this equipment and the conservative attitude of the Industry toward new technology. Gasifier systems should be competitive with Tomlinson boilers by the year 2020.

The Future of the Kraft Process - Based on published opinions, a modified version of the kraft pulping process will continue to dominate the Industry into the next century [16,17,18]. The persistence of the kraft process to the year 2020 is implied in vision statements issued by the U.S. Department of Energy [6] and American Forest and Paper Association [7].

Kraft Recovery Process Alternatives - Although a proven technology, a replacement for the Tomlinson recovery boiler has long been sought. Due to large capital investment and long operating life (40 years), recovery boilers frequently become the "bottleneck" in pulp production. They are also prone to corrosion and catastrophic smelt-water explosions.

Black liquor gasification is a partial combustion process in which the organic compounds are converted into fuel gases. If pressurized and coupled with gas turbines, gasification systems can provide more efficient utilization of black liquor fuel value and produce more electrical power relative to steam [9,19]. This is an attractive feature for future mills where higher on-site electrical generation will be required to operate mechanical pulping and pollution control equipment. Sodium salts are recovered in a solid or molten form and can be causticized to regenerate NaOH. Most BLG concepts avoid the risk of smelt-water explosions. The majority of the sulfur is converted to H₂S during gasification and must be scrubbed from the product gas by caustic solution. This requirement facilitates production of split-sulfidity pulping liquors, which may be favored in future kraft pulping operations [20].

BLG is currently envisioned as the full-scale kraft recovery technology of choice for the next century [7,21,22]. However, it is realistically predicted to be competitive with Tomlinson boilers by the year 2020 [9,11]. Before BLG is accepted, it must be proven in a number of installations and offer process and economic benefits over traditional recovery technology.

BLG with Sulfur Recovery and Conventional Causticizing - Grace and Timmer [3] recently reviewed the status of five BLG concepts under development, and compared the predicted energy efficiency of the gasifiers at projected commercial scale. Given available process data and a

common liquor analysis, they calculated the efficiency of converting the black liquor fuel value into net product gas for an atmospheric pressure gasifier with no combined-cycle power generation. As a basis of comparison, they estimated between 59-61% of the heating value of black liquor is converted to net heat to steam in a Tomlinson recovery boiler. The analysis demonstrated that a low-temperature gasifier (similar to the B&W concept) is the only design, of those evaluated, offering higher energy efficiency than conventional recovery boiler technology.

Future Industry Direction - Assuming the first commercial BLG in North America is effective, this technology may be considered over recovery boiler rebuilds as a means to increase incremental solids processing capacity. However, only after satisfactory demonstration of a number of small-scale installations, including combined-cycle technology, will gasification be considered a viable alternative to new Tomlinson-based recovery systems.

Industry Support for Future Work - Low-temperature black liquor gasification has the potential to benefit the U. S. pulp and paper industry significantly. As noted previously, B&W has continued to evaluate the potential for this technology. Letters of interest from officials of two major pulp and paper companies (The Mead Corporation, and Champion International Corporation) supporting our proposed work under the Agenda 2020 program, and future work in BLG are shown in Figures 4-2 and 4-3. These letters indicate an ongoing interest in developing this technology for the future. Based on 1993 total sales, these companies are ranked 7th and 8th among North American paper products companies [22].

Corporate Engineering
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November 15, 1996

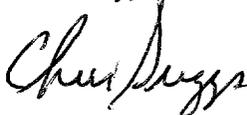
Mr. Jim Dickinson
Manager, Industrial Design Engineering
Babcock & Wilcox Co.
20 S. Van Buren Ave
Barberton, Ohio 44203

Dear Jim:

I am writing to express our support for B&W's efforts to develop a low temperature black liquor gasifier. There is a lot of interest in gasification of liquor as an alternative to traditional recovery boilers. The prospect of installing incremental capacity at lower capital cost is extremely attractive to the industry. We have already seen one commercial installation of a high temperature gasifier that offers capacity but lacks energy efficiency improvements that are sorely needed. A concept that offers potential energy efficiencies suitable for high efficiency combined cycle systems is most desirable.

Please keep us informed of your progress. You may recall that we had some discussions with Hayes Orender in 1994-1995 about using our Chillicothe, Ohio Mill as a host for pilot work. We would be pleased to open discussions again on that possibility.

Sincerely,



Chris Suggs
Manager, Facilities
Power Technology

Figure 4-2. Letter of Interest from The Mead Corporation.

To:
· Jim Dickinson

Date:
November 26, 1996

From:
Ron McCarty

Subject:
Support of the DOE Black Liquor
Gasification Technology

This memo is to indicate Champion's continued interest and support for the effort by B&W, in the DOE, study to advance the proposed low temperature gasification technology that is currently being evaluated.

The potential benefits of the increased process safety, improved energy efficiency, and reduced emissions, will have long term positive implications on our company and the industry. The combined cycle of burning the high heat value gases in a turbine will address the current industry trend of requiring greater levels of electrical energy for the process and increased emission controls systems. The elimination of the smelt from the chemical recovery process will address one of the highest risk areas in our industry.

Champion is interested in the progress of this study and we continue to support the objectives of the program

cc: B. Bannan
E. Kelleher

Figure 4-3. Letter of Interest from Champion International Corporation.

5.0

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6.0

APPROVALS

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W. T. Southards

Section Manager  Date 6/30/97
T. A. Morris

APPENDIX A

BENCH-SCALE TEST UNIT DESIGN AND ARRANGEMENT

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BENCH-SCALE TEST UNIT DESIGN AND ARRANGEMENT

The bench-scale test unit was designed to simulate many of the design features of the B&W gasifier design at a small scale. A simplified schematic of the bench-scale black liquor gasifier is shown in Figure A-1. The key test item is the fluidized bed gasifier reactor (on the left of the figure) in which gasification occurs. The balance of the equipment performs several functions.

- Supplying the reactants and diluents (black liquor solids, and heated air, nitrogen, and steam) to the gasifier.
- Removing gas, solids, and heat from the gasifier.
- Cooling the product gas downstream of the gasifier, and removing tar and water that condense during cooling.

These components and their functions within the system are described in this appendix. Photos of the main components are also provided.

A.1 Gasifier Vessel

The stainless steel gasifier vessel is approximately six inches in diameter and eight feet in length (see Figures 3-2 and A-2). It consists of an inlet air plenum or windbox, an air distributor plate, the bubbling fluidized bed reaction zone, and the free board. Connections to the gasifier include the heated reactants and diluents supply, the black liquor supply, the product gas outlet, and the bed drain. Several minor penetrations are provided for thermocouples and pressure taps used to monitor bed operation. The inlet air plenum is insulated. The fluid bed reaction zone is surrounded by an electric furnace to provide guard heating and additional heat input. The freeboard is surrounded by an air cooling jacket to remove heat.

Gasification reactions occur in the bubbling fluid bed. The bed dimensions and flow rates were set to assure a target residence time of approximately one second while maintaining good fluidization at the operating load. In operation the bed may vary in depth from 12 to 24 inches, with 18 inches being the target operating condition. The initial bed is formed with sand heated to the temperature range desired for testing prior to introduction of the black liquor. The size distribution of sand typically used for these tests is provided in Attachment A.1. The bubbling bed region is surrounded by an electric furnace that provides guard heat to control heat loss, or may be operated at higher power to add heat to the bed zone. The furnace (guard heater) active length extends from approximately three inches above the face of the air distributor plate, which forms the bottom of the bed, to the target operating level for the top of the bed (approximately 18 inches above the air distributor plate).

The bed temperatures are monitored by four thermocouples, at 1- 3/4", 11", 18", and 24.5" above the air distributor plate. These temperatures are monitored and recorded by computer during testing. The overall bed pressure drop from the air inlet plenum or wind box (immediately upstream of the air distributor plate) to the free board outlet of the gasifier vessel is monitored manually using a differential pressure gage. The combined information from the in-bed temperature and bed differential pressure measurements indicate approximate bed height. These measurements are used to determine bed conditions during tests.

Above the fluid bed zone of the gasifier is the free board. The gas cools as it passes through freeboard after exiting the fluid bed reaction zone. The bench-scale test does not simulate overbed spray drying of incoming liquor. This is due to the small size of the unit, which makes forming a spray to dry incoming liquor impractical. Gas exits the free board through a single connection to the cyclone and downstream gas cooling equipment.

A.2 Black Liquor Supply System

Initially, the bench-scale reactor was operated with dried black liquor solids. A screw feeder with a closed and nitrogen blanketed hopper delivered black liquor powder to a drop tube and then, with the aid of a motive nitrogen supply, to the gasifier via a feed tube (Figure A-2). This approach was taken so that the dry liquor solids would simulate liquor reaching the bed at very high solids as is expected to occur with the overbed spray and to avoid the difficulties of handling high-solids black liquor. After attempting two tests; however, this system proved impractical due to problems with preprocessing the black liquor solids and with the solids swelling and plugging the feed tube.

It was decided to pursue the alternate path of feeding black liquor as a liquid at 50 to 55% solids. The solids feed system was removed, and the black liquor feed system was installed. This resulted in unwanted moisture entering the bed, but was considered a practical alternative.

The liquor feed system consists of a liquor container connected to a gear pump with a variable speed controller (see Figure 3-3). The liquor is pumped through a probe, which is cooled with hot water, that discharges the liquor just above the nominal bed surface. The system has the option of bypassing the liquor flow back to the container for startup. It also includes a nitrogen purge available to ensure that the supply probe can be kept open during insertion and removal when there is no liquor flow. Rate checks conducted over a range of pump controller settings demonstrated repeatable flow rates. Pump controller settings and standard analyses for black liquor solids are used to determine the liquor solids flow during testing. Flow rates are checked for each liquor batch to assure adequate performance.

A.3 Air / Steam / Nitrogen Supply System

The reactants and diluents supplied to the gasifier are air, steam, and nitrogen. These incoming constituents are metered separately, mixed and then heated prior to introduction into

the gasifier (see Figure 3-2). Rotameters are used to measure the air and nitrogen flows. An orifice and differential pressure gage is used to measure the steam flow. Temperature and pressures of the air, nitrogen, and steam are measured and used to correct the flows for differences from the calibration conditions to the operating conditions. Air from plant air system is filtered in three stages to remove any water, oil, or oil vapors. The steam is generated by a small dedicated electric boiler. The nitrogen is provided from compressed gas cylinders.

At full operating conditions, only air and steam were planned to be used. However, the gasifier is also supplied with nitrogen to allow independent adjustment of the oxygen content of the incoming gas in proportion to the fuel supply. This enables the operators to maintain a prescribed steam to fuel ratio, stoichiometry, and the target fluidizing gas velocity.

The incoming gases (air / steam / nitrogen) are preheated using electrical heaters, which are easier to construct and install at the bench scale than the gas fired heating planned for the commercial gasifier design. The heated air / nitrogen / steam mixture enters the reactor via a plenum and air distributor plate. The plate uses a bubble cap design proven from prior work in combustion fluidized beds. The criteria for gas flows that produced good fluidization were set based on the methods developed by Battock & Pillai and presented in Reference A.1. This method was confirmed with data from prior fluidized bed combustion work by B&W and others.

A.4 Gas Particulate Removal

Product gas exiting the gasifier is passed through a small cyclone to remove particulate. This is analogous to the first step in gas cleanup in the B&W gasifier design, which also uses a cyclone to remove most of the particulate. The small cyclone in the bench-scale unit drains into a vented drain container that may be isolated and removed from the system if desired. The gas temperature at the gasifier outlet (free board temperature) and the temperature of the gas line exiting the cyclone are monitored manually using thermocouples and a readout device.

A.5 Gasifier Bed Solids Drain

Accumulating solids may be drained from the fluid bed during operation into a sealed bed drain container. The container can be vented, isolated, and removed from the system during operation if desired. Drain containers for the bed and cyclone are interchangeable so that both may be served by a single spare container (Figure 3-2).

A.6 Product Gas Cooling and Sampling

After the product gas leaves the cyclone it is cooled in two stages and is sampled for analysis. Traps for tar and water condensed from the gas are provided. (See Figure A-3 for components described in this section).

Boiling Water Heat Exchanger - The gas is first cooled on the tube side of a single tube boiling water heat exchanger (BWHX) to approximately 180 to 250°F. This approach simulates the gas cooling by the waste heat boiler in the B&W design. The exit gas temperature was selected as conservatively below the practical lower limit for heat recovery and permits evaluation of tar condensation. This temperature is monitored manually during testing. The BWHX also provides a significant degree of operating flexibility in that changing the water level on the shell side provides a means to vary the effective heat transfer surface of the heat exchanger. This in turn provides a means of roughly controlling the exit gas temperature from the heat exchanger.

Tar Trap - Gas leaving the BWHX passes into a one liter glass vessel that serves as a tar trap and partially simulates the tar trap of a full scale unit. However, the vessel has a large surface area per unit volume, and its interior surface temperatures are cooler than would be anticipated in commercial applications. Thus, the conditions are more favorable for tar condensation, than anticipated for commercial applications.

Pressure Relief Protection - Pressure relief protection for the system is provided on the exit side of the BWHX. Due to the very low operating pressure and the small size of the unit, most commercial pressure relief devices are impractical. So pressure relief is provided by a water column. Any gas that may be released is vented to the laboratory fume hood system.

Gas Sampling for Analysis - Gas sampling for gas chromatography (GC) analysis is provided immediately downstream of the tar trap. A branch line for gas sampling connects to a glass filter holder that acts as an initial cooler and trap to remove moisture and tar carryover. The sample line then connects to two glass impingers in series. The impingers are cooled in an ice bath to trap moisture. The gas then flows through a rotameter used to assure that the sample flow rate is sufficiently high. GC samples are withdrawn from the septum between the ice cooled impingers and the rotameter using syringes. The samples are then taken to the GC lab and injected into the GC equipment for analysis. Gas is sampled for several minutes to assure the line is purged of non-representative gas that may be present prior to withdrawing GC samples. The GC analysis methods are designed to detect H₂, N₂, CO, CH₄, CO₂, C₂H₂, C₂H₄, C₂H₆, H₂S, COS, CH₃SH, (CH₃)₂S, and CS₂. The GC analysis methods are described in Attachment A.2 of this appendix. These methods compare the levels attributed to the various species in the samples with those from analyses of known commercial standards containing the species being evaluated.

Condenser and Condensate Trap - Downstream of the tar trap, the gas is further cooled to condense water and lighter tar fractions. Also, some tars that may be conveyed as fine aerosols are captured in the condensing water. This condenser consists of 50 feet of Teflon tubing that is cooled with city water. At the exit of the condenser is a two-liter glass vessel that is the condensate (or water) trap. Its purpose is to collect condensed water and tars. The gas is intended to be cooled to 100-150°F in the apparatus, but may be cooled to a lower temperature with sufficient water flow. The gas temperature at the exit of the condensate trap is monitored manually using a thermocouple during testing. Analogous to the BWHX, the heat transfer surface is arranged such that the effective surface area may be varied by adjusting the water level in the condenser tank. This provides a means to adjust the condenser outlet gas temperature.

Continuous Oxygen Analyzer - Downstream of the condensate trap, a gas sample line branches to an on-line oxygen analyzer used to guide testing. When operating in combustion and moving to gasification, for example, the test personnel can observe the rapid drop in oxygen content as gasification conditions are approached. A zero oxygen reading should be observed when the system is operating in a gasification mode. Gas from the analyzer is vented to the laboratory fume hood system.

A.7 Gas Disposal

The bench-scale system does not include a scrubber system to remove sulfur compounds from the product gas. Instead, the gas is disposed of by thermal oxidization in a fluidized bed combustor that uses limestone as the bed material, which acts as a sorbent for the sulfur in the gas. The combustor is an existing facility typically used for fuel and sorbent testing under fluid bed combustion conditions. The combusted gas is exhausted from the facility in the normal means used for combustion testing, which meets all local codes and emissions regulations.

A.8 Safety Considerations

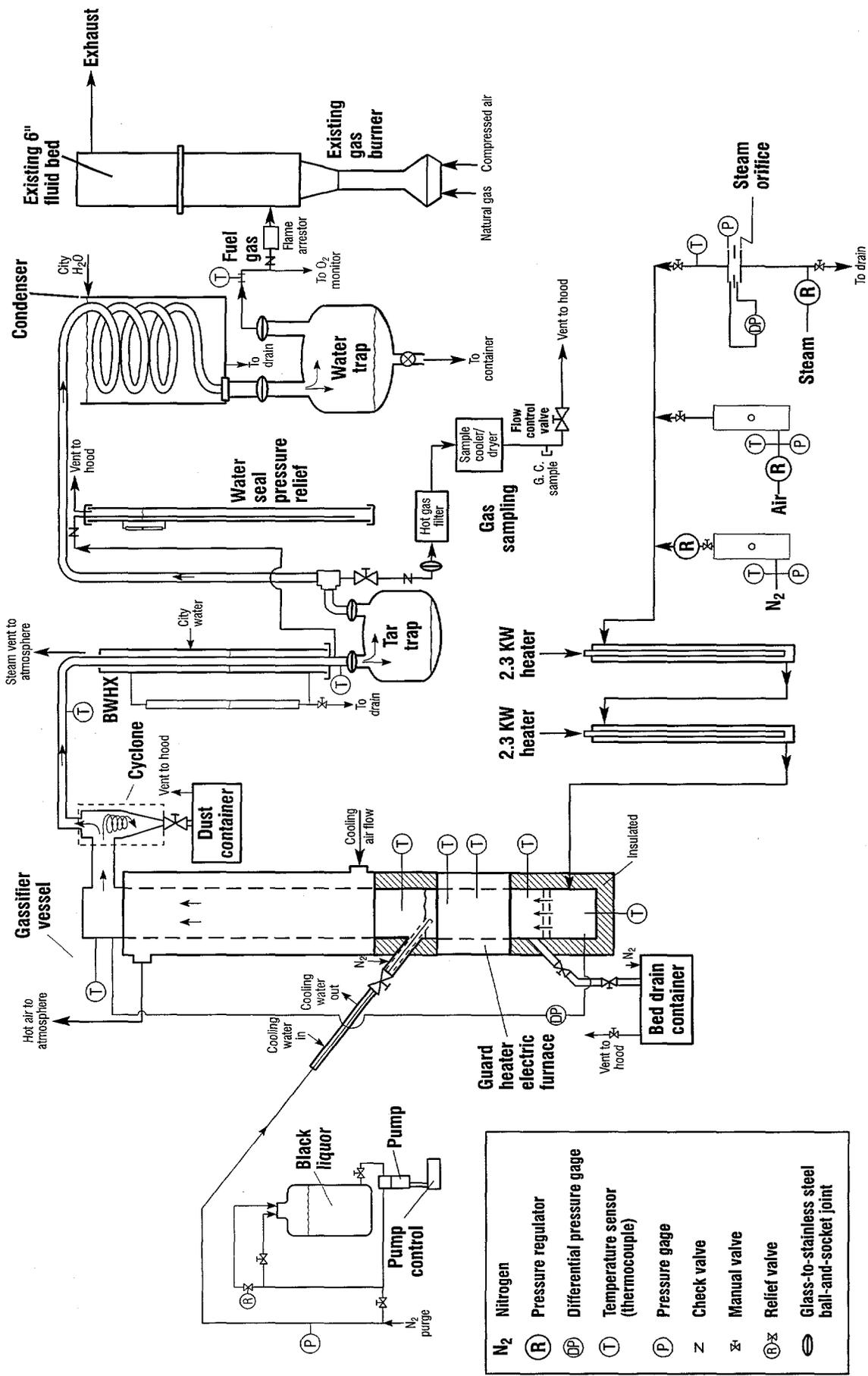
Safety considerations for the bench-scale test are necessary due to the toxic and flammable nature of the gas generated. Although it should be noted that the quantity of gas produced is small. The gasifier system is contained in an enclosure formed by flexible weld curtains vented via the laboratory fume hood system. The intended purpose of the enclosure is to mitigate dispersion of the product gas if there were a leak. A monitoring system to detect any leaks is also in place within the enclosure. All tubing and piping systems were fabricated to the extent feasible from materials that are acceptable under the power piping code. A check valve and flame arrestor were installed in the gas outlet tubing connecting to the fluidized bed combustor. These devices prevent a flame front traveling into the gasifier piping and components.

A.9 Instrumentation

Instrumentation used in the bench-scale facility is described previously in the text. A complete instrument list is included in Attachment A.3.

A.10 References

- A.1 Battock, W. V. And Pillai, K. K., "Particle Size in Pressurized Combustors," Proceedings of the Fifth International Conference on Fluid Bed Combustion, p. 642 (1977).



N ₂	Nitrogen
(R)	Pressure regulator
(DP)	Differential pressure gage
(T)	Temperature sensor (thermocouple)
(P)	Pressure gage
Z	Check valve
⊗	Manual valve
(R)⊗	Relief valve
⊕	Glass-to-stainless steel ball-and-socket joint

Figure A-1. Schematic of the Bench-Scale Black Liquor Gasifier.

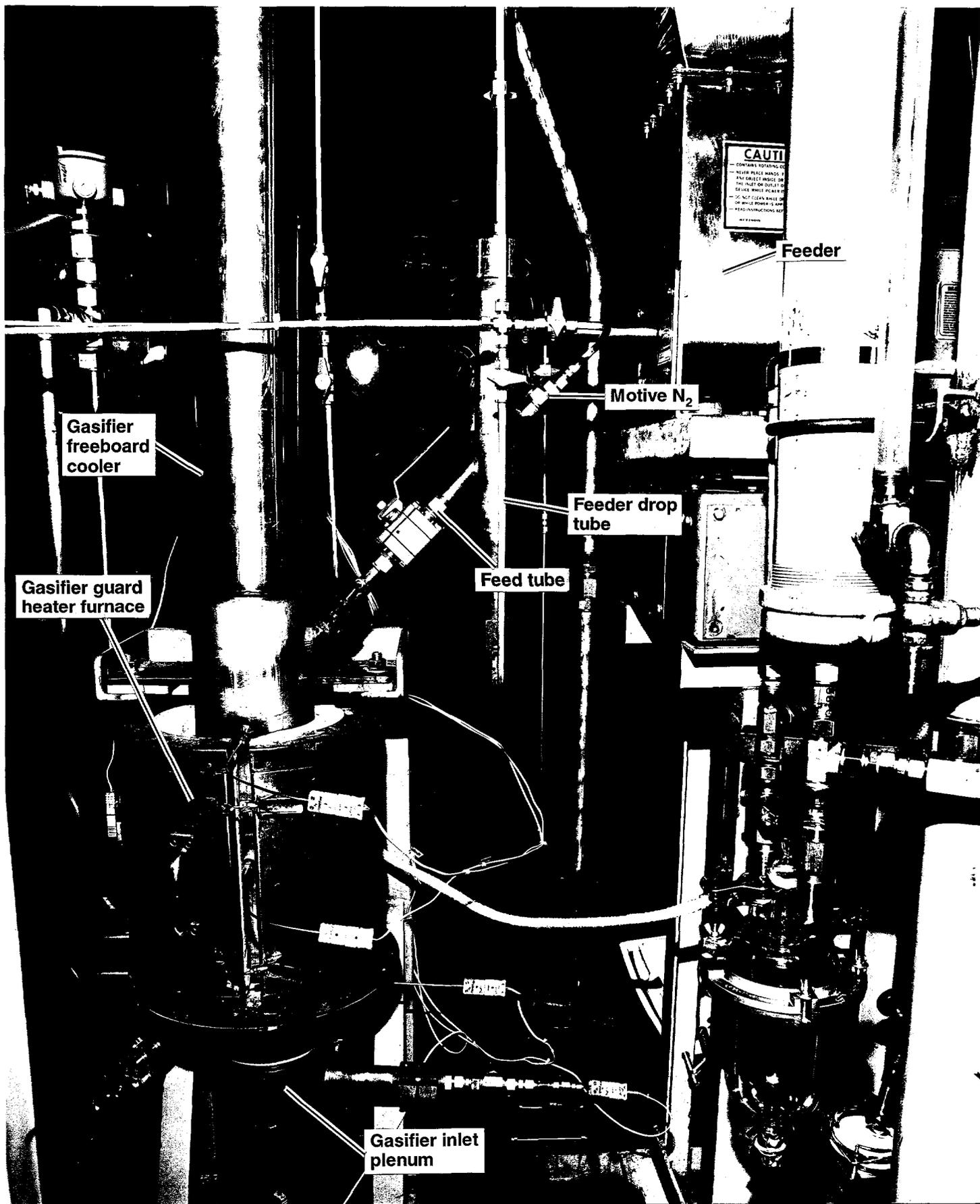


Figure A-2. Bench-Scale Black Liquor Gasifier and Solids Feed System.

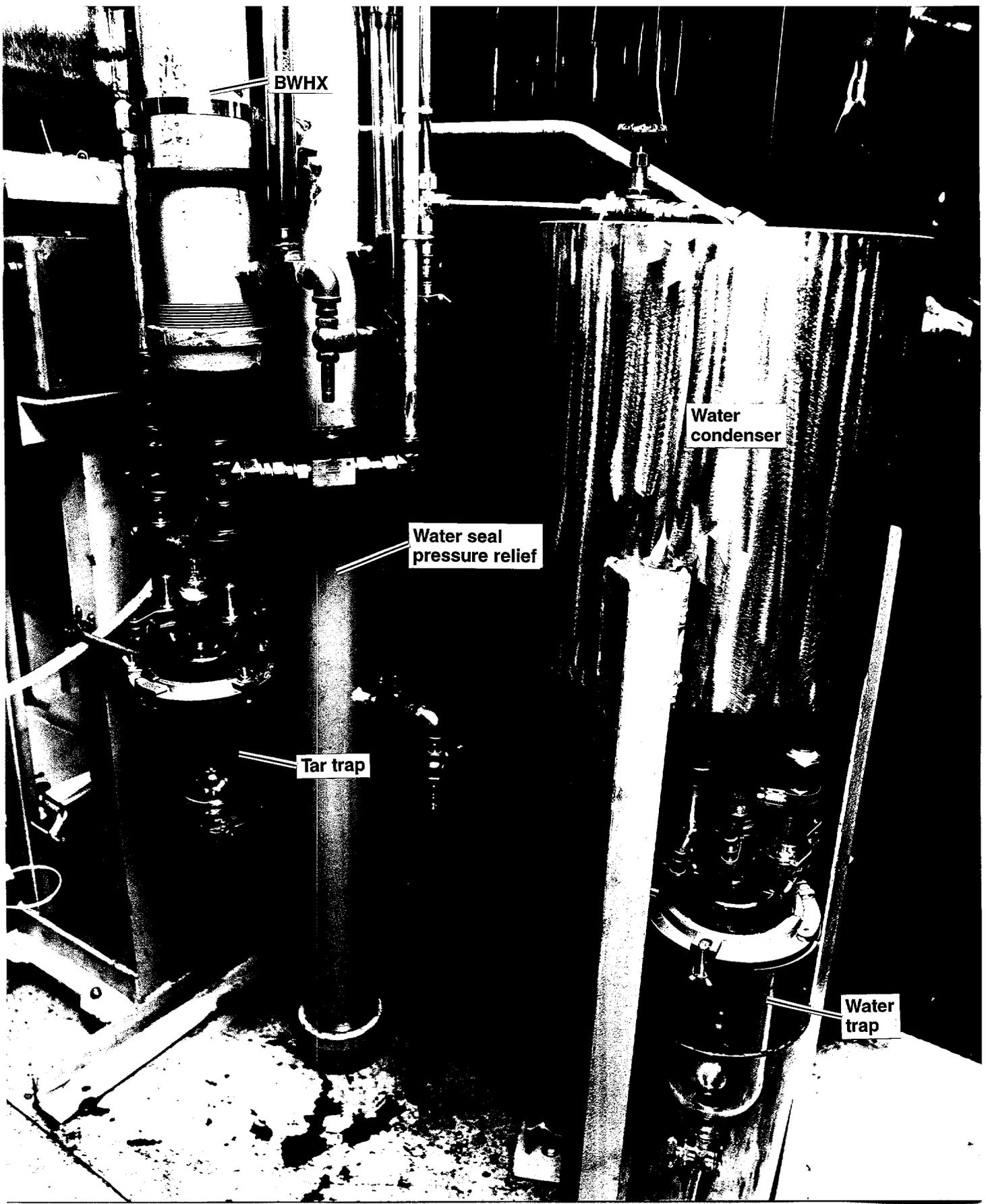
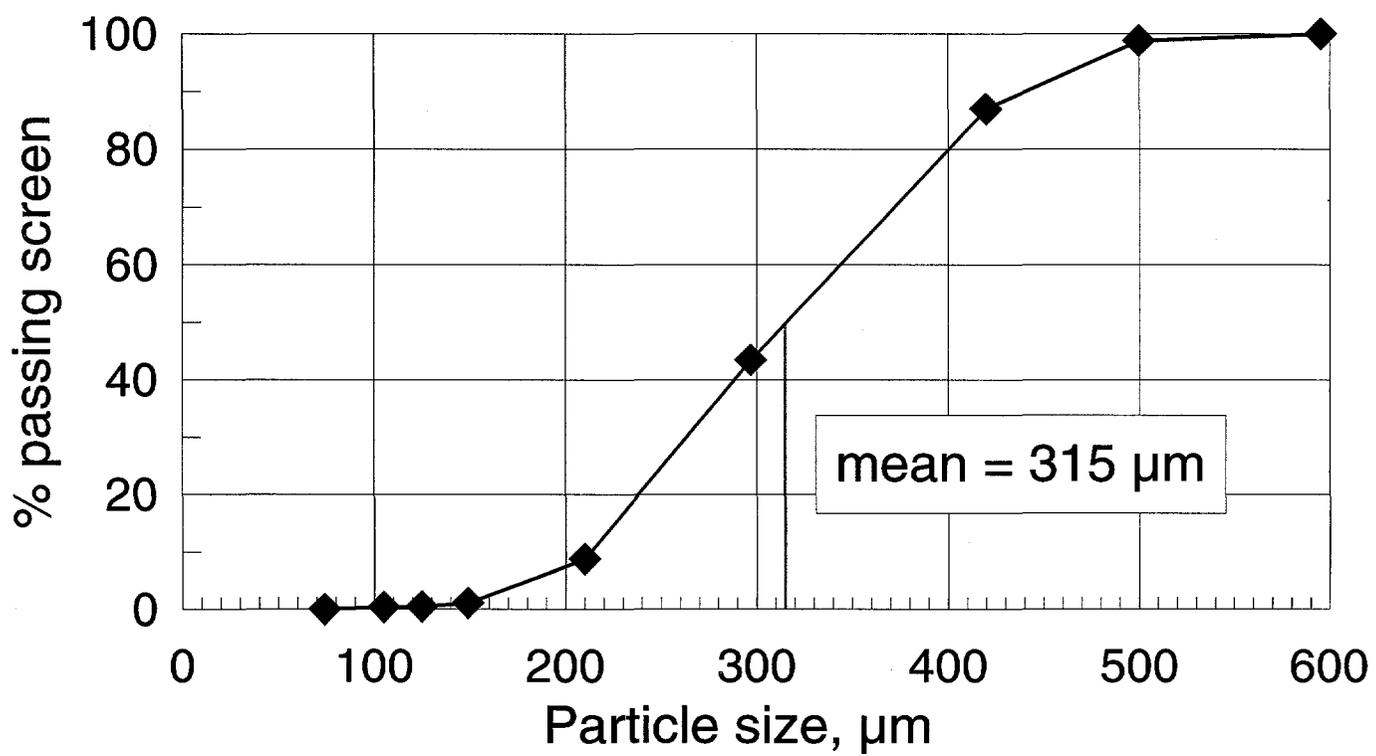


Figure A-3. Bench-Scale Gasifier - BWHX, Tar Trap, Condenser & Condensate Trap.

ATTACHMENT A.1

TYPICAL SIZE DISTRIBUTION OF SAND USED AS STARTING MATERIAL FOR THE
BENCH-SCALE BLACK LIQUOR GASIFIER TESTS

Initial Bed Material (Sand)



ATTACHMENT A.2

GC ANALYSIS METHODS USED FOR
BENCH-SCALE BLACK LIQUOR GASIFIER TESTS

GC Analysis Methods Used for Bench-Scale Black Liquor Gasifier Tests

Sampling Procedure

Gas samples were extracted from the process stream through a septum using 100 ml gas tight syringes. Each gas syringe was equipped with a Teflon valve and side hole needle. The Teflon valve allowed for sealing the sample in the syringe after it was drawn, and the side hole reduced plugging of the needle. Gas samples were taken to gas chromatographs, located nearby, and analyzed expeditiously. Part of each gas sample was injected into a first gas chromatograph designed to perform the sulfur species analysis, and the remainder of the gas sample was injected into a second gas chromatograph setup for the gas mixture analysis.

Sulfur Species Analysis

The GC used to quantify H₂S (<1500 ppmV), COS, CH₃SH, (CH₃)₂S, and CS₂ in each gas sample was a Varian Model 3400 equipped with a flame photometric detector maintained at 230°C (446°F). Nitrogen carrier gas at 30 ml/min flowed through a 30-inch x 1/8 inch OD Teflon (FEP) column packed with Porapak QS 80/100 mesh. Each gas sample filled a constant volume (0.075 ml) sample loop installed in the GC, and was injected onto the column. Linear temperature programming of the column was required from 45°C, 1 min. hold, to 210°C at 30°C/min, and a total analysis time of 10 minutes. Approximate retention time in minutes for each sulfur compound was:

Hydrogen sulfide	1.4
Carbonyl sulfide	2.0
Methyl mercaptan	3.6
Dimethyl sulfide	4.8
Carbon disulfide	6.8

Permanent Gases, C1 to C2 Hydrocarbons, and High Hydrogen Sulfide Analysis

The GC used to quantify H₂, N₂, CO, CH₄, CO₂, C₂H₂, C₂H₄, and high H₂S (>1500 ppmV) in each gas sample was a Hewlett-Packard Model 5890 Series II. It is equipped with a thermal conductivity detector maintained at 290°C. Helium carrier gas at 20 ml/min. flowed through a 30 ft. x 1/8 inch OD SS column packed with HayeSep DB 100/120 mesh maintained at 120°C. Each gas sample filled a constant volume (0.25 ml) sample loop installed in the GC, and was injected onto the column. Polarity of the detector was reversed to positive after elution of the hydrogen peak, and the total analysis time was twenty-one minutes. Approximate retention time in minutes for each compound was:

Hydrogen	3.8
Nitrogen	4.3
Carbon monoxide	4.7
Methane	6.1
Carbon dioxide	8.3
Acetylene	11.8
Ethylene	12.7
Hydrogen sulfide	19.6

Gas Chromatograph Calibration

Varian 3400

The GC was calibrated using N.I.S.T. traceable or primary bottled gas standards in the concentration ranges of 10 to 1500 ppmV for H₂S and 25 to 1100 ppmV for COS.

Certified gas standard mixtures commercially supplied by Matheson Gas Products were used to calibrate the GC in the concentration ranges of 100 to 5000 ppmV for both CH₃SH and (CH₃)₂S and 50 to 500 ppmV for CS₂.

Hewlett-Packard 5890

The GC was calibrated using N.I.S.T. traceable or primary bottled gas standards in the concentration ranges in volume percent:

Hydrogen	10 to 35
Nitrogen	37.3 to 57.5
Carbon monoxide	3 to 25
Methane	0.5 to 6
Carbon dioxide	1 to 20
Acetylene	0.1 to 0.3
Ethylene	1 to 3
Hydrogen sulfide	0.15 to 1.1

ATTACHMENT A.3

INSTRUMENT LIST FOR BENCH-SCALE BLACK LIQUOR GASIFIER TESTS

BENCH SCALE BLACK LIQUOR GASIFIER

INSTRUMENT LIST

Project No 43244-110

Revision No. 8

B&W tag or PO #	Manufacturer	Model No/Mfg No	Item	Description Function	Range
880102	Brooks Inst.	1307D08E1A	Rotameter	Air flow	0.43 - 3.93 SCFM
PO 86519			Type T T/C Elem	Air Temp	
PO 260302 (a)			Type T plug	B1 Air Temp	
650507 (e)	Marshall Town	200-35892	Press Gage	Air Press	0 - 5 psi
900895 (f)	Marshall Town	90640	Press Gage	Air Press	0 - 15 psi
950331	Fisher & Porter	10A455/7	Rotameter	Nitrogen Flow	0.20 - 2.5 SCFM
PO 90400UM			Type T T/C Elem	Nitrogen Temp	
PO 260302 (a)			Type T plug	B2 Nitrogen Temp	
760208	Heise	69037	Press Gage	Nitrogen Press	0 200 psig
850733	Fisher & Porter	10A1755S	Rotameter	Purge N2 Flow	
720054	ITT-Barton	289A-3232	Delta Press	Steam flow	0 - 15 inch H2O
960180	Vickery-Sims	0.750"	Orifice, flow	Steam flow	ID = 0.750 inch
PO 22178			Type K T/C Elem	Steam Temp	
PO 22178			Type K plug	Y6 Steam Temp	
960284 (c)	Omega Eng	PGL-25L160	Press Gage	Steam Press	0 - 160 inch H2O
			Mercury Manometer	Steam Press	0 - 30 "Hg
880063 (g)	Acrison	A401-10-20-105-BB (87561-01)	BLS feeder and control	BLS flow	0 - 100 % of scale
960427	Cole-Parmer		Pump	Liquor Feed Pump	0 - 10 setting
960285 (d)	Omega Eng	PGL-25L160	Press Gage	Gas @ heater inlet	0 - 160 inch H2O
700009	ITT-Barton	289A-3232	Delta Press	Bed Delta P	0 - 30 "H2O
950081	Vaisala	PTB100A	Electronic Bar.	Barometric Press.	24 - 31 "Hg
850734	Matheson	Model 7640 Tube No 603	Rotameter	Sample Gas flow	
PO 178089UD			Type K T/C Elem	Windbox Temp	
PO 22178 (b)			Type K plug	Y7	
PO 178089UD			Type K T/C Elem	In Bed Temp #1	
PO 75604UM (h)			Type K T/C Elem	In Bed Temp #1	
PO 12229 (b)			Type K plug	Y1	
PO 178089UD			Type K T/C Elem	In Bed Temp #2	
PO 12229 (b)			Type K plug	Y2	
PO 178089UD			Type K T/C Elem	In Bed Temp #3	
PO 12229 (b)			Type K plug	Y3	
PO 178089UD			Type K T/C Elem	In Bed Temp #4	
PO 22178 (b)			Type K plug	Y4 (Freeboard in)	
PO 178089UD			Type K T/C Elem	FB exit Temp	
PO 91786 (b)			Type K plug	Y5	
PO 90400UM			Type T T/C Elem	Ambiant	
PO 86819 (a)			Type T plug	B6	
PO 95132UM			Type T T/C Elem	Cyc Ex pipe temp	
PO 260302 (a)			Type T plug	B3	
PO 90400UM			Type T T/C Elem	BWHX exit temp	
PO 95132UM (a)			Type T plug	B4	
PO 260302			Type T T/C Elem	H2O Cond exit T	
PO 93604 (a)			Type T plug	B5	
920023	Strawberry Tree	5732	Type K	Temperatute readout	
850631	Omega	199	Type T	Temperatute readout	
860155	Fluke	3955060	Type K	Temperatute readout	
770481	Metler	P1200	Balance	Liq flow rate check	0 - 1200 grams
860418	Metler	P3600	Balance	Liq flow rate check	0 - 3600 grams
900214	Cole-Palmar		Stopwatch	Liq flow rate check	

Notes:

- (a) All Type T T/C extension wire = PO 73743UM (special limits); Marln No 251000
- (b) All Type K T/C extension wire = PO 17564; Marlin No 47453
- (c) Gauge failed & removed from service 7/9/96
- (d) Gauge switched from air line to heater inlet 7/10/96
- (e) Gauge switched 7/10/96; failed 7/20?
- (f) New gauge 7/22; replaced #650507
- (g) System replaced 8/7/96 by Cole-Parmer pump
- (h) T/C became erratic during preheat 9/17/96; replaced for 9/18 test

APPENDIX B

BENCH-SCALE TEST RESULTS

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BENCH-SCALE TEST RESULTS

This appendix provides a chronological description of the black liquor gasification tests that were conducted in the bench-scale equipment described in Appendix A. Major test data and calculated results are included in the attached tables.

Shakedown Testing (July 3-11, 1996)

The bench-scale equipment underwent a series of shakedown tests prior to beginning the black liquor gasification test program. These tests evaluated the performance of air, steam and nitrogen flow measurement and control devices, reactant gas mixture heaters, product gas cooling equipment, and the solids feeder (operated with sand). Gas leakage tests confirmed that the system was sealed from the fluid bed zone of the gasifier to the product gas combustor. Correlations were developed for black liquor solids feed rate as a function of screw feeder speed and for air distributor plate pressure drop as a function of total gas flow rate. The correlations were used in data reduction during the test program.

Test 1 (July 16, 1996)

The objective of this test was to start up the gasifier at an air/fuel ratio that would support combustion and then increase liquor flow to achieve gasification conditions. Attachment B.1 summarizes test data, calculated gas and liquor flows, windbox oxygen level, oxygen-to-liquor ratio, and superficial gas velocity in the fluidized bed zone.

The first set of data in Attachment B.1 was taken during preheating. Subsequent failure of the guard heater caused the average bed temperature to drop to 860°F. A liquor solids flow of about 1 lb/hr was initiated at 14:33. A rapid rise in bed temperature to 1150°F indicated that combustion of liquor solids was occurring. After 60 minutes of apparently stable operation, the liquor feed was increased to 2 lb/hr. The first sample of product gas was taken for GC analysis at

16:18 (see Attachment B.2); the results indicated that the oxygen content was only slightly lower than the calculated level in the windbox. Because bed temperatures remained steady at 1150-1200°F, it was decided to increase liquor flow to 4 lb/hr at 17:31; this was expected to produce gasification conditions. However, repeated GC analysis of the product gas indicated that essentially no fuel gas species were present and that O₂ content remained close to windbox levels. Liquor feed was terminated at 18:33.

A brown haze began to deposit on the inside of the glass vessel (tar trap) at the exit of the BWHX about 15 minutes after liquor feed was initiated. The deposit slowly darkened, but it was still possible to see through the vessel at the end of the test. Steam flow to the reactor began at 16:30, after which a small amount of oily liquid was present in the water samples drained from the condensate trap.

Twice during the test nitrogen flow was interrupted and temperatures measured at 18" and 24.5" in the bed exceeded 1350°F. After 17:00, the temperature of the lowest in-bed thermocouple dropped to below 1000°F and the differential bed pressure decreased to 5.5-6.5 in. water. These results indicated that some portion of the lowest 10 inches of the bed were defluidized. When the bed material was drained after the test, a few hard agglomerates of sand were observed. It was concluded that the inorganics in the liquor solids melted during the temperature excursions which resulting in a partial defluidization of the bed.

After the test, a blockage was discovered in the black liquor feed tube. Black liquor char plugged the feed tube at the inlet to the fluid bed reactor, the remainder of the feed system was packed with black liquor solids powder. It was concluded that the black liquor powder began to devolatilize and swell at the end of the feed tube, which was heated by conduction from the fluid bed reactor. Based on the quantity of packed solids removed from the feed system, it appears that very little material was actually introduced to the bed. Very little black char was seen in the free-flowing sand which was removed from the reactor after the test. Inspection by fiber optic camera revealed a few char deposits on the distributor plate.

Test 2 (July 23, 1996)

Before beginning the second test, a movable nitrogen lance was installed in the solids feed tube. The lance increased the flow of motive nitrogen to carry the solids into the reactor as well as cooling the feed tube and providing a means to break up blockages that might form during testing. An on-line analyzer was added to the apparatus to allow the oxygen content of the product gas to be monitored during a test. The computer that monitors the six thermocouples located in the windbox, bed zone and freeboard was also programmed to record these temperatures at one-minute intervals.

After air, nitrogen and steam flows were established and the bed zone was preheated to 1050°F, liquor solids flow of 1.65 lb/hr was initiated at 12:05. Test 2 data and calculated quantities are given in the last two columns of Attachment B.1. Gas chromatography analysis is tabulated in Attachment B.2. Temperatures recorded by the data logging computer during the test period are plotted in Figure B.1.

The modified solids feed system was an improvement, but swelling of liquor solids at the entrance to the reactor periodically blocked the flow and resulted in intermittent feed. Within 30 minutes, a sudden increase in differential pressure between the solids feed hopper and the freeboard indicated a char blockage had formed in the feed tube. Increasing the flow of nitrogen to the feed tube and "rodding" with the lance tube was able to clear the blockage. Following this action, the on-line analyzer indicated a rapid drop to near-zero oxygen level in the product gas. The sudden charge of liquor solids to the reactor was apparently enough to consume all the available oxygen in the reactant gas. After several minutes, the oxygen content increased to about 5% and the pressure again begin to rise in the feed system. This procedure was repeated five times during the 180-minute test. The liquor feeder was turned off three times to prevent solids from accumulating in the feed tube, these events are indicated on Figure B.1.

Approximately 20 feet of polyflow line separated the on-line oxygen analyzer from the product gas sample location. Due to the resulting time lag in instrument response, it was not possible to coordinate gas sampling for GC analysis with the brief periods of gasification that were observed. The results in Attachment B.2 indicate that the measured oxygen content of the product gas was 2.9-6.5%, considerably lower than the 10.2-10.9% calculated in the windbox (Attachment B.1). Levels of 1-2% hydrogen, 3-5% CO₂ and H₂S were also determined from GC analysis. The latter species was not quantified by separate sulfur gas analysis; however, a minimum concentration of 2000 ppm was present for H₂S to be detected by GC hydrocarbon analysis.

The thermocouple readings plotted in Figure B.1 show that the average bed temperature increased 100°F about 90 minutes after the first attempt to feed liquor. The average bed temperature oscillated between 1100 and 1200°F as the gasifier cycled between combustion and gasification conditions. The steady increase in temperature from 900 to 1300°F at the 24.5" location may have been due to combustion of volatiles immediately above the bed surface. The reason for the decrease in temperature at 24.5" after 120 minutes is not known.

The calculated superficial gas velocity (Attachment B.1) was about 50% lower than design (1.5 ft/s) which may have resulted in less "vigorous" fluidization. However, uniformity of the three in-bed thermocouple readings (Figure B.1) indicates that the bed material was fluidized throughout the test. The temperature drop at the 1.75" bed elevation after 150 minutes suggests that an incipient defluidization condition may have existed at the end of the test.

A limited amount of tar was detected as a brown haze on the glass tar trap and as oily patches and sticky deposits in the condensate samples.

About 50 g of fine, light char was collected in the cyclone, drain pipe and catch pot. The black liquor powder used in this test was screened to be between 149 and 1680 μm. Calculations indicate that particles of swollen char originating from liquor solids initially smaller than 355 μm

could have been entrained at the superficial gas velocity in the bed zone of the reactor. A portion of these char particles were likely carried over into the cyclone.

Post-test inspection of the equipment revealed a partial char blockage of the feed tube; a hole the size of the ¼ in. nitrogen lance tube was the only opening to the bed. The fiber optic camera was used to inspect the inside of the reactor. A large, friable char mass was resting on the bed surface and a few small char deposits were seen on the reactor walls. There were numerous, distinct 1-5 mm char particles among the sand drained from the reactor. The char deposits were broken up and washed out of the reactor.

When batches of black liquor solids entered the reactor after a blockage was broken up, part of the solids may have devolatilized and swollen together as a large mass. Because gasification kinetics are slow at 1100°F and the gas flow was not high enough for optimum fluidization, it is not surprising that unreacted char remained at the end of the test. Increasing gas velocity to design conditions may improve char burn out by increasing mixing in the bed, but would also increase elutriation of fine black liquor solids and char particles.

Results from this test suggested that black liquor gasification was possible in the bench-scale equipment. The main obstacle impeding progress was sustaining material delivery to the bed. The majority of the dried solids obtained from Resource Recovery Inc. were too fine for use in the equipment. A small amount of useable material was obtained from the powder, most of which was consumed in the first two tests. As discussed in Appendix C, several vendors failed to process this material into a useable form. Accordingly, use of dried liquor solids was discontinued after Test 2 and effort was focused on designing a system to deliver industrial black liquor at 50-60% solids directly into the gasifier.

Test 3 (August 8, 1996)

Preliminary tests with an existing gear pump and available liquor samples demonstrated that transport and injection of 50% solids liquor at room temperature was possible. Two drums of 55% liquor were obtained from a nearby kraft paper mill. A simple, water cooled injection system was fabricated to deliver the liquor through the existing ball valve and coupling. The liquor flowed in a thin stream from the nozzle tip directly into the fluidized bed.

Previous test results indicated that the air heaters alone were only capable of producing windbox temperatures of 900-950°F. It was decided that the desired bed temperature of 1250°F could be reached by increasing heat input from the guard heater and partial combustion of the liquor solids in the fluidized bed.

The third test run demonstrated that uninterrupted delivery of 55% solids black liquor to the reactor was possible and that gasification conditions were achieved. The test was initiated under combustion conditions with a liquor flow of approximately 2 lb solids/hr (Attachment B.3). The thermocouple plots in Figure B.2 show that the average bed temperature increased by about 100°F in the first 30 minutes of liquor feed. The liquor flow was then increased to 4.2 lb solids/hr to attain gasification conditions. The on-line analyzer indicated zero-oxygen conditions were achieved within ten minutes of the liquor flow increase. The product gas sampled at 13:21 and 13:23 contained substantial amounts of fuel components and hydrogen sulfide (Attachment B.2). In Table B.1, the average product gas analysis from the two samples is corrected to commercial conditions by subtracting the excess nitrogen required for fluidization and purge flows in the bench-scale equipment. The undiluted heating value of this gas, calculated from the corrected analysis, is 69 Btu/DSCF. Although the product gas heating value is less than the design target of 100 Btu/DSCF, the results of this test suggest fuel gas can be produced by low-temperature black liquor gasification.

Table B.1. Corrected Analysis of Product Gas from Test 3.

Gas Species	Avg. GC Analysis, % by volume	Corrected Analysis, % by volume
H ₂	7.9	14.1
N ₂	78.0	61.1
CO	1.7	3.1
CH ₄	0.7	1.2
CO ₂	11.2	20.0
C ₂ H ₄	0.1	0.1
H ₂ S	0.2	0.4
Total	99.8	100.0

After about 45 minutes of liquor feed, the temperature at 1.75" above the distributor plate rapidly decreased and leveled out at the windbox temperature. At 14:00 hrs, the temperature at 18" began dropping. Corresponding decreases in pressure drop across the fluidized bed (Attachment B.3) indicate loss of fluidization had occurred in parts of the reactor. The presence of oxygen in the GC samples taken at 14:05 and 14:07, as well as the decrease in fuel gas and reduced sulfur species was a result of channeling in the partially-fluidized bed. Liquor flow was increased at 14:10 and a constant oxygen to black liquor solids ratio was maintained by increasing air flow and decreasing nitrogen accordingly. However, partial combustion of the liquor was not able to balance the increased heat demand for evaporation in the poorly fluidized bed. The GC data in Attachment B.2 show that 13.1% oxygen content of the product gas sampled at the end of the run was close to the windbox level (16.3%). The presence of 680 ppm of reduced sulfur species indicates that some liquor pyrolysis (thermal decomposition) was still occurring.

During Test 3, the quantity of tar formed was small relative to the 20-25 lb. of liquor processed. As in the other tests, a thin haze of brown aerosol collected on the inside of the glass tar trap at the exit of the BWXH (gas temperature 188-193°F); no accumulation of liquid was noted in the tar trap. Post-test inspection revealed that neither the cyclone nor the BWHX pipe were fouled by tar. Thus, the tar condensation temperature is sufficiently low that fouling of the cyclone and heat transfer surface may not be a concern for commercial equipment.

The majority of the organics co-condensed with water at approximately 80°F in the condenser coil. All condensates collected had patches of oil floating on the surface as well as tarry deposits adhering to the glass sample jars. The creamy, opaque appearance of some samples suggests the tendency of the condensate to form stable oil-in-water emulsions (Shaw, 1980). The condensates had a powerful aromatic odor with traces of noxious sulfur compounds. Although no chemical analysis was conducted, these observations and literature information suggest that the organic liquids condensed at room temperature are a mixture of non-volatile tars and low molecular weight aromatics with small amounts of wood alcohols and carboxylic acids.

The check valve and inlet of the flame arrestor (located downstream of the condensate trap) were heavily fouled with a black tarry substance. Because these components were not inspected until after Test 3, some of the material collected may have deposited during the previous tests. As the gas temperature decreased in the heat transfer equipment, condensing water and tar vapors would have formed an aerosol mist. Due to the lack of impingement surface in the condensate trap, the finest mist particles may have been carried through without collection. The check valve and flame arrestor provide a tortuous gas path and extended surface area to facilitate aerosol collection by impingement and impaction. The tars were insoluble in hot water but were easily removed by flushing the hardware with acetone.

After the test, more than 100 g of fine char was removed from the cyclone drain line and catch pot. It is likely that most of this char was deposited during Test 2. Just prior to beginning liquor feed in Test 3, a blockage in the cyclone drain line was discovered. This may have

resulted from excessive carryover of fine char particles in Test 2. High-pressure nitrogen was used to blow out the line, discharging the char into the cyclone catch pot. Any fines elutriated during Test 3 would have subsequently mixed with and been indistinguishable from this material. In all tests the cyclone catch pot accumulated 1-2 quarts of liquid, this results from condensation of water from the slip stream of gas that is vented through the cyclone drain line and catch pot. The liquid was decanted into glass jars. Upon standing for several hours, a thick layer of black solids settled to the bottom of the jars, the liquid was translucent brown with a thin layer of char floating on the surface. There was little evidence of the tarry deposits, floating oils, emulsions or strong aromatic odors that characterized the liquid samples taken from the condensate trap.

A substantial amount of char remained in the bed at the end of Test 3. Numerous char particles, less than 1 mm in size, were mixed with the sand drained from the reactor. Larger agglomerates of char and sand (one-half to one inch in mean diameter) were also removed and may have contributed to the loss of fluidization by settling on the air distributor plate.

Test 4 (August 20, 1996)

During the early tests, it was thought that the initial period of combustion may have been initiating loss of fluidization. Therefore, the objective of Test 4 was to start liquor flow at a high enough rate to attain gasification conditions directly. Additional nitrogen and steam flow were employed to increase superficial gas velocity and improve fluidization in the bed. Other goals of this test were to drain bed material during testing, and operate the BWHX at more realistic minimum temperatures to simulate commercial operation.

With the bed preheated to 1000°F, liquor flow of 3.8 lb solids/hr was initiated at 12:10 hrs (Attachment B.4). At 12:30 the on-line analyzer indicated less than 1% oxygen in the product gas. With bed temperatures stable and bed pressure drop unchanged, the flow was increased to 5.0 lb/hr at 12:35. This change established zero-oxygen conditions. A sudden drop in temperature at 18" elevation (Figure B.3) suggested wet liquor was depositing on the

thermocouple, the flow was stopped at 13:10 hrs and mechanical rapping on the equipment was able to dislodge the deposit and temperatures quickly returned to acceptable levels. A second drop in temperature at the 18" location was followed by a 1500°F peak in temperature in the center of the bed. This may have resulted from combustion of volatiles in the vicinity of this thermocouple when the liquor flow was interrupted a second time. Following this excursion the temperature at the lowest in-bed elevation decreased and remained at the windbox temperature (900°F) for the duration of the test. Deposition on the 18" thermocouple was again indicated in the fourth hour of the test and was successfully cleared by mechanical rapping.

In the second hour of the test, an increase in differential pressure from the windbox to the freeboard indicated the bed was building. A partial bed drain was conducted at 14:12 hrs.; this action reduced the differential pressure from 24 in. to 16.5 in. of water. Suddenly differential pressure increased and the temperature above the liquor injector (24.5") began to drop; a second bed drain was attempted at 14:40 hrs. The test was ended after two more hours of intermittent liquor feed. Oxygen content in the product gas varied between 2 and 7% during the period. Approximately 25 lb of black liquor was injected in the reactor during this time.

GC results (Attachment B.2) show that significant amounts of reduced sulfur compounds were released during the first 60 minutes of testing. It is notable that the concentrations of methyl mercaptan (2203 ppm), dimethyl sulfide (614-677 ppm) and carbon disulfide (560-1803 ppm) are higher than in the gasification period of Test 3. These species, along with H₂S, are principal sulfur compounds released during black liquor pyrolysis (Strohbeen and Grace, 1982). Brink, et al. (1970) demonstrated that mercaptans and disulfides decompose to form H₂S by reaction with H₂ and water vapor; the conversion increases with temperature from 932-1832°F. The higher temperature and H₂ content of the gas produced in Test 3 explains the higher yield of H₂S relative to the heavier reduced sulfur species. Following similar reasoning, any high molecular weight hydrocarbons released during pyrolysis would not have been reformed to CH₄, H₂ and CO at the conditions of Test 4. This could explain the low levels of these fuel gas species and suggests that the unaccounted 3% of the gas analysis was higher molecular weight

compounds containing more than 3 carbon atoms.

Operating the BWHX at 230-254°F assured that no water condensed in the tar trap. The material collected during the test was limited to a thin coating the inside surface of the glass vessel. No accumulation of liquid was observed in the bottom of the tar trap. As in other tests, tars and oily liquids were observed in the samples collected in the trap downstream of the condenser. The total amount of tars formed during this test was very small compared with the 20-25 lbs. of black liquor processed. A check of the flame arrestor at the end of the test revealed much less tar deposition than was noted at the end of Test 3.

The bed was partially drained several times during operation. The bed material was predominantly black in color and contained a mixture of char, residual sand, and very limited amounts of consolidated inorganic material. It was freely flowing under the test conditions. The windbox and distributor plate were removed after this test to inspect the inside of the reactor. A significant number of black lumps were resting on the distributor plate (Figure B.4), these appeared to be a mixture unreacted char and sand. The distributor plate and was fouled with similar material (Figure B.5). The entire cross-section of the reactor above the liquor feed opening was plugged with char (Figure B.6). This material extended well into the freeboard region and contained two distinct layers of deposits, the lower was soft and friable char and the upper was hard, dense char. Clearly an insufficient quantity of material was drained from the bed to balance the build up of char in the reactor. Pluggage of the freeboard was responsible for the unrecoverable increase in differential pressure near the end of the test.

At the feed rate of 5.0 lb solids/hr the heat from partial liquor combustion apparently balanced the heat demand for evaporation and the average bed temperature did not rise above the preheat condition (1000°F). The GC analysis suggests that the test was partially conducted under pyrolysis conditions, but that the rate of gasification at 1000°F was not high enough to consume the char. The theoretical effect of temperature on black liquor gasification rate is illustrated in Figure B.7 for atmospheric pressure steam gasification of black liquor char. An activation

energy of 210 kJ/mol was determined by Li and van Heiningen (1991) for the temperature range 1112-1292°F, i.e., in the low-temperature gasification regime. It is not possible to predict actual carbon gasification rate without defining the pre-exponential factor and several constants for gas composition in the kinetic rate expression. This simple result only indicates relative increase or decrease in gasification rate as the temperature departs from the design value of 1250°F. Figure B.7 clearly shows that at 1000°F the theoretical gasification rate is 0.01 times the rate at 1250°F. This is likely the cause of excessive char build up in the reactor during Test 4.

Heater Performance Test (September 5, 1996)

Based on the results of Test 4, it was deemed important to operate the gasifier at as high a temperature as possible without exceeding the melting point of the inorganics. Prior tests had established that the cartridge heaters were only capable of achieving a windbox gas temperature of 900-1000°F. The objective of this test was to determine if a bed temperature of 1200-1250°F could be achieved by intensive heating with the guard heater furnace.

The system was preheated with maximum flows of air and steam (Attachment B.5) and the guard heater operating at the nominal value from prior tests (1170°F). Once stable operation was achieved, the guard heater set point was increased to 1450°F at 13:05 hrs. The thermocouple measurements plotted in Figure B.8 show that temperature at the 11" and 18" elevations reached 1200°F within 60 minutes. Note that the lowest in-bed temperature (1.75") increased only slightly above the windbox temperature during this period.

It was a serendipitous oversight that sand was not loaded into the reactor prior to this test. This fact became apparent from the low differential pressure measured between the windbox and freeboard (Attachment B.5). Gas flow was interrupted at 14:00 hrs and 15 lb. of sand was added through the liquor feed port. When gas flow was resumed, bed temperatures quickly increased. With sand in the bed, all three temperatures were now equal, as would be expected for a well-fluidized bed. The results of this test are important for three reasons:

- The temperatures recorded after the sand was added verify that the bed was fluidized at a superficial velocity of 1.3-1.5 ft/s.
- Operation without sand revealed that temperature at the base of the bed was significantly lower than in the zone surrounded by the guard heater.
- The test demonstrated that the bed could be preheated to 1200°F by intensive external heating.

Test 5 (September 18, 1996)

The objective of this test was to overcome the low temperature limitation of Test 4 by preheating the fluidized bed to approximately 1200°F. In order to increase windbox temperature, heating tapes were installed upstream of the gas mixture heaters. These proved to be ineffectual due to excessive heat loss from the exit of the second heater to the windbox inlet. The fluidized bed was preheated to an average temperature of 1217°F, by operating the guard heating furnace at 1500°F (Attachment B.6).

A liquor flow of 2.3 lb solids/hr was selected to reduce evaporation heat demand and initiate partial combustion in the fluidized bed. The air flow was set to match the 0.38 oxygen-to-liquor ratio which produced an acceptable product gas in Test 3. Nitrogen dilution was used to increase gas velocity in the bed and ensure good fluidization. Figure B.9 shows that within minutes of starting liquor feed the temperature at the 1.75" elevation began dropping. Partial combustion and gasification of the liquor was expected to reduce the oxygen level in the product gas to zero. However, the on-line analyzer did not indicate an oxygen level in the product gas stream below 2.94% (dry volume basis), the estimated level in the windbox at this condition was 5.12%. After 60 minutes the liquor flow was interrupted and recovery of bed temperature at 11" and 18" was noted. This may have resulted from combustion of char and volatiles at the increased air ratio. A second attempt at starting liquor flow at a higher air ratio was quickly

abandoned when an irrecoverable decrease in differential pressure between the windbox and the freeboard suggested that the bed had become defluidized.

Gasification conditions were not achieved in Test 5; however, the results confirmed that performance of the equipment was limited by low windbox temperature and low temperature surfaces located below the guard heater. These conditions cause the bed region located below the guard heater to be quickly quenched by the significant amount of moisture introduced with the liquor. The temperatures are high enough in this region to dry and partly pyrolyze the liquor which results in char-sand agglomerates forming on top of the distributor plate. As demonstrated in Tests 3 and 4, temperatures of 850-1000°F are insufficient to gasify the char carbon, and the accumulating agglomerates cause partial loss of fluidization and channeling in the bed.

Assuming equilibrium conditions are established in a well-mixed reactor and that the system is perfectly insulated, then the adiabatic bed temperature is a function of liquor heating value, water content and air-to-solids ratio (Backman and Hupa, 1990). Figure B.10 shows that an air ratio of 0.42 would be required to maintain the desired gasification temperature of 1250°F (677°C) at 55% solids. With dried solids, the same condition could be achieved at an air ratio of 0.26. The industrial liquor used in testing had a 15% lower heating value than the hypothetical liquor in Figure B.10 and the conditions in the bench-scale equipment were not adiabatic; however, this example illustrates the advantage of increasing liquor solids. By adding spray-dried black liquor powder to the industrial liquor sample the solids content could be increased to the maximum pumpable level (60-65%) which would substantially reduce evaporative heat demand in future tests.

The results of this test program demonstrate that gasification is feasible in the bench-scale equipment. However, certain improvements are required to maintain gasification conditions for an extended period of time. Reducing the penetration distance of thermocouples into the reactor (currently 1/3 of vessel diameter) would provide less initiation surface for deposits to form. Shortening the unheated length of pipe from the gas mixture heaters to the windbox and

wrapping heating tape around the reactor wall (below the existing guard heater) will allow a minimum temperature of 1200°F to be maintained in the fluidized bed zone. Increasing liquor solids to 60-65% would substantially reduce evaporative heat demand and increase adiabatic temperature at a given air-to-solids ratio. These improvements would prevent wet liquor from accumulating in cool regions of the reactor, forming char deposits and ultimately defluidizing the bed.

References

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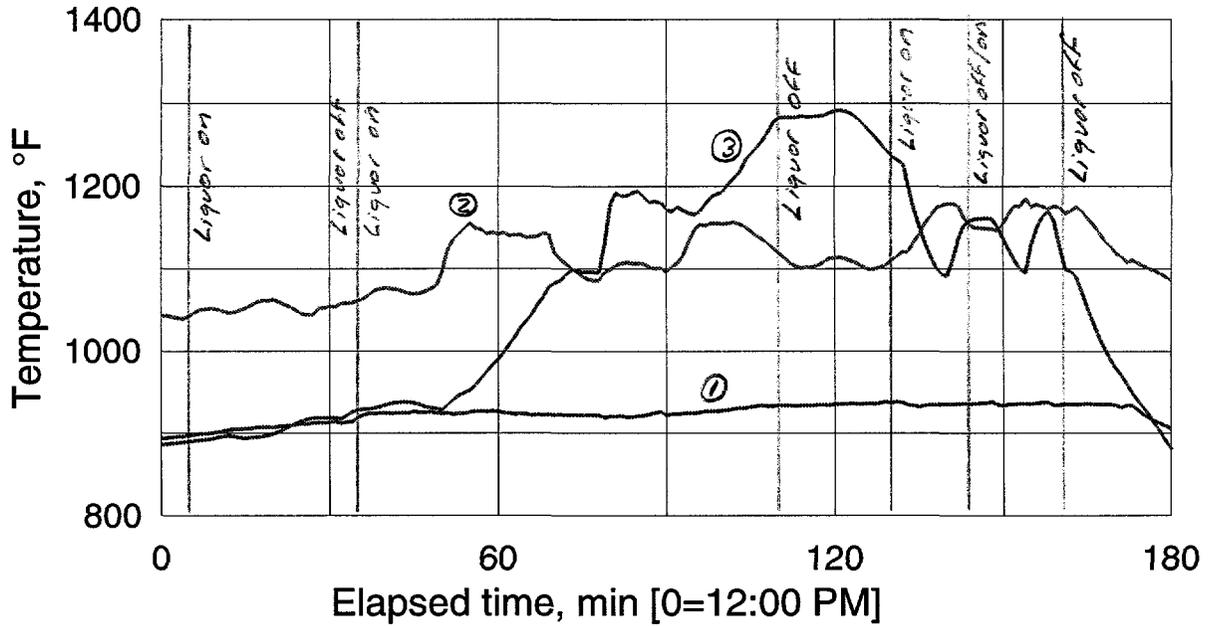
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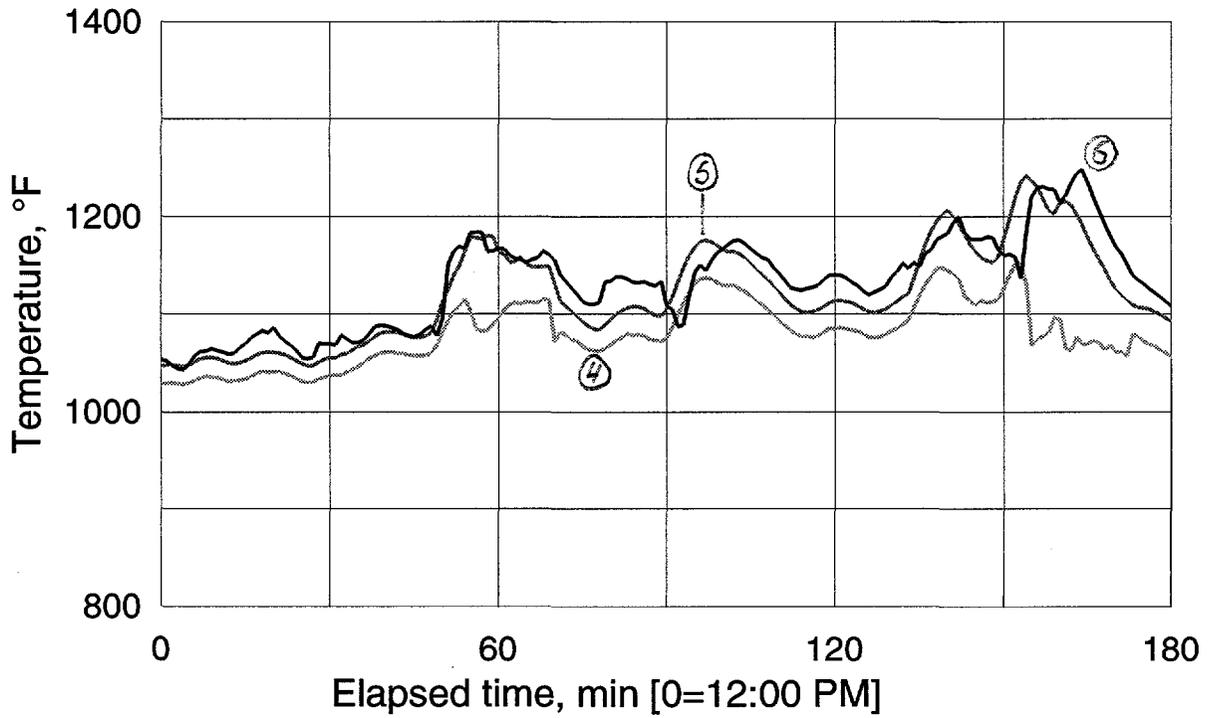
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Figure B.1 - Test 2 (7/23/96)



T/C location

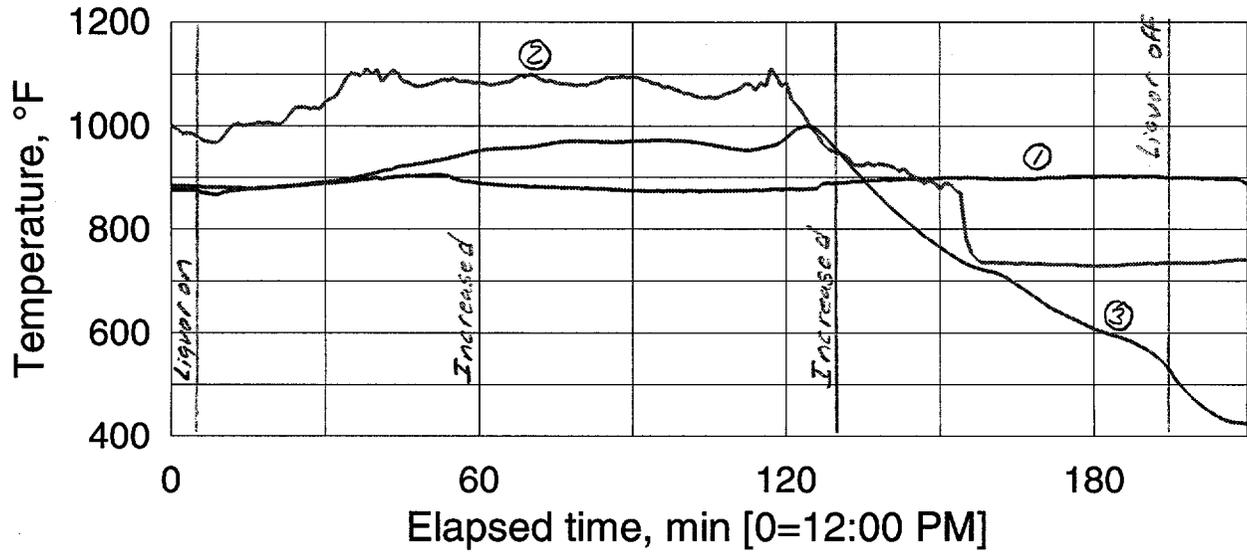
① Windbox ② Bed avg. ③ 24.5"



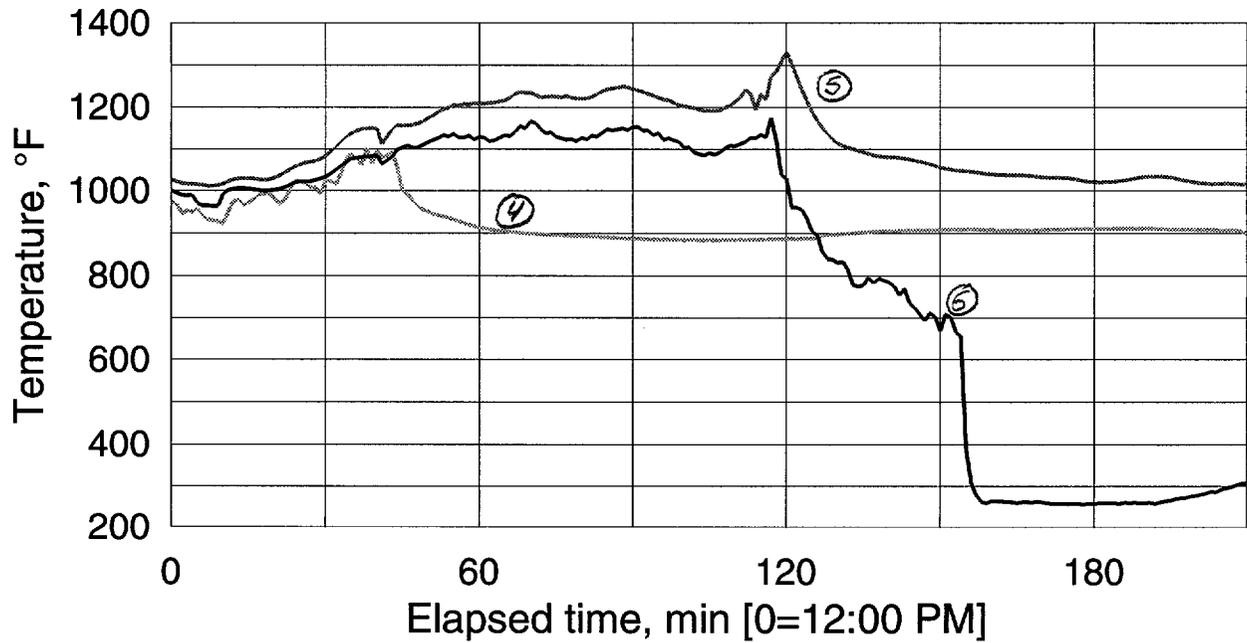
in-bed T/C location

④ 1.75" ⑤ 11" ⑥ 18"

Figure B.2 - Test 3 (8/8/96)

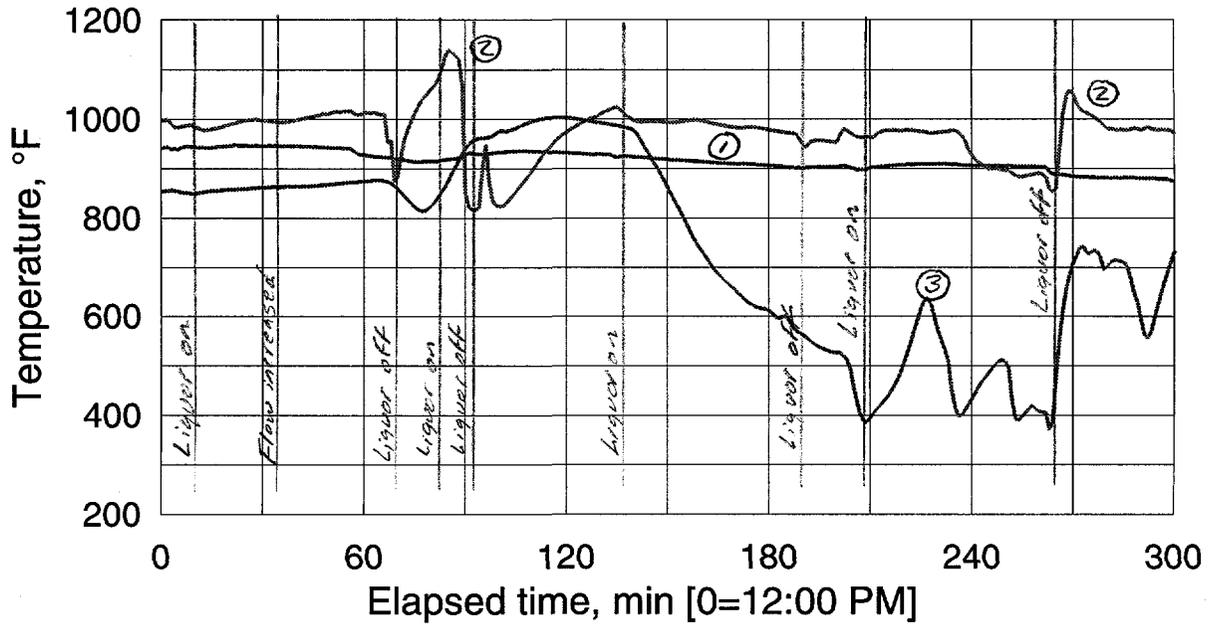


T/C location ① Windbox ② Bed avg. ③ 24.5"

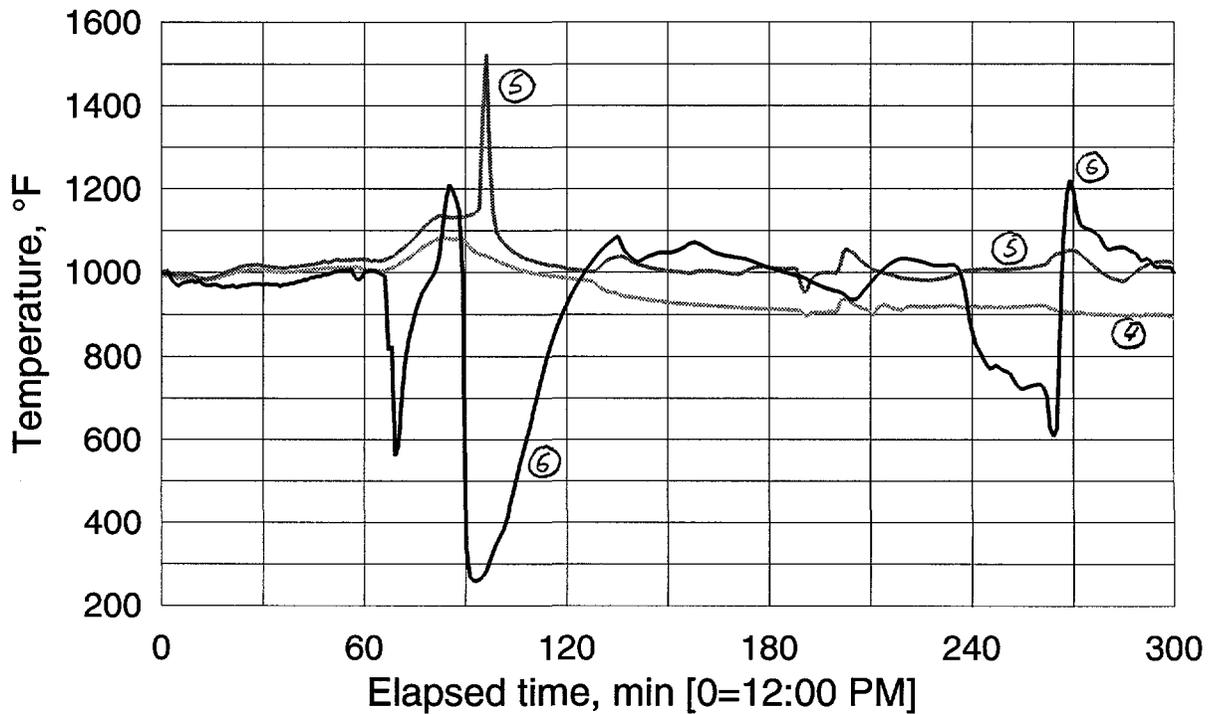


in-bed T/C location ④ 1.75" ⑤ 11" ⑥ 18"

Figure B.3 - Test 4 (8/20/96)



T/C location ① Windbox ② Bed avg. ③ 24.5"

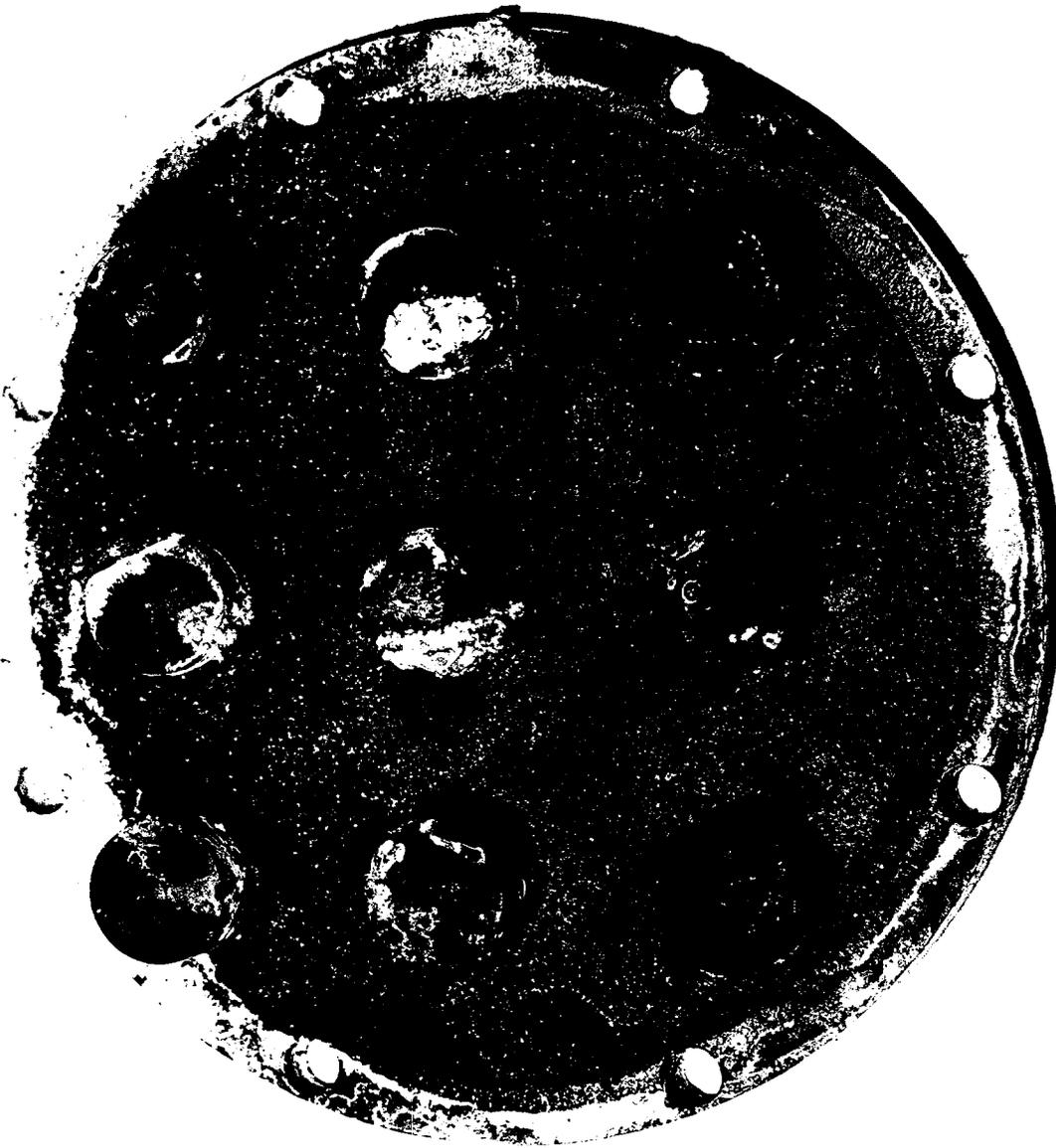


in-bed T/C location ④ 1.75" ⑤ 11" ⑥ 18"



MATERIAL FROM TOP
OF DISTRIBUTOR PLATE
REQ. # 45244 8-22-96

Figure B.4. Char/sand agglomerates removed from top of distributor plate on 8/22/96.



DISTRIBUTOR PLATE
AFTER WATER WASH
Proj. # 43244 8/22/96

Figure B.5. Distributor plate with char/sand deposits, removed for inspection 8/22/96.

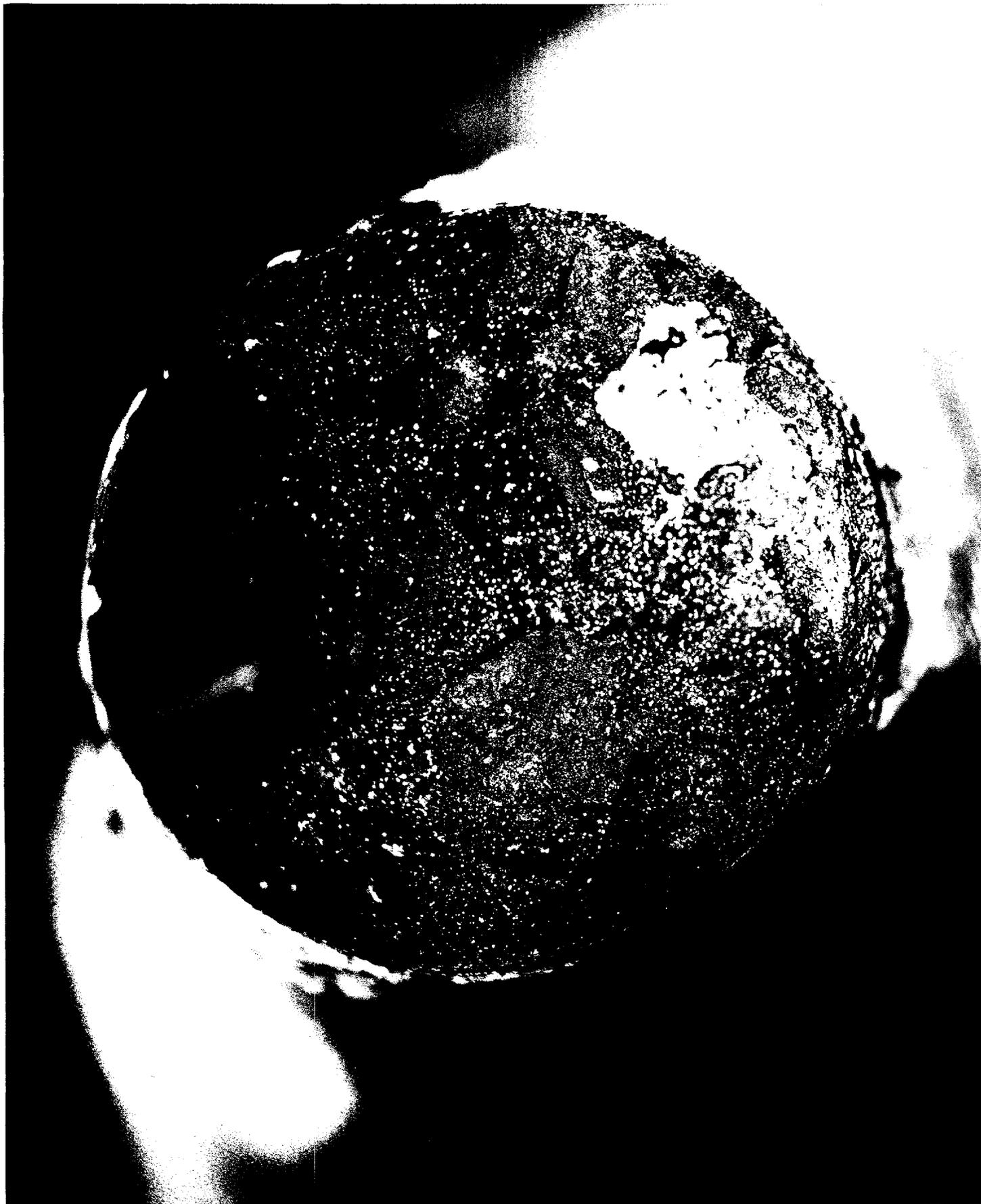


Figure B.6. Reactor view from below, showing complete blockage of cross-section above liquor injection elevation (inspection 8/22/96).

Figure B.7 - Steam Gasification Rate

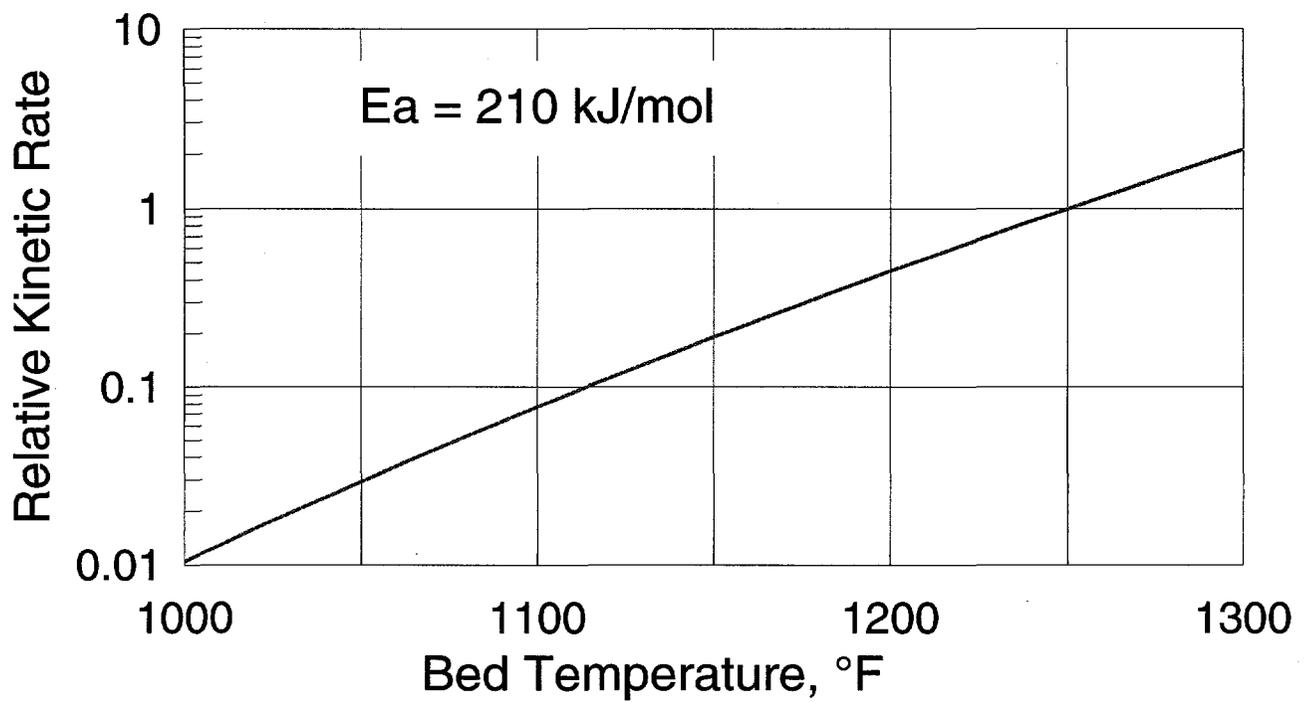
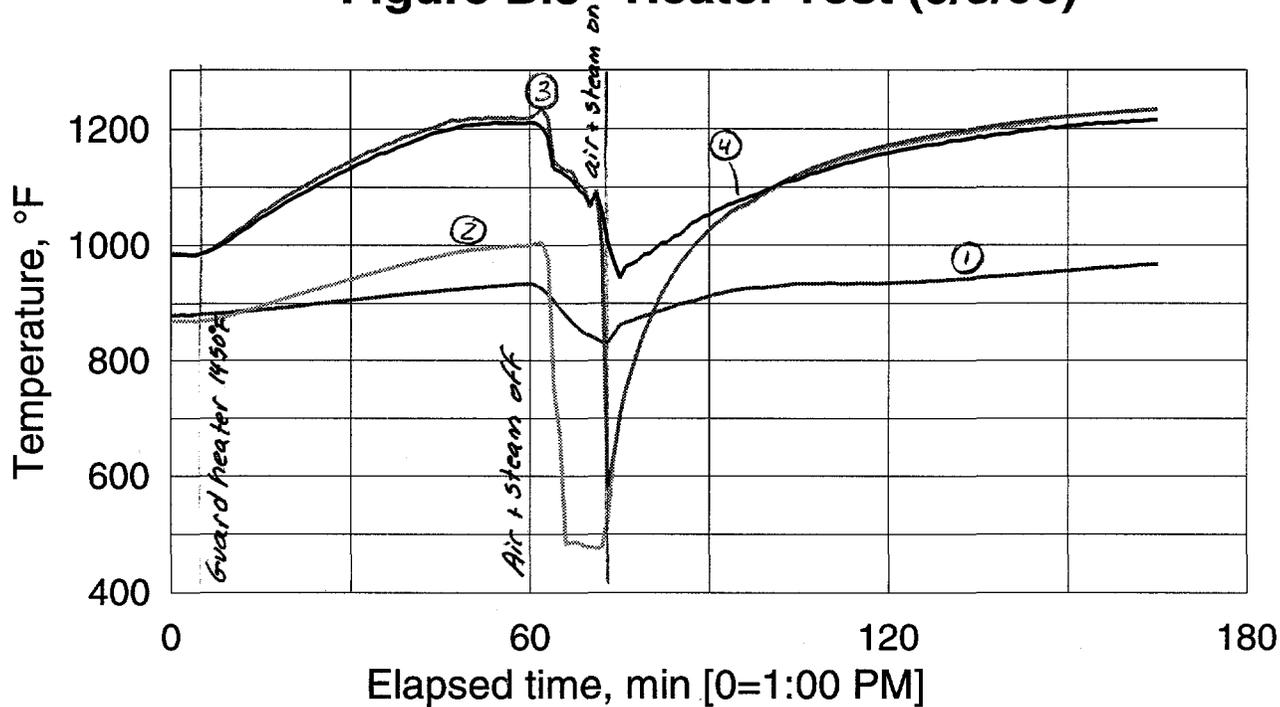
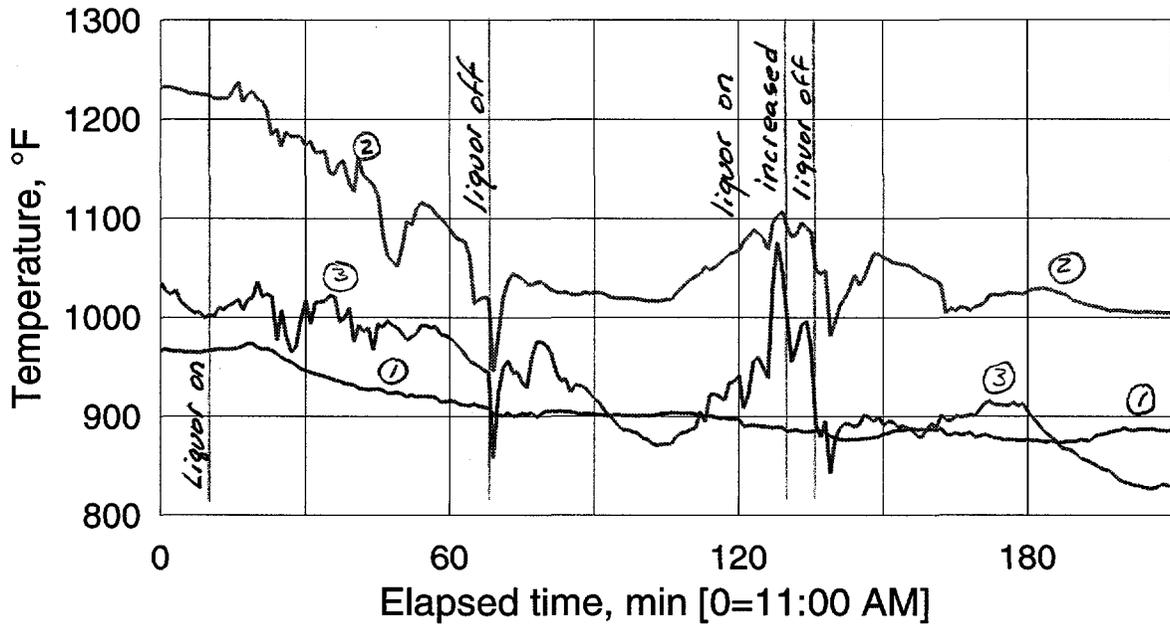


Figure B.8 - Heater Test (9/5/96)

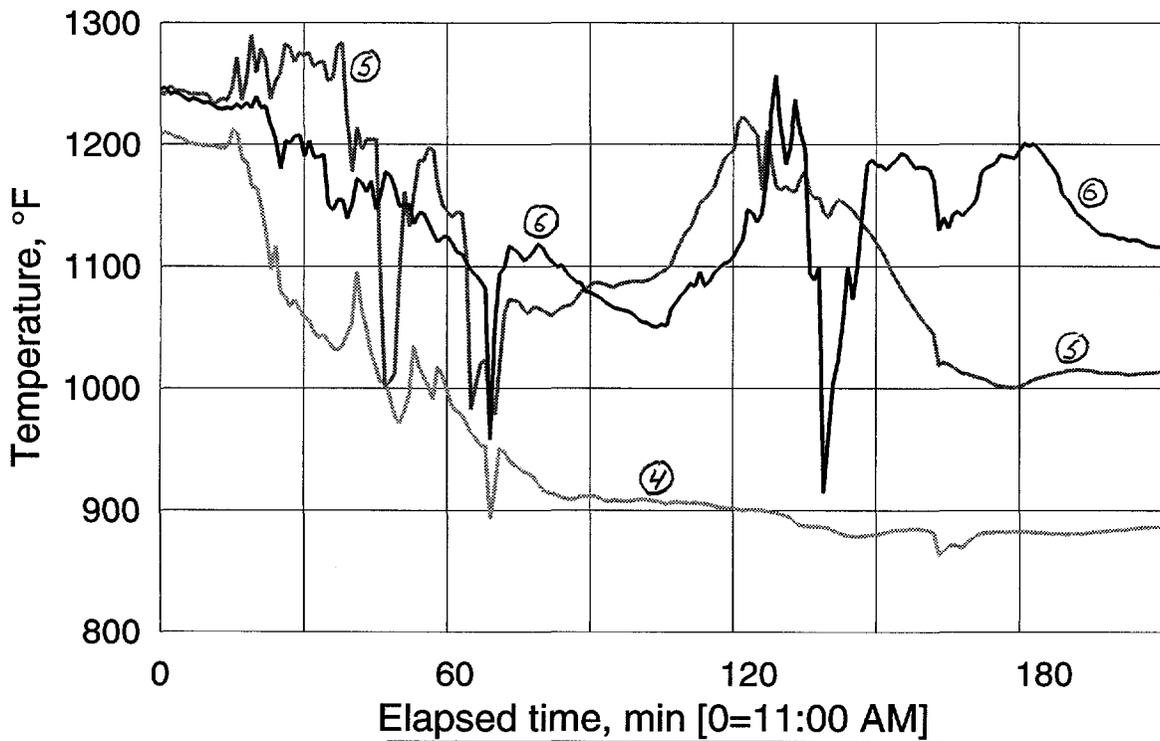


T/C location ① WB ② 1.75" ③ 11" ④ 18"

Figure B.9 - Test 5 (9/18/96)



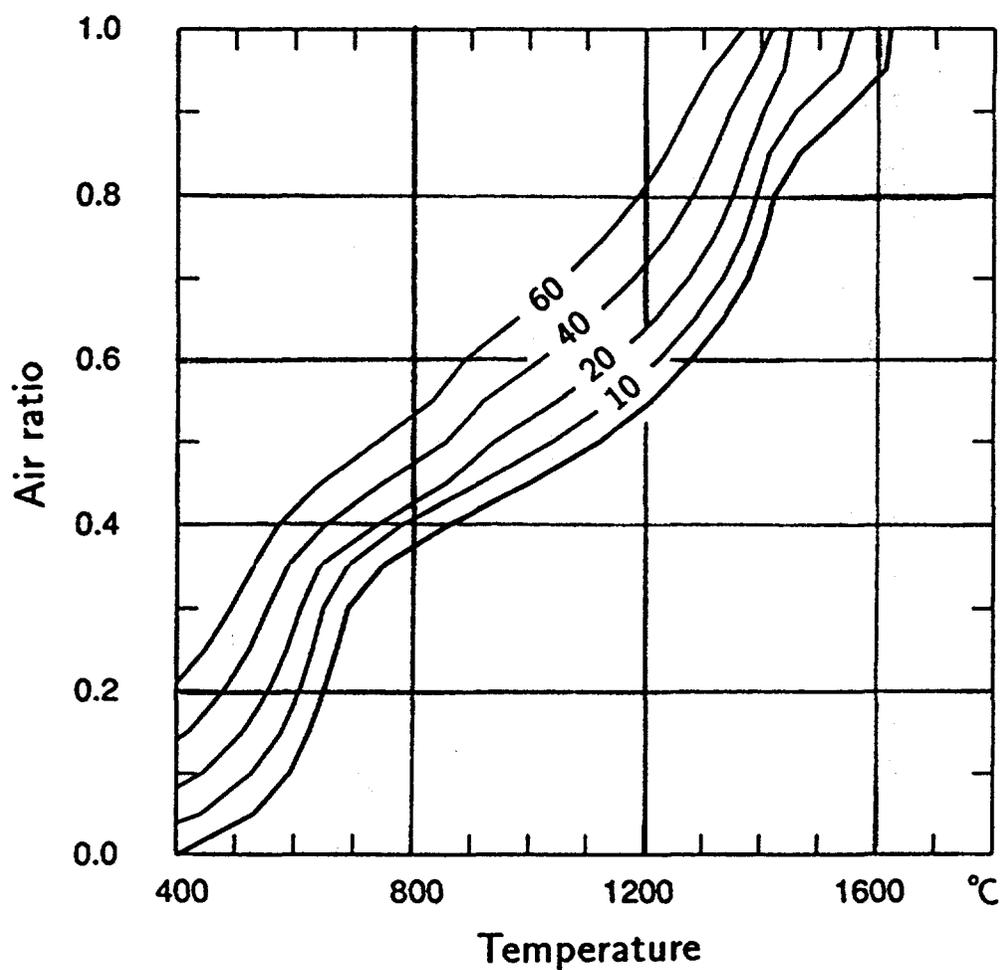
T/C location ① Windbox ② Bed avg. ③ 24.5"



in-bed T/C location ④ 1.75" ⑤ 11" ⑥ 18"

**Figure B.10. Adiabatic Temperature - Air Ratio Diagram
for Black Liquor Gasification (Backman and Hupa, 1990)**

HHV of black liquor = 15350 kJ/kg BLS
Total pressure = 1 bar
Parameter = Water content of BL (weight-%)



ATTACHMENT B.1

Data Reduction

Tests 1&2

Date		16-Jul	23-Jul	23-Jul						
Time		13:30	15:30	16:30	17:00	17:50	18:15	18:30	13:30	14:34
Recorded by		RAM								
Liquor Feeder	setting	0	6.3	12.4	12.4	23.2	23.2	23.2	10.2	10.2
	Solids %	100	100	100	100	100	100	100	100	100
Air Flow	scale rdg	80	83	75	44	40	38	45	41	42
	Press psig	5	4.95	4.8	4.8	4.1	4	4.1	3.4	3
	Temp °F	90	95	96	95	95	95	96	88	92
Steam Flow	"H2O	0.0	0	0	8	7.5	7.4	8.2	9.1	8.6
	Press "Hg				5.75	5.75	5.75	5.8	8.3	8.3
	Temp °F				224.1	224	224	223.7	226.4	226.5
	Cold offset "H2O				1.0	1.0	1.0	1.0	1.0	1.0
N2 Flow	scale rdg	195	205	220	87	95	105	115	87	90
	Press psig	110	110	105	115	112	110	110	111	110
	Temp °F	84	90	91	88	91	91	91	83	88
Purge N2 flow	scale rdg	0.0	4.0	4.1	3.9	3.9	3.9	3.9	7.4	7.4
Air/Steam Heater										
	Inlet Press "H2O	80.0	87.0	86.0	70.0	64.0	64	63	71	72
Bed Pressure Drop										
	WB to FB exit "H2O	24	16.5	17.5	13.5	5.5	6.5	6.3	17	16.5
Temperature										
	Bed @ Windbox °F	870	959	968	957	962	965	968	922	936
	#1 @ 1.75" °F	940	1142	1164	1138	981	972	973	1115	1113
	#2 @ 11.0" °F	939	1182	1191	1339	1183	1267	1140	1150	1177
	#3 @ 18.0" °F	937	1179	1185	1179	1215	1205	1235	1162	1179
	#4 @ 24.5" °F	824	1115	852	908	945	937	935		
	#5 @ Exit °F	420	510	371	358	460	474	496		
	Cyc Exit pipe °F	301	440	364	326	403	415	422		
	BWHX exit °F	145	201	190	192	207	199	210		
	Cond Exit °F	77	80	82	81	82	83	81		
	Ambient °F	81	81	81	81	81	80	80		
Barometric P	"Hg	28.9	28.87	28.86	28.86	28.86	28.86	28.86	28.72	28.72
Calculated Flows										
	Air (Uncor'd) SCFM	3.10	3.22	2.90	1.70	1.54	1.47	1.74	1.58	1.62
	N2 (Uncor'd) SCFM	1.97	2.07	2.21	0.94	1.02	1.11	1.21	0.94	0.97
	Purge N2 (Unc) SLPm	0	4	4.1	3.9	3.9	3.9	3.9	7.4	7.4
	Steam: P(act) psia	14.20	14.18	14.18	17.00	17.00	17.00	17.02	18.18	18.18
	: t(sat) °F	211.64	211.60	211.59	219.58	219.58	219.58	219.65	222.93	222.93
	: Rho lb/ft ³	0.0380	0.0380	0.0380	0.0430	0.0430	0.0430	0.0430	0.0455	0.0455
	Liquor Solids lb/hr	0.00	0.95	2.04	2.04	3.96	3.96	3.96	1.65	1.65
	Cal. accuracy %		± 3.5	± 3.5	± 3.5	± 3.5	± 3.5	± 3.5	± 3.5	± 3.5
	H2O in liquor lb/hr	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
	Vol's (30%) lb/hr	0.00	0.29	0.61	0.61	1.19	1.19	1.19	0.49	0.49
	Char lb/hr	0.00	0.67	1.43	1.43	2.77	2.77	2.77	1.15	1.15
	Air (Corr'd) lb/hr	12.09	12.47	11.20	6.56	5.85	5.54	6.58	5.91	5.96
	N2 (Corr'd) lb/hr	11.08	11.55	12.09	5.37	5.73	6.21	6.75	5.31	5.43
	Steam lb/hr	0.00	0.00	0.00	6.60	6.36	6.32	6.70	7.31	7.08
	Total @ WB lb/hr	23.17	24.02	23.30	18.54	17.94	18.07	20.03	18.53	18.47
	Purge N2 (Cor) lb/hr	0.00	1.75	1.76	1.75	1.72	1.71	1.71	3.28	3.25
	add Purge N2, Evap H2O & Volatile solids									
	Total @ FB lb/hr	23.17	26.06	25.67	20.90	20.85	20.97	22.93	22.30	22.21
	Windbox O2 %, dvb	10.78	10.72	9.93	11.37	10.44	9.73	10.20	10.88	10.82
	O2/BLS lb/lb	ERR	3.05	1.28	0.75	0.34	0.33	0.39	0.84	0.84
	H2O/BLS lb/lb	ERR	0.00	0.00	3.24	1.61	1.59	1.69	4.44	4.30

ATTACHMENT B.1

Data Reduction

Tests 1&2

Date		16-Jul	23-Jul	23-Jul						
Time		13:30	15:30	16:30	17:00	17:50	18:15	18:30	13:30	14:34
Recorded by		RAM								
Molar flow	WB mol/hr	0.8130	0.8429	0.8185	0.7851	0.7598	0.7638	0.8402	0.7995	0.7927
	FB	0.8130	0.9150	0.9017	0.8678	0.8608	0.8643	0.9407	0.9329	0.9251
Static pressure (gauge)										
	@ WB=inlet "H2O	90.0	87.0	86.0	70.0	64.0	64.0	63.0	71.0	72.0
	@ 18"=FB "H2O	66.0	70.5	68.5	56.5	58.5	57.5	56.7	54.0	55.5
Gas Molar Weight										
	@ WB lb/mol	28.50	28.50	28.46	23.62	23.62	23.66	23.84	23.18	23.30
	@ 18.0" lb/mol	28.50	28.48	28.46	24.08	24.22	24.26	24.37	23.91	24.01
Gas Density [Rho] = PM/RT										
	@ WB lb/ft ³	0.0348	0.0324	0.0321	0.0259	0.0255	0.0255	0.0256	0.0261	0.0260
	@ 18.0" lb/ft ³	0.0315	0.0271	0.0268	0.0222	0.0219	0.0221	0.0217	0.0221	0.0220
Gas Superficial Velocity										
	@ 18.0" ft/sec	0.93	1.12	1.09	1.05	1.03	1.03	1.16	1.06	1.06
Fluidized Bed Pressure Drop										
Calc Plate & Bed	*P:									
	W ² /Rho{1} lb/ft ² /hr ²	15412.0	17797.9	16909.6	13253.7	12619.7	12803.5	15679.0	13182.7	13131.8
	Plate DP "H2O	3.7	4.2	4.0	3.2	3.0	3.0	3.7	3.1	3.1
	Bed DP "H2O	20.3	12.3	13.5	10.3	2.5	3.5	2.6	13.9	13.4

ATTACHMENT B.2

Gas Chromatography Analysis

Date		16-Jul	16-Jul	16-Jul	16-Jul	23-Jul	23-Jul	23-Jul
Time		16:18	17:25	18:14	18:25	13:29	14:28	14:43
Analysis by		RFD						
H2	vol%	n.d.	n.d.	n.d.	n.d.	1-2	1.73	2.14
N2	vol%	88.85	91.14	90.35	89.7	88.84	89.86	87.04
CO	vol%	n.d.						
O2 (by diff.)	vol%	10.58	8.48	9.36	10.01	6-7	2.92	6.36
CH4	vol%	0.06	0.01	0.01	0.01	0.17	0.29	0.16
CO2	vol%	0.48	0.37	0.28	0.28	3.01	5.16	4.28
C2H2	vol%	n.d.	n.d.	n.d.	n.d.	n.d.	0.04	0.02
C2H4	vol%	0.03	n.d.	n.d.	n.d.	n.d.	n.d.	n.d.
Total	vol%	100.00	100.00	100.00	100.00	100.00	100.00	100.00

Note: only hydrocarbon analysis was conducted for Tests 1 & 2.

Date		08-Aug	08-Aug	08-Aug	08-Aug	08-Aug	20-Aug	20-Aug	20-Aug
Time		13:21	13:23	14:05	14:07	14:38	12:35	12:59	16:20
Analysis by		RFD							
H2	vol%	7.72	8.04	2	1.68	n.d.	n.d.	n.d.	n.d.
N2	vol%	78.03	77.96	85.31	86.39	86.34	91.15	90.66	91.62
CO	vol%	1.75	1.72	n.d.	n.d.	n.d.	n.d.	n.d.	n.d.
O2 (by diff.)	vol%	n.d.	n.d.	6.73	6.94	13.06	n.d.	n.d.	7.46
CH4	vol%	0.69	0.67	0.1	0.15	0.02	0.39	0.46	0.102
CO2	vol%	11.2	11.21	5.82	4.74	0.51	4.99	5.26	0.821
C2H2	vol%	n.d.							
C2H4	vol%	0.057	0.055	n.d.	0.018	n.d.	0.03	0.03	n.d.
H2S	ppmv	2220	2330	142.7	363	42.6	2624	274	---
CH3SH	ppmv	660	580	135	268	399	2203	2203	---
(CH3)2S	ppmv	389	345	78.5	138	131	614	677	---
CS2	ppmv	189	190	n.d.	50	103	560	1803	---
Tot. reduced S	vol%	0.346	0.345	0.036	0.082	0.068	0.600	0.496	
Total	vol%	99.79	100.00	100.00	100.00	100.00	97.16	96.91	100.00

n.d. = none detected

ATTACHMENT B.3

Data Reduction

Test 3

Date		08-Aug	08-Aug	08-Aug	08-Aug	08-Aug
Time		11:30-12:10	12:50-13:00	13:20-13:30	14:04-14:10	14:30
Recorded by		RAM	RAM	RAM	RAM	RAM
Liquor Pump	setting	0	4	5	5	6
Solids	%	53.55	53.55	53.55	53.55	53.55
Air Flow	scale rdg	44	44	44	45	64
Press	psig	6.5	6.1	6.1	6.1	6
Temp	°F	92	93	92	93	91
Steam Flow	"H2O	8.5	8.6	5.8	5.7	7
Press	"Hg	8.3	8.3	8.3	8.3	8.3
Temp	°F	227	227	227.5	226	226
Cold offset	"H2O	1.0	1.0	1.0	1.0	1.0
N2 Flow	scale rdg	87	90	71	73	40
Press	psig	112	112	112	112	112
Temp	°F	90	90	89	89	88
Purge N2 flow	scale rdg	5.0	5.1	5.0	5.1	5.0
Air/Steam Heater						
Inlet Press	"H2O	76.0	72.0	60.0	56.0	64.0
Bed Pressure Drop						
WB to FB exit	"H2O	16.8	16.6	11.8	7.0	9.0
Temperature						
Bed @ Windbox	°F	870	904	880	877	898
#1 @ 1.75"	°F	1016	960	894	886	907
#2 @ 11.0"	°F	1056	1163	1222	1247	1064
#3 @ 18.0"	°F	1027	1107	1125	958	710
#4 @ 24.5"	°F	876	928	970	997	780
#5 @ Exit	°F	379	411	422	388	386
Cyc Exit pipe	°F	345	372	367	369	365
BWHX exit	°F	181	193	188	189	192
Cond Exit	°F	80	80	79	79	80
Ambient	°F	85	85	82	82	82
Barometric P	"Hg	28.88	28.88	28.87	28.86	28.85
Calculated Flows						
Air (Uncor'd)	SCFM	1.70	1.70	1.70	1.74	2.48
N2 (Uncor'd)	SCFM	0.94	0.97	0.79	0.81	0.49
Purge N2 (Unc)	SLPM	5	5.1	5	5.1	5
Steam: P(act)	psia	18.26	18.26	18.26	18.25	18.25
: t(sat)	°F	223.15	223.15	223.14	223.13	223.11
: Rho	lb/ft ³	0.0457	0.0457	0.0457	0.0457	0.0457
Liquor Solids	lb/hr	0.00	2.12	4.20	4.20	6.28
Cal. accuracy	%		± 18.0	± 18.0	± 18.0	± 18.0
H2O in liquor	lb/hr	0.00	1.84	3.64	3.64	5.45
Vol's (30%)	lb/hr	0.00	0.64	1.26	1.26	1.88
Char	lb/hr	0.00	1.48	2.94	2.94	4.40
Air (Corr'd)	lb/hr	6.87	6.80	6.80	6.95	9.90
N2 (Corr'd)	lb/hr	5.30	5.46	4.45	4.55	2.78
Steam	lb/hr	7.05	7.10	5.64	5.58	6.30
Total @ WB	lb/hr	19.22	19.36	16.89	17.08	18.98
Purge N2 (Cor)	lb/hr	2.21	2.25	2.21	2.25	2.21
add Purge N2, Evap H2O & Volatile solids						
Total @ FB	lb/hr	21.43	24.08	24.00	24.24	28.52
Windbox O2	%, dvb	11.68	11.47	12.53	12.52	16.27
O2/BLS	lb/lb	ERR	0.75	0.38	0.39	0.37
H2O/BLS	lb/lb	ERR	4.22	2.21	2.20	1.87

ATTACHMENT B.3

Data Reduction

Test 3

Date	08-Aug	08-Aug	08-Aug	08-Aug	08-Aug
Time	11:30-12:10	12:50-13:00	13:20-13:30	14:04-14:10	14:30
Recorded by	RAM	RAM	RAM	RAM	RAM
Molar flow WB mol/hr	0.8178	0.8236	0.7066	0.7122	0.7907
FB	0.8966	1.0272	1.0296	1.0369	1.2349
Static pressure (gauge)					
@ WB=inlet "H2O	76.0	72.0	60.0	56.0	64.0
@ 18"=FB "H2O	59.2	55.4	48.2	49.0	55.0
Gas Molar Weight					
@ WB lb/mol	23.51	23.50	23.90	23.99	24.00
@ 18.0" lb/mol	23.90	23.45	23.31	23.38	23.10
Gas Density [Rho] = PM/RT					
@ WB lb/ft ³	0.0279	0.0269	0.0272	0.0271	0.0271
@ 18.0" lb/ft ³	0.0244	0.0226	0.0218	0.0245	0.0297
Gas Superficial Velocity					
@ 18.0" ft/sec	0.99	1.08	0.98	0.88	0.80
<u>Fluidized Bed Pressure Drop</u>					
Calc Plate & Bed •P:					
W ² /Rho(1) lb/ft ² /hr ²	13252.6	13903.3	10497.1	10778.5	13270.0
Plate DP "H2O	3.2	3.3	2.5	2.6	3.2
Bed DP "H2O	13.6	13.3	9.3	4.4	5.8

ATTACHMENT B.4

Data Reduction

Test 4

Date		20-Aug	20-Aug	20-Aug	20-Aug	20-Aug	20-Aug
Time		11:15	12:07-12:09	12:25-12:35	13:00	14:30	15:52
Recorded by		RAM	RAM	RAM	RAM	RAM	RAM
Liquor Pump	setting	0	0	5	5.5	5.5	5.5
Solids	%	53.03	53.03	53.03	53.03	53.03	53.03
Air Flow	scale rdg	88	45	41	41	43	64
Press	psig	7.6	8	8	8	8	7.8
Temp	°F	86	89	91	93	96	95
Steam Flow	"H2O	13.0	12.5	12.9	12.5	7	9
Press	"Hg	8	8.4	8.4	8.8	8.9	8.6
Temp	°F	226.6	226.3	227.6	228	227.6	227.4
Cold offset	"H2O	1.0	1.0	1.0	1.0	1.0	1
N2 Flow	scale rdg	0	45	149	148	200	148
Press	psig	112	110	110	113	110	110
Temp	°F	85	85	87	90	93	91
Purge N2 flow	scale rdg	0.0	5.2	5.2	5.3	5.2	5.2
Air/Steam Heater							
Inlet Press	"H2O	52.0	80.0	96.0	80.0	102.0	96
Bed Pressure Drop							
WB to FB exit	"H2O	12.0	14.5	15.0	15.5	20.0	15
Temperature							
Bed @ Windbox	°F	890	944	947	934	922	909
#1 @ 1.75"	°F	959	994	1003	1013	935	921
#2 @ 11.0"	°F	972	1003	1014	1031	1012	985
#3 @ 18.0"	°F	985	998	967	985	1039	1016
#4 @ 24.5"	°F	800	855	858	873	931	540
#5 @ Exit	°F	356	408	451	470	471	451
Cyc Exit pipe	°F	312	367	409	443	442	419
BWHX exit	°F	172	187	230	253	254	242
Cond Exit	°F	79	80	87	87	82	82
Ambient	°F	79	80	81	81	83	84
Barometric P	"Hg	29.08	29.07	29.06	29.05	29.04	29.01
Calculated Flows							
Air (Uncor'd)	SCFM	3.41	1.74	1.58	1.58	1.66	2.48
N2 (Uncor'd)	SCFM	0.00	0.54	1.53	1.53	2.02	1.53
Purge N2 (Unc)	SLPM	0	5.2	5.2	5.3	5.2	5.2
Steam: P(act)	psia	18.21	18.41	18.40	18.59	18.64	18.47
: t(sat)	°F	223.01	223.56	223.54	224.09	224.21	223.75
: Rho	lb/ft ³	0.0456	0.0460	0.0460	0.0465	0.0466	0.0462
Liquor Solids	lb/hr	0.00	0.00	3.80	5.00	5.00	5.00
Cal. accuracy	%			± 5.3	± 5.3	± 5.3	± 5.3
H2O in liquor	lb/hr	0.00	0.00	3.36	4.43	4.43	4.43
Vol's (30%)	lb/hr	0.00	0.00	1.14	1.50	1.50	1.50
Char	lb/hr	0.00	0.00	2.66	3.50	3.50	3.50
Air (Corr'd)	lb/hr	14.25	7.31	6.65	6.63	6.94	10.31
N2 (Corr'd)	lb/hr	0.00	3.03	8.59	8.62	11.26	8.51
Steam	lb/hr	8.90	8.76	8.91	8.80	6.36	7.32
Total @ WB	lb/hr	23.16	19.11	24.15	24.05	24.56	26.14
Purge N2 (Cor)	lb/hr	0.00	2.29	2.29	2.35	2.27	2.28
add Purge N2, Evap H2O & Volatile solids							
Total @ FB	lb/hr	23.16	21.40	30.94	32.34	32.77	34.35
Windbox O2	%, dvb	21.00	14.70	8.99	8.96	7.84	11.33
O2/BLS	lb/lb	ERR	ERR	0.41	0.31	0.32	0.48
H2O/BLS	lb/lb	ERR	ERR	3.23	2.64	2.16	2.35

ATTACHMENT B.4

Data Reduction

Test 4

Date	20-Aug	20-Aug	20-Aug	20-Aug	20-Aug	20-Aug
Time	11:15	12:07-12:09	12:25-12:35	13:00	14:30	15:52
Recorded by	RAM	RAM	RAM	RAM	RAM	RAM
Molar flow WB mol/hr	0.9865	0.8470	1.0308	1.0252	0.9948	1.0659
FB	0.9865	0.9288	1.3371	1.4053	1.3721	1.4433
Static pressure (gauge)						
@ WB=inlet "H2O	52.0	80.0	96.0	80.0	102.0	96.0
@ 18"=FB "H2O	40.0	65.5	81.0	64.5	82.0	81.0
Gas Molar Weight						
@ WB lb/mol	23.48	22.56	23.43	23.46	24.69	24.52
@ 18.0" lb/mol	23.48	23.04	23.14	23.01	23.88	23.80
Gas Density [Rho] = PM/RT						
@ WB lb/ft ³	0.0262	0.0257	0.0275	0.0269	0.0299	0.0296
@ 18.0" lb/ft ³	0.0238	0.0245	0.0260	0.0246	0.0256	0.0258
Gas Superficial Velocity						
@ 18.0" ft/sec	1.23	0.98	1.17	1.23	1.21	1.28
Fluidized Bed Pressure Drop						
Calc Plate & Bed •P:						
W ² /Rho{1} lb/ft ² /hr ²	20479.9	14205.4	21188.0	21501.9	20195.2	23103.1
Plate DP "H2O	4.9	3.4	5.0	5.1	4.8	5.5
Bed DP "H2O	7.1	11.1	10.0	10.4	15.2	9.5

ATTACHMENT B.5
Data Reduction
Air Heater Performance

		05-Sep 11:45	05-Sep 12:58	05-Sep 15:40
Date		05-Sep	05-Sep	05-Sep
Time		11:45	12:58	15:40
Recorded by		CLV	CLV	CLV
Liquor Pump	setting	0	0	0
Solids	%	50	50	50
Air Flow	scale rdg	100	100	100
Press	psig	5.6	5.8	5.8
Temp	°F	83	83	87
Steam Flow	"H2O	12.8	13.3	12.6
Press	"Hg	9.2	9.3	9.2
Temp	°F	227	226.5	226
Cold offset	"H2O	1.0	1.0	1.0
N2 Flow	scale rdg	0	0	0
Press	psig			
Temp	°F			
Purge N2 flow	scale rdg	0.0	0.0	0.0
Air/Steam Heater				
Inlet Press	"H2O	76.0	72.0	82.0
Bed Pressure Drop				
WB to FB exit	"H2O	5.0	5.0	14.5
Temperature				
Bed @ Windbox	°F	885	879	964
#1 @ 1.75"	°F	871	869	
#2 @ 11.0"	°F	979	986	1230
#3 @ 18.0"	°F	975	984	1212
#4 @ 24.5"	°F	817	829	1021
#5 @ Exit	°F	472	478	543
Cyc Exit pipe	°F	432	441	488
BWHX exit	°F	191	188	190
Cond Exit	°F	80	80	81
Ambient	°F	77	78	80
Barometric P	"Hg	28.9	28.9	28.9
Calculated Flows				
Air (Uncor'd)	SCFM	3.88	3.88	3.88
N2 (Uncor'd)	SCFM	0.00	0.00	0.00
Purge N2 (Unc)	SLPM	0	0	0
Steam: P{act}	psia	18.71	18.76	18.71
: t{sat}	°F	224.43	224.57	224.43
: Rho	lb/ft ³	0.0467	0.0469	0.0467
Liquor Solids	lb/hr	0.00	0.00	0.00
Cal. accuracy	%			
H2O in liquor	lb/hr	0.00	0.00	0.00
Vol's (30%)	lb/hr	0.00	0.00	0.00
Char	lb/hr	0.00	0.00	0.00
Air (Corr'd)	lb/hr	15.45	15.53	15.47
N2 (Corr'd)	lb/hr	0.00	0.00	0.00
Steam	lb/hr	8.92	9.14	8.87
Total @ WB	lb/hr	24.38	24.67	24.34
Purge N2 (Cor)	lb/hr	0.00	0.00	0.00
add Purge N2, Evap H2O & Volatile solids				
Total @ FB	lb/hr	24.38	24.67	24.34
Windbox O2	%, dvb	21.00	21.00	21.00
O2/BLS	lb/lb	ERR	ERR	ERR
H2O/BLS	lb/lb	ERR	ERR	ERR

ATTACHMENT B.5
 Data Reduction
 Air Heater Performance

Date		05-Sep	05-Sep	05-Sep
Time		11:45	12:58	15:40
Recorded by		CLV	CLV	CLV
Molar flow	WB mol/hr	1.0289	1.0437	1.0264
	FB	1.0289	1.0437	1.0264
Static pressure (gauge)				
	@ WB=inlet "H2O	76.0	72.0	82.0
	@ 18"=FB "H2O	71.0	67.0	67.5
Gas Molar Weight				
	@ WB lb/mol	23.69	23.64	23.71
	@ 18.0" lb/mol	23.69	23.64	23.71
Gas Density [Rho] = PM/RT				
	@ WB lb/ft ³	0.0278	0.0276	0.0266
	@ 18.0" lb/ft ³	0.0258	0.0253	0.0220
Gas Superficial Velocity				
	@ 18.0" ft/sec	1.19	1.23	1.40
Fluidized Bed Pressure Drop				
Calc Plate & Bed •P:				
	W ² /Rho{1} lb/ft ² /hr ²	21371.7	22031.8	22255.0
	Plate DP "H2O	5.1	5.2	5.3
	Bed DP "H2O	-0.1	-0.2	9.2

ATTACHMENT B.6

Data Reduction

Test 5

Date		18-Sep	18-Sep	18-Sep	18-Sep
Time		10:43	11:36	13:09	14:13
Recorded by		RAM	RAM	RAM	RAM
Liquor Pump	setting	0	4	4	0
Solids	%	52.56	52.56	52.56	52.56
Air Flow	scale rdg	102	26	54	103
Press	psig	6	6	6	6
Temp	°F	91	90	89	90
Steam Flow	"H2O	13.8	12.7	13.8	14
Press	"Hg	10	10.1	10	10
Temp	°F	229.4	228.7	228	228.7
Cold offset	"H2O	1.0	1.0	1.0	1.0
N2 Flow	scale rdg	0	210	180	0
Press	psig	118	113	112	112
Temp	°F	91	87	89	89
Purge N2 flow	scale rdg	4.8	5.2	5.2	4.9
Air/Steam Heater					
Inlet Press	"H2O	106.0	104.0	100.0	114.0
Bed Pressure Drop					
WB to FB exit	"H2O	11.8	10.0	7.0	6.5
Temperature					
Bed @ Windbox	°F	959	932	885	882
#1 @ 1.75"	°F	1193	1042	891	883
#2 @ 11.0"	°F	1227	1221	1162	1013
#3 @ 18.0"	°F	1232	1139	1207	1129
#4 @ 24.5"	°F	1006	1010	972	845
#5 @ Exit	°F	555	553	531	538
Cyc Exit pipe	°F	514	518	474	490
BWHX exit	°F	256	258	250	257
Cond Exit	°F	81	86	81	81
Ambient	°F	83	85	85	85
Barometric P	"Hg	28.87	28.87	28.87	28.86
Calculated Flows					
Air (Uncor'd)	SCFM	3.95	1.00	2.09	3.99
N2 (Uncor'd)	SCFM	0.00	2.12	1.83	0.00
Purge N2 (Unc)	SLPM	4.8	5.2	5.21	4.9
Steam: P{act}	psia	19.09	19.14	19.09	19.09
: t{sat}	°F	225.50	225.64	225.50	225.49
: Rho	lb/ft ³	0.0477	0.0478	0.0477	0.0477
Liquor Solids	lb/hr	0.00	2.32	2.32	0.00
Cal. accuracy	%		± 14.4	± 14.4	
H2O in liquor	lb/hr	0.00	2.09	2.09	0.00
Vol's (30%)	lb/hr	0.00	0.70	0.70	0.00
Char	lb/hr	0.00	1.62	1.62	0.00
Air (Corr'd)	lb/hr	15.80	4.00	8.36	15.97
N2 (Corr'd)	lb/hr	0.00	11.99	10.31	0.00
Steam	lb/hr	9.40	9.00	9.40	9.48
Total @ WB	lb/hr	25.20	25.00	28.08	25.44
Purge N2 (Cor)	lb/hr	2.17	2.31	2.30	2.17
add Purge N2, Evap H2O & Volatile solids					
Total @ FB	lb/hr	27.37	30.10	33.17	27.61
Windbox O2	%, dvb	21.00	5.12	9.23	21.00
O2/BLS	lb/lb	ERR	0.40	0.84	ERR
H2O/BLS	lb/lb	ERR	4.79	4.96	ERR

ATTACHMENT B.6

Data Reduction

Test 5

Date	18-Sep	18-Sep	18-Sep	18-Sep
Time	10:43	11:36	13:09	14:13
Recorded by	RAM	RAM	RAM	RAM
Molar flow WB mol/hr	1.0676	1.0661	1.1788	1.0774
FB	1.1450	1.2880	1.4004	1.1547
Static pressure (gauge)				
@ WB=inlet "H2O	106.0	104.0	100.0	114.0
@ 18"=FB "H2O	94.2	94.0	93.0	107.5
Gas Molar Weight				
@ WB lb/mol	23.61	23.45	23.82	23.62
@ 18.0" lb/mol	23.91	23.37	23.69	23.91
Gas Density [Rho] = PM/RT				
@ WB lb/ft ³	0.0279	0.0282	0.0294	0.0300
@ 18.0" lb/ft ³	0.0231	0.0239	0.0232	0.0253
Gas Superficial Velocity				
@ 18.0" ft/sec	1.37	1.32	1.52	1.27
Fluidized Bed Pressure Drop				
Calc Plate & Bed ·P:				
W ² /Rho{1} lb/ft ² /hr ²	22754.0	22197.8	26852.2	21582.8
Plate DP "H2O	5.4	5.3	6.4	5.1
Bed DP "H2O	6.4	4.7	0.6	1.4

APPENDIX C
BLACK LIQUOR ANALYSIS

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BLACK LIQUOR ANALYSIS

Two samples of black liquor were used in bench-scale gasification experiments: spray-dried powder and 55% solids content industrial liquor. Use of the powder was discontinued after the second test and no chemical analysis was conducted. The chemical analysis and combustion characteristics of the industrial liquor are given in Attachments C.1 and C.2.

Black Liquor Solids

Weak black liquor was obtained from a northeastern kraft pulp mill and spray dried to a fine powder of approximately 100% solids by a supplier. The quantity of fines in the powder, 55% by weight less than 44 μm (Table C.1), was excessive for use in the bench-scale tests. Calculations suggest that any material less than 50 μm would be elutriated from the fluidized bed reactor operating at design conditions. Tests 1 and 2 were conducted with a limited amount of useable material obtained from several large pieces of solid black liquor provided by the supplier and the lumps of solids that were found in one barrel of the powder. This material was milled and screened. The solids passing a #12 screen and retained on a #100 was used for the tests (149-1680 μm).

Table C.1. Size Classification of Spray-Dried Black Liquor Powder

Sieve No.	Opening, μm	% smaller
16	1190	100.0
30	590	99.2
50	297	98.6
70	210	98.1
100	149	96.6
140	105	92.7
200	74	84.3
270	53	62.4
325	44	55.0

A small sample of the powder was successfully pelletized by a supplier. High pressure and about 5% binder were used to form hard, dense tablets approximately ½ in. diameter by ¼ in. thick. A simple test was conducted to evaluate the swelling behavior of these tablets relative to the black liquor powder. The B&W liquor swelling test is reliably based on years of technician experience. In this test 0.5 g of black liquor is placed in a crucible and loaded into a muffle furnace at room temperature. The furnace is heated to 300°F, and then increased to 600°F in 50°F increments. The extent of swelling is determined by comparing the size of the resulting char to reference samples that are maintained in the laboratory.

In the literature, the extent of black liquor swelling has been characterized by specific swelling factor, which is a measurement of the particle's swollen volume divided by the initial mass of black liquor solids in the drop. Black liquor reaches a maximum swollen volume near the end of devolatilization of 2.2 to 4.6 times initial diameter. This corresponds to specific swelling factors of 10 to 80 cm³/g. The specific swelling of char made from the black liquor powder at 1200°F was 35.8 cm³/g. This approached values reported by Frederick, et al. (1991) for single drops burned in air at 1472°F (43-66 cm³/g). All densified black liquor samples (tablets and milled tablets) swelled significantly less than the spray-dried liquor powder. Chars made from spray-dried liquor powder were soft and friable, i.e., crumbled by gentle pressure. Similar friability is typical of industrial liquor chars. Chars from liquor tablets were much harder and resisted crushing. The pelletizing process apparently reduced the swelling characteristic of the black liquor solids. Similar findings were reported by Miller (1986).

Several vendors of disk pelletizing, briquetting, and agglomeration process equipment were contacted about densifying the powder. A few tried. All failed. Miserably. It became clear that material could not be processed during summer months because of the extreme hygroscopic nature of the powder. Moreover, no supplier was willing to handle the nuisance dust in an air-conditioned production space. Assuming that a densification method was developed, the resulting product would have to be crushed and screened to the desired size for use in the

gasifier. It was decided to discontinue use of dried solids on account of the uncertain affect of densification on swelling and the additional cost of processing.

Industrial Black Liquor

The 55% solid industrial black liquor was obtained from a mid-western kraft mill. This liquor proved to be easy to handle at room temperature. Attachment C.1 summarizes the black liquor chemical analysis and heating value. This liquor is typical of other North American liquors in B&W's extensive database. All elements fall within the normal range of black liquors; the low levels of chlorine and potassium suggest that the liquor inorganics would melt at a relatively high temperature (Tran, 1986). The low amount of sulfur present as sulfate (0.13% of black liquor solids) indicates that the liquor has not undergone oxidation.

A proximate analysis of the black liquor was calculated from the thermogravimetric analysis data (Attachment C.2). On a dry, ash-free basis, the liquor has typical amounts of volatile matter and fixed carbon. An extreme swelling characteristic was noted in the comparative swelling test. This liquor is not unusual in this regard, extreme swelling characteristics have been observed in many other industrial liquor samples.

References

Frederick, W.J., Noopila, T. and Hupa, M., "Swelling of Spent Pulping Liquor Droplets During Combustion," *J. Pulp Paper Sci.* 17(5):J164-J170 (1991).

Miller, P.T., *Swelling of Kraft Black Liquor*. Ph.D. Dissertation. The Institute of Paper Chemistry, Appleton, WI (1986).

Tran, N.H., "How Does a Kraft Recovery Boiler Become Plugged?" *Tappi J.* 69(11):102-106 (1986).

CRD / PGG / DOE
 INDUSTRIAL & UTILITIES SYSTEMS (ARC)
 ACS-97-43244-110-030
 OCTOBER 8, 1996

B&W Sample No. F-6016
 Description BLACK LIQUOR, 9/19/96

BASIS (OVEN DRY)		AS REC'D	DRY
		-----	---
Moisture	(%)	47.04	0.00
Solids	(%)	52.96	100.00
Total Sulfur	(%S)	2.66	5.02
Gross Heating Value	(Btu/lb.)	2981	5629

ULTIMATE ANALYSIS-ELEMENTAL

Moisture	(%)	47.04	0.00
Carbon	(%C)	17.47	32.99
Hydrogen	(%H)	2.12	4.00
Nitrogen	(%N)	0.06	0.11
Sulfur	(%S)	2.66	5.02
Sodium	(%Na)	11.58	21.87
Potassium	(%K)	0.43	0.81
Chlorine	(%Cl)	0.09	0.17
Inerts	(%)	0.00	0.00
Oxygen (by diff.)	(%O)	18.55	35.03
		-----	-----
Total	(%)	100.00	100.00
Sodium	(%Na2O)	15.61	29.48
Sulfate Sulfur	(%S)	0.07	0.13

CRD / PGG / DOE
INDUSTRIAL & UTILITIES SYSTEMS (ARC)
ACS-97-43244-110-030
OCTOBER 8, 1996

B&W Sample No. F-6016
Description BLACK LIQUOR, 9/19/96

PROCESS RATIOS	Include K	Exclude K
	-----	-----
Potential Sulfidity %	32.21	32.92
Salinity %	0.49	0.50
Potassium in Alkali %	2.13	----

Potential Sulfidity = Total S/Total (Na2 + K2), molar
Salinity = Cl/(Na+K), molar
Potassium in Alkali = K/(Na+K), molar

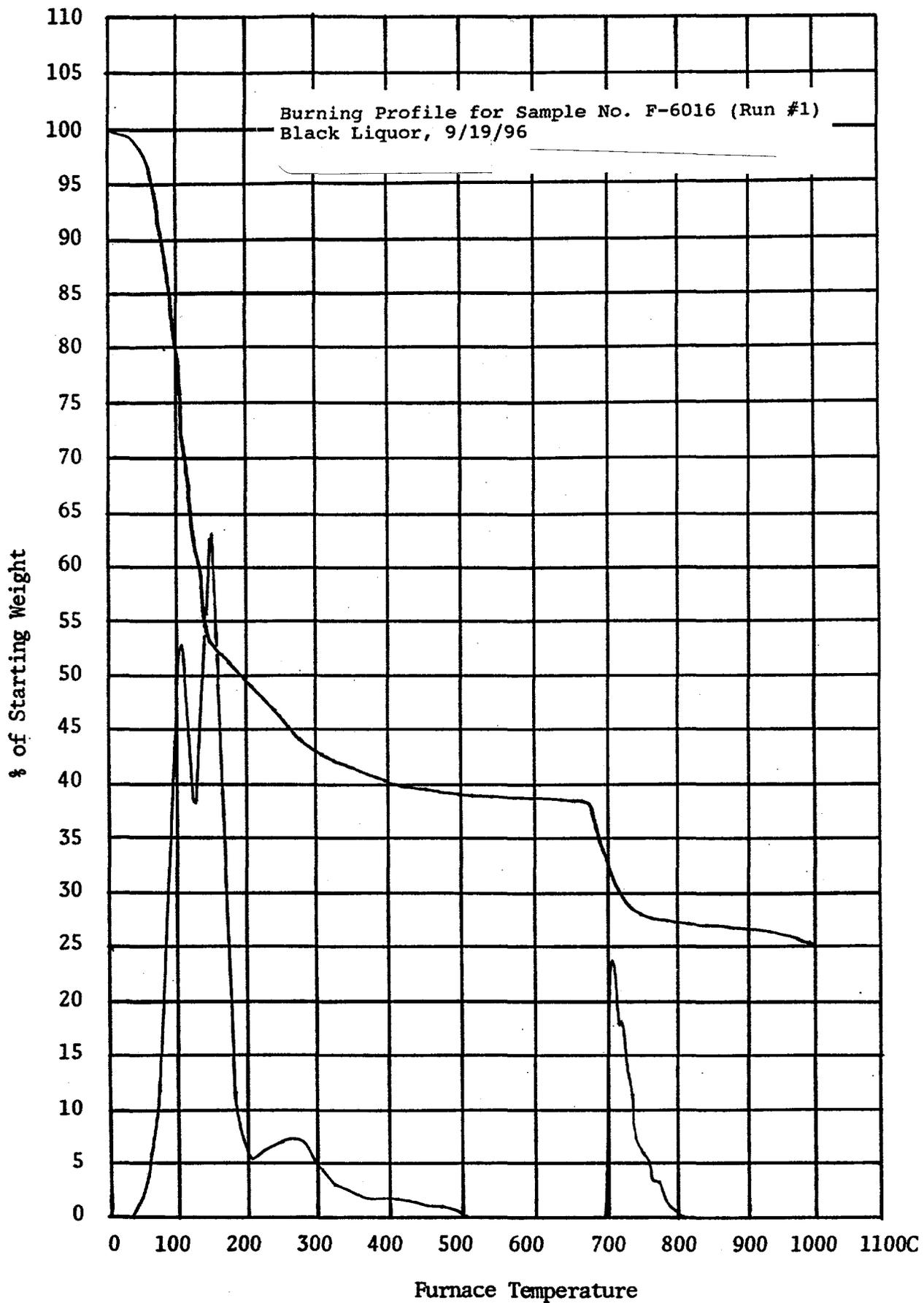
CRD/PGG/DOE
 ACS-97-43244-110-030
 October 8, 1996

<u>Sample No.</u>	<u>Description</u>	<u>As Rec'd Basis Weight Percentage</u>			<u>Dry, Basis Weight Percentage</u>			<u>Dry, Ash Free Basis Weight Percentage</u>			<u>Volatile Ratio (4)</u>	
		<u>H₂O(1)</u>	<u>VM</u>	<u>FC</u>	<u>ASH (2)</u>	<u>ASH (3)</u>	<u>VM</u>	<u>FC</u>	<u>ASH (2)</u>	<u>VM</u>		<u>FC</u>
F-6016	Black Liquor, 9/19/96											
Run #1		47.0	14.8	10.9	27.3	28.4	27.9	20.6	51.5	57.6	42.4	136
Run #2		47.0	14.8	10.7	27.5	28.4	27.9	20.2	51.9	58.0	42.0	138
Average		47.0	14.8	10.8	27.4	28.4	27.9	20.4	51.7	57.8	42.2	137

Note: This black liquor sample was observed to have extreme swelling characteristics.

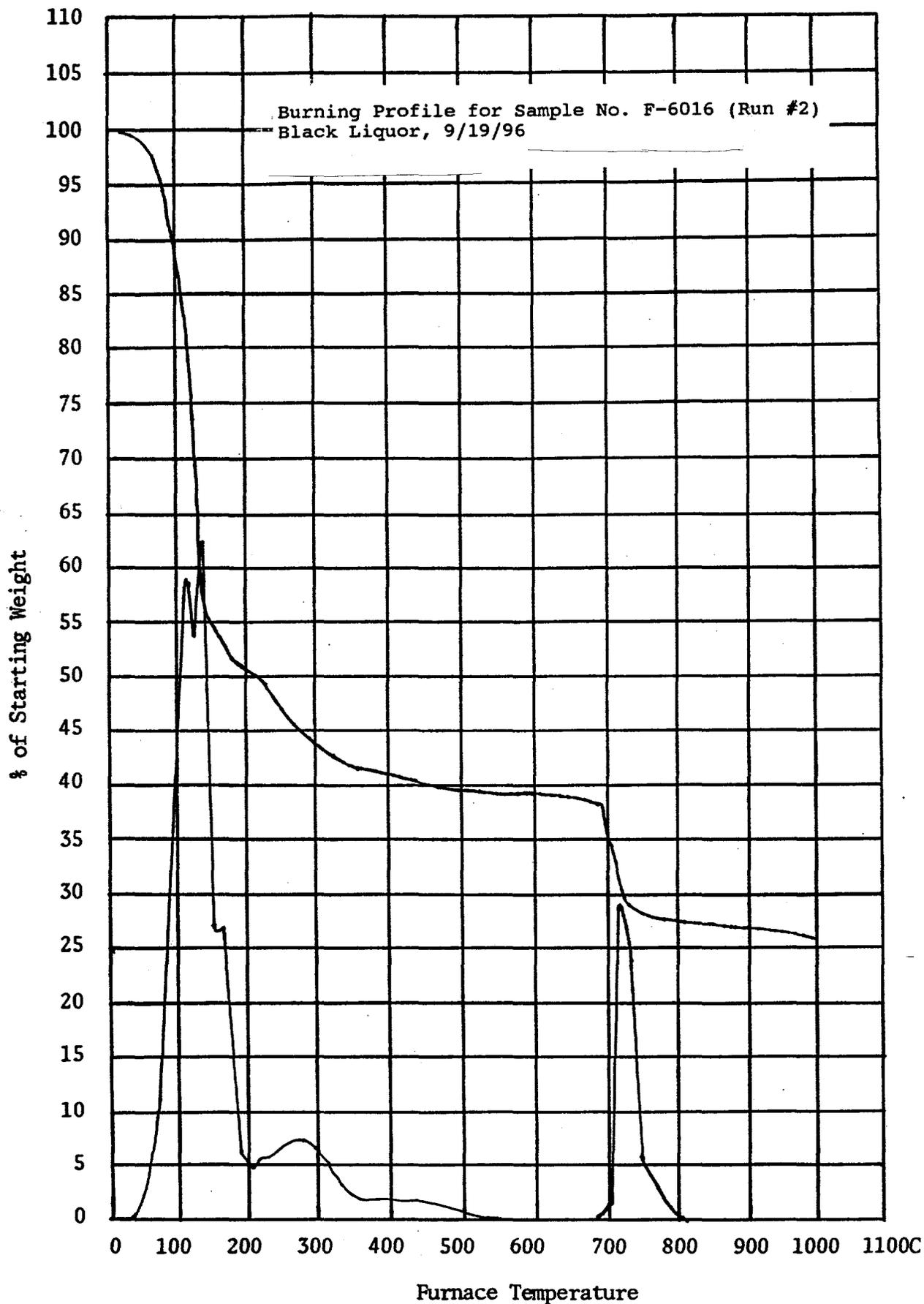
- (1) From Moisture Determination
- (2) By Difference, Not Ultimate
- (3) From Ultimate

(4) $VR = \frac{VM(100)}{FC}$



Sample: F-6016 (Run #1)
Heating Rate: 15°C/Minute
Atmosphere: Air

Thermogravimetric Analysis



Thermogravimetric Analysis

Sample: F-6016 (Run #2)
 Heating Rate: 15°C/Minute
 Atmosphere: Air

APPENDIX D

PILOT-SCALE GASIFIER (PSG) TEST PROGRAM OUTLINE

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PSG TEST PROGRAM REQUIREMENTS

DOE Contract Compliance Requirement

from proposal DE-FC36-94GO10002/A002 Appendix C, page 4

Phase Ia - Task 3: Pilot Plant Test Program Outline

The participant shall prepare an outline of the test program that will be conducted using the pilot unit in Phases II and III. This will include any tests needed to define liquor properties or component performance necessary for final design of the pilot unit or to characterize materials or components that are tested. The test outline will define the approximate number of tests, test objectives and principal information or results sought and the principal measurements or analyses that must be acquired. Development of the test outline at this stage will enable the participant to develop a preliminary pilot unit design suitable for meeting the test program objectives.

B&W Technical Requirements For Pilot-scale Black Liquor Gasifier

The following are excerpts from internal B&W planning documents that must be addressed by the Pilot-Scale Development program:

T2 Reaction kinetics may result in:

- a) high unburned carbon*
- b) incomplete gasification (poor quality product gas)*
- c) excessively large reactor.*

T3 r Overbed "concentration/preheating" of black liquor may be ineffective, increase heat load on reactor and result in carry over (wet or friable).

T4 In-reactor heaters may be required due to T2, reaction kinetics, poor overbed drying or heat losses.

- T5 Concept may be too sensitive to air flow resulting in low caloric gas and high unburned carbon and necessitating higher heat input.*
- T6 Liquor sprayer/nozzle must be understood (pattern, drop size, nozzle design).*
- T7 r Quench code performance (scrubbing).*
- T8 Scrubber performance (< 50 ppm H₂S - no data).*
- T9 Caloric value must be increased (300 Btu/CF) to support future production of electricity.*
- T11 Gas cleanup equipment must emit acceptable levels of particulate, S-compounds, NO_x etc.*
- T12 Sensitivity to black liquor composition and source (type of wood, type of "straw", bamboo, bagasse).*
- T13 Gasifier (reactor) mixing and flow patterns impact reaction, gas quality and amount of unburned carbon.*
- T14 Acceptable/optimum reactor temperature.*
- T15 Product gas combustion may be unstable, require preheating, restrict turndown, require auxiliary fuel.*
- T16 r Fine unsettleable carbon may exist in green liquor.*
- T17 r Scale formation (iron) in green liquor.*
- T18 r Corrosion of reactor heat exchanger materials.*

T19 r Particle size distribution in bed may not be controlled.

Other Technical Issues

Three of the four additional technical concerns (XT1 - XT3) will be addressed by the current bench-scale test program and proposed follow-on work. However, these concerns must also be verified at the pilot scale.

XT1 Tar and mercaptans are likely to be formed at conditions of low-temperature gasification. Cracking of these unwanted substances may be possible given long enough residence time in the fluidized-bed.

XT2 More than 90% of the recovered sulfur must be in a reduced state for acceptable operation of the kraft pulping process. Adequate sulfate reduction is not expected to be a problem in low-temperature gasification; never the less, it must be demonstrated.

XT3 Bed solids melting behavior is sensitive to levels of chlorine and potassium in the black liquor. Acceptable operation of the gasifier must be demonstrated for typical "closed cycle" pulp mill liquors with high Cl and K levels.

XT4 Gasification and sulfur recovery reactions will produce a higher load of sodium carbonate and bicarbonate than conventional recovery. The impact of the additional "load" on the causticization process should be considered in evaluating overall process economics for a gasification-based kraft recovery system.

PSG TEST PROGRAM OUTLINE

Development and use of numerical (gasifier) modeling progresses in parallel with the experimental activities. Primary responsibility for each task is given in parentheses: (420) Combustion Processes Section; (520) Numerical Modeling Section. In the following list, the specific technical requirements addressed by the results are given in brackets [].

PSG Model Development and Design Support

Task 1 (520) - Develop Gasifier Model

- Objectives Develop numerical model of PSG that includes turbulent 3-D flow, drying and devolatilization in above-bed zone, equilibrium reactor in bed zone. Validate model of simplified geometry with results of bench-scale test program.
- Results Predicted amounts and compositions of product gas and char for bench-scale system. Good agreement between predictions and test results is required for model to meet subsequent project needs.
- Measurements None required.

Task 2 (420) - Screen Mill Liquors and Spray Nozzles

- Objectives Screen potential host mill liquors for abnormal properties, e.g., excessive Cl and K levels, using B&W database of analyses. Limited lab tests may be performed if data is lacking. Screen candidate spray nozzles in field by spraying hot, concentrated liquor outside of furnace.
- Results Host site selection criteria. Minor effort avoids subsequent difficulties with low-melting bed solids; identifies other potential difficulties, e.g., low heating value or liquor solids. Initial selection of suitable nozzle(s) for PSG.
- Measurements Sodium, potassium and chlorine content of black liquor (only if results are not

available in database). Temperature, pressure, solids content and flow rate of liquor to test nozzle. Visual observations of spray pattern, preferably recorded by video camera. Duration of nozzle operation without pluggage or deterioration of spray pattern.

Task 3 (420) - Liquor and Spray Characterization

- Objectives** Conduct physical and chemical characterization of liquor from selected host site as inputs for modeling study and pilot gasifier design.
- Results** Liquor characteristics and spray nozzle behavior [T6]. Required inputs for modeling study and pilot gasifier design: liquor composition, heating value, rheological behavior at firing conditions, median drop size, liquor spray pattern, characteristic times for drying and devolatilization, and liquor swelling factor.
- Measurements** Elemental liquor analysis, heating value, thermogravimetric analysis (burning profile), viscosity versus shear rate, and density. Drop size and mass distribution for selected spray nozzle(s); this may be possible from field measurements, or may require limited spray booth testing. Characteristic times for drying and devolatilization of median-sized black liquor drop under simulated gasification atmosphere. Swelling factors for chosen liquor at realistic gasification conditions.

Task 4 (520) - Model Liquor Spray Pattern (4 cases)

- Objectives** Model fate of black liquor spray and gasifier product composition at nominal operating conditions using measured inputs from Task 3. Determine feasibility of over-bed spray approach.
- Results** Summary plots of black liquor distribution versus nozzle elevation above bed surface will be used to set height of freeboard and spray nozzle above bed. Three-dimensional color plots of black liquor particle trajectories and gas

velocity, temperature, and composition in freeboard provide understanding of over-bed spraying, mixing and flow patterns in PSG [T3r; T13].

Measurements None required.

PSG Commissioning and Performance Characterization

Task 5 (420) - Shakedown Tests (8-12 tests)

Objectives Get test facility components running as an integrated system. Establish procedure for safe handling of product streams, e.g., sustained flaring of product gas and carbonated weak wash return to mill's green liquor system.

Results Equipment performance data. Finalized start-up and shut-down procedures. Specific tests may include, but are not limited to:

- Liquor supply system with flow diverted from gasifier.
- Gas heater (test support may be provided by supplier).
- Bubbling bed gasifier with no liquor feed.
- BWHX and quench cooler with no liquor feed; hot steam and air mixture supplied to gasifier.
- Scrubber and green liquor system with product gas simulated by ducting air heater exhaust through gasifier and adding H₂S at exit of BWHX.
- Integrated system with low liquor feed rate, windbox oxygen reduced by ducting air heater exhaust into air supply system.

Measurements Process measurement and sample locations are indicated in Attachment 1.

Task 6 (420) - Baseline Characterization Tests (2-4 tests)

Objectives Conduct short program to establish nominal condition of successful operation and demonstrate material balance closure. Successful operation requires no slagging in bed; control of bed temperature and bed material removal; and reasonable carbon conversion, sulfate reduction and elutriation losses.

Results Sufficient process data to demonstrate material balance closure. Process data and observations to verify suitable over-bed spraying [T3r], gasification performance without in-bed heaters [T2, T4], and solids removal [T19r].

Measurements Flow and composition of reactant and product streams (measurement and sample points indicated in Attachment 1; sample analysis procedures outlined in Attachment 2). Process temperatures and pressures for monitoring equipment performance and correcting gas flow measurements.

Task 7 (520) - Model Baseline Operation (1 case)

Objectives Validate model predictions with results of baseline performance tests.

Results Predicted amounts and compositions of product gas and char; exit gas temperature. Acceptable agreement between predictions and test data will dictate whether or not modeling effort will be of value to balance of program.

Measurements None required.

PSG Performance Optimization and Commercialization Support

Task 8 (520) - Gasifier Performance Modeling Study (9 cases)

Objectives Use modeling to evaluate performance for broad matrix of possible gasifier operating conditions.

Results Predicted amounts and compositions of product gas and char, exit gas temperature and fate of black liquor spray for a fixed black liquor firing rate with varying total gas/solids ratio and oxygen/solids ratio. Reduced level of effort to complete Task 9.

Measurements None required.

Task 9 (420) - Gasifier Performance Optimization Tests (4-8 tests)

- Objectives** Conduct detailed parametric study to evaluate equipment configuration and operating conditions to maximize char carbon conversion and sulfur reduction with a minimum of tar formation and carbon losses. If time allows, investigate effect of oxygen enrichment and vitiation on gasifier performance.
- Results** Optimal operating conditions for subsequent tests. Process data and samples for evaluating carbon conversion [T2, T5]; tar formation and control [XT1]; gas heating value for combined-cycle applications [T9]; sulfate reduction [XT2]; reactor temperature [T14]; and bed behavior [T19r].
- Measurements** Flow and composition of reactant and product streams (measurement and sample points indicated in Attachment 1; sample analysis procedures detailed in Attachment 2). Process temperatures and pressures for monitoring equipment performance and correcting gas flow measurements.

Task 10 (420) - Extended Operation Tests (5-10 tests)

- Objectives** Demonstrate stable operation at optimal conditions over an extended period, e.g., 500 hours. Gain experience for commercial gasifier operation. During steady state gasifier operation, test sulfur removal and green liquor production by replacing NaOH scrubbing solution with carbonated weak wash from gasifier mix tank. Continuously monitor heating value of sulfur-free product gas and combustion in flare burner. Evaluate erosion and/or corrosion rates of material test coupons for conditions in gasifier bed, free-board and gas condenser.
- Results** Pre-commercial marketing demonstration. Process data and samples for evaluating sulfur removal [T7r; T8]; product gas quality [T9, T11, T15]; green liquor quality [T16r; T17r; XT4]; materials suitability for commercial applications [T18r].
- Measurements** Flow and composition of reactant and product streams (measurement and

sample points indicated in Attachment 1; sample analysis procedures outlined in Attachment 2). Process temperatures and pressures for monitoring equipment performance and correcting gas flow measurements. Wastage of sample test coupons located in gasifier bubbling bed, freeboard and quench cooler.

Task 11 (520) - Model Steady-State Optimal Performance (3 cases)

Objectives Validate model predictions of optimal operation with process data gathered during extended operation tests. Model two additional cases based on Task 12 results.

Results Three-dimensional color plots of black liquor particle trajectories and gas velocity, temperature, and composition in freeboard and amounts and compositions of char used to interpret gasifier performance data and provide starting point for modeling commercial-sized units. Good comparison with test results validates model's ability to predict process changes.

Measurements None required.

Task 12 (420) - Test Sensitivity to Process Changes (6-10 tests)

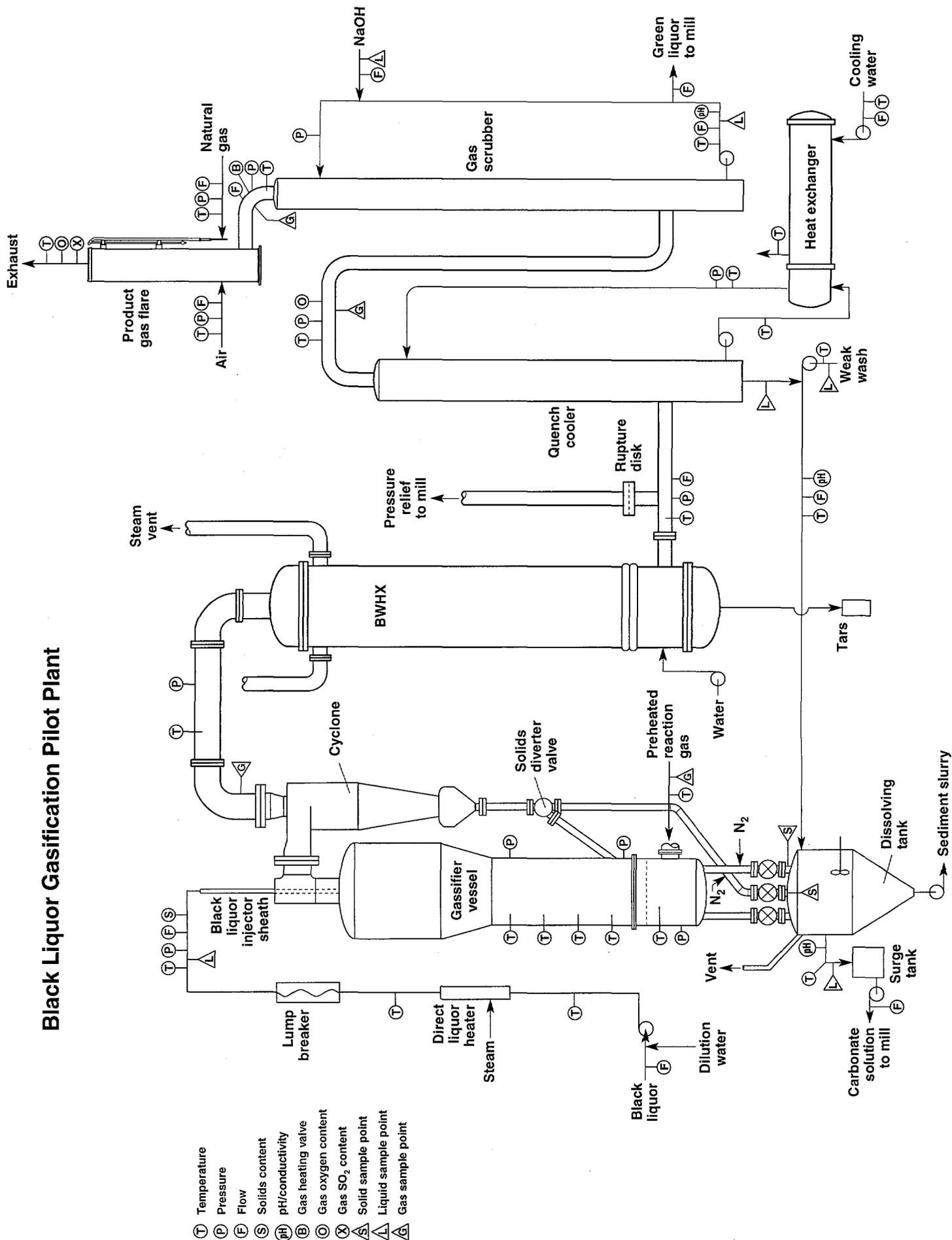
Objectives Evaluate sensitivity of gasifier system to changes in liquor supply, e.g., reduction in liquor solids by dilution, operating at maximum available solids from concentrator system, switching of wood species if practiced at mill, increase in non-process elements by addition of KCl. Adjust operating conditions to run gasifier at reduced loads, e.g., 85%, 70%, 55% of design capacity.

Results Experience with industrial liquor supply fluctuations [T12, XT3]. Process data and samples for predicting commercial system response to load changes and maximum turn-down for acceptable performance [T15].

Measurements Flow and composition of reactant and product streams (measurement and sample points indicated in Attachment 1; sample analysis procedures outlined in Attachment 2). Process temperatures and pressures for monitoring equipment performance and correcting gas flow measurements.

ATTACHMENT D.1
DIAGRAM of PRINCIPAL PROCESS MEASUREMENTS
for BLACK LIQUOR GASIFICATION PILOT PLANT

Black Liquor Gasification Pilot Plant



ATTACHMENT D.2
ANALYTICAL PROCEDURES

Analytical Procedures

The procedures described in this section are those used at B&W's Alliance Research Center (ARC) for standard analysis of kraft recovery process streams. It is assumed that the host mill will provide the facilities to carry out similar procedures. If mill personal will be performing the analysis, some minor adjustments to their standard procedures may be required to insure agreement with ARC results.

Part 1: Black Liquor Analysis

Solids

The solids content of a black liquor is determined gravimetrically using TAPPI T 650 "Solids Content of Black Liquors." In this method the black liquor is dried at 105°C for a minimum of 6 hours. An inert high silica sand is used as a surface extender. High solids black liquors are diluted to reduce scum formation. The method measures the "solids" remaining after removal of water and other nonaqueous volatile materials normally lost in commercial evaporation systems. This test is done in triplicate.

Gross Heating Value

The heating value is determined according to TAPPI T 684 "Gross Heating Value of Black Liquor." In this method the gross (or high) heating value is determined by combusting a sample of the black liquor in an adiabatic oxygen bomb calorimeter. To insure complete combustion, benzoic acid is used as an auxiliary fuel. This procedure is applicable to black liquors containing up to 55% water. This test is performed in duplicate.

Ultimate Analysis - Elemental

A TAPPI procedure is not available for the determination of the ultimate analysis. The ultimate procedure is a B&W procedure based on the analysis of coal, coke and other fuels. Following is a description of each of the determinations.

Carbon and Hydrogen

This procedure is based on ASTM D 3178 "Carbon and Hydrogen in the Analysis Sample of Coal and Coke." In this procedure, a sample of black liquor is burned in a closed system. The combustion products are collected in an absorption train. The carbon and hydrogen are determined gravimetrically. The hydrogen is corrected for the moisture in the sample. This procedure was modified by B&W by lowering the temperature at which the sample is combusted. Lowering the combustion temperature prevents decomposition of the ash. As a result, the carbon value must be corrected for the carbonate in the ash. Care is taken to insure that the temperature is sufficient to burn out all carbon black in the sample. The ash is saved for the determination of carbonate and the inerts. This test is performed in duplicate.

Nitrogen

Nitrogen is determined by ASTM D 3179 "Nitrogen in the Analysis Sample of Coal and Coke." Test Method A (Kjeldahl-Gunning Macro Analysis) is used by B&W. In this procedure, the nitrogen is converted into ammonium salts by destructive digestion of the sample with a hot, catalyzed mixture of concentrated sulfuric acid and potassium sulfate. These salts are decomposed in a hot alkaline solution from which the ammonia is recovered by steam distillation. The nitrogen is then determined by acidimetric titration. The method has been slightly modified in the digestion reagents used. The modification is for laboratory safety reasons and not specific to black liquors. Numerous round robins have shown that the B&W modification has no effect on the nitrogen values obtained. This test is performed in duplicate.

Sulfur

Sulfur is determined following ASTM D 3177 "Total Sulfur in the Analysis Samples of Coal and Coke." The Bomb Washing Method is used. In this method, Sulfur is precipitated as BaSO_4 from the oxygen bomb calorimeter washings. The precipitate is filtered, ashed, and weighed. This test is performed in duplicate.

Chlorine

Chlorine is determined using ASTM D 2361 "Chlorine in Coal." In this method, the chlorine is determined after combustion of the sample in an oxygen bomb calorimeter. The

chlorine is absorbed in alkaline reagent in the bomb. The bomb washings are titrated by a modified Volhard procedure. This test is performed in duplicate.

Sodium and Potassium

Sodium and Potassium are determined by digesting a known weight of black liquor with a mixture of nitric and perchloric acids. After digestion the sample is diluted to a known volume for analysis. Sodium is determined using Inductively Coupled Plasma (ICP) spectroscopy. Potassium is determined using Flame Emission Spectroscopy. These procedures are duplicated.

Inerts

The inerts are determined by correcting the ash (or ultimate ash) obtained in the carbon and hydrogen determination. The ultimate ash is composed of sodium, potassium, sulfate and carbonate. Sodium and potassium were determined previously. The ultimate ash is analyzed for carbonate and sulfate. The inerts are then calculated by difference ($\text{inerts} = 100 - \text{Na}_2\text{O} - \text{K}_2\text{O} - \text{SO}_3 - \text{CO}_2$). The carbonate carbon should be added to the carbon previously determined to obtain the total carbon for the sample. These are single determinations combining the duplicate ashes from the ultimate analysis.

Oxygen

Oxygen is obtained by difference. $\text{Oxygen} = 100 - \text{carbon} - \text{hydrogen} - \text{nitrogen} - \text{sulfur} - \text{sodium} - \text{potassium} - \text{chlorine} - \text{inerts}$. Values are on a "dry" basis. Moisture must also be subtracted if values are on "as received" basis.

Part 2: Bed Solids, Carbonate Liquor and Green Liquor

The following list provides a brief description of procedures used for the analysis of smelts. At ARC, the smelt sample is dissolved in treated water and analyzed as a green liquor. Because the bed solids from low-temperature gasification and the associated process liquors contain essentially the same components as kraft recovery boiler smelt, the same procedures can be used with little or no modification. Analyses consist of a combination of titrimetric, gravimetric and instrumentation methods.

Sodium Carbonate

Sodium carbonate is determined by direct acid titration. A correction must be made for the sodium hydroxide and sodium sulfite present. This method is derived from the Mead Carbonation Recovery Operation methods. An alternate evolution method is contained in TAPPI 624T, "Analysis of Soda and Sulfate White and Green Liquors."

Sodium Hydroxide

Sodium hydroxide is determined by direct titration after precipitation of the carbonate as barium carbonate. A correction must be made for the sodium sulfite present. (See TAPPI 624T)

Sodium Sulfide

Sodium sulfide is determined by an iodine titration for active alkali with corrections made for the presence of thiosulfate and sulfite. (See TAPPI 624T). If the correction yields a negative value, see sodium hydrosulfide.

Sodium Thiosulfate

Sodium thiosulfate is determined by an iodine titration on a solution in which the sulfide has been precipitated in zinc carbonate and the sulfite reaction with iodine has been inhibited by the addition of formaldehyde. (See TAPPI 624T)

Sodium Sulfite

Sodium sulfite is determined by the difference of the combined thiosulfate and sulfite titration and the titration for the thiosulfate alone. (See TAPPI 624T)

Sodium Hydrosulfide

Sodium hydrosulfide appears in smelts when the sodium hydroxide analysis is a negative value. The amount of (negative value) sodium hydroxide is the amount of sodium hydrosulfide present. The sodium sulfide is corrected for the amount of sodium hydrosulfide.

Total Sulfur

Total sulfur is determined by TAPPI 624T using nitric and perchloric acid to digest the

smelt after oxidation with bromine water. The sulfur is precipitated with BaCl. The precipitate is filtered, ashed and weighed.

Sulfate Sulfur

Sulfate sulfur is determined by Mead Carbonation Recovery Operation Method. Analysis consist of a BaCl precipitation after the sulfide sulfur has been boiled off in a solution of excess acid. The precipitate is filtered, ashed and weighed. Sodium sulfate is calculated directly from the sulfate sulfur determined.

Chlorine

Chlorine is determined using the Volhard procedure. In this method, the chlorine is determined by titration with ammonium thiocyanate after the chlorine is absorbed as AgCl₂. Sodium chloride is then directly calculated from the chlorine present.

Sodium and Potassium

Sodium and Potassium are determined by digesting with a mixture of nitric and perchloric acids, then dilute to a known volume. Sodium is determined by Inductively Coupled Plasma (ICP) spectroscopy. Potassium is determined by Flame Emission Spectroscopy.

Carbon

Carbon is determined by a B&W developed method based on ASTM D3178, "Carbon and Hydrogen in the Analysis Sample of Coal and Coke." A known volume, representing a known weight of smelt in solution is filtered and then burned in a closed system purged with oxygen. The combustion products are collected in an absorption train. The carbon is determined gravimetrically.

Reduction and Sulfidity

Reduction and Sulfidity are calculated using the values for the sodium compounds determined above.

Part 3 - Product Gas Analysis

The methods used for product gas sampling and analysis are analogous to those described for bench-scale testing in Attachment 2 of Appendix A. A B&W technician would be present during the testing to calibrate and operate the gas chromatography equipment.

APPENDIX E
PILOT-SCALE GASIFIER PROCESS FLOW DIAGRAM,
HEAT AND MATERIAL BALANCE AND
FUNCTIONAL SPECIFICATIONS

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PART 1 - FUNCTION SPECIFICATION

The pilot-scale gasifier is described in Section 3. A schematic showing the major components is provided in Figure E.1. The functional specifications developed for the following major components/systems follows:

- o Black Liquor Feed System
- o Air Heater
- o Gasifier Vessel
- o Boiling Water Heat Exchanger
- o Dissolving Tank
- o Product Gas Scrubber
- o Product Gas Flame

Sectional side and elevation views for the gasifier vessel and boiling water heat exchanger are also provided in Figures E-2 and E-3.

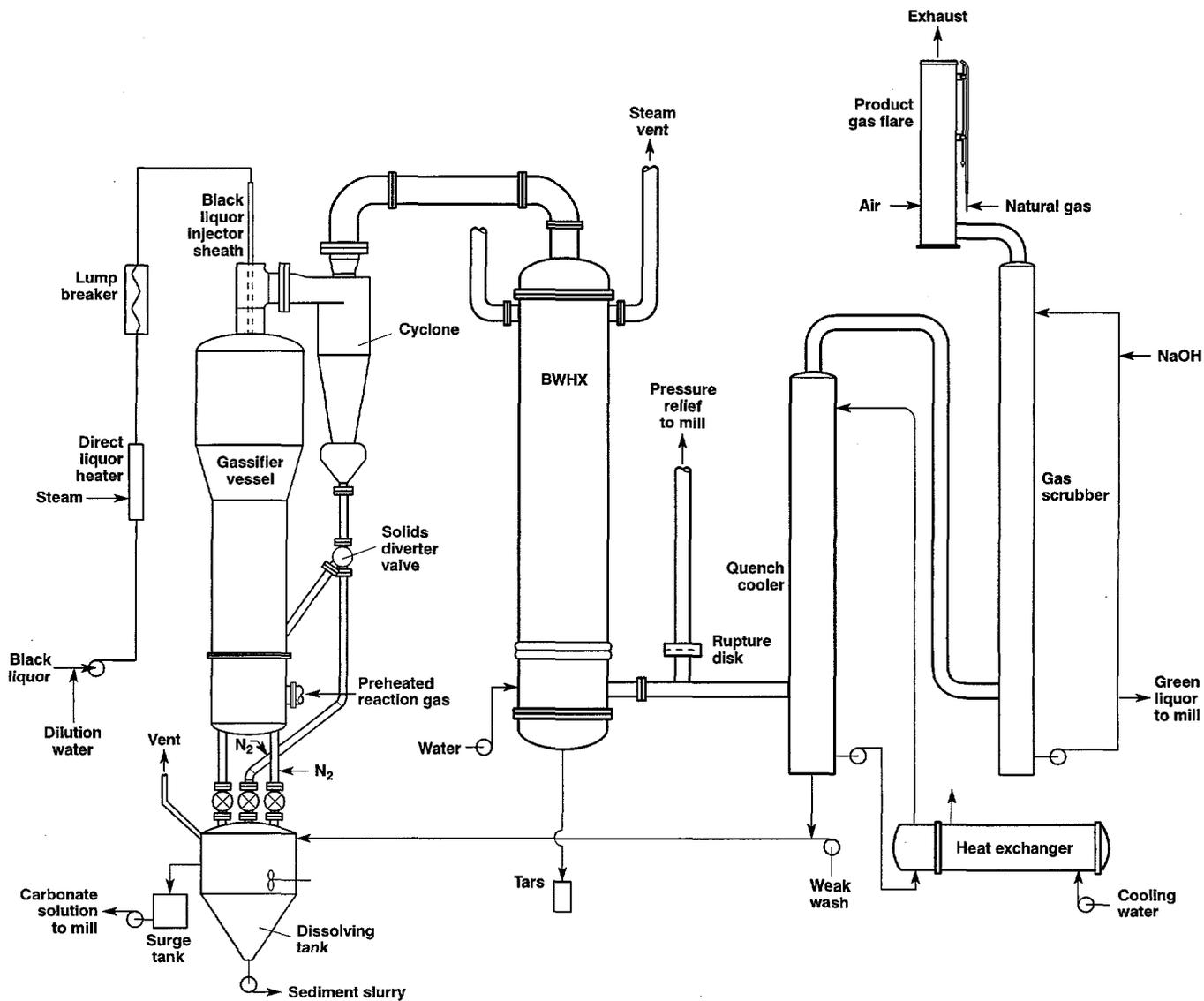


Figure E-1. Pilot-Scale Gasifier Schematic

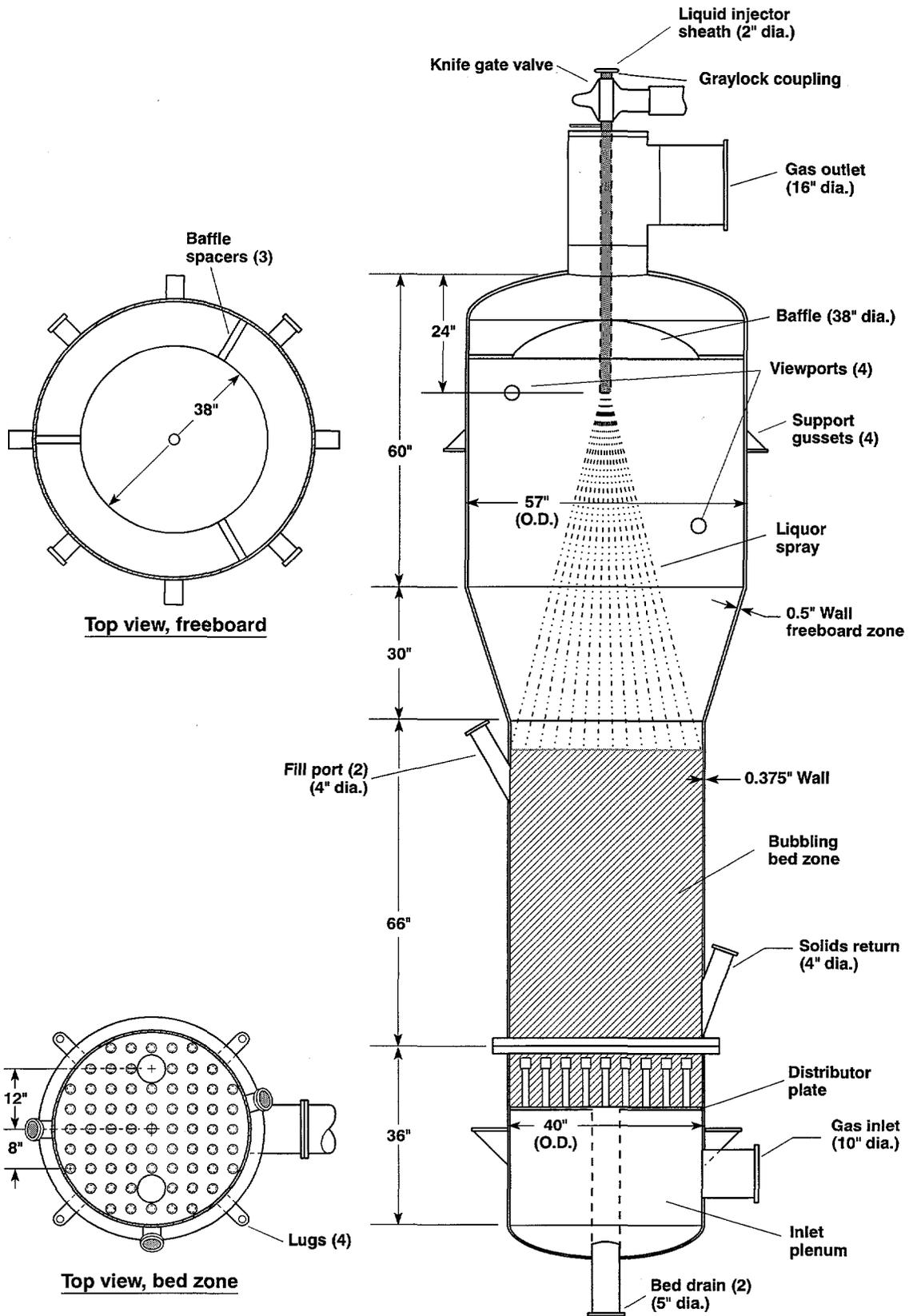


Figure E-2. Pilot-Scale Gasifier Vessel

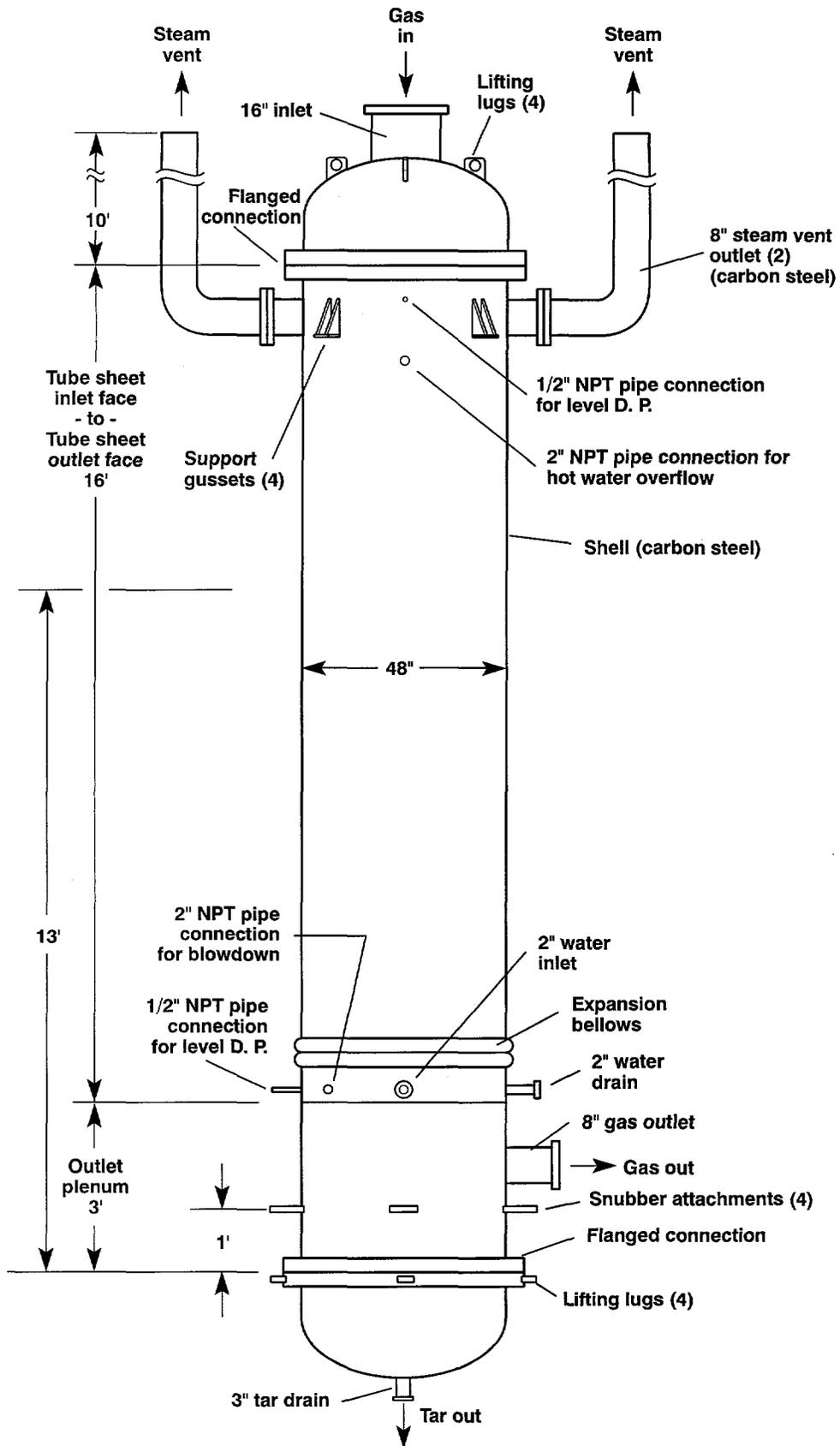


Figure E-3. Pilot-Scale Gasifier Boiling Water Heat Exchanger

Black Liquor Feed System

See Figure E-1.

Conditions

Liquor flow rate	700-1100 lb/hr dry solids 1.5-2.3 gpm (@ 70% solids)
Solids content	50%-73%
Pressure	10-60 psig
Temperature	230-270°F

Materials of Construction

per industry standards

Specifications

pipe size per industry standards

Ancillary Equipment

- 1 Gorator lump breaker
 - 1 direct steam heater
 - 1 steam-cooled injector system (fabricated at ARC)
 - 3 BETE® 3/8" MP spray nozzles (opening size selected in preliminary testing)
 - 1 temperature control for liquor heater
- wash out and drain lines as appropriate (also to serve as dilution point)
- temperature, pressure, solids content (refractometer), and mass flow instruments

<p>This information is for preliminary design considerations only. All engineering data contained herein are preliminary.</p>

Air Heater

Conditions

Inlet air flow	1050-1650 lb/hr
Inlet steam flow	700-1100 lb/hr
Inlet steam pressure	> 50 psig
Inlet exhaust recycle	0-1050 lb/hr
Outlet flow	2500-2800 lb/hr
Outlet temperature	1050-1300°F
Outlet pressure	2.5-10 psig

Specifications

10" Sch 20 outlet pipe with 300 lb class pipe flange
all other dimensions t.b.d. by supplier

Materials of Construction

heat transfer surfaces and piping exposed to hot reaction gas - 304 SS
all other components - minimum necessary to provide 4000 hours service life

Ancillary Equipment

- 2 variable-speed blowers
 - 1 natural gas burner
 - 1 variable-flow ejector for exhaust gas recirculation (preferable to hot gas fan)
 - 2 knife-gate isolation valves; 10" pipe connections
- necessary instruments and control devices to maintain outlet conditions

Gasifier Vessel

See Figures E-1 and E-2.

Code Requirements and Notes

- Code requirements must consider that product is a toxic and flammable gas. All gas-containing piping and vessels must be designed for 50 psig continuous service. Normal operating pressure will be less than 10 psig, as noted below.
- Slip-on flanges are acceptable.
- All piping except for steam or water to be seamless.
- Minimum design metal temperature is 0°F. Design life is 2 years, operating life 4000 hours with 200 thermal cycles.
- Detailed design drawings will be provided by B&W, with supplier comments on preliminary design drawings prior to supplier ordering materials and again prior to B&W issuing released drawing for fabrication.
- All materials of construction 304 SS.

Gasifier - Inlet Plenum

Conditions

Air/steam flow	2500-2800 lb/hr
Composition	52% H ₂ O, 10% O ₂ , 38% N ₂
Pressure	< 10 psig
Temperature	1050-1300°F

Specifications

- 1 40" OD; 3/8" wall; 36" long shell
- 1 40" dia. ring-type pipe flange; 300 lb class
- 1 40" OD - 2:1 ellipsoidal head with 2 - 5 5/8" dia. penetrations for bed drains and 1 - 1" pipe condensate drain line.
- 1 10" Sch 20 gas feed line with 300 lb class pipe flange
- 1 39 1/4" dia. distributor plate; 1/2" thick; 2 - 5 5/8" dia. holes for bed drains; 67 - 2 11/32" dia. holes for bubble caps.
- 2 5" Sch 40 bed drain pipe with 300 lb class pipe flange
- 67 bubble cap air pipes (supplied by B&W)
- 4 lifting lugs
- 2 1/2" thermowells
- 2 1/2" pressure taps

Gasifier - Bubbling Bed Zone

Conditions

Solids discharge	350-5600 lb/hr
Composition	86% Na ₂ CO ₃ , 7% Na ₂ S, 5% Na ₂ SO ₄ , 1% NaCl, 1% C
Pressure	5-7 psig
Temperature	1150-1250°F

Specifications

- 1 40" OD; 3/8" wall; 66" long shell
- 2 4" Sch 40 fill port pipe with 300 lb class pipe flange
- 1 4" Sch 40 solids return pipe with 300 lb class pipe flange
- 8 1/2" thermowells
- 2 1/2" pressure taps
- material test coupons in fluid bed zone

Gasifier - Freeboard Zone

Conditions

Product gas flow	3400-3800 lb/hr
Composition	23% H ₂ , 1% H ₂ S, 5% CO, 13% CO ₂ , 1% CH ₄ , 30% H ₂ O, 27% N ₂
Pressure	< 7 psig
Temperature	1050-1150°F

Specifications

- 1 57" OD; ½" wall; 50" long shell
 - 1 40" OD to 57" OD conical expansion; ½" wall; 30" long
 - 1 38" OD flow baffle with 1 - 2⁷/₁₆" dia. penetration for injector sheath
 - 1 57" OD - 2:1 ellipsoidal head with 1- 16" pipe penetration
 - 1 16" Sch 10 outlet connection to Tee; minimum possible length (e.g. 6")
 - 1 16" straight pipe Tee; butt welding ends
 - 1 16" pipe flange; 300 lb class
 - 1 ½" thick cap plate; 16" dia. with 1- 2⁷/₁₆" dia. penetration for injector sheath
 - 4 4" purged view ports (two opposing sets at different elevations in spray zone)
 - 4 mounting gussets
- material test coupons in freeboard zone

Gasifier - Liquor Injector Sheath

Conditions

- outer, inside dome - same environment as freeboard
- outer, outside dome - mill ambient
- inner - steam purged annulus (contains liquor injector fabricated by B&W)

Specifications

- 1 2" Sch 40 pipe; 50" long
- 1 ½" Sch 40 steam purge line
- 1 2" knife gate valve; 300 lb class; flanged ends
- 1 2" Graylock coupling; 150 lb class

Cyclone and Crossover Piping

See Figure E-1.

Code Requirements and Notes

Per applicable pressure code, same requirements as gasifier vessel.

Inlet Conditions

Gas flow	3400 lb/hr
Gas temperature	1125 °F
Gas density	0.0202 lb/ft ³
Gas viscosity	0.0883 lb/ft·hr
Gas volume	2800 A.C.F.M.
Inlet pressure	1.7 psig
Dust loading	6 Gr/A.C.F.
Particle density	146.7 lb/ft ³
Removal eff.	95% above 5 μm

Specifications

cyclone dimensions t.b.d. by supplier

- 1 gas inlet - 16" Sch 10 pipe with 300 lb class flange connection
- 1 gas outlet - 16" pipe elbow with 300 lb class flange connection
- 1 solids outlet - 4" Sch 40 pipe

Materials of Construction

all materials 304 SS

Ancillary Equipment

- 1 removable pipe length - 16" Sch 10 with 300 lb class flanges
- 1 solids return divert valve; 4" pipe connections
- 2 rotary valves; 5" pipe connections; nitrogen purge line
- 1 rotary valve; 4" pipe connections; nitrogen purge line

Boiling Water Heat Exchanger (BWHX)

See Figures E-1 and E-3.

Code Requirements and Notes

- Code requirements must consider that the gas side of the BWHX contains a toxic and flammable gas. The gas side must be rated for 50 psig at the design temperature. The operating pressure is expected to be less than 5 psig.
- The water side of the BWHX is vented to atmosphere since the function of this unit is to reject heat. Venting to atmosphere removes ASME code requirements for the shell side which is in contact with water and steam only, but must be designed for adequate structural performance as a vented tank. The shell will not be insulated.
- Slip-on flanges are acceptable.
- All piping except for steam or water to be seamless.
- Minimum design metal temperature is 0°F. Design life is 2 years, operating life 4000 hours with 200 thermal cycles.

- Detailed design drawings will be provided by B&W, with supplier comments on preliminary design drawings prior to supplier ordering materials and again prior to B&W issuing released drawings for fabrication.

Conditions

Gas flow	3400-3800 lb/hr
Gas inlet temp.	1050-1150°F
Gas outlet temp.	220-250°F
Gas pressure	< 5 psig operating (tube side), 50 psig rating (tube side)
Water flow	1320 lb/hr operating (includes 20% blowdown)
Fill rate	10,000 lb/hr
Steam flow	1100 lb/hr
Steam temp.	212°F (vents to atmosphere)
Steam pressure	14.7 psia (vents to atmosphere)

Inlet Head Specifications

- 1 48" ID - 2:1 ellipsoidal head; 1 gas inlet connection; 4 lifting lugs; 6" straight section between head and flanged connection to body
- 1 gas inlet connection - 16" Sch 20 pipe with flanged connection. Inlet pipe connection at head to flange face 18"
- 2 ½" thermowells in inlet connection pipe
- 1 ½" pressure tap in inlet connection pipe

Body Specifications

- 1 48" ID x 19' long (16' tubesheet-to-tubesheet and 3' outlet plenum to flange face) shell
- 144 1" OD x 0.065" wall seamless tubes on square pitch, welded connection at tubesheet face
- 1 48" expansion joint to compensate for differential expansion of tubes and shell
- 4 support gussets
- 4 snubber connections

- 2 ½" pipe connections for level transmitter
- 4 2" pipe connections (for water inlet, drain, blowdown, and max level overflow)
- 2 8" pipe flanged steam vents, extending to the elevation indicated (shipped separately)

Lower Head Specifications

- 1 48" ID - 2:1 ellipsoidal head; 6" straight section between head and flanged connection to body; 1- 3" Sch 20 flanged tar outlet connection; 4 lifting lugs

Materials of Construction

All material in contact with the product gas 304 SS (lower flange may be carbon steel with 304 SS clad for surfaces in contact with the product gas, but must allow 40 mils corrosion loss). All material in contact with water or steam only carbon steel (but ASME code grade materials only). All tubing must be seamless.

Ancillary Equipment

- heated drain line to tar collection drum (3" Sch 20, 304 SS)
- rupture disk (10 psig) and pressure relief line (25' long x 8" Sch 20, 304 SS)
- material test coupons located in condensing gas region

Dissolving Tank

See Figure E-1.

Code Requirements and Notes

Same requirements as gasifier vessel except liquid piping and pumps have temperature not to exceed 250°F.

Conditions

Solids flow	350-560 lb/hr
Temperature	1150-1250°F

Liquid flow (in & out)	3.5-6.0 gpm
Liquid temp	140-200°F
Insolubles (sand)	0-250 lb/hr (intermittent)
Vent steam	50-90 lb/hr

Specifications

- 500 gal capacity tank with conical bottom for solids removal
- sealed tank cover with penetrations for solid and liquid transport lines plus steam vent line

Materials of Construction

all materials 304 SS

Ancillary Equipment

- 1 agitator
- 1 vent line
- 1 100 gal surge tank
- 2 liquid transfer pumps
- 1 sediment slurry discharge pump (manual control)
- 1 tank level control
- 1 dilution flow control by solution conductivity
- 1 outlet temperature measurement

Product Gas Scrubber

See Figure E-1.

The Product Gas Scrubber must accomplish two functions:

1. Cool the product gas past the acid dew point (~167°F) to 125°F. Acidic condensate with dissolved H₂S will be neutralized with alkaline weak wash from the pulp mill and sent to the dissolving tank.

2. Remove H₂S from the product gas. Normally 10% NaOH solution from the pulp mill will be used for scrubbing H₂S from the gas stream, but carbonate solution from the dissolving tank will also be evaluated in a limited number of tests.

Two separate components have been shown in Figure E.1, a **Quench Cooler** and a **Gas Scrubber**. Critical process conditions and design specifications for each component are given below. The detailed design of these components is to be determined by the supplier.

Quench Cooler

Code Requirements and Notes

- Code requirements must consider that product is a toxic and flammable gas. All gas-containing piping and vessels must be designed for 50 psig continuous service. Normal operating pressure of these back-end components will be less than 5 psig.
- Minimum design temperature is 0°F. Design life is 2 years; operating life is 4000 hours with 200 thermal cycles.
- The supplier will develop detailed design drawings for review and approval by B&W before materials procurement and release for fabrication.

Nominal Conditions

Inlet gas flow	3420 lb/hr
Inlet gas temp	220°F
Inlet gas pressure	~1 psig
Condensate flow	~950 lb/hr (exact amount t.b.d. by supplier)
Cooling water temp in	80°F
Cooling water temp out	t.b.d. by supplier

Cooling water flow t.b.d. by supplier

Desired outlet gas temp 125°F

Specifications

Gas inlet - 8" Sch 20 pipe flanged connection

Equipment type (e.g., venturi, spray tower) and dimensions t.b.d. by supplier

Materials of Construction

High-end FRP (e.g., Centicast Plus RB-2530) or PVDF (Kynar) have appropriate thermal and chemical resistance for this application.

Carbon steel pipe lined with either polypropylene or PTFE (Teflon) can be used for liquid transfer lines.

Condensing acid gases and temperatures exceeding 200°F preclude the use of CPVC vessels or piping.

Ancillary Equipment

- 1 foul condensate circulation pump
- 1 indirect heat exchanger (sized by supplier)
- spray nozzles (number and arrangement t.b.d. by supplier)
- level control in quench vessel
- exit gas temperature control
- measurement of temperature and pressure in quench vessel

Gas Scrubber

Code Requirements and Notes

Same as Quench Cooler.

Conditions

Inlet gas flow	~2470 lb/hr (depending on condensate yield)
Inlet gas temp	125°F
Inlet gas pressure	~1 psig
Inlet gas comp (wvb)	28% H ₂ , 1% H ₂ S, 5% CO, 16% CO ₂ , 1% CH ₄ , 13% H ₂ O, 36% N ₂
Scrubbant flow	~750 lb/hr (exact amount t.b.d. by supplier)
Scrubbant temp	80°F
Scrubbant conc	12 wt% NaOH
Green liquor flow	~750 lb/hr (exact amount t.b.d. by supplier)
Green liquor temp	~125°F
Green liquor conc	10 wt% Na ₂ S; 3 wt% Na ₂ CO ₃
Outlet gas H ₂ S (dvb)	< 500 ppm

Specifications

Vessel type (e.g., venturi, packed tower), size and connections t.b.d. by supplier

Materials of Construction

High-end FRP (Centicast Plus RB-2530) or PVDF (Kynar) have appropriate thermal and chemical resistance for this application.

Carbon steel pipe lined with either polypropylene or PTFE (Teflon) can be used for liquid transfer lines.

Ancillary Equipment

- 1 green liquor circulation pump
- spray nozzles (number and arrangement t.b.d. by supplier)
- tower packing (if required)
- outlet gas mist eliminator
- level control in scrubber vessel
- thermowell and sampling line in green liquor outlet pipe
- thermowell, pressure tap and gas sampling line in scrubber exit line

Product Gas Flare

See Figure E-1.

The Product Gas Flare must completely burn the fuel gas species in the product gas and fully oxidize the H₂S to SO₂. Critical process conditions for this component are given below. The detailed design is to be determined by the supplier.

Code Requirements and Notes

- Code requirements must consider that product is a toxic and flammable gas.
- Minimum design metal temperature is 0°F. Design life is 2 years; operating life is 4000 hours with 200 thermal cycles.

Conditions

Inlet gas flow	2420 lb/hr
Inlet gas temp	125°F
Inlet gas comp. (wvb)	28% H ₂ , 5% CO, 16.5% CO ₂ , 1% CH ₄ , 13% H ₂ O, 36.5% N ₂
Inlet gas H ₂ S (dvb)	< 500 ppm
Inlet gas HHV	133 Btu/DSCF
Natural gas pilot	~10% of heat input (exact amount t.b.d. by supplier)
Excess air	300% (amount required for thermal oxidation t.b.d. by supplier)
Exhaust SO ₂ (dvb)	< 100 ppm

Specifications

Flare size and connections t.b.d. by supplier

Materials of Construction

Carbon steel

Ancillary Equipment

- 1 air blower
- 1 natural gas pilot
- 1 flash arrestor
- 1 condensate drain line (if required, polypropylene or PTFE lined carbon steel pipe is recommended for corrosion resistance)
- 1 flare stack
- 1 ignition source and burner control
- 1 thermowells and 1 flow orifices in air supply line
- 1 thermowells and 1 flow orifices in natural gas supply line

PART 2 - PROCESS FLOW DIAGRAM AND HEAT AND MATERIAL BALANCE

A process flow diagram and the associated heat and material balance are provided in Attachment E.1.

Supplemental calculations needed to develop the functional specifications for the equipment are provided in Attachment E.2. These were to determine the following:

- o Superficial gas velocity in bed
- o Fluidization of bed solids
- o Gasifier vessel freeboard region sizing
- o Liquor nozzle sizing
- o Sizing for major piping

An instrumentation diagram is included in Attachment 1 of Appendix D. This diagram shows the location of the principal measurements that will be required for the pilot-scale gasifier. A more detailed process and instrumentation diagram could be prepared in the future when the process development at the bench-scale is completed.

The performance of the fluidized bed gasifier is based on the performance criteria from Grace and Timmer (1995) used to design the bench scale unit. The performance will be revised when confirming data is available from the bench scale unit.

The PFD is based on solids flow of 1000 lb/hr to the system. Both a flash tank and a direct steam heater have been included so that we can handle a range of feed liquor temperatures. Provisions have also been included (but not used on the attached PFD) to add flue gas from the gas fired air preheater to the air/steam stream entering the preheater. This will allow operation with reduced oxygen levels. Provisions have also been included but not used to cool the spent bed material before it enters the dissolving tank. This would decrease the amount of evaporation occurring in the tank.

There is no attempt at an integrated sulfur recovery system in this PFD. Foul condensate from the quench cooler is sent to the spent bed material dissolving tank simply to get rid of it. The alkali in the weak wash will neutralize the dissolved H₂S. Fresh make up caustic is used in the H₂S absorption tower to clean up the product gas before it is flared. The product liquors from both the dissolving tank and the tower are sent to the host mill's green liquor system. Note that only 90% absorption is required to get down to about 200 ppm SO₂ at the stack. The flare is assumed to run with very high excess air since there is no attempt at heat recovery.

REFERENCE

Grace, T. M., Timmer, W. M., "A Comparison of Alternative Black Liquor Recovery Technologies," Preprints 1995 International Chemical Recovery Conference, Toronto, Ontario, Canada, 1995.

ATTACHMENT E.1

PRELIMINARY PSG PROCESS FLOW DIAGRAM AND
MATERIAL AND ENERGY BALANCE

H2O 0.0 17.93 0.0

 0.0 100.00

N2 = 78.09 %
 Ar = 0.93 %
 CO2 = 0.03 %

Temp = 298 F

Steam----->

H2O = 1000.0
 Total Enthalpy 1.138 MBtu/hr

EA/S = 1.5000 #/#solids= 34.57 %TA
 H/S = 1.0000 #/#solids
 FGR/S = 0.00 #/#solids

V

HEAT from Nat Gas Comb.

----->| Air Preheater |
 Enth. = 0.842 MBtu

Heat Loss= 0.000 MBtu/hr
 0.0 % of input
 to bed

t = 1178 °F
 p = 1.16 Atm
 Mol wt = 28.74 #/#mol

V

HEAT from Combustor

----->| Fluidized Bed |
 Enth. = 0.000 MBtu +-->| Gasifier |

Bed Diam = 3.271 Ft
 Bed Area = 8.402 Ft²
 Mol wt = 20.98 #/#mol
 Rho{g} = 0.0184 #/ft³
 U{g} = 5.90 ft/sec

t{bed} = 1250 °F
 P{bed} = 1.04 Atm

Heat Loss = 0.060 MBtu/hr
 = 0.7 % of bed input

V

"Dried "
 Solids 76.74 %
 H/S = 1.3031

Equivalent
 B&W Btu-tons/day = 7.3 V

-----<| Freeboard Dryer |-----<

t = 1125.0 °F
 p = 1.04 Atm

FB Diam = 4.667 Ft
 Area = 17.104 Ft²
 Mol wt = 20.85 #/#mol
 Rho = 0.0187 #/ft³
 gas flow = 3045 ft³/min
 U{g} = 2.97 ft/sec

Cyc. Inlet vel. = 50 ft/sec
 Inlet Area 1.015 ft²
 Cyc Diam 34.19 in

V

K{r1} = 1.8144

----->| Dust Separator |
 | (& Bed Drain) |

Hot Product Gas

Spent Bed

Na2CO3	= 395.73	77.78 %
Na2S	34.44	6.77
Na2SO4	23.21	4.56
NaCl	3.63	0.71
NaOH	0.00	0.00
K2CO3	45.78	9.00
KCl	0.00	0.00
Carbon	6.02	1.18

Fuel Gas = 2170.0
 H2S = 35.1
 H2O (v) = 1214.7

 3419.8

HHV = 137.1 Btu/DSCF
 2278.4 (Btu/#dry)
 Enthalpy = 7.675 MBtu/hr

H2S = 63.0 % of input S

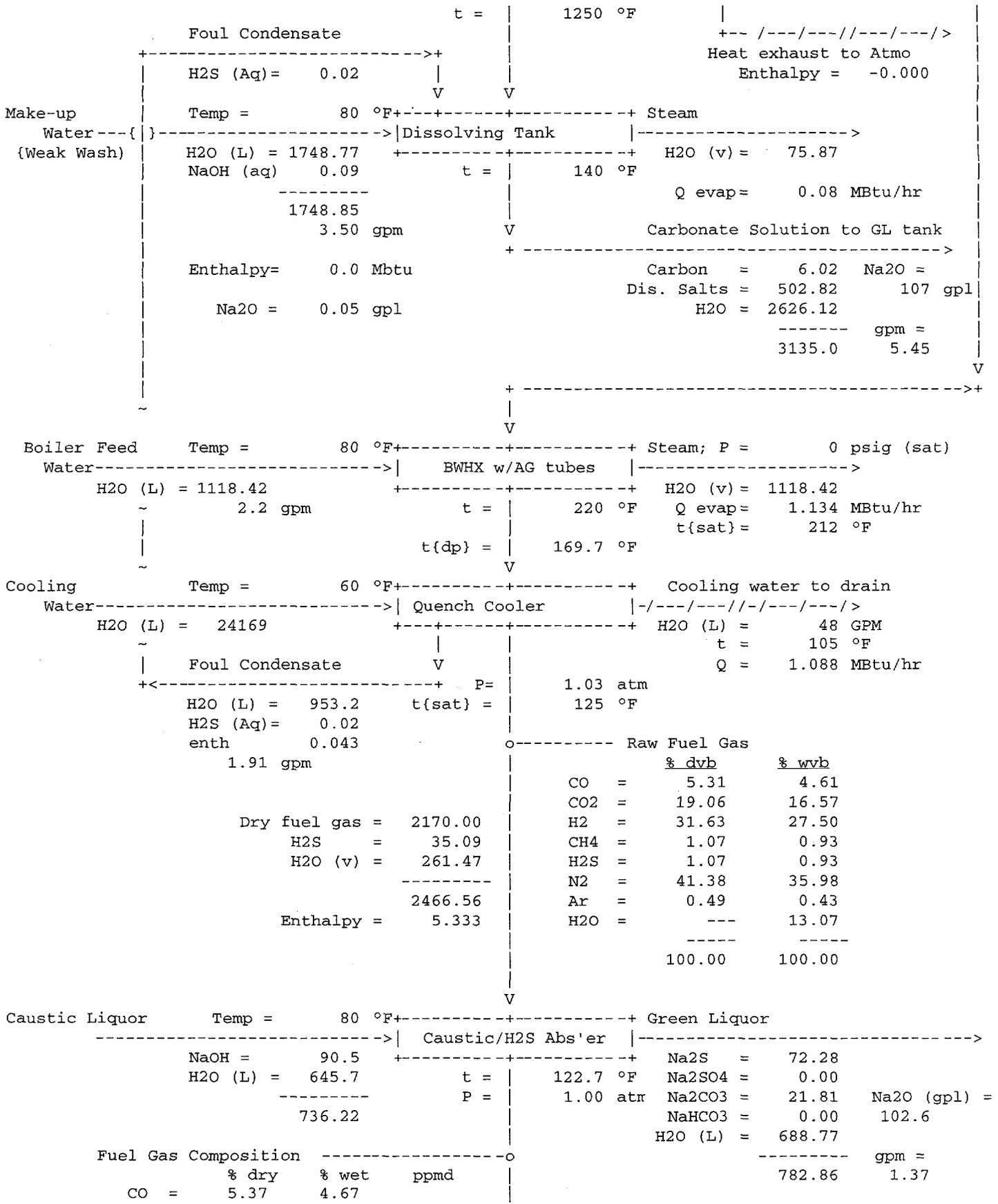
Carbon = 1.7 % of input 508.80
 Enthalpy 0.426 MBtu/hr

t = V 1250.0 °F

----->| Spent Bed Cooler |

Heat input to Comb. Air

Enthalpy = 0.000



CO2 =	19.07	16.60
H2 =	32.01	27.87
CH4 =	1.08	0.94
H2S =	0.11	0.09
N2 =	41.87	36.46
Ar =	0.50	0.43
H2O =	----	12.94
	-----	-----
	100.00	100.00
Mol wt	22.68	0.00

1078

o-----Cleaned Fuel Gas	
Dry gas =	2164.45
H2O (v) =	255.45

	2419.90
HHV =	133 Btu/DSCF
LHV =	115 Btu/DSCF

V

Combustion	Temp =	80 °F	-----+-----	Flue gas to Stack
Air	----->		Gas Burner - Flare	----->
Dry Air =	12469.7		(W/nat Gas pilot)	
H2O (v) =	163.8			

	12633.5		(t{AFT} = 1094 °F)	CO2 = 1133.1 5.43
Enth. =	MBtu		excess air = 300 %	N2 = 10538.6 79.34
			Stack SO2 = 217 ppr	O2 = 2164.6 14.27
Natural Gas	t = 80 °F			SO2 = 6.6 0.02 217
				Ar = 178.9 0.94
				H2O = 1052.9 ---

				15074.7 100.0
				3187 scfm

MATERIAL BALANCE

Pounds/Hour

IN		OUT	
Black Liquor solids	= 1000.0	Flash tank Vapor	= 0.0
H2O in Liquor	= 428.6	Green Liquor	= 782.9
Steam to direct L.Heater	= 0.0	Combustor Stack Gas	= 15074.7
Steam to Fluid bed	= 1000.0	Carbonate liquor	= 3135.0
Air To Gasifier Bed	= 1500.0	Dis. Vent H2O	= 75.9
Oxygen to Gasifier bed	= 0.0		
FGR to Gasifier bed	= 0.0		
Air to Gas Burner	= 12633.5		
Natural gas to pilot	= 21.3		-----
Make up H2O	= 1748.8		19068.4
Caustic solution	= 736.2	Error (in- out) =	-0.0
	-----	% of input =	-0.000
	19068.4	Error (#/100# BLS)	-0.003

Grace Parameters:

=====

Sulfur reduction =	90 %
Reduced S as H2S =	70 %
Carbon Conversion=	98 %
CH4, % of gas C =	6.0 - 0.06 * % Stoich air

Gasifier Operations

=====

Gasifier Exit T =	1250.0 °F
Gasifier Exit P =	1.04 atm

Process: **Pilot Scale Gasifier**

Run ID: **Design Basis**

Solids : 70.00 % to reactor
A/S : 1.500 # Air/# solids
H/S : 1.000 # H2O/# solids

Version: 1.2

ENTHALPY BALANCE

Million Btu/Hour

IN		OUT		
Black Liquor	= 6.012	Reactor Heat loss	0.060	0
Steam to Fluid bed	= 1.138	Steam from HRSG's	1.134	15
Air To Gasifier Bed	= 0.007	Low Level heat	1.168	16
Oxygen to Gasifier bed	= 0.000	Stack Loss	ERR	:
Make up H2O	= 0.000	Green Liquor	ERR	:
Air to Comb	= 0.000			
	-----		-----	-----
	7.157		ERR	:
		Error (in- out) =	ERR	
		% of input =	ERR	

SUMMARY OF DESIGN CONDITIONS:

Process Features: {No = 0, Yes = 1}

=====
 Air Blown Gasify : 1
 H2O Blown Gasify : 1
 O2 enrichment (?): 0
 FGR /Vitiated air: 0
 OverBed Spray (?): 1
 Air Preheat (?): 1
 Sp. Bed Cooler(?): 0

Black Liquor Properties

=====
 As-Rec'ed Solids: 70.0 %
 As-Rec'ed Temp. : 238.0 °F
 As-Rec'ed Press : 50.0 psig
 As-fired Temp. : 238.0 °F

Gasifier Operating Conditions

=====
 Gasifier Exit T : 1250.0 F
 Gasifier Exit P : 1.04 atm
 Air/Solids Ratio : 1.5000 #/# (if used)
 Air Temp. F : 100.0 °F to PreHt
 H2O/Solids Ratio : 1.0000 #/# (if used)
 Steam Temp., F : 298.0 °F to PreHt
 Steam Enthalpy : 1179.1 Btu/# @ T
 Freeboard Evap : 29.3 % of H2O in Liq.
 Air Preheat Temp : 1178.0 °F (if used)
 Bed Cooler exit T: 450.0 °F (if used)
 O2 Content of EA : 40.0 %dwb(if used)
 H2S Absorp eff : 90 %
 FGR /Vitiated air: 1.43 #/# (if used)

Heat Cap. Bed So. = 0.25 Btu/#°F
 Theoretical air = 4.2825 #/# solids
 HHV, Carbon = 14093 Btu/lb

"K" values

=====
 S.D.E. 1000 °F
 Adi.F. 1000 °F
 SB. exit 100 °F
 K{r} 1.8144 --

-> Atmo air supplies: 100.00 mol% of dry "EA"
 FGR gas Temp : 750 °F (if used)

Miscellaneous Data:

=====
 Comb Ex Air = 300 %
 BWHX EG Temp = 220 °F
 Feed water temp= 80 °F
 BWHX Steam Press= 0 {psig,sat}
 Enthlpy = 1150 Btu/lb
 Quench exit Temp = 125 °F

Carbonate/Caustic Liquor Systems:

=====
 Carb liquorl Conc. = 107 gpl as Na2O
 density of G.L. = 1150 gpl
 liquor exit temp = 140 °F
 Excess Caustic = 10 %
 Caustic Conc. = 110 gpl as Na2O
 density of Caus. = 1154 gpl
 Caustic liquor temp 80 °F

Fluid bed reactor products:

				#mole/C# solids
CO	0.5124	Na2CO3	0.3734	
CO2	1.8401	Na2S	0.0441	
H2	3.0546	Na2SO4	0.0163	
CH4	0.1031	NaCl	0.0062	
N2	3.9962	NaOH	0.0000	
H2S	0.1030	K2CO3	0.0331	
H2O	6.0459	KCl	0.0000	
Ar	0.0475	C	0.0501	

ATTACHMENT E.2

SUPPLEMENTAL CALCULATIONS SUPPORTING PSG
FUNCTIONAL SPECIFICATIONS

MATERIAL BALANCE INPUTS

Case study	100%	
Solids flow	0.126 kg/s	1000.0 lb/hr
Solids content fired		70 %
HHV of solids	13984 kJ/kg	6012 Btu/lb
Est. solids reaching bed		76.74 %
Total heat input	1.76 MW	6.01 MBtu/hr
Liquor flow	0.180 kg/s	1428.6 lb/hr
Water evap'd in flight	0.016 kg/s	125.5 lb/hr
Gases from bed	0.100 kg/s	794.1 lb/hr
Total air flow	0.189 kg/s	1500.0 lb/hr
Vitiated air flow	0.000 kg/s	0.0 lb/hr
Total steam flow	0.126 kg/s	1000.0 lb/hr
Bed zone gas flow	0.415 kg/s	3294.1 lb/hr
Est. wet flue gas flow	0.431 kg/s	3419.6 lb/hr

SUPERFICIAL GAS VELOCITY IN BED

Bed diameter	0.997 m	3.27 ft
Bed plan area	0.781 m ²	8.40 ft ²
Av. gas temp. in bed	949.9 K	1250.0 F
Bulk gas mol. wt.	20.98 kg/kmol	20.98 lb/mol
Pressure in bed		1.10 atm
Gas density	0.2961 kg/m ³	0.0185 lb/ft ³
Gas viscosity		0.0927 lb/ft-hr
Superficial gas vel. in bed	1.80 m/s	5.89 ft/s
Fluidized bed height	1.83 m	6.00 ft
Gas res. time in bed	1.02 s	

FLUIDIZATION OF BED SOLIDS

Bed material	Na ₂ CO ₃	
Particle size	710 μm	
Particle density	kg/m ³	137.3 lb/ft ³
Particle Reynolds no.		9.85 -O-
Archimedes no.		1557 -O-
Velocity no.		0.85 -O-
Predicted Fluidization no.		9.59 -O-
Calc. Fluidization no.		9.62 -O-

FREEBOARD CONDITIONS

Freeboard diameter	1.422 m	4.67 ft
Freeboard area	1.589 m ²	17.10 ft ²
Av. gas temp in FB	880.4 K	1125.0 F
Bulk gas mol. wt.	20.85 kg/kmol	20.85 lb/mol
Pressure in freeboard		1.04 atm
Gas density	0.3001 kg/m ³	0.0187 lb/ft ³
Superficial gas vel. in FB	0.90 m/s	2.96 ft/s

LIQUOR SPRAY INPUT VARIABLES

Liquor firing temp.	114.4 C	238.0 F
Liq. density	1365 kg/m ³	85.2 lb/ft ³
Liq. viscosity	0.11 kg/m/s	110 cP
Nozzle type	MP 156 NN	
Nozzle diameter	3.96 mm	0.156 in
Spreading angle	0.52 rad	30 deg
Number of guns	1	
Liquor flow (nozzle)	0.180 kg/s	1429 lb/hr
Vol. liquor flow (nozzle)		2.09 gpm
Equivalent water flow		2.44 gpm H2O

CALCULATE LIQUOR VAPOR PRESSURE

Norm. boiling point	115.9 C	240.6 F
del H(S)/R	5146.989	
C(S)	13.11125	
P(sat)	85679 Pa	12.43 psia

CALCULATE MEDIAN DROP SIZE & NOZZLE PRESSURE DROP

Flashing temp.	125.9 C	258.6 F
Flashing factor	1	
Drop size adj. factor	2.28	
Nozzle liq. vel. (Vn)	10.69 m/s	35.07 ft/s
Mass med. drop size	0.50 mm	
Adjusted d _{mm}	1.14 mm	
Norm. std. dev.	0.263	
Re, liquor	526	

CALCULATE MAXIMUM DROP SIZE FOR ENTRAINMENT

Carryover (probability)	1.5 %	
Variance factor	-2.170	
Max. drop size for ent.	0.21 mm	

CALCULATE SETTLING VELOCITY & CHECK CHOICE OF NOZZLE

Gas viscosity	3.804E-05 kg/m/s	0.0380 cP
Settling vel. for max. drop	0.95 m/s	3.11 ft/s
Re,p	1.6	

Nozzle Size? Acceptable

CALCULATE CRITICAL DROP SIZE FOR NOZZLE HEIGHT

Critical solids	76.74 %
Probability value	50.75 %
Variance factor	0.019
Crit. drop size to dry	1.15 mm

CALCULATE RELATIVE DROP VELOCITIES

Spray angle (from horiz)	90 deg	
Velocity factor for spray	100 %	
Initial liquor vel.	10.69 m/s	35.07 ft/s
Initial slip velocity	12.49 m/s	40.97 ft/s
Final slip velocity	6.59 m/s	21.62 ft/s
Re,p (initial)	112	
Re,p (final)	59	
Init. downward vel.	10.69 m/s	35.07 ft/s
Final downward vel.	4.79 m/s	15.73 ft/s

CALCULATE DROP DRYING TIME

Mean bulk temp. of drop	390.9 K	243.9 F
Hv @ Tb	2206.6 kJ/kg	948.7 Btu/lb
Film temperature	670.4 K	747.0 F
k _{gas} @ Tf	0.0497 W/m/K	0.0287 btu/h/ft/F
Emissivity	1	
Gr	0.8	
Conv. H.T. coeff. (init.)	373 W/m ² /k	65.64 btu/h/ft ² /F
Conv. H.T. coeff. (fin..)	286 W/m ² /k	50.45 btu/h/ft ² /F
Dry. time to crit. solids (init.)	0.20 s	
Dry. time to crit. solids (fin.)	0.25 s	

ESTIMATE GUN HEIGHT

Vert. dist. to dry (initial)	2.15 m	7.04 ft
Vert. dist. to dry (final)	1.19 m	3.90 ft
Min. dist. to dry	1.19 m	3.90 ft
Max. dist. to dry	2.15 m	7.04 ft
Predicted gun elevation	1.83 m	6.00 ft
Allowed gun elevation	1.83 m	6.00 ft
Dia. spray at bed	0.98 m	3.21 ft
Spray coverage?	Okay	

DUCT CALCULATIONS

Inlet pipe i.d.	0.260 m	10.25 in
Flow area	0.053 m ²	0.57 ft ²
Gas temp at exit	909.9 K	1178.0 F
Bulk gas mol. wt.	28.74 kg/kmol	28.74 lb/mol
Pressure		1.18 atm
Gas density	0.4542 kg/m ³	0.0284 lb/ft ³
Gas viscosity	3.73E-05 kg/m-s	0.0902 lb/ft-hr
Superficial gas vel. in exit	13.03 m/s	42.75 ft/s
Reynolds number	41333	41335

Exit pipe i.d.	0.394 m	15.50 in
Injector pipe o.d.	0.060 m	2.38 in
Flow area	0.119 m ²	1.28 ft ²
Gas temp at exit	880.4 K	1125.0 F
Bulk gas mol. wt.	20.85 kg/kmol	20.85 lb/mol
Pressure		1.04 atm
Gas density	0.3001 kg/m ³	0.0187 lb/ft ³
Gas viscosity	3.65E-05 kg/m-s	0.0883 lb/ft-hr
Superficial gas vel. in exit	12.08 m/s	39.63 ft/s
Reynolds number	33121	33122

Crossover pipe i.d.	0.394 m	15.50 in
Flow area	0.122 m ²	1.31 ft ²
Gas temp at exit	880.4 K	1125.0 F
Bulk gas mol. wt.	20.85 kg/kmol	20.85 lb/mol
Pressure		1.04 atm
Gas density	0.3001 kg/m ³	0.0187 lb/ft ³
Gas viscosity	3.65E-05 kg/m-s	0.0883 lb/ft-hr
Gas mass flow	0.431 kg/s	3419.6 lb/hr
Gas volumetric flow	1.44 m ³ /s	3042.0 ACFM
Superficial gas vel. in exit	11.79 m/s	38.69 ft/s
Reynolds number	38196	38191

Estimated carryover	0.00189 kg/s	15 lb/hr
Dust loading (actual)	0.001316 kg/m ³	0.575289 gr/ACF

SUMMARY OF DESIGN CALCULATIONS

100%

1. Given heat input 6.01 MBtu/hr

2. Liquor solids 70 %
Liquor flow (nozzle) 1429 lb/hr

3. Bed diameter 3.27 ft
Bed superficial gas vel. 5.89 ft/s
Bed particle size 710 μm

4. Freeboard dia 4.67 ft
FB superficial gas vel. 2.96 ft/s

5. Nozzle type MP 156 NN
Nozzle size 3.9624 mm
Liquor temp. 238 F
Liq. density 85.2 lb/ft³
Liq. viscosity 110 cP
Flashing conditions? No

6. Mass med. drop size 1.14 mm
Std. dev. 0.28

7. Assumed carryover 1.5 % lost

8. Max. drop size for ent. 0.21 mm
Settling velocity (u_t) 3.11 ft/s

9. Bulk av. vel. - u_t ft/s
(if >0, then add'l carryover)

10. Solids to bed (ass'n) 76.74 %
Med. drop size to dry 1.153 mm
Drying time (min) 0.20 s
Drying time (max) 0.25 s

11. Gun ht. above bed ft
Minimum est. ht. 3.9 ft
Maximum est. ht. 7.0 ft

12. Spray coverage ft

APPENDIX F

MARKET AND ECONOMIC EVALUATION

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ATTACHMENT F.1 - DETAILED RESULTS OF ECONOMIC ANALYSIS

- Exhibit F.1 - Fuel Price Sensitivity - Small Recovery Boiler / BLG Producing Gas
- Exhibit F.2 - Fuel Cost Sensitivity - Small Recovery Boiler / BLGCC Options
- Exhibit F.3 - Capital and Initial Fixed Cost Sensitivity - Small Boiler / BLG Producing Gas
- Exhibit F.4 - Capital and Initial Fixed Cost Sensitivity - Small Recovery Boiler / BLGCC Options
- Exhibit F.5 - O&M Cost Sensitivity - Small Recovery Boiler / BLG Producing Gas
- Exhibit F.6 - O&M Cost Sensitivity - Small Recovery Boiler / BLGCC Options
- Exhibit F.7 - Gasifier Performance Sensitivity - Small Recovery Boiler / BLG Producing Gas
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F.1 PULP AND PAPER RECOVERY EQUIPMENT MARKET

F.1.1 Overall Market Demand

The market for black liquor recovery boilers or alternate black liquor processing capability (such as gasification) is driven largely by growth in the production of virgin fiber wood pulp production.

Over the next 15 years, the pulp and paper market will become increasingly international in nature. The North American share of the world's of virgin fiber wood pulp production is expected to fall from about 54% in 1995 to about 42% by 2010, while still growing at 0.8% per year. The annual worldwide wood pulp capacity is expected to grow from 154 to 217 million metric tons over the next 15 years or a 2.3% average growth rate per year. The resulting black liquor PR boiler capacity additions outside North America will approach the size of the current U.S. installed base. During this period, the North American pulp industry will increase its annual kraft pulp capacity by approximately nine million metric tons. The North American pulp industry is currently recovering from its last down-cycle as capacity utilization and prices rise, but a variety of factors will dampen the near term rate of capacity additions.

F.1.2 Black Liquor Processing Needs

Incremental black liquor processing needs for worldwide wood pulp production between 1995 and 2010 are approximately 540 million #ds/day or the equivalent of 170 large commercial (3,000,000 #ds/day) units (10 or 11-/yr). Additional capacity will also likely be installed as the older units in North America and Europe (>40 years old) are replaced. These will likely take place in association with mill capacity expansions and upgrades. Complying with the pending U.S. environmental regulations may affect the timing of these capacity replacements.

The total capacity, new plus replacement, during the period will average 12-20 units/year worldwide (based on the commercial unit size stated previously). This world total will be divided regionally:

- North American - 4 to 5 units/yr
- South America, Pacific Rim (excluding Japan), Eastern Europe, and CIS - 5 to 8 units/yr
- Europe, Scandinavia, and all other areas - 4 to 7 units/yr

Factors that will restrain the virgin fiber pulp production growth in North America include:

- o Increased recycling
- o Higher cost virgin fiber (at least in the U.S.)
- o Tighter environmental regulations
- o Remaining corporate debt from the last up-cycle of recovery capacity installation
- o Major paper market growth outside North America (especially in the Far East)

The “paperless” workplace will also have an impact but this is being extended in time longer than originally anticipated by many forecasters.

The exact makeup of the capacity additions will depend upon location. Outside North America and Europe, additions will likely take the form of large new units, as new areas are opened and economies of scale dictate larger unit size. In North America and Europe, the mature markets and existing capacity will dictate three strategies:

- 1) expansion of existing units
- 2) addition of small incremental capacity (900,000 #ds/day or less) units
- 3) a few new large (3,000,000 #ds/day or more) units

Based upon current U.S. and Canadian market estimates, the large unit additions may represent approximately one large commercial unit per year. The balance of the market will likely be split between capacity upgrades at existing units and smaller unit additions. This would result in a North American market for 3-7 small (900,000 #ds/day or less) units per year.

F.1.3 Impact of Black Liquor Gasification

Depending upon the cost competitiveness of the final proven gasification technology, black liquor gasification could potentially replace conventional small process recovery boilers in the incremental capacity category and then eventually displace larger units.

A reasonable scenario leading to full commercialization of black liquor gasification at this time is:

- o Startup of the first commercial black liquor gasifier (high-temperature process) in North America (capacity 734,000 #ds/day) is scheduled to occur within the year.
- o Installation of additional small commercial prototype units by 2000 (450,000 to 900,000 #ds/day).
- o Incremental capacity competitive with conventional technology addressed between 2001 - 2005 (unit size 450,000 to 900,000 #ds/day).
- o Installation of several larger units (960,000 to 1,600,000 #ds/day) to become operational by 2005.
- o Sale of the first large scale commercial unit (3,000,000 #ds/day) between 2005 and 2010.

In order for this scenario to occur, black liquor gasification must be successful (cost, performance, and safety) and reliable operation at each step.

F.1.4 Market Constraints

Incremental capacity additions using small units are the anticipated first steps in commercialization of black liquor gasification. These first units will be in competition with other options, including upgrade of existing recovery boilers to increase capacity and installation of small stand alone recovery boilers. Later, combined cycle units will be used. In the economic analysis these options are compared for incremental capacity additions. Some important market constraints are associated with each option.

The small recovery boiler and gasifier producing a fuel gas are more limited than combined cycle options since both options essentially displace purchased fuel. However, significant potential exists to displace purchased fuel within the pulp and paper industry. Note that the quantity of purchased fuel the U.S. pulp and paper industry used in 1992 was approximately equal to the quantity reported for self-generated energy produced by the industry from spent liquor (i.e., recovery boiler operations) [Reference 1]. The option of upgrading an existing recovery boiler is more limited since the mill must have a unit in place that can be upgraded to provide the desired capacity increase. Over time, fewer mills will have this option available, due to boilers already being upgraded and the age of some units.

Black liquor gasification combined cycle technology offers the potential for application without depending on downstream equipment to use low-heating-value fuel gas or a significant quantity of steam. This could expand the applicability of the gasifier for the incremental capacity increases.

Black liquor gasification requires cleanup of a product gas that contains high amounts of CO_2 and H_2S . The gas cleanup can lead to formation of NaHCO_3 in the resulting liquor. The bicarbonate requires twice the lime to causticize. Higher lime requirements per pound of liquor-processed than

conventional technology could result. The additional lime requirement is estimated to be approximately 25% higher for high-temperature gasifiers and 50% higher for low-temperature gasifiers. This may require lime kiln capacity increases to accommodate black liquor gasification versus a conventional approach. Thus, this is a resisting force for market penetration of gasifiers providing incremental capacity increases.

As indicated in the commercialization scenario described previously, black liquor gasification has significant potential beyond incremental capacity additions. Many mills have older, smaller recovery boilers that will be retired, thus yielding a market segment beyond the incremental capacity market noted above. This market segment could apply for each technology compared here since the downstream capability to use the product steam, electricity, or gas may exist, depending on specific mill conditions. Mills that are retiring units may favor conventional technology, particularly where installation or upgrade of a single larger unit is feasible, thus achieving substantial economies of scale. Black liquor gasification has the potential to be competitive for large-scale replacement markets after it has been further developed and integrated with gas turbine combined cycles.

F.1.5 Additional Benefits

Environmental Benefits - The potential environmental benefit of low temperature gasification of black liquor is the reduced emission of sulfur dioxide, hydrochloric acid, and hydrogen sulfide compared with the conventional black liquor recovery process. In addition, the increased overall thermal efficiency that will be possible when combined cycle applications are developed will reduce the use of fossil fuels. Thus, combined cycle applications would also reduce the industry dependence on purchased fuel and the net (or overall) CO₂ production.

- Sulfur Dioxide - The emission of SO₂ from a black liquor gasification process will be significantly lower than from a recovery boiler. In the gasification process, the sulfur in the liquor is converted primarily to hydrogen sulfide (H₂S) in the product fuel gas with some small

amounts remaining in the bed solids as salts. The H_2S in the gas must be recovered before the fuel gas can be burned since the sulfur is an important part of the chemical cycle. Also, the removal efficiency must be high if the fuel is to be burned in a gas turbine in future applications because high sulfur would be detrimental to the turbine. If the fuel gas is burned in a power boiler, lower efficiency can be tolerated. We estimate that the H_2S absorber can easily achieve a level of 50 ppm in the exhaust gas, which is acceptable for combustion in a power boiler. Combustion of this gas will convert the sulfur to sulfur dioxide, but the concentration will be very low because of the dilution effect of the combustion air. We estimate the total SO_2 release resulting for this scenario will be about 1.1 lb/ton of pulp.

- Oxides of Nitrogen - The emission of NO_x from a black liquor gasification process may also be lower than from a recovery boiler, but this is more difficult to quantify. Essential no formation of NO_x will occur in the gasification reactor because of the low temperature and highly reducing conditions. Some ammonia and other NO_x precursors may form from nitrogen in the liquor, but the equilibrium concentrations are predicted to be very low. The most significant source of NO_x will be from combustion of the fuel gas. Gas burner design and NO_x control strategies for natural gas firing can be used to limit NO_x concentration to about 100 ppm. We estimate that this will give a total NO_x release that is about the same as a recovery boiler.
- Chlorine - Thermodynamics predicts that the concentration of chlorine-containing vapors in the fuel gas should be zero, and that the chlorine will all appear as sodium or potassium chloride in the solid phase. On this basis it is estimated that the HCl emissions from the gasification process will be near zero.
- Hydrogen Sulfide - The emission of H_2S from a black liquor gasification process will be nearly zero. It will only be emitted if incomplete combustion of the fuel gas stream occurs either in the gasification air heaters or in the product gas burner.

- Particulate - Solid emissions could result from burning a gasification process fuel gas that contained aerosols, e.g., fine fume from the gasification reactor or mist carryover from the H₂S absorption process. Because of the lower temperature in the gasification reactor, inorganic fume formation would be much less than in a conventional recovery boiler. It is estimated that total solids emissions from the fuel gas burner stack can be less than or equal to the level equivalent to a recovery boiler stack.

Pulping Improvements - Extended delignification is an approach to remove more lignin in the digester vessel by varying the strength of chemicals applied to the wood chips during the "cook." The next generation of extended delignification technology will use two streams of pulping chemicals, one rich in NaOH and the other rich in Na₂S. Producing "split-sulfidity" pulping liquor is possible in several ways. However, since most of the sulfur is converted to H₂S during gasification and must be scrubbed from the product gas by caustic solution, this simplifies production of split-sulfidity pulping liquors. Thus, while split-sulfidity liquor could be produced by other means, black liquor gasification offers a means to achieve this end and thus has an additional market advantage (Reference 2).

F.2 ECONOMIC ANALYSIS

F.2.1 Approach

A simplified analysis was completed to compare the relative economic benefit of black liquor gasification versus conventional recovery boiler technology for adding incremental recovery capacity to a pulp mill. A capacity increase of 150 B&W Tons (493,000 lbs. dry solids /day assuming liquor higher heating value of 6020 Btu/lb dry solids) was used for the comparison cases. The evaluation assumed installation in 2004 and operation in 2005.

The potential options for incremental capacity increases are:

- 1) Upgrade an existing recovery boiler.
- 2) Install a small, stand alone recovery boiler.
- 3) Install a black liquor gasification system.

The option of upgrading an existing recovery boiler would normally be considered first, since it will typically have the lowest capital cost. However, fewer existing units will be available for upgrade in the future, because a significant number of existing units will have already been upgraded to increase capacity. Also, the increasing age of the population of existing units will make upgrading of some older units unattractive. A final factor is the extended outage time associated with recovery boiler upgrades compared with new equipment installation.

This analysis compared three alternative systems for those cases where an upgrade is not feasible for one or more mill specific reasons. Three gasifier system options were compared with installation of a small recovery boiler. Thus, the four specific options considered were:

- 1) A small or mini-recovery boiler.

It was assumed that the steam produced would be used in an existing turbine and downstream equipment. The energy output, high pressure steam, was valued based on the fuel cost that would otherwise be used to raise that steam.

- 2) A black liquor gasifier producing a low-heating-value gas.

It was assumed that the product gas would be consumed in existing equipment, thus displacing purchased fuel and valued at the displaced fuel cost.

- 3) A pressurized black liquor gasifier combined cycle.

It was assumed that the gas would be consumed in a gas turbine combined cycle system. The outputs would be electricity and low pressure (50 psig) process steam.

All systems process the same quantity of liquor, but provide different forms of energy output, and therefore have different values for those outputs. The stand alone recovery boiler was the base case against which the alternatives were compared.

The investment required in each of these cases was evaluated using a cash flow analysis to determine discounted cash flow, internal rate of return, and payback; based on the value of the fuel displaced and electricity or low pressure process steam produced. No credit was taken for the value of processing liquor. The major parameters considered included: capital cost, fuel cost, system performance, [annual] fixed-maintenance cost, one-time-fixed costs for loss of production during installation, and electricity cost. Sensitivity of the financial performance to changes in four key variables was considered.

F.2.2 Analysis Methods

A base case was developed for each of the three options considered in the analysis. A cash flow analysis was conducted for each base case. Sensitivity of the gasifier cases to: energy costs, capital cost plus one-time-fixed cost for installation outage time, fixed-annual operating and maintenance (O&M) cost, and performance was evaluated and compared with the base case for the small recovery boiler. This provided the economic results used to reach the conclusions of the study. The assumptions used for the base cases follow:

Small Recovery Boiler Base Case - The output is all high-pressure steam. That steam was valued as a product based on the cost of the energy that would be required to

raise the same quantity of steam using the assumed fuel price scenario. The energy quantity was determined based on the percentage of energy converted to steam from the high heating value of the energy input energy according to Reference 3. This energy output was valued based on the value of the purchased fuel that could be displaced if this additional steam was produced in the recovery process instead of burning fuel in a conventional power boiler.

Black Liquor Gasifier Producing a Low-Heating-Value Gas - The output was assumed to all be a low-heating-value fuel gas. The percentage of the higher heating value converted to fuel gas was estimated at 66.1% based on methods described in Reference 3. The value of the fuel gas was then based on the fuel price scenario used.

Black Liquor Gasifier / Gas Turbine Combined Cycle (Pressurized Gasifier) - The output is electricity, and low-pressure, process steam. The quantities of the outputs were taken from a proprietary B&W study that includes cycle analyses for this system. The outputs were valued based on the fuel cost scenario used.

Assumptions used to establish the capacity factor, capital costs, variable costs and byproduct credits, fixed costs, and financial parameters are described below.

Capacity Factor - All systems were assumed to perform at a capacity factor of 95%.

Capital Costs - Capital costs for the gasifier system and the small recovery boiler were taken from Reference 4 and escalated at 3.2% per year from the study year, the time frame for the reference, to 2004, the assumed installation date for all cases. Where a range of cost was available, the average was used for the base cases that is the mean cost scenario. Capital cost for the combined cycle used the gasifier costs plus the added costs for the additional major components with allowance for turnkey installation. Gas turbine system costs were adapted from Reference 5.

Additions to Capital Cost - The capital costs described above refer to capital cost of equipment plus installation, or installed capital costs. The capital costs are time-of-performance based. That is, no additional financial costs due to commitment of funds during construction was included. However, other costs considered included:

- 10% of the installed capital costs for A&E costs and owner engineering and management costs.
- 5% of the installed capital costs for project contingencies.
- 2% of the total plant cost for pre-production and startup costs. The total plant cost (TPC) is the installed capital cost plus the A&E and owner cost, plus the project contingencies described above.
- 0.5% of the total plant cost for inventory -- spare parts, etc.

Variable Costs and Byproduct Credits - The fuel input costs are all assumed to be zero (i.e., no charge for black liquor) except in the black liquor combined cycle case. There, the pilot fuel is considered as a variable cost. No additional variable costs were evaluated for adding operating personnel or for the process steam, power, or fuel required to process the incoming weak liquor or to process the product green liquor. However, these costs should be similar for all cases and, while they would change the specific computational results, the impact on the relative economic performance of the four technologies would be minor.

The outputs from the four systems were as described previously. The value for these on a per unit energy basis was established from References 1, 6, and 7 for the high, mean, and low fuel price scenarios described below. The values in Reference 6 are reported in constant 1994 dollars. The values for the first year of operation, 2005, were obtained by escalating the reported values by the GDP Implicit Price Deflator at 3.2% / year. The inflator for subsequent years was set by combining the GDP Implicit Price Deflator with the [real] Annual Growth reported for that fuel price. Values for low-pressure process steam, were conservatively estimated from the value used in Reference 7. Assumptions for each of these fuel cost scenarios are provided in Table F-1.

Table F.1. Fuel Price Scenarios (Basis for Energy Output Value).

Price Scenario	Displaced Fuel Type / Reference	Electricity Price Reference	Process Steam Price Basis
High	Natural Gas, Industrial Price High World Oil Price Table C3, Reference 6.	End-Use Price, Industrial High Economic Growth Table B8, Reference 6.	100% of the price from Reference 7, escalated at 3.2% per year to 2005 and 3.2% per year thereafter.
Mean	Industry weighted, natural gas, residual oil, and coal, Industrial Price - Reference Values - Table C3, Reference 6, weighting based on pulp and paper industry use of these three major fuels - Reference 1.	End-Use Price, Industrial Reference value Table B8, Reference 6.	75 % of the price from Reference 7, escalated at 3.2% per year to 2005 and 3.2% per year thereafter.
Low	Coal Industrial Price Low World Oil Price Table C3, Reference 6.	End-Use Price, Industrial Low Economic Growth Table B8, Reference 6.	50% of the price from Reference 7, escalated at 3.2% per year to 2005 and 3.2% per year thereafter.

All scenarios - Distillate fuel used as pilot fuel - Distillate, Industrial Price - Reference - Table B8, Reference 6

Fixed Costs - In all cases, an annual fixed maintenance cost equal to 1.5% of the total plant cost was assumed [adapted from Reference 8].

A one-time-fixed cost for loss of production during installation was also used. Analogous to the capital costs, the values used for this one-time-fixed cost were taken from Reference 4, and escalated at 3.2% per year from the year of the study to 2004. Since all cases were to install new equipment, the outage time was assumed constant for all cases. Therefore, this cost was the same for all technologies.

Financial Parameters - Constant financial parameters used for all cases included:

Discount Rate (after tax) = 9.0 %

Evaluation Period = 20 years beginning in 2005, capital installed in 2004.

Income Tax Rate (Federal + State) = 38%.

Property Tax and Insurance = 2% of installed capital cost.

General Inflation Rate = 3.2 %.

Tax Depreciation Rate per modified accelerated cost recovery system (MARCS) seven-year class.

F.2.3 Detailed Results

Table F.2 provides direct comparisons of the financial performance for the base cases of the three systems considered. The sensitivity of the base cases for the gasifier options to fuel price, capital cost plus one-time-fixed cost for production loss during installation, annual fixed operating and maintenance cost, and gasifier performance were evaluated. The ranges considered in the sensitivity analysis are provided in Table F.3. Relative financial performance of the two gasifier options compared with the small recovery boiler are provided Table F.4, where "Competitive" indicates that the option has equal or superior financial performance to the small recovery boiler base case. For fuel price sensitivity, the comparison is with the small recovery boiler operating under the same fuel price scenario -- high, mean, or low. Specific numerical results are presented in Exhibits F.1 through F.8 in Attachment F.1. Table F.3 also provides a reference to the specific exhibits that apply for each sensitivity variable. The exhibits show the sensitivity variable used and the Discounted Cash Flow, Internal Rate of Return, and Payback of Sunk Costs, for each option considered.

Table F.2. Direct Comparison of Financial Performance of Base Cases for Small Recovery Boiler, Black Liquor Gasifier Producing Gas, and Black Liquor Gasifier / Gas Turbine Combined Cycle.

Technology	Discounted Cash flow (\$1000s)	Internal Rate of Return (%)	Payback (years)
Small Recovery Boiler	1,626	10.4	8.3
Black Liquor Gasifier Producing Gas	3,071	12.1	7.4
Black Liquor Gasifier / Gas Turbine Combined Cycle	8,792	13.5	6.6

Table F.3. Economic Analysis Sensitivity Variables and Ranges.

Sensitivity Variable	Comments	Range	Results in Exhibit
Fuel Price (Basis for Energy Output Value)	All scenarios evaluated for each option.	High, Mean, & Low Fuel Price Scenarios	F.1 & F.2
Capital Cost	Includes capital cost, additions to capital cost and one-time or initial fixed cost for production loss during installation.	High: Base+15% Mean: Base Low: Base - 15%	F.3 & F.4
Operating & Maintenance Cost	High - 3% of TPC Mean - 1.5% of TPC Low - 1% of TPC	High, Mean, and Low	F.5 & F.6
Gasifier Performance	<p>Approximately the difference between the estimated performance of the B&W design used in mean case and the highest gasifier value presented in Reference 3 set an upper limit.</p> <p>Approximately the difference between the estimated performance of the B&W design used in mean case and the value for conventional technology presented in Reference 3 set a lower limit.</p>	+7% to -7% of energy output	F.7 & F.8

TPC - Total Plant Cost - refers to "Additions to Capital Cost."

Table F.4. Relative Financial Performance of Small Recovery Boiler, Black Liquor Gasifier (BLG) Producing Gas, and Black Liquor Gasifier / Gas Turbine Combined Cycle (BLG/GTCC) for the Sensitivity Variables Considered.

Sensitivity Variable	Technology Competing with Small Recovery Boiler	
	BLG Producing Gas	BLG/GTCC
Fuel Price Scenario (<i>Basis for the value of the energy output</i>)		
- High	Competitive	Competitive
- Mean	Competitive	Competitive
- Low	Competitive	Competitive
Capital & One-Time-Fixed Cost		
- High (Base Case +15%)	Not Competitive	Competitive
- Mean (Base Case)	Competitive	Competitive
- Low (Base Case -15%)	Competitive	Competitive
Annual Fixed O&M Cost		
- High (3% of TPC)	Competitive	Competitive
- Mean (1.5% of TPC)	Competitive	Competitive
- Low (1% of TPC)	Competitive	Competitive
Gasifier Performance		
- High (Base Energy Output + 7%)	Competitive	Competitive
- Mean (Base Energy Output)	Competitive	Competitive
- Low (Base Energy Output -7%)	Competitive	Competitive

F.2.4 Limitations of the Analysis

The reader is reminded that, the equipment described in this report represents a part of a pulp mill. To evaluate it out of the context of the full project for increasing capacity of a pulp mill, these alternatives were considered on a comparative basis. Therefore, the results may only be compared meaningfully with the options considered in this report, and are not meaningful for comparison with other investment options.

Usually, a decision to invest in a pulp mill expansion will be driven by the need for greater pulp production capacity. The decision of which recovery technology to use would be affected by the economic performance of the available options. This includes the value of the energy recovered as considered here and the ability of the mill to use that energy in the form in which it is produced, i.e., steam, product gas, or electricity.

The capital costs used were budgetary estimates. Some consideration was given for lower expected performance of the gasifier at higher operating pressure. However, at this stage of development, this effect cannot be predicted well.

F.2.5 Conclusions

Under the assumptions used to model the performance and relative economics of these systems and the market scenarios and constraints considered, the following conclusions were reached regarding the options for incremental capacity additions for black liquor gasification.

- 1) Upgrades of existing units will remain the most viable option, where an upgrade is feasible and has reasonable technical risk.

- 2) Due to the performance of the current atmospheric pressure, air-blown, high-temperature gasifier designs compared with conventional recovery (refer to References 3 and 9 for details), and the relative risks and high capital costs of those designs, the small stand alone recovery boiler should remain competitive where the high-pressure steam can be used within the mill in the near term.

In addition, the B&W low-temperature gasifier design offers several potential advantages over the current air-blown, high-temperature designs:

- Higher fuel conversion (refer to Reference 3)
- Less low-grade heat rejection, due to the B&W design using the waste heat in the gas to generate steam, while the high-temperature design uses a quench (refer to Reference 10)
- Little or no inorganic fume formation due to the low operating temperature
- Avoids molten smelt
- Complete sulfate reduction.

However, potential disadvantages are:

- Slow gasification kinetics leading to incomplete carbon conversion
- Incomplete conversion of tars
- More H₂S in the product gas that increases causticizing load.

- 3) The black liquor gasifier has merit for incremental capacity increases. It would be more effective using a combined cycle. In either case, whether a stand alone gasifier or combined cycle system, the resulting energy output would be delivered in forms other than higher pressure steam so it can address a market segment that differs from conventional technology and that is potentially larger.

F.2.6 Recommendations

This evaluation was limited to the assumed performance of the black liquor gasifier and a reasonable variation in outputs about that predicted value. Bench-scale testing conducted under a cooperative agreement with DOE provided limited results, indicating that carbon conversion may be low if sufficiently high gasifier temperatures are not achieved. This was partially confirmed by data reported for another low-temperature black liquor gasifier (refer to Reference 11). A novel approach that will emphasize improved tar and char conversion was developed. The first steps to demonstrate the feasibility of the improvements in the design have been successfully completed under a company-funded project. Further bench-scale investigation was also proposed under Agenda 2020.

It is recommended that the economic evaluation be revisited using results of these two projects or other future studies, once the desired carbon conversion and tar utilization are demonstrated at the bench scale. Future work might also consider greater depth to the analysis, but this first level of analysis provides sufficient detail to reach the conclusions drawn in this study.

F.3 REFERENCES

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ATTACHMENT F.1

DETAILED RESULTS OF ECONOMIC ANALYSIS
SENSITIVITY TO: FUEL, CAPITAL, O&M, AND PERFORMANCE

EXHIBITS F.1 - F.8

Exhibit F.1

Fuel Price Sensitivity

Small Recovery Boiler / BLG Producing Gas

Technology	Cost Variable - Displaced Fuel Costs	Price Used (\$/million Btu in 2005)	Inflator (% / year)	Description	Displaced Fuel Savings (\$/hr)	Discounted Cash Flow (\$1000s)	Internal Rate of Return (%)	Payback of Sunk Costs (Years)
Small Conventional Recovery Boiler	High	\$3.87	4.5%	Natural Gas - High	\$336	\$6,047	13.8%	6.8
	Mean	\$3.23	3.9%	Industry Weighted - Mean	\$281	\$1,626	10.4%	8.3
	Low	\$1.81	2.2%	Coal - Low	\$157	(\$7,272)	<1%	19.8
Black Liquor Gasifier Producing Gas	High	\$3.87	4.5%	Natural Gas - High	\$316	\$7,184	15.5%	6.2
	Mean	\$3.23	3.9%	Industry Weighted - Mean	\$265	\$3,071	12.1%	7.4
	Low	\$1.81	2.2%	Coal - Low	\$148	(\$5,323)	1.7%	16.9

Exhibit F.3

Capital and Initial Fixed Cost Sensitivity

Small Recovery Boiler / BLG Producing Gas

Technology	Cost Variable	Capital Cost and Initial Fixed Costs	Initial Fixed Costs (+15% and - 15%)	Description	Displaced Fuel Savings (\$/hr)	Discounted Cash Flow (\$1000s)	Internal Rate of Return (%)	Payback of Sunk Costs (Years)
	Capital and Initial Fixed Costs	Capital Cost (\$1000's in 2004)	Initial Fixed Costs (\$1000's in 2004)					
Small Conventional Recovery Boiler	Mean	\$13,017	\$685	Base	\$281	\$1,626	10.4%	8.3
Black Liquor Gasifier Producing Gas	High	\$12,606	\$788	Base + 15%	\$265	\$1,052	9.7%	8.6
	Mean	\$10,962	\$685	Base	\$265	\$3,071	12.1%	7.4
	Low	\$9,317	\$582	Base -15	\$265	\$5,090	14.7%	6.4

Technology	Cost Variable	Capital Cost and Initial Fixed Costs (+15% and - 15%)	Value of Outputs	Electricity (\$/hr)	Process Steam (\$/hr)	Discounted Cash Flow (\$1000s)	Internal Rate of Return (%)	Payback of Sunk Costs (Years)
		Initial Fixed Costs (\$1000's in 2004)	Displaced Fuel Savings (\$/hr)					
	Capital and Initial Fixed Cost Scenario							
	Mean	\$13,017	\$281			\$1,626	10.4%	8.3
Small Conventional Recovery Boiler								
	High	\$25,825		\$582	\$67	\$4,724	11.2%	7.6
Black Liquor Gasifier Gas Turbine Combined Cycle	Mean	\$22,456		\$582	\$67	\$8,792	13.5%	6.6
	Low	\$19,088		\$582	\$67	\$12,860	16.3%	5.7

Exhibit F.5

O&M Cost Sensitivity

Small Recovery Boiler / BLG Producing Gas

Technology	Cost Variable - Fixed Annual	O & M as % of Capital.	Description	Displaced Fuel Savings (\$/ hr)	Discounted Cash Flow (\$1000s)	Internal Rate of Return (%)	Payback of Sunk Costs (Years)
	O&M Cost (\$1000's in current \$'s for start year)	O&M Cost as a % of Total Plant Cost					
Small Conventional Recovery Boiler	Mean \$226	1.5%	Base (1.5%)	\$281	\$1,626	10.4%	8.3
Black Liquor Gasifier Producing Gas	High \$380	3.0%	High (3%)	\$265	\$1,721	10.8%	8.1
	Mean \$190	1.1%	Base (1.5%)	\$265	\$3,071	12.1%	7.4
	Low \$127	1.0%	Low (1.0%)	\$265	\$3,521	12.5%	7.2

Exhibit F.6

O&M Cost Sensitivity

Small Recovery Boiler / BLGCC Options

Technology	Cost Variable -	Fixed Annual	O & M as % of	Total Plant Cost	Value of Outputs Displaced Fuel Savings (\$/ hr)	Electricity (\$/ hr)	Process Steam (\$/ hr)	Discounted Cash Flow (\$1000s)	Internal Rate of Return (%)	Payback of Sunk Costs (Years)
O&M Cost Scenario	O&M Costs (\$1000s in current \$ for in-service year)	O&M Costs as % of Total Plant Cost	O&M Costs as % of Total Plant Cost	Value of Outputs Displaced Fuel Savings (\$/ hr)	Electricity (\$/ hr)	Process Steam (\$/ hr)	Discounted Cash Flow (\$1000s)	Internal Rate of Return (%)	Payback of Sunk Costs (Years)	
Small Conventional Recovery Boiler	Mean \$226	1.5%	\$281				\$1,626	10.4%	8.3	
Black Liquor Gasifier Gas Turbine Combined Cycle	High \$778	3.0%			\$582	\$67	\$6,026	12.1%	7.1	
	Mean \$389	1.5%			\$582	\$67	\$8,792	13.5%	6.6	
	Low \$259	1.0%			\$582	\$67	\$9,714	13.9%	6.5	

Exhibit F.7

Gasifier Performance Sensitivity Small Recovery Boiler / BLG Producing Gas

Technology	Cost Variable	Gasifier Performance - Gasifier Performance (Energy Output Change of +7% to -7%)	Displaced Fuel Savings (\$/ hr)	Discounted Cash Flow (\$1000s)	Internal Rate of Return (%)	Payback of Sunk Costs (Years)
	Gasifier Performance Scenario	Gasifier Performance Change in % of Net Energy Output				
Small Conventional Recovery Boiler	Mean	N/A	\$281	\$1,626	10.4%	8.3
Black Liquor Gasifier Producing Gas	High	+7%	\$284	\$4,228	13.1%	6.9
	Mean	Unchanged	\$265	\$3,071	12.1%	7.4
	Low	-7%	\$246	\$1,914	11.0%	8.0

