SECTION II

Operations & Economics

5.0   TECHNICAL PERFORMANCE

The technical performance of the plant has been evaluated within this section primarily by a review of system and equipment reliability, availability and maintainability. Each major piece of equipment and process step has been evaluated over the best 12 months of operation during the Demonstration Period. At Wabash River this time period covers March 1, 1998 through February 28, 1999 and is referred to throughout the following section as the "Operating Period". It is significant to note that this period represents the final full 12-month period of operations within the Demonstration Period and is immediately preceding the 100-day outage for the combustion turbine failure.

As the entire Operating Period is encompassed within the defined Demonstration Period for the Project, some of the information contained within this section is redundant when compared with information in Section 4.0 Demonstration Period. This redundancy is intentional, allowing these two sections of the Final Technical Report to be reviewed independently.

Reliability discussions included in Section 5.0 address the consistency in operation of specific equipment items, while the maintainability discussions are associated with specific equipment maintenance issues, which prevented consistent operation. The availability of all subsystems is defined as the percentage of hours that a subsystem was available for operation when compared to the total hours in the Operating Period minus the "scheduled" facility downtime hours. These numbers are compiled and reviewed in detail in Table 5.0A.

Table 5.0A: WRCGRP Operating Period Downtime and Availability

Section 5 Reference	Area	Non-Scheduled Downtime Hours	% of Non-Scheduled Downtime	Availability
5.1	**Air Separation Unit***	**285**	**12.5%**	**96.32%**
5.1.1	Air Compression System	87	3.8%	98.87%
5.1.2	Water Wash System	61	2.7%	99.21%
5.1.3	Air Purification System	10	0.4%	99.87%
5.1.4	Air Cooling and Liquefaction System	0	0.0%	100.00%
5.1.5	Cryogenic Distillation System	77	3.4%	99.00%
5.1.6	Oxygen Mixing System	0	0.0%	100.00%
5.1.7	Nitrogen Handling and Storage System	15	0.7%	99.81%
5.1.8	Oxygen Compression System	70	3.1%	99.09%
5.2	**Coal Handling**	**36**	**1.6%**	**99.54%**
5.2.1	Coal Hopper System	0	0.0%	100.00%
5.2.2	Rod Mill System	0	0.0%	100.00%
5.2.3	Slurry Storage Tank System	3	0.1%	99.96%
5.2.4	Slurry Feed System	33	1.4%	99.58%
5.3	**Gasification**	**1590**	**69.9%**	**79.45%**
5.3.1	First Stage Gasifier	1001	44.0%	87.06%
5.3.2	Second Stage Gasifier	169	7.4%	97.82%
5.3.3	Raw Syngas Conditioning	0	0.0%	100.00%
5.3.4	Slag and Solids Handling	66	2.9%	99.15%
5.3.5	Syngas Cooling/Steam Generation System	158	6.9%	97.96%
5.3.6	Particulate Removal System	152	6.7%	98.03%
5.3.7	Chloride Scrubbing System	10	0.4%	99.87%
5.3.8	COS Hydrolysis System	0	0.0%	100.00%
5.3.9	Low Temperature Heat Recovery System	7	0.3%	99.90%
5.3.10	Syngas Moisturization System	0	0.0%	100.00%
5.3.11	Acid Gas Removal System	22	1.0%	99.72%
5.3.12	Sulfur Recovery System	5	0.2%	99.94%
5.3.13	Sour Water Treatment System	0	0.0%	100.00%
5.3.14	Cooling Tower System	0	0.0%	100.00%
5.4	**Power Block**	**363**	**16.0%**	**95.30%**
5.4.1	Combustion Turbine	140	6.1%	98.19%
5.4.2	Heat Recovery Steam Generator	201	8.8%	97.40%
5.4.3	Water Treatment Facility	13	0.6%	99.83%
5.4.4	Steam Turbine	9	0.4%	99.88%
	Totals	**2274**	**100.0%**	**70.61%**
	Coal Hours	**5462**		
	Hours for 3 scheduled outages	**1022**		
	Non-scheduled downtime hours	**2274**		
	Period Hours	**8760**		

*Thirty-five hours of downtime affected both sections 5.1.1 and 5.1.8. It has been counted only once in the ASU total but in each section independently.

Each of the plant areas shown in Table 5.0A are discussed from three perspectives in this section of the report. First, system modifications, which include significant mechanical, procedural or process operating changes made to the system prior the Operating Period. Second, operating overview, which summarizes operation, downtime and maintenance associated with each piece of equipment in an area during the Operating Period. Third, summary and conclusions, which highlight past and present operation, describe changes made within the system after the Operating Period, and recommends changes for future plants.

5.1   Air Separation Unit

5.1.1 Air Compression System

Air, utilized as feedstock in the air separation plant, is harvested from the atmosphere and pressurized by a three-stage, centrifugal, main air compressor. Subsystems for the main air compressor include a three-phase, high-voltage, AC synchronous motor driver, lube oil system with vapor recovery, anti-icing protection and vibration monitoring equipment, an inlet air plenum that filters particulate prior to compression, inlet guide vanes, discharge vent valves and inter-stage cooling air coolers that remove heat of compression after each of the three compressor stages.

The flow of air through the compressor is controlled via modulation of guide vanes on each of the three compressor stages. The machine is protected from surge via automatic vent valves on the discharge.

Figure 5.1.1A: Main Air Compressor

5.1.1.1 System Modifications

No system modifications were made to the air compression system prior to the Operating Period. However, as a result of reliability issues with the main air compressor's inlet guide vane and surge protection subsystems, several procedural changes were made in an attempt to increase the main air compressor's reliability. Also, concerns over moisture detected within the motor driver enclosure prompted steps to guard against a potential future failure of the motor windings. Details of these significant procedural changes and preventative actions taken prior to the Operating Period follow.

Inlet Guide Vanes

The inlet guide vane subsystem was recognized very early in system operation as the biggest reliability issue for the main air compressor. Specifically, guide vane corrosion as a result of moisture, infrequent guide vane movement, and inadequate weatherproofing of the guide vane actuators and their controllers were all noted as concerns resulting from improper design. Fifty-three hours of downtime were suffered beginning on January 25, 1998 due to the inlet guide vane subsystem. In addition to this specific downtime incident, maintenance performed on the guide vane subsystem was frequent and costly. On one occasion prior to the Operating Period, one of the three guide vanes required complete extraction from the compressor in order to allow repair to a twisted actuator shaft suffered while attempting to stroke the guide vanes after a prolonged period of inactivity.

The following actions were taken prior to the Operating Period in order to address concerns over the inlet guide vane subsystem of the main air compressor. First, usage of the subsystem's capability to provide an on-line wash of the inlet guide vanes via a water spray was abandoned. Results from lab analysis of both the supply and return water indicated that on-line washing might have been increasing, rather than decreasing, the amount of deposits on the guide vanes. Second, a policy of exercising the guide vanes periodically during outages was instituted in order to prevent them from binding due to long periods of inactivity. Third, inspections of all critical components of the inlet guide vane subsystem were performed during extended outages. Weatherproofing efforts were undertaken as needed during those inspections. Finally, efforts to upgrade or procure a replacement for the inlet guide vane subsystem were initiated.

Discharge Vent Valves

Failure of the main air compressor's discharge vent valves caused 35 hours of downtime beginning on January 11, 1998. These valves provide surge protection for the main air compressor and are normally held in their fully closed position while the machine is running. These surge valves are subjected to extended time periods without being exercised, and therefore, are susceptible to corrosion that causes them to be stiff and difficult to operate. The experienced downtime as a result of their non-responsiveness prior to the Operating Period prompted the implementation of a preventative maintenance plan. The plan includes lubrication of the valve actuators and periodic stroking of the valves through their full range of motion during outages.

Motor Driver

A routine inspection of the main air compressor's motor driver prior to the Operating Period showed evidence of moisture within the motor enclosure. The moisture was visible in the bottom of the enclosure as well as detected in the insulation of the motor windings through standard electrical testing. Although it did not lead to downtime, moisture within the windings of such a large, high-voltage motor was addressed with top priority due to its potential to cause a catastrophic failure of the windings. It was determined that the fiberglass motor enclosure was inappropriately designed for outdoor service. Hairline cracks and pinholes around the rivet points and seams in the enclosure provided possible entrance points for moisture. As an immediate action, care was taken to seal the enclosure by application of a commercially available sealant. Further investigation of the system revealed that the electric heaters within the motor enclosure were improperly installed during construction and had never functioned. The installation error was rectified, and the heaters began functioning as designed. As an added precaution, a procedural change was made to include removal of the cooling water flow from the motor immediately after a shutdown of the machine. Cooling water flow significantly counteracted the positive effects of the motor enclosure heaters, and may have promoted moisture condensation during the winding cooldown period immediately following the shutdown of the main air compressor.

5.1.1.2 Operating Experience Overview

The total downtime associated with the air compression system during the Operating Period was 87 hours. Operating experience for individual pieces of equipment follows.

On August 9, 1998 an electronic card supplying power to the programmable logic controller for the inlet guide vane actuator subsystem failed causing 47 hours of downtime. The failed card was sent to the original equipment manufacturer for analysis. Their investigation concluded that the failure resulted from a high voltage surge consistent with a lightning strike. This analysis seems reasonable due to the fact that the incident occurred during severe weather. No corrective actions other than a like-for-like replacement of the card were undertaken.

On August 15, 1998 a blown 15kV electrical fuse in a potential transformer located in the motor control center's high-voltage switchgear triggered a trip of the main air compressor (the oxygen compressor also tripped for the same reason). The incident led to 35 hours of plant downtime. The driver motors for both the main air compressor and the oxygen compressor rely upon a signal from this potential transformer in order to control their synchronous fields. Loss of the transformer's fuse forced them both to trip instantaneously. Rigorous electrical testing of the potential transformer confirmed no perceivable problem with the transformer. The possibility of a defective fuse was also ruled out, as this was the second incident involving the same fuse location in just over eight months (the first caused 33 hours of downtime prior to the Operating Period on February 5, 1998). Although no definitive explanation could be found for the blown fuse, the potential transformer was swapped with an identical one from a less critical service. No further problems with this device have been experienced or are anticipated.

On October 8, 1998 an interruption in the power supply to the vibration monitoring subsystem for the main air compressor was experienced causing 5 hours of plant downtime. Investigation revealed that power was accidentally shutoff by a technician working inside the cabinet housing of the power source. Work in that cabinet is now restricted to plant outages and a caution sign has been posted on the cabinet door in order to warn of the shutdown potential.

A project to maximize the effectiveness of the main air compressor's inlet air plenum de-icing mechanism was implemented during the Operating Period. This mechanism operates by routing a small slipstream of warm air from the compressor discharge back to the inlet air plenum. The piping for this slipstream had been exposed to ambient weather conditions since the completion of plant construction. The addition of insulation to the line minimized the flow of air required for de-icing the plenum, thus increasing the amount of air available for feedstock to the cryogenic plant during cold weather.

A vibration data collection system was installed on the main air compressor during the Operating Period. The new system added historical archiving, event capturing, and analytical tools to the already existing vibration monitoring equipment. The increased functionality of the vibration equipment now allows for predictive maintenance techniques to be employed through monitoring of data trends and enhances the accuracy of troubleshooting in response to vibration related events.

Maintenance on the main air compressor during the Operating Period was limited to normal preventative maintenance procedures (PM's) performed during planned outages. Some of the tasks performed during these PM's included: inspection, cleaning, and replacement of the air intake filters, electrical measurements on the motor, motor enclosure inspections and weatherproofing, coupling and bull-gear inspections, lube oil system inspection and minor leak repair, pressure safety valve calibrations, inspection of surge valves, and inlet guide vane actuator calibrations.

5.1.1.3 Summary and Conclusions

In summary, the overall reliability of the main air compressor has been disappointing. Specifically, the inlet guide vane subsystem has been a point of almost constant concern leading up to and through the Operating Period. Several aspects of the subsystem were poorly designed by the original equipment manufacturer.

Actuators have been highly susceptible to weather conditions, the electronic controllers for the actuators have proved to be imprecise and difficult to manage, critical components have worn

quickly and repeatedly and replacement parts that should be commonplace have been difficult to procure. Furthermore, all of these difficulties have been compounded by the fact that the original equipment manufacturers for both the compressor and guide vane actuators have not been cooperative with regard to corrective actions. Future gasification projects requiring the installation of air compressors of this magnitude need to pay particularly close attention to the detailed design of the main air compressor's inlet guide vane subsystem.

The actuators on the inlet guide vane subsystem for the main air compressor were replaced in October 1999, with a new third party design (Figure 5.1.1B). Some of the benefits of the new actuators include the following. The new system is designed for all weather conditions, which should eliminate plant downtime associated with moisture. The new design completely eliminates the field electronics inherent with the old system, thus simplifying guide vane control. The new actuators are designed for more cycles of operation than the plant could reasonably require in the course of several decades. All three actuators have a mechanism to allow them to be manually positioned and locked anywhere within their full range of motion providing the capability to perform actuator maintenance while the main air compressor remains on line. The new system incorporates only parts made in the USA, thus reducing cost and delivery lead-times. All three stages now utilize identical components affording a reduction of spare inventory. The actuators have caused no downtime since installation, and are expected to eliminate the air separation unit's largest source of downtime.

Figure 5.1.1B: Close-up of a Newly Installed Guide Vane Actuator

While installing the new inlet guide vane project for the main air compressor, all three of the guide vanes needed to be removed from the compressor for inspection and repairs. During this process several small improvement projects were implemented in order to prevent damage to the inlet guide vanes and reduce costs of future guide vane repairs. They include the addition of lift points to the large diameter second and third stage air pipe spools to reduce rigging and lifting efforts while increasing safety and the addition of drain valves for each guide vane flange to prevent internal guide vane corrosion. Also, new flanges were added to the compressor oil return line to improve access for future guide vane extractions. Finally, proper mechanical design was implemented via materials selection for the guide vane shaft and shaft key, providing future equipment protection.

In addition to the problems related to the inlet guide vane subsystem, difficulties in preventing moisture from penetrating the motor enclosure have been a serious concern with regard to protecting against a costly motor rewind. A dry gas purge installed in the motor enclosure during the first quarter of 1999 coupled with heater repairs, enclosure-weatherproofing efforts, and procedural changes pertaining to cooling water flows have brought the moisture problem under control. Future gasification projects requiring the installation of outdoor air compressors need to pay particularly close attention to the quality of design and material selection used in the motor driver's enclosure.

One other point of concern with regard to the reliability of the main air compressor involves the lack of redundancy of vibration probes located within the machine. On February 26, 2000, a total of 15 hours of downtime was suffered due to a false vibration trip of the main air compressor. Investigation identified that loose electrical connections and moisture penetration into those connections were the root causes of the downtime. The problem has been rectified with precautions that should prevent a repeated occurrence. However, this incident coupled with sporadic vibration-related nuisance trips suffered prior to the Operating Period have prompted evaluation of the need for added redundancy in the vibration monitoring subsystem. The evaluation is currently underway.

Major maintenance that has occurred with the main air compressor since the Operating Period has been limited to leak repairs within the inter-cooler between the first and second stages. During the course of leak repairs, evidence clearly identified a design flaw associated with the internal mist elimination equipment. A robust brace assembly was designed and implemented in order to prevent similar future failures. Minor leaks with the main air compressor's inter-stage coolers continue, but they are currently of no detriment to the performance of the plant. They will be addressed in future outages, but only if they become more severe. All other maintenance performed since the end of the Operating Period, and anticipated in the future, is limited to the preventative maintenance work discussed at the end of Section 5.1.1.2.

5.1.2 Water Wash System

The water wash system within the ASU includes:

Water chiller tower (Figure 5.1.2A)	Water chiller pumps (Figure 5.1.2A)
Wash water tower (Figure 5.1.2A)	

Figure 5.1.2A: Water Wash System

5.1.2.1 System Modifications

No system modifications were made and no downtime was attributed to the water wash system prior to the Operating Period.

5.1.2.2 Operating Experience Overview

The total downtime associated with the water wash system during the Operating Period was 61 hours. Operating experience for individual pieces of equipment follows.

Water Chiller Tower

On June 8th and 9th, 1998 packing fires inside the water chiller tower were accidentally ignited from welding slag during vessel entry work. These fires accounted for all 61 hours of downtime attributed to the water wash system during the Operating Period. Evidence suggested the fires to be the result of inadequate usage of fire barriers while work was being performed, as well as failure to utilize a low energy welding technique. The use of heli-arc type welding in place of common stick welding and better usage of fire barriers will prevent recurrence of this incident if weld repairs inside the vessel ever need to be repeated.

The scope of work being performed in the water chiller tower at the time of the aforementioned fires was to correct oversight in the water distribution system design, repairing and reinforcing hardware within the tower. The water distribution system increases the efficiency of the water chiller tower by maximizing the contact between the water and the waste nitrogen. Years of operation with inadequate design had taken a toll upon the distribution system to the point that warmer water temperatures were beginning to limit the production of the air separation unit. As a result of the maintenance to the water chiller tower, water temperatures were significantly reduced, thus increasing oxygen and nitrogen production in the air separation unit.

Water Chiller Pumps

There were no major maintenance, downtime or operational problems associated with the water chiller pumps during the Operating Period. However, minor maintenance during the Operating Period included periodic replacement of suction strainers and typical preventative maintenance inspections and procedures.

<u>Water Wash Tower</u>

There was no maintenance, downtime or operational problem associated with the water wash tower during the Operating Period.

5.1.2.3   Summary and Conclusions

The water wash system has been a very reliable system. With the exception of delays caused by accidental fires in June of 1998, the water wash system has caused no plant downtime. However, the system does struggle to maintain cool water temperatures during unusually hot summer conditions. These periods have the potential to limit plant production because warmer water temperatures increase moisture carryover to the downstream air purification adsorber beds. This increased carryover can occasionally limit the throughput capabilities of the adsorber beds.

The occasional difficulties in cooling the chiller water are attributed to the following. First, the air temperature from the discharge of the main air compressor is elevated during the summer, because cooling tower water used in the compressor inter-coolers is warmer than desired. Secondly, rust and corrosion products tend to build up, over time, in the water distribution system within the water chiller tower. This allows channeling of the water through portions of the tower, thus reducing contact between the water and the waste nitrogen. Periodic cleaning and repair of the distribution system is all that is anticipated in order to prevent channeling. However, limits in the capabilities of the cooling tower are more difficult to overcome. As these limits are present only during periods of higher ambient temperatures, a cost versus benefit analysis for designing the cooling tower for these conditions should always be performed for any future projects.

It is also worth mentioning that corrosion of the vessel walls within the water chiller tower is occurring and will likely necessitate the replacement of the vessel several years in the future. It is unlikely that very much of the corrosion is taking place during plant operation, because the vessel is padded with waste nitrogen during those times. However, plant shutdowns allow air to contact the wet vessel internals. It would be wise to evaluate the cost of using an alternative material for the internals of the water chiller tower for future gasification projects.

5.1.3 Air Purification System

The air purification system includes:

Adsorber beds (Figure 5.1.3A)	Regeneration heater (Figure 5.1.3B)

The adsorbers are two-stage, radial-flow media beds with one adsorber on line at all times while the other is being regenerated off line. Once regenerated and cooled, the off-line bed is brought on line and the other is heated for regeneration. This batch cycle is repeated constantly during plant operation.

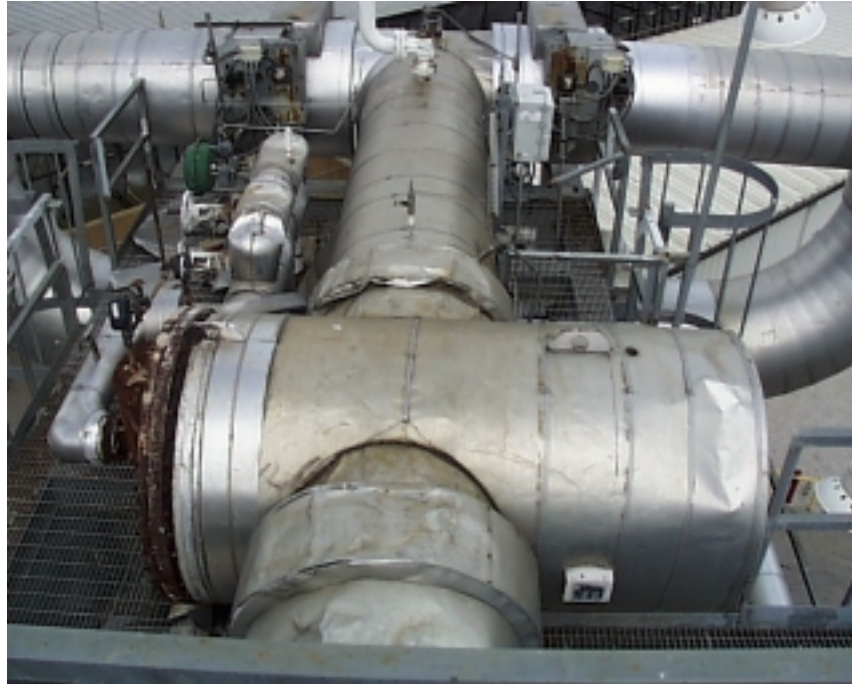**Figure 5.1.3A: Adsorber Beds for the Air Purification System**

Figure 5.1.3B: Regeneration Heater for the Air Purification System

5.1.3.1 System Modifications

Modifications in the air purification system prior to the Operating Period were limited to the adsorber vessel valves. Several large, automated block valves are employed by the air purification system to control the adsorption and regeneration cycles of each adsorber bed. These must operate several times per day and their performance is critical to plant operation. Each valve must stroke quickly when operation is requested and valves in the closed position must maintain a seal even with high differential pressure across them. Failures in the performance of these valves were frequent and often threatened the production of the facility prior to the Operating Period. Investigation of the failures concluded that the valve bushing material utilized by the original equipment manufacturer was inadequate. The material was upgraded in response to the investigation findings.

5.1.3.2 Operating Experience Overview

The total downtime associated with the air purification system during the Operating Period was 10 hours. Operating experience for individual pieces of equipment follows.

Adsorber Beds

On October 27, 1998 ten hours of downtime were suffered as the result of actuator problems on one of the automatic block valves that control the adsorber batch cycle. The actuator apparently worked itself loose from the valve. When the valve was called close, the loose actuator did not allow the valve's limit switch to prove the valve closed and prevented a successful regeneration of the off-line adsorber bed. The regeneration delay went unrecognized by operating personnel until monitoring equipment detected carbon dioxide breakthrough in the air purification system. This initiated a manual halt of operation until a full regeneration cycle could be completed for each of the adsorber beds. All of the following action items were identified and completed in response to the downtime. The valve actuator was repaired; new control system alarms were implemented to detect similar problems; and operating personnel were trained in techniques that would have allowed them to recognize valve issues before they become a threat to the process. No further downtime of this type is expected in the future.

A project was implemented during the Operating Period to reduce noise levels produced during the venting step of the regeneration cycle of the off-line adsorber bed. An ancillary silencer was placed on the common adsorber bed, high-pressure exhaust vent. This reduced peak noise levels in the area from 105 decibels (dB) to below 87 dB. After the project was implemented it became difficult to accurately measure the noise produced by venting because it was at, or below, the normal plant background noise. There was no major maintenance associated with the adsorber beds during the Operating Period.

Regeneration Heater

The gas distribution hardware of the regeneration heater was modified during the Operating Period to include enhanced stiffening supports to prevent collapse of the regeneration heater gas baffles. The additional support restored the proper gas contact and thus increased the efficiency of the regeneration heater, resulting in improved regeneration heat peaks on the absorber beds. Prior to the modification, the air separation unit struggled to maintain production levels during periods of hot and humid weather as a result of increased moisture carryover to the adsorber beds. The increase in heater efficiency relieved production limitations during those times by providing more heat energy to drive off the excess moisture in the regeneration cycle. During all

other periods of time, the improved heater performance allowed regenerated adsorber beds to remain on line for longer periods, thus reducing the number of equipment cycles per year and operating energy requirements.

In addition to the gas distribution design remediation project, the regeneration heater required extensive and regular maintenance to repair leaking steam tubes. The regeneration gas exchanger has an external pass-partitioned collector box and the tube segments connecting the external collector box to the exchanger body are exposed to high thermal stresses and were prone to fatigue failures. Although no downtime was attributed to the regeneration heater during the Operating Period, multiple and regular tube repairs were required.

5.1.3.3 Summary and Conclusions

Several large block valves in the air purification system have made the system difficult to operate almost from the beginning of plant operation. Bushing material upgrade implemented prior to the Operating Period temporarily alleviated problems, but was conceived as a patch while the valve manufacturer committed to correct the design flaw for all valves under warranty. The patch did not eliminate operational difficulties entirely. In fact, problems with these valves caused the only downtime associated with the air purification system during the Operating Period (see Section 5.1.3.2 for details) and instances of production curtailment in the air separation unit have been experienced after the Operating Period as a result of valve performance issues.

An example of valve problems immediately following the Operating Period occurred in the 3rd quarter of 1999 when fifteen hours of downtime resulted from failure of an air purification valve to properly seal. The valve leak prevented depressurization of one absorber bed causing the regeneration cycle to go into hold. This briefly interrupted the feedstock of clean dry air into the front end of the plant, ultimately resulting in the gasifier coming off coal due to a lack of oxygen supply. In response to the incident, a third bushing design was implemented, within that specific valve, after inspection revealed failed valve bushings that placed the internal mechanism in a bind, which prevented the valve from closing. The new design incorporates the use of oxygen-compatible grease with free-floating bushings and is the plant's second attempt at improving the

life span of these bushings (the first attempt was the material upgrade implemented prior to the Operating Period, see Section 5.1.3.1 for details). It is expected that this design, or a minor modification thereof, will be a permanent, long-term fix for all valves in the air purification system should their bushings begin to fail in the future.

Other significant problems were attributed to the actuators that automate the batch sequencing valves. One problem involved the actuator solenoids, which were not designed for outdoor service and used substandard foreign coil voltage. Weak coil response resulted in poor solenoid activation reliability and consequently poor valve actuation. At each activation event, manual intervention was required to operate the valve and to maintain the solenoids. A retrofit dual solenoid with upgraded coil voltage, more compatible with the ASU control system, was implemented on all actuators.

A second problem involved the actuator support design. When actuating, the pneumatic cylinder encounters rotational displacement against the supporting trunnion bushings. The bushing design was inadequate for outdoor service and utilized ineffective materials. The trunnions of one actuator actually sheared off, completely preventing valve operation, and other actuators experienced similar problems related to these bushings. A new trunnion bushing design was specified utilizing a grease-able trunnion, a grease pocket in the bushing, and increased bushing clearances.

Solutions pertaining to these valve issues have been progressively developed and implemented throughout the history of the plant. It is now believed modifications to the materials and design of the valve actuator bushings and trunnions and upgrades to the electrical solenoids that modulate the instrument air signals for the valves will greatly increase the reliability of the valves, and, hence, the entire air purification system. The instrument air solenoid upgrades have been installed for all air purification valves; however, the new bushing and trunnion design will be progressively implemented on an as-needed basis for valves that show signs of wear. It is also worth noting that efforts to develop and implement solutions to issues surrounding these valves have been significantly hampered over the years due to a lack of cooperation from the

original equipment manufacturer. Future gasification projects should be very selective with regards to choosing a valve type and manufacturer for this particular application.

Other projects and modifications within the air purification system that have been implemented since the completion of the Operating Period have included the following: the outlet tubes of the regeneration heater were sleeved in the areas of thermal stress and tube fatigue to prevent high O&M costs associated with thermal fatigue and leaking tubes; automatic bottle switchers were installed on the analyzer subsystem in an effort to reduce nuisance analyzer problems; and thermowells and RTD's were installed at selected locations within the air purification system in order to facilitate data collection during the plant derime procedures. All of these system modifications are worthy of consideration for any potential future gasification project.

Finally, the impact of process transients induced by swapping adsorber beds at the end of every regeneration cycle is typically felt throughout the air separation plant. The consequence is that the Operations team often must operate the air separation plant at slightly below capacity in order to brace the plant for these periodic upsets. Although these transients have never caused downtime, they do represent a bottleneck in the air separation process. Future gasification projects should take the transient affects of a batch-type air purification process into account when sizing plant equipment.

5.1.4 Air Cooling and Liquefaction System

The air-cooling and liquefaction system includes:

Main cryogenic exchangers (Figure 5.1.4A)	Compressor/Expander refrigeration system (Figure 5.1.4B)
Nitrogen vaporizer (Figure 5.1.4B)	

Figure 5.1.4A: Nitrogen Vaporizer and Enclosure for Main Exchangers

Figure 5.1.4B: Compressor/Expander Skid

5.1.4.1 System Modifications

Two major modifications were made to the air-cooling and liquefaction system very early in the Project. The first modification was the addition of a supplemental liquid nitrogen vaporizer attached to the main exchanger cold box, utilizing warm, purified air from the air purification system to vaporize liquid nitrogen (LIN) from the nitrogen storage and handling system (Figure 5.1.4A). Prior to this equipment installation, LIN vaporization for the purpose of supplying the facility's gaseous nitrogen requirements was insufficient. Only a fraction of the LIN vaporization was handled through the main exchangers. The bulk of the high-energy required to generate LIN was discarded with an increased energy penalty by utilizing a steam vaporizer. The installation of the ancillary nitrogen vaporizer increased the efficiency of nitrogen vaporization and provided an additional refrigeration stream for a portion of the air feedstock to the air separation plant.

After installation of the ancillary LIN vaporizer, a significant problem arose. Refrigeration being placed back into the purified air stream via the ancillary exchanger resulted in unexpected air stream condensation and imbalance in the main exchanger hot and cold approach temperatures. Additionally, feed to the refrigeration expander from the main exchangers was too cold for proper expansion. Hot, boosted air bypass was necessary to keep the expander operational resulting in inefficient expander operation and lost refrigeration.

To restore the main exchanger approach temperatures and restore refrigeration expander performance, a second major modification was made to the air-cooling and liquefaction system with the addition of a high-pressure oxygen recycle line. The ancillary LIN vaporizer performed well enough to reconfigure the smaller vaporizer within the main exchanger as a condenser to remove surplus refrigeration. To achieve this, a side-stream of warm high-pressure oxygen from the discharge of the oxygen compressor system, when available, is now condensed in the main exchangers. Its pressure is dropped across a control valve, and it is routed into the low-pressure column as a feedstock. This re-established the temperature approaches of the main exchanger, restored operating efficiency to the refrigeration expander and recycled the otherwise lost refrigeration (in the condensed oxygen), directly to the low-pressure column. The net result is increased liquid nitrogen production for the air separation unit.

The only major maintenance job performed in the air-cooling and liquefaction system prior to the Operating Period was the replacement of the refrigeration expander's rotating assembly. An incident that occurred prior to commercial operation severely damaged the expander assembly. Root cause investigation revealed two possible causes of the incident. The unit may have been operating at a rotational speed that coincided with its natural resonance frequency, hence promoting catastrophic internal vibrations. Another possibility is that the unit's performance may have been pushed to the point at which liquefaction began occurring in the air stream prior to exiting the expander. Either condition could have quickly caused the type of damage that was found in the post-incident investigation. The unit was repaired and procedural changes were instituted to prevent a re-occurrence. The unit is now operated at a different rotational speed in order to avoid resonance frequencies, and the outlet temperature is closely monitored and controlled in order to prevent liquefaction within the expander.

5.1.4.2 Operating Experience Overview

There were no major maintenance, downtime or operational problems associated with any of the air-cooling and liquefaction equipment during the Operating Period. Minor maintenance performed during this time included periodic removal of ice build-up around the control actuator of the expander guide vanes, instrumentation calibrations and lubrication oil changes.

5.1.4.3 Summary and Conclusions

The implementation of the ancillary nitrogen vaporizer and high pressure oxygen condensing recycle line and the procedural changes regarding operation of the compressor/expander allowed the air cooling and liquefaction system to be very reliable and efficient in the period leading up to and through the Operating Period.

However, an air leak inside the enclosure that houses the main exchangers forced an unplanned outage resulting in 299 hours of lost production during the 3rd quarter of 1999. Repair of the leak required the complete removal of insulation inside the exchanger housing. The leak was the result of three failed welds on one of four identical headers within the exchanger enclosure (Figures 5.1.4C&D). The three process piping welds on each of the four headers provided the only structural support for the headers, causing the welds to be under constant stress. Furthermore, close inspection of the welds determined that they were not held to acceptable quality control standards during the original installation. Even the welds that had not yet failed displayed the potential for future problems if left uncorrected. The fact that these low quality welds were responsible for both the physical support of the headers as well as maintaining pipe integrity led the investigation team to conclude that equipment design and poor quality control methods employed during installation were the root causes of the incident. As a result, all twelve welds were re-worked and additional structural supports for each of the four headers were installed. No future downtime as the result of similar failures is expected, but it is worth noting that future installations of similar exchanger systems need to incorporate designs that provide sufficient structural support and maintain strict quality control standards when making cryogenic welds.
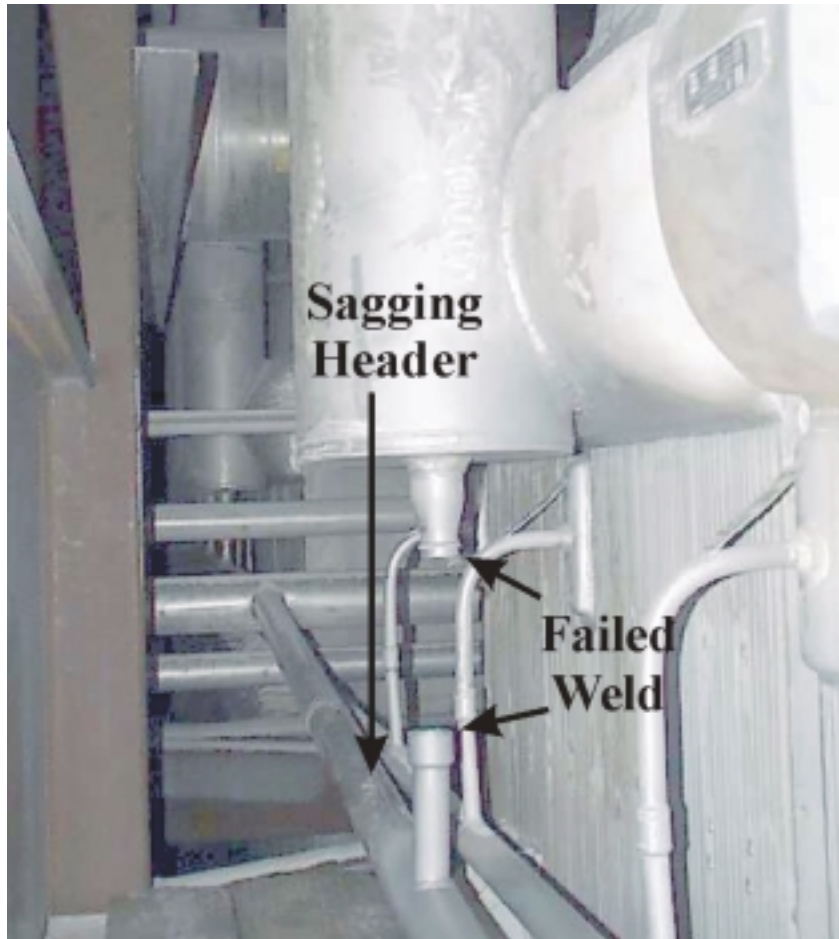
Figure 5.1.4C: Damage Inside Enclosure for Main Exchangers

Figure 5.1.4D: Derime Header Failed Weld

Other suggestions for future installations of air cooling and liquefaction systems include added instrumentation redundancy with regard to temperature monitoring around the main exchangers and a more reliable means for automatic control of the compressor/expander system outlet temperature. The currently installed temperature instrumentation for the main exchangers is buried deep within the insulation of the enclosure and is inaccessible for troubleshooting, repair or replacement. Over time, the temperature instrumentation has begun to fail, making plant start-ups increasingly more difficult. Considering the low cost of temperature instrumentation when compared to the value of accurately knowing the temperature of a cryogenic process, it would be appropriate to install as many as two times the current number of indicators in future projects. Another suggestion concerns the importance of very tight expander outlet temperature control. It is absolutely critical to maintain the temperature above the point of liquefaction (see Section 5.1.4.1 for details), but operation too far above liquefaction prevents the plant from being optimized. However, a manual throttling valve that allows some of the flow to bypass sub-cooling in the main exchangers is the only means for controlling the expander outlet temperature. An automated method of temperature control should be specified as part of the design package of the next compressor/expander subsystem.

With the exception of the main exchanger leak that occurred after the Operating Period, the air cooling and liquefaction system has been reliable and relatively maintenance-free for the entire commercial operation of the Project. The air cooling and liquefaction system is not expected to cause any further downtime or lost production. Furthermore, maintenance performed in the future is expected to be limited to minor periodic jobs similar to the ones presented in Section 5.1.4.2.

5.1.5 Cryogenic Distillation System

The cryogenic distillation system includes:

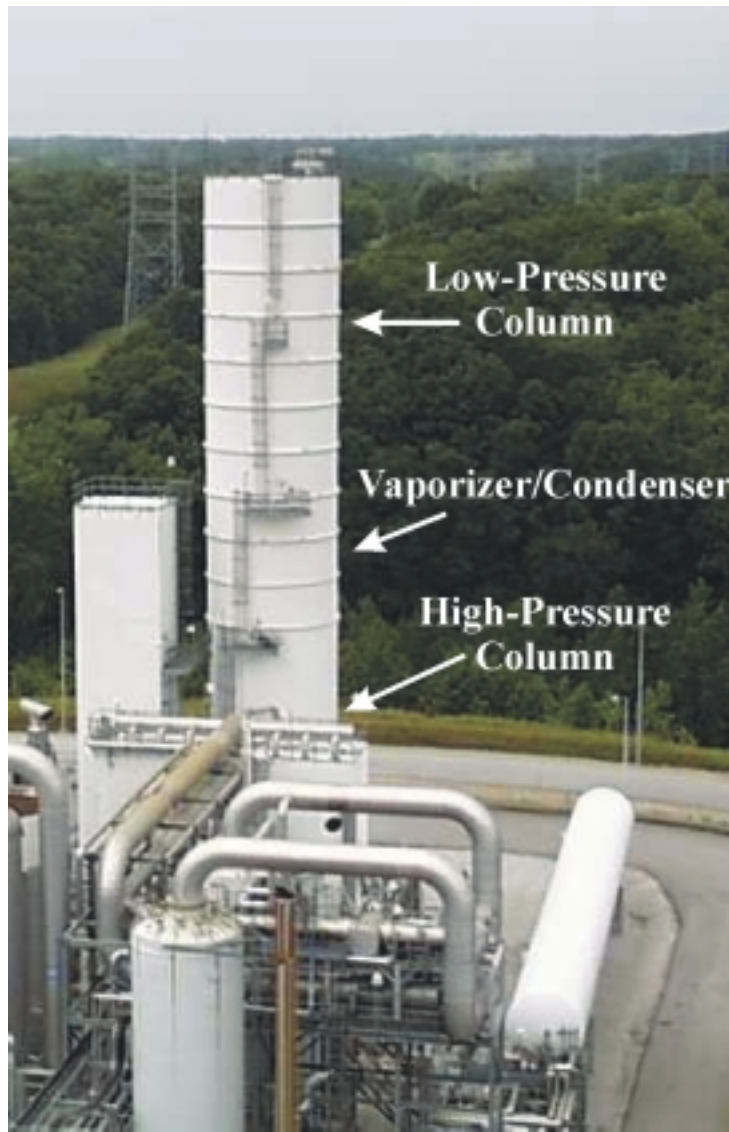| High/Low pressure distillation columns (Figure 5.1.5A)	Vaporizer/Condenser (Figure 5.1.5A)
Nitrogen reflux sub-cooler	Rich liquid sub-cooler
Rich liquid separator	Lean oxygen separator

Figure 5.1.5A: High-Pressure and Low-Pressure Columns

5.1.5.1 System Modifications

The only system modification implemented in the cryogenic distillation system was the addition of a high-pressure oxygen recycle line from the oxygen compression system to the low-pressure column. This modification allows a slip-stream of product oxygen to be taken from the discharge of the oxygen compressor, condensed in the main exchanger, expanded to the point of liquefaction, and emptied into the low-pressure column (see Section 5.1.4.1 for more details.). The modification was implemented early in 1995, prior to the commercial operation of the plant.

No downtime was attributed to the cryogenic distillation system and no major maintenance was performed on any of the system equipment prior to the Operating Period.

5.1.5.2 Operating Experience Overview

The total downtime associated with the cryogenic distillation system during the Operating Period was 77 hours. Operating experience for individual pieces of equipment is given below.

Low-pressure Column

On July 19, 1998 four hours of lost production were experienced as a result of a loss of level indication for the low-pressure column. Investigation revealed that the transmitter level tap had plugged with solid carbon dioxide. The level indication began reading normally after the level tap was bled to atmosphere. It was common for this particular level tap to periodically plug and require a purge to atmosphere as a corrective action. However, a level tap plug had not been experienced for an extended period prior to this incident, consequently, the Operations team failed to recognize this action as a viable response and manually initiated a production interruption. The root cause was determined to be human error and training was performed in order to prevent a recurrence. There were no other significant operational problems nor any major maintenance performed in association with the low-pressure column during the Operating Period.

Rich Liquid Separator

Sixty-two (62) hours of lost production were experienced beginning on August 11, 1998 due to difficulties and delays in recovering from a plant upset that had occurred as a result of an

incident in the air compression system on August 9, 1998. (See Section 5.1.1.2 for details of the August 9, 1998 incident.) There were several contributing factors to the difficulties and delays and they spanned more than one system within the air separation plant. However, two problems localized to the cryogenic distillation system are noteworthy. Specifically, the flow of rich liquid effluent from the sump of the high-pressure column to the low-pressure column was interrupted due to either carbon dioxide pluggage or vapor-lock within the rich liquid separator. Flow interruption upset operation of both the low-pressure and high-pressure columns for several hours before efforts to re-establish flow proved to be successful. Also, an actuator failure on the pressure control valve for rich liquid feed to the low-pressure column resulted in a significant period of time where column pressure control could be achieved only through manual throttling of the valve. Although no downtime was suffered as a result of the actuator failure, the increased difficulty in pressure controlling the low-pressure column contributed to inefficiencies and delays in getting the overall air separation plant back into full production. Corrective actions identified and implemented in response to the lost production included personnel training on methods for re-establishing flows and improved preventative maintenance for control valve actuators. No other operational difficulties were experienced nor was there any maintenance associated with the rich liquid separator during the Operating Period.

Vaporizer/Condenser

Ten (10) hours of lost production were voluntarily sacrificed on February 2, 1999 in order to perform a safety procedure known as a "vaporizer wash". The vaporizer wash was recommended by the original equipment manufacturer of the air separation plant. The procedure was developed in response to a relevant explosion incident at an international air separation plant owned and operated by other parties. The plant suffering the explosion operated a falling film vaporizer/condenser similar to the equipment employed within the cryogenic distillation system. The purpose of the vaporizer wash was to investigate and quantify hydrocarbon accumulation on the oxygen-side heat exchange surface of the vaporizer/condenser.

Results of the vaporizer wash clearly showed that safe levels of hydrocarbon accumulation had occurred. Safe levels for hydrocarbons are determined to be within the solubility limits of liquid oxygen at operating conditions after the wash test, ruling out the potential for hydrocarbon

precipitation and deposition. Operation of the vaporizer/condenser was completely trouble free and no maintenance was required during the Operating Period.

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: the high-pressure column, the nitrogen reflux sub-cooler, the rich liquid sub-cooler and the lean oxygen separator.

5.1.5.3 Summary and Conclusions

Operation of the cryogenic distillation system has been very reliable and efficient throughout the life of the plant. The only lost production associated with the system occurred during the Operating Period, and most of it was the result of minor problems that are not expected to re-occur and are due to voluntary performance of a safety procedure. No major maintenance has been required and no system modifications have been performed in the cryogenic distillation system since commercial operation of the plant began in 1995.

A total plant "derime" of the cold end of the plant was optionally performed during the 2nd quarter of 1999 as a safety precaution following the vaporizer wash test performed during February of 1999 (see Section 5.1.5.2 for details of the vaporizer wash). Derime is a procedure in which the cryogenic distillation, air cooling and liquefaction, and oxygen mixing systems are warmed with hot, dry, purified air in order to purge all traces of moisture, carbon dioxide, and hydrocarbons from the cold end of the plant. It is a costly procedure in terms of energy consumption and time commitment. However, it must be performed any time there is a potential for moisture, carbon dioxide, or hydrocarbons to be present within the plant. Opening of process equipment and failure to maintain a pressure above ambient within the cold end of the plant during an outage are examples where derime may be necessary. In this particular case, derime was voluntarily performed as a result of the plant's high commitment to safety. However, derime was again performed during the 3rd and 4th quarters of 1999.

The 3rd quarter derime was in response to a leak repair within the air-cooling and liquefaction system (see Section 5.1.4.3 for details of this incident). The 4th quarter derime was due to a long

outage where loss of liquid level in both the high-pressure and low-pressure columns prevented the ability to maintain positive pressure on the plant.

The addition of better instrumentation redundancy is a noteworthy suggestion for future plants requiring a cryogenic distillation system. A redundant level transmitter tap would have prevented four hours of lost production during the Operating Period. Also, a redundant pressure indication for the low-pressure column would have facilitated production optimization on a few occasions. Finally, redundant temperature indications on process streams around the high-pressure and low-pressure columns would promote long-term operability of the plant. The currently installed temperature instrumentation in the cryogenic distillation system has begun to fail, making efficient start-up and operation of the plant difficult. The fact that the temperature indicators are inaccessible for troubleshooting, repair, or replacement, means that the difficulties will probably only worsen with time. Redundancy should have been designed and installed prior to plant start-up, and definitely needs to be considered for future projects.

One final suggestion for future installations is the addition of an improved means for filling the low-pressure column with liquid nitrogen. Currently, liquid nitrogen can be transported into the low-pressure column only if the main air compressor is available to provide the motive force. Several instances occurred during outages, indicating the benefit of filling the low-pressure sump without the aid of the main air compressor (e.g., addition of liquid nitrogen dilutes sump liquid increasing hydrocarbon solubility and prevents the need to purge during outages). However, it is difficult to justify the energy consumption as well as the additional start imposed upon the large air compressor motor just to prevent disposal of the liquid.

5.1.6 Oxygen Mixing System

The oxygen mixing system includes:

Liquid oxygen pumps	Mixing column (Figure 5.1.6A)
Liquid oxygen filter (Figure 5.1.6A)	Reactivation filter
Lean oxygen sub-cooler	Liquid oxygen reactivation heater

Figure 5.1.6A: Equipment Associated with Oxygen Mixing System

5.1.6.1 System Modifications

The only system modifications and maintenance performed in the oxygen mixing system prior to the Operating Period were limited to the liquid oxygen pumps, the flanges and valves immediately surrounding the pumps, and the pump enclosures. Both of the liquid oxygen pumps necessitated re-builds near the end of 1997 as a result of bearing failures.   The original equipment manufacturer of the pumps was unable to determine the failure mode from post-mortem investigation. The first pump that failed was re-built and returned to operation without modification. However, a new impeller shaft bearing design was available at the time of the second pump failure, and was implemented as an upgrade prior to placing it back in service.

In addition to pump re-builds, the insulated pump enclosures were entered on several occasions in order to repair valves and flange leaks. Leaks occurred so often that portions of the enclosures were damaged due to thermal stresses and received metallurgy upgrades in an effort to minimize future damage.   Other minor modifications were made to the enclosures in order to facilitate better access for pump extractions and leak repairs.

5.1.6.2 Operating Experience Overview

There was no downtime associated with the oxygen mixing system during the Operating Period. Operating experience for individual pieces of equipment follows.

Liquid Oxygen Filter

Inspection of the liquid oxygen filter, performed on a routine outage during the Operating Period, revealed attrition of the hydrocarbon adsorption media. Analysis by the original equipment manufacturer of this and similar problems at other plants resulted in the conclusion that fracturing of the filter media might have occurred during periodic filter regeneration. A new filter media with a higher thermal shock and fracture resistance was installed while the filter regeneration procedure was modified to minimize the thermal stresses contributing to fracture.

Liquid Oxygen Pumps

Maintenance issues surrounding the liquid oxygen pumps continued throughout the Operating Period. Only one year after the first pump re-build, both pumps required a second re-build. Once again, the original equipment manufacturer was unable to determine a root cause for the failures. However, the redesigned bushing arrangement was implemented in one of the pumps in order to evaluate its potential as a permanent solution.

Operational problems with the liquid oxygen pumps also persisted throughout the Operating Period. Several instances of high differential pressure across the suction strainers for the liquid oxygen pumps during operation forced frequent on-line swaps of the pumps. Possibilities for the source of the suction strainer restrictions included build-up of sublimated carbon dioxide, moisture (as ice), or the presence of solids suspended within the liquid oxygen. However, it was often possible to temporarily alleviate the suction strainer differential pressure by means of liquid oxygen back-flush through the suction strainer. Had the source of the differential pressure been carbon dioxide or ice, then a complete derime procedure (in which the pump circuit is warmed and purged with dry, hot gas) would have been required in order to significantly affect the differential pressure. Those back-flush experiences, coupled with the fact that small amounts of solids were recovered from the suction strainers during maintenance procedures, suggested that the source of the differential was indeed solids and may have been contributing to the poor reliability of the liquid oxygen pumps. Lab analysis performed on solids removed from the suction strainers of the liquid oxygen pumps showed that the properties of the solids were somewhat consistent with the properties of the hydrocarbon-adsorption filter media contained within the liquid oxygen filter. It is feasible that the attrition of fractured filter media observed during the Operating Period may have resulted in filter media re-circulating back into the suction of the liquid oxygen pumps (see the operating experience for the liquid oxygen filter for more details regarding how this attrition was addressed.). However, it is also feasible that carbon dioxide adsorption media contained in the air purification system may have migrated through the cold end of the plant and found its way into the suction of the pumps. Regardless of the source of the solids, a project was implemented on both pumps during the Operating Period to allow solids purge from the suction strainer through a low-point drain to the cryogenic liquid disposal system during a back-flush of the pump.

<u>Other Equipment</u>
There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: the mixing column, the lean oxygen sub-cooler, and the liquid oxygen reactivation heater.

5.1.6.3 Summary and Conclusions

All equipment within the oxygen mixing system, with the exception of the liquid oxygen pumps and their isolation valves, has been reliable, maintenance-free, and easy to operate throughout the history of the plant. However, the extremely poor reliability of the liquid oxygen pumps, coupled with the fact that they are not easily accessible for maintenance due to their cryogenic service, has made them a recurring source of expensive maintenance as well as an ever-present threat to plant availability. Both pumps were re-built prior to the Operating Period, re-built again one year later during the Operating Period, re-built a third time during the 3rd quarter of 1999, and a fourth re-build was required in the 1st quarter of 2000. Despite efforts to eliminate solids within the liquid oxygen stream (see the operating experience for the liquid oxygen filter for more details), manufacturer recommended bushing modifications, and continuing investigation efforts involving the original equipment manufacturer, the reliability of these pumps has not increased over the life of the plant. Thus far, an effective means for increasing the reliability of the pumps has neither been implemented nor developed.

Future installations of oxygen mixing systems within gasification plants should closely scrutinize the design and configuration of the liquid oxygen pumps. Specifically, the ability to perform an on-line swap of a liquid oxygen suction strainer and a finer mesh size for the suction strainer quite possibility could extend the life of the pumps. Also, selection of a fracture resistant adsorption media for the liquid oxygen filter, as well as a conservative regeneration procedure, should help to prevent solids accumulation in the first place.

5.1.7 Nitrogen Handling and Storage System

The nitrogen handling and storage system includes:

Liquid nitrogen storage tank (Figure 5.1.7C)	Liquid nitrogen pumps (Figure 5.1.7A)
Nitrogen steam vaporizer (Figure 5.1.7B)	High-pressure nitrogen fill pump (Figure 5.1.7B)
High-pressure nitrogen cylinders (Figure 5.1.7C)	

Figure 5.1.7A: Liquid Nitrogen Pumps

Figure 5.1.7B: Liquid Nitrogen Pump Skid

Figure 5.1.7C: Liquid Nitrogen Storage Tank and High-pressure Cylinders

5.1.7.1 System Modifications

Several small modifications were made in the nitrogen handling and storage system prior to the Operating Period. Most of them were very minor in nature and included: enlarging drains on pressure safety valve vents; additions of small manual block valves for better isolation; condensate piping modifications; temporary hoses replaced with permanent pipe; addition of insulation; purge valve improvements; and firewater sprinkler system improvements.

Major maintenance performed in the nitrogen handling and storage system prior to the Operating Period was limited to re-builds of both liquid nitrogen pumps due to bearing failures. Other minor maintenance included: repair of oil leaks; valve repairs; pressure safety valve calibrations; repair of seal leaks on the high-pressure nitrogen fill pump; and replacement of the original vaporizer condensate sump pump due to temperature rating.

Although the nitrogen handling and storage system was very reliable prior to the Operating Period, difficulty maintaining sufficient inventory of liquid nitrogen was recognized early in the Project. Lack of nitrogen availability often caused production delays for the plant. For example, one such documented incident occurred on January 16, 1998 just prior to the start of the Operating Period. The liquid nitrogen storage tank was depleted of inventory on that day, resulting in delayed heat-up of the dry char system for a total of 12 hours lost production. Poor management of liquid nitrogen reserves coupled with frequent periods of high volume nitrogen consumption within the gasification plant preceding the incident were the identified root causes. In general these root causes were the same issues that plagued the plant from the early days of the Project and needed to be overcome. A third-party nitrogen supplier was contracted to supply liquid nitrogen to the plant when low reserves and high volume usage necessitated purchase of liquid nitrogen. Also, it was assumed that periods of high volume nitrogen usage would become less frequent over time as operating experience was gained in the gasification process.

5.1.7.2 Operating Experience Overview

The total downtime associated with the nitrogen handling and storage system during the Operating Period was 15 hours. Operating experience for the entire system follows.

On January 4, 1999, the effects of a severe snowstorm culminated with the only lost production attributed to the nitrogen handling and storage system during the Operating Period. Several weather-related problems in the Gasification Island and a utility power blackout forced the plant into an extended period of abnormally high nitrogen consumption. Unfortunately, the utility power blackout also placed the entire air separation plant into an upset condition, restricting its ability to produce nitrogen. The high level of consumption coupled with the production difficulties led to a depletion of the liquid nitrogen tank inventory. Efforts to procure liquid nitrogen through third-party suppliers were unsuccessful due to emergency road conditions consisting of ice and snow. Fifteen hours of lost production were suffered as a result of the delays in nitrogen availability. However, it is worth reiterating that the production loss did not occur as a result of any operational, reliability, or performance issue related to equipment within the nitrogen handling and storage system, but rather was fully attributable to weather conditions beyond the control of plant personnel. As a result of this event, agreements were secured with two backup liquid nitrogen suppliers to facilitate delivery if the contracted supplier failed to meet plant needs.

A telemetry system was installed on the liquid nitrogen storage tank during the Operating Period in order to allow the contracted third-party liquid nitrogen supplier to monitor the tank's inventory. The system allows the nitrogen supplier to better anticipate the plant's delivery needs, thus helping to alleviate production delays suffered as a result of liquid nitrogen unavailability.

A project to reduce maintenance costs as well as to increase efficiency in the nitrogen handling and storage system was implemented during the Operating Period. The nitrogen steam vaporizer bellows trap and condensate pump systems were eliminated in favor of a float and thermostatic steam trap. The bellows trap system required sub-cooled condensate for effective steam separation, which resulted in poor vaporizer performance due to condensate holdup within the vaporizer shell. In addition, the condensate sump pump, which had been replaced prior to the Operating Period, continued to suffer poor reliability and high maintenance cost. The new steam trap system installation provided a less complicated system that is easier to operate, enhances nitrogen delivery, results in energy and maintenance savings from the elimination of the unreliable condensate pump, sump and field control station.

A final project, implemented during the 4th quarter of 1998, positively affected safety in the nitrogen handling and storage system. The vent for the liquid nitrogen storage tank discharges vaporous nitrogen to atmosphere at a very cold temperature (as low as –250°F). As the vapor is vented, moisture condenses on all the surrounding cold surfaces and on the ground near the discharge point. Frequent pluggage of the vent often resulted in intervention by the Operations team utilizing steam hoses to remove ice from the vent release area. The potential for burns resulting from ruptured steam hoses was significant. Installation of a steam-jacketed tip on the vent now prevents discharge vapor from promoting ice formation. A jet of steam is also directed to the ground in the discharge vicinity to prevent buildup from below. It is not safe to discharge from a vent above ground as the potential for cryogenic liquid spray and serious burns is unacceptable, though this approach would more easily have remedied the tip frosting problems.

There was no downtime or operational problems associated with any specific piece of equipment within the nitrogen handling and storage system during the Operating Period. Maintenance performed was limited to a minor leak repair on the high-pressure nitrogen cylinders.

5.1.7.3 Summary and Conclusions

The nitrogen handling and storage system has been very reliable and reasonably easy to operate over the history of the Project. The only major maintenance performed thus far has been limited to a re-build of both liquid nitrogen pumps prior to the Operating Period. The only other modification to the system that has occurred since the end of the Operating Period involves the addition of flanges on the discharge piping of the liquid nitrogen storage tank in order to eliminate the need to make line cuts for minor pressure safety valve maintenance.

Operating experience has revealed two specific opportunities for future improvements of the nitrogen handling and storage systems. First, a larger capacity storage tank may have provided significant dollar savings over the life of this facility. Several occurrences of a shortage of nitrogen caused great difficulties for the entire plant. Although deficiency in nitrogen availability rarely culminated in documented production losses, availability problems were quite

often the source of heat-up delays during plant start-ups and cooldown delays when entering outages.

Furthermore, extended plant run-time often afforded excess nitrogen production to the point that the storage tank was filled to the limit. Production excesses in those cases had to be wastefully vented for the remainder of the run-time due to a lack of storage capacity. Finally, when nitrogen consumption exceeds generation capability, typically during heat-up, cooldown and occasionally during outages, a package nitrogen generating system independent of the ASU could have proven cost effective and warrants consideration for the next plant.

Secondly, the addition of a smaller capacity liquid nitrogen pump has the potential to greatly reduce nitrogen consumption on frequent occasions. The currently installed pumps have a minimum turndown associated with them. In other words, the pump must flow a minimum amount of liquid nitrogen or it will trip off line. If operation of a liquid nitrogen pump is required in order to provide vaporized nitrogen for plant consumption, then either an amount of liquid nitrogen greater than the minimum turn-down must be vaporized or some of the pump discharge must be recycled back to the nitrogen storage tank. In cases where consumption requirements are very small, it is wasteful to do either. Excess vaporized nitrogen is vented to the atmosphere and high-pressure liquid nitrogen recycled back to the tank promotes flashing, which is also vented. However, it would not be wise to replace the current pumps with smaller ones, because their full capacity is often required. Installation of an additional smaller pump whose maximum capacity is slightly greater than the minimum turndown of the current pumps would be ideal.

5.1.8 Oxygen Compression System

The oxygen compression system includes:

AC synchronous motor driver (Figure 5.1.8A)	Lube oil system (Figure 5.1.8A)
Vibration monitoring equipment	Two inter-process-stage coolers
Oxygen bypass cooler	

Product oxygen that has been purity controlled in the mixing column and warmed in the main exchangers is received and pressurized by a six-impeller, three-process-stage radial turbo-compressor for oxygen service (each process stage contains two impellers).

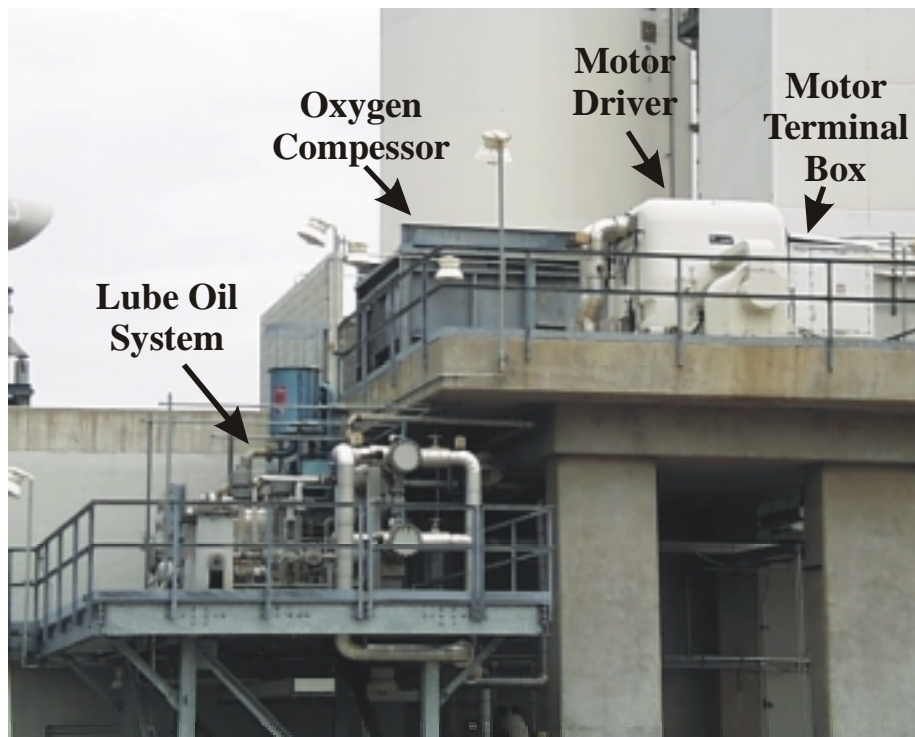**Figure 5.1.8A: Oxygen Compressor**

5.1.8.1 System Modifications

No system modifications were made to the oxygen compression system prior to the Operating Period. However, concerns over moisture detected within the motor driver enclosure prompted

steps to guard against a potential future failure of the motor windings. Also, several instances of plant downtime prior to the Operating Period caused by electrical fuse failures within the control and instrumentation subsystem prompted a full audit of fuse coordination within the entire subsystem. Details of these significant procedural changes and preventative actions taken prior to the Operating Period follow.

A routine inspection of the oxygen compressor's motor driver prior to the Operating Period showed evidence of moisture within the motor enclosure. The moisture was visible in the bottom of the enclosure as well as detected in the insulation of the motor windings through standard electrical testing. Although it did not lead to any downtime prior to the Operating Period, moisture within the windings of such a large, high-voltage motor was addressed with top priority due to its potential to cause a catastrophic failure of the windings. It was determined that the fiberglass motor enclosure was inappropriately designed for outdoor service. Hairline cracks and pinholes around rivet points and seams in the enclosure provided possible entrant points for moisture. As an immediate action, care was taken to seal the enclosure through the application of a commercially available sealant. The moisture presence was severe enough that steps were also taken to dry out the insulation of the motor windings via a process in which direct electrical current was forced through the windings. The current generated heat within the windings driving off the moisture. Further investigation revealed that the electric heaters within the motor enclosure were improperly installed during construction and had never functioned. The installation error was rectified, and the heaters began functioning as designed. As an added precaution, a procedural change was made to include removal of the cooling water flow from the motor immediately after a shutdown of the machine. It is believed that the cooling water flow significantly counteracted the positive effects of the motor enclosure heaters, and may have promoted moisture condensation during the winding cooldown period immediately following the shutdown of the oxygen compressor.

Blown electrical fuses within the control and instrumentation subsystem prior to the Operating Period caused several instances of downtime and production delays for the oxygen compression system. Exact numbers for the amount of lost production related to these fuses are not known for the time prior to the Operating Period. This is mainly because many of them occurred very early

in the Project and most of the production delays were just a few hours in duration as well as spread across several of the air separation systems. Nevertheless, all incidents of fuse failures prior to the Operating Period can be attributed to one overall root cause: the control and instrumentation subsystem was poorly designed and improperly installed by the project subcontractor. This is evidenced by the fact that instances of undersized fuses resulted in nuisance trips of critical valves. Instances of oversized fuses allowed minor instrumentation issues to overload upstream power supplies resulting in cascading trips of equipment. Also, instances where the wrong fuses were installed, even though the proper fuse size was specified in the design, resulted in similar failures. Load imbalance also played a role in production loss. It was found that some power supply circuits were overloaded while others were under-utilized. A full audit of fuses, coordinated with redistribution of some power supply circuits within the air separation unit's control and instrumentation subsystem, was performed prior to the Operating Period in response to the multiple failures. All fuse issues pertaining to the original design and installation of the control and instrumentation subsystems were addressed prior to the commencement of the Operating Period.

5.1.8.2 Operating Experience Overview

The total downtime associated with the oxygen compression system during the Operating Period was 70 hours. Operating experience for the oxygen compressor follows.

On June 17, 1998 a total of 26 hours of lost production occurred as a result of a blocked orifice in the oil return line from the oxygen compressor coupling. This return line allows oil to drain out of the coupling housing and flow back to the oil reservoir, under gravity, assisted by the vacuum oil recovery system. The orifice in the line had become blocked with debris promoting oil accumulation within the coupling housing.

The rotation of the coupling at over 11,000 rpm (revolutions per minute) churned and agitated the pooled oil to the point that boiling oil vapor began to spray from the coupling guard vent. No damage to the machine was encountered, but the presence of an oil cloud in the immediate vicinity of pressurized high-purity oxygen resulted in shutdown of the oxygen turbo-compressor in the interest of safety. Investigation determined that placement of the orifice at the bottom of

the orifice plate had allowed residual debris from previous preventative coupling maintenance to obstruct the flow path of the oil. The situation was remedied by locating a second orifice in the center of the plate. No further downtime as a result of similar causes is expected.

On August 4, 1998 a trip of the oxygen compressor caused by a loose power supply connection resulted in 9 hours of production loss. Several of the trip interlocks for the oxygen compressor are fed from the same power supply. According to high-speed data archived by the control system, all of those interlocks tripped the oxygen compressor at precisely the same instant. Investigation revealed that the power supply wire to the interlocks had come loose from its termination point. The location of that particular terminal strip is in a high vibration area of the plant. The power supply wire was re-secured and all relevant termination points were tightened.

On August 15, 1998 a blown 15kV electrical fuse in a potential transformer located in the motor control center's high-voltage switchgear triggered a trip of the oxygen compressor (the main air compressor also tripped for the same reason). The incident led to 35 hours of plant downtime. The driver motors for both the oxygen compressor and the main air compressor rely upon a signal from this potential transformer in order to control their synchronous fields. Loss of the transformer's fuse forced them both to trip instantaneously. Rigorous electrical testing of the potential transformer confirmed no problem with the transformer. The possibility of a defective fuse was also ruled out, as this was the second incident involving the same fuse location in just over eight months (the first caused 33 hours of downtime prior to the Operating Period on February 5, 1998). Although no definitive explanation could be found for the blown fuse, the potential transformer was swapped with an identical one from a less critical service. No further problems with this device have been experienced or are anticipated.

A vibration data collection system was installed on the oxygen compressor during the Operating Period. The new system added historical archiving, event capturing, and analytical tools to the already existing vibration monitoring equipment. The increased functionality of the vibration equipment now allows for predictive maintenance techniques to be employed through monitoring of data trends and enhances the accuracy of troubleshooting in response to vibration related events.

Maintenance on the oxygen compressor during the Operating Period was limited to normal preventative maintenance procedures (PM's) performed during planned outages. Some of the tasks performed during these PM's included: electrical measurements on the motor, motor enclosure inspections and weather proofing, coupling and bull-gear inspections, lube oil system inspection and minor leak repair, and pressure safety valve calibrations.

5.1.8.3 Summary and Conclusions

In summary, the oxygen compressor has been very reliable with regard to all of the physical hardware associated with it. However, the overall reliability of the oxygen compression system has suffered tremendously as a result of issues surrounding electrical fuses and power supplies. Most of the downtime credited to the oxygen compression system prior to, within, and after the Operating Period could have been avoided had more care been exercised in the design and installation of the control and instrumentation subsystem. Future gasification projects should pay particularly close attention to quality control in the detailed design and installation of the control and instrumentation subsystem, especially if the work is to be contracted to a third party.

In addition to the problems related to fuses and power supplies in the control and instrumentation subsystem, difficulties in preventing moisture from penetrating the oxygen compressor's motor enclosure have been a serious concern with regard to protecting against a costly motor rewind. A dry gas purge installed in the motor enclosure during the first quarter of 1999, enclosure heater repairs, enclosure weatherproofing efforts, and procedural changes pertaining to cooling water flows have all been credited with preventing moisture from entering the motor enclosure.

One noteworthy project that has been implemented in the oxygen compression system since the Operating Period is the installation of baffle plates in the oxygen compressor coupling guard. The purpose of these new baffle plates is to prevent heat buildup caused by air friction with the wall of the coupling guard resulting from high-speed coupling rotation (known as windage). High coupling guard temperatures tend to carburize (or coke) the oil, reducing its lubrication properties. The new baffle plates rectified the problem by breaking up the airflow that causes the heat build up (eliminating windage).

One last point of concern with regard to the reliability of the oxygen compressor involves the lack of redundancy of vibration probes located within the machine. A few sporadic, vibration related, nuisance trips suffered throughout the history of the oxygen compressor have prompted evaluation of the need for added redundancy in the vibration monitoring subsystem. The evaluation is currently underway.

All maintenance performed on the oxygen compression system since the end of the Operating Period, and anticipated in the future, is limited to the preventative maintenance work discussed at the end of Section 5.1.8.2. The system is expected to be very reliable in the future.

5.2   Coal Handling

5.2.1   Coal Hopper System

The coal hopper system includes:

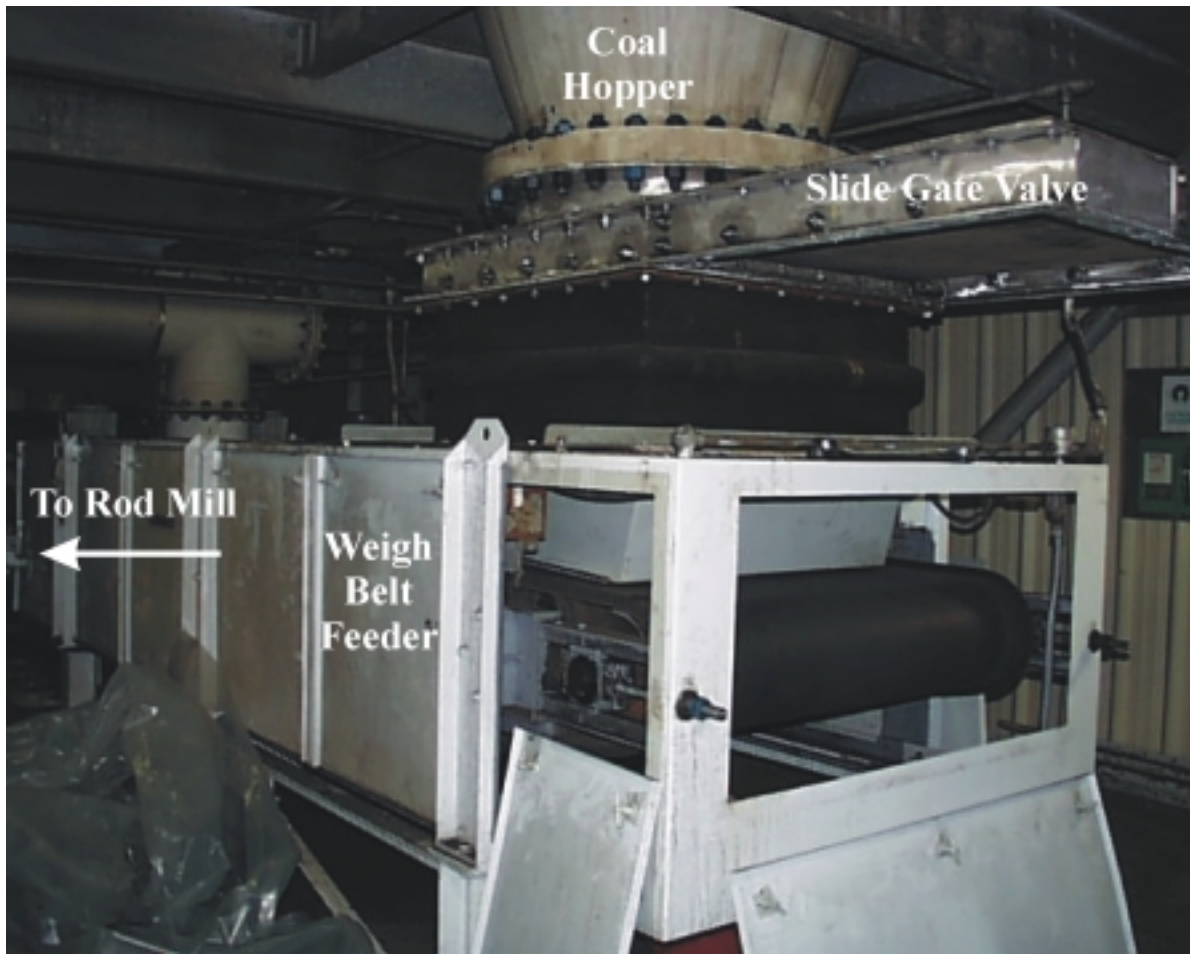| Coal sampler	Coal hopper (Figure 5.2.1A)
Weigh belt feeder (Figure 5.2.1A)	

Figure 5.2.1A: Weigh Belt Feeder

5.2.1.1 System Modifications

The coal hopper system performed as expected since plant start-up in 1995 with only minor modifications. This might be expected, as all of the types of equipment used in this process system are well proven in previous gasification plants operated in the development of the E-Gas^{TM} technology, as well as in similar solids processing industries. No plant downtime or production turndown has been attributed to this system. However, erosion of the coal hopper inlet chute was a recurring maintenance problem throughout the first two years of operation. In August of 1997, the inlet chute was replaced and lined with an erosion-resistant material, eliminating the erosion problem and subsequent near-term maintenance expenditures.

The weigh belt feeder motor required extensive maintenance attention in late 1996 and early 1997 due to failure of the coal hopper bottom outlet isolation slide gate valve. The resulting weight of the coal residing on the weigh belt feeder belt created an overload condition for the weigh belt feeder motor and subsequent attempts to start the motor resulted in motor failure. To address the overload condition, operating procedures for shutting down the plant were modified to empty the coal hopper and, thus, avoid starting the weigh belt feeder under load. Since replacement of the motor in August of 1997 and implementation of the new plant shutdown procedures, the problem has not recurred.

Although the coal sampler resulted in no plant downtime, it had a history of poor reliability and unacceptably high operating and maintenance costs. The coal sampler was taken out of service (i.e., bypassed) in late 1997. To provide process stream data and to ensure sufficient quality control for operating the plant, procedures were modified and slurry samples from the slurry storage tank were utilized after the removal of the automated coal sampler.

5.2.1.2 Operating Experience Overview

There was no downtime associated with the coal hopper system during the Operating Period. Operating experience for individual pieces of equipment follows.

Coal Sampler

The coal sampler was returned to service in the summer of 1998 when a new coal was introduced to the plant. Plugging problems continued with the unit, but were not as severe as in the past. This was due to almost continuous "hands-on" operation of the coal sampler by laboratory and Operations personnel. Once an analytical baseline was established for the new coal, the sampler was again taken out of service.

Coal Hopper

A small leak occurred on the inlet chute in May of 1998. It was repaired without incident during a subsequent outage. Later in the Operating Period, the coal sampler, although out of service, began promoting erosion at the inlet of the coal hopper due to impingement caused by the disruption of the coal dropping from the coal feed belts. This problem was addressed following the Operating Period.

Other Equipment

There was no maintenance, downtime or operational problem associated with the weigh belt feeder during the Operating Period.

5.2.1.3 Summary and Conclusions

The coal hopper system has operated satisfactorily since its commissioning in 1995. Erosion problems were mitigated by the addition of erosion-resistant linings. Following the Operating Period, the coal sampler was physically dismantled and removed from the system to allow the installation of an erosion-resistant inlet chute to the coal hopper.

Early problems with the weigh belt feeder were resolved procedurally until replacement of the hopper slide gate valve was initiated during the fall of 1999. Corrosion of the slide gate valve carbon steel components started becoming evident throughout 1999 and was identified as a factor in the failure of the slide gate valve to be able to be operated through normal means. Critical slide gate valve components have been upgraded with improved metallurgy to better deal with the corrosive environment, which is likely attributable to the acidic water drainage from the high sulfur coal stored in the hopper above the isolation slide gate valve.

5.2.2 Rod Mill System

The rod mill system includes:

Rod mill (Figure 5.2.2A)	Recycled solids tank
Recycled solids pumps	Recycled water tank
Recycled water pumps	Rod mill product tank
Rod mill product pumps (Figure 5.2.2B)	

The rod mill system was installed as part of the original plant design. It has operated as designed with only minor modification.

Figure 5.2.2A: Rod Mill

Figure 5.2.2B: Rod Mill Product Tank Pumps

5.2.2.1 System Modifications

Modifications to this system were limited to installation of erosion-resistant materials in high wear areas due to erosive nature of the coal and coal slurry and installation of covers over the rod mill outboard trunnion bearing and lube oil reservoir (Figure 5.2.2C) to prevent solids contamination of the rod mill bearing lube oil system.

Figure 5.2.2C: Rod Mill Lube Oil Skid

5.2.2.2 Operating Experience Overview

There was no downtime associated with the rod mill system during the Operating Period. Operating experience for individual pieces of equipment follows.

Rod Mill

The rod mill operated as expected during the Operating Period with the exception of holes being worn into the inlet and outlet chutes due to erosion, and contamination of the bearing lube oil system with solids. Erosion-resistant materials were added to the chutes to alleviate the wear and have performed satisfactorily to date. The covers installed over the outboard trunnion bearing and lube oil reservoir (Figure 5.2.2C) to prevent solids contamination of the rod mill lube oil system have also performed satisfactorily. Routine preventative maintenance during the Operating Period included removing of worn grinding rods, installation of new grinding rods, refilling of gaps between the rod mill liner plates, inspection and measurement for wear of the rod mill liner plates, and tightening of rod mill liner plate bolts that loosen over time.

Recycled Solids Pumps

During the Operating Period, the common discharge piping on these pumps required cleaning, during a plant shutdown, due to accumulation of scale in the piping from minerals dissolved in the water. The scale buildup resulted in diminished flow rates. The diminished flow rates caused some concern while operations ensued, but did not result in downtime for the plant.

Rod Mill Product Pumps

The pumps operated as expected during the Operating Period. Other than preventative maintenance, the only maintenance required was replacement of some fluid end parts due to erosive wear from coal slurry. This is within expectations and is considered normal for the type of service that these pumps are performing.

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: recycled solids tank, recycled water tank, recycled water pumps, and the rod mill product tank.

5.2.2.3 Summary and Conclusions

The rod mill system has operated as designed since plant start-up in 1995. With the exception of having to install wear-resistant materials in high wear areas, only routine and preventative

maintenance has been required to this system. After the Operating Period concluded, repairs were required to the agitator in the rod mill product tank as well as repairs to the floor of the tank. In both cases damage was due to abrasion from the coal slurry. Abrasion-resistant material was also installed on the floor of the rod mill product tank beneath the agitator. With this modification, only routine maintenance is anticipated for this system in the foreseeable future.

5.2.3 Slurry Storage Tank System

The slurry storage tank system includes:

Slurry storage tank	Slurry storage tank agitator (Figure 5.2.3A)
Slurry recirculation pumps (Figure 5.2.3B)	

Figure 5.2.3A: Slurry Storage Tank Agitator

Figure 5.2.3B: Slurry Recirculation Pumps

5.2.3.1 System Modifications

No modifications were made since commissioning and start-up of the system prior to the Operating Period. Prior to the Operating Period, on one occasion, the slurry storage tank was emptied for removal of settled solids. Erosion damage to the blades on the slurry storage tank agitator was repaired at that time.

5.2.3.2 Operating Experience Overview

The total downtime associated with the slurry storage tank system during the Operating Period was 3 hours. Operating experience for individual pieces of equipment follows.

Slurry Recirculation Pumps

The slurry recirculation pumps caused a total of 3 hours downtime during the Operating Period due to settled solids plugging the pump suction. Also, wear-resistant material was applied to the discharge piping of the pumps to combat erosive wear from the coal slurry. Normal, expected replacement of some fluid end parts due to wear from coal slurry was also necessary during this period.

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: slurry storage tank, slurry storage tank agitator.

5.2.3.3 Summary and Conclusions

The slurry storage tank system operated as designed during the Operating Period, with the exception of some unwanted settling of solids in the slurry storage tank, creating a "fillet" around the inner circumference of the tank wall. This situation is not uncommon in industries that process solids-containing slurries. However, to address the situation, after the Operating Period, the blades on the slurry storage tank agitator will be replaced with larger diameter blades to increase the agitation of the coal slurry, within the horsepower limitations of the existing agitator drive motor, in an effort to mitigate the amount of solids settling along the tank walls.

5.2.4 Slurry Feed System

The slurry feed pump system includes:

First stage gasifier feed pumps (Figure 5.2.4A)	Second stage gasifier feed pumps
Primary slurry heater	Secondary slurry heaters
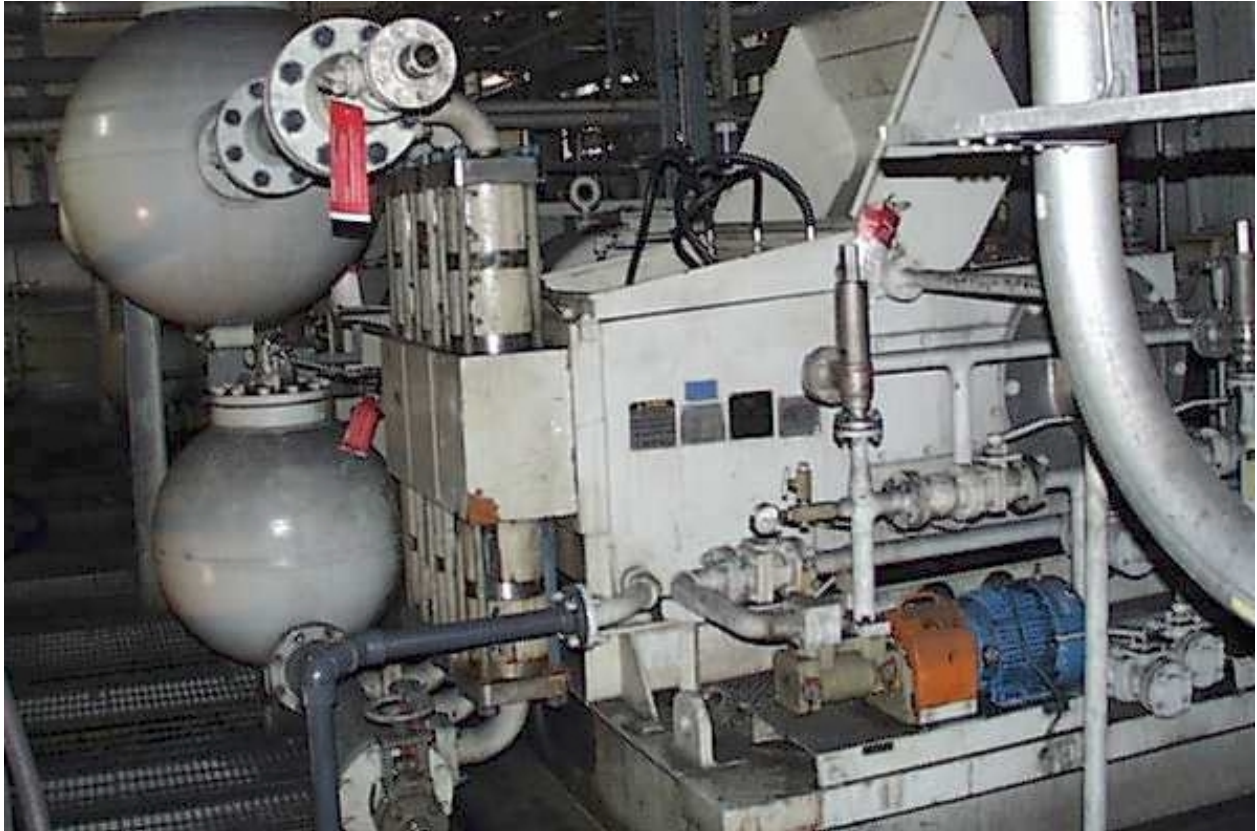
Figure 5.2.4A: Typical First Stage Reactor Feed Pump

5.2.4.1 System Modifications

During the Operating Period, two modifications were made to this system. Selected valves in the system were replaced with upgraded metallurgy and some of the flow meters in hot slurry service were removed and other means to monitor the slurry flow to the gasifier were initiated.

The slurry heater system performed adequately with minimal modifications. No plant downtime or production turndown has been attributed to this system. Fouling of the slurry heaters began to

manifest itself during the second year of operation. The deposits were readily removed by hydro-blasting with a high-pressure water source in August of 1997.

5.2.4.2 Operating Experience Overview

The total downtime associated with the slurry feed pumps system during the Operating Period was 33 hours. Operating experience for individual pieces of equipment follows.

First Stage Reactor Feed Pumps

The first stage reactor feed pumps were responsible for a total of 30 hours of downtime for the Operating Period. The pumps operated as expected during the Operating Period, but several problems unrelated to the core gasification technology resulted in loss of syngas production. Slurry lines that froze during a winter outage were not completely thawed after the shutdown and accounted for 18 hours of downtime. Rupture disks, installed to protect the reactor feed pumps' discharge piping from overpressure situations, blew on two occasions, resulting in 3 hours of loss production. Leaking slurry flow meters accounted for 3 hours of downtime. A failed piston flush on one of the pumps also contributed 5 hours of downtime.

Second Stage Gasifier Feed Pumps

The second stage gasifier feed pumps were responsible for 3 hours of downtime during the Operating Period. This downtime was due to a leaking pump discharge slurry flow meter.

Secondary Slurry Heaters

Hydro-blasting was repeated on the slurry heaters in May of 1998. In June of 1998, the softening of the process water supply was discontinued. This resulted in increased fouling of the slurry heaters, especially secondary heaters due to their higher operating temperature. In an effort to reduce maintenance costs, the slurry flow meters downstream of the slurry heaters were removed from service in the fall of 1998 and were replaced with an alternate means of slurry flow measurement. Removal of these temperature-limited flow meters allowed higher slurry operating temperatures. This allows for more efficient operation of the gasifier.

Other Equipment

There was no maintenance, downtime or operational problem associated with the primary slurry heater during the Operating Period.

5.2.4.3 Summary and Conclusions

The slurry feed pumps have operated as designed since the plant went on line. This is partly due to the vast amount of operating and maintenance experience obtained from operating a pilot plant, a prototype plant and another commercial-scale demonstration plant in Louisiana in the course of the development of the current E-Gas^{TM} technology. Other than routine maintenance and replacement of fluid-end parts due to erosion from the coal slurry, no changes were made during the Operating Period. Since the Operating Period ended, metallurgical upgrade of a minor portion of the piston flush lines on each pump was made. It is anticipated, also, that in the future, some of the coal slurry piping will require replacement due to the erosive effects of transporting coal slurry.

The slurry heater system has operated satisfactorily since its commissioning. Fouling of the tubes continues to be a problem and is only aggravated by the hard water supplying the slurry preparation system. Hydro-blasting is required during every major plant outage to mitigate the problem. Chemical treatments to control the rate of scale formation are currently being investigated.

5.3   Gasification

5.3.1   First Stage Gasifier System

The first stage gasifier system includes:

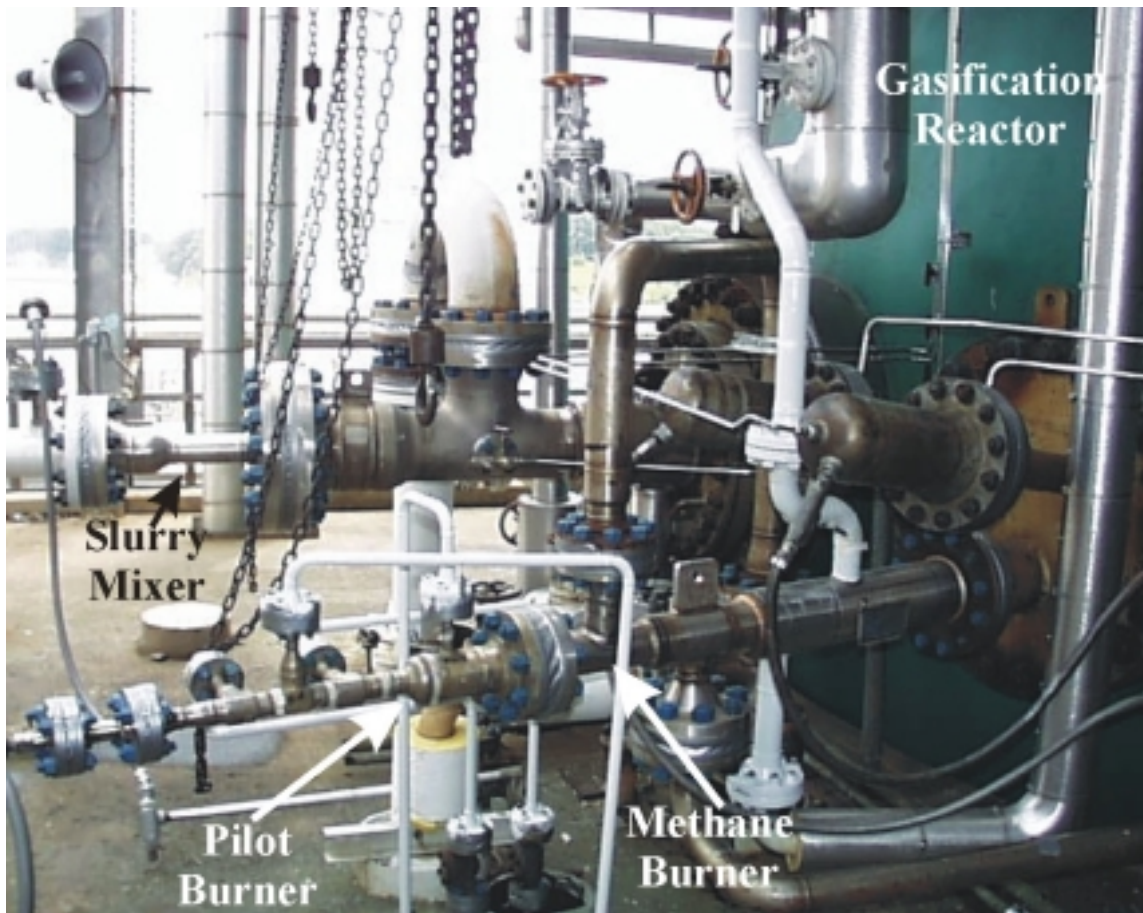| First stage gasifier	Slurry mixers
Methane burners	Water-cooled device system
Dry char recycle nozzles	

Figure 5.3.1A: First Stage Gasifier Feed Nozzles

5.3.1.1 System Modifications

In the first year of operation, the gasifier refractory required extensive maintenance, which included 559 hours of downtime for partial re-bricking. Due to the nature of first year plant operations, the initial gasifier refractory was subjected to an above-normal number of thermal cycles contributing to the extensive damage. While a gasifier cooldown after a gasifier trip and purge could not be optimized further, the methane operation cooldown after a transfer off of coal operation was lengthened to minimize the thermal stresses caused by cycling operations. This procedural enhancement lengthened refractory life of the initial liner while subsequent gasifier liners benefited from this enhancement as well as improved plant operations.

Slurry mixer failures prior to the Operating Period were continually addressed for cause and corrective actions were implemented to extend mixer life as described below. Although slurry mixers can be replaced in less than 24 hours, this requires an interruption in syngas delivery.

During 1996, 353 downtime hours were attributable to slurry mixer failures. The major cause of failures during this period was due to hot gasifier contents backing up into the mixer during start-ups and shutdowns, which was corrected through improved control valve sequence timing during these transitions. In 1997, mixers were replaced proactively based on projected life and, as a result, downtime from slurry mixer failures was reduced to 11 hours. Premature slurry mixer failures caused 92 hours of downtime in the first two months of 1998. Investigation revealed that the transition to coal operations was being made at rates nearly 40% higher than in the past. Slurry flow instability caused by flow meter problems contributed to the flow deviations and resulted in higher than normal flame temperatures during start-up, which accelerated slurry mixer failures. Flow meter calibration and a modification to the coal transfer procedure eliminated these types of failures. In late 1996 and early 1997, the slurry mixers were modified to inject slurry into the gasifier at a different angle, attempting to improve carbon conversion and reduce ash deposition downstream of the first stage gasifier. No appreciable improvement was attributed to this change and the slurry mixers were restored to the previous injection angle. In early 1997, the slurry mixer was modified to allow full oxygen flows at lower backpressures.

System modifications were also required to correct problems in the gasifier water-cooled device system. Excessive flashing across the make-up water control valve resulted in piping failures in October of 1996 and March of 1997, accounting for 288 and 79 hours of downtime, respectively. During a subsequent outage in the spring of 1997, the make-up water supply was altered, which eliminated the destructive flashing.

Although it made no contribution to downtime, the dry char recycle nozzles often required expensive repairs due to cracking and pitting of the nozzle tip. The fact that the nozzles were water-cooled contributed to the expense of the repairs due to the precise internal clearances required for adequate water flow distribution. An uncooled nozzle was installed in December 1997 on a trial basis. An inspection during an outage prior to the Operating Period revealed little damage after about 900 operating hours. The cost advantage of the uncooled design was readily apparent; fabrication costs for a new (uncooled) unit were about the same as the repair cost for a water-cooled dry char nozzle.

5.3.1.2 Operating Experience Overview

The total downtime associated with the first stage gasifier during the Operating Period was 1001 hours. Operating experience for individual pieces of equipment follows.

First Stage Gasifier

The largest single contributor to gasifier downtime during the Operating Period was slag removal difficulties (plugged taphole). Plugging was a result of the introduction of higher ash-fusion temperature coal to the gasifier and constant changing of the feedstock blend. Taphole plugs resulted in 762 hours of downtime during the Operating Period. Gasifier operation was modified to include higher operating temperatures when operating on the new coal and additional training was given to operators to provide for better recognition of which coal blends were being fed to the gasifier, thus ensuring adequate slag flow.

Improper shutdown of the gasifier in December of 1998 resulted in refractory damage in the unit. The need for controlled cooldown was re-iterated to Operations personnel and specific instructions added to the shutdown procedure. Preventative repairs to the refractory extended an

outage and resulted in 42 hours of downtime in early 1999, but probably averted greater downtime in the future.

Slurry Mixers

During the Operating Period, 134 hours of downtime were attributed to the slurry mixers. In early March of 1998, an internal failure of a slurry mixer caused by erosion forced a transfer off of coal operations resulting in 27 hours of downtime. The problem was addressed by proactively changing the internal parts at a higher frequency while more erosion-resistant metallurgy was evaluated. Three scheduled mixer changes occurring in April, October and November of 1998 resulted in 58 hours of downtime. Proactive mixer changes progressively reduced downtime associated with this activity through improved planning and preparation. In August of 1998, following a slurry mixer trip forced by loss of oxygen, numerous attempts to return the plant to coal operation were ended by excessive temperatures in the gasifier. Investigation and correction of this problem resulted in 49 hours of downtime. The cause was attributed to slag accumulation at the discharge of the mixer, which prevented proper mixing of the coal slurry and oxygen. The oxygen preferentially reacted with the syngas, resulting in the internals of the mixer being exposed to excessive temperatures. This problem has been addressed by automating a trip of the slurry mixers at the onset of an oxygen loss. This minimizes the addition of unburned coal and ash, which can obstruct the mixer's flow path on a subsequent start-up.

Methane Burners

Prior to transferring the first stage gasifier to coal operations, the methane burners are used to pre-heat the gasifier system to operating temperatures. Methane burner light-off failures increased during the Operating Period, resulting in 43 hours of downtime. The worst occurrence was in January of 1999 when multiple failed light-off attempts resulted in excessive nitrogen consumption due to the introduction of unreacted oxygen to the system. The root cause of the problem was frozen syngas condensation in the methane lines, which completely obstructed methane flow. The problem has been addressed through improved operating discipline as well as some nuisance alarm elimination. A nuisance alarm during light-off attempts was a causal factor in the methane flow problems going undetected.

Following a taphole-plugging incident in January of 1999, a methane burner failed externally. This resulted in 20 hours of downtime to repair piping components associated with the failed burner. An insufficient purge of the burners during the transfer off of coal operation was a key factor in the incident. The purge has been modified to maintain a positive forward flow at all times when the burners may be exposed to a slagging environment.

Water-Cooled Device System

Although it made no contribution to downtime during the Operating Period, the cooler in the water-cooled device system required maintenance during every outage in 1998 to repair tube leaks. The root cause of the tube leaks was corrosion caused by high operating temperatures with the water-cooled device system in early 1997. Due to the increasing frequency of repairs, the exchanger tube bundle was replaced during the winter 1998 outage. Cooler operating temperatures have prevented any repeat failures.

Dry Char Recycle Nozzles

The uncooled design proved to be largely successful, as it required no maintenance during the Operating Period other than routine inspections. During an outage in May 1998, the face of the remaining water-cooled dry char nozzle was found partially melted. It was replaced with a spare from inventory and remained in place during the remainder of the Operating Period.

5.3.1.3 Summary and Conclusions

The first stage gasifier in and of itself has met, if not exceeded, reliability expectations. The refractory life has improved to the point that the necessity of a spare gasifier in future projects should be evaluated carefully as it presents an opportunity to significantly decrease total plant installed capital cost. Following the Operating Period, neither the gasifier nor its refractory were the cause of any downtime. Refractory work in the original gasifier was completed shortly after the Operating Period and the unit was cured out in the fall of 1999, preparatory to removing the spare gasifier from service and swapping back to the original unit in the spring of 2000.

Slurry mixer reliability has proven to be a major challenge due to the abrasive characteristics of the coal, coupled with being exposed to a hostile environment (i.e., high temperatures, slagging

conditions). Early on, typical mixer lives were well under half that achieved at Dow/Destec's LGTI facility. A major breakthrough was made in slurry mixer design following the Operating Period. A modification was made to the mixer throat and face that more than doubled the best mixer life to that point. Additional mixer improvements under evaluation include the addition of erosion-resistant materials to mixer parts contacting slurry and reduction of pressure drop while maintaining adequate atomization.

The reliability of methane burners has been adequate, although some improvements on purges during coal operation are in order. In the slagging environment present during coal operations, the burners tend to get plugged with slag, making subsequent attempts to return them to service difficult. Additional burner purges during coal operation are being evaluated and will be given due consideration in future designs. Additionally, the use of oxygen to fire the methane burners warrants re-evaluation in future projects. Using oxygen makes the operation of the methane burners contingent on oxygen being available from the air separation unit. This can be both costly due to utility usage to power the ASU and can add to the start-up time when the ASU is on the shutdown maintenance critical path. Air-blown burners would not have this limitation and could provide a more effective heat-up of the gas path due to the higher mass flow.

The reliability of the pilots for the methane burners has been somewhat disappointing. Part of the problem is obstruction of the flame scanner sight paths that occurs during coal operations. The vulnerability of flame scanner systems in a slagging environment is a problem that needs to be addressed in future projects. The pilots themselves are of a rather complicated design; four different feeds supply the pilot, which makes troubleshooting a rather complex task. Like the methane burners, the pilots are dependent on oxygen from the air separation unit. For future projects, a simpler pilot design, capable of running without enriched oxygen, should be evaluated. Of course, both the methane burners and the pilots operating on air versus oxygen would add new equipment in the form of a pressurized air source, so the trade-offs would need to be thoroughly evaluated from an economic perspective.

The uncooled dry char nozzle was deemed a success. It remained in service until fall 1999 (a period of nearly two years) when the dry char recycle piping was rebuilt (the nozzle was sent out

for rebuild and replaced with a new uncooled nozzle). At this time, the remaining water-cooled nozzle was permanently removed from service and likewise replaced with an uncooled nozzle. Some opportunity still exists to further simplify the design of the nozzles, as thermal growth is no longer a consideration in the absence of water-cooling.

5.3.2 Second Stage Gasifier System

The second stage gasifier system includes:

Second stage gasifier	Second stage gasifier slurry feed nozzle
Second stage gasifier quench nozzles	Soot blowers

5.3.2.1 System Modifications

The second stage gasifier performed adequately with minimal modifications. During the first year of operation ash deposition in the second stage was a continual maintenance problem. Reduction of the operating temperature had a very positive effect and decreased the deposition to the point that it was no longer a run-limiting concern. Failure of the slurry feed and quench nozzles by backflow of hot gas was a problem in early operation. Installation of nitrogen purge lines to ensure forward flow when the nozzles are not in service, but are exposed to high temperatures, has prevented this problem from recurring.

Soot blowers were installed in the original second stage gasifier due to a history of severe ash deposition problems at the LGTI facility. Repeated failures resulted in two of the four being removed from service in 1996. No adverse effect on ash deposition severity was noted following the change. Ash management efforts (i.e. reduced operating temperature) were so successful that the remaining two soot blowers were permanently removed from service when the new post gasifier pipe spool (See 5.3.3 for details) was installed in late 1997.

The spare second stage gasifier was rebricked to a larger diameter to reduce gas velocities, thus minimizing impingement of molten ash. This new configuration was brought on line in November 1997 and resulted in a reduction of ash deposition. The redesigned refractory lining also created one "hot spot" (i.e., overheating) on the vessel shell resulting from insufficient refractory lining at an expansion joint. While the hot spot caused no operational problem, cooling water applied to one side of the gasifier shell created uneven thermal growth causing the vessel to shift. The movement of the gasifier caused misalignment of the quench reactor and slag grinders, resulting in a failed fluid coupling on the grinder and a subsequent 40-hour shutdown. To prevent recurrence, hot spot water is added only as a mist so as to prevent

differential thermal growth. In re-bricking the original gasifier, this particular expansion joint was modified to avoid the occurrence of any hot spots due to the thinner refractory.

Immediately prior to the Operating Period, 13 hours of downtime was attributed to the second stage gasifier slurry feed nozzle. During start-up of the gasification plant, it was noted that the recycle syngas flow into the nozzle was restricted. The clearance for gas flow was found to be tighter than specification due to a manufacturing error. An additional root cause of the problem was identified as overheating of the nozzle due to disabling of the nitrogen purge valve to limit excessive nitrogen use. This problem has been addressed by throttling non-critical purges before introducing nitrogen to the recycled syngas system, thus alleviating concerns of excessive nitrogen use.

5.3.2.2 Operating Experience Overview

The total downtime associated with the first stage gasifier during the Operating Period was 169 hours. Operating experience for individual pieces of equipment follows.

Second Stage Gasifier

The second stage gasifier was responsible for 169 hours of downtime during the Operating Period. During start-up in December of 1998, an ash deposit that had formed dropped from the second stage gasifier and covered the gasifier taphole. The remedy has been removal of large ash deposits from the second stage gasifier during outages. Prior to December 1998, maintenance activities were mainly limited to routine refractory inspections.

Second Stage Gasifier Slurry Feed Nozzle

Maintenance of the second stage gasifier slurry feed nozzle was limited to routine inspections and periodic replacements during the Operating Period. The replacement is dictated by the wear caused by the coal slurry and atomizing gas flows. No downtime was attributed to this equipment during the Operating Period.

Figure 5.3.2B: Second Stage Gasifier Slurry Feed Nozzle

Other Equipment

There was no maintenance, downtime or operational problem associated with the second stage gasifier quench nozzles during the Operating Period.

5.3.2.3 Summary and Conclusions

The second stage gasifier has operated satisfactorily, both during and following the Operating Period. The unit continues to be 100% reliable, maintenance is limited to preventative inspections of the second stage gasifier and the slurry feed nozzle. The quench nozzles have exhibited outstanding reliability without need for maintenance.

5.3.3 Raw Syngas Conditioning System

The raw syngas conditioning system includes:

Post gasifier pipe spool	Post gasifier residence vessel
Raw syngas de-superheater	Boiler inlet transfer piping

5.3.3.1 System Modifications

In the first year of operation, 567 hours of downtime were associated with ash deposition in this section of the plant. The post gasifier pipe spool that connects the second stage gasifier to the post gasifier residence vessel caused the majority of this downtime. The post gasifier pipe spool was an abrupt transition between the two connecting vessels resulting in high velocity zones where molten ash could accumulate due to impingement upon gas flow direction change. Accumulation of these deposits in the spool increased the differential pressure across the spool, which would lead to hot gas (1800ºF) flowing behind the refractory and contacting the carbon steel shell. Hot spots on the shell required that water and steam be applied to the external vessel wall to cool the metal. The refractory in the spool was changed in November of 1996 to a type of refractory that did not form as tenacious a bond with the molten ash. Although the accumulation of deposits in the post gasifier pipe spool were slightly decreased with the refractory change, ash deposits continued to form in the pipe spool and on the upper sections of the post gasifier residence vessel. As the deposit mass increased, and as the gasification system went through thermal cycles, deposits would break free, shatter and be carried downstream where they caused plugging in the boiler inlet transfer pipe and boiler tubes. This deposition caused 381 hours of downtime in the second year of operation.

Flange leaks associated with the post gasifier pipe spool were also common during the first two years of operation due to differences in thermal growth between the second stage gasifier and the post gasifier residence vessel. Several attempts to alleviate the problem were met with limited success.

By early 1997, it was clear that a major piping change would be required to eliminate the foregoing problems with this system. In September of 1997, the post gasifier pipe spool was

permanently removed from service and replaced with a new post gasifier pipe spool, which incorporated a long sweeping refractory lined 180-degree bend. The new spool provides significantly lower velocities between the second stage gasifier and the post gasifier residence vessel than the previous pipe spool and is lined with refractory that does not form tenacious bonds with the ash. The new spool also accommodates more thermal growth between the second stage gasifier and the post gasifier residence vessel. Consequently, since this modification, not a single hour of downtime has been attributed to ash deposition in this system and syngas leaks from this area have been virtually eliminated.

5.3.3.2 Operating Experience Overview

There was no downtime associated with the pre-boiler gas conditioning system during the Operating Period. Operating experience for individual pieces of equipment follows.

Post Gasifier Pipe Spool

During three routine outages within the Operating Period, the manway on the post gasifier pipe spool was opened for inspection and to facilitate equipment ventilation. In each case, only small amounts of ash were removed from the vessel walls. A thermocouple inserted in the spool to monitor gas temperature was replaced during one of those outages. No other maintenance, downtime or operational problems have been associated with this piece of equipment.

Post Gasifier Residence Vessel

During three routine outages within the Operating Period, the manway on the post gasifier residence vessel was opened for inspection and to facilitate equipment ventilation. During these outages, minor amounts of ash were removed from the vessel and the refractory expansion joints were packed with insulating material. The thermocouple inserted in the post gasifier residence vessel to monitor gas temperature was replaced twice during the Operating Period. No other maintenance, downtime or operational problems have been associated with this piece of equipment.

Raw Syngas De-superheater

During each of the three routine outages within the Operating Period, the raw syngas de-superheater was removed from service, refurbished and returned to service. Plugging in the de-superheater occurs over time, which limits its ability to quench the gas prior to the heat recovery boiler. Although no downtime was associated with the de-superheater during the Operating Period, plugging in the de-superheater occasionally resulted in elevated gas temperatures at the boiler inlet, which contributed to a higher fouling rate for the boiler and a higher blinding rate for the primary char filters.

Boiler Inlet Transfer Piping

During three routine outages within the Operating Period, the manway and the top head of the boiler inlet transfer piping were opened for inspection and to facilitate equipment ventilation. Solids are routinely removed from the bottom manway of this pipe during outages, however the accumulation of solids is manageable. In March of 1998, a refractory breach in this piping system directly opposite the raw syngas de-superheater resulted in a hot spot on the steel piping wall. The hot spot was cooled with steam on line and the refractory was repaired in a subsequent outage. None of these events resulted in any downtime or operational problems associated with this piece of equipment.

5.3.3.3 Summary and Conclusions

Since the installation of the new post gasifier pipe spool in September of 1997, the pre-boiler gas conditioning system has not been responsible for a single hour of downtime. With the exception of the raw syngas de-superheater, there have been no operational problems. Since the Operating Period, the strainers on the water feeding the de-superheater have been replaced with a finer mesh material. Preliminary run-time since the replacement indicates less plugging of the de-superheater, which has enabled the boiler inlet temperature to be controlled at the desired set point. Routine maintenance associated with this system is now limited to routine inspections and cleaning of the equipment during scheduled outages.

5.3.4 Slag and Solids Handling System

The slag and solids handling system includes:

Slag pre-crusher	Slag fines settler bottoms pumps (Figure 5.3.4C)
Slag crusher	Slag recycle water tank
Slag dewatering tanks	Slag recycle water pumps
Slag water storage tank (Figure 5.3.4A)	Slag recycle water cooler
Slag water pumps	Slag quench feedwater pumps (Figure 5.3.4D)
Slag fines settler (Figure 5.3.4B)	

Figure 5.3.4A: Slag Dewatering Tank Building and Slag Water Storage Tank

Figure 5.3.4B: Slag Fines Settler
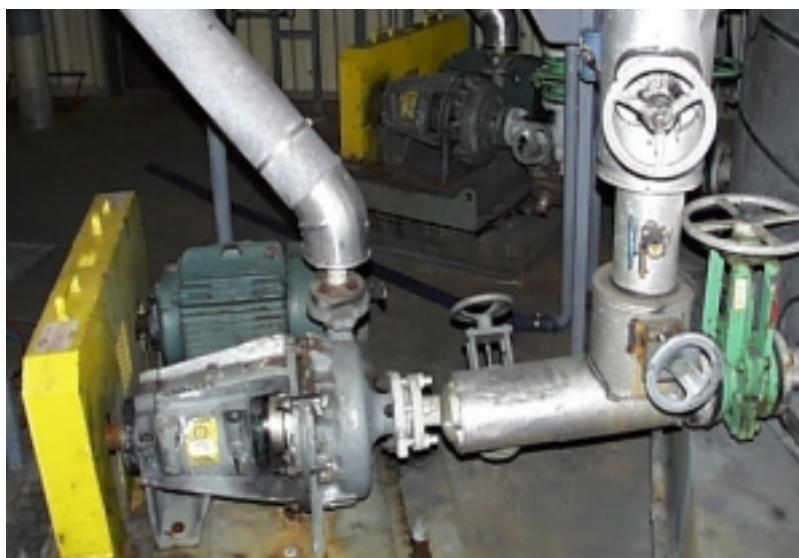
Figure 5.3.4C: Slag Fines Settler Bottoms Pumps

Figure 5.3.4D: Slag Quench Feedwater Pumps

5.3.4.1   System Modifications

Prior to the Operating Period, an anti-scale treatment system was added to slag and solids handling to help prevent scale build up in the piping associated with the slag water piping. This addition proved to be effective in preventing excessive scale build up in this area.

5.3.4.2   Operating Experience Overview

The total downtime associated with slag and solids handling during the Operating Period was 66 hours. Operating experience for individual pieces of equipment follows.

Slag Pre-Crusher

The slag pre-crusher was responsible for 66 hours of downtime during the Operating Period. An incorrect wiring installation for the slag pre-crusher electrical motor caused it to run in reverse. As a result of this error, that occurred some time during construction of the plant, oversized slag or refractory chunks would not pass through the pre-crusher and caused slag to fill the quench section of the gasifier. This eventually caused the plant to be shutdown to take corrective action.

Slag Fines Settler

The slag fines settler did not cause any downtime during the Operating Period, but maintenance was required on the slag fines settler solids rake due to a failed bearing in the rake gearbox and also on a failed stud on the rake steady bearing.

Slag Recycle Water Tank

The slag recycle water tank did not cause any downtime during the Operating Period, but had to be emptied and cleaned of solids during scheduled outages. The accumulation of solids was due to process upsets and failure of the decant filters in one of the slag dewatering tanks.

Slag Recycle Water Cooler

The slag recycle water cooler did not cause any downtime during the Operating Period, but had to be cleaned on several scheduled outages. Fouling of the tubes and accumulation of solids on the tubesheet resulted in higher-than-desirable water temperature in the system and necessitated the cleaning.

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: slag dewatering tank, slag fines settler bottom pumps, slag crusher, slag water storage tank, slag water pumps, slag recycle water pumps, slag quench feedwater pumps.

5.3.4.3 Summary and Conclusions

The slag and slag handling area of the plant continued to run as designed after the Operating Period with one notable exception. A metallurgical problem with the shaft sleeves on the slag crusher caused 104 hours of downtime after the Operating Period. The root cause of the problem was a poor metallurgical bond between the slag crusher shaft coating and the shaft base metal. Delamination of the coating resulted in a rough surface, which destroyed the crusher shaft packing necessitating a shutdown to facilitate repairs. The problem was addressed with a change in metallurgy and further problems have not been experienced.

Piping and, valve material loss from erosion by the slag continues to be a problem in the slag piping that is not lined with materials designed specifically to combat erosion. However, this is currently being addressed with use of more abrasion-resistant materials than those used in the original plant design.

5.3.5 Syngas Cooling/Steam Generation System

The syngas cooling/steam generation system includes:

Heat recovery boiler	High-pressure steam drum
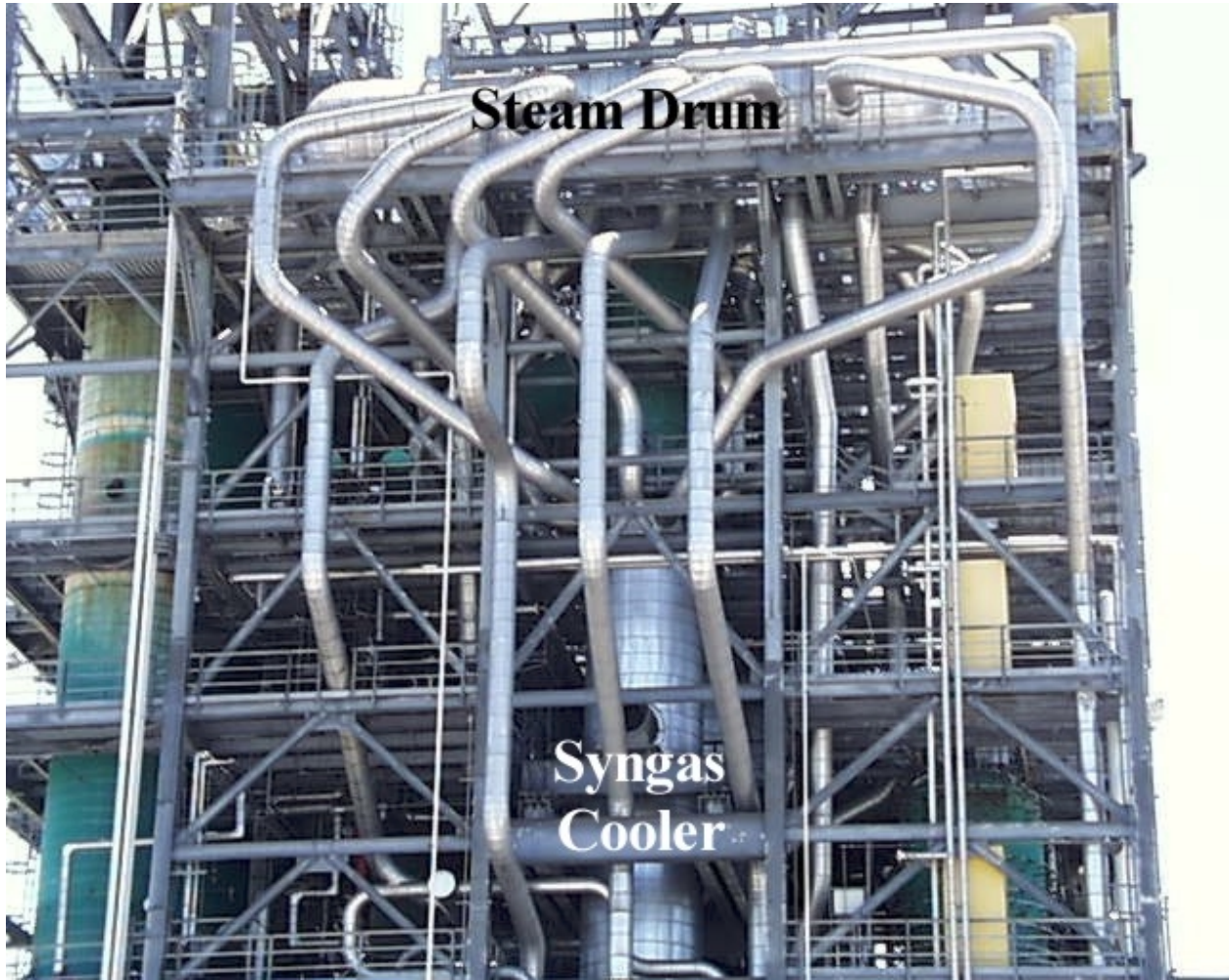
Figure 5.3.5A: Syngas Cooler and Steam Drum

The syngas cooling/steam generation system is used to remove high-level heat from the syngas stream as it exits the raw syngas conditioning system. The heat removed is used to produce high-pressure saturated steam. Major equipment in the syngas cooling/steam generation system includes a firetube heat recovery boiler and high-pressure steam drum. Refer to the process description section for additional details concerning the operation of this system.

5.3.5.1 System Modifications

Prior to the Operating Period, one of the leading contributors to gasification plant downtime associated with the syngas cooling/steam generation system was plugging in the boiler tubes. The plugging was a result of ash deposits spalling off of upstream piping and equipment and being carried downstream into the gas inlet of the boiler. (See Section 5.3.3.1) Particles too large to pass through the boiler tubes would then lodge in them and restrict flow through them. As the boiler tubes plug, the velocity in the least-restricted tubes could increase to a level at which erosion from the entrained char begins to occur. To avoid excessive tube wall thinning, the gasification plant is forced to terminate operation once a predetermined differential pressure across the gas side of the boiler is reached.

To help manage ash-particle-plugging of the boiler tubes, a screen was installed in the gas inlet to the boiler. The screen was modeled after a similar device used in the LGTI plant and designed to capture ash particles too large to pass through the boiler tubes. Due to the high operating temperature and corrosive nature of the syngas, extensive metallurgical testing was undertaken to find materials of construction that could provide acceptable life for the boiler screen. Several design enhancements and material changes were implemented to achieve a high level of reliability and useful life for the screen. Implementation of the screen and reducing the likelihood of forming large ash deposits upstream of the boiler has virtually eliminated boiler tube plugging due to ash particulate deposition spalling.

Another cause of gasification plant downtime associated with the syngas cooler/steam generation system is boiler tube fouling. Trace metals contained in the gasifier feedstock are vaporized during the gasification process. A portion of these volatile trace metals is present in the syngas exiting the gasifier system. As the gas stream flows through the boiler tubes, some trace metal compounds condense onto the inner tube wall surfaces reducing boiler heat transfer. The loss in heat transfer can eventually cause the gas temperatures exiting the boiler to exceed the design temperature of downstream equipment.

Prior to the Operating Period, it was determined that the boiler inner tube wall deposition was so tenacious and hard that conventional cleaning methods such as high-pressure hydro-blasting, mechanical scraping and impact drilling could not effectively remove the deposits. As a result,

special cleaning methods were developed to remove these deposits in a minimal amount of time. Additional methods that were evaluated to remove these deposits are discussed in more detail in Section 5.3.5.2

Plant thermal cycles exacerbated the spalling of ash deposition upstream of the boiler and accentuated installation flaws with the piping system on the outlet of the heat recovery boiler. A flanged piping spool piece on the boiler outlet was particularly problematic, developing gas leaks on several occasions. Enhancements to the spool flange torquing procedure were incorporated to improve the reliability of this system. This was only moderately successful. As a permanent solution, both flanged joints in this spool were replaced with welded connections to eliminate this potential leak point.

The high-pressure steam drum pressure transmitter sensing lines were upgraded as a result of several plant trips caused by erratic measurements in this equipment. As most of these occurrences took place during extremely cold weather conditions, improved heat trace and insulation systems were added to maximize the reliability of these systems. Additionally, control logic for the steam drum was upgraded to make the level control less sensitive to the idiosyncrasies of flow measurement.

Material upgrades were required for several nozzles located on the boiler gas inlet channel to control excessive corrosion at dissimilar metal welds. The alloy enhancements eliminated the aggressive corrosion present at these locations. Upgrades in alloy materials were also implemented at the boiler gas inlet channel manway insulation plugs to extend the service life of these components.

In 1997, syngas leakage through a valve located in a line connected to the boiler gas outlet piping eroded a small hole in the valve body forcing the plant off line for a short period of time. The valve was rebuilt with enhancements to the materials of construction. Plant operating procedures for this system were also changed to provide additional means of isolation downstream of this valve.

Excessive gas side boiler tube fouling resulted from a brief operating test period on a petroleum coke feedstock in late 1997. Inadequate temperature control for the gas entering the boiler was determined to be the root cause for accelerated tube fouling. Changes to plant operating procedures and control set points were incorporated to minimize tube fouling while on this feedstock.

5.3.5.2 Operating Experience Overview

The total downtime associated with the syngas cooling/steam generation system during the Operating Period was 158 hours. Operating experience for individual pieces of equipment follows.

Heat Recovery Boiler

The heat recovery boiler contributed to plant downtime during the Operating Period on two occasions. The first occasion was an operational trip of the recycle syngas compressor that caused an interruption in cooling at the boiler gas inlet. This interruption resulted in a brief acceleration in the tube-fouling rate for the unit, which required 6 additional hours of downtime to remove the deposition created from this temperature excursion. The second occasion extended a planned outage by 152 hours to remove process-side tube fouling.

Just prior to the Operating Period, fouling deposits previously removed from the boiler tubes were used to evaluate a number of tube cleaning methods. The most effective method was selected for a small-scale field test on one boiler tube during an outage in February of 1999. Data taken from this study was used to design a full-scale chemical cleaning system that was used to remove a significant amount of process-side tube deposition during July 1998 of the Operating Period. The chemical cleaning was effective in restoring the boiler heat transfer to new conditions.

From an operational point of view the chemical cleaning method was deemed quite successful. However, when evaluating the economics, timeliness and disposal issues associated with this process, it was evident that there needed to be another method. As a result, another method of mechanically cleaning the tubes was developed and tested during the Operating Period. This

particular method proved to be much more effective and less costly than the chemical cleaning method. Process and equipment optimization continues to be an ongoing high priority for this method of cleaning as cleaning of the boiler can be on the shutdown maintenance plan critical path.

Steam Drum

Other than routine inspections, there was no maintenance, downtime, or operational problem associated with the steam drum during the Operating Period.

5.3.5.3 Summary and Conclusions

During the Operating Period, the significant contributor to plant downtime and maintenance expenditures associated with the syngas cooling/steam generation system was deposition in the gas side of the boiler tubes. An acceptable mechanical cleaning method was developed to effectively remove these deposits. Optimization of this process continues to be made to further reduce the time required when performing this procedure. Studies also continue to determine if an additive in the gas stream prior to the boiler can be used to mitigate deposition in the unit. The major challenge for reliable operation of the syngas cooling/steam generation system is to effectively manage deposition on the gas side of the boiler tubes.

5.3.6 Particulate Removal System

The particulate removal system includes:

Primary char filters	Primary char recycle ejectors
Secondary char filters	Secondary char ejectors
Pulse gas accumulator drum	

The particulate removal system is used to remove char and ash from the syngas stream and recycle it back to the first stage gasifier. The system is located in the syngas path just downstream of the syngas cooler/steam generation process. The particulate removal system consists of two parallel-configured primary filter vessels with char recycle ejectors, secondary or backup char filters with char ejectors, and a pulse gas accumulator drum. Refer to the process description section for additional details concerning the operation of this system.

5.3.6.1 System Modifications

Prior to the Operating Period, problems in the particulate removal system contributed significantly to the total plant downtime. This section will describe modifications made to this system to improve its reliability and reduce operating and maintenance expenditures.

During the 1995-1996 start-up and commercial Operating Period, a number of hot gas filter system reliability problems were quickly identified and corrected. During the initial plant start-up char breakthrough was experienced in the primary candle filter system, which initially incorporated ceramic candles mounted to fixing hardware in a tie-rod type system. The source of this leakage was found to have occurred from inadequate sealing gasket at both the tie-rod element hardware and filter tubesheet flange connections. These problems were easily resolved by changing gasket materials in the tie-rod elements and by implementing a number of small changes in the tubesheet flange design. As run-time accumulated, more difficult challenges were realized as ceramic candle failure and bridging of solids between the filters began to occur.

During plant start-up, it was soon realized that backpulse pressures measured within the candle filters were insufficient and therefore could not provide the proper amount of backflow necessary

to effectively remove the char cake from the filters. As a result, the filter elements were not cleaned sufficiently and solids accumulated and bridged between the elements. This eventually led to a loss of filter area, high system differential pressures and failures of multiple filter elements.

Investigation into this problem showed that a number of key components within this system were undersized in the original design. To provide optimal system operation, the backpulse system components were re-designed and replaced.

Bridging of solids between the elements also occurred as a direct result of mixing filters with various hours of operation on them. This practice also led to failures in the elements. It was soon realized that elements of different operating "age" have different permeabilities, i.e., resistances to flow through them. Mixing these elements led to flow imbalances within the filter system. This flow imbalance could be quite severe immediately following the backpulse event. As a result, both bridging and failures occurred in the newer elements. Filter conditioning periods during plant start-up and changes in vessel element loading practices were used to mitigate this problem.

Frequent element failures during the first year of operation caused at least five (5) major plant outages. Element breakage resulted from a number of sources, such as those mentioned in the preceding paragraphs as well as improper handling during maintenance, and design problems in the hardware used to affix the filters to the tubesheets. After many attempts to improve the filter hardware system, it became evident that many of these design problems were quite complex and, as a result, the system was retrofitted with metal candles late in 1996. Metal candles immediately improved the reliability of the particulate removal system. Several years of operating experience were necessary to develop an optimal metal alloy and filter construction to provide adequate filter life.

Certain types of candle filters display a much more rapid blinding rate (loss of permeability) than others when operated in the particulate removal process. Two factors have been found to contribute significantly to this phenomenon. They are vapor phase condensation of trace metal

compounds on the element surface, and small particle penetration into the element surface pore structure.

The mechanism for how these two interact to blind the candle continues to be an area of intense study. This blinding results in a gradual increase in the baseline differential pressure across the primary filter vessels. At some point, the differential pressure becomes so high that effective backpulsing cannot be achieved. The "seasoning" or "conditioning" of a new filter in operation involves the accumulation of char on the candle surface that is not dislodged during the backpulse event. This permanent, or residual, layer causes the upward shift in baseline differential pressure across the filter during the conditioning period. During normal candle filter operation, once the conditioning period is over, the increase in baseline differential pressure across the primary filter should be minimal. However, when operating certain types of filters in the Wabash River process, the permeability is reduced rather rapidly in both the residual char layer and the filter substrate as a result of condensation of metal compounds. This rapid rise in primary filter differential pressure has forced the plant into a number of unplanned outages. Numerous blinding studies have been conducted for virtually all commercially available filters to determine which ones will provide an optimal life. As a result, filters that provide adequate operating time between scheduled plant outages have been found. Off-line cleaning methods have also been developed to allow many of these filters to be re-used.

Another form of filter blinding that happened on at least one occasion prior to the Operating Period was caused by condensation of organics (tars) on the primary candle filter surface. This resulted in a rapid loss in filter permeability. Tars were removed from the filters utilizing off-line cleaning techniques. Improved procedures were used to reduce tar production and eliminate this type of blinding.

Another source of candle breakage and bridging was caused by inadequacies within the primary vessel internal gas/solids distribution system. High filter impingement velocities from the particulate-laden syngas exiting the internal gas distributor caused erosion of filter surface and resultant element blinding at certain locations. Filter permeability was totally lost in these high wear areas on the element surfaces. High erosion rates on some distributor parts also yielded

low component life. Computational flow dynamics (CFD) modeling was used to optimize the design of the existing distributor system. A number of improvements were implemented to this system to significantly improve component life and reduce filter impingement velocities.

Operational problems in several backpulse valves caused some plant downtime prior to the Operating Period. Sticking instrument air slider/solenoid mechanisms were the first major problem areas to develop in these valves. The valves were retrofitted with an alternate manufacturer's slider/solenoid assembly to eliminate this problem. Pre-start-up audits to verify normal backpulse valve operation also helped to minimize outage time caused by this equipment. Additional backpulse valve problems developed as a result of a number of control disk seat failures that happened prior to the Operating Period. Refer to the next section for more detail concerning the seat failures.

Prior to the Operating Period, high erosion rates in the char recycle ejectors and associated piping systems caused a number of plant outages. High solids loading in the char recycle gas was found to be the root cause for this problem. Since design recycle gas velocities were already at the minimum required to keep the solids moving, material changes were necessary in the piping systems at key high wear locations. Harder, more abrasion-resistant materials were incorporated in the piping at these locations to mitigate this problem.

Most char ejector failures were caused by interference problems between internal parts caused by thermal expansion differences and design problems in the interlocking mechanisms of these parts. The life-limiting internal components were redesigned and upgrades in materials of construction were used to improve char recycle ejector life.

The secondary filtration system is designed to serve as a backup to the primary filter vessels. Its purpose is to contain any solids that may breakthrough the primary filter system due to filtration inefficiencies or failures. This system was installed to protect downstream equipment such as COS catalyst beds and heat exchangers from char particulate contamination. Early on in plant operation, several key design problems became evident. The holding mechanism that affixed the filters within their containing vessels was found to be inadequate. Filters were also being

damaged during the backpulse event. Stronger holding mechanisms and better filter element constructions were implemented to improve the reliability of this system.

Another problem that became evident early on in this system was low filter internal pressures being generated during the backpulse event. As in the primary filter system, this resulted in low reverse flows and resultant poor cleaning efficiencies. Key backpulse system components were resized to overcome this problem.

5.3.6.2 Operating Experience Overview

The total downtime associated with the particulate removal system during the Operating Period was 152 hours. Operating experience for individual pieces of equipment follows.

Primary Char Filter

During the Operating Period, there were two separate events in which primary filters failed and forced the plant to be taken off line. The first event that forced the plant into an outage was caused by a failure, and subsequent char breakthrough, of several metal candle filters. The initial char breakthrough was first noticed approximately two weeks prior to a scheduled plant outage and, as a result, only accounted for 25 hours of plant downtime. The plant remained on line for a period of time until the solids breaking through finally overwhelmed downstream equipment. In an effort to gather additional data on metal filter corrosion rates, a test panel of two previously operated filters was installed in one of the primary vessels prior to the campaign. All other filters in the primary vessels were new. Metallurgical support personnel had predicted the filters could last well beyond the Operating Period. However, the filters became weakened by corrosion and eventually failed during a backpulse event. During this campaign it became evident that changes in process conditions had accelerated corrosion rates and rendered the previous data used for life predictions inaccurate. Knowledge gained from this experience was used to implement a number of process changes to better control the rate of corrosion in these metal filters. The use of fail-safe devices was also implemented for all test filters in subsequent operating campaigns. A fail-safe is a device that protects against particulates entering downstream equipment in the event of a primary filter element failure.

During the second forced plant outage caused by solids breaking through a primary filter a fail-safe device was used. This particular event involved a test panel of ceramic filters. A new and novel way to affix these filters into the tubesheet had been developed and was being tested during this campaign. Several days into the campaign, a small amount of char was found to be breaking through the primary filter. It remained fairly steady at a low rate and was manageable. After approximately 35 days of operation the fail-safe located over the leaking filter failed and the solids breaking through quickly overwhelmed downstream equipment. The plant downtime associated with this failure was approximately 61 hours. The root cause for the filter failure was determined to be from dimensional interferences in the hardware used to affix the ceramic elements into the tubesheet. The manufacturer of these parts supplied a number of them out of dimensional specifications. As a result, one test filter was damaged during assembly and the location of the damage was such that it could not be detected during assembly and installation. Evaluation and restructuring of supplier quality control procedures were implemented to mitigate these problems. A manufacturing method to increase fail-safe media strength was also developed to enhance the reliability of these devices.

Repairs to the primary filter internal gas distributor system also accounted for 18 hours of lost production during the Operating Period. Improvements made to this system prior to the Operating Period have increased component life significantly. However, high erosion rates remain in certain key locations within the distributor system. New areas of wear that had previously not been detected were found during a scheduled plant outage. Gas impingement from holes in the distributor system was directed toward the vessel walls and resulted in unanticipated repairs that extended the scheduled outage time. Panels of wear-resistant materials were installed in these locations to extend component life and mitigate this problem.

During the Operating Period, approximately 24 hours of downtime was associated with a flange leak and subsequent fire that developed in the gas inlet piping to the primary char filter vessel. The leak was found to have been caused by movement in the piping system encountered when new inlet ball valves were installed during a previous outage. Installation of the new valves did not include inspection or pressure testing of flange connections that were not disturbed. Inspection procedures were changed to prevent this sort of problem from occurring again.

The original installation used butterfly valves to provide isolation in the gas inlet piping to the primary filters. These valves suffered high rates of erosion resulting from the gate being located in the center of the particulate laden gas stream when in the open position. Additionally, high erosion rates were found in the piping just downstream of these valves. Gas turbulence from flow around the butterfly gates was determined to be the root cause for this high rate of material loss. Excessive repairs were required for both the butterfly valves and associated downstream piping during scheduled plant outages. The isolation valves were replaced with new high alloy full-port ball valves thereby eliminating the associated maintenance requirements. During the initial campaign with these new valves, problems were encountered with the hydraulic actuators that accounted for 6 hours of additional plant downtime. The supplier implemented a number of changes to the hydraulic system to improve its reliability.

There were 6 hours of plant downtime that were attributable to problems in the backpulse system for the primary particulate removal filter. A backpulse valve failed early in an operating campaign resulting in the above-mentioned downtime. The valve was found to be leaking pulse gas through at an excessive rate. The valve was found to have a damaged control disk seat. Earlier investigation into similar problems had determined that valve body distortion caused by thermally induced pipe stress was the most likely cause of these failures. A pulse gas pre-heater used during system start-up was left in operation and the elevated temperature likely contributed to higher induced stresses and the resultant seat failure. New and improved valve seats were installed in the valve to make them more reliable. The new seats are designed to better handle this valve body distortion and, as a result, there have been no failures since they were installed.

Another problem associated with these backpulse valves was in the retaining nut that connects the control disk to the stem. When replacing the control disk and seat with new and improved parts, this nut was removed, reinstalled and tightened to within the manufacturer's specifications. Several of the nuts came loose during operation. This problem did not force the plant to be taken off line or cause a capacity reduction, but it did result in removal of some filtration area for the campaign. New higher torque specifications and replacement nuts and lock washers were used to rectify this problem.

Primary Char Recycle Ejectors

On three separate occasions during the Operating Period, problems in the char recycle ejectors contributed to plant downtime. The first event, which accounted for 6 hours of plant downtime, resulted from a manufacturing error in one of the ejector internal components. A dimensional discrepancy and resultant thermally induced interferences led to a premature failure in this ejector. More stringent quality control was implemented to address this problem. The last two events were a result of degradation in ejector performance caused by high wear rates in key internal motive gas section parts. These two events combined accounted for 7 hours of plant downtime. Upgrades in materials of construction were used to extend the life of these parts.

Secondary Char Filters

There was no downtime attributable to the secondary filter system during the Operating Period. However, continued failures in the filter holding mechanisms led to investigations into better ways to backup the primary filter. Fail-safe devices are continuously being evaluated both for reliability and efficiency in the particulate removal slipstream and primary systems. Through accumulated operating experience and design enhancements, a reliable and efficient fail-safe was developed. During the Operating Period, fail-safes began to be used regularly in all filters of questionable reliability and eventually replaced the secondary filter system.

Other Equipment

There was no maintenance, downtime or operational problem associated with the pulse gas accumulator drum during the Operating Period.

5.3.6.3 Summary and Conclusions

Figure 5.3.6A is a chart that shows the percentage of total downtime hours attributable to the particulate removal system as well as plant outage hours caused by other gasification plant processes and the combined cycle power generation facility. In 1996, difficulties in the particulate removal system made up over 39% of the total plant downtime. As particulate removal system improvements were implemented, its operation improved in 1997 with the total attributable downtime being reduced to around 17%. Additional improvements in 1997 and

early 1998 have further reduced the particulate removal system downtime hours to less than 6% of the total.

**Wabash River Plant
Downtime Summary**

Downtime Hours

- "Other" Gasification Plant
- Combined Cycle Power Generation
- Particulate Removal

1996 1997 1998
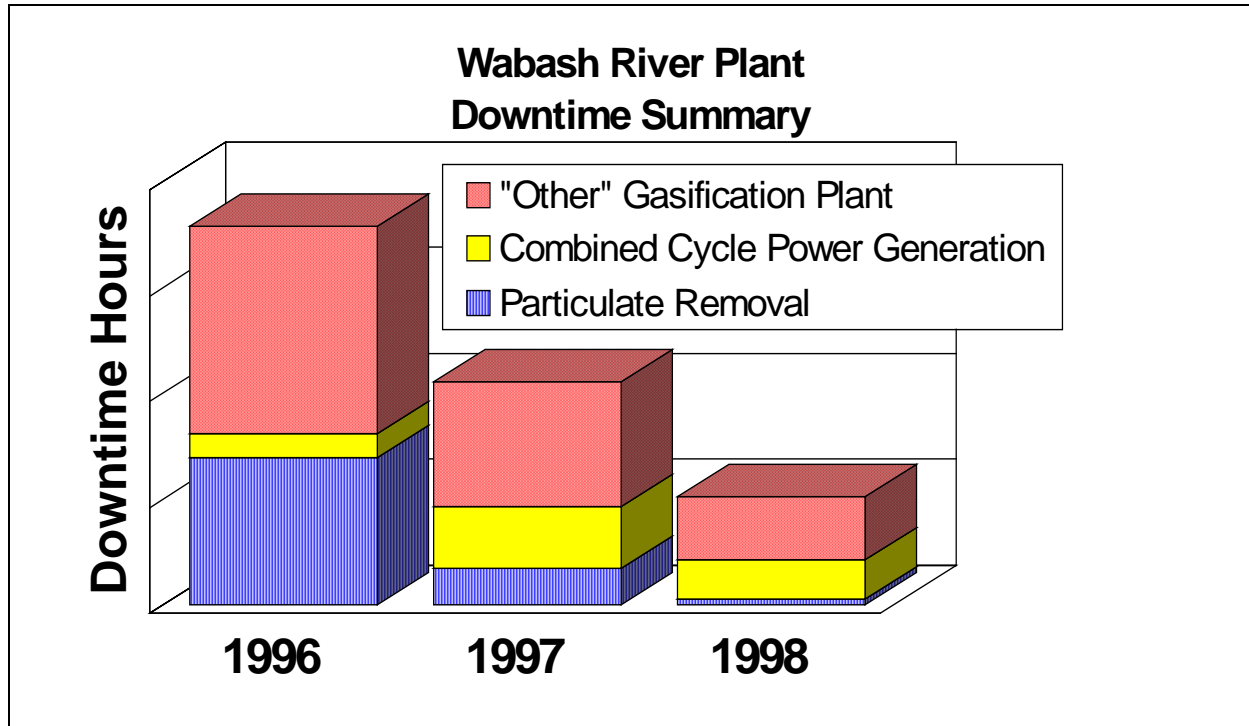
Figure 5.3.6A: Wabash River Plant Downtime Summary

Since plant start-up, there have been many design and materials of construction changes that have been implemented to significantly increase the reliability of the Wabash River particulate removal system. However, opportunities remain for improving the system in order to meet specific performance goals that will further reduce operating and maintenance expenditures. A number of key areas continue to be intensely studied and improved. Accelerated filter blinding continues to be a concern and is an area of keen interest in filter development work. Filter corrosion is also a concern when operating with metal candle filters, especially if higher sulfur-containing feedstocks are used. There is still a need to further improve the internal gas distribution system to extend component life and minimize filter gas impingement velocities.

5.3.7 Chloride Scrubbing System

The chloride scrubbing system includes:

Syngas cross-exchanger (Figure 5.3.7A)	Syngas cooler (Figure 5.3.7A)
Chloride scrubber (Figure 5.3.7A)	Knockout drum (Figure 5.3.7A)
Scrubber knockout level control drum	Sour water cross-exchangers

The chloride scrubbing system was installed in September of 1996 to eliminate two problems. First, chloride in the gas stream was causing chloride stress-corrosion-cracking of the low temperature heat exchangers and piping systems. Second, the chlorides and trace metals, primarily arsenic, were blinding the carbonyl sulfide (COS) hydrolysis catalyst. Installation of the system successfully eliminated both of these problems.

Figure 5.3.7A: Major Equipment Associated with the Chloride Scrubbing System

5.3.7.1 System Modifications

The chloride scrubbing system came on line in October of 1996. The system performed as expected with very few modifications. In the first seventeen months of operation, the packing in the chloride scrubber required removal for cleaning three times due to the accumulation of tars. On one occasion, the differential pressure across the vessel forced the plant off line, resulting in 249 hours of downtime to clean the packing. In early 1998, process conditions upstream of the scrubber were altered to reduce the formation of tars. Since that time, the packing has not required removal for cleaning. Several syngas leaks from the inlet piping flange of the chloride scrubber resulted in 38 hours of downtime and one additional syngas leak from an improper gasket on the outlet flange of the syngas cross-exchanger resulted in 8 hours of downtime in the first year of its operation. During a planned outage in December of 1997 the chloride scrubber inlet piping flange was removed, re-worked to ensure a proper gasket-sealing surface, and placed back in the piping system. As a result of the re-work, this piping flange has operated trouble-free since that time.

5.3.7.2 Operating Experience Overview

The total downtime associated with this chloride scrubbing system during the Operating Period was 10 hours. Operating experience for individual pieces of equipment follows.

Syngas Cross-Exchanger

With one exception, the syngas cross-exchanger operated as expected during the Operating Period. In December of 1998, the cross-exchanger experienced tube fouling that prevented it from maintaining the desired syngas superheat temperature entering the COS hydrolysis system. Allowing a small amount of particulate-laden syngas to bypass the dry char system into the cross-exchanger facilitated on-line cleaning of the cross-exchanger but the downstream sour water carbon filter was quickly plugged. Fortunately, the spare sour water carbon filter was available so that the plugged filter could be back-washed without interruption to syngas production. Consequently, no downtime was associated with this equipment in the Operating Period.

Syngas Cooler

The original installation included a water spray at the inlet of the syngas cooler, which could be used for cleaning the exchanger tubes on line. This system proved to be unsuccessful in cleaning on more than one occasion. Use of the water spray resulted in syngas leaks at the exchanger flanges due to thermal growth difference caused by the cold spray on the hot exchanger. This system was completely removed after a failure of the spray nozzle caused the plant to come off line in January of 1999, resulting in 10 hours of downtime. With the exception of this failure, the system operated without incident for the Operating Period.

Chloride Scrubber

Other than routine inspections, maintenance associated with the chloride scrubber during the operation period was limited to two repairs to the outlet piping. A defective piping ell, combined with downtime corrosion, led to polythionic stress-corrosion-cracking in the ell. The sour water leaking from the ell was stopped with the placement of a temporary pipe repair clamp while the plant was on line and then replaced during a subsequent outage. No downtime was associated with this equipment during the Operating Period.

Scrubber Knockout Level Control Drum

Although no downtime was associated with this equipment during the Operating Period, three sour water leaks from the outlet piping system required repair. These piping failures occurred in the welds as a result of high mechanical stresses on this line. The outlet isolation valve on the drum was relocated to allow on-line maintenance of the piping system and an additional piping support was added to reduce the line stress. In addition, debris is routinely cleaned from this drum during planned outages.

Sour Water Cross-Exchangers

During the three planned outages in the Operating Period these two exchangers were flushed and cleaned by Operations personnel to remove sludge that accumulates during operation. No downtime or operational problems were associated with these exchangers during the Operating Period.

<u>Other Equipment</u>

There was no maintenance, downtime or operational problem associated with the knockout drum during the Operating Period.

5.3.7.3 Summary and Conclusions

The chloride scrubbing system has proved effective in removing chlorides and trace metals that damage downstream equipment and piping and that deactivates the COS hydrolysis catalyst. The system has operated as expected since it was installed in September of 1996. Early problems with excessive tar formation in the chloride scrubber were resolved early in 1998. Occasional removal of the packing from the chloride scrubber for cleaning will still be necessary but should not be required more frequently than biennially. After the Operating Period, a shutdown procedure was developed to flush the liquid outlet piping from the chloride scrubber so that downtime corrosion could be minimized. Flanges were also installed in this piping to facilitate inspections of the piping to insure downtime corrosion is under control. Routine maintenance associated with this system is now limited to routine inspections of the equipment and piping and cleaning of the sour water cross-exchangers and the sour water level control drum.

5.3.8 Carbonyl Sulfide (COS) Hydrolysis System

The carbonyl sulfide (COS) hydrolysis system includes two COS reactors. Refer to the process description for more process details of this system.

Figure 5.3.8A: Carbonyl Sulfide Reactors

5.3.8.1 System Modifications

The chlorides and trace metals contained within the coal caused premature equipment failures downstream and deactivation of the original COS hydrolysis catalyst. The catalyst used at start-up lasted approximately 700 hours due to chlorides and trace metals occupying active sites on the catalyst. The sampling and replacement of the catalyst occurred frequently prior to the successful implementation of the chloride scrubbing system (Refer to Section 5.3.7). The

chloride scrubbing system installed in September of 1996 effectively addressed these concerns as well as metallurgy concerns with chlorides further downstream in the process.

Additional efforts were undertaken in early 1996 to identify cost-effective catalysts for the carbonyl sulfide reactors. To test and evaluate the COS hydrolysis catalyst of various manufacturers, a small slipstream unit was implemented at the Wabash River site. Several manufacturers' COS hydrolysis catalysts were tested in the slipstream unit to provide additional data into the selection of a replacement COS hydrolysis catalyst.

Other process improvements were implemented on the COS reactors prior to the Operating Period that reduced downtime and maintenance. These process modifications targeted improvements to the COS reactors' heat-up and cooldown system, extension of COS catalyst life, and equipment reliability.

The COS reactors are in the critical path during maintenance outages due to the duration of time required to heat and cool these vessels. The COS hydrolysis system logged 277 hours of downtime associated with the heat-up and cooldown of equipment prior to the Operating Period. The reactors were modified in October 1995 to accommodate activated carbon in addition to the COS hydrolysis catalyst. The carbon acts as a guard bed to prevent trace compounds in the syngas from fouling the catalyst and downstream equipment. Improvements were implemented on the COS reactors heat-up and cooldown system in April of 1997 to increase plant availability, decrease maintenance expenses, and reduce nitrogen consumption. Nitrogen is used for heat-up and cooldown of the system. The improvements included the addition of a heat exchanger and permanent piping system. The addition of this equipment reduces the heat-up and cooldown duration, which conserves nitrogen and eliminates the need to install temporary equipment each maintenance outage, resulting in a safer operation. An additional piping modification was made to this system in December of 1999 that enables gasifier maintenance to commence while the COS reactors continue in the cooldown process. This improvement further reduces the plant maintenance outage duration.

The manual block valves on the COS reactors' inlet and outlet piping were modified to improve the operability of the valves. Bearing protectors and flush connections to inject grease were added to prevent the build-up of solids and to assist the smooth operation of the valves.

5.3.8.2 Operating Experience Overview

There was no downtime associated with the COS hydrolysis system during the Operating Period. Operating experience for individual pieces of equipment follows.

Carbonyl Sulfide Reactors

Other than routine inspections, the maintenance associated with the COS hydrolysis system during the Operating Period was limited to replacement of the activated carbon and a piping modification that enables the carbon to be removed in an inert atmosphere. When the carbon was replaced during the June 1998 outage, additional carbon was added to each of the COS reactors. This additional volume of carbon, plus improvements made in plant operations, has tripled the useful carbon bed life and decreased the frequency of change outs. No downtime was associated with this equipment during the Operating Period.

5.3.8.3 Summary and Conclusions

Since start-up of the chloride scrubber system and replacement of the COS hydrolysis catalyst, this system has performed per design. The process improvement plan to identify alternative hydrolysis catalysts, less prone to poisoning from both chlorides and trace metals within the syngas was successful. An alternative catalyst was identified and installed in October of 1996 and has demonstrated high performance in the hydrolysis process with minimal degradation in performance over extended run-time. Routine maintenance associated with this system is now limited to the replacement of COS catalyst and activated carbon. The operating life projection for the COS catalyst and activated carbon is 5 years and 1.5 years, respectively.

5.3.9 Low Temperature Heat Recovery System

The low temperature heat recovery system includes the following equipment:

Amine boiler (Figure 5.3.9A)	Sour gas condensate condenser (Figure 5.3.9A)
Sour water level control drum	Sour water cooling tower water condenser (Figure 5.3.9A)
Sour water condenser (Figure 5.3.9A)	Sour gas knockout pot (Figure 5.3.9B)
Sour water receiver (Figure 5.3.9B)	Syngas Recycle compressor knockout drum (Figure 5.3.9B)
Second stage gasifier quench water pumps	Syngas recycle compressor (Figure 5.3.9B)
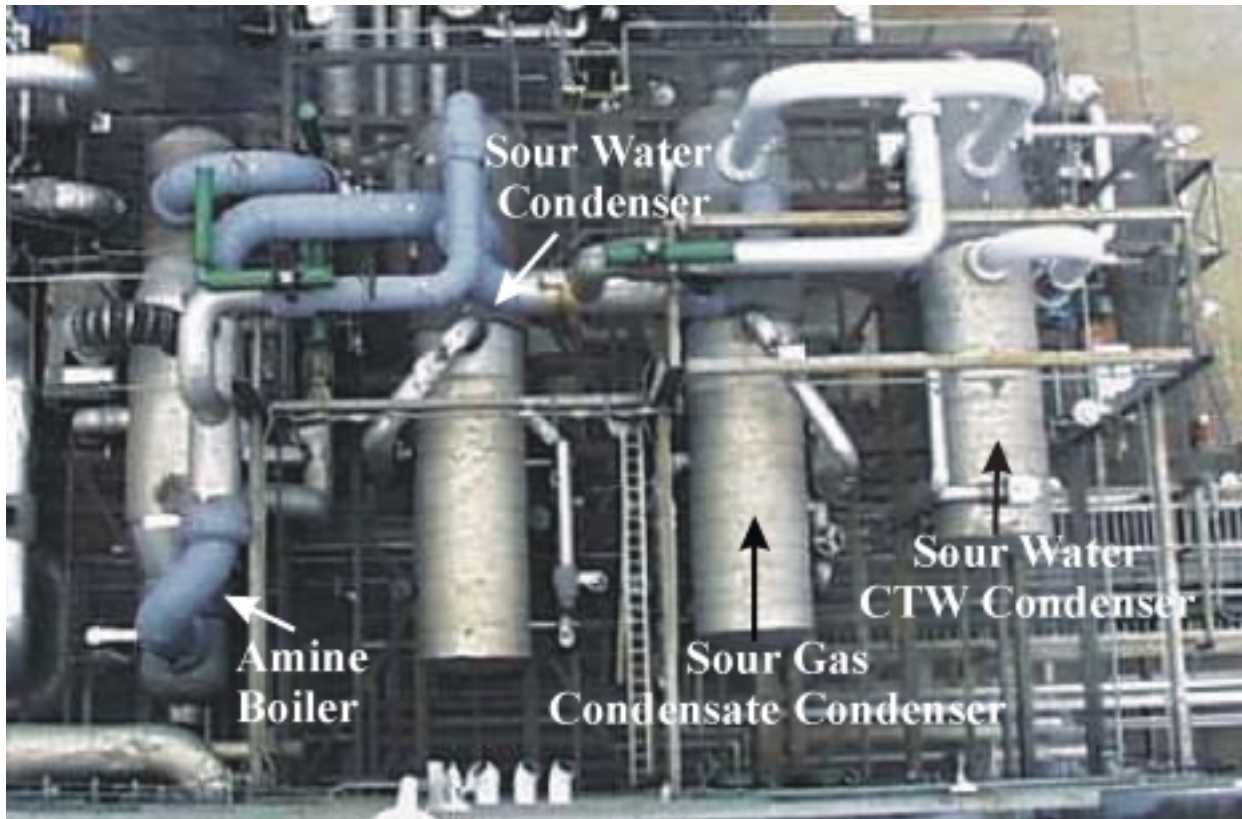
Figure 5.3.9A: Heat Exchanger Deck for Low Temperature Heat Recovery

5.3.9.1 System Modifications

During the first year of operation, the low temperature heat recovery system accounted for 1,682 hours of plant downtime. Fouling of exchangers in this system was frequent and the location and

size of these exchangers required large cranes every time maintenance was needed. Obtaining tight seals on the exchanger heads after maintenance also proved difficult, resulting in syngas leaks on more than one occasion. Chlorides in the syngas aggressively attacked the exchanger tubes and tube leaks often brought the plant off -line. In June of 1996, a new amine boiler, constructed from tube material impervious to chloride stress-corrosion-cracking, was installed. A month later the tube bundle in the sour water condenser was replaced and, in September of 1996, the chloride scrubbing system was installed. The chloride scrubbing system virtually eliminated all of the chloride problems in the low temperature heat recovery system. The chloride scrubbing system also removed the material that caused the exchangers to foul, thus eliminating the need for frequent and expensive exchanger cleaning.

During the first year of operation, the sour water cooling tower water condenser (the last of the exchangers in this system) developed several tube leaks not associated with chlorides. The inlet velocity to the exchanger was so high that the tubes near the entrance vibrated, resulting in mechanical fatigue failures of the tubes. In June of 1996, an impingement baffle was installed to disperse the incoming gas flow. There have been no tube failures following the installation of the baffle.

Of the total downtime listed, 642 hours was associated with the syngas recycle compressor. Much of this downtime resulted from the failure of one of the impellers of this 10,500-rpm machine. The root cause of the failure could not positively be identified, but was likely due to off-spec dimensional properties on the failed impeller coupled with fatigue loading in a corrosive environment. Installation of the chloride scrubbing system significantly reduced the corrosiveness of the syngas environment and thus has increased the reliability of this machine.

The machine was also subjected to severe fouling on several occasions, ultimately necessitating disassembly of the machine to replace the fouled rotor with a spare. Repeated failure of one of the seals in the machine also prompted the original equipment manufacturer to redesign the seal and install the modifications in the compressor. Operating procedures on the compressor were modified to mitigate fouling and a procedure was developed to clean the machine on line. Additionally, the recycle compressor knockout drum was fitted with a mist eliminator to reduce

liquid entrainment into the machine further reducing the fouling potential of the syngas. Prior to the Operating Period, the recycle compressor knockout drum occasionally experienced high liquid levels due to partial plugging in the outlet line. An atmospheric blowdown line was installed from the drum to the waste water sump to allow restrictions in the line to be safely cleared while the plant was on line.

The syngas recycle compressor also experienced high vibrations that caused the machine to trip on several occasions. Replacement of some components on the high-speed coupling alleviated the vibration, although the coupling manufacturer could not find any defects in the part returned for failure analysis

5.3.9.2 Operating Experience Overview

The total downtime associated with the low temperature heat recovery system during the Operating Period was 7 hours. Operating experience for individual pieces of equipment follows:

Amine Boiler

Although no downtime, maintenance or operational problems were associated with the amine boiler during the operational period, the piping connected to this exchanger was removed and inspected for residual cracks that were formed prior to the installation of the chloride scrubbing system. The cracks that were inspected did not show appreciable growth since they were last measured.

Sour Water Level Control Drum

During a vessel inspection in December of 1998, severe pitting below the normal liquid level of the sour water level control drum was discovered. The pitting was initiated by high chlorides in the gas prior to installation of the chloride scrubber. The lower section of the vessel was re-clad and no additional pitting has been noted since the repair. There was no downtime or operational problems associated with this piece of equipment during the Operating Period.

Sour Water Receiver

During the three planned outages in the Operating Period, the sour water receiver was flushed and cleaned by Operations personnel to remove sludge that accumulates during operation. No downtime or operational problems were associated with this equipment during the Operating Period.

Second Stage Quench Water Pumps

During the Operating Period, the mechanical seal failed once on each pump and the coupling on one of the pumps was also replaced. During a routine outage, the pump was disassembled for inspection and the wear rings were replaced. No downtime or other operational problems were associated with this equipment during the Operating Period.

Sour Gas Condensate Condenser

During the Operating Period the outlet temperature transmitter from this exchanger failed, which caused the syngas to completely bypass the exchanger. Since the bypass valve is much smaller than the main syngas valve, the downstream syngas recycle compressor was starved of gas and tripped on low-low first stage flow. The plant lost 4 hours of production due to the transmitter failure. The control code was re-written to prevent total closure of the main syngas valve, which will prevent recurrence of this problem. No other downtime, maintenance or operational problems were associated with this equipment during the Operating Period.

Syngas Recycle Compressor Knockout Drum

Early in the Operating Period the level transmitter on the recycle compressor knockout drum failed and, during the ensuing troubleshooting, the syngas recycle compressor was inadvertently tripped due to the human error. The resulting downtime was 3 hours. The recycle compressor knockout drum was opened one time during the Operating Period for inspection and to remove solids from the drum. No other downtime, maintenance or operational problems were associated with this equipment during the Operating Period.

Syngas Recycle Compressor

The syngas recycle compressor shutdown during methane operations on three different occasions. The compressor tripped each time on low-low first stage flow as a result of start-up conditions, which starved flow to the machine. Because the plant was not on coal operation, none of the trips resulted in downtime. Modification to the start-up procedure as well as operator awareness of the potential low flow condition has alleviated this problem. As mentioned previously, instrument problems associated with the sour gas condensate condenser and the recycle compressor knockout drum forced the syngas recycle compressor off line two times during coal operation. During the latter of these two incidents, the syngas recycle compressor failed to restart on three successive attempts due to high thrust conditions. The root cause of the high thrust could not be determined and the machine started normally on the next attempt after allowing the motor windings to cool for one hour. During a subsequent routine outage, the axial thrust proximity probe was reset in an effort to mitigate the forgoing problem. No other downtime, maintenance or operational problems were associated with the syngas recycle compressor during the Operating Period.

Figure 5.3.9B: Syngas Recycle Compressor and Knockout Drum

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: the sour water condenser, the sour water cooling tower water condenser, and the sour gas knockout pot.

5.3.9.3 Summary and Conclusions

During the first operating year, the low temperature heat recovery system experienced significant maintenance and was responsible for 1,682 hours of plant downtime. The root cause of most of this downtime was the high chloride content in the gas that caused rapid failures of the exchanger tubes and contributed to corrosion in the syngas recycle compressor. Since installation of the chloride scrubbing system in September of 1996 and other modifications made to the low temperature heat recovery system, the total downtime associated with this system has only been seven hours. Likewise, operational problems have been very few, and recent maintenance in this system has been limited to minor repairs to rotating equipment and cleaning and inspections of vessels during routine outages.

5.3.10 Syngas Moisturization System

The syngas moisturization system includes:

Syngas water saturator	Syngas water saturator recirculation pumps
Syngas preheater	

Refer to the process description for more process details of this system.

5.3.10.1 System Modifications

Since start-up the syngas moisturization system has operated without any mechanical changes or process modifications to the system.

5.3.10.2 Operating Experience Overview

There was no downtime associated with the syngas moisturization system during the Operating Period. Operating experience for individual pieces of equipment follows.

Syngas Water Saturator

With one exception, the syngas water saturator has preformed flawlessly during the Operating Period. In February of 1999, the taps for the differential pressure transmitter froze. Operations personnel successfully thawed the taps by applying steam and no further problems were noted.

Syngas Water Saturator Recirculation Pumps

Other than replacement of the suction valve on one of the pumps and normal preventative maintenance, these pumps have operated trouble-free for the Operating Period.

Syngas Preheater

During the Operating Period, no maintenance or operational problems were associated with this piece of equipment.

5.3.10.3 Summary and Conclusions

The syngas moisturization system has performed well since start-up. Occasionally the system has had difficulty reaching the set point temperature but no other problems, downtime or operational concerns have been associated with this system since start-up.

5.3.11 Acid Gas Removal System

The acid gas removal system includes:

Acid gas absorber (Figure 5.3.11A)	Acid gas stripper (Figure 5.3.11A)
Lean amine pumps	Acid gas stripper knockout drum
MDEA storage tank	Acid gas stripper reflux drum
MDEA cross-exchangers (Figure 5.3.11A)	Acid gas stripper quench pumps
MDEA pre-filter	Acid gas stripper reboiler
MDEA carbon bed (Figure 5.3.11A)	Acid gas stripper carbon beds
MDEA post-filter	Acid gas stripper recirculation cooler
MDEA CTW coolers	Acid gas preheater
Lean amine transfer pumps	Acid gas stripper overheads filters
MDEA reclaim unit	

Refer to the process description for more process details of this system.

Figure 5.3.11A: Acid Gas Removal System Major Equipment

5.3.11.1 System Modifications

Prior to the Operating Period, modifications were made to the equipment within the acid gas removal system to increase reliability, decrease maintenance costs, or improve the safety of the equipment. Modifications to individual pieces of equipment are given below.

MDEA Storage Tank

One significant process change prior to the Operating Period was the addition of an oxygen scavenger to the amine solution. Injected at the discharge of the MDEA storage tank, the oxygen scavenger serves two purposes. The oxygen scavenger consumes oxygen in the amine solution

while also acting as a corrosion inhibitor by forming a thin passivation layer on the piping wall. Oxygen has been proven to breakdown amine, and leads to heat stable salt formation.

Lean Amine Pumps

Automatic recirculation valves were installed immediately downstream of the lean amine pumps to replace existing orifice plates in the lean amine pump recirculation line. When the H_2S concentration in the syngas is low, lower amine flow rates are required. Under this condition, the recirculation line passes a large fraction of the flow from the pump. With the original piping system, the pressure drop across the orifice plate created cavitation, which damaged the downstream piping creating a potential safety and environmental concern. Installation of the automatic recirculation valves eliminated the cavitation.

Acid Gas Absorber

Stiffening members were added to four of the lower trays to increase the mechanical integrity. Under upset conditions, such as high differential pressure or high gas or liquid flow, the lower trays within the column suffered damage. Upon inspection, several of the trays were found to be resting on the tray below. Additional supporting members were installed to provide more resistance to the forces acting to collapse the trays.

To reduce heat stable salt accumulation, caustic was introduced into the amine solution. The caustic reacts with the heat stable salt liberating the MDEA for absorption of H_2S. Excess addition of caustic; however, causes the system to reach the saturation concentration of sodium bicarbonate leading to salt precipitation onto the trays of the acid gas absorber creating high differential pressure. As a result of this type of incident, 30 hours of downtime were attributed to this system during the Operating Period. Consequently, the introduction of caustic as a means of reducing heat stable salt accumulation was terminated.

MDEA Cross-Exchangers

There are manual block valves located upstream and downstream on the rich amine side of the MDEA cross-exchangers. If these valves were to be closed, there existed the potential to overpressure the rich amine side of the exchanger as a result of thermal expansion of the amine.

Therefore, a pressure safety valve was installed upstream of the block valve on the discharge side of the exchanger. The discharge from the pressure safety valve was directed to the MDEA storage tank to prevent any MDEA losses and personnel exposure.

Acid Gas Stripper

The acid gas stripper consists of a stripping section and a reflux section separated by a sump tray. In the stripping section, steam (generated by boiling water out of the amine solution) flows up through the column and strips the H_2S and CO_2 from the amine. The amine solution is introduced on the tray below the sump tray. Upon start-up of the system, a significant amount of amine was being entrained with the steam and acid gas stream. The entrained amine solution caused appreciable MDEA losses and prevented proper blow down of the acid gas stripper reflux drum, which is necessary for ammonia control. Therefore, a demister pad was installed directly below the sump tray to knockout the entrained amine. Additionally, modifications were made to the sump tray to improve the seal and prevent quench water from falling into the stripping section and diluting the amine.

Prior to the Operating Period, a potential overpressure scenario was identified in the acid gas removal system. To protect these systems against this overpressure scenario, a pressure rupture disk was installed downstream of the acid gas stripper. Gases released from this pressure rupture disk would then be burned at the flare.

MDEA CTW Cooler and Acid Gas Stripper Recirculation Cooler

For personnel safety and environmental considerations, the pressure safety valves on these two heat exchangers were removed. The discharge of these pressure safety valves was directed to the grade in an area where operators frequently walk during their rounds. The liquid discharged would contain dangerous levels of H_2S. Therefore, the pressure safety valves were removed and the manual block valves were locked open and tagged to prevent the possibility of over-pressuring the respective equipment.

MDEA Reclaim Unit

The MDEA reclaim unit uses an ion exchange resin that has a thermal degradation temperature of approximately 200°F. The resin manufacturer warns that continuous exposure to temperatures in excess of 150°F will cause appreciable damage. Cold condensate is used to rinse the resin after the MDEA, displacing solution, and regenerating solution are passed over the resin. The cold condensate temperature at times can reach the 150°F limit recommended by the resin manufacturer. Therefore, a rinse water cooler was installed on the MDEA reclaim unit to avoid thermally degrading the resin. The heat stable salt regenerant solution originally recommended by the MDEA reclaim unit manufacturer was not able to remove completely the salts that have high affinity for the ion exchange resin. An alternate regenerant solution was used with good performance.

Several modifications were made to the piping configuration within the MDEA reclaim unit as well. The balance of resin canisters being fed by MDEA, heat stable salt displacing solution, regenerating solution, and rinse water were changed over several iterations to attain the maximum heat stable salt removal rate. The flow direction through the canisters was changed as well. The ion exchange resin within the canister frequently became packed. This created high differential pressure limiting the flow through the resin. At specific points in the cycle of the unit the flows were reversed to "fluff" the resin eliminating the high differential pressure condition.

Also prior to the Operating Period, a pump was installed at the MDEA storage tank to supply amine to the MDEA reclaim unit during plant outages. With the installation of the alternate pump, the reclaim unit could be operating and removing heat stable salts during plant outages, which was impossible with the previous equipment arrangement.

5.3.11.2 Operating Experience Overview

The hydrogen sulfide removal efficiencies from the product syngas for the Operating Period were as follows:

	1st Quarter	2nd Quarter	3rd Quarter	4th Quarter	Average / Total
H$_2$S Removal Efficiency (%)	98.9	98.5	98.5	98.5	98.6
Syngas Processed (MMscf)	11,038	6,264	10,255	5,619	33,176

The removal efficiency calculation uses total combustion turbine stack and flared syngas emissions (as sulfur) compared to the total sulfur feed to the gasification plant (dry weight percent), for the most conservative estimate of performance. Syngas processed is calculated from the monthly total produced syngas in MMBtu and the average syngas heating value (Btu/scf), while operating on coal. The total downtime associated with the acid gas removal system during the Operating Period was 22 hours. Operating experience for individual pieces of equipment follows.

Acid Gas Absorber

During the Operating Period, it was determined that by reducing the MDEA concentration in the circulating absorbent solution, higher recovery efficiency could be attained. At lower MDEA concentrations the properties of the amine solution become closer to those of water. The solution becomes less viscous, contributing to improved column hydraulics and, the composite heat capacity of the solution is increased allowing for more heat from the exothermic adsorption of H$_2$S to be absorbed without increasing the solution temperature. Since the amine solution becomes more selective toward H$_2$S than CO$_2$ at lower temperatures, the H$_2$S removal efficiency is increased.

Upon start-up of the facility on January 5, 1999, a pressure test was delayed while Operations investigated and identified a premature pressure safety valve release. This delay constituted all 22 hours of downtime attributed to this system. Subsequent investigation revealed that moisture

had condensed and frozen in the pilot sensing line for the pressure safety valve on the acid gas absorber causing the valve to relieve significantly below the relief set point. Additional insulation and heat tracing have been applied to this valve to reduce the probability of a repeated incident.

One significant project was implemented during the Operating Period. The acid gas absorber has six separate amine feedpoints. As the H_2S concentration in the product syngas increases, the amine solution is introduced at a higher feedpoint to provide more contact time between the syngas and the amine solution. Linkages between these feedpoint valves and remote operating stations were installed to allow for operation of these valves without utilizing a man-lift.

MDEA Storage Tank

A new level control philosophy for the MDEA storage tank was adopted during the Operating Period. During normal operation of the acid gas removal system, water is constantly removed from the amine solution as overheads from the acid gas stripper and blowdown from the acid gas stripper reflux drum. The old procedure for replacing this water was to "batch-in" a large amount of water when the operator noticed the tank level was low. However, this batching of water effectively reduced the amine concentration enough to cause upsets downstream in the sulfur recovery unit. To correct this problem an MDEA storage tank level controller was added to the system to automatically add water to maintain a constant MDEA storage tank level and, consequently, a constant MDEA concentration.

MDEA Carbon Bed

During the Operating Period, routine core samples taken from the MDEA carbon bed filter revealed carbon attrition and excess antifoam accumulation within the carbon bed were causing periods of high and/or increasing differential pressure. To alleviate this problem, the antifoam feed rate was reduced and the carbon bed was back-flushed with water to remove carbon fines.

Acid Gas Stripper Reflux Drum

One significant project was implemented on the acid gas stripper reflux drum during the Operating Period. During plant start-ups and shutdowns, there is insufficient heat supplied to the

column to generate the vapor necessary to maintain level in the acid gas stripper reflux drum. Therefore, the drum loses level. During the Operating Period, a dedicated line was installed to supply make-up water to the reflux drum to prevent a low-level condition.

Acid Gas Stripper Reboiler

During the Operating Period, Operations experienced difficulties starting up the acid gas stripper reboiler. Under low syngas flow conditions the amine boiler alone cannot supply the necessary heat for proper operation and the acid gas stripper reboiler must supply the additional heat duty. Often, the reboiler appears to be vapor-locked requiring Operations to vent it for several minutes before it will accept steam.

Additionally, the level control valve at the bottom of the reboiler has experienced leak-by during the Operating Period. This allows the level in the reboiler to drop and steam to flow in, adding heat to the column. Additional heat into the column results in an upset in the sulfur recovery system. The leak-by was repaired and no further problems have been experienced with this valve.

Acid Gas Stripper Carbon Beds

During the Operating Period, routine core samples were taken from the acid gas carbon beds. The results from activity tests revealed little tar accumulation on the carbon. Similar to the MDEA carbon bed, carbon attrition was contributing to slightly increasing differential pressure across the bed. This differential pressure has been resolved by employing a bed expansion procedure using nitrogen and water.

MDEA Reclaim Unit

Although no plant downtime or turndown can be attributed to the MDEA reclaim unit during the Operating Period, this unit has experienced significant reliability and performance problems. The original canisters containing the ion exchange resin were of fiberglass construction. The MDEA acted as a solvent on the resin in the fiberglass exposing the fibers. Caustic, used to regenerate the ion exchange resin, would then attack the fibers. This process continued until the canister wall became weak and ruptured. A project was implemented which increased the

volume of the canisters by approximately 20% and upgraded the material of construction. No further rupture problems have been experienced.

Over the Operating Period, several piping configuration changes were made to the MDEA reclaim unit, increasing the heat stable salt removal capacity by approximately 35%. Also during the Operating Period, process changes were made which extended the life of the ion exchange resin from a few weeks to more than a year.

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period:

Lean amine pumps	Acid gas stripper
MDEA storage tank	Acid gas stripper knockout drum
MDEA pre-filter	Acid gas stripper quench pumps
MDEA post-filter	Acid gas stripper recirculation cooler
MDEA CTW coolers	Acid gas preheater
Lean amine transfer pumps	Acid gas stripper overheads filters

5.3.11.3 Summary and Conclusions

With a few exceptions, the acid gas removal system has high reliability and low maintenance, relative to other systems within the facility. The most significant maintenance issue, which continues to be a problem, is tray damage within the acid gas absorber. As discussed in the system modification section, stiffeners have been installed to provide extra support for the trays. However, damage continues to occur. The design of a future acid gas absorber for syngas sweetening should include tray support rings extending farther out to support more of the load on the tray. The tray design should include a stiffening device to aid in supporting the tray load. Finally, the design of the retaining clips securing the tray to the support ring should be evaluated. Because of vibration, these clips frequently work lose allowing the tray to shift. This created an unequal distribution of the load, which contributed to tray damage.

Another significant maintenance issue within the acid gas removal system is the maintenance on the MDEA reclaim unit. The complexity of the continuous-feed and rotating-resin canister carousel design have contributed to poor reliability. Spatial limitations have also made maintenance difficult. Future designs for an MDEA reclaim unit should consider a three cell system employing sequenced automatic block valves for directing flow. The automatic block valves would swap the flow from the online, regeneration, and stand-by resin beds as heat stable salt breakthrough was detected.

In terms of cost of operation, the MDEA reclaim unit is also the most costly piece of equipment within the acid gas removal system. Since the Operating Period, installation of a new heat stable salt displacing solution storage tank has reduced the chemical costs associated with the reclaim unit, because shipments can now be received in bulk.

Since the Operating Period, optimization efforts on the MDEA reclaim unit have appreciably increased the heat stable salt removal capacity such that salt removal has exceeded salt generation.

5.3.12 Sulfur Recovery System

The sulfur recovery system includes:

Acid gas burner	Hydrogenation cooler (Figure 5.3.12A)
Acid gas burner residence vessel	Hydrogenation steam drum
Acid gas heat recovery boiler	Condensate flash drum
Burner sulfur condenser (Figure 5.3.12A)	Tail gas quench column
Sulfur storage tank (Figure 5.3.12A)	Quench column pumps
Sulfur pump	Quench filter
Sulfur loading arm	Quench strainers
No. 1 gas heater	Quench column cooler
Claus 1st stage reactor (Figure 5.3.12A)	Tail gas recycle compressor
No. 2 sulfur condenser (Figure 5.3.12A)	Tank vent blowers
No. 2 gas heater	Combustion air blowers
Claus 2nd stage reactor (Figure 5.3.12A)	Suction vent gas blower knockout pot
No. 3 sulfur condenser (Figure 5.3.12A)	Tail gas burner
No. 3 gas heater	Tail gas incinerator residence vessel
Claus 3rd stage reactor (Figure 5.3.12A)	Tail gas quench cooler
No. 4 sulfur condenser (Figure 5.3.12A)	Tail gas dispersion stack
Hydrogenation gas heater	Low-pressure steam drum
Hydrogenation reactor (Figure 5.3.12A)	
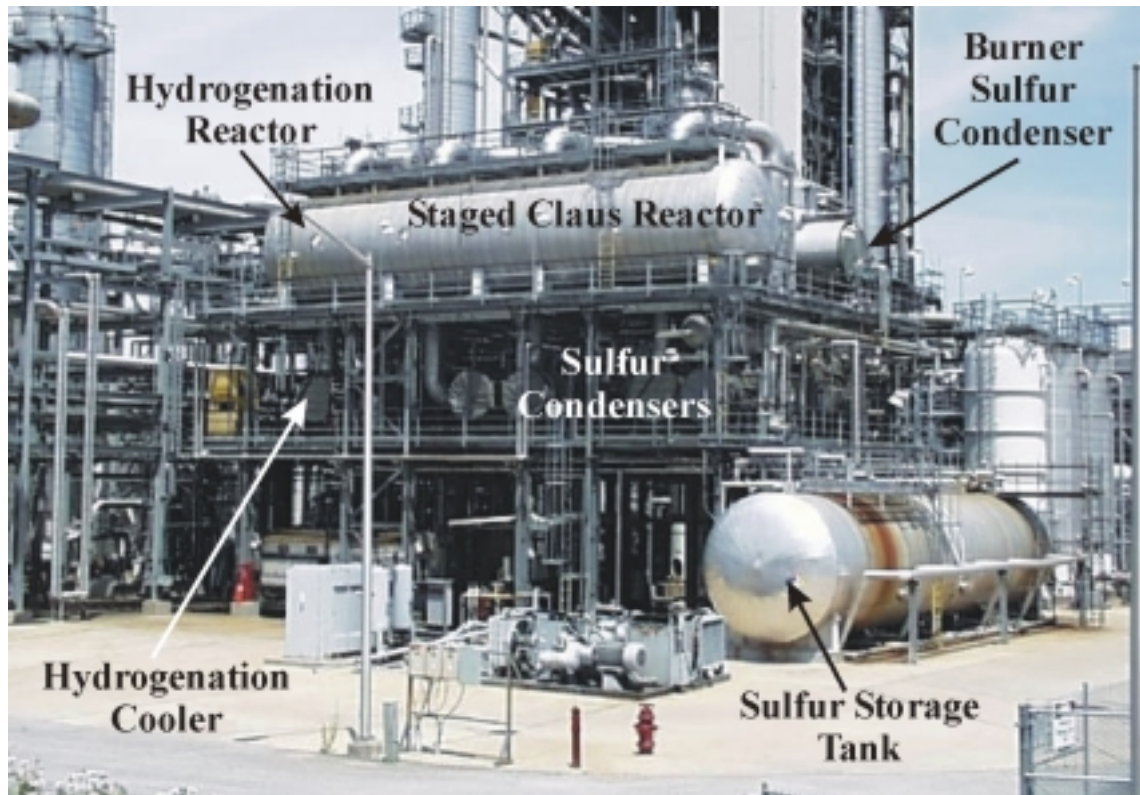
Figure 5.3.12A: Sulfur Recovery Unit Major Equipment

5.3.12.1 System Modifications

Prior to the Operating Period, many modifications were made to the sulfur recovery system. These modifications are discussed in order of upstream to downstream equipment.

Acid Gas Burner

Shortly after start-up of the facility, it was discovered that the automatic block valve in the acid gas line feeding the acid gas burner was severely leaking through. During a subsequent outage, the valve was pulled and inspected and the seat of the valve was found to be significantly corroded. At that time, the decision was made to change the seat material to a more corrosion resistant material. Since then, there have been no further leaking problems with the valve.

No. 1 Gas Heater

During winter conditions when the No. 1 gas heater was running on high-pressure steam and the block valves for the intermediate pressure steam feed were closed, the intermediate steam line froze and failed to function when needed. To prevent this from happening again, a steam trap

was installed on the intermediate pressure steam header. Instead of accumulating and freezing in the header, condensing liquid will flow through the trap keeping the header steam filled and hot.

No. 4 Sulfur Condenser

During operation of the sulfur recovery system prior to the Operating Period, it was discovered that there was a moderate degree of entrainment of liquid sulfur from the No. 4 sulfur condenser to the hydrogenation gas heater. This entrainment occurred mainly during upset conditions. Therefore, a seal leg was installed at the entrance of the hydrogenation gas heater to drain the sulfur from the head of the exchanger. Though the sulfur flow from this seal leg is infrequent, if it were allowed to accumulate within the hydrogenation gas heater it could create excessive differential pressure within the sulfur recovery unit.

Hydrogenation Reactor

After changing the catalyst in the hydrogenation reactor, it is necessary to ensure the catalyst is in the sulfide state. To sulfide the hydrogenation catalyst, a dedicated line was installed from the acid gas inlet to the sulfur recovery system to the inlet to the hydrogenation reactor. Since start-up, this line has been used on both catalyst replacement occasions.

Sulfur Storage Tank

The sulfur storage tank vent line was also modified prior to the Operating Period. Originally, the vent line from the sulfur storage tank tied into the inlet of the hydrogenation gas heater. However, this created a backpressure scenario, which caused sulfur to overrun the liquid sulfur inspection boxes whenever the level in the sulfur storage tank was greater than approximately 50%. Consequently, the sulfur storage tank vent line was rerouted to the tail gas burner eliminating the backpressure.

Sulfur Pump

Due to thermal cycling of the sulfur storage tank and the suction piping leading to the sulfur pump, stresses were continually imposed on the pump coupling. These stresses eventually led to a failure of the coupling. Therefore, a flexibly mounted bedplate was installed to allow the pump and motor to move together as one unit. This eliminated the stress, preventing future failures.

Sulfur Loading Arm

To prevent overfilling a sulfur railcar while pumping down the sulfur storage tank, an instrument air bubbler was installed on the sulfur loading arm. Previously, there were temperature elements protruding from the loading arm into the railcar. Once sulfur reached these temperature elements, the railcar fill would be terminated. With the addition of the bubbler, Operations had an analog indication of the level. This made the control system associated with preventing overfilled railcars more reliable.

Tail Gas Quench Cooler

Prior to the Operating Period, several modifications were made to the tail gas quench cooler. While operating the tail gas burner at elevated firing rates, a high level of vibration noise can be observed from the tail gas quench cooler. It is assumed that the noise is coming from vibration of the tubes inside the exchanger. Consequently, binding rods were installed to limit the movement of the tubes. The first set of binding rods was successful in reducing the vibration. However, the material of construction was inadequate to resist corrosion within the vessel. After selecting the proper metallurgy, the binding rods have continued to significantly dampen the vibration from the tail gas quench cooler.

Samples from the tail gas quench cooler drum need to be routinely tested to determine conductivity. The original sampler allowed part of the sample to boil off if a large amount of sample were taken. This boiling rendered the sample useless because it altered the contaminant concentrations. Therefore, a sample cooler was installed to allow Operations to withdraw a continuous sample without risk of boiling.

Changes were also made to two of the overpressure protective devices on the tail gas quench cooler prior to the Operating Period. Below two sets of pressure safety valves, rupture disks were installed to prevent leakage and multiple repairs to the pressure safety valve.

Tail Gas Recycle Compressor

Related to the tail gas compressors, a piping change was necessary prior to the Operating Period. The original design had the compressor second and third stage suction drum liquid drain lines tied into the discharge of the area sump pump. When this pump was running, water could back flow through the compressor suction drum into the compressor. Once water reached the compressor, there would be a great risk of damaging the equipment as well as potential for personnel injury. Therefore, dedicated lines and control valves were installed routing the suction drum drains to the recycled solids tank.

Suction Vent Gas Blower Knockout Pot

Two minor modifications were made to the suction vent gas blower knockout pot before the Operating Period. A gate valve was installed on an unused nozzle on the knockout pot to be used for troubleshooting. Previously, water accumulation was thought to have contributed to the inadvertent shutdown of the tail gas burner. This gate valve could be used to quickly determine if any water was present and could be used to de-inventory the vessel if necessary. Additionally, the reference point for the liquid level transmitter on the knockout pot was changed. Originally, the transmitter was referenced to atmosphere. This resulted in unreliable and false drum level readings. Relocation of the high side tap to the knockout pot seal leg allowed for earlier detection of increases in level.

Tank Vent Blowers

The downstream explosive gas analyzer for the tank vent blower was moved to upstream of the suction vent gas blower knockout pot. Reliability of the analyzer was compromised by being under pressure during operation and calibrated under a vacuum. The analyzer downstream of the blowers typically recorded poor reliability compared to the redundant analyzer upstream of the knockout pot.

A final modification to the tank vent blowers was the alteration to the shrouds on the outboard bearing. This modification was to allow for vibration readings to be taken without removing the shroud. Shroud removal while the machine is in operation could cause overheating of the bearings, shortening bearing life in addition to presenting a personnel safety issue.

5.3.12.2 Operating Experience Overview

The sulfur recovery unit and plant overall efficiencies for the Operating Period are as follows:

Month	Sulfur Recovery Efficiency	Overall Efficiency
March 1998	98.7	97.8
April 1998	99.3	98.4
May 1998	99.6	98.5
June 1998	99.3	98.1
July 1998	99.8	98.5
August 1998	99.4	98.1
September 1998	99.3	98.3
October 1998	99.7	98.3
November 1998	99.7	98.1
December 1998	98.0	97.6
January 1999	99.6	98.2
February 1999	99.8	98.3
Operating Period Average	99.4	98.2

The efficiency calculation is the total stack emissions and acid gas flared emissions (as sulfur) compared to the total feed to the sulfur recovery system. The overall efficiency corresponds to the total project emissions (as sulfur) compared to the total sulfur feed to the gasifier.

The total downtime associated with the sulfur recovery system during the Operating Period was 5 hours. Operating experience for individual pieces of equipment follows.

Acid Gas Burner

There were no significant maintenance issues related to the acid gas burner during the Operating Period. However, there was a change in operating philosophy related to the firing rate for the burner. Previously, the operating philosophy was to achieve the complete destruction of ammonia. However, after receiving results from tests performed on the Claus 1st stage reactor, ammonia destruction was no longer targeted. The results from the catalyst activity tests indicated a heavy amount of coking. The excessive supplemental fuel required to reach

temperatures to completely destroy the ammonia causes uncombusted carbon to foul the Claus catalyst. Consequently, the philosophy was changed to one of fuel conservation.

One potentially serious operational incident did take place concerning the acid gas burner during the Operating Period when an operator, during a routine inspection, found a leak on the acid gas burner pilot. While attempting to tighten the pilot retaining nut the pilot became lose and the leak increased. The retaining nut was tightened, securing the pilot and gasket material was wrapped around the pilot. The gasket material slowed the leak enough for sulfur vapor, escaping from the acid gas burner, to condense and solidify, affecting a seal of the leak.

Acid Gas Heat Recovery Boiler

A minor maintenance issue occurred related to the acid gas heat recovery boiler during the Operating Period. At the outlet of the boiler, there is a steam-jacketed plug valve used for troubleshooting. During periods of elevated pressure within the sulfur recovery system, valves such as these are used to determine where a flow restriction is likely. During a routine inspection round, an operator noticed a sulfur formation hanging from this valve indicative of an acid gas leak. An H_2S monitor was used to determine the severity of the leak and it was decided that the leak was minimal and the valve could be, and was, repaired on the subsequent outage.

Burner Sulfur Condenser

At the bottom of each of the sulfur condensers within the sulfur recovery system, there is a seal leg that prevents acid gas from reaching the sulfur storage tank. During the Operating Period, these seal legs routinely became plugged, requiring Operations to clear the obstruction by use of a hand held rod. The material accumulating in the seal legs has been analyzed and found to be an amorphous mixture of ammonia salts, iron corrosion products and soot from the excess firing of the acid gas burner. A clean-out flange has been installed on the bottom of the seal leg to minimize the maintenance costs of cleaning the seal leg on the burner sulfur condenser. Once an appreciable amount of material has accumulated in the bottom of the seal leg, the flange is removed on the following outage and the material is removed. This cleaning method has eliminated seal leg plugging in the burner sulfur condenser.

Sulfur Storage Tank

There was little significant maintenance work during the Operating Period related to the sulfur storage tank. The notable exception was clearing of a restriction in the storage tank vent line. The liquid sulfur flow to the sulfur storage tank is saturated with H_2S. Once the liquid reaches the storage tank, degassing occurs, liberating the H_2S. This gas, in conjunction with a nitrogen purge placed on the tank, causes the tank pressure to rise if the sulfur storage tank vent line is not clear. During the Operating Period, a significant plug in the vent line occasionally created operational difficulty. In a major maintenance outage during the Operating Period, a sulfur plug was detected and cleared.

After the Operating Period, it was discovered that a portion of the steam-jacketed vent line had collapsed. It was thought that the steam jacketing froze. The force from the expansion of water turning to ice was great enough to crush the vent line. This restriction was contributing to the sulfur plugging during the Operating Period.

Sulfur Pump

During the Operating Period, there was no maintenance activity related to the sulfur pump and only one notable operational concern. While attempting to transfer liquid sulfur from the sulfur storage tank to a sulfur railcar, a steam trap of the steam-jacketed line was misaligned. This created a cold spot in the transfer line allowing the sulfur to solidify and plug the line. Since the plug prevented sulfur flow from the pump, the liquid sulfur within the pump heated up. As the sulfur heated up, the viscosity increased. This is a unique characteristic of sulfur. Consequently, the sulfur pump continuously tripped because of high amperage. The misaligned steam trap was soon discovered, repaired, and proper operation resumed.

Sulfur Loading Arm

During the Operating Period, the sulfur railcar supplier changed. The new railcar supplier had railcars within the fleet that were retrofitted corn syrup cars. To accommodate the change in fill level for the new cars, modifications to the sulfur loading arm and control code were necessary. The level bubbler tube on the sulfur loading arm was extended to protrude further into the railcar. This allowed the operators to accurately monitor the fill volume of the new railcars.

Claus 1st Stage Reactor

During the Operating Period, the performance of the Claus 1st stage reactor degraded. The temperature profile of the bed began to drift deeper into the bed, indicating deactivation from the top. During the next major maintenance outage, the catalyst was replaced and tested. As previously mentioned, the test results indicated a high degree of coking on the catalyst.

No. 2 Sulfur Condenser

Two notable maintenance activities occurred relating to the No. 2 sulfur condenser during the Operating Period. First, a steam leak was observed coming from the jacketed condenser boot valve. During the next maintenance outage, a steam union was serviced and the leak was stopped. Second, a steam jacketed plug valve, similar to the valve on the acid gas heat recovery boiler, was found to be leaking. The valve was replaced on a subsequent outage and no further maintenance was necessary.

No. 4 Sulfur Condenser

The No. 4 sulfur condenser has been a source of significant maintenance costs within the sulfur recovery system. Because of temperature and ammonia concentration considerations, the ammonia salt formation is greatest in this condenser. It has not been uncommon to open the head of the condenser and find a few feet of solidified sulfur (Figure 5.3.12B). This sulfur is trapped because of a plug in the seal leg or the run down line to the seal leg. Routinely, during the Operating Period, the liquid sulfur piping downstream of the No. 4 sulfur condenser was disassembled and cleared of plugging deposits.

To reduce the maintenance costs associated with cleaning this piping, the welded steam unions to the steam jacketing were replaced with flanges. Additionally, to reduce costs associated with cleaning the seal leg, a clean out flange similar to that installed on the burner sulfur condenser seal leg has been installed.

Figure 5.3.12B: Sulfur in No. 4 Sulfur Condenser as a Result of a Plugged Seal Leg

Hydrogenation Reactor

At the same time the catalyst was changed out in the Claus 1st stage reactor, the catalyst was changed in the hydrogenation reactor. The temperature profile of the bed indicated a potential reduction in bed activity and downstream gas sampling confirmed the reduction. Upon start-up of the hydrogenation reactor, the catalyst needed to undergo a sulfiding process. During this process, the sulfur recovery efficiency was inadequate to maintain plant operation. Consequently, 5 hours of downtime was attributed to the hydrogenation reactor while the catalyst was being prepared. This constitutes all of the downtime associated with the sulfur recovery system during the Operating Period.

The hydrogenation reactor bypass valve experienced a leak during the Operating Period. Consequently, SO_2 was allowed to bypass the reactor. Recombining with the H_2S from the outlet of the hydrogenation reactor, the SO_2 underwent an aqueous Claus reaction within the tail gas quench column. This created significant operational difficulty as well as appreciable maintenance costs. The results will be further discussed in the tail gas quench column operating experience section. Once the leak was identified, the valve was serviced. Upon inspection, it was found that corrosion products formed a deposit and damaged the seat of the bypass valve preventing a proper seal.

Tail Gas Quench Column

As a result of the leaking hydrogenation reactor bypass valve, an aqueous phase Claus reaction occurred with the tail gas quench column forming a pale yellow paste of elemental sulfur. This paste plugged every piece of equipment, to some degree, within the tail gas quench column loop. The packing within the tail gas quench column was packed with the sulfur paste. This created high differential pressure across the column. The differential pressure across the column starved the tail gas compressors of flow causing them to trip on low suction pressure. This, in turn, created high pressure within the sulfur recovery unit causing elevated sulfur emissions at the tail gas dispersion stack.

After water flushing methods were exhausted, a costly chemical cleaning of the column and related equipment was performed. During this cleaning, hot caustic was circulated throughout the quench column loop to dissolve the elemental sulfur. The chemical cleaning was very successful, but the maintenance and caustic disposal costs were quite high.

Quench Column Pumps

Because of the elemental sulfur paste, the quench column pumps experience operational difficulty. The seal flush on these pumps typically comes from the discharge of the pump. However, the sulfur paste plugged the seal flush, contributing to failure of the pump seal. This occurred several times before the seal flush was replaced with a clean water source.

Quench Filter and Quench Strainers

These vessels are responsible for removing entrained particulate matter from the water in the tail gas quench column loop. Because of the sulfur paste incident, the operational cost of these vessels was high for a portion of the Operating Period. The additional load on the Operations team as a result of constant changing of the quench filters was the single greatest operational difficulty experienced in the sulfur recovery system during the Operating Period. As a result of constant use in a two-phase application, the isolation valves on the quench strainer became difficult to operate. Once the sulfur formation problem was solved, the valves were serviced.

Quench Column Cooler

During the sulfur paste formation incident, the quench cooler became packed with sulfur. This created an appreciable flow restriction as well as significantly reducing the heat removal from the quench loop. A system of blowing down the cooler and making up with fresh water was devised to satisfy temperature requirements within the column.

Tail Gas Recycle Compressor

The tail gas recycle compressors have been a source of significant maintenance during the Operating Period. Much of the maintenance has been related to the internal check valves within the compressor. Frequent check valve failures have contributed to poor compressor reliability. However, changes in check valve design throughout the Operating Period have helped increase the reliability of the machines.

During the Operating Period, the seal leg on the first stage suction drum to the compressors was extended. Due to small suction pressure fluctuations, the seal leg would de-inventory allowing tail gas to escape to a nearby sump. The seal leg was extended to provide greater resistance to the pressure fluctuations.

Figure 5.3.12C: Tail Gas Recycle Compressors

Tank Vent Blowers

During the Operating Period, the tank vent blowers also experienced appreciable operational reliability problems and high maintenance. A sulfur compound routinely fouled the impellers of these blowers. To reduce the fouling, an epoxy coating was applied. This coating proved unsuccessful; however, as corrosion underneath the coating loosened it, causing it to flake off.

Tail Gas Burner

During the Operating Period, an alternate mode of operation was identified and tested for the tail gas burner. It was discovered that fuel costs could be reduced if the tail gas burner could be fired on syngas derived from natural gas during plant start-ups rather than burning the syngas at the flare and firing the burner on supplemental fuel gas. The test was successful in that the burner was able to sustain operation. However, because of piping limitations and reduced gas path pressures during start-up, Operations found it difficult to maintain the required temperature within the tail gas incinerator residence vessel.

Tail Gas Dispersion Stack

During the Operating Period, one noteworthy maintenance activity occurred related to the tail gas dispersion stack. A large gap was discovered in a nozzle located on the stack. Upon inspection, it was determined that the combination of sulfur compounds and condensed moisture contributed to downtime corrosion of the nozzle. The nozzle was temporarily repaired until the next maintenance outage when it was replaced.

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period:

Acid gas burner residence vessel	Condensate flash drum
No. 1 gas heater	Hydrogenation gas heater
No. 2 gas heater	Hydrogenation cooler
Claus 2nd stage reactor	Combustion air blowers

No. 3 sulfur condenser	Suction vent gas blower knockout pot
No. 3 gas heater	Tail gas incinerator residence vessel
Claus 3rd stage reactor	Tail gas quench cooler
Hydrogenation steam drum	Low-pressure steam drum

5.3.12.3 Summary and Conclusions

Since the Operating Period there have been a few noteworthy system modifications. As a result of an analyzer failure, the plant experienced a release of a reportable quantity of SO_2 from the permitted tail gas dispersion stack. To aid in troubleshooting the sulfur recovery system and detecting failures of the analyzer used for control of the acid gas burner, a secondary tail gas analyzer has been installed. Thus far, Operations personnel have not had an opportunity to utilize the new analyzer.

To prevent the sulfur tank vent line from plugging with corrosion products, a nitrogen sweep has been installed. Since installation, the sulfur tank has experienced no elevated pressure problems.

A new method of cleaning the tail gas dispersion stack has been developed and tested since the Operating Period. The inner wall of the dispersion stack continuously becomes layered with a sulfur compound as a result of the combustion of tail gas and subsequent natural cooling of the gas as it flows up the stack. Under upset plant conditions, high gas flow rates through the dispersion stack cause the sulfur compounds to dry and flake off and exit the stack.

To remove the layers of sulfur and prevent the flaking, an explosive expert team was contracted to clean the stack. The contractor constructed an assembly using wire hoops and approximately 500 feet of detonation cord and erected the assembly into the stack. Upon detonation, a small plume of fine dust exhausted from the top of the dispersion stack. However, the vast majority of the material fell to the bottom of the stack and collected in the tail gas quench cooler, from which it was collected and removed. The explosive cleaning of the tail gas dispersion stack was considered a complete success and will be employed in the future as a viable replacement for chemical or high-pressure water cleaning methods.

Finally, there is currently a project under review, which would eliminate a large portion of the ammonia feeding the sulfur recovery system. The largest source of maintenance costs for the sulfur recovery unit is the continuous removal of debris in the seal legs. The plugging material is composed of two compounds, ammonia salt and an amorphous mixture of iron corrosion products and carbon. The temperature can be increased in the acid gas burner to effect the destruction of the ammonia thus preventing ammonia salt formation. However, this decreases the life of the No. 1 stage Claus reactor and increases the carbon residue. The ideal solution, which should be considered while designing future sulfur recovery systems within coal gasification facilities, would be to eliminate the ammonia source for the sulfur plant.

5.3.13 Sour Water Treatment System

The sour water treatment system includes:

Sour water carbon filter (Figure 5.3.13A)	Stripped water transfer pumps
Sour water filter	Ammonia stripper bottoms cooler
Condensate degassing column (Figure 5.3.13A)	Quench column (Figure 5.3.13A)
Sour water transfer pumps	Quench water transfer pumps
Degassing column reboiler	Quench column bottoms cooler
Degassing column bottoms cooler	Supplemental quench column bottom cooler
Ammonia stripper (Figure 5.3.13A)	Outfall water day tanks
Ammonia stripper reboiler	Discharge water pumps

Figure 5.3.13A: Sour Water Treatment System Major Equipment

5.3.13.1 System Modifications

Prior to the Operating Period, the sour water carbon filters routinely experienced periods of high differential pressure. Due to small amounts of vapor flashing in the sour water stream feeding the carbon beds, frequent venting of the vessel head spaces was required to avoid vapor-locking the beds. To remedy this problem, a dedicated line was installed from the top of the carbon bed vessels to the gas inlet of the quench column. This small project solved the vapor-locking problem and prevented operators from having to manually vent the vessels to atmosphere.

As a result of the installation of the chloride scrubbing system, the chloride concentration in the feed stream to the sour water treatment system was significantly increased. The chloride in the sour water formed complexes with the ammonia, preventing the ammonia from being stripped in the ammonia stripper. It was discovered that caustic addition would dissolve the ammonia chloride complexes allowing the ammonia to be stripped. Therefore, a metering pump was installed to supply a continuous feed of caustic. Later, a carbon storage vessel was converted into a caustic storage tank. This modification decreased the operating cost of the caustic feed system allowing caustic to be purchased in bulk.

The ammonia stripper experienced a small amount of damage to the top tray upon start-up. It was thought that this damage was the result of increased feed stream velocity because of two-phase flow. To eliminate this mechanism of damage, an internal distribution system with an impingement plate was installed. Since installation, no further damage has been observed.

Finally, prior to the Operating Period, Operations experienced difficulty achieving adequate cooling of the quench column bottoms stream. Consequently, the quench column experienced high overhead ammonia flow. The effects of the elevated bottoms temperature within the rod mill building were of great concern.

The ammonia rich stream from the quench column is recycled to the rod mill to make coal slurry. Since the temperature of this stream was elevated, high levels of ammonia flashed from the liquid and entered the building housing the rod mill. To eliminate the personnel exposure concern, the supplemental quench column bottom cooler was installed. Although the installation

of the supplemental cooler made a significant improvement, the ammonia concentration within the rod mill building has occasionally been problematic.

5.3.13.2 Operating Experience Overview

The water volume processed by the sour water treatment system during the Operating Period is as follows:

Month	Volume (millions of gallons)
March 1998	8.2
April 1998	8.6
May 1998	7.5
June 1998	7.3
July 1998	7.0
August 1998	8.4
September 1998	7.8
October 1998	3.9
November 1998	3.8
December 1998	0.7
January 1999	2.0
February 1999	2.8
Operating Period Total	68.0

No downtime was associated with the sour water treatment system during the Operating Period. However, coal operation was terminated as a result of high differential pressure across the sour water carbon beds. The shutdown time for this incident was later attributed to the particulate removal system. Operating experience for individual pieces of equipment follows.

Sour Water Carbon Filter

The intent of the carbon beds is to absorb tars that have condensed along with the sour water in the low temperature heat recovery system. Because of the nature of the carbon beds; however, they also act as a filter to remove any particulate matter that is suspended in the sour water stream. During the Operating Period, a failure in the particulate removal system allowed char to pass into the particulate-free gas stream where it was then removed from the syngas by the chloride scrubber. When the water from the chloride scrubber reached the sour water carbon filter, the char was collected in the carbon beds. This char packed into the void spaces in the carbon beds, creating high differential pressure. As a result, the screens, responsible for retaining the carbon in the vessel, were crushed into the floor of both vessels. The distribution systems in each vessel were also severely damaged. Internals from both vessels were salvaged to reconstruct a complete set so that operations could continue with one vessel. The remaining vessel was isolated for repairs. Internals for the off-line vessel were fabricated and installed while the facility was operating. Built into the new distribution system were additional supporting members, which could withstand greater differential pressures. The carbon in this vessel was also replaced. Imbedded in the spent carbon from the damaged vessel were accumulations of tars that had been washed from the chloride scrubber. As discussed in the chloride scrubbing system section, tar accumulation was a problem in the chloride scrubber prior to the Operating Period.

As a result of using the carbon unload line on the sour water carbon filter, another maintenance issue arose during the Operating Period. Carbon from the vessel damaged the seats of the unload line block valves on each vessel. Shortly after placing both carbon filters in service, leaks were discovered from the unload lines of the vessels. These valves were serviced on a subsequent plant outage and have since been maintenance-free.

Sour Water Filters

The water filters are designed to filter any particulate material that breaks through the screens of the sour water carbon filter. Because of the damage to the sour water carbon filter described previously in this section, large amounts of carbon filled and packed into the water filters. The filter cartridges collapsed and carbon was allowed to pass into the sour water treatment system.

The carbon packed into the water filters caused minor damage to the filter cartridge supporting struts. The carbon was removed from these vessels and the damage was repaired. There were no other maintenance concerns with the water filters during the Operating Period.

Condensate Degassing Column

Carbon from the collapsed water filter, previously discussed, passed into the condensate degassing column. Although this caused no damage to the degassing column, it created operational difficulties. Over a period of several days frequent back-flushing of both sour water transfer pumps was necessary to prevent a high water level condition within the condensate degassing column.

Sour Water Transfer Pumps

Throughout the Operating Period, the sour water transfer pumps occasionally experienced poor performance. It was determined that the suction strainers were plugging with foreign material. Once the suction strainers were cleared of debris, the sour water transfer pumps exhibited acceptable performance.

Degassing Column Reboiler

During routine daily inspections, Operations discovered a small steam leak from the head of the degassing column reboiler. During the next planned outage, the head of the exchanger was removed and a small amount of pitting due to corrosion was discovered on the sealing surface between the head and the exchanger body. After the pitting was ground out, the gasket was replaced and the reboiler was placed back into operation without further leakage.

Degassing Column, Ammonia Stripper, and Quench Column Bottoms Coolers

Following the carbon breakthrough incident, each of the coolers was flushed to remove any remaining carbon. After flushing, each was opened for inspection. The degassing column bottoms and quench column bottoms coolers appeared to be in proper condition. The ammonia stripper bottoms cooler; however, was heavily covered with corrosion by-product deposition. The exchanger was satisfactorily chemically cleaned.

Ammonia Stripper

As with the level control problems in the degassing column following the carbon breakthrough incident, Operations was forced to frequently back-flush both stripped water transfer pumps in order to maintain proper level within the ammonia stripper.

Over the Operating Period, the pressure rupture disk on the ammonia stripper was replaced. Due to poor stripper water transfer pump performance, Operations experienced level control problems within the ammonia stripper. As a result of the variable level and abnormal overheads flow, the column experienced a high-pressure condition and relieved the pressure through the pressure rupture disk. The pressure rupture disk was replaced on the subsequent plant outage.

Because tray hardware, such as clamps, bolts and bubble caps, were occasionally found in the suction strainer to the stripped water transfer pumps, the ammonia stripper was inspected for tray damage. Upon inspection, minor tray damage was found and repaired. There was no other maintenance concern associated with the ammonia stripper during the Operating Period.

Stripped Water Transfer Pumps

During the Operating Period, the sour water transfer pumps frequently experienced poor performance due to plugging of the suction strainers with foreign material. This phenomena was intensified during the carbon breakthrough incident. However, unlike the sour water transfer pumps, the stripper water transfer pump performance continued to suffer after the suction strainers were cleaned. To resolve the stripped water transfer pump poor performance problems experienced during the Operating Period, an epoxy coating was applied to the pump impeller casing. A significant amount of corrosion was discovered on the casing, which increased the clearance between it and the impeller. The additional clearance allows water to recirculate to the suction side of the pump instead of out the pump discharge, lowering performance. The coating reduced the clearance between the impeller and casing significantly reducing recirculation. The same coating was also applied to the sour water transfer pumps.

<u>Discharge Water Pumps</u>

The upstream manual block valve for the discharge water pumps failed to provide a positive seal during the Operating Period. The valves were removed from the lines and inspection revealed that the seats of the valves were severely corroded. The seats on the valves were replaced like-for-like and the pumps were put back into service.

<u>Other Equipment</u>

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: quench column, quench water transfer pumps, supplemental quench column bottom cooler, and the waste water discharge day tanks.

5.3.13.3 Summary and Conclusions

Since the Operating Period, the sour water treatment plant has experienced few operational difficulties and a low level of maintenance activity.

Two noteworthy alternative modes of operation have been developed for the sour water treatment plant since the Operating Period. First, the continued operation of the treatment system during a plant outage was implemented. During major maintenance outages water management becomes a problem within the facility. Often, rented storage tanks are required due to the amount of water generated during shutdown and maintenance activities. To avoid the expense of these rented tanks, Operations developed a method to continue operation of the sour water treatment system during an outage to clean and discharge the water and eliminate the water management issue. Though Operations proved that this mode of operation was possible, it is not likely that it will be employed often. Because of flow limitations of the discharge water pumps, operation of the treatment plant proved to be quite labor intensive and water filter cartridge changes became relatively expensive.

The second alternative mode of operation proved to be far more successful. During periods of low plant throughput or during petroleum coke operation, the water feed to the sour water treatment plant is significantly reduced. This creates operational difficulties because proper column hydraulics are not maintained. However, Operations developed a method of recycling

water from the ammonia stripper to the degassing column. This method creates a false loading of the sour water treatment plant and maintains proper column hydraulics.

Significant corrosion to the discharge check valves of these same pumps also contributed to poor pump performance. The corrosion damage to these check valves allowed water from the discharge of the on-line pump to flow through the back-up pump, back to the suction of the on-line pump. To eliminate this corrosion, the metallurgy of the check valves was changed.

The design of a sour water treatment system for a future coal gasification facility should consider the following alterations: the metallurgy of the system should be evaluated, considering the maintenance history and operational experience of this facility. Corrosion to the heat exchangers, pumps and piping within the treatment system has created operational difficulties and appreciable maintenance costs.

Finally, the destination of the overhead streams from the degassing column and the ammonia stripper should be re-evaluated. As discussed in the sulfur recovery system section of this document, ammonia salt formation in the sulfur recovery system has been a significant source of maintenance costs. A large fraction of the ammonia in the sulfur recovery system comes from the sour water treatment system. If this ammonia could be directed to an alternative destination, much of the maintenance within the sulfur recovery system could be avoided.

5.3.14 Cooling Tower System

The cooling tower system includes the following equipment:

Cooling water pumps (Figure 5.3.14A)	Cooling tower (Figure 5.3.14A)

The cooling tower is a three-cell, induced-draft, counter-flow type tower used to remove low-energy, non-recyclable waste heat from the air separation unit and gasification process equipment. Two cooling water pumps circulate cooling water to plant process equipment. Air is drawn through the packing by three induced draft fans, located atop each cell within a sealed plenum. Makeup water from PSI's water treatment system is used to replenish water losses from evaporation and cooling tower blowdown.

Figure 5.3.14A: Cooling Tower Water System

5.3.14.1 System Modifications

Start-up of the cooling tower during plant commissioning in 1995 was mechanically uneventful. However, the supply water to the tower was high in total solids necessitating the use of side-stream filtration units to prevent fouling and corrosion of process equipment. Unexpectedly high corrosivity and solids levels in the supply water resulted in the complete revision of the planned treatment program for the cooling tower. Control of deposition and under-deposit corrosion was partially alleviated by use of an iron dispersing agent and tighter pH control by PSI's water treatment facility. The supply water quality to the cooling tower remained problematic through 1996, 1997 and 1998. Maintenance for the cooling tower equipment was limited to preventative and minor maintenance during the period 1996 through 1998.

5.3.14.2 Operating Experience Overview

There was no downtime associated with the Cooling Tower during the Operating Period. Operating experience for individual pieces of equipment follows.

Cooling Tower Pumps

The cooling tower pumps are set up with two operating pumps and an on-line spare. To maintain comparable operating hours on all three pumps, the operating pumps are rotated as part of an established predictive/preventive maintenance plan. The cooling tower pumps were in service without major maintenance for the entire Operating Period. Maintenance items were limited to preventative oil changes, packing gland leak repairs and lantern ring changes.

Cooling Tower Fans

The three cooling tower fans were in service for the entire Operating Period. Several fuses were replaced in fan motor speed controllers before a mismatch between the Programmable Logic Controller (PLC) language for the controllers and the Central Control System language was recognized as the cause of this problem. Other maintenance was preventative only during the Operating Period.

Other Equipment

There was no maintenance, downtime or operational problem associated with the following equipment during the Operating Period: structural timbers, polymer honeycomb fill, drift eliminators, distribution system, fan stacks, casing, partitions and windwalls, sprinklers and hangers for the fire protection wet down system.

5.3.14.3 Summary and Conclusions

The cooling tower has performed to design specifications. However, the load placed on the system by a higher-than-expected ASU demand has limited the cooling capacity available for other parts of the facility. In particular, the higher-than-expected temperature of the cooling tower water curtails the amine unit absorption capacity during hot, summer days. This necessitates running the amine absorption system at a higher contact rate than desirable. A larger capacity cooling tower or a separate refrigeration unit for the amine system should be evaluated for the next plant. After the Operating Period, the entire cooling tower system was taken out of service and inspected. The only finding of significance was that the original carbon steel hangers for the fire protection distribution piping were badly corroded. The hangers were replaced.

5.4 Power Block

5.4.1 Combustion Turbine

The combustion turbine (CT) is a General Electric MS-7001F single-shaft combustion turbine, designed for operation as a simple-cycle unit or in a combined steam and combustion turbine cycle. The MS-7001F combustion turbine assembly contains the following major sections:

Air inlet plenum (Figure 5.4.1A)	Compressor
Combustion system	Turbine/Generator
Exhaust duct (Figure 5.4.1A)	

Figure 5.4.1A: Combustion Turbine

5.4.1.1 System Modifications

Prior to the Operating Period, the combustion liners failed due to extensive cracking. The manufacturer was contacted and the liners were replaced with like-for-like material until an upgrade or resolution could be found. Recent inspection revealed continued cracking of the liner and deterioration of the coating material. Problems continue to plague the combustor liners in the turbine due to the elevated firing temperatures of the 7FA turbine. While the unit continues to operate within the temperature specifications outlined by the manufacturer, cracking and liner wear continue to be a problem. The manufacturer is currently working on a new liner design that will incorporate an upgraded liner coating, which should increase liner life.

The original NO_x steam injection control valves have had a tendency to stick or become sluggish during operation. These valves play a critical role in the control of nitrogen oxide emissions and modulate to vary the amount of steam injection, depending on emission levels. To resolve the problem, the equipment manufacturer has recommended a change in these valves and they will be replaced with an upgraded version in 2000.

5.4.1.2 Operating Experience and Overview

Combustion System

In late March of 1998, a stop ratio fuel control valve failed on the combustion section of the CT due to failure of a hime joint actuator. The original equipment manufacturer was contacted and the hime joint design was changed. There have been no further failures of this valve since replacement of the hime joint; however, the OEM has recommended installation of an upgraded valve with an improved seal design be installed at a later date.

In mid-April, start-up of the facility was delayed due to an exhaust temperature differential warning that prevented synchronization of the turbine with the generator. Outlet temperature thermocouples are located around the circumference of the combustor exhaust to ensure even temperature distribution and proper fuel flow/ignition in the combustion section of the CT. During start-up on fuel oil, the temperature indications revealed an uneven distribution of temperature in the exhaust and the CT was prevented from synchronizing with the

turbine/generator. Due to a delay in determining the cause of the failure, the gasification plant had to be taken off of coal operations. The exact cause of the uneven temperature distribution was never determined. The OEM recommended removal of all fuel inlet nozzles for inspection and flow testing before operation could be resumed. After conducting the inspection, the unit was restarted with no further problems.

Other Equipment

There was no maintenance downtime or other operational problem associated with the air inlet plenum, compressor, turbine generator or exhaust ducting during the Operating Period.

5.4.1.3 Summary and Conclusions

Overall, the combustion turbine and its components have operated as expected since the project began operation in 1995. While increased operating temperatures on the 7FA turbine have identified areas requiring additional engineering, the unit has operated within the design specifications of the manufacturer. Operational difficulties associated with initial start-up and transfer procedures from fuel to syngas operation have been resolved and critical transferring equipment internal to the CT has been identified and put on appropriate preventative maintenance schedules based on operational history. Some combustion turbine systems merit discussion regarding continued concern or modifications that have taken place since the Operating Period.

It was recognized, early in the operation of the CT, that the temperature of the inlet air during the summer months curtailed power output. During the spring of 1999, PSI installed a water mist system in the inlet air plenum to cool the incoming air, thus regaining a portion of lost power output during the hottest months. This system ran only briefly during 1999 and will be further evaluated in 2000.

In March 1999, the compressor section of the combustion turbine failed. Investigation revealed that the failure occurred in the 14th stage stator blades and propagated downstream, causing extensive damage to the compressor section. Damage from the 14th stage downstream was catastrophic in nature and required complete replacement of all rotating and stationary parts.

Due to schedule considerations and opportunities to upgrade the compressor, the decision was made to purchase and install a new upgrade from the manufacturer. The root cause of this failure has not been made public by the combustion turbine manufacturer and PSI. Other 7FA combustion turbines operating on natural gas appear to have experienced similar failures, and it has been concluded and publicized that this failure was not related to operation on syngas.

The benefits of "off-line" versus "on-line" water washing of the compressor section continues to be evaluated. Currently, the compressor is water washed off line even though some minimum amount of power degradation is realized during continuous operation. This degradation is considered minimal at this time, but on-line washing will be re-evaluated as continuous run-time on the turbine increases.

5.4.2 Heat Recovery Steam Generator

The heat recovery steam generator (HRSG) has a design steam flow of 753,180 lbs/hr at a superheater pressure of 1725 psig. The HRSG design is a bottom supported structure, which contains six circuits of convection heat transfer surfaces. There was no downtime attributable to the HRSG during the Operating Period.

5.4.2.1 System Modifications

Prior to the Operating Period it was determined by the equipment manufacturer that the feedwater heater section could not be operated. During start-up conditions the superheat, reheat and evaporator sections would not absorb sufficient heat from the combustion turbine exhaust to prevent boiling in the tubes of the feedwater heater. Additionally, when operating in full syngas operation, the feedwater heater would vaporize the water in this section which would create a flashing of steam and quenching of tube material which could eventually lead to tube damage. The feedwater heater was removed from service almost immediately after start-up in 1995 and has never been used in the HRSG.

5.4.2.2 Operating Experience

The total downtime associated with the HRSG during the operating period was 201 hours. In April, June, August and September, HRSG tube leaks prevented or interrupted gasifier operation for a total of 201 hours. All these leaks are similar due to a design oversight that did not allow for proper thermal expansion of the HRSG piping. Efforts to address these failures are further discussed in Section 5.4.2.3 below.

5.4.2.3 Summary and Conclusions

The superheat and reheat sections of the HRSG have experienced an abnormal number of tube leaks. In an effort to resolve the problems, the design engineering firm has engineered and installed several modifications; however, these modifications have not been successful to date. Engineering review of the original design has revealed that the superheater and reheater outlet headers and external piping did not allow for proper piping movement. The external piping had to have additional hangers and supports installed to ensure headers and piping move together.

Another major design concern is support of all piping circuits within the HRSG. All of the piping circuits in the HRSG are bottom supported; consequently, movement of piping during heat-up and shutdown is limited and does not provide for sufficient expansion and contraction of the tubes without excess stress being applied at tube turns. This has been the cause of numerous failures during the life of the project. It is recommended that in the installation of future HRSG's of this type that tubes be hung from the ceiling of the upper shell rather than rigidly supported from the bottom of the unit. This should allow for proper expansion and contraction with a minimum of stress on the tube turns and piping. The plant is currently pursuing a replacement design for the superheat and reheat sections of the HRSG to limit further downtime associated with this design.

Another source of power block downtime has been the failure of superheat and reheat section header expansion joints at the HRSG casing. Manufacturer inspection revealed that the penthouse roof panels were binding the headers on these two sections. When these expansion joints have failed, it has caused fires in adjacent electrical cable trays. To correct this problem, the expansion joint design has been modified and heat shielding has been installed on the electrical cable trays and conduit to protect it against heat damage. Additionally, the roof panels and retaining clips were trimmed to allow for additional expansion and contraction. Currently, these expansion joints still leak and additional expansion joint design changes are planned to correct this problem.

The HRSG inlet duct and boiler casing have experienced several hot spots and the inlet duct liners have shifted in the past, causing gas leakage in the expansion joints. To correct the problem, the inlet duct was repositioned and new duct liner locating pins were installed. The expansion joint was repaired and additional insulation was installed in the overheated areas of the duct and casing to prevent further hot spot damage.

5.4.3 Water Treatment/Handling System

The water treatment/handling system consists of the following equipment:

Clarifier	Multimedia filtration
Demineralization	Feedwater supply
Pumping equipment	

Figure 5.4.3A: Water Treatment/Handling System

5.4.3.1 System Modifications

With the exception of minor chemical treatment applications associated with the potable water system, there have been no major modifications to this system after the original start-up date.

5.4.3.2 Operating Experience

In mid-October 1998, 13 hours of downtime were due to a piping failure of the main feedwater line from the main power station to the repowering water treatment facility. Investigation revealed that the buried high-pressure feedwater supply line had split and operations at the site had to be suspended until it was repaired. The location of the failure was identified and a piping clamp was installed as a temporary measure to enable the facility to resume operation. The cause of the failure is suspected to be either isolated internal or external corrosion to the carbon steel line. The affected area of piping was repaired or replaced during the next outage.

5.4.3.3 Summary and Conclusions

While the design and operation of this system has been adequate for the water demands of the facility, several minor problems still impact overall continuous operations. Early in the Demonstration Period it was determined that proper clarifier operation provides process feedwater with low alkalinity. While this is ideal for demineralizer feedwater, this low alkalinity water aggressively attacks carbon steel piping; when pH is controlled between 8.0 to 8.2, there is low buffering ability. While chemical treatment has reduced corrosion rates significantly, process and potable water piping going to the gasification process still experiences elevated corrosion rates and piping failures due to the corrosive nature of the water. To resolve this problem, additional chemical treatment and metallurgy considerations are being reviewed and implemented, where applicable.

Each of the two demineralizer trains is designed to produce an effluent with a conductance of less than 0.1 microsiemens when operating at a rate of 480 gallons per minute. Packed bed units of this size proved to be challenging for the manufacturer to produce. The design was based on a scaled-up version of a smaller demineralizer. While simply making some components bigger to achieve performance specification seemed realistic, these units have never realized the economic benefits that packed bed units should provide. Throughput and quality limits are achievable, but at a cost that is high by industry standards. Operations personnel continue to investigate ways to reduce total chemical costs during regeneration of these units while also seeking methods to enhance overall performance.

An unexpected event occurred immediately after start-up of the demineralizer trains when Operations personnel were informed that the original equipment manufacturer was going out of business. This created a situation where engineering support and operational assistance were non-existent and made optimization of the process very difficult and time consuming. Additionally problems have been encountered with spare parts availability and any potential design changes to the system, including system expansion if required, will be negatively impacted by the lack of original equipment manufacturer input.

5.4.4   Steam Turbine

The steam turbine, located in PSI's existing main station powerhouse, was originally supplied by Westinghouse and went into commercial operation in 1953 at a nominal rating of 99 MW. The turbine was designed for reheat operation with five levels of extraction steam used for feedwater heating. In the reconfigured design, feedwater is heated in both the HRSG and the gasification process, eliminating the need for the extraction steam. Therefore, the steam previously extracted now passes through the steam turbine to generate 105 MW of power. Minor modifications to the original turbine steam path were made to ensure acceptable steam path velocities. All other equipment remains as original. There was no downtime associated with the steam turbine during the Operating Period.

5.4.4.1   System Modifications

There have been no modifications to the reconfigured design of the steam turbine package prior to, or after, the Operating Period.

5.4.4.2   Operating Experience

The steam turbine and its subsystems operated within design specifications during the Operating Period. The total downtime associated with the steam turbine during the Operating Period was 9 hours. In October of 1998, failure of several tubes in the steam ejector condenser delayed start-up of the facility for 9 hours. The Tubes were repaired and no further problems have been noted.

5.4.4.3   Summary and Conclusions

The operation of the steam turbine has been uneventful during the Demonstration Period with only minor preventative maintenance performed on the unit. The only event that led to downtime in the four-year operation was due to a warm-up vent valve packing failure, which occurred in January 1996. The vent valve was replaced with an identical valve and the turbine returned to service. Additionally, Operations continued to deal with repeated fouling and leak repair of the steam ejector condenser exchanger tubes due to corrosion and corrosion by-products depositing on the tube walls. Preventative maintenance and appropriately timed flushing of the exchanger have reduced the frequency of fouling and tube replacement.

6.0   ENVIRONMENTAL PERFORMANCE

Project compliance monitoring includes all monitoring that is required by Federal, State and local governmental agencies to satisfy statutes, regulations and terms of leases, permits, grants and other requirements. The permits dictate the monitoring and documentation requirements and reports to regulating agencies. In addition, the Project concluded an Environmental Monitoring Plan (EMP) as required by the Cooperative Agreement.   Through the permit and EMP requirement, monitoring of the environmental performance of the Project has been well documented in the reporting requirements of these programs and will be summarized here for the Demonstration Period. The basis for the information contained herein, will be the sampling and analysis program outlined under the EMP and will include information on the non-proprietary process streams and environmental streams of concern. For several of the streams, only years 1997 and 1998 of the Demonstration Period were reported here as the years 1996 and 1999 both had limited operation.   Under the requirements of the EMP, Table 6.0A lists the streams monitored under the Cooperative Agreement while Figure 6.0A illustrates their location in the process flow map.

Table 6.0A: Key to Monitoring Locations

Location Designator	Proprietary/ Non-Proprietary Status	Description of Monitoring Location
1	NP	Coal Slurry
2	P	Raw Syngas
3	P	Sour Syngas
4	P	Sour Water
5	P	Acid Gas
6	P	Tail Gas
7	NP	Tail Gas Incinerator Stack Gas
8	NP	Sweet Syngas
9	NP	GT/HRSG Stack Gas
10	NP	Slag
11	NP	Sulfur
12	NP	Non-Contact Cooling Water (Outfall 001)
13	NP	Process Waste Water (Gasification Plant)
14	NP	Treatment Pond Discharge to Ash Pond (Internal Outfall 102)
15	NP	Ash Pond Effluent (Outfall 002)
16	NP	Equipment Leak Fugitive Emissions
17	NP	Slurry Facility Fugitive Emissions
18	NP	Slag Handling Fugitive Emissions
19	NP	Coal Handling Fugitive Emissions
20	NP	Slag Transport and Storage Fugitive Emissions
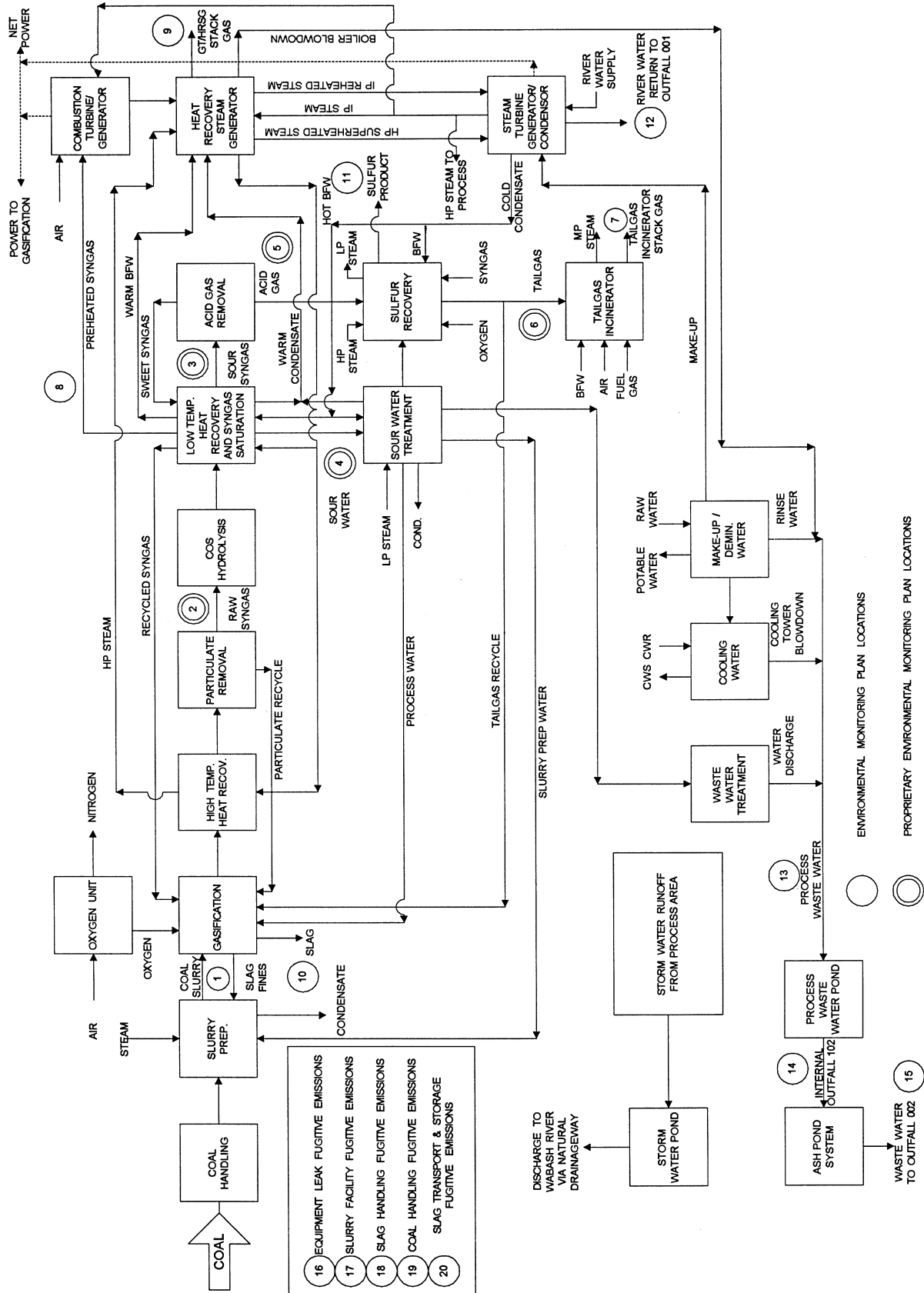
Figure 6.0A: Monitoring Locations

6.1   Non-Proprietary Streams

Analytical process information of the non-proprietary streams as designated under the EMP was compiled by on-site laboratory personnel and by independent laboratories to fulfill the requirements under the EMP.

Stream 1: Coal Slurry to the Gasifier

A series of coal slurry samples were analyzed for determination of Group I and Group II Metals analysis at the beginning of each operational campaign on the coal, blunted coal, and petroleum coke feed stocks (approximately 50 samples per year). Table 6.1A illustrates the average values for the years of 1997 and 1998:

Table 6.1A: Coal Slurry Analysis

	Units	1997 Average Analysis	1998 Average Analysis
Group I Metals			
Antimony	mg/kg	<10.0	0.75
Arsenic	mg/kg	<10.0	1.78
Cadmium	mg/kg	<0.50	0.31
Chromium	mg/kg	4.55	5.55
Cobalt	g/kg	1.67	0.932
Manganese	mg/kg	12.07	12.33
Mercury	mg/kg	<0.080	ND
Nickel	mg/kg	4.97	15.99
Selenium	mg/kg	<10.0	0.08
Group II Metals			
Aluminum	mg/kg	323.33	814
Barium	mg/kg	6.34	5.82
Beryllium	mg/kg	1.18	1.81
Boron	mg/kg	99.20	106
Calcium	mg/kg	1787	1361
Copper	mg/kg	3.26	5.9
Iron	mg/kg	5827	12233
Lead	mg/kg	<5.0	17.3

Table 6.1A: Coal Slurry Analysis (Continued)

	Units	1997 Average Analysis	1998 Average Analysis
Group II Metals (Cont.)			
Magnesium	mg/kg	289	227
Molybdenum	mg/kg	<1.0	1.53
Phosphorous	mg/LP	18.67	10.52
Potassium	mg/kg	118	483
Silicon	mg/kg	152	450
Silver	mg/kg	<1.0	0.553
Sodium	mg/kg	394	212
Thallium	mg/kg	<20	1.25
Vanadium	mg/kg	9.57	9.30
Zinc	mg/kg	26.70	49.3
Ultimate Analysis			
Carbon	wt %	69.0	71.12
Hydrogen	wt %	4.72	4.95
Nitrogen	wt %	1.46	1.41
Oxygen	wt %	12.9	8.51
Sulfur	wt %	0.29	2.89
Ash	wt %	11.56	11.12

Stream 7: Tail Gas Incinerator Stack Gas

The tail gas incinerator stack operates under the provisions of State and Federal Prevention of Significant Deterioration (PSD) air permits administered by the Vigo County Air Pollution Control agency. There are separate air permits for gasification plant sources and combined-cycle power plant air emission sources.

Under the guidelines of this permit, pollution standards have been set for various emission points for the facility. Listed in Table 6.1B that follows is an outline of the permit limits for the Tail Gas Incinerator.

Table 6.1B: Tail Gas Incinerator Permit Limits

Process	Pollutant	lb/hr	tons/yr
Tail Gas Incinerator	Sulfur Dioxide	527	944
Tail Gas Incinerator	Carbon Monoxide	56	221
Tail Gas Incinerator	Sulfuric Acid Mist	3.79	6.8

During the second quarter of 1996, stack testing of the tail gas incinerator was conducted to prove the reliability of the monitoring system. Testing was done during a period of relatively high emissions prior to optimization of the facility's COS hydrolysis unit and AGR/SRU systems and is not indicative of longer term average emissions from this source. Table 6.1C illustrates the results of the test work.

Table 6.1C: Initial Compliance Stack Testing

Process	Pollutant	lb/hr	Remarks
Tail Gas Incinerator	Sulfur Dioxide	342.66	Value represents an average of 10 separate 30-minute readings. Monitor proved relative accuracy within 6.89 lb/hr. Values varied from a low of 145 lb/hr to a high of 487 lb/hr. This unit was then qualified under 40 CFR parts 60 and 75 as an acceptable monitor with no bias.
Tail Gas Incinerator	Carbon Monoxide	0.46	Well below permitted value and demonstrates unit in compliance with permit. CO is not continuously monitored.
Tail Gas Incinerator	Sulfuric Acid Mist	2.69	Demonstrates compliance with the permit. H_2SO_4 is not continuously monitored.

Following this initial test work, quarterly "Relative Accuracy" tests were conducted on the stack monitoring equipment to ensure the reliability of the data collected. The tail gas stack is required

to monitor sulfur dioxide (SO$_2$) emissions on a continuous basis, along with flow and temperature monitoring, to ensure compliance with permitted limits.

Based on the results of this stack testing, and additional testing conducted for the EMP, calculated values for carbon monoxide (CO) and sulfuric acid (H$_2$SO$_4$) mist are reported to the administrative agency on an annual basis. From November of 1997 to December 31, 1999, Wabash River continued to operate within the guidelines of both Federal and State regulations and only exceeded emission limits when equipment malfunctions occurred. Specific information regarding these events is covered, in full detail, in the Environmental Monitoring Plan quarterly and annual reports and will not be reproduced here. It is essential to note that none of these emission exceedances resulted in violations of the air permits because each malfunction was properly addressed under the requirements of the regulations.

In both 1997 and 1998, comprehensive stack testing was done to meet the requirements of the EMP. The results of those tests are detailed in Table 6.1D.

Table 6.1D: 1997 and 1998 Stack Testing Results

Constituent	Units	1997 Testing	1998 Testing
Antimony	lb/hr	<0.000239	<0.00000667
Arsenic	lb/hr	0.000358	0.000224
Cadmium	lb/hr	<0.000119	0.0000551
Cobalt	lb/hr	0.000222	0.00000741
Manganese	lb/hr	<0.0000314	0.000188
Mercury	lb/hr	0.000930	0.000161
Nickel	lb/hr	0.0000971	0.000137
Selenium	lb/hr	0.0000958	<0.00000383
PM$_{10}$	lb/hr	0.28	0.57
Sulfuric Acid	lb/hr	1.98	0.24
Hydrogen Sulfide	lb/hr	<0.00147	<0.00217
Benzene	lb/hr	<0.0125	<0.00541
Toluene	lb/hr	<0.0471	<0.00639
Ethyl Benzene	lb/hr	<0.00742	<0.00736
Total Xylenes	lb/hr	<0.004	<0.00541
Ammonia	lb/hr	<0.081	<0.0031
Cyanide	lb/hr	<0.003	<0.0000606
Phenol	lb/hr	<0.011	<0.0002

Hydrogen	lb/hr	<0.413	<0.283
Methane	lb/hr	<0.367	<0.270
Carbon Monoxide	lb/hr	<0.030	<0.445
Sulfur Dioxide	lb/hr	67.4	14.64
Nitrogen Oxides	lb/hr	8.82	5.33

It should be noted that all values preceded by the < symbol are below the detection limits of the analytical procedure utilized. All analytical procedures utilized for this test work are in compliance with EPA accepted and/or recommended practices for environmental gaseous testing and are either outlined in EPA guidance documents or standard ASTM methodology.

Annual emission rates in tons/year were also provided to Vigo County Air Pollution Control and the Indiana Department of Environmental Management on a yearly basis as totalized from the continuous emissions monitor on the tail gas incinerator. Table 6.1E outlines the tons/year of permitted pollutants reported to these agencies for 1997 and 1998. The years of 1995, 1996 and 1999 were well below the permitted limits for the facility due to reduced operation and are not presented here, but can be found in the EMP reports for those years. Nitrogen oxides, PM_{10}, and volatile organic compounds will be referenced later in this section.

Table 6.1E: Annual Emission Inventory – Tail Gas Incinerator Stack (Tons/Year)

Constituent	1997 Total Emissions	1998 Total Emissions
Sulfur Dioxide	139.77	152.49
Carbon Monoxide	0.029	0.588
Sulfuric Acid Mist	3.84	0.633
Nitrogen Oxides	17.13	14.07
PM_{10}	3.85	1.51
Volatile Organics	0.0019	0.0021

Stream 8: Sweet Syngas

During the Demonstration Period, the quality of the product (sweet) syngas delivered to the combustion turbine was relatively consistent despite differing coal/fuel feeds. Table 6.1F illustrates the average values for the primary constituents in the sweet syngas. Variations in the quantity of sulfur compounds in the syngas are directly related to the operation of the COS

hydrolysis unit and the AGR/SRU systems. Sections 4.0 and 5.0 provide more insight into the direct cause of these annualized variances as seen in the table.

Table 6.1F: Sweet Syngas Quality

Constituent	1996	1997	1998	1999
Hydrogen, vol %	33.6	33.65	33.35	32.66
Carbon Dioxide, vol %	15.83	16.75	15.62	15.75
Carbon Monoxide, vol %	44.44	44.45	45.54	45.52
Methane, vol %	1.82	1.92	2.06	1.99
Hydrogen Sulfide, ppmv	39.39	42.48	75.24	95.98
Carbonyl Sulfide, ppmv	64.89	37.34	26.78	15.92

The sweet syngas product can also be directed to the process flare. Under the guidelines of the air permit for the flare source, pollution standards have been set and are listed in Table 6.1G below.

Table 6.1G: Flare Permit Limits

Process	Pollutant	lb/hr	tons/yr
Flare	Carbon Monoxide	11,099	1,401

<u>Stream 9: Gas Turbine and Heat Recovery Steam Generator</u>

The combustion turbine stack operates under the provisions of State and Federal Prevention of Significant Deterioration (PSD) air permits administered by the Vigo County Air Pollution Control agency. Table 6.1H illustrates the emission limits in the permit for the combustion turbine stack.

Table 6.1H: Combustion Turbine Permit Limits

Process	Pollutant	Requirement on Syngas	Requirement on Oil	Comments
Gas Turbine Stack	Nitrogen Oxides	25 ppmdv	65 ppmdv	Corrected to 15% oxygen and dry conditions.
Gas Turbine Stack	Sulfur Content of Fuel	$\leq$ 360 ppmdv	$\leq$ 0.05% by weight	Satisfies the requirements of 40 CFR 60.333
Gas Turbine Stack	Opacity	20% maximum	60% maximum	Fuel oil limit is based on a time limit of 10 minutes per 24 hours allotted for start-up.
Gas Turbine Stack	Carbon Monoxide	15 ppm	25 ppm	Values corrected to 15% oxygen and 75% or greater load.
Gas Turbine Stack	Sulfuric Acid Mist	0.01 lb/MMBtu		Controlled by limiting fuel gas to less than or equal to 360 ppmdv of sulfur and ensuring that exhaust stack temperature is maintained at or above 264ºF.

Included in PSI's operational permit for the power block are limits on their combustion turbine and auxiliary boiler, not included in this report, as part of the overall permit for the main generating station. Compliance testing for these units was conducted in mid-year 1996 and both units met emission requirements as outlined in the operating permit. Some minor difficulties were initially encountered with the data logger system for the emission points, but were rectified early in the Demonstration Period with the installation of a more compatible software program.

Emissions, reported on an annualized basis in the form of an Annual Emission Inventory Report to the Indiana Department of Environmental Management (IDEM), are listed in Table 6.1I on a tons/year basis for the years of 1997 and 1998. These figures represent total emissions from the Power Block to include stack emissions and total fugitive emissions based on EPA publication AP-42 "Compilation of Air Pollutant Emission Factors" calculated methods and from direct emission monitoring equipment. The quantity of total annual stack emissions is directly related to the coal sulfur content, the efficiency of the gasification plant sulfur removal and recovery systems, and the quantity of syngas and fuel oil burned in the combustion turbine during the year.

Table 6.1I: Power Block Emissions (Tons/Year)

Constituent	1997	1998
Sulfur Dioxide	666.99	872.34
Nitrogen Oxides	76.87	107.96
Carbon Monoxide	33.67	29.68
PM_{10}	41.26	55.98
Volatile Organics	8.83	3.13

Stream 10: Slag

Vitrified frit, also known as slag, is a by-product of coal gasification and represents the non-combustible portion of the coal leaving the gasifier. Table 6.1J represents the analytical data results for this stream for the operational years of 1997 and 1998. The slag has been determined through laboratory analysis (TCLP (total), which is representative of the total value of all leachable material in the slag) to be non-leachable and is classified as a non-hazardous by-product, consistent with slag from many different feedstocks (lignite, sub-bituminous coal, bituminous coal, petroleum coke) processed through the E-Gas^{TM} gasification process during the past three decades.

Table 6.1J: Slag Analysis

Constituent	Units	1997 Average Analysis	1998 Average Analysis
Carbon Content	wt %	15.67	7.91
Moisture Content	wt %	30.17	28.04
Group I Metals			
Antimony	mg/kg	<10.0	4.8
Arsenic	mg/kg	<10.0	6.39
Cadmium	mg/kg	<0.50	0.76
Chromium	mg/kg	29.6	120
Cobalt	g/kg	5.35	1.017
Manganese	mg/kg	33.3	67
Mercury	mg/kg	<0.08	<0.01
Nickel	mg/kg	22.9	146.7
Selenium	mg/kg	<10.0	3.02

Table 6.1J: Slag Analysis (Continued)

Constituent	Units	1997 Average Analysis	1998 Average Analysis
Group II Metals			
Aluminum	mg/kg	8527	49600
Barium	mg/kg	37.9	80
Beryllium	mg/kg	2.03	10.9
Boron	mg/kg	128	283
Calcium	mg/kg	9253	3246
Copper	mg/kg	12	54.1
Iron	mg/kg	17267	105333
Lead	mg/kg	8.19	97
Magnesium	mg/kg	1383	2783
Molybdenum	mg/kg	3.30	8.63
Phosphorous	mg/LP*	<39	2.0
Potassium	mg/kg	1687	7623
Silicon	mg/kg	1207	2000
Silver	mg/kg	<1.0	3.0
Sodium	mg/kg	743	1037
Thallium	mg/kg	<0.4	16
Vanadium	mg/kg	25.10	156
Zinc	mg/kg	32.97	213
TCLP (total)	mg/l	<0.682**	<0.12**

*mg/LP=milligrams/Liter as Phosphate
**These values represent the total of all leachates present in the analyzed sample. Barium
  was the only constituent tested that presented leachable characteristics in the slag samples.

Stream 11: Sulfur

Sulfur quality has remained very consistent throughout the entire Demonstration Period. The
purity of the sulfur for the Demonstration Period was consistently above 99.999% on every
quarterly analysis submitted to the purchaser of the by-product.

Stream 12: Non-Contact Cooling Water

No analysis was required or conducted for this stream. There were no reported problems
associated with this stream from either an environmental or operational standpoint.

Streams 13 and 14: Process Waste Water (Outfall 102)

Process waste water includes all waste water streams generated or captured during normal operations and equipment purges/wash-downs during maintenance activities. Stream 13 is specific to the internal monitoring of the gasificiation process while stream 14 represents the total waste water streams from the entire Repowering Project. The combined streams include, but are not limited to: cooling tower blowdown; gasification plant process waste water; regeneration waste water from the demineralizer system in the power block; rainwater collected in the process blocks for both gasification and the power block; equipment purges (blowdowns) and water wash-downs during maintenance preparation procedures; and un-recycled condensed water from the process.

Due to the inconsistencies of plant operation immediately after start-up and through 1996, IDEM agreed to suspend Outfall 102 monitoring and reporting requirements until April of 1997. Outfall 102 is the NPDES internal pond that was added to the Wabash River Generating Station water discharge permit for the sole purpose of monitoring the waste water discharge from the Repowering Project. Table 6.1K lists the permitted constituents of the process stream along with the set limitations for discharge.

Table 6.1K: Process Waste Water Permit Limits

Parameter	Units	Quality or Concentration	
		Monthly Average	Daily Maximum
Flow	mg/l	Report	Report
Ammonia, (as Nitrogen)	mg/l	27.14	54.29
Arsenic	mg/l	0.018	0.043
Cadmium	mg/l	0.010	0.025
Chromium	mg/l	3.47	8.07
Hexavalent Chromium	mg/l	0.014	0.032
Copper	mg/l	0.040	0.093
Cyanide	mg/l	0.019	0.044
Lead	mg/l	0.260	0.606
Mercury	mg/l	0.0005	0.001
Nickel	mg/l	2.91	6.78
Selenium	mg/l	0.017	0.040
Zinc	mg/l	0.241	0.560
pH		6.0 to 9.0	6.0 to 9.0

Table 6.1L illustrates the monitoring results for each of the years indicated based on an annual average of monthly averages. The annual average for 1997 represents only the months of April through December. Daily maximum values can be reviewed in the Annual Environmental Monitoring Report for each of the years indicated.

Table 6.1L: Process Waste Water Discharge

Parameter/Constituent	Units	Monthly Average	Daily Maximum	1997 Monthly Average	1998 Monthly Average	1999 Monthly Average
Ammonia, (as Nitrogen)	mg/l	27.14	54.29	3.93	6.56	8.8
Arsenic	mg/l	0.018	0.043	0.0077	0.0199	<0.01
Cadmium	mg/l	0.010	0.025	<0.0038	<0.008	<0.01
Chromium	mg/l	3.47	8.07	<0.006	<0.0108	<0.0167
Hexavalent Chromium	mg/l	0.014	0.032	<0.01	<0.0120	<0.01
Copper	mg/l	0.040	0.093	<0.01	<0.0145	0.0185
Cyanide	mg/l	0.019	0.044	0.107	0.2798	0.1438
Lead	mg/l	0.260	0.606	<0.08	<0.08	<0.08
Mercury	mg/l	0.0005	0.001	<0.0005	<0.0005	<0.0006
Nickel	mg/l	2.91	6.78	<0.02	<0.0236	<0.1140
Selenium	mg/l	0.017	0.040	0.0714	0.230	0.1380
Zinc	mg/l	0.241	0.560	0.05	0.0414	0.1363
pH		6.0 to 9.0	6.0 to 9.0	7.99	8.4	7.5

Elevated levels of selenium, cyanide and arsenic in the waste water have caused the process waste water to be out of permit compliance. Daily maximum values, though not indicated in the table above, were routinely exceeded for selenium and cyanide, and only occasionally for arsenic. The Project participants have been working closely with IDEM on remediation plans to bring the Project waste water in compliance.

Trace elements from the coal feedstock are volatilized in the gasification process, and later condensed from the syngas during low temperature heat recovery. Further processing of this sour condensate to remove dissolved gases is accomplished by steam-stripping; however, trace elements remain in the sweet processed water stream and must be removed via other means if levels are above the stringent waste water permit limits.

Removal of certain trace elements, such as selenium, arsenic and cyanide can be effectively accomplished through the use of evaporation systems; however, these systems add cost and complexity to the gasification facility. Continued development of alternate methods to remove trace constituents, such as chemical precipitation, bio-remediation, or reverse osmosis may offer more cost-effective ways to bring these waste water streams into compliance and are currently under investigation. However, laboratory tests have demonstrated that the bio-remediation option will probably require pondage area far beyond the space available at the project site (requires over 50 acres of vegetation). Laboratory tests and field trials of chemical precipitation methods have also proved unsatisfactory, in that the removal efficiency of the system seems to vary in certain plant operating modes, and the maximum removal efficiency achieved to date (about 90% for selenium) is less than that needed to achieve compliance. The remaining two methodologies, reverse osmosis (with drum dryer) and evaporation are still under study. Field trials on reverse osmosis systems are currently being conducted, and design development of an evaporation system has begun. Installation of either of these systems is expected to resolve all wastewater effluent compliance issues. However, the evaporation system, a more conventional approach, is expected to be significantly more capital intensive (approximately $5 million versus $3.5 million for the reverse osmosis – dryer system) and is also known to have a significantly higher operating cost. These evaluations are still under way at the time of issue of this report, with determination of the technology to be complete by October 2000. The Project and IDEM have agreed to proceed with a technology selection and system installation schedule to achieve permit compliance by September of 2001.

Stream 15: Ash Pond Effluent (Outfall 002)

Fly ash and bottom ash from the existing Wabash River Generating Station are conveyed to an ash pond system by high-pressure service water. A channel is maintained by dredging the first pond to facilitate waste water flow to the subsequent ponds and ultimately for discharge to the Wabash River via permitted Outfall 002. Since the ash is managed in the NPDES permitted impoundment system, a solid waste permit is not required for ash management practices.

Table 6.1M: Ash Pond Effluent (Outfall 002) Permit Limits

Parameter	Quality or Concentration	
	Monthly Average	Daily Maximum
Flow	Report	Report
Total Suspended Solids	30 mg/l	70 mg/l
Oil and Grease	15 mg/l	20 mg/l
Ammonia (as Nitrogen)	Report	Report
Arsenic	Report	Report
Cadmium	Report	Report
Hexavalent Chromium	Report	Report
Chromium	Report	Report
Copper	0.04 mg/l	0.09 mg/l
Cyanide	Report	Report
Iron	Report	1.0 mg/l
Lead	Report	Report
Mercury	Report	Report
Nickel	Report	Report
Selenium	Report	Report
Zinc	Report	Report
pH	6.0 to 9.0	6.0 to 9.0

Ash pond discharge has remained within permitted limits for the Demonstration Period. Refer to Table 6.1M for permitted parameters and limits.

Stream 16: Equipment Leak Fugitive Emissions

As part of the gasification facility's air permit, equipment leak fugitive emissions are monitored on a continuous basis with the aid of 39 continuous area monitors that measure low levels of carbon monoxide (CO) and hydrogen sulfide (H_2S) throughout the plant. The units have low alarm settings of 30 ppm for CO and 9 ppm for H_2S. These units also have high alarm settings of 60 ppm for CO and 18 ppm for H_2S. While included in the monitoring program for calculating total CO air emissions, these units are also utilized to help protect workers in the locations from encountering harmful levels of these toxic air contaminants.

Total CO values are calculated from these readings to provide information to the IDEM on total fugitive emissions from the plant process. As an integral part of the emission reporting process, these levels are reported on an annual basis in tons per year and added to the total emissions inventory.

Fugitive emissions reported in tons per year to the IDEM for the two years of 1997 and 1998 are indicated in Table 6.1N and represent typical values during normal operation for all years of the Demonstration Period. While there are no specific permitted limits for these compounds outlined in the facility's air permit, fugitive emissions must be quantified and reported on an annualized basis to IDEM as part of the Title V permit reporting process.

Table 6.1N: Fugitive Emission – Tons/Year

Constituent	1997	1998
Benzene	0.0058	0.000017
Carbon Monoxide	1.0275	0.0030
Hydrogen Sulfide	0.158	0.000425
Ammonia	0.0374	0.00011
Total Sulfurs	0.158	0.00058
Volatile Organics	0.68	0.002038

As can be noted, there was a significant reduction of the fugitive emissions reported in 1998 compared to 1997 primarily due to accurate testing that took place at the end of 1997 and the subsequent focused follow-up actions to address the findings. A main contributor to the reduction in reportable emissions was the implementation of improved operational and maintenance procedures directed at initiating immediate repairs to any leaking valves and equipment.

Stream 17 through 20: Fugitive Emissions for the Coal, Slag, Slurry and Transport

These sites focus on fugitive dust emissions from the facility for both industrial hygiene concerns and pollution from particulates. Independent testing of these process areas reveals that, as would be expected from a wet slurry process, fugitive dust emissions are of little concern in this process. Particulate concentrations of 39 to 99 microgram per cubic meter are common in most processing and handling areas. The highest concentration of particulate, as one would expect, is upstream of the beginning of the wet slurry process in the rod mill coal feed hopper building and has been measured with a level of 224 micrograms per cubic meter. PM_{10} concentrations ranged from 22.9 to 165 micrograms per cubic meter in all areas sampled. Both total particulate and

PM$_{10}$ concentrations were well below the American Conference for Governmental Industrial Hygiene (ACGIH) threshold limit value (TLV) for coal dust, which is 2,000 micrograms per cubic meter over an 8-hour period. Triplicate visible emission tests have also been completed for each area by a certified smoke reader who indicated that there were no visible emissions detected during any of the tests. Visible emission limits have been set in our Operating Air Permit for the facility to be maintained at or below 20% opacity. As indicated by these tests the plant was well within limits of its permit.

6.2   Conclusions

6.2.1   Process Waste and Waste Water

As can be seen by the information presented in this section, the components that have contributed to the environmental performance issues of this facility stemmed from volatile components in the coal feedstock. Sulfur, arsenic, selenium and cyanide, among others, either fully or partially partition into the gas phase. Arsenic, selenium and cyanide, typically end up in the condensed vapor stream and preferentially leave the process via the process waste water stream. At the end of the Demonstration Period, the participants had a plan well underway to bring the waste water into permit compliance for these constituents.

Also, due to the presence of arsenic in some of the deposits in the process, additional industrial hygiene programs were developed to ensure worker safety while doing maintenance in localized areas of the process.

Ensuring procedures for safe, cost effective, and environmentally sound options for the recycle, treatment, or disposal of all excavation/demolition materials and process waste was a top priority during the development of the waste handling program. Of particular importance was ensuring that each waste stream, whether hazardous or non-hazardous, was managed according to all State and Federal regulations. The programs established at Wabash River have been recognized by the State of Indiana government officials as models for other companies against which to benchmark.   Additionally, Wabash River operated a comprehensive recycling program throughout the Demonstration Period and is continuing that effort today. For its efforts, the

Wabash River Coal Gasification Plant was named the 1998 recipient of the state of Indiana Governor's Award for Excellence in Recycling. Over 358 tons of materials were recycled and eliminated from being land filled during the course of the Demonstration Period.

6.2.2 Air Emissions

The Wabash River Coal Gasification Repowering Project was designed to exceed the requirements of the 1990 Clean Air Act Amendments for 2000 while at the same time offering an alternative means to generate power from mid-to-high sulfur-containing coals. Based on the emissions reported to the IDEM, the Project has met or exceeded every expectation outlined in the pre-construction literature. Table 6.2.2A represents total emissions, in pounds emitted per MMBtu of dry coal feed to the gasifier, based on all sources monitored or calculated at the Wabash River Coal Gasification Repowering Project during the years of 1997 and 1998. It should be noted that sulfur dioxide emissions from the Wabash River site have demonstrated nearly an order of magnitude reduction from the 1990 Clean Air Act Amendment year 2000 target of 1.2 lb/MMBtu of feed.

Table 6.2.2A: Component Emissions in Pounds per MMBtu of Dry Coal Feed

	1997	1998
Sulfur Dioxide	0.13	0.13
Nitrogen Oxides	0.024	0.021
Carbon Monoxide	0.056	0.033
Volatile Organics	0.002	0.0021
PM$_{10}$	0.012	0.011

7.0   ECONOMICS

7.1   Actual Installed Costs for Wabash River

The initial budgeted cost for the construction of the Wabash River facility was $248 million for the syngas facility (Destec scope) and approximately $122 million for the new power block and modifications to the existing Wabash River Generating Station (PSI Energy scope). The power block budget also included the initial infrastructure work for the site, including leveling the hill where the Project was sited, remediation of the abandoned mineshaft on the site, all roads, and all underground drainage and fire-protection lines.

Both Destec and PSI acted as general contractors for their individual scopes of work. Detailed design engineering and procurement was subcontracted (Dow Engineering for the syngas facility and Sargent & Lundy for the power block). In the field, Destec and PSI each acted as construction manager, hiring local and national contractors to perform specific areas of the work (excavation, foundation, structural steel, piping, electrical and instrumentation, refractory, etc.). Additionally, Destec subcontracted the air separation unit as a turnkey contract from Liquid Air Engineering Corp. (Air Liquide).

Table 7.1A provides the total costs for the power block and for the syngas facility. The syngas facility is additionally broken down functionally. These costs are equivalent to what is commonly referred to as the EPC (engineering–procurement-construction) cost. This cost includes conceptual, process and detailed engineering, environmental permitting, equipment procurement, project management, construction management, construction and start-up, and the costs of hiring and training Operations personnel up to the commercial operations date of November 17, 1995. The cost as shown excludes fuel supply and owner's costs such as interest during construction, financing fees and license fees. There were no significant interconnection costs for water or power since the IGCC facility was built at the site of an existing power station.

Table 7.1A: Project Costs

Cost Area	Budget *	Actual *
SYNGAS FACILITY		
Engineering & Project Management	29.6	27.3
Equipment Procurement	98.3	84.5
Construction	55.5	106.1
Construction Management	7.9	8.1
ASU	36.9	32.8
Pre-Operations Management	19.8	21.7
POWER BLOCK	121.8	136.2
Total	**369.8**	**416.6**

*$MM, 1994 Average

These values do not include any offset for the DOE CCT Program contributions to the Project capital cost, but reflect actual expenditures. On an EPC basis, including start-up, the installed cost of the overall IGCC facility was about $1,590/kw (1994$). Allowing for new equipment that would have been required if this had been a greenfield project instead of repowering, the installed cost figure on this Demonstration Project was $1,700/kw (1994$).

It should be noted that one area that contributed to the Project cost, but that is somewhat hidden, is the cost of sulfur removal and recovery beyond the normal requirements. As discussed elsewhere, the Wabash River facility is the cleanest coal-fired power plant of any technology in the world. It was designed to operate at one sixth of the SO_X emissions allowed under the Clean Air Act Amendment rules for the year 2000, and normally runs at one half to one fifth of the design point. These low emissions have been achieved on coal and pet coke supplies with up to 5.5% sulfur content. This design philosophy for the Project, included in the budget numbers, added approximately $100/kw to the installed cost.

As shown in Table 7.1A, nearly all cost areas within the syngas facility were completed under budget or virtually at budget, with the exception of the construction cost and the pre-operations management cost of the syngas facility. Overruns in the power block budget were in the same areas. The construction cost was nearly double the budgeted amount, due to several factors, many of which were beyond the control of the Project participants.

- Weather delays at the outset of the Project: The initial dirt work and site forming on the Project, originally expected to take five months, took nine months due to extreme and constant rainfall through the spring, summer and fall of 1993. This was the year of the 500-year flood of the Mississippi River, as a sign of the extreme rainfall. The constant rain took its toll on the progress of the dirt work and initial job activities. Even into November of 1993, Indiana recorded the wettest November since 1888. The late start of civil work on the Project was further compounded by a severely cold winter, frequented by both snow and ice storms that shutdown the Project, in addition to daily temperatures that often did not rise above zero degrees Fahrenheit. (The lowest recorded temperature at the jobsite during this period was minus 28 degrees Fahrenheit, for instance, with a wind chill of about 50 below zero.) Both the rainfall and the cold winter were far from the norm for central Indiana.

- Equipment shipment problems: In an effort to save both time and money, several equipment systems were shipped as modularized pieces of equipment, where possible within the prevailing union craft rules. These modules were generally designed to be as large as possible, up to 200 tons and 20 feet wide, based on initial traffic studies. After the design phase of this equipment, while it was being fabricated, three states in the traffic pattern changed their superload laws. This resulted in extended permitting times, change of routes and even changes to barge freight in some cases. Delays of up to four months were experienced with the equipment in some plant areas. There were also delays on a variety of other items, including alloy piping which was ordered well in advance of the start the civil construction, but nonetheless impacted the Project schedule after being delayed.

- Mechanical Contracting: Both Destec and PSI experienced contractual, schedule and budget problems with their respective mechanical contractors. Destec's problems were severe, complicated by an initial piping estimate that was insufficient. When Destec sought to accelerate construction to regain a portion of the Project schedule lost due to the weather and equipment delays, labor shortages in the summer and fall of 1994 compounded the problem. At one point, Destec was forced to change mechanical contractors, losing about a month of schedule, which was subsequently regained by

construction overtime. The overtime costs contributed significantly to the construction cost overrun.

- Start-up: During start-up, there were significant problems with some of the equipment related to the novel technology employed on the Project, in particular the dry char filtration system (as discussed elsewhere in this report). These problems led to a prolonged start-up period, during which the fixed costs for both the site contractors and Destec's own Operations personnel accrued to the construction budget, as did the costs for equipment modifications, dry char filter replacement, and other non-warranty equipment related costs. These costs are estimated at $15 million, or approximately 6% of the gasification facility cost (excluding the ASU and power block).

Having noted the major areas of cost overruns on the Wabash River Project design and construction, it is only fair to note that it was also by far the least expensive of the first generation coal gasification combined cycle plants that were built in the 1992 to 2000 timeframe. The other coal IGCC's, two in the U.S. and two in Europe, all first generation at this scale, have been reported to have cost $2,000/kw and over.

An additional $14 million was spent in the first two years of operation to improve plant availability, especially with respect to the operation of certain components of the novel technology incorporated into the facility. The primary areas that this improvement program focused on were the addition of a chloride scrubber, a change in the geometry of the syngas piping exiting the second stage of the gasifier and numerous changes in the dry char filtration system.

7.2 Forecast Costs for Year 2000 Installation

The Wabash River Coal Gasification Repowering Project was constructed in the 1993 to 1995 timeframe, at an equivalent EPC cost of about $417 million in 1994 dollars ($1,590/kw), as discussed in detail in the preceding section.

As part of another study being conducted with Nexant LLC (a Bechtel technology and consulting company), the Wabash River facility costs were examined to determine the cost of the same facility being built today in the same location. All cost components were escalated, reflecting differential rates of escalation for different commodities. However, adjustments were made to remove the costs of construction acceleration that caused some of the construction budget overruns above. The result of this evaluation, which is the estimated cost of an identical plant with no benefit from what was actually learned at Wabash River, was a greenfield project with an estimated cost of about $390 million (2000$) or about $1,444/kw.

A more realistic value for a current generation plant, that does incorporate lessons learned and improvements to the technology, would be $1,250 - $1,300/kw (2000$) with a heat rate of approximately 8,250 Btu/kwh (HHV) for a greenfield coal gasification facility (inclusive of costs for the new steam turbine and other equipment that were available at the Wabash River Station). A new, stand-alone greenfield IGCC to produce power, but no other products, and utilizing petroleum coke as fuel would have an approximate installed EPC cost of $1,100 - $1,200/kw (2000$), based on reduced equipment requirements with petroleum coke feeds. An approximate breakdown of these costs is shown in the Table 7.2A.

Table 7.2A: Costs of Near Term IGCC Projects, $/kW

Cost Area	Coal *	Petcoke *
SYNGAS FACILITY	720	607
AIR SEPARATION UNIT (ASU)	148	137
POWER BLOCK	407	406
Total	1,275	1,150

$/kW (2000$)

7.3   Operational Costs

7.3.1   Fuel Cost

The Wabash River Coal Gasification Repowering Project operates at a nominal heat rate of 8,910 Btu/kWh.   Fuel consumption for the nominal 262 MWe facility is, therefore, 2,334 MMBtu/hr. The Project can run on medium to high sulfur Midwestern bituminous coals, which typically cost in the range of $1.00 - $1.25 per MMBtu delivered to the West Terre Haute plant site ($22 - $27 per ton). At an annual availability of 75%, total fuel cost while running on this coal feedstock would be in the range of $15.3 - $19.2 million.

7.3.2 Non-Fuel Operations and Maintenance Costs

When constructed, the Wabash River syngas facility was over 800 miles from the nearest Destec-operated facility.   The only other syngas facility in operation by Destec was in the Louisiana Division of Dow Chemical (Louisiana Gasification Technology, Inc., or LGTI); all other Destec operated plants were conventional natural gas-fired cogen power plants. The LGTI plant benefited from the substantial infrastructure and support base of the Dow complex for labor and support services. At Wabash River, much of this support base needed to be established, requiring additional staffing and infrastructure to set up this stand-alone location for Destec.

Using 1999 budgeted costs as a basis, the non-fuel operating and maintenance (O&M) cost for the syngas facility (excluding the Power Block) was 6.8 % of installed capital based on a 75% operating rate.   Included in this figure are all operating labor and benefits, technical and administrative support labor from on-site staff as well as technical and managerial support from the corporate office.

Also included in this cost figure is administrative cost (including insurance), routine and major maintenance, chemicals, waste disposal, operating services and supplies, and a sizable budget for plant betterments (about 5% of the total O&M cost).   Projected O&M cost for a mature greenfield IGCC facility (including the Power Block) is 5.2% of installed capital. Additional operating cost definition has been provided to DOE in Appendix C. (Appendix C, the operational cost information, is classified as "Protected Clean Coal Technology Data" under the Wabash

River Cooperative Agreement and is not included in the public issue of the Final Report at this time.)

7.4 Economic Analysis

7.4.1 Historical Perspective

The Wabash River Coal Gasification Repowering Project was originally approved by the Indiana Utility Regulatory Commission in 1993 based on the results of a financial model considering projected avoided costs for other types of generation and projected load growth in what was then a totally regulated electrical utility environment. In a model for the entire PSI-owned electrical grid, these avoided costs were calculated considering factors such as performance and incremental operation costs at other power generation facilities, costs of new facilities to meet expected load growth, fuel and maintenance costs at the syngas facility and repowering project, and revenues from slag and sulfur sales from the gasification facility. Other factors, such as the avoided cost of retrofitting flue gas desulfurization scrubbers to the Wabash River station for compliance with the Clean Air Act, were considered as well. All factors were viewed in the perspective of the impact to the entire grid of electric generation in the state of Indiana.

7.4.2 Evaluation of Future Power Generation Projects

Prospective new plants are evaluated in a much different way, of course. Most markets are fully or partially deregulated, and power is marketed across the state or across the country to meet the requirements of peak demands. In the current situation, utilities are not proposing to build new generation that would be rate based, because this structure has or will soon disappear under deregulation. Even in states that have not yet deregulated, federal or state mandates for such a structural change are expected in the next five years – and utilities would risk that any new generation would become a stranded asset before the capital costs were retired via electricity rate-based capital cost recovery. A greenfield, base load IGCC facility, owned and operated by an independent power producer rather than a regulated electric utility, must be evaluated against a market that is increasingly dominated by new natural gas-fired peaking plants and older, fully depreciated, coal-fired utility plants that have been restructured as wholesale generators. Most

new generation capacity additions must stand alone on their own merits in a market place without guaranteed long-term power purchase agreements.

The following economic analysis was made utilizing the projected year 2000 capital and operating costs discussed previously to determine the potential for coal and pet coke based IGCC plants in the current environment. The IGCC model developed by Nexant LLC for the DOE was used to evaluate the rate of return for IGCC facilities utilizing today's fuel and power prices with conventional project financing methods. Expansion of the IGCC into coproduction plants, capable of producing revenue streams other than electric power, were not evaluated. No advancements in the IGCC technology, as are expected from the advanced gas turbine programs already in place, were included in this analysis.

Inputs to the economic model:

- Fuel cost: Coal $ 20/ton   with   1.0% annual escalation

  Coke $ 5/ton   with   0% annual escalation
- Power: $ 35/MWh   with   3.0% annual escalation
- Discount Rate: 10%
- Financing: 70% Debt, 30% Equity; 70% Senior Debt at 10% interest on 15 year term with 2% financing fee, 30% Subordinated Debt at 10% interest on 15 year term with 3% financing fee, no debt reserve fund
- Additional Capital Costs (% of EPC costs): Owner's Contingency 5%, Development Fee, Owner's Costs and Licenses 4%, Start-up 1%
- Construction Depreciation 7 years, Financing Depreciation 12 years
- Initial Working Capital: 3% of first year revenue
- Three year construction period, excluding start-up. Start-up ramping considered for revenues
- Plant Economic Life: 20 years from January 1, 2002 Start-up
- Slag $1/ton, Sulfur $25/ton, Escalation of Slag and Sulfur 3%
- Coal Heating Value 11,400 Btu/lb HHV (25,080 Btu/kg), 3% Sulfur, 11% ash
- Petcoke Heating Value 14,800 Btu/lb (32,560 Btu/kg), 6% Sulfur, 0.5% ash
- Escalation of O&M costs 2%
- Income Tax Rate 40%

Net generation, heat rate, capital and operating costs differ for the coal and petroleum coke cases reviewed, as noted in Table 7.4A.

The heat rate values utilized in the analysis reflect results from coal and petroleum coke operation at the Wabash River plant. Operation in near-future plants with Global Energy's advanced gasifier design and/or advanced gas turbine models is expected to be at a heat rate of 8,000 – 8,300 Btu/kWh (HHV).

Likewise, the availabilities used reflect that achieved with the Wabash River design on coal and projected for long-term operation on petroleum coke, but are not indicative of future plants incorporating planned improvements. Plant availabilities in the 90-95% range can be achieved by installation of multiple smaller units, with some incremental increase to the capital and operating costs.

Lower capital and operating costs and a higher availability are used for the base case for petroleum coke operation. Operation on 100% petroleum coke at Wabash River in 1997 and 1999 has indicated that certain pieces of equipment can be reduced in size or eliminated in a plant designed for petroleum coke, as opposed to coal. Operating costs and availability both benefit from petroleum coke operation as well, due primarily to the improved consistency of the feed (as opposed to blended coal or alternating coal supply trains from different mines, as experienced at Wabash River).

Table 7.4A: Results of Economic Analysis for Wabash River Style IGCC

Single Train

	Coal	Petroleum Coke
Plant Net Generation, MWe	270	271
Plant Heat Rate, Btu/kWh (HHV)	8,910	8,790
Plant EPC Capital Cost, $/kw	1,275	1,150
Plant Operating Cost, % of capital	5.2	4.5
Annual Availability	75%	80%
NPV_{10}, Millions $	(59)	79
Internal Rate of Return, at $35/MWh for power	5%	17%
Sensitivity Analysis Cases, 12% IRR	Required Power Pricing in first year $/MWh	
Base Case	42	30
10% reduction in capital	38	27
10% reduction in O&M	41	30
10% increase in availability	38	28
10% reduction in capital, O&M 10% increase in availability	34	24
Minemouth ($10/ton fuel)	39	NA

As evidenced in Table 7.4A, the strongest drivers of overall plant economics are fuel type and capital cost. The economic analyses of project returns with coal as a feedstock reach a credible economic condition of 12% IRR at power pricing of $34 - $40/MWh, depending on how capital, operating and maintenance costs are set and the availability that is assumed. Plant design and operation based on petroleum coke is economically stronger, due not only to the lower fuel cost, but also the incrementally improved capital and operating costs for a plant designed for petroleum coke initially.

References:

"IGCC Economic and Capital Budgeting Evaluation Model", prepared by Nexant LLC for the U.S. DOE under Contract No. DE-AMOI-98FE64778, May 2000

"Improved Performance of the Destec Gasifier", Dr. David L. Breton, 1999 Gasification Technologies Conference, San Francisco, California, Oct. 18-20, 1999

"Commercial Operation of the Wabash River Gasification Project", Phil Amick, AIChE Spring Meeting 2000, Atlanta, Georgia, March 9, 2000

"Gasification Plant Cost and Performance Optimization", Nexant LLC and Global Energy Inc., DOE Cooperative Agreement No: DE-AC26-98FT40342 (expected to be published 1Q2001)

"IGCC Power Plants – EPRI Design & Cost Studies", Neville Holt, Electric Power Research Institute, 1998 Gasification Technologies Conference, San Francisco, California, October, 1998

SECTION III

Commercialization

and

Recommendations

8.0   COMMERCIALIZATION POTENTIAL AND PLANS

The 1999 World Gasification Survey is a database of all the gasification projects in the world, on all fuels (natural gas, coal, petcoke, biomass) and producing all products (power, hydrogen, heat and chemicals). The output of all of these facilities was presented in equivalent megawatts of power generation. The total of all these projects, dating back to 1952 and including those projected to start-up by 2004, is 30,325 equivalent megawatts. Figure 8.0A shows the cumulative increase in the installed gasification capacity over this period, and indicates a significant increase in activity in just the past decade.

Particularly of note is that the majority of the recent increase in installed gasification capacity is in solid fuel gasification (coal and petroleum coke). Overall, 16,505 MW (equivalent) of the 30,325 MW total equivalent capacity is coal or petroleum coke based (approximately 54% of the total). Of the 16,505 MW of solid fueled capacity, nearly half (43%) is starting up in the ten-year period ending in 2004.
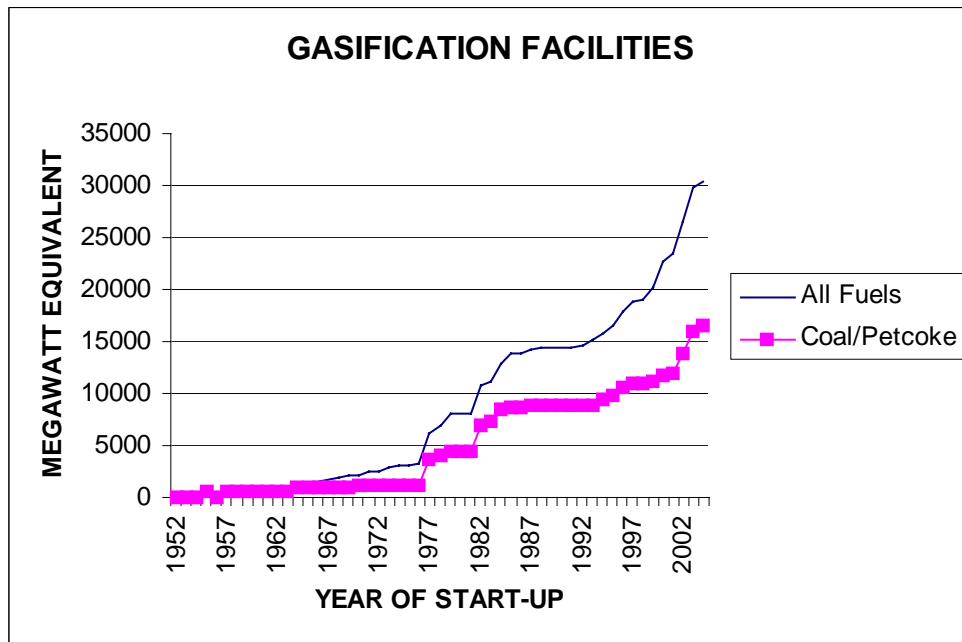
Figure 8.0A: World Gasification Facility Capacity

(from data in the 1999 World Gasification Survey)

The impetus for this growth includes the costs of environmental compliance, the drive to improve efficiencies in refinery settings, the availability of low cost alternative feedstocks and the need to utilize indigenous coals in markets without access to natural gas. The maturation of gasification technologies in large-scale facilities, following completion of several demonstration projects, has made this technology a popular and viable alternative (technology-wise) to the historically conventional combustion technologies.
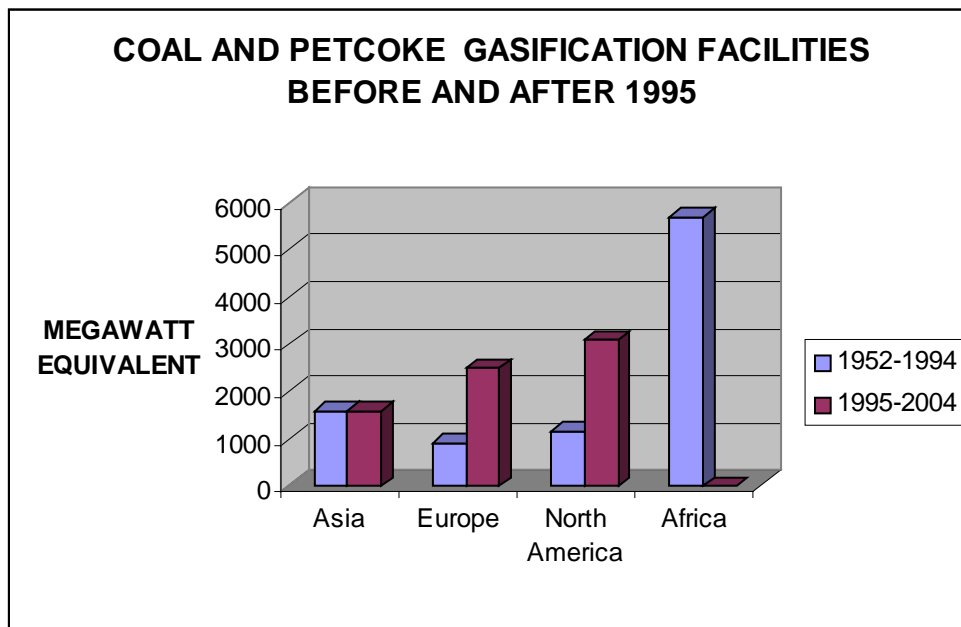
Figure 8.0B: Solid Fueled Gasification Facilities Starting Up Before and After 1995

(from data in the 1999 World Gasification Survey)

Figure 8.0B examines the solid fueled projects geographically and chronologically. Interestingly, the break point chronologically is 1995, the year that Wabash River started commercial operation. This figure also captures the last ten years of data from the survey, 1995-2004, to illustrate the substantive growth. Excluding South Africa, where gasification has been driven by an abundance of coal and lack of almost all other fuels, the solid fueled gasification capability is nearly *tripling* in the last decade of the survey period. The area with the greatest forecast growth, both capacity and percentage wise, is North America, namely the United States.

According to still other industry sources, there are over 1,600 MW of gasification-for-power projects in operation and another 2,700 MW that have started construction or been financed for start-up by the year 2001. Historically, and certainly before the widespread use of combustion turbines, gasification-for-power was much less common than the chemical and fuel gas production applications. This represents better than a tenfold increase over the gasification-for-power market as it existed ten years ago. General Electric (GE) reported in a recent Gasification Technology Council meeting and at the Sixth Annual Clean Coal Conference that there are about 5,000 MW of gasification projects for power generation that have proceeded to the point of placing orders for combustion turbines. Many of these projects include co-production facilities. GE states also that it is in discussions with various refiners, developers and other parties about projects totaling another 50,000 MW (This includes unannounced projects and projects starting up after the 2004 end date in the World Gasification Survey).

Reviewing this market information, it is clear that market trends support the creation of additional opportunities for solid fuel gasification technologies. Trends of rising energy prices and tightening environmental controls are most evident in Europe and North America, and the growth in environmentally superior gasification facilities in these areas is a result of those trends. The single strongest market in either of these areas is the United States. Indications are that many of these domestic projects will be refinery based, utilizing petroleum coke and other low cost refinery by-products to produce power, steam and hydrogen for the refinery and additional power or chemicals for export.

Global Energy has prepared the E-Gas™ Technology for these applications through a petroleum coke operational program at the Wabash River facility, now in its second phase. In the first phase of the program, two trials of full load operation (about 2000 tons per day of pet coke with no coal blending) were completed. All environmental and operational parameters were proven in these trials. The second phase of the program is the long-term operation of the facility on petroleum coke, commencing in August 2000.

The E-Gas™ Technology is also being readied for co-production applications through other study and co-operative evaluation programs, some of which are taking place under other DOE programs. Extensive investigations are also being made into optimizing plant performance and improving the operating life of certain components of the plant.

Global Energy expects that the E-Gas™ Technology, as demonstrated at the Wabash River facility on both coal and petroleum coke feedstocks, will be utilized in a significant portion of new solid fuel based gasification facilities. It should be particularly suited to those refinery based applications utilizing petroleum coke in areas that have been declared as non-compliance areas for one or more air emissions, because of the superior environmental performance of the E-Gas™ Technology.

Somewhat further in the future, as the price of other forms of energy, particularly natural gas, increase domestically and as existing conventional coal-fired power-generating stations strive to meet future environmental regulations, it is also expected that a repowering market will develop in the United States. This market may include both repowering of the coal plants of the 1960's and 70's and repowering of the natural gas-fired combined cycle plants of the 1990's and present day. The E-Gas™ Technology is expected to be well positioned to recreate the Wabash River repowering scenario with improved technology in either of those applications.

References:

Luncheon Keynote speech, Doug Todd, General Electric, Sixth Annual Clean Coal Technology Conference

1999 World Gasification Survey, SFA Pacific Inc., published by the Gasification Technologies Council (www.gasification.org)

9.0   CONCLUSIONS AND RECOMMENDATIONS

9.1   Success of the Demonstration Project

The Wabash River Coal Gasification Repowering Project began as a Joint Venture between PSI Energy and Destec in the summer of 1990. At that time, there was relatively little experience with gasification for power generation (Texaco's Cool Water Plant and LGTI), and none of this experience had been aligned with what was then a regulated utility generation and distribution system. The primary goal of the Wabash River Project was thus a formidable one: commercial operation on the utility grid, matching the utility reliability and availability requirements for baseload operation. The associated technology goals to achieve this were daunting as well:

- Demonstrate Bituminous Coal Operation

- Prove the Repowering of a 40 year old Utility Plant

- Demonstrate Advance Combustion Turbine Application for Syngas

- Demonstrate Superior Environmental Performance

- Prove Particulate Removal System/Reliability

- Establish Availability on the Utility Electrical Grid

Each of the following sections summarizes the challenge, the lessons learned and the outcome in these areas. All are discussed in more detail elsewhere in the report.

Bituminous Coal Operation

Operation at LGTI had been primarily on sub-bituminous coals. The sizing of the sulfur system and constraints on the air permit did not allow demonstration testing at LGTI with the high sulfur bituminous coals envisioned for Wabash River. Operation on bituminous coal would mean changes in the gasifier design, the slag removal system design, operation of the gasifier systems, and unknown impacts in ash deposition control and heat recovery. The high sulfur content of the coal would result in COS levels in the syngas that would have to be removed to meet the stringent emission standards set for the Project. Catalysts for COS conversion had been tested in the laboratory and certain other limited applications, but had never been tested with syngas or at the scale needed at Wabash River.

Through four years of operation, several differences were noted in comparison with sub-bituminous coal operation, particularly in requirements for gas path design and first stage gasifier operation as well as differences in gasifier subsystems. None of these differences are now an impediment to operation and plant run-times have been extended beyond the records set at LGTI.

Ash deposition in the hot gas path, which had never really been solved at LGTI, has been eliminated as a cause of unplanned outages. Initial problems with slag quality have been overcome and this by-product is now being successfully placed in construction applications. The COS catalyst, after some initial problems, is now achieving desired conversion levels and evidencing a substantially longer life than originally expected. Operation of the acid gas removal and sulfur recovery systems has been demonstrated with up to 5.5% sulfur-bearing feedstocks.

Repowering of a 40-Year Old Utility Plant as an IGCC

Even repowering of older steam turbine plants with natural gas-fired gas turbines was fairly rare in the 1990's. Integration with the gasification plant, and the high level of integration on the steam system with the gasification high temperature heat recovery unit, the combustion turbine HRSG and the reheat steam turbine had never been attempted.

Operation of the steam turbine with steam from the gasification and gas turbine plants and the integration of the combustion turbine and gasification boilers has been successful. The Project operates well below its design heat rate. Fuel swaps occur in a matter of minutes. P procedures for starting up and shutting down the combined cycle and swapping to syngas have become routine.

Syngas Application with an Advanced Combustion Turbine

Wabash River utilized one of the first ten General Electric F class machines produced. It was the first GE Frame 7FA to operate on syngas as well. With its 192 MW rating on syngas, the Wabash River combustion turbine was the largest combustion turbine in North America when it was installed. The 7FA turbine has been measured as running comparatively smoother and cooler on syngas than it does on fuel oil operation. The predicted 192 MW output has been

achieved with syngas at a heat rate significantly better than the design guarantee. Emission levels have been met as well, and the plant has operated at as low as one-fortieth of the Clean Air Act Year 2000 standard for sulfur dioxide emissions.

Environmental Performance

Despite its planned design to use up to 5.9% sulfur coal, Wabash River also set out to be one of the cleanest coal-fired power plants in the world. A sulfur dioxide emission level was targeted that was less than one-fifth the emissions allowable under the year 2000 standards of the Clean Air Act. The plant has met all of its emission criteria for SO_x, NO_x, CO and particulate emissions, and it has done this while operating with both coal and high sulfur petroleum coke. The Project is the cleanest coal-fired power plant in the world.

Particulate Removal System

In order to meet the efficiency goals set for the plant, a 20% improvement on the heat rate of the existing unit, the plant would utilize dry barrier filtration to remove char and particulates from the raw syngas stream, rather than a wet quench operation as had been utilized at the LGTI facility. The dry char barrier filtration installation would be only the third in the world at this scale, and with a much higher dust loading and finer char than the other systems.

In the first year of operation, the particulate removal system was the chief cause of facility downtime. Today, it has been virtually eliminated as a cause of unplanned outages and the campaign life of the elements has been extended to a level comparable with the other run-limiting factors.

Availability

In order to function within the utility grid system, incremental increases to an annual availability factor of 75% were planned for the syngas facility. This level of operating availability was a significant increase from the operating record of any single train, solid-fueled gasification facility in the early 1990's.

Despite severe initial problems with some aspects of the novel technology, the syngas facility achieved an average availability of approximately 70% over the last two years of the Demonstration Period. It achieved as high as 77% availability over one nine-month period, and should be able to achieve this level of performance on an annualized basis in the future

The Project proved successful in meeting the challenges of each of these areas of technical advancement (superior environmental performance, particulate removal system application, and commercial availability levels) as demonstrated in four years of operation as part of the Clean Coal Technology Program and documented in this report. This success was not quick and easy, but rather came as a result of a dedicated, persistent effort to meet the challenges posed by these lofty objectives. In the end, all of the goals of the Project were met.

Significant challenges were met and overcome in many other areas outside of the primary demonstration objectives, including technical, commercial and organizational challenges. The Project has also demonstrated success in some areas that were not planned at the outset – operation on petroleum coke, for instance, and operation on a blend and combination of coals that sometimes changes daily. The Project operates today as part of the utility power generation system, competing with Cinergy's alternative market options for on-peak and off-peak power. Competitive market-based pricing allows the syngas facility to run at base-load in Cinergy's system.

9.2   Commercialization Barriers and Areas of Recommended Research

The Wabash River Project has successfully met the challenges set forth for it under the Clean Coal Technology Program. As a result of the experience at Wabash River and other projects like it, in the year 2000 gasification has become established as a viable technology for power generation. It has not historically been as widely accepted as being economically competitive in these applications, however. Gasification remains somewhat of a niche application, currently seeing utilization where feedstock costs are well below the baseline price of coal or where products more valuable than electricity can be produced. Tremendous progress has been made in solid fuel gasification technology applications since 1991 when the Wabash River Project was selected for inclusion in the Clean Coal Technology Program. This is largely due to the initiative of and incentives through the United States Department of Energy in their partnership arrangements with industrial technology developers.   However, major inroads to commercialization of gasification and a significant share of the U.S. electric power generation market are dependent on additional progress in several related areas.

Based on the experiences at the Wabash River Coal Gasification Repowering Project, five areas are suggested for continued research and development to promote gasification technologies.

- Availability Improvement
- Capital Cost Reduction
- Operating Cost Reduction
- Fuel Flexibility
- Product Flexibility

The first three principal areas can be viewed as barriers to widespread market penetration of solid fuel gasification technologies. These areas are the availability, capital cost and non-fuel operating cost of the IGCC facility. The other two bulleted items are "opportunity" areas, namely fuel flexibility and product flexibility, or coproduction, where additional progress will aid the economics of future projects. Obviously, these barriers and concepts have been discussed elsewhere in this report and in the industry in general, and are the basis of many of NETL's current programs, but warrant review here in the context of the Wabash River experience.

Availability

An availability target of over 85% may be needed to successfully compete in the power generation industry against older technology baseload facilities. The refinery industry demands availabilities of "95% plus" for the products it gets from gasification facilities. Meeting these higher availabilities can be achieved with the current technology, although multiple trains at an increased capital cost would currently be required. Exploration into more cost-effective redundancy within the plant and low-cost back-up systems of conventional technologies needs to continue. Global Energy and Nexant LLC are exploring these areas as part of the Gasification Performance and Optimization Study, DOE Contract No. DE-AC26-98FT40342. Also needed are advances in the current campaign and run-limiting process units and equipment in the gasification facility. In addition to Global Energy's internal development of advanced burners and mechanisms and operating disciplines to reduce ash deposition and boiler fouling, Global Energy has been studying materials and designs for better candle filter operation. Wabash River is operating a slipstream unit of char filtration toward this end (DOE Cooperative Agreement No. DE-FC26-97FT34158).

Capital Cost

An integrated gasification combined cycle will always cost more than a natural gas-fired combined cycle plant; the increase in output is not expected to overcome the cost difference between the gasification plant required for IGCC and the mere pipeline construction cost for the natural gas plants. Nonetheless, substantial inroads are being made on reducing capital cost, both by integration of new technologies and through the optimization process of studies like the one mentioned previously.

Capital costs are expected to decrease, as well, due to increases in the capacity and decreases in the cost of the combustion turbines that are the prime movers in IGCC facilities. The advent of even more advanced gas turbines will improve economics both on the capital cost (to $1000/kW) and efficiency parameters (45-50% HHV).

Further work is needed to develop the H class turbines for syngas applications.

Operating Costs

The Wabash River syngas facility reduced its operating costs by over 25% in its first three years of operation while simultaneously increasing run-times and availability. Efforts continue to further reduce operating and maintenance costs, including programs for operator cross-training, materials improvements in critical high cost areas such as refractory and char filtration, and general continuous improvement in O&M practices. Reduction of non-fuel operating and maintenance costs below 5% of total installed cost is achievable for the next generation of solid-fueled gasification facilities.

Fuel Flexibility

Fuel flexibility may be the area of the greatest opportunity for solid fuel gasification. Operation with low cost feedstocks such as petroleum coke greatly enhances project economics. Incorporation of other fuels, especially those opportunity fuels with tipping fees (negative cost feedstocks), can turn the fuel component of the cost of power into a revenue stream. Global Energy is pursuing identification and characterization of these feedstocks and is planning future testing at Wabash River as well as incorporating them into the design basis of other projects. There are particular opportunities at Wabash River to utilize the second stage of the gasifier for operation of these opportunity fuels, especially biomass feedstocks that may not be available in great quantities in the local area.

Product Flexibility

As well as reducing the costs of the inputs to the plant (fuel and operating costs), there are opportunities to increase the revenue streams for the plant products. Global Energy is studying coproduction economics for Wabash River and future plants as part of NETL's Early Entrance Coproduction Plant program (in Cooperative Agreement DE-FC26-99FT40659). Other opportunities have been identified to improve the value of the slag by-product. The already extremely pure sulfur by-product is a bulk commodity with pricing determined by market availability.

Further research aimed at reducing the costs of sulfur removal equipment to achieve the high syngas purities needed for utilization in hydrogen production and gas-to-liquid process units will be needed.

9.3 Outlook

All of the aforementioned areas, in essence, are economic barriers and not "technology" barriers. The novel technologies developed at Wabash River are advanced enough to allow the Project to operate commercially. Reducing the cost of operation is part of the economic improvement needed. The primary barriers are economics and competition with new gas-fired peaking plants and cogens, and in the off-peak hours, with existing baseload facilities.

Market penetration for gasification technologies is obviously picking up speed, especially for solid fuel applications and particularly in the United States (as discussed in Section 8.0). Gasification-produced megawatts will increase tenfold from 1992 to 2002, based on plants already in operation or construction. The current opportunities are not primarily in power generation, however. They are in co-production facilities, especially those able to use opportunity fuels. Looking at low-cost feedstocks and high-value products stretches both ends of the economic equation. These facilities seem to be primarily in the refining sector, and it is expected that most of the next generation of solid fuel gasification plants will be inside the fences of refineries, as opposed to the entire first generation of greenfield and repowering applications for power generation facilities.

The power market, especially for repowering of older plants as demonstrated at Wabash River, still exists, but may not be significant in the near term. Eventually, increasing fuel costs, the aging of the coal-fired fleet, and increasingly stringent environmental regulation will create opportunities in the generation system.

The reality of a market for power generated from gasification facilities is coming into focus however, and the beginnings of this market are here with us now. The demonstration plants are operating on utility grids in the U.S. and in other countries. The need and interest in co-production plants for refining and chemical industry complexes is increasing steadily. It appears that the current reality has caught up with the Department of Energy's vision of the future as it was fifteen years ago when the Clean Coal Technology Program was initiated. The "Vision 21" Program, looking ahead ten to fifteen years from now, certainly holds the same promise for

additional increases in the utilization of gasification technologies to meet market demands for environmentally superior ways to generate power and produce other products.

APPENDIX A

Glossary of Acronyms, Abbreviations and Symbols

APPENDIX A

Glossary of Acronyms, Abbreviations and Symbols

ºC	-	degrees Celsius
ºF	-	degrees Fahrenheit
$	-	dollars (U.S.)
$MM	-	millions of dollars (U.S.)
$/kw	-	dollars per kilowatt
$/ton	-	dollars per ton
$/MMBtu	-	dollars per million British thermal unit
%	-	percent
AC	-	alternating current
acfm	-	actual cubic feet per minute
ACGIH	-	American Conference for Governmental Industrial Hygiene
AGR	-	acid gas removal
AIChE	-	American Institute of Chemical Engineers
AP-42	-	EPA publication "Compilation of Air Pollutant Emission Factors"
API	-	American Petroleum Institute
Ar	-	argon, elemental
As	-	arsenic, elemental
ASU	-	air separation unit
atm	-	atmosphere(s)
avg.	-	average
ASTM	-	American Society for Testing and Materials
bbl	-	barrels
bbl/d	-	barrels per day
BFW	-	boiler feed water
Btu	-	British thermal unit(s)
Btu/kWh	-	British thermal unit(s) per killowatt-hour
C	-	carbon, elemental
CA	-	cooperative agreement
CAA	-	Clean Air Act
CAAA	-	Clean Air Act Amendments of 1990
CCT	-	clean coal technology
CCT IV	-	fourth CCT program solicitation
CCT Program	-	Clean Coal Technology Demonstration Program
CEM	-	continuous emission monitoring
CFD	-	computational flow dynamics
CFR	-	Code of Federal Regulations

CH_4	-	methane
CO	-	carbon monoxide
CO_2	-	carbon dioxide
Corp.	-	corporation
COS	-	carbonyl sulfide
CSS	-	chloride scrubbing system
CT	-	combustion turbine
CWR	-	cooling water return
CWS	-	cooling water supply
d	-	day(s)
dB	-	decibel(s)
DBB	-	double block and bleed
DCS	-	distributed control system
DCSS	-	dry char slipstream system
DOE	-	Department of Energy
DP	-	differential pressure
dscf	-	dry standard cubic feet
dscfm	-	dry standard cubic feet per minute
EIA	-	Energy Information Administration
E-Gas^{TM}	-	current name of the gasification technology at the WRCGRP
EPA	-	Environmental Protection Agency
EPC	-	engineer(ing)-procure(ment)-construct(ion)
EMP	-	environmental monitoring plan
EPRI	-	Electric Power Research Institute
FAA	-	Federal Aviation Authority
FCC	-	Federal Communications Commission
ft, ft^2, ft^3	-	foot (feet), square feet, cubic feet
gal	-	gallon(s)
g	-	gram(s)
gr	-	grain(s)
gr/dscf	-	grain(s) per dry standard cubic foot
GE	-	General Electric
GSA	-	Gasification Services Agreement
GT	-	gas turbine
GTC	-	Gasification Technologies Council
H	-	hydrogen, elemental
H_2O	-	water
H_2S	-	hydrogen sulfide
H_2SO_4	-	sulfuric acid
HHV	-	high heating value

HP	-	high pressure
hr	-	hour(s)
HRSG	-	heat recovery steam generator
HSS	-	heat stable salts
HTHRU	-	high temperature heat recovery unit
IDEM	-	Indiana Department of Environmental Management
IGCC	-	integrated gasification combined-cycle
IPP	-	independent power producer
IRR	-	internal rate of return
ISEP	-	ion separation unit
IURC	-	Indiana Utility Regulatory Commission
JV	-	joint venture
in, in^2, in^3	-	inch(es), square inches, cubic inches
kg	-	kilogram(s)
kW	-	kilowatt(s)
kWh	-	kilowatt-hour(s)
l	-	liter
lb or lbs	-	pound or pounds
lb/hr	-	pound(s) per hour
lb/MMBtu	-	pound(s) per million British thermal units of heat input
LGTI	-	Louisiana Gasification Technology, Inc.
LIN	-	liquid nitrogen
LLC	-	limited liability corporation
LOX	-	liquid oxygen
LP	-	low pressure
Ltd.	-	limited
LTHRU	-	low temperature heat recovery unit
MCC	-	motor control center
MDEA	-	metyldiethanolamine
mg	-	milligram
min	-	minutes(s)
Mlb	-	thousand pound(s)
MMBtu	-	million British thermal units
MMSCF	-	millions of standard cubic feet
mo	-	month(s)
MP	-	medium pressure
MSW	-	municipal solid waste
MW	-	megawatt(s)
MWe	-	megawatt(s)-electric
MWh	-	megawatt-hour(s)

N_2	-	nitrogen, gas
ND	-	non-detectable
NEPA	-	National Environmental Policy Act
NETL	-	National Energy Technology Laboratory
NH_3	-	ammonia
NiO	-	nickel oxide
No.	-	number
NO, NO_2, NOx	-	nitric oxide, nitrogen dioxide, nitrogen oxides
NP	-	non-proprietary
NPDES	-	National Pollutant Discharge Elimination System
NPV_{10}	-	net present value at 10% discount rate
NSPS	-	new source performance standards
O&M	-	operating and maintenance
OEM	-	original equipment manufacturer
O_2	-	oxygen, gas
P	-	proprietary
P&ID's	-	piping and instrument drawings
pH	-	the negative logarithm of the effective hydrogen ion concentration
PLC		programmable logic controller
PM_{10}	-	particulate matter less than 10 microns in diameter
PM	-	preventative maintenance
PMP	-	project management plan
PON	-	program opportunity notice
ppm	-	part(s) per million (mass)
ppmdv	-	part(s) per million by volume, dry basis
ppmv	-	part(s) per million by volume
PSD	-	prevention of significant deterioration
psia	-	pound(s) per square inch absolute
psig	-	pound(s) per square inch gauge
PT	-	potential transformer (electrical)
RDF	-	refuse derived fuel
rpm	-	revolutions per minute
RTD	-	resistance temperature device
S	-	sulfur, elemental
scf	-	standard cubic feet at 60°F and 1 atmosphere
scfm	-	standard cubic feet per minute at 60°F and 1 atmosphere
SO_2	-	sulfur dioxide
SO_3	-	sulfur trioxide
SO_X	-	combined sulfur dioxide and sulfur trioxide
SRU	-	sulfur recovery unit

TCLP	-	toxicity characteristics leaching procedure
TLV	-	threshold limit value
U.S.	-	United States (of America)
USA	-	United States of America
V_2O_5	-	vanadium pentoxide
VOC	-	volatile organic compounds
vol	-	volume
WRCGRP	-	Wabash River Coal Gasification Repowering Project
WREL	-	Wabash River Energy Ltd.
wt	-	weight
Y2K	-	the year 2000
yr	-	year(s)
ZLD	-	zero liquid discharge

APPENDIX B

Monthly Plant Production Data

Table B.1: 1996 Monthly Plant Performance Data

	JAN	FEB	MAR	APR	MAY	JUN	JUL	AUG	SEP	OCT	NOV	DEC	TOTAL
PERFORMANCE DATA													
Coal Gas Efficiency (%)	62.3	68.5	69.5	68.2	70.4	69.4	72.7	59.0	72.3	69.7	0.0	71.1	69.5
Gasifier on Coal (Hours)	151.18	307.00	188.55	165.29	21.80	33.91	293.61	16.40	13.16	339.43	0.00	371.87	1902
PRODUCTION DATA													
Syngas on Spec (MMBtu)	218,821	457,091	297,791	225,047	14,058	42,352	446,012	18,679	15,374	480,007	0	554,451	2,769,683
160# Steam (Mlbs)	68,219	134,650	86,241	64,198	9,945	13,496	125,484	8,053	6,075	144,265	0	159,997	820,623
Sulfur (Mlbs)	624	902	607	585	200	65	881	42	39	1,183	0	1,451	6,579
Slag, Moisture Free (Tons)	1,818	3,829	2,454	1,914	197	356	3,687	160	139	4,006	0	4,728	23,288
DELIVERED PRODUCTION													
Actual Syngas Delivered (MMBtu)	166,496	377,304	249,531	183,002	7,142	13,564	418,404	9,609	12,126	404,169	0	455,139	2,296,486
160# Steam (Mlbs)	62,446	129,039	76,841	56,905	4,147	7,706	119,883	4,121	4,359	126,221	0	135,218	726,886
MATERIAL/ENERGY USED													
Coal, Moisture Free (Tons)	12,709	26,757	17,154	13,375	1,375	2,494	25,765	1,120	974	27,998	0	33,036	162,756
Coal (MMBtu)	316,502	671,868	430,748	337,043	34,849	63,312	645,172	28,053	24,388	701,116	0	827,292	4,080,141
Intermediate Pressure Steam (Mlbs)	13,303	14,353	13,669	14,139	6,458	5,437	12,524	7,750	5,685	15,001	2,714	13,244	124,277
Electrical Power, Total (MWh)	18,034	18,853	19,835	19,993	13,970	11,726	22,679	16,423	12,722	20,387	9,269	22,660	206,551
Oxygen, (Tons)	14,590	26,159	18,006	13,387	2,310	3,218	24,256	2,214	1,472	25,671	56	29,234	160,573
Fuel Gas (Mlbs)	2,658	2,617	3,631	2,167	1,973	1,390	1,902	1,111	488	1,760	48	826	20,571
PLANT EMISSION DATA													
Average Total Sulfur in Syngas (ppm)	116	216	170	176	60	97	183	103	13	24	0	102	134
Total SO2 Emissions (lbs)	87,830	153,881	66,666	65,968	3,081	7,933	119,105	8,325	1,773	21,547	0	65,298	601,436
SO2, (Total Plant lbs/MMBtu of Coal Feed)	0.278	0.229	0.155	0.196	0.089	0.125	0.185	0.297	0.073	0.031	0	0.079	0.147
POWER PLANT PRODUCTION DATA													
Combustion Turbine Generator (MWh)	24,864	53,310	31,530	22,548	1,967	4,540	49,192	5,255	1,685	49,545	10	56,755	301,201
Steam Turbine Generator (MWh)	12,580	24,932	15,872	11,639	1,035	3,603	21,417	1,885	796	26,219	0	28,740	148,718
Total Gross Generation (MWh)	37,444	78,242	47,402	34,187	3,002	8,143	70,609	7,140	2,481	75,764	10	85,495	449,919
Total Syngas Generation (MWh)	19,500	52,200	28,650	20,334	783	1,363	47,313	565	11,746	44,528	0	51,182	278,164

Table B.2: 1997 Monthly Plant Performance Data

	JAN	FEB	MAR	APR	MAY	JUN	JUL	AUG	SEP	OCT	NOV	DEC	TOTAL
PERFORMANCE DATA													
Coal Gas Efficiency	65.8	71.2	72.0	72.9	71.8	75.0	72.4	71.8	73.2	74.9	71.4	71.0	72.1
Gasifier on Coal (Hours)	151.15	390.94	393.38	197.85	311.66	262.33	376.68	585.20	413.39	175.54	370.89	255.95	3885
PRODUCTION DATA													
Syngas on Spec (MMBtu)	198,422	632,819	663,669	328,199	507,537	426,957	611,182	959,155	626,837	258,603	572,826	427,660	6,213,864
1600# Steam (Mlbs)	62,445	169,460	169,047	84,045	140,811	121,732	176,070	259,997	182,914	77,590	156,637	119,472	1,720,219
Sulfur (Mlbs)	802	1,254	1,820	1,113	1,271	1,112	1,662	2,550	1,496	923	2,117	1,053	17,171
Slag, Moisture Free (Mlbs)	1,786	5,316	5,473	2,743	4,316	3,475	5,096	7,952	5,276	2,188	4,337	3,461	51,418
DELIVERED PRODUCTION													
Actual Syngas Delivered (MMBtu)	160,091	607,035	633,747	317,328	478,091	406,566	582,472	936,909	600,515	249,046	529,647	415,397	5,926,845
1600# Steam (Mlbs)	53,890	166,676	163,834	82,511	134,278	113,389	174,728	257,471	171,188	70,245	141,423	111,597	1,641,232
MATERIAL/ENERGY USED													
Coal, Moisture Free (Tons)	12,467	37,148	38,241	19,165	30,157	24,280	35,603	55,558	38,864	15,292	30,315	24,185	359,294
Coal (MMBtu)	312,687	930,250	957,632	479,940	755,179	608,023	875,848	1,366,762	906,861	376,201	745,770	594,960	8,910,111
Intermediate Pressure Steam (Mlbs)	13,085	16,773	15,821	7,187	13,331	12,340	13,574	16,616	13,770	8,670	14,513	8,961	154,641
Electrical Power, Total (MWh)	21,373	21,446	23,717	12,483	19,312	22,499	25,472	25,472	24,651	22,413	23,283	18,016	260,138
Oxygen, (Tons)	11,699	33,339	34,010	16,868	27,599	22,316	32,373	49,358	33,236	14,375	31,360	22,069	328,601
Fuel Gas (Mlbs)	1,863	1,392	1,951	871	2,036	1,723	1,500	1,419	1,056	1,476	3,030	813	19,129
PLANT EMISSION DATA													
Average Total Sulfur in Syngas (ppm)	103	123	110	104	88	121	151	170	225	201	66	79	132
Total SO2 Emissions (lbs)	35,181	91,311	79,597	34,067	48,355	60,750	130,027	211,044	123,108	61,806	73,649	69,657	1,018,555
SO2, (Total Plant lbs/MMBtu of Coal Feed)	0.113	0.098	0.083	0.071	0.084	0.100	0.148	0.154	0.136	0.164	0.099	0.117	0.114
POWER PLANT PRODUCTION DATA													
Combustion Turbine Generator (MWh)	27,105	72,754	77,071	39,760	59,145	53,235	70,501	108,784	70,671	30,132	66,918	48,978	725,054
Steam Turbine Generator (MWh)	12,282	36,843	37,175	18,501	30,174	26,409	35,912	55,249	37,350	16,012	31,639	24,277	361,823
Total Gross Generation (MWh)	39,387	109,597	114,246	58,261	89,319	79,644	106,413	164,033	108,021	46,144	98,577	73,255	1,086,897
Total Syngas Generation (MWh)	27,830	104,020	107,338	53,755	82,957	67,781	99,191	106,411	101,672	44,042	75,931	69,437	940,365

Table B.3: 1998 Monthly Plant Performance Data

	JAN	FEB	MAR	APR	MAY	JUN	JUL	AUG	SEP	OCT	NOV	DEC	TOTAL
PERFORMANCE DATA													
Coal Gas Efficiency	73.9	74.1	69.3	69.97	70.61	72.31	71.7	72.1	72	73	73	72.8	71.8
Gasifier on Coal (Hours)	386.21	250.32	686.48	624.42	503.17	336.78	272.72	455.69	307.05	667.99	694.96	93.18	5279
PRODUCTION DATA													
Syngas on Spec (MM/Btu)	638,885	415,197	1,163,287	1,057,500	823,729	553,705	447,787	719,963	480,371	1,155,078	1,215,447	160,062	8,831,011
1600# Steam (Mlbs)	182,479	105,969	301,501	265,524	197,757	133,854	119,336	184,989	120,700	279,726	284,480	38,088	2,214,393
Sulfur (Mlbs)	1,724	1,187	3,303	2,978	2,271	1,533	1,297	2,016	1,412	3,359	3,436	386	24,902
Slag, Moisture Free (Tons)	5,122	3,333	9,368	8,535	6,648	4,347	3,555	5,700	3,863	8,979	9,548	1,230	70,228
DELIVERED PRODUCTION													
Actual Syngas Delivered (MMBtu)	619,052	405,804	1,128,977	1,028,532	803,478	537,357	433,914	674,473	473,241	1,134,140	1,182,704	156,746	8,578,518
1600# Steam (Mlbs)	164,968	105,428	301,009	265,134	197,741	132,120	119,162	183,203	119,334	277,888	282,319	36,504	2,184,810
MATERIAL/ENERGY USED													
Coal, Moisture Free (Tons)	35,794	23,292	65,454	59,637	46,446	30,375	24,842	39,835	26,994	62,738	66,742	8,592	490,741
Coal (MMBtu)	880,539	572,989	1,610,214	1,467,091	1,142,605	747,240	611,131	979,954	664,061	1,543,395	1,641,138	211,371	12,071,728
Intermediate Pressure Steam (Mlbs)	15,042	9,746	8,173	17,719	12,651	11,244	9,620	13,516	9,602	16,471	16,320	6,317	146,421
Electrical Power, Total (MWh)	22,244	20,344	26,022	25,194	21,232	20,144	21,109	23,516	21,202	25,699	25,343	16,743	268,792
Oxygen, (Tons)	33,138	21,125	59,198	53,277	40,570	27,725	22,580	36,459	24,787	56,270	59,076	8,117	442,322
Fuel Gas (Mlbs)	2,155	845	1,149	1,021	302	1,306	461	1,635	786	620	396	847	11,522
PLANT EMISSION DATA													
Average Total Sulfur in Syngas (ppm)	88	79	76	87	101	146	96	97	87	116	133	132	102
Total SO2 Emissions (lbs)	96,328	60,666	161,714	111,780	85,575	89,772	43,412	86,815	60,082	130,601	148,110	23,733	1,098,588
SO2, (Total Plant lbs/MMBtu of Coal Feed)	0.109	0.106	0.100	0.076	0.075	0.120	0.071	0.089	0.090	0.085	0.090	0.112	0.091
POWER PLANT PRODUCTION DATA													
Combustion Turbine Generator (MWh)	73,935	47,905	132,768	120,078	89,708	68,394	59,783	84,634	54,447	132,709	136,975	21,778	1,023,114
Steam Turbine Generator (MWh)	35,432	22,691	66,187	59,214	44,832	33,400	26,908	37,109	25,722	63,601	65,738	9,681	490,515
Total Gross Generation (MWh)	109,367	70,596	198,955	179,292	134,540	101,794	86,691	121,743	80,169	196,310	202,713	31,459	1,513,629
Total Syngas Generation (MWh)	102,870	67,160	189,659	172,069	134,495	89,119	72,179	106,622	74,794	190,146	202,083	27,959	1,429,155

Table B.4: 1999 Monthly Plant Performance Data

	JAN	FEB	MAR	APR	MAY	JUN	JUL	AUG	SEP	OCT	NOV	DEC	TOTAL
PERFORMANCE DATA													
Coal Gas Efficiency	73.7	73.7	72	0	0	76.6	73.3	72.4	73.8	74.5	73.1	73.9	73.8
Gasifier on Coal (Hours)	354.47	465.25	31.08	0.00	0.00	205.56	430.54	497.21	720.00	384.65	4.79	401.96	3496
PRODUCTION DATA													
Syngas on Spec (MMBtu)	589,864	797,006	45,732	0	0	333,039	682,052	825,483	1,204,573	637,766	6,139	681,498	5,813,151
1600# Steam (Mlbs)	145,995	188,306	12,614	0	0	93,412	188,199	213,422	296,513	156,114	2,383	182,951	1,480,908
Sulfur (Mlbs)	1,607	2,480	104	0	0	963	1,845	2,532	3,875	1,792	78	1,837	17,113
Slag, Moisture Free (Tons)	4,608	6,208	365	0	0	2,571	5,500	6,494	9,292	4,895	52	5,234	45,216
DELIVERED PRODUCTION													
Actual Syngas Delivered (MMBtu)	571,433	779,107	39,409	0	0	322,896	643,640	794,628	1,152,059	602,461	4,931	649,920	5,560,483
1600# Steam (Mlbs)	142,712	187,821	11,147	0	0	86,876	175,828	204,157	287,675	151,060	1,624	182,337	1,431,236
MATERIAL/ENERGY USED													
Coal, Moisture Free (Tons)	32,196	43,375	2,550	0	0	17,962	38,429	45,378	64,924	34,203	384	36,569	315,951
Coal (MMBtu)	792,049	1,067,051	62,731	0	0	441,884	945,382	1,116,314	1,597,184	841,423	8,947	899,623	7,772,568
Intermediate Pressure Steam (Mlbs)	13,562	14,162	5,900	1,837	1,135	6,725	10,784	11,625	11,436	9,867	2,934	13,422	103,390
Electrical Power, Total (MWh)	23,650	22,213	17,626	965	843	19,125	19,863	24,093	25,821	20,767	12,304	24,110	211,369
Oxygen, (Tons)	29,480	39,527	2,588	0	0	16,742	35,135	41,266	59,282	31,766	6	34,143	289,935
Fuel Gas (Mlbs)	1,044	826	602	0	0	662	856	553	81	1,048	0	1,146	6,818
PLANT EMISSION DATA													
Average Total Sulfur in Syngas (ppm)	113	128	115	0	0	118	116	109	120	104	110	105	115
Total SO2 Emissions (lbs)	66,468	92,642	8,622	0	0	38,712	83,915	88,589	129,020	68,813	12,460	70,967	660,207
SO2, (Total Plant lbs/MMBtu of Coal Feed)	0.084	0.087	0.137	0	0	0.088	0.089	0.079	0.081	0.082	1.393	0.079	0.085
POWER PLANT PRODUCTION DATA													
Combustion Turbine Generator (MWh)	70,524	91,395	5,624	0	0	36,586	92,524	102,116	132,522	71,184	1,058	77,677	681,210
Steam Turbine Generator (MWh)	33,794	43,734	2,770	0	0	19,443	43,282	47,047	62,623	31,702	483	37,765	322,643
Total Gross Generation (MWh)	104,318	135,129	8,394	0	0	56,029	135,806	149,163	195,145	102,886	1,541	115,442	1,003,853
Total Syngas Generation (MWh)	94,300	129,048	6,466	0	0	54,062	112,726	136,610	195,028	92,250	956	110,507	931,943